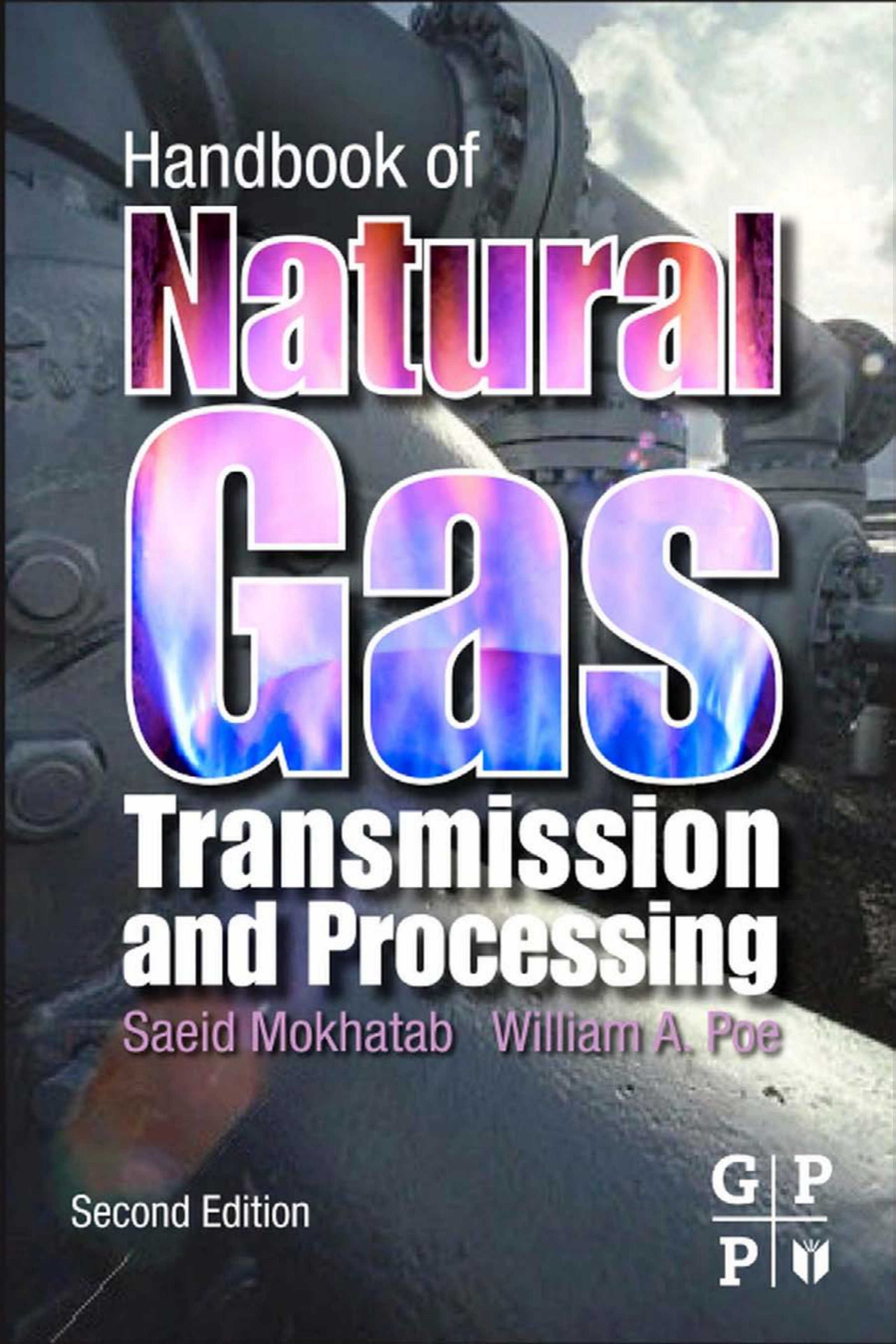

Handbook of Natural Gas Transmission and Processing

Saeid Mokhatab William A. Poe

Second Edition

ENDORSEMENTS FOR THE SECOND EDITION

"Today's natural gas industry is evolving and the projects are technically challenging. This book is an excellent reference for all professionals, engineers and scientists working in the midstream and downstream sectors of natural gas industry. It fills a considerable void."

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"This well-balanced handbook is the only book of its kind, covering all aspects of natural gas transmission and processing in more details. I believe it will serve as a valuable desk reference for practicing gas engineers and technologists, and as a text for graduate students in the gas engineering curriculum."

J.C. Kuo, Senior Advisor of Gas Team, Process Technology Unit, Chevron ETC, USA

"This high quality, comprehensive book gives an accurate picture of where the natural gas transmission and processing industry stands today, as well as indicating some relatively new technologies that could become important in the future. I recommend this book for any professional gas processing engineer and technologist."

David Messersmith, Bechtel Fellow and Manager of LNG Technology and Services Group, Bechtel, USA

"This is a valuable handbook to both the experienced engineer and the graduate just commencing in natural gas engineering. It provides practical advice for design and operation based on sound engineering principles and established techniques as well as introducing process solutions based on new and emerging technologies."

Adrian Finn, Technology Manager, Costain Energy & Process, UK

"This is a wide ranging book providing the reader with much more information than that necessary to just support the engineering of a gas plant. Having introduced the fundamental science of natural gas, it covers sources, products, transportation and economics of natural gas production, before covering all the unit operations involved in gas treatment to meet product specifications. Additional sections are included covering control systems, dynamic simulation, environmental aspects, maximizing gas plant profitability and gas plant project management, and these contribute to what is a holistic handbook that will educate all those who wish to learn about the subject."

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"This book is very well written with very good format to make it an excellent reference book for engineers and other professionals working in the natural gas transmission and processing industry. The book addresses basic fundamentals, basic concepts in natural gas transmission and processing, new technologies as well as challenges ahead. It presents field and literature data in a form that can be easily followed and understood by engineers involved in this industry. In addition, the book presents a chapter on the theory of natural gas pricing making it a more comprehensive resource for all engineers as well as textbook for students entering in the natural gas engineering field."

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"This book addresses the advanced technologies, new issues and challenges related to natural gas transmission and processing, which have not been addressed in depth in any existing books. The format of the book makes it a particularly valuable reference work for all engineers in the natural gas business as well as a textbook for students in petroleum and chemical engineering curricula and in the training departments of a large group of companies."

Dr. Douglas G. Elliot, President and Chief Operating Officer, IPSI LLC, USA

"This is an excellent document that combines all facets of engineering within the natural gas business from source to supply. As well as supporting our engineering design industry, this work will offer a comprehensive education to our process engineers of the future."

Dr. John H. Hargreaves, Chief Process Engineer, Wood Group PSN, UK

"Though there are many publications and books that deal with the gas processing industry, this book presents comprehensive coverage. The book also covers some topics rarely discussed and hard to find in the literature and also has valuable references for those interested in further reading of the various topics. As a result, this book is a valuable reference, recommended for any practitioner in the gas processing industry."

Dr. Wayne Monnery, President, Chem-Pet Process Technology Ltd, Canada

"This book comes at a critical time when many nations are shifting to an increasingly higher percentage of natural gas use within their range of energy sources. Giving the reader a comprehensive insight into the natural gas transmission and processing industry, this book will prove invaluable in orienting the newcomer and extending the scope of understanding of the veteran."

Dr. Lorenzo Micucci, Manager of Technology and R&D, Siirtec Nigi, Italy

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"For natural gas professionals working with transmission and processing, this book provides an excellent reference that covers a wide range of topics. From a brief chapter on basic fundamentals to advanced topics such as plant project management and environmental aspects of the business, this book provides a comprehensive look at post-production aspects of the natural gas industry. I believe this book will be a great companion for active natural gas professionals."

Dr. Shahab D. Mohaghegh, Professor of Petroleum & Natural Gas Engineering, West Virginia University, USA

"Natural gas will not only continue to be a significant part of the energy resources for many years to come, but its use will also be increasing. This book documents the latest technology in all areas of natural gas engineering. It is a useful book for every engineer, scientist, and researcher who has ever faced the challenge of investigating natural gas gathering, processing, and transport. I recommend it highly, as a reference and textbook."

Dr. Brian F. Towler, Professor of Chemical and Petroleum Engineering, University of Wyoming, USA

Handbook of NATURAL GAS TRANSMISSION AND PROCESSING

SECOND EDITION

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DEDICATION

This book is dedicated to all of those persons who have come before us as family, scientists, and engineers and from whom we have derived comfort and knowledge.

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FOREWORD

Natural gas is an environmentally friendly, clean-burning fuel, offering important environmental benefits compared to other fossil fuels. It is also a remarkably safe source of energy when transported, stored, and used. Therefore, gas usage surely will increase worldwide in the coming years. The emergence of natural gas in a global market underscores the importance of transmission and processing. In this regard, it is fitting that technical books covering these subjects become required reading for technical specialists. While several books exist for natural gas upstream engineering, the literature has lacked a comprehensive book that captures the midstream and downstream sectors of the natural-gas value chain. This book provides well-organized, theoretical, and technical background knowledge on natural gas transmission and processing; and covers subject areas missed by other books in these areas. Because the authors have either personal or supervisory experience in the design and troubleshooting of natural gas processing plants throughout the world, the text has attempted to concentrate upon what they perceive to be modern and common practices. To keep things interesting for the experts, they have included the most current developments reported in the latest published works, although the selection of some material also reflects their own research involvement in problems related to natural gas transmission and processing. Therefore, this book is a major contribution to the professional literature in this very crucial area of the emerging energy scene, and will serve as a valuable desk reference for scientists, researchers, and engineers working in the midstream and downstream sectors of natural gas utilization. Literature references for in-depth study enhance the reference aspect of this comprehensive work. The organization of materials also permits flexibility in designing courses in this field for senior university students in chemical/petroleum engineering curricula.

Professor Kenneth R. Hall, PhD
Senior Associate Dean for Research
College of Engineering
Jack E. & Frances Brown Chair and Professor
Artie McFerrin Department of Chemical Engineering
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PREFACE TO THE SECOND EDITION

Many interesting and exciting developments in the natural gas industry have occurred since the publication of the first edition in 2006 that make it appropriate to revise portions of the book and to present additional material of interest to our professional readers in order to preserve its quality. This realization has led us to embark on the new revision of our work to present an updated and comprehensive description of all major aspects of natural gas transmission and processing. The major guideline for the work has been to update and add information where we felt it was relevant. This way, we hope that the main text shall remain understandable and useful even for newcomers in the field, while at the same time provide access to a more complete discussion for the more experienced practitioners.

The preparation of a book that covers such a broad subject requires many sources of information. As with any text, many more names were involved in its creation than those mentioned in the book. We gratefully acknowledge our indebtedness to all of the individuals who contributed to the ideas that form the background of this text. An invaluable contribution to this edition is the insight provided to the authors by experts in certain applications over the past 6 years. Special thanks are due to friends and colleagues, who encouraged, assessed, and made this book possible. Among them are Dr. Louis Mattar and Dr. Mehran Pooladi-Darvish of Fekete Associates Inc., Canada, who prepared the section on "Natural Gas Exploration and Production" in Chapter 1. We also appreciate Dr. Vivek Chandra for preparing the section on "Dynamics of Global Gas Business" in that chapter. We express our heartfelt thanks to Mr. Ed Wichert, President of Sogapro Engineering Ltd – Canada, who provided valuable input in preparing the sulfur recovery and handling chapter. We greatly acknowledge the co-authorship of Dr. Rainer Kurz, Manager of Systems Analysis and Field Testing of Solar Turbines Inc – USA, in updating the natural gas compression chapter. Our special thanks and gratitude go to Mr. Sidney P. Santos, Senior Consultant of Petrobras Gas & Energy – Brazil, for his excellent comments and additions on the sales gas transmission chapter. We also express our sincere thanks to Professor Kenneth R. Hall of Texas A&M University who agreed to write its foreword. Finally, we appreciate the editorial staff members of Elsevier; Kenneth P. McCombs, Jill Leonard, and Jeff Freeland; who have been an excellent source of strong support during the preparation and publication of this book.

The authors know that this work is not perfect. But we also know that this handbook had to be written to promote the sharing of knowledge and experience. We were inspired to accelerate the delivery of information contained in this handbook to the community of past, present, and future practitioners who have and will continue to contribute to a better world for all through the effective use of natural gas. Our greatest hope is that we have given those who will follow us in future editions of this handbook sound basic material to work with.

Saeid Mokhatab
William A. Poe

AUTHORS' BIOGRAPHICAL SKETCHES

Saeid Mokhatab is one of the most instantly recognizable names in the natural gas industry with dedicated technical expertise in the midstream sector. He has been on the international advisory board of a number of petroleum/energy consulting firms around the world and has been actively involved in several large-scale gas-field development projects, concentrating on design, precommissioning, and startup of processing plants. He has over 200 technical publications to his credit, including two reference books. He founded the Elsevier peer-reviewed *Journal of Natural Gas Science and Engineering,* has held editorial positions in many prestigious journals/book publishing companies for the hydrocarbon processing industry, and served on various IGU, SPE and ASME technical committees and conferences worldwide. As a result of his work, he has received a number of international awards, including the Albert Einstein Gold Medal of Honor and the Chilingar Gold Medal of Scientists Without Borders; and is listed in several international biographical directories.

William A. Poe is a business consultant for Invensys Operations Management in Houston, TX, USA. He has over 30 years of experience in chemical and gas processing plants design and operations and has developed business relationships with several domestic and international gas processors, chemical plants, and gas transmission pipelines owners. His main area of focus the past two decades has been automation, multivariable predictive control (MPC), advanced process control (APC), and optimization design and implementation for the natural gas processing and treating, sulfur recovery, ammonia synthesis, methanol synthesis, liquefied natural gas, and power generation industries. Working under contracts to the Gas Research Institute, he developed new multivariable control applications in the natural gas, ammonia, and methanol industries. Poe developed APC and Optimization Master Plans for international companies such as Saudi Aramco, Statoil, and PDVSA, as well as automation and advanced process control feasibility studies for about 100 natural gas processing, power generation, ammonia, and methanol plants worldwide. He spent over a decade in natural gas processing operations and engineering experience with Shell Oil Company. Poe has project management experience with Shell Oil and GE Continental Controls with multimillion-dollar projects. He assumed the technical leadership of Continental Controls, a privately held company, through acquisition by GE. After joining GE as part of the Continental Controls acquisition, he became vice president of this division of GE, which included direction of product development, projects, technical sales support, and customer service. He has authored over 40 papers, presented at numerous technical conferences, received the GE Innovators Award in 1999, and attained the Invensys Circle of Excellence in 2011.

CHAPTER 1

Natural Gas Fundamentals

1.1 INTRODUCTION

Natural gas is the most energy-efficient fossil fuel; it offers important energy-saving benefits when it is used instead of oil or coal. Although the primary use of natural gas is as a fuel, it is also a source of hydrocarbons for petrochemical feedstocks and a major source of elemental sulfur, an important industrial chemical. Its popularity as an energy source is expected to grow substantially in the future because natural gas can help achieve two important energy goals for the twenty-first century: providing the sustainable energy supplies and services needed for social and economic development and reducing adverse impacts on global climate and the environment in general. Natural gas consumption and trade have been growing steadily over the past two decades, and natural gas has strengthened its position in the world energy mix. Although natural gas demand declined in 2009, as a result of the economic slowdown, it is expected to resume growth in both emerging and traditional markets in the coming decades. Such increase in the near future will be driven because of additional demand in current uses, primarily power generation. There is yet little overlap between the use of natural gas and oil in all large markets. However, there are certain moves in the horizon, including the electrifying of transportation, that will push natural gas use to even higher levels.

This chapter gives the reader an introduction to natural gas by describing the origin and composition of natural gas, gas sources, phase behavior and properties, and transportation methods.

1.2 NATURAL GAS HISTORY

The discovery of natural gas dates from ancient times in the Middle East. It was considered by ancients to be a supernatural manifestation. Noticed only when ignited, it appeared as a mysterious fire bursting from fissures in the ground. Natural gas seeps were discovered in Iran between 6000 and 2000 B.C. The practical use of natural gas dates back to the Chinese of 2,500 years ago, who used bamboo pipes to collect it from natural seeps and convey it to gas-fired evaporators, where it was used to boil ocean water for the salt. Apparently, natural gas was unknown in Europe until its discovery in England in 1659. However, since manufactured gas (coal gas) was already commercially available, natural gas remained unpopular. In 1815, natural gas was discovered in the United States during the digging of a salt brine well in Charleston, West Virginia. In 1821, an American gunsmith named William Aaron Hart drilled the first natural gas well in the United States. It was covered with a large barrel, and the gas was directed through wooden pipes that were replaced a few years later with lead pipe. One of the earliest attempts to harness it for economic use occurred in 1824 in Fredonia, New York, and led to the formation of the first natural gas company in the United States, the Fredonia Gas Light Company, in 1858.

The nineteenth century is considered as the starting point of the gas industry. In the early 1900s, huge amounts of natural gas were found in Texas and Oklahoma, and in the 1920s modern seamless steel pipe was introduced. The strength of this new pipe, which could be welded into long sections, allowed gas to be carried under higher pressures and, thus, in greater quantities. As the technology to create seamless steel pipe and related equipment advanced, the size and length of pipelines increased, as did the volumes of gas that could be transported easily and safely over many miles. The first natural gas pipeline longer than 200 miles was built in 1925, from Louisiana to Texas.

Steady growth in the use of gas marked the early and mid-twentieth century. However, it was the shortages of crude oil in the late 1960s and early 1970s that forced major industrial nations to seek energy alternatives. Since those events, gas has become a central fossil fuel energy source. Today, natural gas has become extremely important as a concentrated, clean fuel for home heating and cooking and electrical power generation, and is sought after almost as much as oil.

1.3 NATURAL GAS ORIGIN AND SOURCES

There are different theories as to the origins of fossil fuels. The most widely accepted theory of the origin of natural gas assumes that natural gas hydrocarbons come from organic matter (the remains of land and aquatic plants, animals and micro-organisms) that was trapped within sediments as they were deposited and transformed over long periods of time into their present form. Two main mechanisms, namely, biogenic and thermogenic, are responsible for the degradation of fossil organic material in sediments (Rojey et al., 1997). Biogenic gas is formed at shallow depths and low temperatures due to the action of bacteria on the organic debris accumulating in the sediments. In contrast, thermogenic gas is formed at deeper depths by degradation of organic matter, called kerogen, accumulated in fine-grained sediments, especially clays and shales. This degradation occurs through the combined effects of temperature and pressure. Thermogenic gas is believed to be produced through two mechanisms, namely, direct thermal cracking of sedimentary organic matter and secondary thermal cracking of oil that is formed in the first stage. The former is called primary thermogenic gas, which co-exists with oil, while the latter is called secondary thermogenic gas, which co-exists with insoluble solid matter, called pyrobitumen. Both mechanisms involve thermal cracking with some degree of sustained pressure, mainly through the weight of the sedimentary formation. Little information is available on the time required to generate thermogenic gas other than the general assumption that it is a long time.

Natural gas comes from both "conventional" (easier to produce) and "unconventional" (more difficult to produce) geological formations. Conventional gas is typically "free gas" trapped in multiple, relatively small, porous zones in various naturally occurring rock formations such as carbonates, sandstones, and siltstones. Conventional natural gas generally occurs in deep reservoirs, either associated with crude oil (associated gas[1]) or in reservoirs that contain little or no crude oil (nonassociated gas[2]). Natural gas from coal (also known as coal-bed methane, CBM), tight gas sands, gas shales, geopressurized aquifers, and gas hydrates[3] are often referred to as unconventional gas resources. The common characteristic of the different types of unconventional gas resources is that they contain large quantities of natural gas, but it is usually more difficult to produce this gas as compared to conventional reservoir rocks. New technologies are continually being developed to allow more accurate estimations of the amount of gas in these unconventional reservoirs and to stimulate these rocks to produce the gas.

1.4 NATURAL GAS COMPOSITION AND CLASSIFICATION

Natural gas is a complex mixture of hydrocarbon and nonhydrocarbon constituents and exists as a gas under atmospheric conditions. Virtually hundreds of different compounds may be present in natural gas in varying amounts.

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[1] Associated gas is produced with the oil and separated at the casinghead or wellhead. Gas produced in this fashion is also referred to as casinghead gas or oil well gas.

[2] Nonassociated gas is sometimes referred to as gas-well gas.

[3] Gas hydrates are ice-like structures of water and gas located under the permafrost.

Even two wells producing from the same reservoir may produce gases of different composition as the reservoir is depleted.

While natural gas is formed primarily of methane (CH_4), it can also include significant quantities of ethane (C_2H_6), propane (C_3H_8), butane (C_4H_{10}), and pentane (C_5H_{12}), as well as traces of hexane (C_6H_{14}) and heavier hydrocarbons. Many natural gases often contain nitrogen (N_2), carbon dioxide (CO_2), hydrogen sulfide (H_2S), and other sulfur components such as mercaptans (R-SH),[4] carbonyl sulfide (COS), and carbon disulfide (CS_2). Trace quantities of argon, hydrogen, and helium may also be present. Trace quantities of metallic substances are known to exist in natural gases including arsenic, selenium, mercury, and uranium.

According to the proportion of hydrocarbons heavier than methane, different types of natural gas (dry, wet, and condensate) can be considered. Natural gas is considered "dry" when it is almost pure methane, having had most of the other commonly associated hydrocarbons removed. When other hydrocarbons are present, the natural gas is "wet," where it forms a liquid phase during production at surface conditions. "Condensate" gases have a high content of hydrocarbon liquids and form a liquid phase in the reservoir during production, during the depletion process.

Natural gases commonly are classified according to their liquids content as either lean or rich and according to the sulfur content as either sweet or sour.

The lean and rich terms refer to the amount of potentially recoverable liquids. The term usually applies to ethane and heavier components but sometimes applies instead to propane and heavier components (if ethane is not regarded as a valuable liquid component). To quantify the liquids content of a natural gas mixture, the industry uses GPM, or gallons of liquids recoverable per 1,000 standard cubic feet (Mscf) of gas. Lean natural gas has less than 2.5 GPM at 60°F. Moderately rich natural gas has between 2.5 and 5 GPM, and very rich natural gas has greater than 5 GPM (Ewan et al., 1975).

The sweet and sour terms refer to the H_2S content. Strictly speaking, "sweet" and "sour" refer to both acid gases (H_2S and CO_2) but are usually applied to H_2S alone. A sweet gas contains negligible amounts of H_2S, whereas a sour gas has unacceptable quantities of H_2S. The terms are relative, but generally, sweet means the gas contains less than 4 ppmv of H_2S. Carbon dioxide can be tolerated to much higher levels, say 3–4 mole%, as long as the heating value of the sales gas is satisfactory.

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[4] R signifies an organic group such as a methyl, ethyl, propyl, or other group.

1.5 NATURAL GAS PHASE BEHAVIOR

Natural gas is a naturally occurring hydrocarbon mixture that is found underground and at elevated conditions of pressure and temperature. Therefore, there is an essential need to know a priori how the gas fluid will behave under a wide range of pressure and temperature conditions, particularly in terms of its volumetric and thermophysical properties that are required in simulating reservoirs, evaluating reserves, forecasting production, designing production facilities and designing gathering and transportation systems. In fact, an accurate knowledge of hydrocarbon fluid phase behavior is crucial in designing and operating the gas-engineering processes efficiently and optimally. This means having advanced predictive tools for the characterization of hydrocarbon phase behavior with the highest accuracy possible is the key to mastering the economics of natural gas systems.

The natural gas phase behavior is a plot of pressure versus temperature that determines whether the natural gas stream at a given pressure and temperature consists of a single gas phase or two phases, gas and liquid. The phase behavior for natural gas with a given composition is typically displayed on a phase diagram, an example of which is shown in Figure 1-1. The left-hand side of the curve is the bubble point line and divides the single-phase liquid region from the two-phase gas-liquid region. The right-hand side of the curve is the dew point line and divides the two-phase gas-liquid region

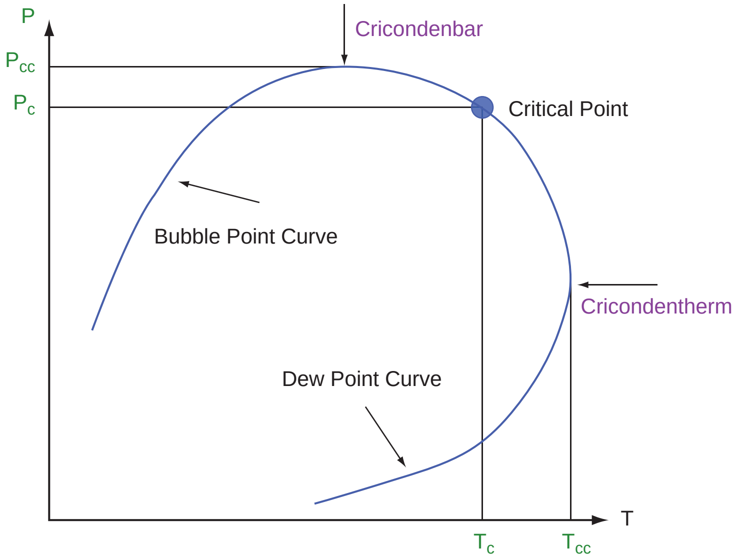

Figure 1-1 Pressure-Temperature diagram for a typical natural gas mixture.

and the single-phase gas region. The bubble point and dew point lines intersect at the critical point, where the distinction between gas and liquid properties disappears. The maximum pressure at which liquids can form is called the cricondenbar (P_{cc}), and the maximum temperature at which liquids can form is called the cricondentherm (T_{cc}). However, there is something very interesting going on within the region $T_c < T < T_{cc}$, where we will be moving from a 0% liquid to another 0% liquid condition (both on the dew point curve) in an isothermal compression. This different behavior of a vapor under compression is called retrograde (contrary to expectation) condensation. It is also important to see that a similar behavior is to be expected within the region $P_c < P < P_{cc}$. In this case, we talk about retrograde vaporization since we will be moving from a 100% liquid to another 100% liquid condition (both on the bubble point curve) in an isobaric heating.

The natural gas phase behavior is a function of the composition of the gas mixture and is strongly influenced by the concentration of the heavier hydrocarbons, especially hexane plus.[5] The presence of heavier hydrocarbons will increase the phase envelope and failure to include them in a phase calculation will underpredict the phase envelope. There is also an essential need for proper characterizing the heavy ends. In fact, although some different fluid descriptions match to some extent the behavior of the reservoir fluids at reservoir conditions, they exhibit larger variations once surface simulators are used and the fluids are subjected to process conditions.

1.6 NATURAL GAS PROPERTIES

1.6.1 Chemical and Physical Properties

Natural gas is colorless, odorless, tasteless, shapeless, and lighter than air (see Table 1-1). The natural gas after appropriate treatment for acid gas reduction and after moisture and hydrocarbon dew point adjustment would then be sold within prescribed limits of pressure, heating value, and possibly *Wobbe Index* (often referred to as the *Wobbe Number*).

One of the principal uses of natural gas is as a fuel, and consequently, pipeline gas is normally bought and sold on the basis of its heating value that can be produced by burning the gas. The heating value of natural gas is variable and depends on its accumulations, which are influenced by the amount and types of gases they contain.[6] The gas industry always uses the gross

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[5] Hexane plus or C_6^+ is a terminology used in natural gas engineering which indicates a grouping of heavier hydrocarbons including the hexane and heavier hydrocarbons such as heptane, octane, nonane, etc.

[6] Since the heat energy of the natural gas is related to the relative proportion of "lighter" methane versus "heavier" ethane, propane, butane, pentane, and other components, heat energy is not a constant value between different natural gas sources (Chandra, 2006).

Table 1-1 Properties of Natural Gas

Properties	**Value**
Relative molar mass	17–20
Carbon content, weight %	73.3
Hydrogen content, weight %	23.9
Oxygen content, weight %	0.4
Hydrogen/carbon atomic ratio	3.0–4.0
Relative density, 15°C	0.72–0.81
Boiling point, °C	–162
Autoignition temperature, °C	540–560
Octane number	120–130
Methane number	69–99
Stoichiometric air/fuel ratio, weight	17.2
Vapor flammability limits, volume %	5–15
Flammability limits	0.7–2.1
Lower heating/calorific value, MJ/kg	38–50
Stoichiometric lower heating value, MJ/kg	2.75
Methane concentration, volume %	80–99
Ethane concentration, volume %	2.7–4.6
Nitrogen concentration, volume %	0.1–15
Carbon dioxide concentration, volume %	1–5
Sulfur concentration, weight % ppm	<5
Specific CO_2 formation, g/MJ	38–50

heating value (frequently called higher heating value) in custody transfer. Obviously, the numerical difference between the two net and gross heating values is the heat of condensation of water at the specified conditions. Heating values for custody transfer are determined either by direct measurement, in which calorimetry is used, or by computation of the value on the basis of gas analysis (Gas Processors Association, 1996). The heating value of natural gas is measured in British thermal unit (Btu). A British thermal unit is the energy required to raise the temperature of 1 pound of water by 1°F. For larger industrial customers, the abbreviations MBtu (thousand Btu) or MMBtu (million Btu) are more commonly used. Since meters measure volume and not energy content, a conversion factor of 1,000 Btu/ft^3 is commonly used by gas companies to convert the volume of gas used to its heat equivalent, and thus calculate the actual energy use.

The Wobbe Index (defined as the gross heating value of the gas divided by the square root of the specific gravity) gives a measure of the heat input to an appliance through a given aperture at a given gas pressure. Using this as a vertical coordinate and the flame speed factor as the horizontal coordinate, a combustion diagram can be constructed for an appliance, or a whole range of appliances, with the aid of appropriate test gases. This diagram shows the area within which variations in the Wobbe Index and flame speed factor of gases may occur for the given range of appliances without resulting in either incomplete combustion, flame lift, or the lighting back of pre-aerated flames. This method of prediction of combustion characteristics is not sufficiently accurate to eliminate entirely the need for the practical testing of new gases.

Since natural gas as delivered to pipelines has practically no odor, the addition of an odorant is required by most regulations in order that the presence of the gas can be detected readily in case of accidents and leaks. This odorization is provided by the addition of trace amounts of some organic sulfur compounds to the gas before it reaches the consumer. The sulfur compound, a chemical odorant (a *mercaptan* also called a *thiol* with the general formula R-SH and the odor of rotten eggs) is added to natural gas so that it can be smelled if there is a gas leak. The standard requirement is that a user will be able to detect the presence of the gas by odor when the concentration reaches 1% of gas in air. Since the lower limit of flammability of natural gas is approximately 5%, this requirement is equivalent to one-fifth the lower limit of flammability. The combustion of these trace amounts of odorant does not create any serious problems of sulfur content or toxicity.

In the following section we discuss important gas properties including specific gravity, compressibility factor, formation volume factor, density, isothermal compressibility, and viscosity.

1.6.1.1 Gas-Specific Gravity

Specific gravity of a natural gas is defined as the ratio of gas density to the density of air, both defined at the same pressure and temperature. These densities are usually defined at standard conditions (14.7 psia and 60°F). Therefore, specific gravity of gas is defined as

$$\gamma_g = \frac{M}{M_{air}} \tag{1-1}$$

where M is the molecular weight of gas, and M_{air} is the molecular weight of air that is equal to 29. Once we can calculate the value of molecular weight of mixture, we can calculate the specific gravity of mixture. For a gas mixture, we can calculate the molecular weight as

$$M = \sum_{i=1}^{n} y_i M_i \tag{1-2}$$

where M_i is the molecular weight of component i, y_i is the mole fraction of component i, and n is the total number of components.

Various gas properties, including the molecular weights for pure components, are given in Table 1-2.

1.6.1.2 Gas Compressibility Factor

The volume of a real gas is usually less than what the volume of an ideal gas would be, and hence a real gas is said to be super-compressible. The ratio of the real volume to the ideal volume, which is a measure of the amount the gas deviates from perfect behavior, is called the super-compressibility factor, sometimes shortened to the compressibility factor. It is also called the gas deviation factor, and given the symbol Z. The gas deviation factor is, by definition, the ratio of the volume actually occupied by a gas at a given pressure and temperature to the volume it would occupy if it behaved ideally.

The real gas equation of state is then written as

$$PV = ZnRT \tag{1-3}$$

where P is the pressure, V is the volume, T is the absolute temperature, Z is the compressibility factor, n is the number of kilo-moles of the gas, and R is the gas constant.

The gas deviation factor, Z, is close to 1 at low pressures and high temperatures, which means the gas behaves as an ideal gas at these conditions. At standard or atmospheric conditions, the gas Z factor is always approximately 1.

Empirical correlations for Z factor for natural gases were developed before the advent of digital computers. Although their use is in decline, they can still be used for rapid estimates of the Z factor. Chart lookup is another means of determining Z factor of natural gas mixtures. These methods are invariably based on some type of corresponding states development. According to the theory of corresponding states, substances at corresponding states will exhibit the same behavior. The theory of corresponding states dictates that the Z factor can be uniquely defined as a function of reduced pressure and reduced temperature. The reduced pressure and reduced temperature are defined as

$$P_r = \frac{P}{P_c} \quad \text{and} \quad T_r = \frac{T}{T_c} \tag{1-4}$$

where P_r and T_r are reduced pressure and reduced temperature, respectively; and P_c and T_c are critical pressure and critical temperature of the gas,

Table 1-2 Physical Constants for Pure Components (Whitson and Brule, 2000)

Compound	Formula	Molecular Weight	Critical Constants			
			P_c, Psia [*kPa*]	T_c, °R [°K]	V_c, ft^3/lb [m^3/kg]	Z_c
Methane	CH_4	16.043	667.8 [4,604]	343 [190.6]	0.0991 [0.0062]	0.2884
Ethane	C_2H_4	30.070	707.8 [4,880]	549.8 [305.4]	0.0788 [0.00492]	0.2843
Propane	C_3H_8	44.097	616.3 [4,249]	665.7 [369.8]	0.0737 [0.0046]	0.2804
n-Butane	C_4H_{10}	58.124	550.7 [3,797]	765.3 [425.2]	0.0702 [0.00438]	0.2736
Isobutane	C_4H_{10}	58.124	529.1 [3,648]	734.7 [408.2]	0.0724 [0.00452]	0.2824
n-Pentane	C_5H_{12}	72.151	488.6 [3,369]	845.4 [469.7]	0.0675 [0.00422]	0.2623
Isopentane	C_5H_{12}	72.151	490.4 [3,381]	828.8 [460.4]	0.0679 [0.00424]	0.2701
Neopentane	C_5H_{12}	72.151	464.0 [3,199]	781.11	0.0674 [0.00421]	0.2537
n-Hexane	C_6H_{14}	86.178	436.9 [3,012]	913.4 [507.4]	0.0688 [0.0043]	0.2643
n-Heptane	C_7H_{16}	100.205	396.8 [2,736]	972.5 [540.3]	0.0691 [0.00432]	0.2633
n-Octane	C_8H_{18}	114.232	360.6 [2,486]	1,023.9 [568.8]	0.0690 [0.0043]	0.2587
n-Nonane	C_9H_{20}	128.30	332 [2,289]	1,070.3 [594.6]	0.0684 [0.00427]	0.2536
n-Decane	$C_{10}H_{22}$	142.30	304 [2,096]	1,111.8 [617.7]	0.0679 [0.00424]	0.2462
Ethylene	C_2H_4	28.054	729.8 [5,032]	508.6 [282.6]	0.0737 [0.0046]	0.2765
Propene	C_3H_6	42.081	669.[4,613]	656.9 [364.9]	0.0689 [0.0043]	0.2752
Acetylene	C_2H_2	26.038	890.4 [6,139]	555.3 [308.5]	0.0695 [0.00434]	0.2704
Carbon Dioxide	CO_2	44.010	1071. [7,382]	547.6 [304.2]	0.0342 [0.00214]	0.2742
Hydrogen Sulfide	H_2S	34.076	1306 [9,005]	672.4 [373.6]	0.0459 [0.00287]	0.2831
Sulfur Dioxide	SO_2	64.059	1145. [7,894]	775.5 [430.8]	0.0306 [0.00191]	0.2697
Nitrogen	N_2	28.013	493 [3,399]	227.3 [126.3]	0.0514 [0.00321]	0.2916
Water	H_2O	18.015	3208.[22,105]	1,165.0 [647.2]	0.0500 [0.00312]	0.2350

respectively. The values of critical pressure and critical temperature can be estimated from the following equations if the composition of the gas and the critical properties of the individual components are known.

$$P_c = \sum_{i}^{n} P_{Ci}\, y_i \text{ and } T_c = \sum_{i}^{n} T_{ci}\, y_i \tag{1-5}$$

where P_{ci} and T_{ci} are the critical pressure and critical temperature of component i, respectively; and y_i is the mole fraction of component i.

Once critical properties of the mixture are calculated as stated in Equation 1-5, we can use Equation 1-4 to calculate the reduced properties of the mixture.

The values of critical pressure and critical temperature can be estimated from its specific gravity if the composition of the gas and the critical properties of the individual components are not known. The method uses a correlation to estimate pseudocritical temperature and pseudocritical pressure values from the specific gravity. There are several different correlations available. The most common is the one proposed by Sutton (1985), which is based on the basis of 264 different gas samples. Sutton (1985) used regression analysis on the raw data to obtain the following second order fits for the pseudocritical properties.

$$P_{pc} = 756.8 - 131.07\gamma_g - 3.6\gamma_g^2 \tag{1-6}$$

$$T_{pc} = 169.2 + 349.5\gamma_g - 74.0\gamma_g^2 \tag{1-7}$$

These equations are valid over the range of specific gas gravities with which Sutton (1985) worked $0.57 < \gamma_g < 1.68$.

The most commonly used method to estimate the Z factor is the chart provided by Standing and Katz (1942). The Z factor chart is shown in Figure 1-2. The chart covers the range of reduced pressure from 0 to 15, and the range of reduced temperature from 1.05 to 3. This chart is generally reliable for sweet natural gases with minor amounts of nonhydrocarbons. It was developed using data for binary mixtures of methane with propane, ethane, butane, and natural gases having a wide range of composition. None of the gas mixtures had molecular weight in excess of 40. For low molecular weight gases, it was found that the Z factor estimated from Standing and Katz (1942) chart has error in the order of 2% to 3%. However, for gas mixtures whose components differ greatly in molecular weight from 40, this chart provides inaccurate Z factors (Elsharkawy et al., 2001).

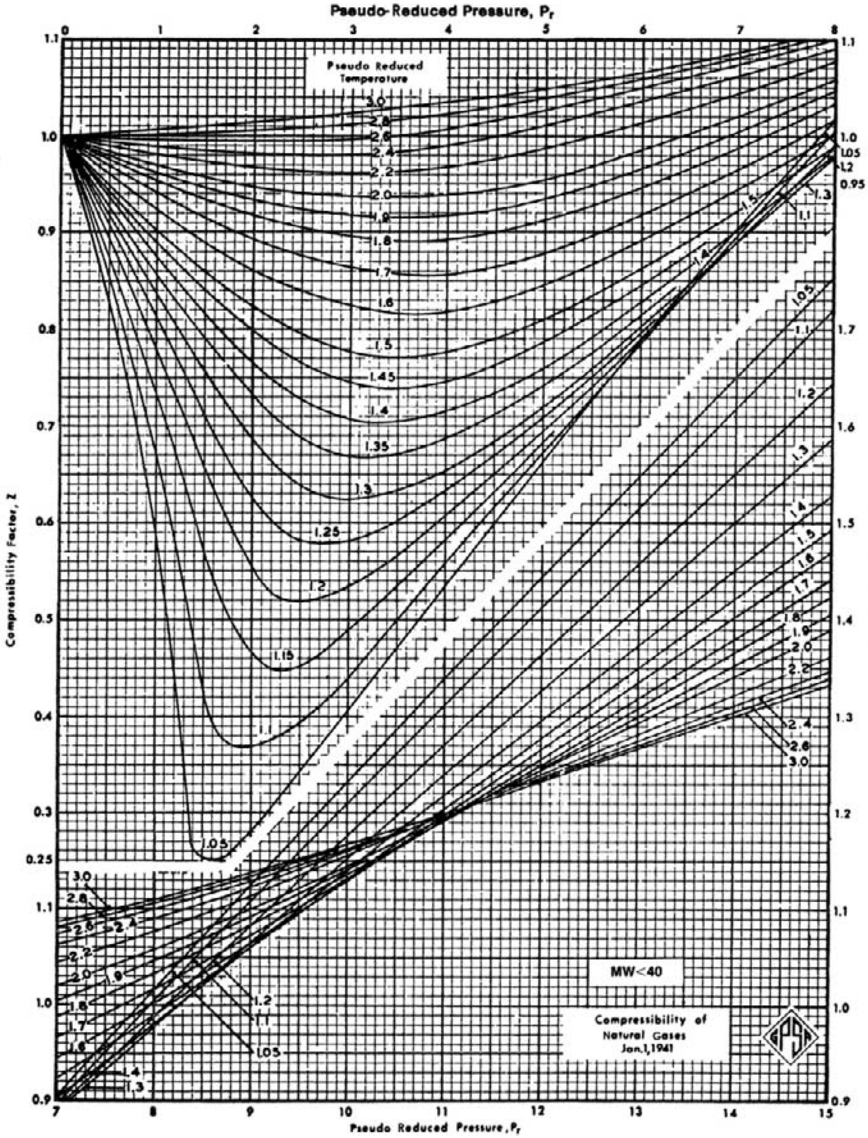

Figure 1-2 Compressibility of natural gases as a function of reduced pressure and temperature (Standing and Katz, 1942).

Wichert and Aziz (1972) developed a correlation to account for inaccuracies in the Standing and Katz chart when the gas contains significant fractions of acid gases, specifically carbon dioxide (CO_2) and hydrogen sulfide (H_2S). The Wichert and Aziz (1972) correlation modifies the values of the pseudocritical temperature and pressure of the gas. Once the modified pseudocritical properties are obtained, they are used to calculate pseudoreduced properties and the Z factor is determined from Figure 1-2. The Wichert and Aziz (1972) correlation first calculates a deviation parameter ε:

$$\varepsilon = 120(A^{0.9} - A^{1.6}) + 15(B^{0.5} - B^4) \tag{1-8}$$

where A is the sum of the mole fractions of CO_2 and H_2S in the gas mixture, and B is the mole fraction of H_2S in the gas mixture.

Then ε is used to determine the modified pseudocritical properties as follows:

$$T'_{pc} = T_{pc} - \varepsilon \tag{1-9}$$

$$P'_{pc} = \frac{P_{pc} T'_{pc}}{[T_{pc} - B(1 - B)\varepsilon]} \tag{1-10}$$

The correlation is applicable to concentrations of $CO_2 < 54.4$ mole% and $H_2S < 73.8$ mole%. Wichert and Aziz (1972) found their correlation to have an average absolute error of 0.97% over the following ranges of data: $154 \text{ psia} < P < 7026 \text{ psia}$ and $40°\text{F} < T < 300°\text{F}$.

Methods of direct calculation using corresponding states have also been developed, ranging from correlations of chart values to sophisticated equation sets based on theoretical developments (Elsharkawy et al., 2001; Bahadori et al., 2007; Heidaryan et al., 2010). However, the use of equations of state (EOS) to determine Z factors has grown in popularity as computing capabilities have improved. Equations of state represent the most complex method of calculating Z factor, but also the most accurate.

1.6.1.3 Gas Formation Volume Factor

The formation volume factor for gas is defined as the ratio of volume of one mole of gas at reservoir conditions to the volume of one mole of gas at standard conditions (P_s and T_s). Using the real gas law and assuming that the Z factor at standard conditions is one, the equation for formation volume factor (B_g) can be written as

$$B_g = \frac{V_R}{V_s} = \frac{nZRT}{P} \frac{P_s}{nZ_sRT_s} = \frac{P_sZT}{T_sP} \tag{1-11}$$

When P_s is 1 atmosphere (14.6959 psia or 101.325 kPa) and T_s is 60°F (519.67°R or 288.71°K), this equation can be written in three well-known standard forms:

$$B_g = 0.0283 \frac{ZT}{P} \tag{1-11-1}$$

where B_g is in ft^3/scf, P is in psia, and T is in °R. Alternately,

$$B_g = 0.3507 \frac{ZT}{P} \tag{1-11-2}$$

where B_g is in m^3/Sm^3, P is in kPa, and T is in °K.

In some cases, it is more convenient to define the value of B_g in bbl/scf. The equation can be written as

$$B_g = 0.005 \frac{ZT}{P} \tag{1-11-3}$$

where T is in °R and P is in psia.

1.6.1.4 Gas Density

The gas density is defined as mass per unit volume and so can also be derived and calculated from the real gas law:

$$\rho_g = \frac{m}{V} = \frac{PM}{ZRT} \tag{1-12}$$

Knowing that molecular weight of gas is the product of specific gravity and molecular weight of air, and the value of R is 10.73 in field units [8.314 in SI units], we can write the equation for density as

$$\rho_g = 2.7 \frac{P\gamma_g}{ZT} \tag{1-12-1}$$

where ρ_g is in lbm/ft^3, P is in psia, and T is in °R. Alternately, we can write

$$\rho_g = 3.49 \frac{P\gamma_g}{ZT} \tag{1-12-2}$$

where ρ_g is in kg/m^3, P is in kPa, and T is in °K.

The density can also be written as

$$\rho_g = 0.0764 \frac{\gamma_g}{B_g} \tag{1-12-3}$$

where ρ_g is in lbm/ft^3, and B_g is in ft^3/scf. Alternately, we can write

$$\rho_g = 1.224 \frac{\gamma_g}{B_g} \tag{1-12-4}$$

where ρ_g is in kg/m^3, and B_g is in m^3/Sm^3.

1.6.1.5 Isothermal Compressibility of Gases

The isothermal gas compressibility, which is given the symbol c_g, is a useful concept that will be used extensively in determining the compressible properties of the reservoir. The isothermal compressibility is also called the bulk modulus of elasticity. Gas usually is the most compressible medium in the reservoir. However, care should be taken so that it is not confused with the gas deviation factor, Z, which is sometimes called the super-compressibility factor:

$$c_g = -\frac{1}{V_g}\left(\frac{\partial V_g}{\partial P}\right)_T \tag{1-13}$$

where V and P are volume and pressure, respectively, and T is the absolute temperature. For ideal gas, we can define the compressibility as

$$c_g = \frac{1}{P} \tag{1-13-1}$$

whereas, for nonideal gas, compressibility is defined as

$$c_g = \frac{1}{P} - \frac{1}{Z}\left(\frac{\partial Z}{\partial P}\right)_T \tag{1-13-2}$$

If the relationship between the Z factor and pressure at a given temperature is known, we can calculate the compressibility of gas. Since we already know the relationship between Z and P, we can numerically calculate the derivative and, hence, the value of the compressibility.

1.6.1.6 Gas Viscosity

A number of methods have been developed to calculate gas viscosity. The method of Lee et al. (1966) is a simple relation that gives quite accurate results for typical natural gas mixtures with low nonhydrocarbon content. The Lee et al. (1966) correlation was evaluated by Jeje and Mattar (2004), the form from which is shown here:

$$\mu_g = 1.10^{-4} K \exp\left[X\left(\frac{\rho_g}{62.4}\right)^{Y}\right] \tag{1-14}$$

where

$$K = \frac{\left(9.4 + 0.02M_g\right)T^{1.5}}{209 + 19M_g + T} \tag{1-14-1}$$

$$X = 3.5 + \frac{986}{T} + 0.01M_g \tag{1-14-2}$$

$$Y = 2.4 - 0.2X \tag{1-14-3}$$

In this expression, temperature is given in (°R), the density of the gas in lbm/ft^3 (calculated at the pressure and temperature of the system), and the resulting viscosity is expressed in centipoises (c_p).

The viscosity of gas mixtures at one atmosphere and reservoir temperature can be determined from the gas mixture composition:

$$\mu_{\mathrm{ga}} = \frac{\sum_{i=1}^{N} y_{\mathrm{i}} \mu_{\mathrm{i}} \sqrt{M_{\mathrm{gi}}}}{\sum_{i=1}^{N} y_{\mathrm{i}} \sqrt{M_{\mathrm{gi}}}} \tag{1-15}$$

where μ_{ga} is the viscosity of the gas mixture at the desired temperature and atmospheric pressure, y_{i} is the mole fraction of the *i*th component, μ_{ga} is the viscosity of the *i*th component of the gas mixture at the desired temperature and atmospheric pressure, M_{gi} is the molecular weight of the *i*th component of the gas mixture, and N is the number of components in the gas mixture.

This viscosity is then multiplied by the viscosity ratio to obtain the viscosity at reservoir temperature and pressure.

1.6.2 Thermodynamic Properties

The principles of thermodynamics find very wide application in predicting the properties of hydrocarbons. The properties of greatest interest are specific heats of natural gas and its Joule–Thomson coefficient.

1.6.2.1 Specific Heat

Specific heat is defined as the amount of heat required raising the temperature of a unit mass of a substance through unity. It is an intensive property of a substance. It can be measured at constant pressure (C_{p}) or at constant volume (C_{v}), resulting in two distinct specific heat values. In terms of basic thermodynamics quantities

$$C_{\mathrm{p}} = \left(\frac{\partial H}{\partial T}\right)_{\mathrm{p}} \tag{1-16}$$

$$C_{\mathrm{v}} = \left(\frac{\partial U}{\partial T}\right)_{\mathrm{v}} \tag{1-17}$$

where H is the molal enthalpy (Btu/lb mole), U is the molal internal energy (Btu/lb mole), C_{p} is the molal specific heat at constant pressure (Btu/lb mole-°R), and C_{v} is the molal specific heat at constant volume (Btu/lb mole-°R).

Both heat capacities are thermodynamically related. It can be proven that this relationship is controlled by the P-V-T behavior of the substances through the following relationship:

$$C_p - C_v = T\left(\frac{\partial V}{\partial T}\right)_p \left(\frac{\partial P}{\partial T}\right)_v \tag{1-18}$$

For ideal gases:

$$C_p - C_v = R \tag{1-19}$$

where R is the universal gas constant.

1.6.2.2 Joule–Thomson Coefficient

When a nonideal gas suddenly expands from a high pressure to a low pressure, there is often a temperature change. Note that this is far from a reversible effect. It is, however, an adiabatic effect due to the fact that the pressure change occurs too quickly for significant heat transfer to occur. Thermodynamically, the Joule–Thomson coefficient is defined as

$$\eta = \left(\frac{\partial T}{\partial P}\right)_H \tag{1-20}$$

Using thermodynamic relationships, alternative expressions can be written. For example, using the cycling rule, we may write

$$\eta = \frac{RT^2}{PC_p}\left(\frac{\partial Z}{\partial T}\right)_P \tag{1-21}$$

An interesting observation from the preceding expressions for the Joule–Thomson coefficient is that the Joule–Thomson coefficient of an ideal gas is identically equal to zero. However, real fluids take positive or negative Joule–Thomson values.

1.7 NATURAL GAS RESERVES

As natural gas is essentially irreplaceable, it is important to have an idea of how much natural gas is left in the ground to use. Measuring natural gas in the ground involves a great deal of inference and estimation. There is no single way that every industry player uses to quantify estimates of natural gas. Therefore, it is important to delve into the assumptions and methodology behind each study to gain a complete understanding of the estimate itself. With new technologies, these estimates are becoming more and more reliable; however, they are still subject to revision. The Energy Information Administration (referred to as the EIA) estimates world proved natural gas reserves to be around 5,210.8 Tcf.[7] The biggest reserves are located in the Middle East with 1,836.2 Tcf (or 34% of the world total), and Europe and the former U.S.S.R. with 2,158.7 Tcf (or 42% of total world reserves), but very significant reserves exist also in other regions. Note should be made that the most recent data on the world proved natural gas reserves can be found by visiting the EIA website (www.eia.doe.gov).

Natural gas reserves have grown rapidly in recent years. However, the share of natural gas reserves located onshore, easy to produce, and close to consumers is decreasing, and the share of reserves located offshore and in hostile environments appears to be increasing. The availability of these reserves for the end user is therefore hampered by production and transportation costs, which can exceed the price at which the gas can be sold. In such cases, innovative technical options are required for reducing the costs and providing new outlets for natural gas.

1.8 NATURAL GAS EXPLORATION AND PRODUCTION

There are four major activities involved when developing a natural gas resource. They are exploration, drilling, completion, and production. These activities vary, depending on whether the natural gas is "conventional" or "unconventional," and accordingly, the discussion will follow these two broad categories. In the discussion of production, the emphasis will be on the flow characteristics within the reservoir. However, as the gas needs to travel from the bottom of the well to the wellhead, the wellbore's delivery capacity will also be reviewed.

1.8.1 Conventional Gas

1.8.1.1 Exploration

Gas is contained in porous rocks at different depths in different locations. So, the first step is to locate where the gas is likely be found. This involves the two disciplines of geophysics and geology.

A seismic survey (geophysics) is conducted to determine the layering of the different rock strata. This helps determine where an accumulation of gas (or oil) may occur, and the potential extent of this accumulation. A seismic

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[7] Proved reserves are those estimated quantities of natural gas that can be commercially recoverable from a given date forward, from known reservoirs and under current economic conditions.

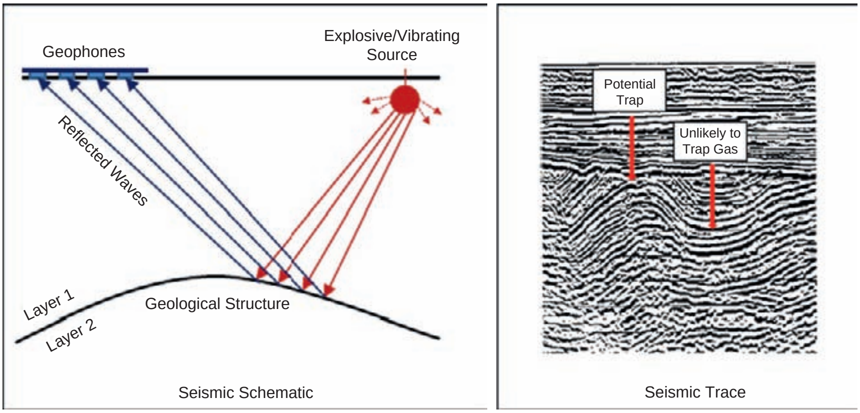

Figure 1-3 Seismic survey used in exploration.

survey consists of a wave generator (an explosive device or a vibrator) and a series of geophones, all located at the surface. The geophones detect reflections of the wave from the various layers. The geophysicist interprets the seismic data to generate a 2-D or 3-D picture of the layers in the earth (see Figure 1-3).

The seismic survey defines only the structure of the rock layers. It does not identify what is contained within the rock. However, it indicates where the highs and lows of the structure are or if there are any faults. Because of buoyancy effects, the hydrocarbons are more likely to be found in the highs of the structure. With assistance from the geoscientist, who brings regional experience and knowledge of the hydrocarbon bearing potential of these layers, a potential drilling location is determined. The chances of success vary significantly, depending on the degree of oil and gas development in the region. Where there is little knowledge (wildcat exploration), the chances of drilling a successful well are 15%–30%, but where there has already been a lot of successful drilling (development wells), the chance of success improves to 75%–100%.

1.8.1.2 Drilling

The mechanics of drilling are complex, and are significantly more so for an offshore well than for one on land. The well depth can vary from 1,000 feet to 20,000 feet. The well may encounter several layers of gas, oil, or water-bearing rocks, but usually, there is one particular layer that is the primary target of interest. While drilling, the wellbore is filled with drilling mud.

The hydrostatic pressure of the mud counterbalances the reservoir's pressure and avoids blowouts during drilling.

Once the intended depth is reached, a suite of electric logs is run from the bottom to the top of the well, to determine the porosity, shale content, fluid saturations, and thickness of the rock layers (see Figure 1-4). If a potentially gas-bearing reservoir is indicated, flow tests can be performed to estimate the permeability of the reservoir. If no productive reservoirs are found, the well is abandoned. If one or more productive layers is found, a steel pipe (casing) is placed in the wellbore and cemented. A "wellhead," which is an assembly of control valves, is placed on top of the well to control the flow from the well.

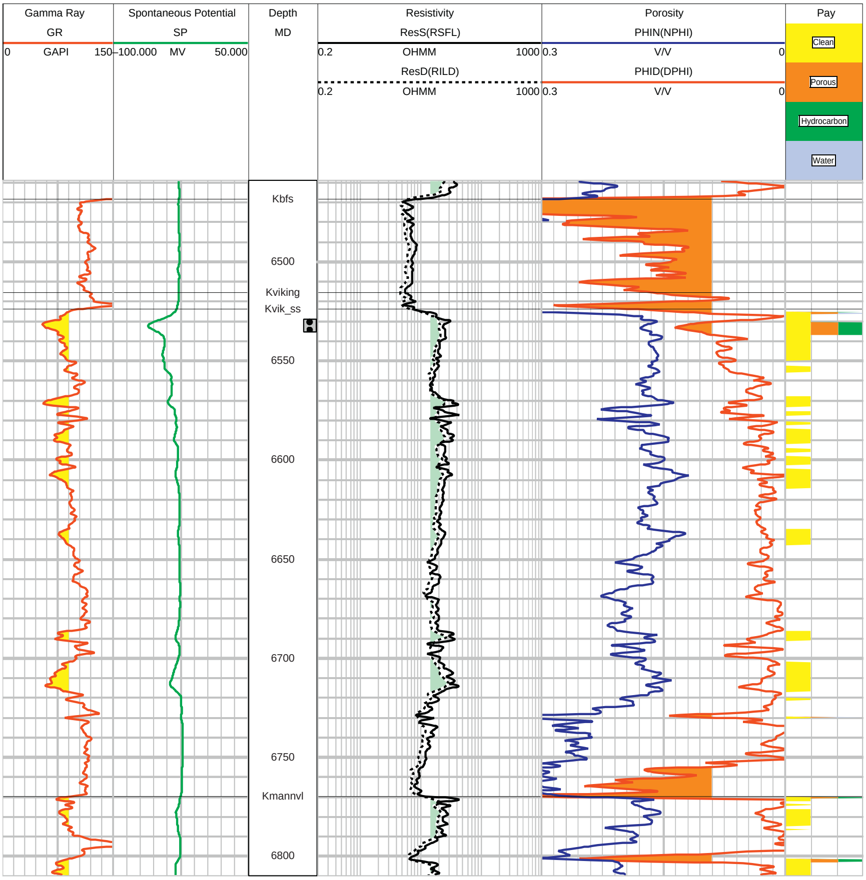

Figure 1-4 Electric logs used to determine presence of hydrocarbons.

1.8.1.3 Completion

The casing is perforated where the gas reservoir is known to exist. The perforations penetrate both the steel casing and the surrounding cement sheath, thus creating a flow path, allowing the gas to flow from the reservoir into the well (see Figure 1-5). When the reservoir rock has a low permeability, the deliverability of the well can be improved significantly by acidizing or hydraulic fracturing. Acidizing increases the permeability in the vicinity of the well. Fracturing increases the contact area between the well and the reservoir. To fracture a well, a liquid (usually gelled water with a proppant) is injected at high rates and pressures until the rock cracks. After the completion treatment, flow and shut-in tests are conducted to determine the deliverability potential of the well and to estimate the permeability of the reservoir and the effectiveness of the completion.

1.8.1.4 Production

The rate at which a gas well can produce depends, principally, on two fundamental equations, namely, Darcy's law and the Material Balance Equation (Dake, 1978). Darcy's law relates the flow rate in a reservoir to the driving force (pressure difference) and the resistance (permeability). Material Balance relates the decrease in the average reservoir pressure to the size of the reservoir and production volumes (hence to time). The principal factors that determine the deliverability of a well are, therefore, reservoir pressure, flowing pressure, permeability, completion effectiveness, Original-Gas-In-Place, and wellbore configuration.

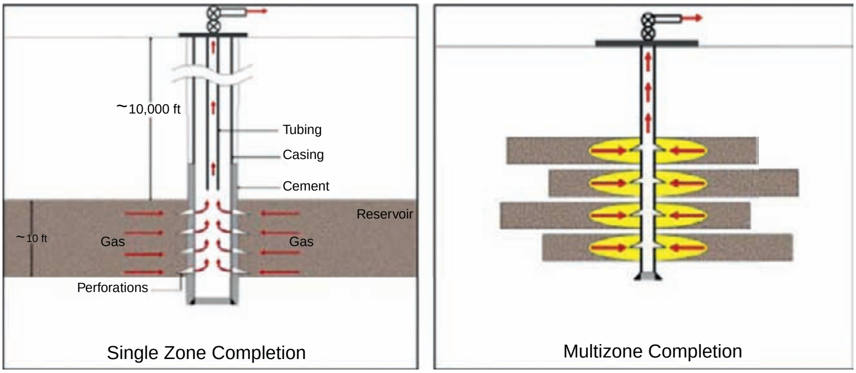

Figure 1-5 Different well completions.

For conventional gas wells, the deliverability equation can, for all practical purposes, be expressed as (ERCB, 1975)

$$q = C\left(P_R^2 - P_{wf}^2\right)^n \tag{1-22}$$

where q is the flow rate; C is a constant that encompasses permeability and completion effectiveness and other gas properties such as gravity, viscosity, and temperature; P_R is the reservoir pressure; P_{wf} is the pressure at the well; and n is an exponent (between 0.5 and 1) that accounts for Darcy and non-Darcy flow in the reservoir (commonly referred to as laminar and turbulent flow).

The constants C and n are determined from a flow test (often called Absolute Open Flow, AOF). The P_{wf} depends on surface facility restrictions, and usually reflects the pressure at which the gas is to be delivered to the transportation system or to the gas plant.

The material balance equation relates the reservoir pressure to the amount of gas produced:

$$P_R/Z = P_i/Z_i\left(1 - G_p/G\right) \tag{1-23}$$

where Z is the gas deviation factor (compressibility factor); G_p is the gas produced; and G is the Original-Gas-In-Place. Subscript "$_R$" represents current reservoir pressure, and subscript "$_i$" represents initial reservoir pressure.

The preceding two equations appear to be independent of time. However, that is not so. In Equation 1-23, G_p changes with time. Hence, the reservoir pressure changes with time. This, in turn, causes the flow rate calculated in Equation 1-22 to decrease as the reservoir depletes.

Combining Equations 1-22 and 1-23 and superimposing operating restrictions such as contract rates and facilities limitations (compression/pipeline/processing) results in a forecast of production rate versus time. Production stops when the flow rate becomes uneconomic. The total gas that will have been produced at the time of abandonment is called the Expected Ultimate Recovery (EUR).

The Original-Gas-In-Place (G) is determined from geological estimates of the areal extent and thickness of the reservoir, as well as its porosity and saturation, or from analysis of production data.

The recovery factor is the ratio of the Expected Ultimate Recovery to the Original-Gas-In-Place (EUR/G). It depends primarily on the reservoir pressure at the time of abandonment. The abandonment pressure itself is directly related to the delivery pressure and to the distance of transmission. As a consequence, the recovery factor can range from 50% (for remote locations) to 90% for wells adjacent to fully developed transportation systems.

1.8.2 Unconventional Gas

1.8.2.1 Exploration

Unlike with conventional gas, the location of unconventional resources is generally known, often as an indirect result of conventional mining or oil field activity. These unconventional gas resources are generally laterally extensive, and therefore, less exploration is required to locate them.

1.8.2.2 Drilling

Whereas most conventional gas wells are vertical, unconventional gas wells are either horizontal wells or multiple well pads (see Figure 1-6). There are two principal reasons for this:

(i) Either the productivity of the vertical well is too low; this is improved by using a horizontal well (single or multilateral); or

(ii) The area of the reservoir that is drained by a vertical well is so small that it would require a large number of wells to effectively drain the whole reservoir; this could be very intrusive on the surface land activity and is resolved by drilling multiple directional wells from a single pad, which significantly minimizes the surface footprint.

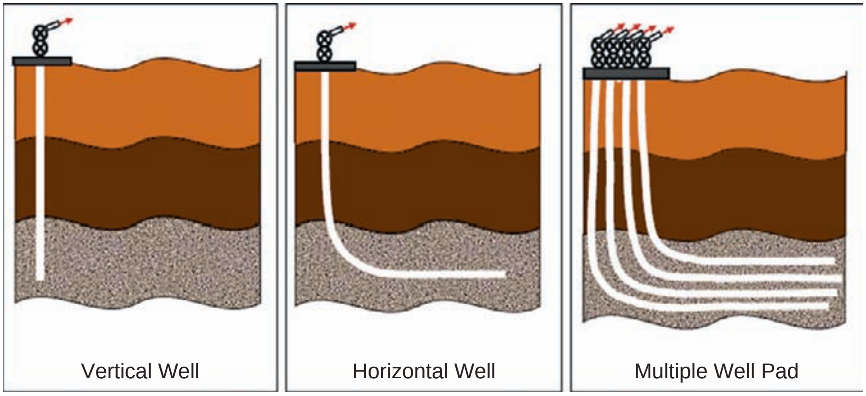

Figure 1-6 Different drilling patterns.

1.8.2.3 Completion

Typically, the permeability of unconventional gas reservoirs is extremely small. As a consequence, with conventional completions, the deliverability per well can be uneconomic, in spite of the known presence of large quantities of gas. The best way to improve productivity per well is to increase the area of contact between the well and the reservoir. This is what underlies the success of unconventional gas completions. The single most significant contribution to the commercialization of "shale" gas and "tight" gas has been the multiple-stage hydraulic fracturing of horizontal wells to create this increased contact area (see Figure 1-7).

1.8.2.4 Production

The deliverability of gas from unconventional gas wells is governed by a combination of Darcy's law and Material Balance (just as for conventional reservoirs). However, Equations 1-22 and 1-23 are not very useful because they are applicable only after the flow from the reservoir has stabilized. When a well is first opened to flow, a pressure transient travels through the reservoir (transient flow). It is only when this pressure transient reaches the boundaries of the reservoir (Boundary Dominated Flow) that the flow is considered to be stabilized and Equations 1-22 and 1-23 become applicable. The time to reach stabilization is a function of permeability, and in conventional reservoirs, this time is, for all practical purposes, relatively short (<3 months). However, in unconventional reservoirs, the permeability is so low that it can take many years before Equation 1-22 becomes applicable. Accordingly, more complex equations that are applicable during transient

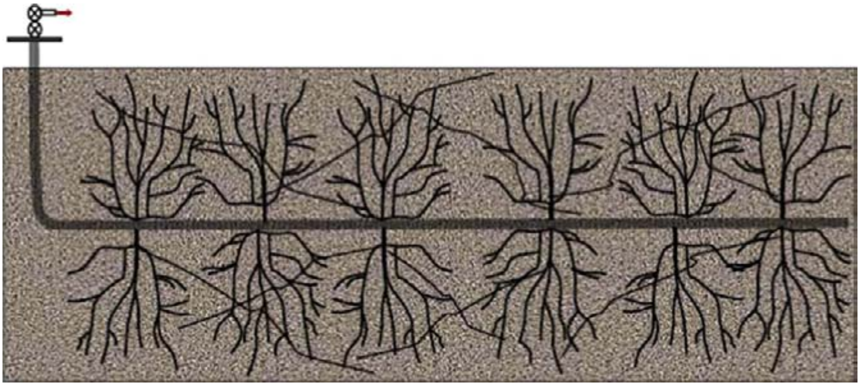

Figure 1-7 Very large contact area caused by multistage hydraulic fracturing.

flow are needed. Such equations will usually incorporate time explicitly and typically require computer programs. The form of such an equation is

$$q = f(p_i, p_{wf}, \text{permeability, well type, completion type, fluid properties, reservoir geometry and time})$$

Unconventional gas reservoirs are areally extensive, but because of the very low permeability, the effective drainage area per well is very limited. Accordingly, the Original-Gas-In-Place is often an “assigned” value, directly linked to the well spacing.

The recovery factor is the ratio of the Expected Ultimate Recovery to the Original-Gas-In-Place (EUR/G). As a consequence, the reported recovery factor depends on the spacing of the wells and the assigned drainage area. Accordingly, it can range from 20% to 70%.

1.8.3 Well Deliverability

The flow of gas from the reservoir into the wellbore, described previously, is often called the Inflow Performance Relationship (IPR). It is usually represented as a graph of flowing bottom-hole pressure versus gas rate (see Figure 1-8a). This graph demonstrates that the flow rate depends on the flowing pressure.

From a production perspective, the flowing pressure is controlled at the surface. The bottom-hole (sand-face) flowing pressure that corresponds to the wellhead flowing pressure depends on the flow rate, the pressure level, the gas composition, the pipe diameter, and the depth of the well. For a given wellbore configuration and a specified wellhead pressure, the flowing bottom-hole pressure can be calculated at various arbitrary flow rates. This is shown in Figure 1-8b, and is called a Tubing Performance Curve (TPC).

Since the reservoir and the wellbore must interact, the “operating point” is obtained by overlaying the TPC and the IPR curves (see Figure 1-8c).

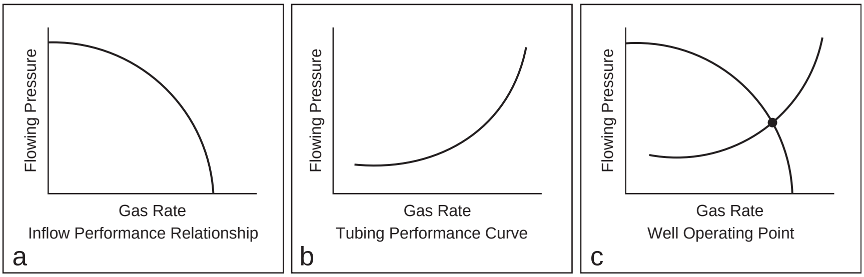

Figure 1-8 Well deliverability.

The point of intersection is the production rate from the well at the specified conditions. Several TPCs can be generated by varying the specifications of the wellhead conditions or the tubulars. These are overlaid on the reservoir's IPR, and different operating points will result. The choice of operating conditions is based on the economic factors associated with each TPC.

Calculation of the TPC is based on the friction and hydrostatic pressure drop calculations outlined in Chapter 3. In general, for single phase flow of gas, the bigger the tubing diameter, the larger the operating point flow rate, and the choice of tubing diameter is straightforward and is directly linked to the cost of tubing. However, many gas wells produce gas and liquids (water or condensate). Multiphase calculations are complex and often show that the operating point rate can be increased by "reducing" the tubing diameter.

When a gas well produces liquids, care must be taken to efficiently remove all the liquids; otherwise, they accumulate in the wellbore, and eventually the increasing hydrostatic back pressure "kills" the well. Various mechanisms exist for removing the liquids. These range from using "siphon strings" (small diameter tubing) or "plunger lift" to reducing the wellhead pressure to using bottom-hole pumps. Many gas wells stop producing (die), even though there is still a lot of gas left in the reservoir. They die simply because it is not economically viable to remove the liquids from the wellbore.

The production rate of a gas well decreases with time because the reservoir pressure depletes. Eventually, the flow rate becomes uneconomic, and the well is abandoned. Typically, to maintain a gas supply contract, companies drill additional wells over time to supplement the decreasing deliverability of the wells.

1.9 NATURAL GAS TRANSPORTATION

Transportation is an essential aspect of the gas business, since gas reserves are often quite distant from their main markets. For almost a century, natural gas has been transported safely, reliably, and economically via pipeline, and pipelines were ideally suited to the supply and market conditions of the twentieth century, when large reservoirs of gas could be found in accessible locations that provided the stability and long-term security that pipeline projects demand. Now, in the twenty-first century, the vast majority of the large, easy-to-produce, and pipeline gas reserves plays have been already tapped, and attention is shifting to stranded reservoirs that were previously thought too small, too remote, or technically too difficult to develop.

The liquefied natural gas (LNG) industry has commercialized many large remote gas fields over the past three decades and developed gas markets commercially unreachable by pipeline.

Over the last two decades several technologies have been also evaluated and proposed for monetizing hitherto remote gas reserves (see Figure 1-9). These include a number of technologies converting natural gas into a range of easily transported and marketed hydrocarbons (grouped under the generic term "gas-to-liquids," or GTL, technologies); using gas to produce electricity at the producing field and then transporting it by high-voltage direct current (HVDC) transmission lines over long distances (generically referred to as "gas-to-wire," or GTW, technologies); using compressed natural gas (CNG) that avoids the cost of gas liquefaction; converting gas, particularly associated gas, into solids or slurries formed of gas hydrates for storage and transportation (generically referred to as "gas-to-solid," or GTS, technologies). The latter two technologies are still in the research and development stage, and although much work has been done to verify the potential of these options in the past decade, no commercial projects exploiting them have yet been sanctioned (Mokhatab and Wood, 2007; Wood et al., 2008).

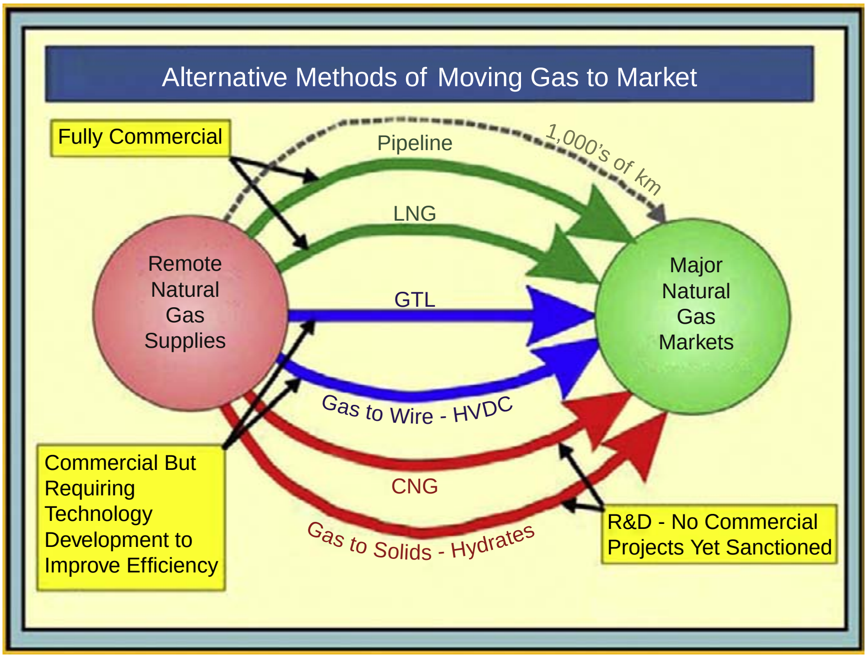

Figure 1-9 Technologies available to transport natural gas long distances (Wood and Mokhatab, 2008). *(Figure created by David Wood & Associates).*

The following sections examine some of these technical methods by which natural gas energy can be transported and cover many of the essential points needed to enter the discussion.

1.9.1 Pipelines

Pipeline transportation has been employed to deliver gas to markets for a long time. However, supply of natural gas to markets via gas pipelines is often faced with technical, economic, and even political limitations.

Pipelines are a very convenient method of transport but are not flexible, as the gas will leave the source and arrive at its (one) destination. If the pipeline has to be shut down, the production and receiving facilities and processing plant often also have to be shut down because gas cannot be readily stored, except perhaps by increasing the pipeline pressure by some percentage (Cranmore and Stanton, 2000).

For onshore and near-shore gas, pipeline is an appropriate option for transporting natural gas to market. However, as transportation distances increase, pipelines become uneconomical, where large-diameter and long-distance pipelines imply very high capital investment. They require both large, high-value markets and substantial proven reserves to be economically viable. Pipeline technology has advanced significantly in the past two decades, however, overcoming challenges of pipe size, distance, and maximum water depths that lead to lower gas-transportation costs.

Pipelines usually entail the crossing of a number of countries and borders, which means that several governments may be involved, increasing the relative complexity of development. This may also raise the potential of supply interruptions through diversion of gas volumes, attack on the infrastructure, or having the intermediary country close the valve for political reasons.

1.9.2 Liquefied Natural Gas (LNG)

Liquefied natural gas (LNG) technology has been proven to be effective over the past thirty years. When natural gas is cooled to approximately –260°F (–162°C) at atmospheric pressure, a condensed liquid called liquefied natural gas forms. The volume reduction is about 1/600 the volume of natural gas at the burner tip. The physical properties of LNG allow for its long-distance transport by ship across oceans to markets and for its local distribution by truck onshore. Occasionally, liquefaction of natural gas also provides the opportunity to store the fuel for use during high consumption periods close to demand centers, as well as in areas where geologic conditions are not suitable for developing underground storage facilities.

The much lower volume of liquefied natural gas relative to gaseous natural gas can reduce transportation costs by allowing delivery using cargo ships or transport trucks instead of pipelines. However, the costs of building a liquefied natural gas plant are typically high relative to comparable energy projects for a number of reasons, including remote locations, strict design and safety standards, large amounts of cryogenic material required, and a historic tendency to overdesign to ensure supply security. Costs throughout the value chain have been declining in the LNG industry in recent years. Further advances in LNG technology can be expected in liquefaction and shipping, which could lead to lower overall project costs (Cornot-Gandolphe et al., 2003). In the longer term, the floating LNG (FLNG) concept, where processing and storage facilities are based on a vessel moored offshore in the vicinity of the producing fields, could reduce costs even more and make the development of some small and remote gas reserves or deep offshore gas feasible. This technology can reduce costs by minimizing the cost of offshore platforms and pipelines, eliminating the need for port facilities, and reducing the time needed to build the plant. Vessel construction can be also carried out in a low-cost location. However, there are potentially many commercial and technical challenges that need to be addressed during the development of FLNG projects. The key for the industry now is to delineate these challenges and discuss means to resolve these problems (Chiu and Quillen, 2006).

1.9.3 Compressed Natural Gas (CNG)

Gas can be transported in containers at high pressures, typically 1,800 psig for a rich gas (significant amounts of ethane, propane, etc.) to roughly 3,600 psig for a lean gas (mainly methane). Gas at these pressures is termed compressed natural gas (CNG). CNG offers proven technology that has the potential to provide an early-to-market, economic solution for remote offshore gas developments with small to medium reserves or for associated gas reserves in large oil field developments. They could work where subsea pipelines are not viable because of distance, ocean topography, limited reserves, modest demand, or environmental factors, but where LNG is also not economical due to its high cost of liquefaction and regasification facilities, or feasible due to community or safety issues. Safety and security are contentious issues for CNG, but by locating the offloading facilities tens of kilometers offshore, it is possible to remove many of the community-related safety and environmental issues that have dogged the permitting for LNG receiving terminals located onshore United States in populous areas in recent years.

Energy consumed in operating a CNG project is about 40% of an LNG project. The threshold volume of gas reserves is relatively low for commercial viability, providing shipping costs can be kept low and work effectively. Greater than 85% of CNG project costs are likely to be associated with the ships, which are based upon conventional bulk carriers with at least four competing certified containment designs (Hatt, 2003): EnerSea (U.S.) VOTRANS carbon steel pressure cells; Trans Ocean Gas (Canada) fiber-reinforced plastic (FRP) covering high-density polyethylene cells; Trans-Canada CPV steel liner cell overwrapped with a glass fiber laminate; Sea NG (Canada) patented Coselle of coiled X70 high-strength steel pipe wound into a cylindrical storage container.

In January 2007, Japan's Marubeni and Teekay joined Sea NG in a global alliance to develop CNG ships. Coselle ships in this alliance will target projects to transport moderate volumes of natural gas (30 to 700 MMscfd) over medium distances (200 to 1,500 miles). The involvement of these major backers could finally get a commercial CNG project off the ground. Certainly with most of the capital costs of the technology in the vessels, it is important to have a large and experienced shipping company such as Teekay at the helm (Wood and Mokhatab, 2008).

An additional emerging possibility for CNG is the distributed "hub-and-spoke" concept currently under consideration in the Greek islands and the Amazon river tributaries in Brazil. These approaches appear to be even more attractive than all other natural gas transportation possibilities for even smaller markets (Wood et al., 2008).

1.9.4 Gas-to-Liquids (GTL)

In gas-to-liquids (GTL) transport processes, the natural gas is converted to a liquid and transported as such. The technology of converting natural gas to liquids is not new. In the first step, methane is mixed with steam and converted to syngas or synthetic gas (mixtures of carbon monoxide and hydrogen) by one of a number of routes using suitable new catalyst technology (Keshav and Basu, 2007). The syngas is then converted into a liquid using a Fischer–Tropsch (FT) process (in the presence of a catalyst) or an oxygenation method (mixing syngas with oxygen in the presence of a suitable catalyst). The produced liquid can be a fuel, usually a clean-burning motor fuel (syncrude) or lubricant, or ammonia or methanol or some precursor for plastics manufacture; e.g., urea, dimethyl ether (DME), which is also used as a transportation fuel, LPG substitute, or power-generation fuel as well as a chemical feedstock (Knott, 1997; Skrebowski, 1998; Thomas, 1998; Apanel, 2005). The environmental benefits of GTL products have been well demonstrated and have much potential in improving air quality in cities polluted with transportation fuel emission. However, on a full-cycle environmental analysis, GTL fuels do not significantly outperform refinery fuels because the GTL plants and their components involve substantial emissions. The problem is primarily with the low energy efficiency of syngas generation and the low carbon efficiencies of the conversion processes such as FT (O'Rear and Goede, 2007). Technology breakthroughs are required to improve these efficiencies, which will lead to both cost and environmental benefits.

The Fischer–Tropsch GTL (FT-GTL) method is an application of the basic Fischer–Tropsch process, where synthesis gas (or syngas) is reacted in the presence of an iron or cobalt catalyst. End products are determined by the length of the hydrocarbon chain, which, in turn, is determined by catalyst selectivity and reaction conditions. FT-GTL is essentially a three-stage process (syngas generation, F-T transformation, and product upgrade). Possible end products include kerosine, naphtha, methanol, dimethyl ether, alcohols, waxes, synthetic diesel, and gasoline, with water or carbon dioxide produced as a byproduct. FT-GTL has reached a key milestone as an industry in that worldscale capacity plants are in the process of being planned, constructed, and commissioned. However, FT-GTL has a lot left to prove about its technical, economic, and environmental efficiency before it can considered as a rival to LNG for developing and monetizing stranded gas reserves on a large scale (Fleisch et al., 2003; Al-Saadoon, 2005; Apanel, 2005). Hundreds of modifications and patents have been applied to this complex, energy-intensive process, and further developments continue. Most recent technical advances are focused on lowering capital expenditures, establishing economies of scale, and improving operating and energy efficiencies for large-scale FT-GTL plants. Efforts to reduce the footprint, energy consumption, and efficient and safe use of oxygen in syngas reforming plants have also received attention in recent years. Much research is still ongoing with respect to other GTL processes, but none has yet reached large-scale commercialization although further developments, particularly in methanol to gasoline plants, should be expected in this high-price gasoline market (Wood and Mokhatab, 2008; Wood et al., 2008).

GTL technology development has reached a stage where its marinization, enabling smaller footprints and more compact plants than other technologies, may be considered for FPSO application. Floating GTL will provide new opportunities for companies to produce, transport, and market gas reserves that would otherwise remain stranded. Also, floating GTL plants could be moved from location to location, enabling access to small fields that would otherwise not justify building a dedicated GTL facility. The GTL FPSO concept, which is still in the development phase, could greatly improve potential project economics (Loenhout et al., 2006).

1.9.5 Gas-to-Solid (GTS)

Gas can be transported as a solid, with the solid being gas hydrate (Børrehaug and Gudmundsson, 1996). Natural gas hydrates (NGHs), which are essentially natural gas in a "frozen" state, form when water and natural gas combine at low temperatures and high pressures. Gas hydrates are clathrates where the guest gas molecules are occluded in a lattice of host water molecules. They are most commonly encountered in the industry as a production problem in pipelines to be avoided as part of flow assurance.

Gas-to-solid involves three stages: production, transportation, and regasification. Conceptually, hydrate slurry production is simply mixing chilled water and gas. In practice, processed gas is fed to a hydrate production plant, where a series of reactors convert it into hydrate slurry. Each reactor further concentrates the hydrate slurry. It is then stored and eventually offloaded onto a transport vessel (insulated to near adiabatic conditions). At the receiving terminal, the hydrate is dissociated, and the gas can be used as desired. The water can be used at the destination if there is a water shortage, or returned as ballast to the hydrate generator and, since it is saturated with gas, will not take more gas into solution.

Gas storage in hydrate form becomes especially efficient at relatively low pressures where substantially more gas per unit volume is contained in the hydrate than in the free state or in the compressed state when the pressure has dropped. When compared to the transportation of natural gas by pipeline or as liquefied natural gas, the hydrate concept has lower capital and operating costs for the movement of quantities of natural gas over adverse conditions. NGHs are chemically stable at about $-20°C$ compared with $-162°C$ for LNG, reducing the costs of transportation and storage. One cubic meter of NGH contains approximately 160 m^3 of natural gas, while one cubic meter of LNG contains 600 m^3 of natural gas, thus limiting the quantity of gas that can be transported with the NGH technology. This is a considerable volume penalty (and hence transport cost) if considered in isolation; with the cheaper ships for hydrate transport, the process could be economic.

"Gas-to-solids" options for transporting and storing stranded gas to market have been extensively researched and laboratory tested for more than a decade. BG Group, Marathon, NTNU (Norway), and others have worked on a range of gas-to-solids transportation technologies (Fischer, 2001) and have taken them to small-scale pilot plants. Concepts include storage and transport of gas either as atmospheric hydrates or as a paste in pressurized insulated containers or as frozen hydrates are mixed with refrigerated crude oil for atmospheric pressure transport. Even though these studies have proved the concept of storing natural gas in hydrates technically feasible, applications have not progressed beyond the laboratory stage because of complexities of the process, slow hydrate formation rates, and costs. No projects are close to commercialization, but technological advances continue to be made, e.g., forming the hydrates from a surfactant solution (Rogers et al., 2005) that may make this possible in the long term.

1.9.6 Gas-to-Wire (GTW)

Currently, much of the transported gas destination is fuel for electricity generation. Electricity generation at or near the reservoir source and transportation by cable to the destination(s) (gas-to-wire) is possible. Thus, for instance, offshore or isolated gas could be used to fuel an offshore power plant (may be sited in less hostile waters), which would generate electricity for sale onshore or to other offshore customers.

High-voltage direct current (HVDC) transmission lines offer the most technically viable solution to moving large quantities of electric power over large distances (up to about 1,500 km) keeping line losses less than 10%. However, HVDC is capital intensive and requires costly converter stations at either end of the transmission line. Additional costs for installing and then operating and maintaining gas turbines at the remote site would be incurred.

While HVDC cables are now being used more widely to transport electricity tens of kilometers, no projects to develop remote gas fields in this way have yet been sanctioned. Distance, cost, and efficiency of remote generation make other options currently more attractive. There are other practical considerations to note; for example, if the gas is associated gas, then if there is a generator shutdown and no other gas outlet, the whole oil production facility might also have to be shut down or the gas released to flare. Also, if there are operational problems within the generation plant, the generators must be able to shut down quickly (in around 60 s) to keep a small incident from escalating. Additionally, the shutdown system itself must be safe so that any plant that has complicated processes that require a purge cycle or a cool-down cycle before it can shut down is clearly unsuitable (Ballard, 1965). Finally, if the plant cannot shut down easily and/or be able to start up again quickly (perhaps in an hour), operators will be hesitant to ever shut down the process, for fear of financial retribution from the power distributors.

1.9.7 Comparison Between Various Methods

As discussed previously, there are a number of options of exporting natural gas energy from oil and gas fields to market. However, distance to market and potential production volumes (dependent on the reserve size of the gas resource) influence the technologies that might viably be used to exploit remote gas fields (see Figure 1-10).

Any gas energy export route requires a huge investment in infrastructure, and long-term "fail proof" contracts, covering perhaps 20 years or more. However, improvements in technology, economies of scale, and synergies will doubtless lower capital costs and further improve the project economics over the next few years. Most of these technologies have reached a stage

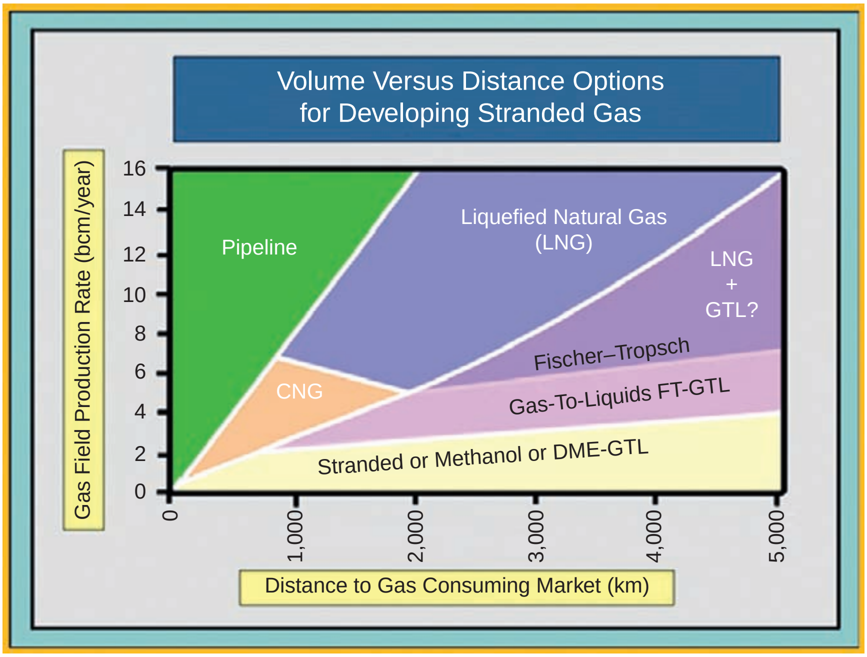

Figure 1-10 Production volume versus distance to market framework for gas technologies (Wood and Mokhatab, 2008). *(Figure created by David Wood & Associates).*

where commercialization on a wide scale is only a few years away. However, their use carries risks in terms of technology, creditworthiness, revenue security, and market competition risks, each of which needs to be appropriately mitigated.

Gas is currently transported to markets primarily via two long-established methods (pipeline and LNG technology). Alternative technologies, which have been refined and developed in recent years, are yet to make serious inroads into the challenges of developing remote gas fields. Figure 1-11 shows that as the distance over which natural gas must be transported increases, usage of LNG has economic advantages over usage of pipelines. As can be seen, for short distances, pipelines—where feasible—are usually more economical. LNG is more competitive for long-distance routes, particularly those crossing oceans or long stretches of water since overall costs are less affected (Economides and Mokhatab, 2007).

LNG, an effective long-distance delivery method, constitutes only 25% of global gas movement. However, LNG can offer flexibility, diversity, and security of supply advantages over pipeline alternatives. LNG projects require large investments along with substantial natural gas reserves and

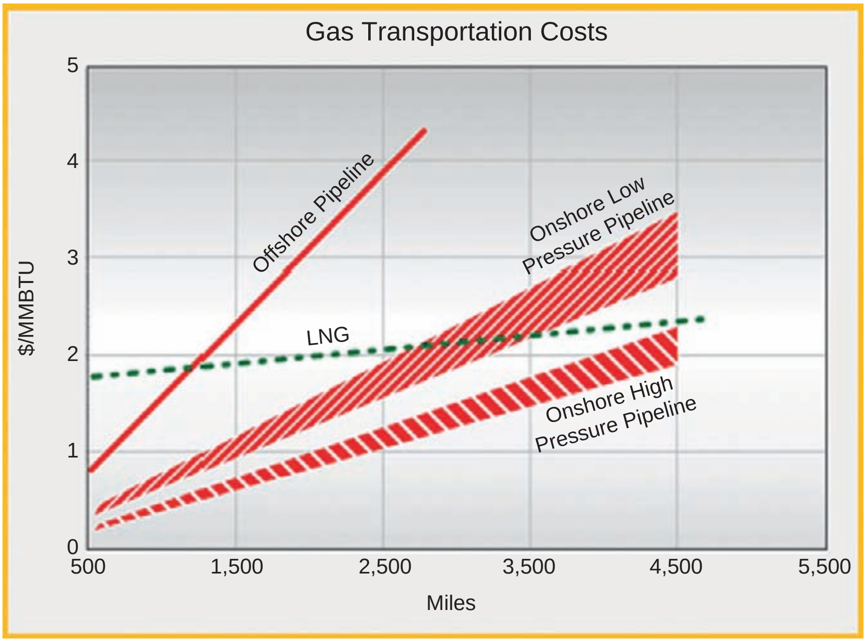

Figure 1-11 Comparison of the cost of transporting gas via pipeline and LNG; for 1 tcf/year and including regasification costs (Economides and Mokhatab, 2007).

are economically viable for transporting natural gas in offshore pipelines for distances of more than 700 miles or in onshore pipelines for distances greater than 2,200 miles (Mokhatab and Purewal, 2006). However, further improvements in LNG technology can be expected, which could lead to lower overall project costs.

Note that CNG is an economic alternative to monetize stranded gas reserves and creates new market where pipelines and LNG deliveries are not practical. CNG technology provides an effective way for smaller volumes and shorter-distance transport of gas. The results show that for distances up to 2,200 miles, natural gas can be transported as CNG at prices ranging from \$0.93 to \$2.23 per MMBtu compared to LNG, which can cost anywhere from \$1.50 to \$2.50 per MMBtu depending on the actual distance. At distances above 2,200 miles, the cost of delivering gas as CNG becomes higher than the cost for LNG because of the disparity in the volumes of gas transported with the two technologies (Economides et al., 2005). Also, unit gas transportation comparisons for pipeline, LNG, and CNG (see Figure 1-12) suggest that pipelines will match the CNG costs at transportation volumes of about 750 MMcf/d or greater over intermediate distances. At volumes below this, CNG will offer lower unit transportation costs. LNG will match CNG unit transportation costs at larger volumes or longer distances. Therefore, CNG technology may have the potential to challenge LNG transportation for some niche markets. CNG can start with much smaller initial volumes than either a pipeline or LNG and can grow incrementally to meet demand by simply adding ships. It should be

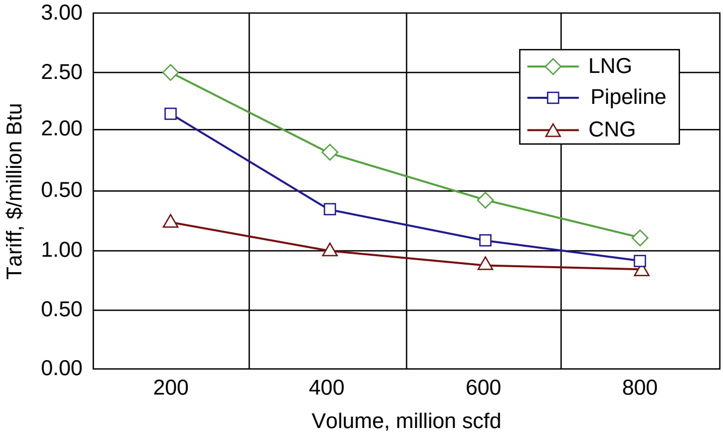

Figure 1-12 Transportation tariffs (Economides et al., 2005).

noted that, while pipeline and LNG are proven either as concepts or technologies, CNG is still awaiting its first commercial application. Nevertheless, reflecting on the system's simplicity, the level of engineering efforts invested, and developed reliable cost estimates, it is believed that CNG technology is ready for commercialization (Economides and Mokhatab, 2007).

In addition to CNG technology, there are numerous technology alternatives to the currently commercial pipeline and LNG options for transporting natural gas to markets on a wide range of scales, which have been described before. Unfortunately, few of them are yet receiving the levels of investment and commitment required to bring them into widespread development. High prevailing oil and gas prices around the world could change this lack of enthusiasm for embracing novel technologies.

The transport options preferred by governments and companies not only must take the economic risks into account but also consider the negative effects of possible terrorist activity, political changes, and trade embargos over long periods of time. Thomas and Dawe (2003) covered many of the essential technical points and broad economic pointers needed to enter the discussion of gas rich states that do not need the gas for domestic use but wish to monetize their reserves by export.

1.10 DYNAMICS OF GLOBAL GAS BUSINESS

The global natural gas market has undergone massive changes over the past two decades. These changes have allowed the market to be more dynamic and more competitive, encouraging countless new market entrants, both on the sell and the buy side of the equation. The process is expected to continue as the role of gas in the global energy mix continues to grow at 2%pa—at the expense of oil (growing at 0.7%pa) and, to a limited extent, coal (also growing at 0.7%pa) and nuclear energy. This growth is both within national boundaries as well as across borders and overseas. Gas usage is growing across all sectors, including power, heating, industrial, petrochemicals, and transport usage. New technologies, such as GTL and CNG, may also unlock further gas demand in the future once challenges and efficiencies unique to these concepts can be overcome on a larger scale.

In the past, gas contracts were relatively simple and straightforward. A gas resource owner would typically build and operate production and processing facilities. The facilities would remove natural gas liquids from the gas stream for separate sale—at prices higher than lean gas—and pipeline-quality lean gas (with acceptable quantities of impurities) would be sold to a gas wholesaler or consumer. A tolling fee would be paid to the pipeline owner to transport the gas, and the gas wholesaler would be responsible for ensuring that the gas was available to the consumer at the specified volume and time.

As gas markets matured, national grids (as in the United States) and multinational grids (as in Europe and other select regions of the world) were developed. These grids connected large numbers of sellers and buyers. These parties often contracted with the pipeline owner or with an intermediate trading company to buy and sell gas. Different quality gas was often comingled in the pipeline, and the terms of the contract became increasingly more flexible.

A similar evolution has occurred in the LNG space. In the past, there were limited numbers of LNG exporters. LNG importers were limited to those companies that had access to LNG-receiving terminals—and most of these terminals were in Japan with a handful in Europe and North America. LNG was sold on a long-term basis (typical contracts were over 20 years' duration) with very limited flexibility. The buyers had to take their contracted volumes without the right to divert cargoes, and ships, either owned by the buyers or the sellers, were dedicated to the contracts. LNG prices were typically set according to a formula related to oil prices with limited scope of modifications.

As the LNG world expanded dramatically from 2000 onwards, the number of countries involved in the trade expanded to around 50 countries today. Between 2008 and 2015, the number of countries importing LNG will grow from fewer than 20 to around 35 (a growth of 75%), and the number of countries exporting LNG will grow from 15 to 20 (25% growth). New entrants are no longer bound by the rules of the past and are more willing to accept shorter duration contracts, more flexible terms, pricing formulas not linked to oil prices, and nonrigid shipping options. The growth of the worldwide LNG spot market, from 5%–10% of the global trade (in 2000) to around 25% of the trade in 2011, is a testament to this change.

In the past, LNG projects would only be developed once all the produced volumes were contracted to blue-chip creditworthy buyers. As LNG prices began to rise around 2007 onwards, growing economies such as Korea and others (such as Spain) were willing to pay a premium for individual or short-term cargoes. These spot cargoes allowed the buyers to augment their long-term contracted volumes for seasonal variations or for unusual events. For example, during Spain's drought in 2008, Spanish utilities were willing to pay a very high premium to import LNG to make up for the shortfall in hydroelectric power generation. In turn, LNG sellers began to reserve a proportion of their outputs for these 'spot' sales. A growing proportion of LNG projects currently under development have such allocated volumes. Unfortunately, the premium that spot market prices commanded over long-term contracts cannot be guaranteed, and there have been periods where the spot market prices have been less than the long-term contracted prices.

As would be expected in commodity markets, as the number of buyers and sellers increased, LNG trading companies became increasingly active. These companies may be involved in individual trades, where they buy LNG cargo at a certain price for resale at a higher price, or they could contract to purchase volumes over multiyear periods, whereby they, in turn, would enter into LNG sales contracts to end users. Energy companies, such as Shell, Gazprom, Total, BG, and others, are active LNG traders. Specialist trading companies have also entered into this business. An LNG consumer (such as a utility company in Singapore or Dubai) would contract with a trading company such as BG or Shell to supply LNG to its terminals. BG and Shell would procure these cargoes from their own equity-owned LNG projects or buy them from other projects on an opportunistic basis. Owners of the LNG exporting plants would sell a cargo to BG or Shell without any real control where the cargo would be ultimately unloaded. A cargo sold to a trading company may, in fact, compete with another cargo sold on a longer-term contract basis by the same plant to the same end consumer.

Emerging LNG buyers, such as India and China, are using LNG to augment to their existing gas supply. Though they have signed some long-term LNG and pipeline contracts, both markets are active players in the LNG spot market. They buy spot cargoes when these cargoes are priced less than their contracted supply. These markets are also reluctant to accept oil-price linkages to their gas contracts. As oil prices stay high, large exporters such as Qatar continue to push their buyers toward oil-linked contracts; however, this is meeting increasing resistance, both from existing and emerging buyers.

The dramatic growth of U.S. unconventional gas production has been a significant "game changer" in the global gas market. Unconventionals, which include shale gas, coal seam methane gas, and tight gas (gas produced from low-permeability reservoirs), have grown from low levels to over 50% of current U.S. production. This abundant production has resulted in depressed gas prices in the United States, increasing the disparity between North American prices and European/North Asian prices. This has come at an unfortunate time since a number of large energy companies (and some aggressive independent energy companies) built LNG-receiving terminals in anticipation of a domestic gas shortage—and resulting spike in gas prices—expected in North America. These terminals are now running at very low volumes since it is no longer feasible to import large volumes of relatively expensive LNG into the U.S. market—though on a seasonal basis it may be feasible for limited volumes. As a result, a number of these receiving terminals are evaluating the option to invest in new liquefaction facilities. This would allow them to import LNG during periods when U.S. pipeline grid prices are relatively high and export LNG when U.S. prices are low enough to support purchase of gas from the grid plus liquefaction and shipping costs to global LNG markets. Promoters of these proposals have quoted prices of less than $10/MMBtu for delivered LNG to Europe and Japan. If they can deliver LNG at these prices, U.S. LNG would be competitive with new LNG suppliers in Australia and elsewhere.

The number of LNG exporting projects has expanded dramatically over the past few years. Unfortunately, this dramatic expansion—which resulted in an increase of global LNG production from 175 Mta to over 250 Mta in the period from 2007 to 2011—coincided with a sharp slowdown in world economic growth. This overhang will likely lead to a period of overcapacity and flat to declining prices. New and growing gas markets, especially in Asia, will absorb this new capacity and by 2015–2016, supply and demand are expected to regain equilibrium. This will be the same period that new projects, mainly in Australia and Africa, will begin production. It remains unclear whether these new projects, many of them based on relatively expensive coal seam methane, will be able to compete during this period of price recovery. If they are able to do so and are not significantly delayed, they may enjoy a period of higher LNG prices, beginning 2016 or so.

The global gas markets are very dynamic. Contracts are becoming more flexible, shorter term, and increasingly delinked from oil prices. These changes are being promoted by both established consumers as well as emerging buyers in Asia and elsewhere. The future promises to be increasingly vibrant.

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CHAPTER 2

Natural Gas Energy Pricing

2.1 INTRODUCTION

The economics and the engineering of the production of natural gas intersect in the marketplace—the locale in which this energy source eventually clears at a price that enables sale and distribution of natural gas as a commodity in its own right. No one is unaware or in any doubt as to the fact that the engineering of the production of gas, like the production of any other raw material that eventually enters the market as a commodity for sale, must have its effect on the market price. What is less clear but at least as, if not more, important is the fact and the modalities by which the price of gas (or, more properly speaking, the *trend* in the market price of this resource) subsequently exerts its own impacts on the further engineering of how gas is produced—and this from more than one direction. Helping clarify some key aspects of how this operates is the main rationale for including such a chapter in this book.

Regardless of the fact that the engineering of conventional natural gas production has become highly standardized around the world, there exists no such thing as a world price for the natural gas commodity. This is quite unlike the situation with refined petroleum, with which natural gas is usually bracketed together under the "fossil fuel" rubric, and, like petroleum, is also believed to derive some portion of marketable value from its apparently "nonrenewable" character. In most of the world down to the present day, the wholesale-level market prices of natural gas and refined petroleum operate in different ranges unconnected with one another,[1] and this divergence is no accident. As long as access to world markets for those in the oil-rich Middle Eastern producing states extracting so much of the world's supply of crude oil upstream remains gated by refining and retailing interests downstream, some kind of world price is bound to be maintained for crude oil. Natural gas is produced in commercial quantities mostly for local and regional markets in many parts of Europe and the Americas; a world price

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[1] As discussed further in Section 2.5 and later, from the 1980s to date, the price of natural gas in U.S. markets has tracked that of crude oil.

is thus neither necessary nor likely. Although the actual energetic content of equivalent volumes of gas and refined petroleum is easy enough to compare arithmetically, the usefulness of such a comparison in establishing an acceptable market-clearing price for each is limited by the fact that the set of end uses are increasingly distinct for each commodity and these end uses are by no means uniformly distributed. For example, over the last two decades, natural gas has considerably increased its profile as a transport fuel in Europe and Russia, but this end use has yet to make a significant dent in North America, Africa, or most parts of Asia. On the other hand, over the same period, natural gas has considerably displaced refined petroleum as a fuel source in electric power plant operations throughout the Americas and the Eurasian land mass. In addition to this complexity, there is the additional fact that the elasticities of supply and demand for each of these energy sources, which have considerable impact on the movement of their wholesale prices, are also very different. For example, the elasticities of supply for refined petroleum and produced natural gas diverge quite sharply. Whereas market supplies of crude oil for refineries, as well as of refined petroleum products for the wholesale and retail market, include a very large amount maintained in various forms of storage, ranging from tank farms to contango fleets, the situation of natural gas could not be more different. While a certain portion of marketable gas product is stored as LNG, most reaches markets more or less directly whether from regional pipeline branches and local spur lines or through pipeline connections from LNG import-handling facilities.[2]

Currently at the international level, from Asia to the Americas, the natural gas business appears at least on the surface to be booming globally everywhere, from just about any angle. Beneath that surface, meanwhile, since at least 2005–2006, several contradictory tendencies have emerged. Some are new, whereas others are coming into greater prominence than before. Interacting within the contexts of the exploration, production, and marketing of natural gas, some of these trends are altering the direction of certain fundamental drivers of the main scenarios discussed in the first edition of this chapter.

The trigger for this chain of reactions has been provided by the ongoing global financial and economic crisis. Its bubble burst ruinously during 2007–2008 upon the economies of the leading developed industrial countries, as

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[2] Beyond this divergence, it is also necessary to mention here that, for both the supply and demand elasticities of refined petroleum, as well as the demand elasticity of produced natural gas, further cogent comment regarding trends in these respective elasticities is necessarily constrained by the large component of corporate self-interest that necessarily infests any claims they have issued regarding their own actual reserves of such resources.

well as upon the economies of some (but not all) of their leading trading partners. By 2011, none of these economies appeared to be definitively putting this crisis or its effects behind them, or showing even credible evidence of stabilization, however temporary. It should come as no surprise, then, that such an event and its differential impacts should have made mincemeat of many predictions earlier in the last decade, especially predictions concerning the likeliest provenances and outcomes of "resource wars" that seemed to loom so large on the horizon back at the start of this century.

In similarly unexpected ways, this crisis and its impacts have heightened the very different effects of the unevenness of political and economic development as between the developing and developed countries. Notably in this regard, most of the predictive power routinely assumed to inhere in the prevailing assumptions used for modeling the international economy has been vitiated as one prediction after another is disproven by events.

It is while that movement itself passes through contemporary historical time that the authors of the following chapter examine and attempt to summarize natural gas production and marketing as a component of the global economic movement of this period. The mode of presentation adopted here necessarily aims, therefore, to unfold a two-phase cycle before the reader. The two phases comprise

1. Information concerning the rolling out of new technologies and productive capacities and their impacts on the profitable delivery of energy sources in world markets; and
2. The reaction or other impacts from the further movement of actual political and economic events on the development and relative importance of natural gas among the various energy sources now accessing markets.

Data published regarding natural gas reserves and reserves-to-production ratios supplied over the previous decade would seem to spell boom times ahead.[3] However, to guard against any tendency to get carried away by, and overinterpret, such elaborate efforts at quantitative measurement, this chapter addresses the qualitative developments in contemporary historical time by rolling out a brief *tour d'horizon* of salient natural gas developments within the global picture.

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[3] See, for example, the *BP Statistical Review of Oil and Gas* for 2004 and for 2010. Natural gas statistics in that series of volumes usually begin around page 20. These data sheets may be viewed on, or downloaded from, the internet at http://www.bp.com. Since the BP source data gives the R/P ("reserves-to-production") ratios only for the year preceding the year cited in the report-title, however—e.g., the 2010 report gives the R/P ratio for the end of 2009—other sites need to be consulted to recover the R/P ratios for earlier years. For that purpose, the website http://www.peakoil.net/statistics is useful.

To help the reader better appreciate the importance of asserting such a differentiation of the qualitative from the quantitative representation of developments, consider the nature and scale of the sources of data about natural gas production and marketing. It is hardly an overstatement to point out that the United States leads the entire world both in the complexity and extent of marketing mechanisms in the natural gas sector. This is to the extent that the Energy Information Administration (EIA) of the United States Department of Energy itself is probably the premier collector and processor of natural gas marketing and production data from around the world. Its databases are used as a source by the International Energy Agency in Paris, an arm of the Organization for Economic Cooperation and Development comprising all the most industrially developed economies of the planet. Its data-gathering capabilities provide a powerful backstop for the already-extensive data gathering in this field undertaken by the major international banks based in the United States that participate in financing, or syndicating the financing of, numerous energy projects around the world, not just within the United States. Although the EIA adds considerable qualitative material in the form of "country analyses," the impression a researcher takes away from studying their materials is that the amount of qualitative disclosure provided is actually driven by the amount and assumed import of the quantitative data available, rather than the other way round. For example, the EIA's coverage of Russia's oil and gas sector appears to cover the most significant features of the industrial structures that ultimately generate the quantitative data compiled by the EIA. But they manage to miss the rather significant fact that the earnings of the post-Soviet oil and gas giant Gazprom comprise 8% of the GDP of present-day Russia. This fact by itself would seem to mark Gazprom clearly as the key link of the entire Russian oil and gas sector, but the EIA's country analysis is "neutral" about this or any other evidence that might point to any particular driving force behind Russia's ever-expanding production and export of oil and gas. Its take on the matter seems to be that the tail wags the dog, i.e., that the country's oil and gas output and scale of its exports is the main story and the structures that ensure this output derive from that.[4]

When it comes to assembling a meaningful qualitative picture of the leading developments in natural gas production and marketing, another tendency that unconsciously asserts itself is rooted in an assumption that

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[4] EIA's Russia country analysis is publicly accessible at http://www.eia.gov/countries/cab.cfm?fips=RS.

the primacy in global terms of the quantitative picture that can be assembled for the United States implies similar primacy for the qualitative picture that can be related concerning current developments in natural gas production and marketing in that country. However, here a most interesting anomaly leaps out: notwithstanding the widescale participation to be found involving U.S. financial institutions, the supplies and production of every other major natural-gas-producing country on the planet participate heavily in export markets often quite distant from countries immediately contiguous to their borders. However, the United States' own supplies of natural gas show up almost nowhere in export markets beyond Canada and Mexico. This simple fact suggests that the qualitative picture of U.S. natural gas production and marketing is highly singular. It tends to reflect no general trend commonly found among other natural-gas-producing countries. This makes it difficult to assign any particular status to the quality of developments in the U.S. natural gas sector in general, either as leading or lagging indicators of what is happening elsewhere in the world natural gas business. At the same time, certain singular developments in the commercialization of natural gas production and distribution in the United States do indeed appear to hold out warnings and/or lessons for the further globalization of certain financial instruments that need pointing out. Similarly, it is important for practicing engineers to be aware of certain regulatory activities of the U.S. Environmental Protection Administration (EPA) that may be playing a present and future role in standardizing how certain novel or unconventional technologies are applied to natural gas production.

Examining particular natural gas markets according to criteria that start from the present and look back over the last couple of decades, and more acutely since the middle of the last decade, to locate and identify sources for the principal dynamic factors operating in the present, profound transformations emerge. In general, market-cornering attempts and activity associated with such attempts have tended increasingly to trump the soundest judgments as to the best and highest engineering to apply to the provision of this energy source. This same chapter in the previous edition of the present work captured and reported some snapshots of pieces of these phenomena. Since then, a natural gas price bubble blamed on speculative activity by a North American hedge fund grew and burst. Natural gas prices plunged, and new investment from the leading Western-developed economies in natural gas exploration and production was deeply discouraged. This happened in the same moment as the fallout from the U.S.-based financial crisis over subprime mortgages and derivatives in 2007 and 2008 fueled a quadrupling of the world oil price to just under US$150 per barrel before crashing to about US$30 a barrel at the start of 2009.

Dramatic changes have happened, meanwhile, with natural gas processing and development in the Middle East that were not even on the horizon of the previous edition of this chapter. Russian participation has developed from a focus on supplying European markets into directions such as central and east Asia as well as the far-eastern Mediterranean that were not previously anticipated, with a serious effort being mounted to unfold a Russian-led international natural gas cartel that reportedly includes the Emirate of Qatar. The Russian pipeline giant Gazprom, whose activities were reported in 2010 to account for 8% of Russia's gross domestic product, has extended its activities in the direction of connecting Syria and thus the Mediterranean by pipeline to Russia, as well as Libya's coastal natural gas fields in North Africa by pipeline under the Mediterranean to a southern Italian terminal of the Eurogas network.

Meanwhile, according to the 2010 survey issued by the U.S. Department of Energy (NETL, 2010), the shale gas boom currently underway on the territory of the United States is also displacing, to an unanticipated degree, imports of LNG by that country. The enthusiasm to adapt hydraulic fracturing techniques to liberate gas from rock—so-called shale gas plays—has spread beyond the United States to locales as varied as Canada (on the northern borders of the United States, particularly in the Rocky Mountains trench extension of the Western Canada Sedimentary Basin [WCSB]), the Blackpool area of Britain's Irish Sea coast (Peck, 2011), and the People's Republic of China, among others.

In Canada, the U.S. enthusiasm for shale gas has spread to Quebec, where strong candidates for shale gas development are being investigated at various locales within the basin of middle and upper reaches of the St. Lawrence River valley. Early in 2011, the Quebec government successfully capped a 2-year effort to win federal government support for its assertion of a jurisdictional claim over exploration and development of the "Old Harry" offshore natural gas show discovered at the marine boundary of the coastal waters of the provinces of Quebec, Nova Scotia, Newfoundland, and Labrador in the southeastern corner of the Gulf of St. Lawrence. Eastern Canadian gas production in the northwest Atlantic off the coast of Nova Scotia continues meanwhile with EnCana's development of its Deep Panuke property near Sable Island. Short of a major new find, however, production in this area is still expected to taper off over the coming decade. At the same time, companies still active in Canada's tar sands sector in the WCSB of northern Alberta and northwestern Saskatchewan have set aside most of the previous plans that were still on the boards in 2005 for the maintenance or further development of local capacity to upgrade bitumen to synthetic crude. Instead, the bitumen itself is increasingly to be sent by pipeline to upgrader facilities south of the border. Thus has Enbridge's Keystone XL pipeline system, running from Alberta across the Great Plains to the Mississippi and then south to refinery facilities on the coast of the Gulf of Mexico, become the big story coming out of this sector since 2010. One consequence of this development is that the natural gas market picture in Canada, where syncrude upgraders were initially developed in northeastern Alberta on the basis of harnessing local supplies of natural gas, has become increasingly incoherent. An alarming prediction from the EIA in 2005–2006 that Canada had less than 8 years' supply of natural gas at then-current rates of consumption (including the fueling of syncrude upgraders) provoked a flurry of interest in replacing natural gas with nuclear power as the energy source for the upgrading facilities. This was a flurry that disappeared as quickly as a snow flurry in the morning sun, however, after the numbers for a basic cost comparison of natural gas and nuclear power were crunched. At the same time, in the wake of the news circulating about the imminent disappearance of its natural gas resources, the Alberta provincial government came under pressure, *pro* and *con*, regarding the scheduling of increases in the royalty rates it collects on oil and gas production within its jurisdiction.

The pricing of natural gas as an energy commodity presents a number of features of historic significance. These historical elements relating to the politics and economics of the development of oil and gas resources are highly significant in their own right. Indeed, natural gas is playing an increasingly important role as an alternative to crude oil and the gasoline and numerous other byproducts refined from petroleum. At the same time, a number of nonhistorical components also enter into the pricing of natural gas as an energy commodity. Some of these elements, especially those implicated in environmental pollution, have an intangible connection to gas pricing although one would be hard-pressed to ferret out its exact contribution to consumers' expense. Appropriate intervention by the working engineer can do wonders for preserving the environment and all of us living in it from serious harm in the short term and the long term. Nothing—including the most precious energy commodity—can have a higher value than human life, and that forms the starting point of this chapter. Although natural approaches in general, and their usefulness for addressing environmental problems created by current energy development technologies and approaches in particular, are beyond the scope both of this chapter and of conventional concerns about gas pricing, throughout the rest of this chapter some of the obstacles that have been strewn along and throughout the various paths toward the goal of knowledge-based, all-natural, researched solutions to what some have called "our technological disaster"—solutions that would prove innovative, economically attractive, environmentally appealing, and socially responsible—will be pointed out. The principal aim of this chapter is to examine natural gas energy pricing with the aim of answering the following question: considered from an engineering standpoint, how consistent with their actual potential are the patterns, changes, and trends in the supply of and demand for so-called nonrenewable energy commodities?

2.2 ENERGY PRICING, SUPPLY, AND DEMAND

In recent years there has been an increasing recognition that the actual supply and demand for energy commodities diverge significantly from the conventional economic notions and expectations concerning the supply and demand of commodities. Much effort has been expended to define and account for the divergence as an exception relating mainly to the strategic character of energy as a commodity in a globalized marketplace. The problem with an analysis that goes only this far, however, is its failure to explain why current economic theory as a body of theory fails to account for, or predict, the energy exception. Without such comprehension at the level of theory, it becomes difficult if not impossible to discover or elucidate any paths that could or might overcome the numerous contradictions and downsides inherent in the present marketing and pricing arrangements in place for energy commodities.

What do such considerations have to do with the price of gas? Consider what happens with the hydrogen sulfide (H_2S) stream recovered from the natural gas processing scenario. The recovered hydrogen sulfide gas stream itself is considered an object of "waste management." It "may be either vented, flared in waste gas flares or modern smokeless flares, incinerated, or utilized for the production of elemental sulfur or sulfuric acid" (EPA, 1995). Any one of the first three options requires regulation as potentially hazardous emissions. Again, what is being proposed is to burden society with a further problem arising from the original technological decision to "sweeten" the gas as "efficiently" as possible rather than innovate delivery systems that would avert creation in the first place of additional waste management expense. What these waste management expenses actually represent for those selling natural gas is a further cost factored into the final delivered price of the product—to pay for meeting the U.S.'s Environmental Protection Agency (EPA) standard. (The fourth option, which would not only potentially generate new revenue but do so in a manner that undergirds ongoing production of natural gas with an additional productive purpose as a feedstock generator, is neither discussed nor framed in EPA discussions as part of the planning of natural gas processing.)

If the recovered H_2S gas stream is not to be utilized as a feedstock for commercial applications, the gas is usually passed to a tail gas incinerator in which the H_2S is oxidized to SO_2 and is then passed to the atmosphere out a stack (Mullins et al., 1975). Although there have been major mitigation efforts over the last 30 years that seem to be bearing fruit across most of the acid-rain catchment areas of eastern North America, future gas engineering projects will still have to take note of the fact that smokestack wastes are one of the most endemically costly and health-threatening environmental hazards of modern living and accordingly undertake a waste conversion plan capable of returning a revenue stream that allows the production and transmission costs as well as the delivered price of natural gas either to be reduced, or its rate of increase contained. A generation ago, these wastes were especially strongly implicated in the return and increasing occurrence of asthma not only in children but in many sections of the adult population—with the smokestack waste from the Ohio Valley region in particular being fingered as the source of the well-known problem of "acid rain." This is an almost entirely anthropogenic phenomenon widely detected across eastern Indiana state; in southern Michigan; in the Ohio regions of the catchment zone of the Ohio River Valley; in southwestern Ontario and southern and eastern Quebec areas of central Canada; in the northern New York, Vermont, and New Hampshire parts of New England; in the catchment area of the St. John River system in Maine, Quebec, and New Brunswick; and in the lakes and streams of the Cobequid Hills of northern Nova Scotia. A number of ingenious methods were developed to neutralize the quite varied effects of such precipitation on everything from groundwater quality to the health of the southern boreal forest cover of eastern North America in general and its maple and other hardwood tree growth in particular. Notably, however, the main positive changes were noticed only after coal-burning processes were replaced with natural gas in most of the Ohio Valley plants. To render conversion of these plants profitably from coal to natural gas and petroleum as sources of thermal energy, and largely eliminate at the same time the burden of potentially ruinous financial liability facing the utilities operating these facilities, an SO_2 emission credits exchange was initiated between the utility operators and gas and petroleum suppliers. A further careful and reasoned calculation of all the costs and benefits derivable from taking the particulates vented as smokestack waste in this major acid-rain vulnerable zone and recapturing them for production of nanomaterials has yet to be organized or implemented, but such a process would likely prove most instructive as well as enlightening on how to proceed with removing the remaining inflexibilities in current production, delivery, and pricing arrangements for natural gas.

Science and engineering issues of technological development have become consciously disconnected from rational considerations of the overall aims and needs of human social life, a disconnection that leads to, and feeds, an incoherence that becomes increasingly intolerable in our current "Information Age," when information as the building block of useful knowledge and understanding has become available in quantities and detail as never before—but the research and thinking needed to make such a flood of information coherent, comprehensible, and useful remains an extremely rare quantity indeed. The working engineer must never forget that there are alternatives, but, remarkably, while the EPA website generated 5,443 "hits" in response to a query for article titles from its libraries that deal with "glycol and natural gas processing," a request for article titles discussing "glycol substitutes" returned a grand total of 2 "hits."

In the course of addressing gas-pricing issues in this chapter, some key elements of the massive disinformation attending the entire notion of "renewable" versus "nonrenewable" energy sources will be mentioned. The price mechanism and price formation for these resources as energy commodities do not depend mainly or solely on any of their inherent material characteristics as raw materials, nor on their particular properties as energy sources. Nor do these price mechanisms and price formation depend purely or even mainly upon the conditions limiting or enabling how humanity accesses them. Not the prices themselves, but certainly the processes of energy commodity price formation and associated price movements, remain circumscribed to an uncannily large degree by the evolution of energy policy in the context of world-altering developments at the level of international politics and economics as they emerged at the end of the First World War. The decisive role of the overall economic system in these price mechanisms and price formation has been greatly influenced by this context, even into our own time. The leading features of this overall economic system have exercised distorting influences over the engineering approaches taken to the development of these resources. Practical as the concerns must be of the working engineer, an essential barrier to humanity achieving a truly functional as well as just solution to our current energy supply-demand conundrum is to be found in this historic legacy.

2.3 SUSTAINABILITY AND THE INCREASING FASCINATION WITH NATURAL GAS

The entire question of the sustainability of economies dependent on fossil fuels as their principal energy source—be it for transportation, for domestic uses such as heating, cooking etc., or for the provision of electrical power for industrial or domestic markets—is affected and influenced by changes over time in the pricing structure of the production and delivery of these resources. However, there still seems to have occurred no price at which humanity will give up seeking and harnessing these resources. Thus, as influential as shifts in prices and production costs may appear in the short term, the price mechanism as such, left to its own devices, is unlikely to play a determining role in how choices are made between different energy alternatives for the long term. As the long term is exactly what sustainability is all about, however, and as evidence has accumulated about the finitude of existing conventional reserves that are either in production or under exploration, a great hubbub has broken out over whether the world is about to enter the era after which oil production will have "peaked." Matters of this kind insert an element beyond conventional notions of risk assessment that become highly problematic for predicting price trends and resource "futures." At the same time it seems likely that a sustainability factor will have to be considered for longer-term contracts of the kind.

In the previous edition of the present book, virtual annexation of Canadian natural gas by insatiable U.S. demand seemed to loom large. The stagnation that has beset Canadian natural gas exports to the United States, coupled with the unchecked expansion of bitumen exports delivered by pipeline from the Alberta tar sands to synthetic crude oil upgrader facilities in the United States today and to Chinese-bound tankers docking at Pacific coast ports tomorrow, suggests that the role Canadian natural gas will play in the North American energy picture is headed for long-term decline.

2.4 IS NATURAL GAS ALWAYS "NONRENEWABLE"?

Natural gas is often considered, like crude oil, as being of finite availability, since it is "nonrenewable" (unlike solar or wind power, for example). This designation slides in alongside the "fossil fuel" designation commonly also used to describe natural gas. To think, however, of natural gas as a "fossil fuel" is actually a starting point for considerable confusion. It is true that natural gas is found frequently in connection with crude oil. It is also true that crude oil is deposited as the decayed remains of ancient flora, and that these are found in very ancient geological layers of the earth in which fossilized remains also frequently appear.

As a fuel/energy source, however, the valuable active element of "natural gas" is methane. Methane is present throughout the Milky Way galaxy. It comprises a large portion of entire celestial bodies, such as "gas giant" planets like Jupiter. On Earth, it occurs in many places, having no connection whatever with fossil layers in the earth's geology, for example, as a major byproduct of landfills. In the earth itself at very great depths both on land and in the oceans, there are significant quantities of trapped gas seeking a way to be vented. Finally, there is no set formula for "crude oil," which appears in raw material form as an inhomogeneous, usually liquid, mass of carbon and hydrogen atoms in a wide variety of proportions—but the chemical formula of methane is always the same, CH_4, and, moreover, this chemical configuration of a single carbon atom with four available bonding sites filled each by a hydrogen atom is remarkably stable. Only when some pressure confining it, or some level of thermal energy exciting the atoms as a gas, is removed does its physical state become unstable.

In the case of a generally considered fossil fuel reservoir, "nonrenewable" is taken to mean that once a reservoir is commercially exhausted—often it is only possible to extract 20%–40% of the reservoir's estimated crude oil and/or gas contents—it cannot be replaced in a time frame that would be useful to humanity within the time frames that normally condition or dictate modern economic existence. It was formed over geological time spans of billions of years; even the hugest oil sources located on this planet would last less than 200 years at current rates of economic activity.

Of course, much natural gas comes from these same reservoirs, formed over the same lengthy geological time span and existing under pressure at various depths in the bowels of the earth. But whereas crude oil really needed that geological time span to form in liquid state, decaying organic material anywhere can begin to produce methane, the active energetic component of natural gas. Natural gas, in addition to being available in reservoir formations as an energy source, can also be extracted as an output of a process of waste conversion.

There is a much homelier example of exactly the same idea for a source of natural gas: landfills. It appears that there is no way modern human societies are not going to produce landfills. One can treat them purely as a waste burden on society, or research their possibilities and innovate. Russia, already rated as one of the world's largest single reserves of conventional sources of natural gas, has a further reserve in municipal solid waste landfills. Although enormous quantities continue simply to be incinerated, the largest biogas complex in Europe comprises two plants on the outskirts of Moscow fed from some of this waste (Izrael et al., 1997).

A previous version of this chapter published in the last edition of the present book cited the evidence of a 1993 EPA study not published until 1999 pointing to the "downsides of further development of natural gas as a substitute for refined petroleum" that have "become undeniable and that emanate from the cartelized development of the petroleum industry." It was also mentioned in this connection how "some partisans of this line of development assert that natural gas can fill the gap so long as it is developed and made available to end users on a competitive, deregulated and non-cartelized basis," and that this itself was "in fact a starting point of yet further disinformation." Now, with the subsequent takeoff in shale gas development based on horizontal drilling and hydraulic fracturing, and massive opposition that this is generating throughout the United States (Barone, 2010), it now turns out that our own antidisinformation league was itself disinformed and had been barking up the wrong tree. The justifications pressed on public opinion—"world oil prices skyrocketing out of control" coupled with blizzards of superficially credible reports about how many byproducts of petroleum refining can be replaced by natural gas liquids (NGLs)—have turned many a head away from noticing the serious environmental blight of compromised groundwater resources, widescale soil destruction, and disruption that have come in hydraulic fracturing's wake.

The actual evolution of the expanded production and distribution of natural gas, organized according to the deregulated model pioneered in the United States, was indeed premised on enabling a "soft landing" for Big Oil on the basis of restructuring government incentives to shift investment from Big Oil into natural gas. Unfortunately for tens of millions of U.S. citizens subsequently affected, this comforting premise tripped over an orgy of hedge fund speculation in natural gas futures on an electronic commodities exchange left unregulated by the U.S. Congress that sent the price of natural gas skyrocketing into outer space until some hedge funds holding very large positions in natural gas futures contracts were caught by a tidal wave of short selling in the market, wiping out billions of dollars "on paper"(Levin and Coleman, 2007).

Once the fullest and most unbiased inventory of all sources of natural gas—be it methane emissions from waste or disused coalbeds, etc.—is compiled, the question as to whether it is indeed economically sounder to continue with conventional approaches or to conduct research into an innovative delivery system can be settled. Elucidating the history of actual development is quite important for shedding light on where thinking became trapped in the past down various blind alleys, and especially where various assumptions about crude oil as a nonrenewable fossil fuel were simply transferred and reapplied willy-nilly to natural gas.

Another seed of yet further confusion is the very label "natural" in "natural gas." The methane fresh from the cow and the methane trapped billions of years ago beneath the earth's surface are doubtless equally "natural," but are their actual potentials as sources of useful energy truly equal? Does pointing out that they both contain CH_4 shut down all further discussion of this temporal dimension of the very formation of these compounds?

Overwhelmingly, the main natural gas sources discussed by and within the oil and gas industry, and within the engineering fields of work connected with that industry, are those that can be extracted in connection with crude oil, using pipeline rights of way or related distribution systems developed initially by oil companies, etc. In the context of the political and economic implications of how strategic energy commodities in general are disposed of on the world scale, these same elements are also significant as this entire aspect emerges as a prominent feature of the overall picture. However, in discussions or considerations in purely engineering-based technical writing on the subject, it is still ordinarily and, it must be said, systematically excluded. This separation of engineering from politics and economics has enabled considerable confusion to grow into outright disinformation. This has spread in the form of two key messages: first, that the oil and gas industry is a hostage to "nature," which possesses only finite quantities of usable or accessible energy sources to meet humanity's infinitely expanding needs; and second, that humanity is the hostage of the oil and gas industry for the foreseeable future as no serious alternative sources of comparable energy content are available or conceivable.

2.5 U.S. NATURAL GAS: PRICING, MARKETS, RISK MANAGEMENT, AND SUPPLY

2.5.1 Some Ongoing Features of Natural Gas Pricing in the United States

As mentioned earlier in this chapter, an outstanding feature of natural gas pricing is the absence of a world price in the sense of the world price that exists for crude oil. For conventional sources of natural gas, the processing into marketable outputs is far less costly than what needs to be done to turn crude oil into refined products, and a great deal of that processing itself is connected relatively closely in time and space with the extraction of the raw material from the ground. Vast tankers or pipelines full of unprocessed natural gas are not required to take unrefined raw material to distant processing centers (refineries). Imports of crude oil are essential to the downstream sector, whereas imports of natural gas generally supplement domestic production and need little further processing apart from possibly having to be converted from liquid back to gaseous form. It is therefore hardly surprising to see the price of natural gas displaying no tendency to converge around a single figure that becomes the market benchmark simultaneously across a wide swath of economies from both the developed and the developing world.

At the same time, the fact is that in the U.S. economy, the rate of growth in the demand for natural gas continues to outstrip the rate of growth in its supply. Thus, traders and wholesale-level interests alike involved in buying and selling natural gas contracts treat the pricing of natural gas much as they treat the buying and selling of crude oil contracts that enter the market according to a similar underlying scenario in which the rate of change in the demand for the commodity continually exceeds the rate of changes in its supply. Thus, basing themselves on an energetic equivalency factor that equates 5.80 million Btu produced from natural gas to the heat equivalent of one barrel of crude oil, natural gas spot-market participants buy and sell contracts at prices whose changes mirror directly current spot-market price movements for crude oil. Of course, it cannot be overemphasized that it is not the absolute price, but rather the *changes* in the price of natural gas that track not the absolute price, but rather the *changes* in the price of crude oil. Reflecting the overall decline in demand for oil as residential heating fuel relative to demand for natural gas, the large increase in winter heating season demand for natural gas, especially in the northern portions of the United States during the northern hemisphere winter, adds higher seasonal variability in demand for natural gas, just as the northern hemisphere summer season is usually accompanied by an uptick in the price of crude oil contracts and a simultaneous decline in the price of natural gas contracts. Apart from these seasonal influences on the contract price, the rates of increase/decrease in the price of natural gas contracts track upward/downward mostly in conjunction with the trend in crude oil supply contracts based on the world price.

The temporary tripling of the Henry Hub spot price for natural gas disclosed in Figure 2-1 demonstrates the extreme potential effect of a short-term event—in this case, Hurricane Katrina which, in the end, was found to have caused little consequential damage to the gas pipeline network along the Louisiana coast, whose throughput at Henry LA forms the basis for the so-called hub price. This single event and speculation before, during, and after it about the potentially disastrous consequences of the 2005 hurricane

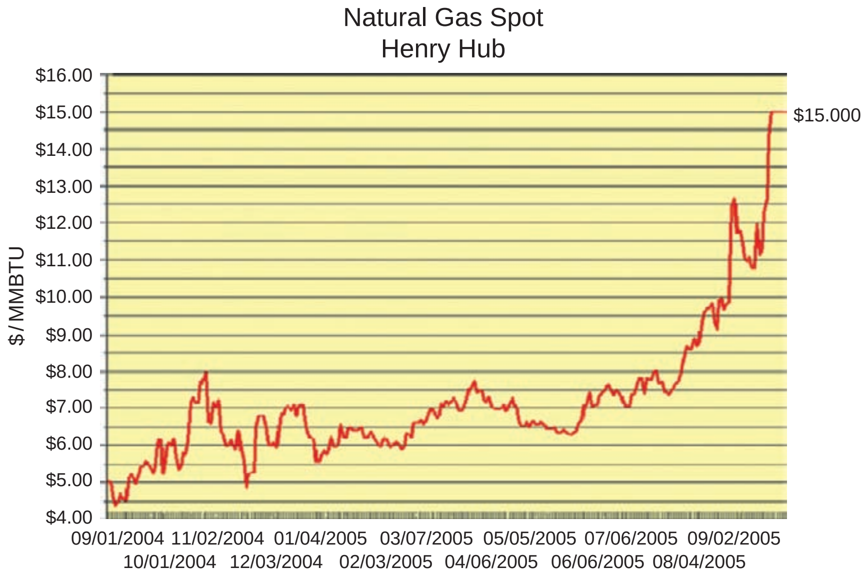

Figure 2-1 Henry Hub natural gas spot market price between September 1, 2004, and September 1, 2005. *(Source: WTRG Economics).*

season in the Gulf of Mexico nevertheless enabled sellers into markets basing themselves on the Henry Hub to double the price from the US$5.00–$7.50/Mcf band in which gas had traded from September 1, 2004, until the third week of August 2005. If there were such a thing as a world price for natural gas, the oscillation in the market price would likely be considerably dampened. For example, there was in the same time period an even shorter-lived "speculative premium" that bumped the price of oil from US$65.00 to US$70.00 per barrel, but it lasted less than 72 hours.

Occasional short-term perturbations in the gas market punctuate the overall picture, but these are actually the moments when the anarchy inherent in the market system unveils its whip hand. Thus, contrary to what was promised and predicted, deregulation of gas and electricity prices has essentially failed to bring about any significant orderliness in gas marketing. Indeed, this conclusion could certainly have been suspected from a brief consideration of the financial earthquake unleashed by the bankruptcy of Enron, a pure "deregulated market" player, and its various consequences and ripple effects throughout corporate boardrooms across North America. The subsequent disaster played out mostly during 2006–2007 at the Amaranth hedge fund speculating in natural gas futures further serves to reinforce this conclusion.

2.5.2 U.S. Energy Markets: The Regulation-Deregulation Nexus

As Table 2-1 illustrates, from the outset, the rising importance of gas within the U.S. energy supply picture could not long be separated from, and was actually decisive in bringing about, challenges to existing regulation. The first 30 years of this process developed from the supply of gas as a customer service being regulated as a residential utility, like electrical service, to its delivery being regulated across state lines, on the basis of preventing the emergence and dictate of oligopolistic firms operating as a restraint upon free competition. The exigencies of wartime sequestering of petroleum, to meet the demands of modern motorized forms of warfare on land,

Table 2-1 Initial Development and Regulation of Natural Gas in the U.S.
A Timeline, Compiled from Fleay (1998) and Natural Gas Organization (2005)

	Development and Regulation of Natural Gas
1859	Col. Drake, U.S. Army, drills first commercial oil well at Spindletop, Pennsylvania. Ambient shows of natural gas are “flared off.” This technique comes to be applied generally in oil fields until the start of World War II.
1900s	Starting in New York and Wisconsin, municipalities enable articulation of gas pipeline networks into urban neighborhoods to provide an alternative to coal for domestic cooking and heating.
1935	U.S. Federal Trade Commission (FTC) report notes with concern the increasing merger of financial interests from the electric and gas utility industries and recommends intervention to regulate in the interests of consumers.
1938	U.S. Congress passes Natural Gas Act (NGA) in response to the FTC recommendations and continuing widespread public agitation over cartelization of utilities raising the cost of delivering an essential service on an uncompetitive basis at a time of widespread impoverishment and want. Drawing on the model of the Interstate Commerce Commission Act of 1913, the NGA proposes to arrest the negative consequences of their excessive concentration by empowering the Federal Power Commission to regulate interstate development of pipeline networks and sale of gas.
1941–1945	The United States enters the war against Japan. Within a year, as it becomes clear that domestically produced oil will be needed for the war, especially for the fleet, the U.S. government mandates widespread development of natural gas in the domestic market as a wartime substitute for refined petroleum wherever possible. In the U.S., a Supreme Court ruling regulates the wellhead price of gas where the utility and the pipeline are part of the same financial entity.

Continued

Table 2-1 Initial Development and Regulation of Natural Gas in the U.S.—cont'd

1954	Supreme Court renders its decision in *Phillips Petroleum Co. v. Wisconsin* (347 U.S. 672 [1954]). In this decision, the Supreme Court ruled that any natural gas producer selling natural gas into interstate pipelines fell under the classification of "natural gas companies" in the NGA, and were subject to regulatory oversight by the FPC, simultaneously streamlining regulation of both the rates at which producers sold natural gas into the interstate market and the rates at which natural gas that was sold by interstate pipelines to local distribution utilities. This sets the stage for the issue on which the deregulation melodrama unfolds, as a liberation of the producers from the tributary yoke of the pipeline companies backed by the FPC.
1970	U.S. domestic oil production reaches a peak from which it has declined without interruption to date. Energy industry planning of domestic supply centers hereafter around natural gas.
1998	The role of oil in supplying the commercial energy of the entire globe had fallen to about 40%; natural gas had risen to about 22%. Some 60% of the consumption of oil has been taken up by transport systems—rail, sea, and air.

sea, and in the air, opened the prospect of gas being substituted for oil in the supply of electric power to industry and local utilities. By the 1950s, the bias of regulatory effort shifted toward attenuating tendencies among producer oligopolies, pipeline oligopolies, and supplier oligopolies to poach one another's turf. Conventional economics presents natural gas production and distribution as a "natural monopoly," where the costs of business entry and start-up, and the likely diminution of any newcomer's margin of profit, discourage establishment of competing enterprises in the same market. This fails, however, to explain and actually confuses historical facts. Natural gas went from being highly regulated to being deregulated as its industrial structure was transformed from scores of small local distributors to a network dominated by a handful of major pipeline operators.

As the point of market saturation and the peak in the increase of the rate of return on investment in natural gas production, processing/transport, and delivery were passed, the pressure grew to consider deregulation. This pressure greatly intensified as it became apparent after 1970 that the United States could no longer profitably increase its domestic production of petroleum. Table 2-2 documents the manner in which, since 1978–1985, deregulation has been introduced.

Table 2-2 Timeline of Natural Gas Deregulation in the United States
Compiled from LIHEAP Clearinghouse (2005)

	A Condensed Deregulation Timeline
1978	Natural Gas Policy Act ends federal control over the wellhead price of "new" gas as of January 1, 1985, but keeps in place wellhead price controls for older vintages of gas. According to the American Gas Association, the leading natural gas producers' cartel in the United States, with this administrative change of everything except the wellhead price for existing supplies, "the laws of supply and demand begin to work again in the natural gas industry." *[This logic brings to mind the old American one-liner: "But apart from that, Mrs. Lincoln—how was the play?" – Ed.]*
1985	Order 436, issued by the Federal Energy Regulatory Commission (FERC), establishes a voluntary program that encourages natural gas pipelines to be "open access" carriers of natural gas bought directly by users from producers. This order brings the separation of pipelines' merchant and transportation functions, and it initiates a restructuring of the natural gas industry's regulatory structure.
1989	Natural Gas Wellhead Decontrol Act lifts the remaining wellhead price controls on natural gas.
1992	FERC Order 636 orders interstate natural gas pipelines to "unbundle," or offer separately, their gas sales, transportation, and storage services. The goal of this order is to ensure that all natural gas suppliers compete for gas purchasers on equal footing.
1996	The first residential natural gas customer choice programs are implemented. By 1997, local natural gas utilities in 17 states and the District of Columbia had proposed and/or implemented such residential customer choice policies or pilot programs. On August 31 and September 1, the Legislature in Sacramento, California, the most populous state in the United States, authorizes deregulation of electric power rates to begin January 1, 1998.
1999–2000	During one of the hottest summers on record in the State of California, a private corporation in Houston, Texas, set up under the name Enron to purchase electric power from deregulated utilities and resell it to customers in other deregulated jurisdictions, corners the California market, and people's air conditioning bills skyrocket. Mustering the facts of the brownouts and related power cuts and outages caused by the unprecedented demand in conditions of such artificially restricted supply as evidence of the need for massive increases in power supplies generally, lobbyists for the power engineering cartel in Washington, DC, put the case for expanded federal subsidies and incentives to construct new power plants across the United States. As an unexpectedly long cold snap hits the populous regions of the U.S. eastern seaboard and southeast in late 2000, the Federal Energy Regulatory Commission permits a fourfold increase in the wellhead price (Henry Hub) of natural gas—the main proposed feedstock for the proposed future plants.

2.5.3 Energy Price Volatility and Derivatives

It has been clear for some time that when it comes to energy sources and the maintenance of upstream investment and downstream revenue, the market left to its own devices cannot deliver. One of the main ways developed to overcome this gap has been the development of risk management strategy as a major part of managing corporate-scale interventions in energy markets.

Derivatives, conventionally defined as "financial instruments (contracts) that do not represent ownership rights in any asset but, rather, *derive [Emphasis added – Ed.]* their value from the value of some other underlying commodity or other asset," have been around since the late 1800s in the United States as a tradable agricultural commodity, the most famous example being probably the "pork belly" contract (EIA, 2002). Within the envelope of the previously described events, and tending to overstimulate the marketing and pricing of both natural gas and electricity, a derivatives market was developed extensively as an instrument for the hedging of risk among entities investing in areas of the energy market undergoing deregulation, especially electricity and natural gas.

Modernized, streamlined derivatives trading, and especially the associated practices of indexing arbitrage and what is known as "program trading" (the preset, computer-regulated selloff by syndicates of banks, investment dealers and other large-scale purchasers of shares, usually near the end of a financial quarterly reporting period, of tens and even hundreds of millions of shares of stock in a vast range of entities), were first seen in the stock exchanges of the United States on a large scale during and following the October 1987 stock market crash. However, indictments and convictions shortly thereafter of Michael Milken, Ivan Boesky, and other exchange insiders for abusing the latitude available for those prepared to indulge new areas of risk in the market, far from discouraging their development, generated new interest in taming and stabilizing the playing field to facilitate new entrants. The risks involved are greatly multiplied by (1) the mutual and often reciprocal dependence of different contract performance promises on one another, and (2) the legal system's general posture that contracts between legal entities are deemed to have been entered into freely until otherwise proven in a court of law, and therefore it is no outside body's business to inquire into or approve in advance any of the terms or requirements of such contracts.

The single most dramatic recent example of what can happen when "contracts" are simply invented without any intention of performing or completing them was seen in the incredibly rapid expansion and equally dramatic implosion of Enron, the derivatives-based energy commodities trader in Houston, Texas, between 1997 and 2001. After the implosion of Enron, the EIA commissioned an authoritative study that exculpated derivatives trading as such of any responsibility. Entitled "Derivatives and Risk Management in Energy Industries," the report documented the litany of disasters that have trailed in the wake of liberating deregulated energy commodity markets into the eager and waiting arms of the derivatives traders thus:

> *Although derivatives meet legitimate needs, they have also been implicated in tremendous losses. For example, Orange County, California, lost $1.7 billion in 1993; Metallgesellschaft lost about $1.3 billion in 1993 in energy trading; and in 1998 the Federal Reserve Bank of New York organized a rescue of Long Term Capital Management in order to avoid disrupting international capital markets. And in 2001 Enron became at that time the largest bankruptcy in American history. Enron was a large user and promoter of derivative contracts. Although Enron's failure was not caused by derivatives, its demise raised significant concerns about counterparty (credit) risk and financial reporting in many energy companies.*
>
> ***EIA, 2002***

The report explains the overall situation of energy commodity risk management and derivatives from the government's point of view as follows in the report's Summary:

> *Natural gas and oil pipelines and residential natural gas prices are still regulated. Operating under government protection, these industries had little need for risk management before the wave of deregulation that began in the 1980s—about the same time that modern risk management tools came into use.*
>
> *There are five general types of risk that are faced by all businesses: market risk (unexpected changes in interest rates, exchange rates, stock prices, or commodity prices), credit/default risk; operational risk (equipment failure, fraud); liquidity risk (inability to buy or sell commodities at quoted prices); and political risk (new regulations, expropriation). Businesses operating in the petroleum, natural gas, and electricity industries are particularly susceptible to market risk—or more specifically, price risk—as a consequence of the extreme volatility of energy commodity prices. Electricity prices, in particular, are substantially more volatile than other commodity prices.*
>
> ***EIA, 2002***

Of course, with the increased use of natural gas as a fuel for new electric power plants, the risk-reward "game matrix" that might be developed to predict investment outcomes for the widened variety of possible scenarios became ever more convoluted, with changes in electricity prices capable of leading or following changes in natural gas prices, and vice versa.

> *Price volatility is caused by shifts in the supply and demand for a commodity. Natural gas and wholesale electricity prices are particularly volatile, for several reasons. Demand shifts quickly in response to weather conditions, while 'surge production' is limited and expensive. In addition, electricity and natural gas often cannot be moved to areas where there are unexpected increases in demand, and cheap local storage is limited, especially for electricity.*
>
> ***EIA, 2002***

Figure 2–2, comparing the standard deviation of percentage changes in annual average prices for a number of different traded commodities over the past decade or so, demonstrates the potential profitability of short-term investing in certain electricity markets over investing in practically any significantly traded other "commodity play" in U.S. markets.[5]

The conventions of risk-management strategy are thus very much in flux in U.S. energy markets, a situation that has become increasingly fluid as interest generally in the field of econometrics about the relationships elucidated by game theory regarding market uncertainty has extensively developed.

2.5.4 Natural Gas Supply in North America

The role of the United States in natural gas production, marketing, and pricing differs in significant respects from the picture just described. It has developed a cartelizing model that has successfully and—in many respects—falsely presented itself as a model and high point of deregulated competition. Figure 2–3a gives an idea of how the United States supplies

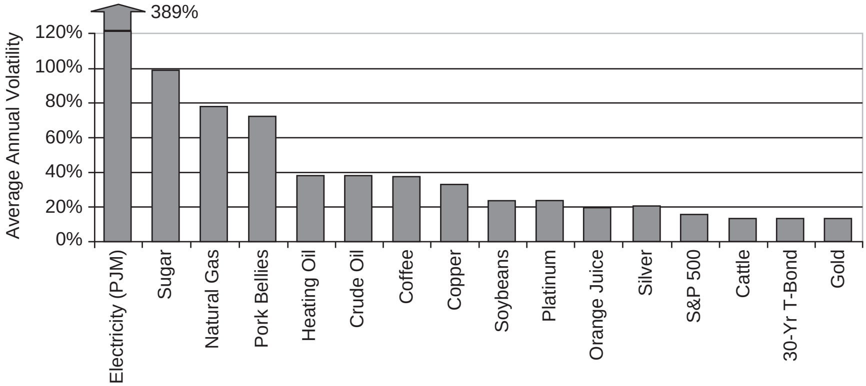

Figure 2-2 Average spot-market price volatility for selected commodities (Bolinger et al., 2003).

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[5] Figure 2–7, from the EIA (2002), was created by Bolinger et al. (2003) prior to the latest natural gas price spike in February 2003.

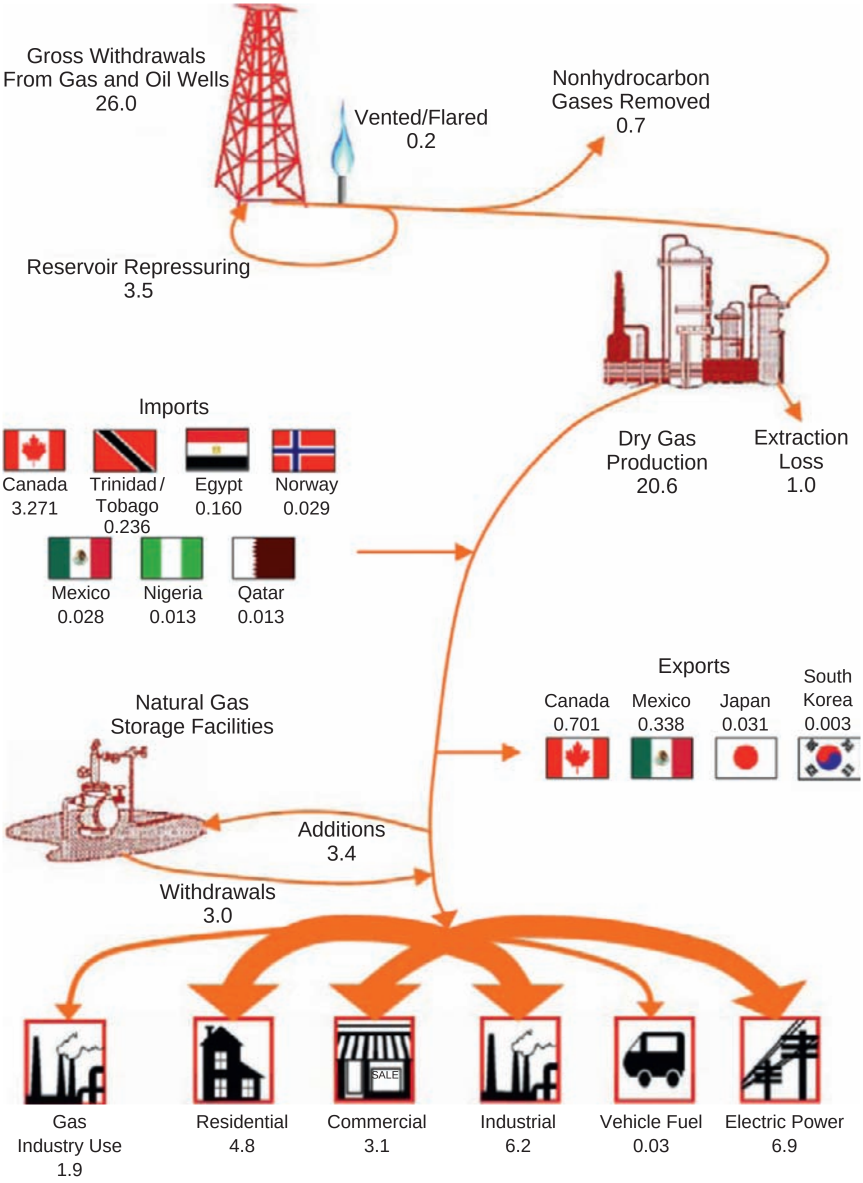


Figure 2-3a U.S. production, consumption, import, and export of natural gas (2009). Quantities are in trillion cubic feet. *(Source: Energy Information Administration).*

its market and hints at how this develops in a manner that also integrates entire portions of the economies of other countries, notably its geographically contiguous neighbors Canada and Mexico, into the economy of the United States.

2.5.4.1 The Special Position of the United States

Annual market consumption of natural gas in the United States increased from about 5 trillion cubic feet in 1949 to a peak of more than 22 trillion cubic feet in 1973. After falling back throughout most of the 1980s and 1990s, it resumed a steady upward climb again to its next peak, at about 23.3 trillion cubic feet in 2000 (U.S. Department of Energy, 2005).

Two factors account for most of the post-war expansion: the explosion of the housing market throughout the United States, especially during the main years (1945–1960) of the so-called "baby boom"; and the expansion of medium and heavy industry beyond the coal-and-steel belts of the U.S. northeast and Great Lakes to new urban growth centers of the West Coast, the "right-to-work" states in the South, and the tremendous expansion of population and industry in Texas and Florida.

By the mid-1990s, however, some 70% of new homes in the United States were being heated with natural gas. Essentially, this market had reached saturation point. At the same time, as a result of the so-called "deregulation" of natural gas pricing begun in the U.S. in 1978 and taking effect after 1985, a situation was created whereby the natural gas producers, after having divested themselves of any significant further financial involvement pipeline and/or other distribution networks and infrastructure, were deemed to be no longer in a cartelizing situation (Udall and Andrews, 2001).

At this point, arguing that they would now need to be able to compensate their operations for the future losses in future revenue that could be anticipated from the decline in household demand, the gas producers became free to combine in setting the wellhead price. This—the so-called "Henry Hub" price, based on delivery at the main hub on the U.S. Gulf Coast, where comes some 60% of U.S-produced natural gas—quadrupled in 2000. Far from ensuring themselves against future losses of revenue from declining home heating demand, the producers were in fact harvesting the windfall created by deregulation of electricity prices, provision of services, and markets in major populated states, starting with California. To keep electric power generation profitable in a deregulated market, natural gas was increasingly substituted for oil and coal. Natural gas would thereafter be produced increasingly as a feedstock commodity for the electricity-generating market, with home heating markets concentrated in the Northeast—where population was in long-term decline—contributing a diminishing portion of overall revenue (Udall and Andrews, 2001).

Deregulation of electrical power supply and markets opened massive opportunities for outright swindling of the public on an unprecedented scale. The most dramatic result, which came quickly in 2000–2001, was the rise and collapse of Enron—headquartered in Houston at the center of the natural gas boom, but doing most of its business in California in the form of Ponzi-type "contracts for future supplies of electric power." With its collapse, natural gas prices fell back 50% from their historic highs, but superprofits were garnered in any event from the happy confluence during the winter of 2000–2001 of an extremely cold and extended heating season in the U.S. northeast along with the Enron-induced brownouts and blackouts visited on residents of California (Udall and Andrews, 2001).

The United States' "long-standing glut of natural gas and electrical capacity, along with the world's spare oil capacity, vanished simultaneously in spring 2000. . . . Fueled by cheap energy, the U.S. economy grew 60 percent since 1986, an astounding five percent in 2000 alone. Gas consumption grew 36 percent over that period (+43.8% over the period 1986 to 2000). But it was the demand for electricity . . . up 5.4 percent in 1998, an astounding rate for such a large economy . . . that has had the biggest impact on gas prices. To meet . . . growing electricity needs, utilities have ordered 180,000 Megawatts of gas-fired power plants to be installed by 2005. It was a logical thing to do: gas is the cheapest, cleanest way to convert fossil fuel to electricity. But if ordering one gas turbine makes perfect sense, ordering 1,000 is a recipe for disaster. No one in the utility industry asked the key question: can we produce enough gas to run all those plants? Many experts think the answer is no. . . ." Reproducing a previously published 20-year projection of natural gas consumption in the United States for the period 1997–2017 estimating relatively moderate increases in annual industrial consumption (from about 8.7 to slightly more than 9 trillion cubic feet), residential use (from 5 to 5.5 trillion cubic feet), and commercial applications (from 3.2 to 3.5 trillion cubic feet), this same source indicates that gas for electricity generation is predicted to nearly triple over the same period, from about 3.2 to more than 9.1 trillion cubic feet per annum (Udall and Andrews, 2001).

The importance of a distribution network cannot be overstated when it comes to natural gas.

> *Without a pipeline, natural gas is worthless, a constraint first recognized by the Chinese. They were drilling for gas in 1000 A.D. . . . but their pipeline materials were limited to bamboo [sic]. American pipelines today could stretch to the Moon. Most date to post-World II, when Gulf Coast supplies were tied to markets in the Midwest and New England. Since pipelines are prone to corrosion, beer keg-sized diagnostic tools called "pigs" are pushed through the lines to search for weak spots, not always in time. In August, 2000, a pipeline exploded in New Mexico, killing 10 people, and crimping gas deliveries to California. Many aging pipelines need to be rebuilt, replaced, or expanded to deliver more gas to urban areas, where the new fleet of gas-fired power plants will be moored. In December 2000, gas delivered to L.A. briefly fetched $69, equivalent to $400 for a barrel of oil.*
>
> **Udall and Andrews, 2001**

By 2009, i.e., about a decade since the data just discussed were recorded, the relative positions of the various offshore sources and export destinations of natural gas in the U.S. market had changed somewhat relative to one another, as the data in Figure 2-3b disclose. That portion of U.S. domestic residential and commercial heating end uses for natural gas had rapidly expanded and crashed in the wake of the Enron implosion in 2001 and the aftereffects of the Amaranth hedge fund collapse in 2006–2007, i.e., just before the current recession took hold in the United States. (The further development of industrial end uses for natural gas within the United States presents a more mixed picture, reflecting the fact that so much of the country's manufacturing base moved abroad in the last decade.)

The most remarkable feature to emerge in U.S. natural gas usage over the last half-decade, however, has been the clearly rising trend that has become established with regard to the expanded reliance on domestic sources of natural gas production (clearly illustrated in Figure 2-3c). That reflects the trend that continues increased U.S. production of so-called unconventional sources of natural gas, meaning gas extracted from shales, utilizing the newest technical developments in horizontal drilling in combination with the highly controversial practice of hydraulic fracturing heavily criticized in the popular 2010 movie *Gasland*.

2.6 NATURAL GAS IN EURASIA: THE SPECIAL POSITION OF POST-SOVIET RUSSIA

When the situation of natural gas is compared to oil, the contrast could not be more striking. Even for what are commercially defined as natural gas sources, there is no shortage of supply, just some distribution constraints

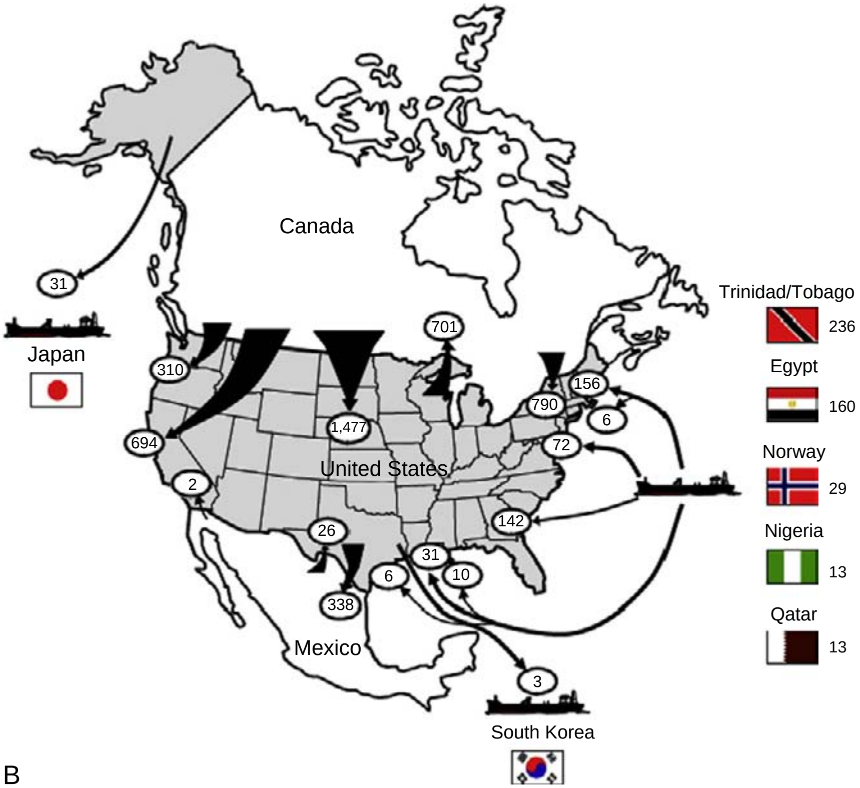


Figure 2-3b U.S. import and export of natural gas (2009). Quantities are in billion cubic feet. *(Source: Energy Information Administration).*

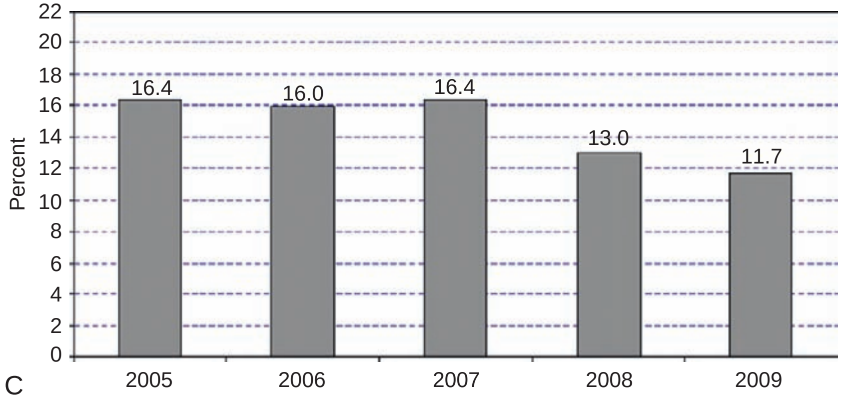


Figure 2-3c U.S. net imports as a percentage of total domestic consumption of natural gas, 2005–2009. *(Source: Energy Information Administration).*

available for some degree of artificial and largely temporary manipulation. Most remarkably, one of the zones of greatest commercial supplies is the territory of the former Soviet Union, in which there never developed the upstream-downstream separation that facilitated U.S. domination, the manipulation of excess demand for oil (EDO) and the ever-expanding gap between the stagnation in real constant-dollar costs of production in oil-exporting countries and the price extracted for products in the consuming countries.

Another factor further transforming the picture stems from the absence of the corporate as well as government-owned sectors of the oil and gas industries of post-Soviet Russia from any of the existing U.S. or U.S.–European production or marketing cartels. The territories of the former Soviet Union (including the separate republics of Central Asia that have emerged since 1991) possess the largest conventional oil reserves outside OPEC and the second-largest proven reserves of natural gas anywhere on earth. Its emergence as a major world-market "player" in its own right in the production and sale of crude and refined petroleum for markets in Europe, and especially for the fastest-growing markets in China and the Asian subcontinent, challenges both the demand-management model used by OPEC and backed by the U.S. oil majors to set and maintain a world oil price. As for natural gas, Russia provides the geographic linchpin for delivering to markets anywhere in the Eurasian land mass.

Post-Soviet Russia's situation represents one of the foremost features of the world picture of commercial production of natural gas. Russia's natural gas industry, a monopoly run by Gazprom, which controls about 86% of the country's marketed production, is the second major source of the world's commercial production of natural gas and the world's leading exporter of natural gas. Furthermore, within Russia, natural gas is the predominant industrial fuel, accounting for nearly half of the country's domestic energy consumption.

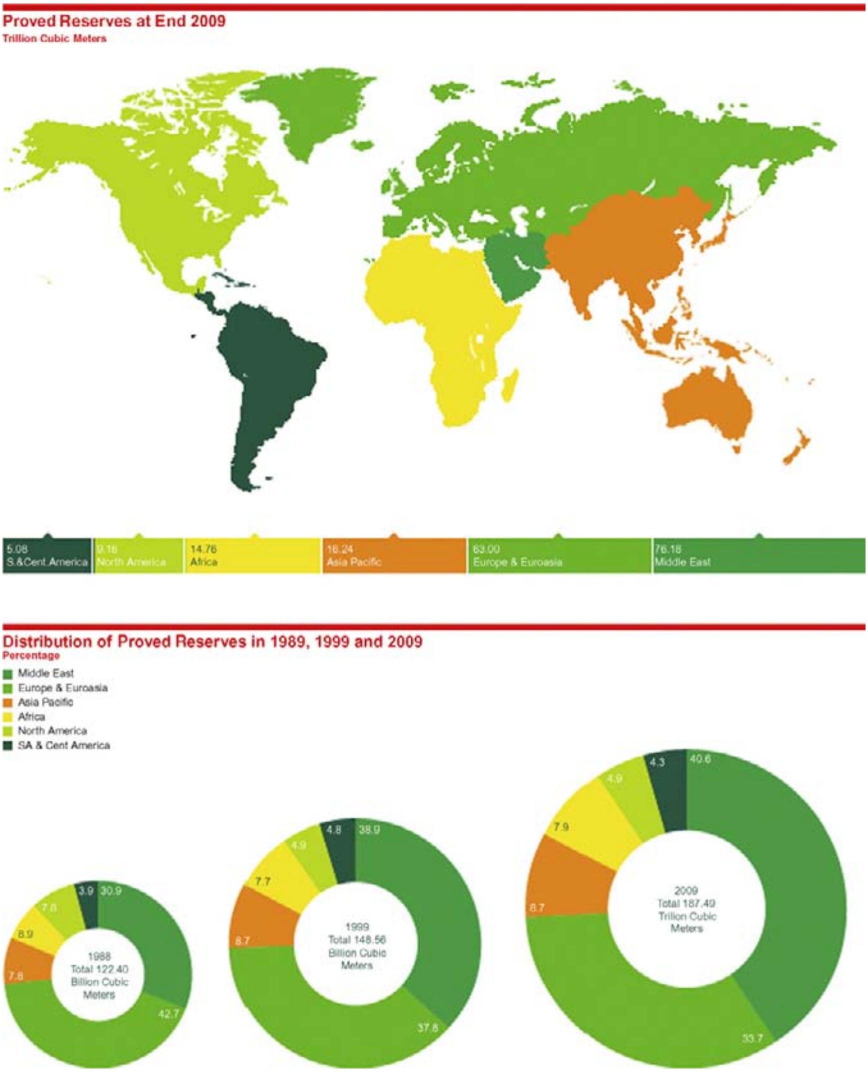
Another remarkable piece of the natural gas marketing picture related to this is the situation presented by Western Europe, which possesses only about 5% of commercial natural gas resources but has an insatiable appetite for this form of energy. The gas industry in Europe consists mainly of downstream activities undertaken by transmission and distribution companies. Before the current economic downturn, more than 30% of gas consumption was being met by pipeline imports from the former Soviet Union and Algeria as well as liquefied natural gas (LNG) imports from North Africa.

For the more than 4 billion people living on that land mass, this fact must over time erode any notion of any purely local market for natural gas.

The global distribution of commercial natural gas and its production are described later.

Proven natural gas reserves—those that could be economically produced with the current technology—are illustrated in Figure 2-4.

At the end of 2009, the former Soviet Union held the world's largest natural gas reserves, 31.2% of the world's total. Together with the Middle



Proved Reserves at End 2009

Trillion Cubic Meters

Distribution of Proved Reserves in 1989, 1999 and 2009

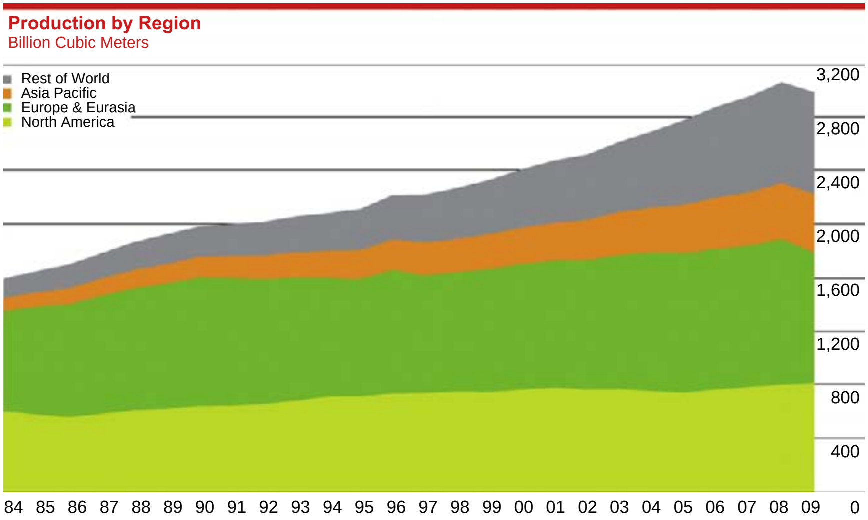
Percentage

Figure 2-4 Distribution of proven natural gas reserves in 2009. *(Source:* BP Statistical Review of World Energy, *2010).*

East, which holds 40.6% of total reserves, they accounted for 71.6% of world natural gas reserves. In 2009, total world reserves were 187.49 trillion cubic meters; global reserves increased more than 50% in the last 20 years (from 122.4 trillion cubic meters). The world's ratio of proven natural gas reserves to production at current levels—the time that remaining reserves would last if the present levels of production were maintained—has remained at about 60 years through the past decade.

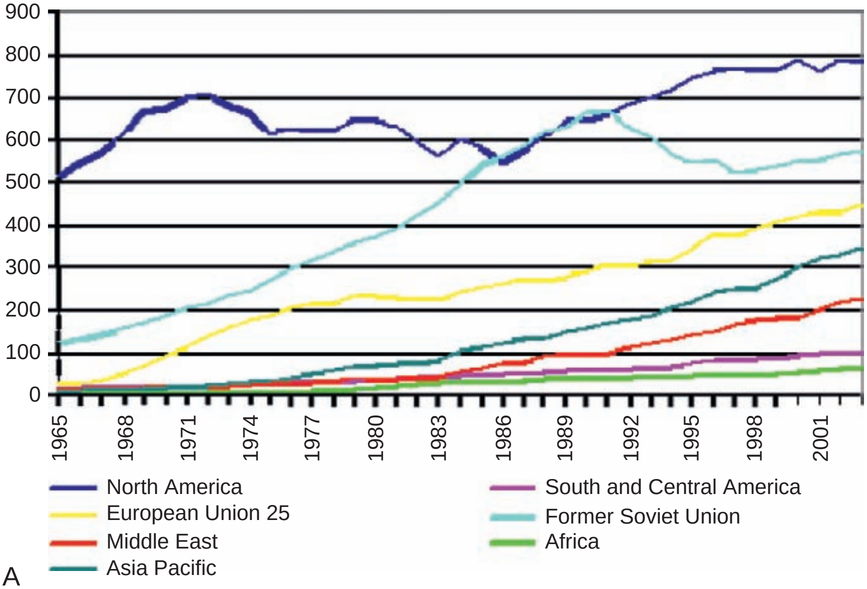
Figure 2-5 illustrates the relative proportional contributions to the world's total production of natural gas in 2009. According to the data published in the *BP Statistical Review of Oil and Gas* for 2009, among the top 10 producers were the United States (20.1% of world production) and the Russian Federation (17.6% of total production). Other major producing countries are Canada (5.4%), Iran (4.4%), Norway (3.5%), Qatar (3.0%), China (2.8%), Algeria (2.7%), Saudi Arabia (2.6%), Indonesia (2.4%), and the Netherlands (2.1%). These 10 countries accounted for two-thirds of total natural gas production in 2009.

Natural gas accounts for almost a quarter of the world's energy consumption. As clearly shown in the Figures 2-6a and b, consumption of natural gas has increased considerably in the last 30 years.



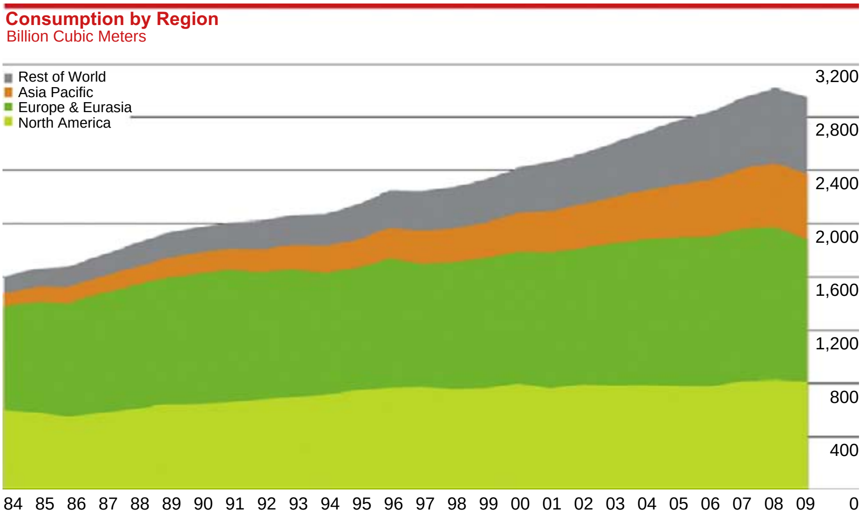
Global natural gas production fell by 2.1% in 2009, the first decline on record. Production fell sharply in Russia (–74.2 bcm) and Turkmenistan (–29.7 bcm), in each case the largest decline on record. The U.S. recorded the largest increase in the world for the third consecutive year.

Figure 2-5 Natural gas production 1984–2009, billion cubic meters. *(Source:* BP Statistical Review of World Energy, *2010).*



A

Figure 2-6a Natural gas consumption, billion cubic meters. *(Source: UNCTAD based on data from BP Amoco, Statistical Review of World Energy, June 2004).*



World natural gas consumption fell by 2.1%, the most rapid decline on record and the sharpest decline among major fuels. Russia experienced the world's largest decline, with consumption falling by 26.3 bcm. Consumption growth was below average in every region.

B

Figure 2-6b Natural gas consumption, billion cubic meters to 2009, clearly showing the dip in global consumption. *(Source:* BP Statistical Review of World Energy, *2010).*

As Figure 2-6a illustrates, the world's main consuming countries in 2000 were the United States, accounting for 27.2% of total consumption, and the Russian Federation, with 15.7% of total consumption. Total world consumption back at that time was 2,404.6 billion cubic meters. This consumption trend, including the dominant position of the U.S. and the Russian Federation in first and second place, respectively, continued steadily upward until 2008, at which point the impact of the global financial crisis was felt and, as illustrated in Figure 2-6b, total consumption declined.

Behind what the scale of Figure 2-6b renders as a seemingly small downward hitch, however, lurks the loss or reassignment of tens of billions of dollars in investment capital from the enormous financial crash of 2008–2009. The commanding heights of the Russian natural gas industry, meanwhile, remain largely under central government direction. Moreover, as this is being written (in mid-2011), recovery in most areas of basic industrial production still seems months and possibly years away in much of the United States, Canada, and Europe. Such recovery is fundamental to investment banks and other consortia taking major decisions regarding expansion and upgrading of the European connections to these pipeline networks that have so insinuated themselves into the center of international trade in natural gas. The Russian system has in the meantime stepped up its connections and exports to China and Japan and increased frontier exploration activities in its offshore Arctic zone.

2.7 LOOKING TO NATURE FOR A NEW MODEL

Normally, we would end the chapter on gas pricing right here. However, as much as we may document how it works, the fact remains that the pricing of natural gas up to now has reflected neither the true energy content of the gas, nor the true costs of its production and delivery to consumers. Many costs have come to be added that also added further layers of opacity to the problem. For example, there is a problem with hydrate formation in gas pipelines that can severely restrict flow. Tens of billions of dollars annually are expended on toxic chemicals that are injected into pipelines to arrest, break up, or slow down hydrate formation. This cost is passed on to consumers, and then sometimes the pipeline company—but more often society, usually in the form of government—is then also confronted with paying the costs of environmental damage and cleanup from the toxic chemicals. This adds to the gross domestic product by simply increasing spending, but is it improving the standard of living of society as a whole?

The adding of costs in this manner—and with them, further layers of opacity—is reflecting no significant improvement in either of the two most intangible aspects of the energy commodity, namely, its efficiency or quality. Indeed, this process so effectively marginalizes all discussion or consideration of these aspects that the deeper question starts to pose itself: is any of this entire proceeding on the right path? Resolving this issue, however, is by no means as straightforward as it might at first appear. The commonest approach is the familiar one of "cost-benefit" analysis: the costs and benefits of various paths are totted up, compared, and finally one presumably optimal path is selected. Such quantitative comparison has the advantage of being highly tangible, but what if the problem resides precisely in the tangibility of all the paths thus examined, i.e., the fact that they all quantify on the basis of putting dollar values on external manifestations?

There are alternatives to a number of the sources of added cost that follow a path quite different from that which can be directly engineered through adding chemicals. For example, bacteria can also break up hydrates and return to the environment without doing pollution damage. So far no one has found a way to turn such a thing into a multibillion dollar industry that would deliver a useful result. Some conclude from this state of affairs that such a course is uneconomic, but the fact of the matter is that doing this well would in the end be low cost, its essential mechanism (like many natural solutions) quite self-contained and thus unlikely to be spun into a grand new profit center. Of course, this would also mean that a major increased charge on the final price of gas to the customer would be that much harder to justify in the name of defeating the hydrates menace and saving the environment.

The choices of path cannot be limited to rearranging and manipulating external features, elements that are extraneous to providing an energy source on a continuous and reliable basis. Natural paths are to be preferred over engineered paths, but the driver of the selection process remains "return on investment." Not all engineered paths will deliver an optimum return on investment, but when natural paths are examined, the problem becomes far more complicated. When "return on investment" is the driver, every possible natural path looks hopelessly utopian and impossible to justify.

What is the essence of the problem posed by continuing with energy commodity production on the basis that has developed up to the present? Many question whether any actual social need of society or its individual members is being addressed within the confines of any presently available social or economic model. At the same time, who or where anywhere in this modern world can survive without a system of providing and meeting energy needs? Could it be the very fact that the return on investment is the driver is ensuring that, in the end, no actual need or needs of society or its individual members is or are being, or can be, addressed?

Hydrates, like many of the other hazards attending the exploration, production, distribution, and consumption of natural gas, petroleum, and its byproducts, become irremovable singularities that can only be dealt with either at the cost of private corporate profit or at public government expense. They are labeled "downsides," and entire forests have been felled to produce the thousands upon thousands of pages of statutes and regulations that are supposed to "manage" the problems created by these "downsides," but nothing fundamental is transformed in the result. Why? Could the answer be: because these approaches all follow one and the same antinature model?

The antinature model takes on too many specific forms to be easily pigeon-holed with a concise definition. However, in the case of energy commodity production, the production of energy mainly as a feedstock to produce energy in another form—oil or natural gas to make electricity, or renewable solar energy for the same purpose—rather than for their highest and best uses directly as energy sources in their own right, is one of the most commonly seen indicators of an antinature path and agenda.

The real issue is firmly and finally to step off this antinature path and renew the knowledge-based, researched approach to extracting solutions that are innovative, economically attractive, environmentally appealing, and socially responsible.

The production, distribution, and supply of energy is the key to modern social and economic existence everywhere on this planet. There is no corner left on this earth where the solution to these problems can be purely local or not connected with the wider world. How problems arising in the production, distribution, and supply of energy are addressed ought to be the heart and soul of solving the problem of raising the true standard of living everywhere, the notion crudely and incompletely estimated by the well-known index forming the ordinate axis in Figure 2-7, namely, the "per capita gross domestic product."

Of course, spending of all kinds is incorporated in the per capita GDP, from prostitution, gambling, and tobacco consumption to armaments production and export to outer space research and medical research, etc. The idea underlying per capita GDP is that even if individuals, institutions,

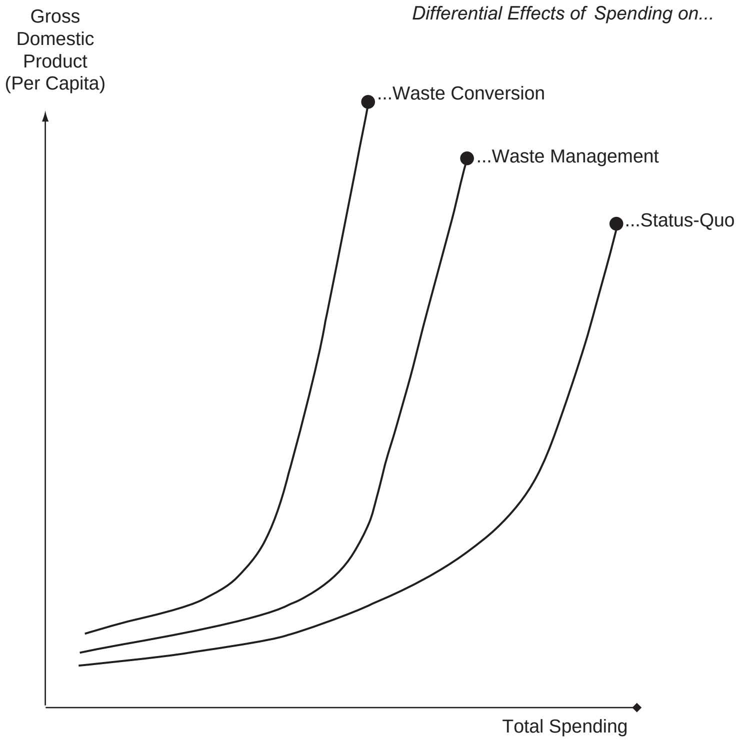


Figure 2-7 "Deeds are but by intention"—illustrating the longer-term effects on GDP of redirecting spending to socially useful ends.

corporations, etc., each have their own priorities, personal or private or social, a meaningful statistic about the general state of social development as a whole can nevertheless be generated by averaging spending on all these wildly divergent agendas over the entire population. The legitimate question raised by this approach is: what social progress other than mere arithmetic increases in spending can be registered by such a statistic? Mixing together privately directed and socially directed forms of spending on the assumption that their very direction, i.e., the agendas informing the actual spending decisions, are a matter of indifference to economic science ensures that spending that is indeed harmful and informed by some private agenda is placed on a par with spending that is socially beneficial and informed by an agenda that is not purely self-serving. Yet it is widely understood as a fact of life anywhere in this world that "good for some and bad for others is bad for everyone." Allegedly "scientific"-minded indifference to this crucial distinction acts as a serious blind to understanding the fundamental issue of path. What the graph brings out is how crucial it is for collective well-being that social concerns are assigned their proper priority: any change in this direction, even just a focus on waste management, would being about an improvement compared to the status quo.

The pro-nature path is quite readily attained by taking a knowledge-based and researched approach. This means starting from the position that nothing that presently exists is the "last word" in technology. One must be ready fearlessly to work out arrangements based on what is actually found to work in nature, not on the basis that nature is broken and needs to be "fixed." Nothing that originates from nature can pollute the natural environment if it is used and returned to nature without in the meantime being artificially reconstituted or otherwise engineered outside the range within which it serves its customary natural purpose(s). Nothing that originates in nature is linear in its native state; if it is linearized and then returned to nature, it will either pollute or cease to be useful. Natural constituents utilized according to their customary functions within nature are, in fact, time-tested. This often ensures overcoming, by entirely avoiding, the costly burden of having to re-engineer short-term solutions developed away from a nature-based path as they inevitably break down and start to fail.

What is not being suggested here is either any romanticized notion about nature as a mysterious force or mysterious treasure house, nor a theological or teleological assumption about the purposes of nature. It is an unavoidable starting point of all scientific investigation, into either social or natural-physical phenomena, that the universe at any moment is a function of space and time: $U = f(x, y, z; t)$. The actual space relevant to any proper specific investigation is not to be confused with, or held to be coequal with universal space, the "x, y, z" of U as just defined. A proper investigation depends on finding ways of investigating—perceiving and measuring—the portion of space that is relevant. The temporal component, however, is a different matter.

The temporal component of some finite process or system being observed or investigated, which can be defined as commencing at t_{initial} and ending at t_{final}, is one thing. However, the temporal component of processes in nature-proper, so to speak, is very different, precisely because—frequently if not indeed customarily—t_{final} is simply unknown. All engineered processes can be very nicely handled according to the paradigm of the $[t_{\text{initial}}, t_{\text{final}}]$ interval, whereas most natural processes can be investigated at best only according to an interval that looks something like $[t_{\text{initial}}, t_{\text{final}\,\{\text{best-guess}\}}]$, where the right-hand side of the interval cannot be closed.

It could very well be that it closes at $t = \infty$. The one thing that is certain is that right-hand ")" can only be replaced with a "]" *after* a true t_{final} value is established, and not before. From the standpoint of establishing scientific knowledge, it will not advance matters simply to assert and assume a t_{final} value and a right-hand "]" without some testing procedure and associated criteria that either establish a t_{final} value or demonstrate that t_{final} only appears or converges at $t = \infty$. The temporal component of a natural process is thus inherently intangible. Usually, at this point, accusations are heard about "introducing matters of metaphysics and religion into science," etc. However, should acknowledging such intangibility have to be the end of further scientific discussion and investigation, or could it not represent a new starting point?

Here, it becomes important to clarify that what is being set forth is very far removed from the notions of intangibility presently attending the entire matter of natural gas pricing. These are quite concentrated, for example, in the specific area of risk management, specifically using derivatives (see Section 2.5.3):

> *The asset that underlies a derivative can be a physical commodity (e.g., crude oil or wheat), foreign or domestic currencies, treasury bonds, company stock, indices representing the value of groups of securities or commodities, a service, or even an intangible commodity such as a weather-related index (e.g., rainfall, heating degree days, or cooling degree days). What is critical is that the value of the underlying commodity or asset be unambiguous; otherwise, the value of the derivative becomes ill-defined [Emphases added – Ed.]*
>
> ***EIA, 2002***

The problem here lies with how intangible elements have been selected in the first place for inclusion among the derivative's components. The effective value of the derivative as a tradable instrument is entirely a function of time rendered in tangible form, namely, the date(s) on which the underlying contract(s) must be completed. The intangibility of the "weather-related index" mentioned in the preceding example, however, is intended to provide the seller of the instrument with an "out," an escape from liability, in the event of a failure to perform the contract. Time as the essential intangible of processing of anything within nature, on the other hand, either adds or subtracts effective value. Issues of liability, much less distancing contracting parties from its reach, do not arise.

Viewed in the light that has been cast here, the time-tested character of natural constituents and sources, before and without further chemically or industrially engineered processing, poses an interesting conundrum: are the best sources of energy locked in nature's newest products or its oldest?

The conventionally trained scientist's response would be that all of this is resolved by establishing the chemical structure: CH_4 from the living cow at around 1 KPa pressure or CH_4 from material that decayed eons ago, lying thousands of meters beneath the earth's surface under confining pressures of thousands of KPa is also CH_4: end of story.

Consider, however, the implications of this claim:

- It is accepted that the entire planetary system and galaxies far beyond this planet and solar system are continuously undergoing change, and,
- it is accepted that animal species have undergone and continue to undergo evolutionary change, speciation etc., but:
- it is not acceptable to assume anything consequential might have happened at the molecular level over the billions of years that passed while some electron continued revolving about some nucleus.

The weight of this logic alone suggests that an important piece of the picture remains to be further investigated. Despite what has become commonly accepted, and that forms the basis of the entire energy extraction and processing sector, the energetic content of various well-defined molecular arrangements of carbon and hydrogen atoms may in the final analysis indeed *not* have been due after all, either mainly or entirely, to how these compounds as raw material were engineered (i.e., processed or refined). Rather, these molecular remains of "dead" matter may have further transformed internally as raw material in the natural environment according to processes still not understood. Just because such material is no longer living organic material may not mean that it could not undergo further internal transformation on top of the transformations being imposed externally by continental drift and plate movement, climate change, etc. Within the field of petroleum engineering research, to dismiss any consideration of the possibility of subatomic intramolecular change in posthumous organic matter left undisturbed over geological periods of time cannot be considered a sign of wisdom. In the present context, it would even seem to serve as evidence for the conclusion that ongoing fanatical adherence to views that subsequent progress in actual research and discovery rendered unsustainable is a syndrome that cannot be attributed to any particular belief system, national or ethnic origin. All that is required to catch this disease is a strong will to remain ignorant.

As energy is in continual demand, solutions are needed on a continuous basis and that which can be shown to work in the long term can therefore take care of the short term as well. That is why a pro-nature-oriented scientific research effort could reasonably be expected to produce innovative breakthroughs not confined to managing pollution and waste but to conversion of so-called waste into useful new social products. The energy potential locked within many so-called waste materials, researched and developed on a nature-friendly basis, could produce a tremendous improvement in the overall standard of living by rendering energy supplies affordable and available to broad sections of the population that are currently underserved and overcharged.

When it comes to problems of international economic development, the development of energy sources cannot be discussed seriously or honestly without addressing the critical issues of groundwater supply. This is the most fundamental metric of true living standards in the still overwhelmingly rural societies of Asia, Africa, and Latin America. The entire bias in favor of engineered solutions as "modern" and "sophisticated" has served to accelerate the departure from the pro-nature path. Before the arrival, starting in the seventeenth century (1600s), of colonizing powers from Europe and the United States in these regions, this path was the norm of many so-called traditional societies; it was the same story among the native peoples of North and Central America, who suffered an outright genocide that began a century earlier.

A powerful movement has spread far and wide in recent years in many of these countries to restore traditional technologies, renewed appropriately and sufficiently to deal adequately with the speed of information and development in the modern world but otherwise traditional in their fundamental approach, which was and had to be pro-nature by definition. This accounts for the success of such initiatives as the "Water University," which has developed in several drought-stricken regions of the Asian subcontinent to recapture and reapply ancient traditional knowledge—dismissed for the last century and more as "old wives' tales," "folklore," "anecdotes but not science" etc.—of how to use the ambient natural conditions to conserve precious supplies of groundwater. The rapidity and scale of this success represents a most telling indictment as well of the signal failure of allegedly "advanced technological solutions" to lift more than a tiny elite from these societies into the ranks of the "civilized" colonizers.

But there is something yet more fundamental here than the confrontation between traditional societies and European colonization: the dependence of modern societal infrastructures on energy supplies is matched only by, and ultimately even underpinned by, the broader overall dependence of human social existence on the supply of water. The intimate connections between groundwater supplies and oil reservoirs were well understood in ancient times throughout Mesopotamia. One of the gravest indictments of the arrival of "advanced modern technology" to exploit the oil wealth of the region was the serious messing up of groundwater supplies that followed. The significance was greatly multiplied by the circumstance that these regions were literally at the edge of the westward-moving front of the central Asia desert. In the American southwest in particular, as activity in the civil courts serve to indicate, commercial natural gas development has enormously disrupted groundwater supplies and aquifers on the traditional lands of many native American tribes.

The corporate mantra of the true believers in "return on investment" as the Final Arbiter of the Possible seems to be: "if oil and water don't mix, forget the water!" Today, this stands increasingly exposed as a path imperiling the very future of humanity. The fact that one cannot sacrifice either the one or the other in the short term or the long term is increasingly well understood among all peoples everywhere. At the same time, nature ultimately cannot accept the confines of any system of so-called private property. It has its own logic that proceeds utterly oblivious to such inventions and conventions. That is why those following the corporate road must inevitably at some point take the antinature path—the path of "fixing" nature rather than learning its secrets and working with it, a path with a doubtful future even in the short term, and no future whatsoever in the long term.

Earlier in this chapter, this question was posed: considered from an engineering standpoint, are the patterns, changes, and trends in the supply of and demand for so-called nonrenewable energy commodities consistent with their actual potential? Our best answer at the moment is that the role of "return on investment"—as the driver matching the supply of energy commodities to an asserted demand that, more often than not, has been manipulated by hidden hands—seems to have blocked humanity's access to the pro-nature path. This is the only path on which the actual potential for these commodities can be properly researched and the problems involved in releasing their potential solved in ways that will be innovative, economically attractive, environmentally appealing, and socially responsible.

In general, natural gas has a key role to play as an energy source in the 21st century and the demand, which will be driven by the development of new markets for natural gas, is steadily growing. The share of natural gas reserves located onshore, easy to produce and close to consumers is decreasing and the share of natural gas reserves located offshore in difficult or distant areas is growing. Hence, the major present challenge of the gas industry is to be able to bring gas to the consumers at an economic cost. Substantial cost reductions achieved through technical progress have resulted in longer distance transports. LNG developments have also contributed to this evolution. However, in this case, gas transportation involves heavy and expensive infrastructures, which result in rather rigid ties between the producer and the consumer.

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CHAPTER 3

Raw Gas Transmission

3.1 INTRODUCTION

Natural gas is often found in places where there is no local market, such as in the many offshore fields around the world. For natural gas to be available to the market, it must be gathered, processed, and transported. Quite often, collected natural gas (raw gas) must be transported over a substantial distance in pipelines of different sizes, due to drive for reduced field processing facilities particularly for offshore fields. These pipelines vary in length between hundreds of feet to hundreds of miles, across undulating terrain with varying temperature conditions. Liquid condensation in pipelines commonly occurs because of the multicomponent nature of the transmitted natural gas and its associated phase behavior to the inevitable temperature and pressure changes that occur along the pipeline. Condensation subjects the raw gas transmission pipeline to multiphase, gas-condensate-water, flow transport.

Multiphase transportation technology has become increasingly important for developing marginal fields, where the trend is to economically transport unprocessed well fluids via existing infrastructures, maximizing the rate of return and minimizing both capital expenditure and operational expenditure (Klemp, 1999). In fact, when multiphase well fluid is transported in a single pipeline, separate pipelines and receiving facilities for separate phases, costing both money and space, are eliminated, which reduces capital expenditure. On the other hand, phase separation and reinjection of water and gas save both capital expenditure and operating expenditure by reducing the size of the fluid transport/handling facilities and the maintenance required for the pipeline operation (Hill, 1997). Given the savings that can be available to the operators using multiphase technology, the market for multiphase flow transportation is an expanding one. Hence, it is necessary to predict multiphase flow behavior and other design variables of gas/condensate pipelines as accurately as possible so that pipelines and downstream processing plants may be designed optimally. This chapter covers all the important concepts of multiphase gas/condensate transmission from a fundamental perspective.

3.2 MULTIPHASE FLOW TERMINOLOGY

The following sections define the variables commonly used to describe multiphase flow. Definitions for these variables are described based on Figure 3-1, which is the ideal flow of three fluids. It is assumed that the water is heavier than the oil and flows at the bottom, while the oil flows in the middle and the gas is the top layer.

3.2.1 Superficial Velocity

The superficial velocity is the velocity of one phase of a multiphase flow, assuming that the phase occupies the whole cross-section of pipe by itself. It is defined for each phase as follows:

$$V_{\mathrm{SW}} = \frac{Q_{\mathrm{W}}}{A} \tag{3-1}$$

$$V_{\mathrm{SO}} = \frac{Q_{\mathrm{O}}}{A} \tag{3-2}$$

$$V_{\mathrm{SG}} = \frac{Q_{\mathrm{G}}}{A} \tag{3-3}$$

where

$$A = A_{\mathrm{W}} + A_{\mathrm{O}} + A_{\mathrm{G}} \tag{3-4}$$

The parameter A is the total cross-sectional area of pipe, Q is volumetric flow rate, V is velocity, and the subscripts are W for water, O for oil, G for gas, and S for superficial term.

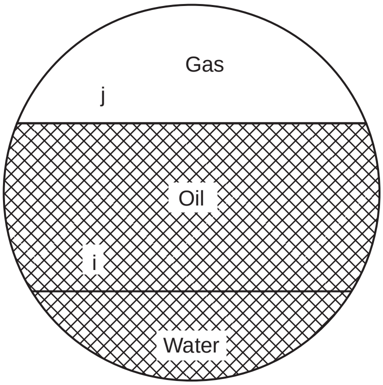


Figure 3-1 Three-phase flow pipe cross-section.

3.2.2 Mixture Velocity

The fluid mixture velocity is defined as the sum of the superficial velocities:

$$V_{\mathrm{M}} = V_{\mathrm{SW}} + V_{\mathrm{SO}} + V_{\mathrm{SG}} \tag{3-5}$$

where V_{M} is the multiphase mixture velocity.

3.2.3 Holdup

Holdup is the cross-sectional area, which is locally occupied by one of the phases of a multiphase flow, relative to the cross-sectional area of the pipe at the same local position.

For the liquid phase,

$$H_{\mathrm{L}} = \frac{A_{\mathrm{L}}}{A} = \frac{A_{\mathrm{W}} + A_{\mathrm{O}}}{A} = H_{\mathrm{W}} + H_{\mathrm{O}} \tag{3-6}$$

For the gas phase,

$$H_{\mathrm{G}} = \frac{A_{\mathrm{G}}}{A} \tag{3-7}$$

where the parameter H is the phase holdup, and the subscripts are L for the liquid and G for the gas phase.

Although "holdup" can be defined as the fraction of the pipe volume occupied by a given phase, holdup is usually defined as the *in situ* liquid volume fraction, while the term "void fraction" is used for the *in situ* gas volume fraction (Hasan and Kabir, 2002).

3.2.4 Phase Velocity

Phase velocity (*in situ* velocity) is the velocity of a phase of a multiphase flow based on the area of the pipe occupied by that phase. It may also be defined for each phase as follows:

$$V_{\mathrm{L}} = \frac{V_{\mathrm{SL}}}{H_{\mathrm{L}}} = \frac{V_{\mathrm{SW}} + V_{\mathrm{SO}}}{H_{\mathrm{L}}} \tag{3-8}$$

$$V_{\mathrm{G}} = \frac{V_{\mathrm{SG}}}{H_{\mathrm{G}}} \tag{3-9}$$

3.2.5 Slip

Slip is the term used to describe the flow condition that exists when the phases have different phase velocities. In most two-phase flow pipelines, the gas travels faster than the liquid. Under this condition, there is said to be slippage between the phases.

The slip velocity is defined as the difference between the actual gas and liquid velocities, as follows:

$$V_S = V_G - V_L \tag{3-10}$$

If there is no slip between the phases, $V_L = V_G$, and by applying the no-slip assumption to the liquid holdup definition, it can be shown that

$$H_{L,\text{no-slip}} = \lambda_L = \frac{V_{SL}}{V_M} \tag{3-11}$$

Investigators have observed that the no-slip assumption is not often applicable. For certain flow patterns in horizontal and upward inclined pipes, gas tends to flow faster than the liquid (positive slip). For some flow regimes in downward flow, liquid can flow faster than the gas (negative slip).

3.2.6 Mixture Density

The equations for two-phase gas/liquid density used by various investigators are

$$\rho_S = \rho_L H_L + \rho_G H_G \tag{3-12}$$

$$\rho_{nS} = \rho_L \lambda_L + \rho_G \lambda_G \tag{3-13}$$

where the subscripts S and nS represent the slip and the no-slip conditions, respectively.

In these equations, the total liquid density can be determined from the oil and water densities and flow rates if no slippage between these liquid phases is assumed:

$$\rho_L = \rho_O f_O + \rho_W f_W \tag{3-14}$$

where

$$f_O = \frac{Q_O}{Q_O + Q_W} = 1 - f_W \tag{3-15}$$

where the parameter f is the volume fraction of each phase.

3.2.7 Mixture Viscosity

To determine the mixture viscosity, three approaches have been proposed (Brill and Beggs, 1991):

$$\mu_S = \mu_L H_L + \mu_G H_G \tag{3-16}$$

$$\mu_S = \mu_L^{H_L} \cdot \mu_G^{H_G} \tag{3-17}$$

$$\mu_{nS} = \mu_L \lambda_L + \mu_G \lambda_G \tag{3-18}$$

where μ_L and μ_G are the liquid and gas viscosity, respectively.

The liquid viscosity can be the viscosity of water, oil, or water-oil mixture. Normally, the water-oil mixture viscosity can be calculated as

$$\mu_L = \mu_O f_O + \mu_W f_W \tag{3-19}$$

Studies have shown that Equation 3-19 often is not valid for the viscosity of two immiscible liquids, such as oil and water. For some oil/water systems, the emulsion viscosity can be many times higher than the individual phase viscosities. Peak viscosities (over 30 times the oil phase value) typically occur near the inversion point where the emulsion reverts from a water-in-oil dispersion to an oil-in-water dispersion. The inversion point usually prevails at water cuts in the range 20% to 50% (Arirachakaran, 1983).

Over the years, quite extensive research has been conducted to develop simplified correlations for the water-oil emulsion viscosity (Becher, 2001). But since there are so many parameters that affect the emulsion viscosity (Benayoune et al., 1998), none of these correlations can be universally applied to engineering calculations. Instead, the best way to determine water-oil emulsion viscosity is to perform lab measurements of emulsions of different water cut at elevated pressure and temperature conditions.

3.2.8 Mixture Pressure Drop

The general pressure drop equation for multiphase (two- and three-phase) flow is similar to that for single-phase flow except some of the variables are replaced with equivalent variables that consider the effect of multiphase. The general pressure drop equation for multiphase flow is as follows (Brill and Beggs, 1991):

$$\left(\frac{dP}{dx}\right)_{\text{tot}} = \left(\frac{dP}{dx}\right)_{\text{ele}} + \left(\frac{dP}{dx}\right)_{\text{fri}} + \left(\frac{dP}{dx}\right)_{\text{acc}} \tag{3-20}$$

where

$$\left(\frac{dP}{dx}\right)_{\mathrm{ele}} = \rho_{\mathrm{tp}}\left(\frac{g}{g_{\mathrm{c}}}\right)\sin\theta \tag{3-20-1}$$

$$\left(\frac{dP}{dx}\right)_{\mathrm{fri}} = \frac{\rho_{\mathrm{tp}} f_{\mathrm{tp}} V_{\mathrm{tp}}^2}{2g_{\mathrm{c}}D} \tag{3-20-2}$$

$$\left(\frac{dP}{dx}\right)_{\mathrm{acc}} = \frac{\rho_{\mathrm{tp}} f_{\mathrm{tp}}}{g_{\mathrm{c}}}\left(\frac{dV_{\mathrm{tp}}}{dx}\right) \tag{3-20-3}$$

where dP/dx is flow pressure gradient, x is pipe length, ρ is flow density, V is flow velocity, f is friction coefficient of flow, D is the internal diameter of the pipeline, θ is the inclination angle of the pipeline, g is gravitational acceleration, and g_c is the gravitational constant. The subscripts are tot for total, ele for elevation, fri for friction loss, acc for acceleration change terms, and tp for two- and/or three-phase flow.

The pressure-drop component caused by acceleration is normally negligible and is considered only for cases of high flow velocities.

Many methods have been developed to predict multiphase, flowing-pressure gradients. They differ in the manner used to calculate the three components of the total pressure gradient. Section 3-4 describes these methods.

3.2.9 Mixture Enthalpy

When temperature-change calculations are performed for multiphase flow in pipelines, it is necessary to predict the enthalpy of the multiphase mixture. If enthalpies of the gas and liquid phases are expressed per unit mass, the enthalpy of a multiphase mixture can be calculated from

$$h_{\mathrm{M}} = H_{\mathrm{L}}h_{\mathrm{L}} + (1 - H_{\mathrm{L}})\mathrm{h}_{\mathrm{G}} \tag{3-21}$$

3.3 MULTIPHASE FLOW REGIMES

Multiphase flow is characterized by the existence of interfaces between the phases and discontinuities of associated properties. The flow structures are rather classified in “flow regimes” or “flow patterns,” whose precise characteristics depend on a number of parameters. Flow regimes vary depending on operating conditions, fluid properties, flow rates, and the orientation and geometry of the pipe through which the fluids flow. The transition between different flow regimes may be a gradual process. Due to the highly nonlinear nature of the forces that rule the flow regime transitions, the prediction is nearly impossible. In the laboratory, the flow regime may be studied by direct visual observation using a length of transparent piping (Wallis, 1969). However, the most utilized approach is to identify the actual flow regime from signal analysis of sensors whose fluctuations are related to the flow regime structure. This approach is generally based on average cross-sectional quantities, such as pressure drop or cross-sectional liquid holdup. Many studies have been documented using different sensors and different analysis techniques (Dukler and Hubbard, 1966; Jones and Zuber, 1975; Lin and Hanratty, 1987; Rajkovic et al., 1996; Soldati et al., 1996).

3.3.1 Two-Phase Flow Regimes

The description of two-phase flow can be simplified by classifying types of "flow regimes" or "flow patterns." The distribution of the fluid phases in space and time differs for the various flow regimes and is usually not under the control of the pipeline designer or operator (Amdal et al., 2001).

Dukler and Hubbard (1966) suggested three basic flow patterns: separated, intermittent, and distributed flow. In separated flow patterns, both phases are continuous, and some droplets or bubbles of one phase in the other may or may not exist. In the intermittent flow patterns, at least one phase is discontinuous. In dispersed flow patterns, the liquid phase is continuous, while the gas phase is discontinuous.

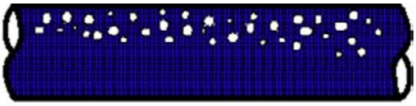
Due to a multitude of flow patterns and the various interpretations accorded to them by different investigators, the general state of knowledge on flow patterns is unsatisfactory, and no uniform procedure exists at present for describing and classifying them. In the following sections, the basic flow patterns in gas-liquid flow in horizontal, vertical, and inclined pipes are introduced.

3.3.1.1 Horizontal Flow Regimes

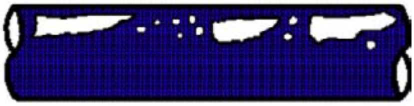
Two-phase, gas-liquid flow regimes for horizontal flow are shown in Figure 3-2. These horizontal flow regimes are defined as follows.

Dispersed Bubble Flow: At high liquid flow rates and for a wide range of gas flow rates, small gas bubbles are dispersed throughout a continuous liquid phase. Due to the effect of buoyancy, these bubbles tend to accumulate in the upper part of the pipe.

Plug (Elongated Bubble) Flow: At relatively low gas flow rates, as the liquid flow rate is reduced, the smaller bubbles of dispersed bubble flow coalesce to form larger bullet-shaped bubbles that move along the top of the pipe.

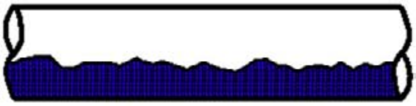


Bubble Flow

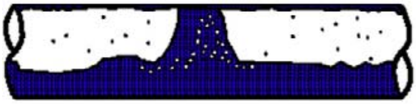


Plug Flow

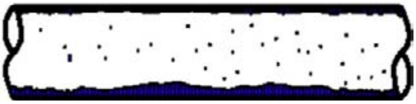
Stratified Smooth



Stratified Wavy



Slug Flow



Annular Flow

Figure 3-2 Horizontal two-phase, gas-liquid flow regimes.

Stratified (Smooth and Wavy) Flow: At low liquid and gas flow rates, gravitational effects cause total separation of the two phases. This results in the liquid flowing along the bottom of the pipe and the gas flowing along the top, where the gas-liquid surface is smooth. As the gas velocity is increased in stratified smooth flow, the interfacial shear forces increase, rippling the liquid surface and producing a wavy interface.

Slug Flow: As the gas and liquid flow rates are increased further, the stratified liquid level grows and becomes progressively more wavy until eventually the whole cross-section of the pipe is blocked by a wave. The resultant "piston" of liquid is then accelerated by the gas flow, surging along the pipe and scooping up the liquid film in front as it progresses. This "piston" is followed by a region containing an elongated gas bubble moving over a thin liquid film. Hence, an intermittent regime develops in which elongated gas bubbles and liquid slugs alternately surge along the pipe. The major difference between elongated bubble flow and slug flow is that in elongated bubble flow there are no entrained gas bubbles in the liquid slugs.

Annular Flow: When gas flow rates increase, annular (also referred to as annular-mist) flow occurs. During annular flow, the liquid phase flows largely as an annular film on the wall with gas flowing as a central core. Some of the liquid is entrained as droplets in this gas core. The annular liquid film is thicker at the bottom than at the top of the pipe because of the effect of gravity and, except at very low liquid rates, the liquid film is covered with large waves.

3.3.1.2 Vertical Flow Regimes

Flow regimes frequently encountered in upward vertical two-phase flow are shown in Figure 3-3. These flow regimes tend to be somewhat simpler than those in horizontal flow. This results from the symmetry in the flow induced by the gravitational force acting parallel to it. A brief description of the manner in which the fluids are distributed in the pipe for upward vertical

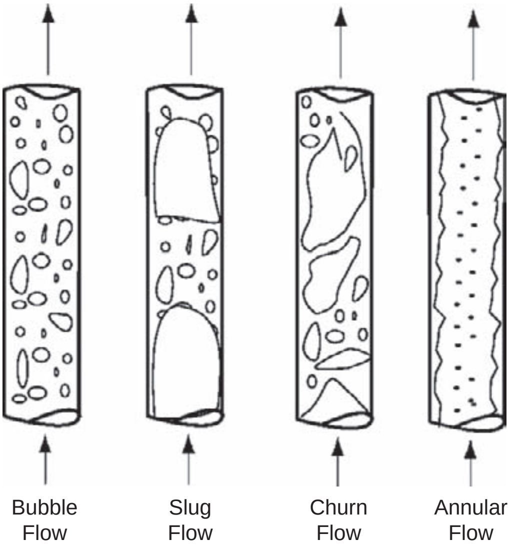


Figure 3-3 Upward vertical two-phase flow regimes.

two-phase flow is as follows. It is worth noting that vertical flows are not so common in raw gas systems (i.e., wells normally have some deviation, and many risers are also inclined to some extent).

Bubble Flow: At very low liquid and gas velocities, the liquid phase is continuous, and the gas phase travels as dispersed bubbles. This flow regime is called bubble flow. As the liquid flow rate increases, the bubbles may increase in size via coalescence. Based on the presence or absence of slippage between the two phases, bubble flow is further classified into bubbly and dispersed bubble flows. In bubbly flow, relatively fewer and larger bubbles move faster than the liquid phase because of slippage. In dispersed bubble flow, numerous tiny bubbles are transported by the liquid phase, causing no relative motion between the two phases.

Slug Flow: As the gas velocity increases, the gas bubbles start coalescing, eventually forming large enough bubbles (Taylor bubbles) that occupy almost the entire cross sectional area. This flow regime is called slug flow. Taylor bubbles move uniformly upward and are separated by slugs of continuous liquid that bridge the pipe and contain small gas bubbles. Typically, the liquid in the film around the Taylor bubbles may move downward at low velocities although the net flow of liquid can be upward. The gas bubble velocity is greater than that of the liquid.

Churn (Transition) Flow: If a change from a continuous liquid phase to a continuous gas phase occurs, the continuity of the liquid in the slug between successive Taylor bubbles is repeatedly destroyed by a high local gas concentration in the slug. This oscillatory flow of the liquid is typical of churn (froth) flow. It may not occur in small diameter pipes. The gas bubbles may join, and liquid may be entrained in the bubbles. In this flow regime, the falling film of the liquid surrounding the gas plugs cannot be observed.

Annular Flow: As the gas velocity increases even further, the transition occurs, and the gas phase becomes a continuous phase in the pipe core. The liquid phase moves upward partly as a thin film (adhering to the pipe wall) and partly in the form of dispersed droplets in the gas core. This flow regime is called an annular flow or an annular mist flow.

Although downward vertical two-phase flow is less common than upward flow, it does occur in steam injection wells and down-comer pipes from offshore production platforms. Hence, a general vertical two-phase flow pattern is required that can be applied to all flow situations (Hasan, 1995). Reliable models for downward multiphase flow are currently unavailable, and the design codes are deficient in this area.

3.3.1.3 Inclined Flow Regimes

The effect of pipeline inclination on the gas-liquid two-phase flow regimes is of a major interest in hilly terrain pipelines that consist almost entirely of uphill- and downhill-inclined sections. Pipe inclination angles have a very strong influence on flow pattern transitions. Generally, the flow regime in a near horizontal pipe remains segregated for downward inclinations and changes to intermittent flow regime for upward inclinations. An intermittent flow regime remains intermittent when tilted upward and tends to segregate flow pattern when inclined downward. The inclination should not significantly affect the distributed flow regime (Scott et al., 1987).

3.3.1.4 Flow Pattern Maps

In order to obtain optimal design parameters and operating conditions, it is necessary to clearly understand multiphase flow regimes and the boundaries between them, where the hydrodynamics of the flow as well as the flow mechanisms change significantly from one flow regime to another (Cheremisinoff, 1986). If an undesirable flow regime is not anticipated in the design, the resulting flow pattern can cause system pressure fluctuation and system vibration and even mechanical failures of piping components.

Most early attempts to predict the occurrence of the various flow patterns in pipes were based on conducting experimental tests in small-diameter pipes at low pressures with air and water. The results of experimental studies were presented as a flow pattern map. The respective patterns were represented as areas on a plot, the coordinates of which were the dimensional variables (i.e., superficial phase velocities) or dimensionless parameters containing these velocities (Collier and Thome, 1996). For horizontal flows, the classical flow pattern map is that of Mandhane et al. (1974), as shown in Figure 3-4. This particular map is based on air-water data at atmospheric pressure in 0.5- to 6.5-inch pipes.

Figure 3-5 shows the flow pattern map developed by Aziz et al. (1972) for vertical upward flow. Equations 3-22 and 3-23 define the coordinates:

$$N_x = V_{SG}\left(\frac{\rho_G}{0.0764}\right)^{1/3}\left[\left(\frac{72}{\sigma_L}\right)\left(\frac{\rho_L}{62.4}\right)\right]^{1/4} \tag{3-22}$$

$$N_y = V_{SL}\left[\left(\frac{72}{\sigma_L}\right)\left(\frac{\rho_L}{62.4}\right)\right]^{1/4} \tag{3-23}$$

The bracketed terms in the preceding equations are attempts to validate the flow pattern map for fluids other than air and water.

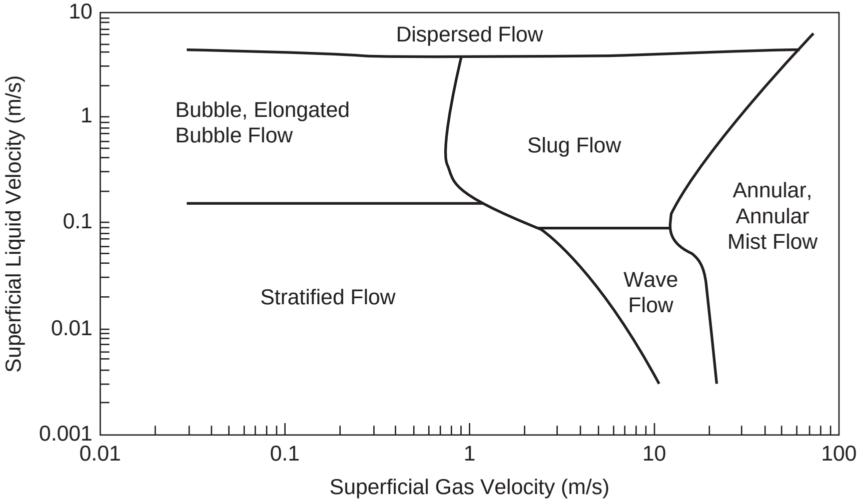


Figure 3-4 Mandhane et al. (1974) flow regime map for horizontal flow.

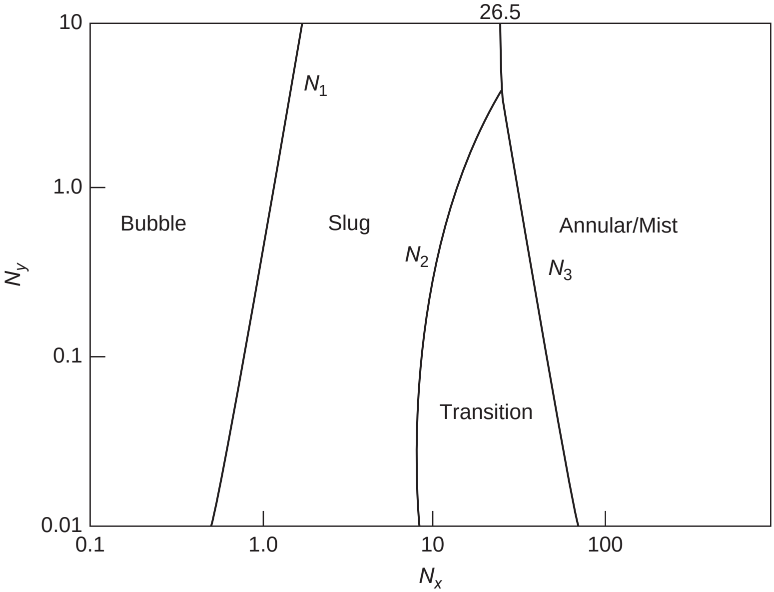


Figure 3-5 Aziz et al. (1972) flow pattern map for vertical upward flow.

Equations 3-24 to 3-26 represent the flow pattern transitions in Figure 3-5:

$$N_1 = 0.51\left(100N_y\right)^{0.172} \tag{3-24}$$

$$N_2 = 8.6 + 3.8N_y \tag{3-25}$$

$$N_3 = 70\left(100N_y\right)^{-0.152} \tag{3-26}$$

where superficial velocities are in feet per second, liquid and gas densities are in pounds per cubic feet, and liquid surface tension[1] (σ_L) is in dynes per centimeter.

Empirically based flow maps are not particularly accurate for systems where the fluid properties, pipe size, and inclination are different from those for which the flow map was originally produced. In addition, in these flow maps a change in flow rate can lead to a change of flow pattern between the old and new steady states. However, an interesting and important effect that frequently occurs is the existence of temporary flow pattern changes between the respective steady states (Shaha, 1999). A more flexible method, which overcomes this difficulty, is to examine each transition individually and derive a criterion valid for that particular transition. There have been some attempts to evaluate the basic mechanisms of flow pattern transitions and thus to provide a mechanistic flow pattern map for estimating their occurrence (Taitel and Dukler, 1976; Barnea, 1987; Taitel et al., 1980; Petalas and Aziz, 2000). In these transition models, the effects of system parameters are incorporated; hence, they can be applied over a range of conditions. However, most of them are somewhat complex, and they do require the use of a predetermined sequence to determine the dominant flow pattern (Hasan and Kabir, 2002).

3.3.2 Three-Phase Flow Regimes

The main difference between two-phase (gas/liquid) flows and three-phase (gas/liquid/liquid) flows is the behavior of the liquid phases. In three-phase systems, the presence of two liquids gives rise to a rich variety of flow patterns (Hall, 1997). Basically, depending on the local conditions, the liquid phases appear in a separated or dispersed form (Brauner and Maron, 1992; Brauner, 2001). In the case of separated flow, distinct layers of oil and water can be discerned, though there may be some interentrainment of one liquid phase into the other. In dispersed flow, one liquid phase is

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[1] Assuming no slippage, the liquid surface tension is calculated from $\sigma_L = \sigma_O f_O + \sigma_W f_W$.

completely dispersed as droplets in the other, resulting in two possible situations: namely, an oil continuous phase and a water continuous phase (Chen and Guo, 1999). The transition from one liquid continuous phase to the other is known as phase inversion. If the liquid phases are interdispersed, then prediction of phase inversion is an important item. For this purpose, Decarre and Fabre (1997) developed a phase inversion model that can be used to determine which of two liquids is continuous.

Due to the many possible transport properties of three-phase fluid mixtures, the quantification of three-phase flow pattern boundaries is a difficult and challenging task. Acikgoz et al. (1992) observed a very complex array of flow patterns and described 10 different flow regimes. In their work, the pipe diameter was only 0.748 inch and stratification was seldom achieved. In contrast, the Lee et al. (1993) experiments were carried out in a 4-inch diameter pipe. They observed and classified seven flow patterns, which were similar to those of two-phase flows: (1) smooth stratified, (2) wavy stratified, (3) rolling wave, (4) plug flow, (5) slug flow, (6) pseudo slug, and (7) annular flow. The first three flow patterns can be classified as stratified flow regime and Lee et al. noted that the oil and the water are generally segregated, with water flowing as a liquid layer at the bottom of the pipe and oil flowing on top. Even for plug flow, the water remained at the bottom, because the agitation of the liquids was not sufficient to mix the oil and water phases. Note that turbulence, which naturally exists in a pipeline, can be sufficient to provide adequate mixing of the water and oil phase. However, the minimum natural turbulent energy for adequate mixing depends on the oil and water flow rates, pipe diameter and inclination, water concentration, viscosity, density, and interfacial tension. Dahl et al. (2001) provided more detailed information on the prediction methods that can be used to determine whether a water-in-oil mixture in the pipe is homogeneous or not.

3.3.3 Gas/Condensate Flow Regimes

Gas/condensate flow is a multiphase flow phenomenon commonly encountered in raw gas transportation. However, the multiphase flow that takes place in gas/condensate transmission lines differs in certain respects from the general multiphase flow in pipelines. In fact, in gas/condensate flow systems, there is always interphase mass transfer from the gas phase to the liquid phase because of the temperature and pressure variations. This leads to compositional changes and associated fluid property changes (Ayala and

Adewumi, 2003). In addition to that, the amount of liquid in such systems is assumed to be small, and the gas flow rate gives a sufficiently high Reynolds number that the fluid flow regime for a nearly horizontal pipe can be expected to be annular-mist flow and/or stratified flow (Boriyantoro and Adewumi, 1994; Asante, 2002). For other inclined cases, even with small quantities of liquids, slug-type regimes may be developed if liquids start accumulating at the pipe's lower section.

In water wet gas/condensate systems, it can normally be assumed that the condensate and water phases are well mixed. However, when the water loading is a substantial proportion of the total liquid loading, the water and condensate can separate out at low flow rates, leading to the accumulation of stagnant water at low points in the line. The presence of stagnant water can lead to the production of water slugs during normal line operation or flow-rate changes.

3.4 DETERMINING MULTIPHASE FLOW DESIGN PARAMETERS

The hydraulic design of a multiphase flow pipeline is a two-step process. The first step is the determination of the multiphase flow regimes, because many pressure drop calculation methods rely on the type of flow regime present in the pipe. The second step is the calculation of flow parameters, such as pressure drop and liquid holdup, to size pipelines and field processing equipment such as slug catchers.

The analysis of multiphase flow phenomena in pipeline systems is usually classified into steady-state and transient approaches. In steady-state flow, there are no major changes transgressing the pipeline network. In transient or dynamic flows, the flow behavior is changing on a regular and significant basis. Both steady-state and transient analyses shall be performed to study the pipeline system performance under different operating conditions. In preliminary engineering, a design engineer often uses steady-state analysis. However, the dynamic response of the system to start-up, shutdown, and flow-rate changes must also be considered. For example, liquid slugs are often produced during flow-rate increase or line depressurization. The volume of these transient slugs must be determined to ensure that they can be handled by the downstream separation and processing facilities. If there is insufficient capacity, operating procedures can be introduced to limit the rate of change of flow rate and therefore the size of any produced slugs.

3.4.1 Steady-State Two-Phase Flow

The techniques most commonly used in the design of two-phase flow pipeline can be classified into three categories: the single-phase flow approaches, the homogeneous flow approaches, and the mechanistic models. Within each of these groups are subcategories that are based on the general characteristics of the models used to perform design calculations.

3.4.1.1 Single-Phase Flow Approaches

In this method, the two-phase flow is assumed to be a single-phase flow having pseudo-properties arrived at by suitably weighting the properties of the individual phases. These approaches basically rely solely on the well-established design equations for single-phase gas flow in pipes. Two-phase flow is treated as a simple extension by use of a multiplier—a safety factor to account for the higher-pressure drop generally encountered in two-phase flow. This heuristic approach was widely used and generally resulted in inaccurate pipeline design (Dukler, 1969; Oranje, 1973; Schweikert, 1986).

In the past, single-phase flow approaches were commonly used for the design of wet gas pipelines. When the amount of the condensed liquid is negligibly small, the use of such methods could at best prevent underdesign, but more often than not, the quantity of the condensed liquid is significant enough that the single-phase flow approach grossly overpredicted the pressure drop (Ullah, 1987). Hence, this method is not covered in depth here but can be reviewed elsewhere (Uhl, 1965).

3.4.1.2 Homogeneous Flow Approaches

The inadequacy of the single-phase flow approaches spurred researchers to develop better design procedures and predictive models for two-phase flow systems. This effort led to the development of homogeneous flow approaches to describe these rather complex flows. The homogeneous approach, also known as the friction factor model, is similar to that of the single-phase flow approach except that mixture fluid properties are used in the determination of the friction factor. Therefore, the appropriate definitions of the fluid properties are critical to the accuracy of the model. The mixture properties are expressed empirically as a function of the gas and liquid properties as well as their respective holdups. Many of these correlations are based on flow regime correlations that determine the two-phase (gas/liquid) flow friction factor, which is then used to estimate pressure drop. While some of the correlations predict pressure-drop reasonably well, their range of applicability is generally limited, making their use as a scale-up tool marginal.

This limitation is understandable because the database used in developing these correlations is usually limited and based on laboratory-scale experiments. Extrapolation of these data sets to larger lines and hydrocarbon systems is questionable at best (Cindric et al., 1987).

Historically, empirical correlations have been successful in terms of being relatively easy to employ as design tools. Brill and Beggs (1991) and Collier and Thome (1996) provided in-depth comparative analyses of the available correlations. However, two of the better known, more widely applicable correlations are outlined here.

3.4.1.2.1 Lockhart and Martinelli Method

The Lockhart and Martinelli method was developed by correlating experimental data generated in horizontal isothermal two-phase flow of two-component systems (air-oil and air-water) at low pressures (close to atmospheric) in a 1-inch diameter pipe. Lockhart and Martinelli (1949) separated the data into four sets dependent on whether the phases flowed in laminar or turbulent flow, if the phases were flowing alone in the same pipe. In this method, a definite portion of the flow area is assigned to each phase, and it is assumed that the single-phase pressure drop equations can be used independently for each phase. The two-phase frictional pressure drop is calculated by multiplying by a correction factor for each phase, as follows:

$$\frac{dP}{dx} = \varphi_G^2 \left(\frac{dP}{dx}\right)_G = \varphi_L^2 \left(\frac{dP}{dx}\right)_L \tag{3-27}$$

where

$$\left(\frac{dP}{dx}\right)_G = \left(\frac{f_G \rho_G \ V_{SG}^2}{2\, g_C\ D}\right) \tag{3-27-1}$$

$$\left(\frac{dP}{dx}\right)_L = \left(\frac{f_L \rho_L \ V_{SL}^2}{2\, g_C\ D}\right) \tag{3-27-2}$$

The friction factors f_G and f_L are determined from the Moody (1944) diagram using the following values of the Reynolds number:

$$N_{Re,G} = \frac{\rho_G \ V_{SG}\ D}{\mu_G} \tag{3-28}$$

$$N_{Re,L} = \frac{\rho_L \ V_{SL}\ D}{\mu_L} \tag{3-29}$$

The two-phase flow correction factors (φ_G, φ_L) are determined from the relationships of Equations 3-30 and 3-31 (Lockhart and Martinelli, 1949; Chisholm and Sutherland, 1969):

$$\varphi_G^2 = 1 + CX + CX^2 \tag{3-30}$$

$$\varphi_L^2 = 1 + \frac{C}{X} + \frac{1}{X^2} \tag{3-31}$$

where

$$X = \left[\frac{\left(\dfrac{dP}{dx}\right)_L}{\left(\dfrac{dP}{dx}\right)_G}\right]^{0.5} \tag{3-32}$$

where parameter C is given in Table 3-1.

Note that the laminar flow regime for a phase occurs when the Reynolds number for that phase is less than 1,000.

In this method, the correlation between liquid holdup and Martinelli parameter, X, is independent of the flow regime and can be expressed as follows (Collier and Thome, 1996):

$$H_L^{-2} = 1 + \frac{20}{X} + \frac{1}{X^2} \tag{3-33}$$

The fluid acceleration pressure drop was ignored in this method. However, the extension of the work covering the estimation of the accelerative term was done by Martinelli and Nelson (1948). Although different modifications of this method have been proposed, the original method is believed to be generally the most reliable (Collier and Thome, 1996).

3.4.1.2.2 Beggs and Brill Method

The Beggs and Brill method was developed from 584 experimental data sets generated on a laboratory scale test facility using an air-water system. The facility consisted of 90 feet of 1- or 1.5-inch diameter acrylic (smooth) pipe,

Table 3-1 C Parameter

Liquid Phase	Gas Phase	C
Turbulent	Turbulent	20
Laminar	Turbulent	12
Turbulent	Laminar	10
Laminar	Laminar	5

which could be inclined at any angle (Beggs and Brill, 1973). The pipe angle was varied between horizontal to vertical, and the liquid holdup and the pressure were measured. For each pipe size in the horizontal position, flow rates of two phases were varied to achieve all flow regimes. Beggs and Brill (1973) developed correlations for the liquid holdup for each of three horizontal flow regimes and then corrected these for the pipe inclination/angle.

The following parameters are used for determination of all horizontal flow regimes (Brill and Beggs, 1991):

$$N_{\mathrm{Fr}} = \frac{V_{\mathrm{M}}^2}{g\,D} \tag{3-34}$$

$$L_1 = 316\,\lambda_{\mathrm{L}}{}^{0.302} \tag{3-35}$$

$$L_2 = 0.0009252\lambda_{\mathrm{L}}{}^{-2.4684} \tag{3-36}$$

$$L_3 = 0.10\,\lambda_{\mathrm{L}}{}^{-1.4516} \tag{3-37}$$

$$L_4 = 0.5\lambda_{\mathrm{L}}{}^{-6.738} \tag{3-38}$$

And the flow regimes limits are as follows:

Segregated:

$$N_{\mathrm{Fr}} < L_1 \quad \text{for} \quad \lambda_{\mathrm{L}} < 0.01$$

$$N_{\mathrm{Fr}} < L_2 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.01$$

Transition:

$$L_2 \leq N_{\mathrm{Fr}} \leq L_3 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.01$$

Intermittent:

$$L_3 \leq N_{\mathrm{Fr}} \leq L_1 \quad \text{for} \quad 0.01 < \lambda_{\mathrm{L}} \leq 0.4$$

$$L_3 < N_{\mathrm{Fr}} \leq L_4 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.4$$

Distributed:

$$N_{\mathrm{Fr}} \geq L_1 \quad \text{for} \quad \lambda_{\mathrm{L}} < 0.4$$

$$N_{\mathrm{Fr}} > L_4 \quad \text{for} \quad \lambda_{\mathrm{L}} \geq 0.4$$

Also, the horizontal liquid holdup, $H_{\mathrm{L}}(0)$, is calculated using the following equation:

$$H_{\mathrm{L}}(0) = \frac{a\lambda_{\mathrm{L}}^{\mathrm{b}}}{N_{\mathrm{Fr}}^{\mathrm{c}}} \tag{3-39}$$

Table 3-2 a, b, and c Parameters

Flow Regime	**a**	**b**	**c**
Segregated	0.98	0.4846	0.0868
Intermittent	0.845	0.5351	0.0173
Distributed	1.065	0.5824	0.0609

where parameters a, b, and c are determined for each flow regime and are given in Table 3-2, with the constraint that $H_L(0) \geq \lambda_L$.

When the flow is in the transition region, the liquid holdup is calculated by interpolating between the segregated and intermittent flow regimes as follows:

$$H_L(0)_{\text{transition}} = A\,H_L(0)_{\text{segregated}} + (1-A)\,H_L(0)_{\text{intermittent}} \tag{3-40}$$

where

$$A = (L_3 - N_{Fr})/(L_3 - L_2) \tag{3-41}$$

The amount of liquid holdup in an inclined pipe, $H_L(\theta)$, is determined by multiplying an inclination factor (ψ) by the calculated liquid holdup for the horizontal conditions:

$$H_L(\theta) = H_L(0) \times \psi \tag{3-42}$$

where

$$\psi = 1 + \alpha[\text{Sin}\,(1.8\theta) - 0.333\,\text{Sin}^3(1.8\theta)] \tag{3-42-1}$$

where θ is the pipe angle and α is calculated as follows:

$$\alpha = (1-\lambda_L)\,\text{Ln}\,[d\,\lambda_L{}^e\,N_{LV}{}^f\,N_{Fr}{}^g] \tag{3-42-2}$$

where

$$N_{LV} = V_{SL}\left(\frac{\rho_L}{g\sigma}\right)^{0.25} \tag{3-42-3}$$

The equation parameters are determined from each flow regime using the numbers from Table 3-3.

Table 3-3 d, e, f, and g Parameters

Flow Regime	**d**	**e**	**f**	**g**
Segregated uphill	0.011	–3.768	3.539	–1.614
Intermittent uphill	2.96	0.305	–0.4473	0.0978
Distributed uphill	$\alpha = 0$, $\psi = 1$			
All flow regimes downhill	4.7	–0.3692	0.1244	–0.5056

The two-phase pressure gradient due to pipeline elevation can be determined as follows:

$$\left(\frac{dP}{dx}\right)_{\text{ele}} = \frac{g}{g_C}\{\rho_L\ H_L(\theta) + \rho_G[1 - H_L(\theta)]\} \tag{3-43}$$

Also, the two-phase frictional pressure drop is calculated as follows:

$$\left(\frac{dP}{dx}\right)_{\text{fri}} = \frac{f_{tp}\rho_n V_M^2}{2\, g_c\, D} \tag{3-44}$$

where

$$f_{tp} = f_n\ \text{Exp}(\beta) \tag{3-44-1}$$

where f_n is the no-slip friction factor determined from the smooth pipe curve of the Moody diagram using no-slip viscosity and the density in calculating the two-phase flow Reynolds number. In other words,

$$f_n = \frac{1}{\left[2\ \text{Log}\left(\dfrac{N_{Re,n}}{4.5223\ \text{Log}\ N_{Re,n} - 3.8215}\right)\right]^2} \tag{3-44-1-1}$$

where

$$N_{Re,n} = \frac{\rho_n\ V_M\ D}{\mu_n} = \frac{[(\rho_L\lambda_L + \rho_G(1 - \lambda_L))\ V_M\ D]}{[\mu_L\lambda_L + \mu_G(1 - \lambda_L)]} \tag{3-44-1-2}$$

The exponent β is given by Equation 3-44-2:

$$\beta = [\text{Ln}Y]/\left\{-0.0523 + 3.182\ \text{Ln}Y - 0.8725\ (\text{Ln}Y)^2 + 0.01853\ (\text{Ln}Y)^4\right\} \tag{3-44-2}$$

where

$$Y = \frac{\lambda_L}{[H_L(\theta)]^2} \tag{3-44-3}$$

The pressure drop due to acceleration is significant only in gas transmission pipelines at high gas flow rates. However, it can be included for completeness as follows:

$$\left(\frac{dP}{dx}\right)_{\text{acc}} = \left[\frac{\rho_S\ V_M\ V_{SG}}{g_c\ P}\right]\left(\frac{dP}{dx}\right)_{\text{tot}} \tag{3-45}$$

The Beggs and Brill (1973) method can be used for horizontal and vertical pipelines, although its wide acceptance is mainly due to its usefulness for inclined pipe pressure drop calculation (Brill and Beggs, 1991). However, the proposed approach has limited applications dictated by the database on which the correlations are derived.

Note that the Beggs and Brill (1973) correlation is not usually employed for the design of wet-gas pipelines. Specifically the holdup characteristic (holdup versus flow rate) is not well predicted by this method. This makes design difficult because one is unable to reliably quantify the retention of liquid in the line during turndown conditions.

3.4.1.3 Mechanistic Models

While it is indeed remarkable that some of the present correlations can adequately handle noncondensing two-phase flow, they give erroneous results for gas/condensate flow in large diameter lines at high pressures (Osman and El-Feky, 1985; Battara et al., 1985). This shortcoming of existing methods has led to the development of mechanistic models that are based on fundamental laws and thus can offer more accurate modeling of the pipe geometric and fluid property variations. All of these models firstly predict stable flow pattern under the specified conditions and then use the momentum balance equations to calculate liquid holdup, pressure drop, and other two-phase flow parameters with a greater degree of confidence than that possible by purely empirical correlations (Hasan and Kabir, 1986, 1992; Collier and Thome, 1996; Hasan and Kabir, 1999; Holt et al., 1999). The mechanistic models presented in the literature are either incomplete in that they only consider flow pattern determination or are limited in their applicability to only some pipe inclinations or small-diameter, low-pressure two-phase flow lines (Taitel and Dukler, 1976; Barnea, 1987; Xiao et al., 1990; Ansari et al., 1994; Taitel et al., 1995; Wilkens, 1997). However, new mechanistic models presented by Petalas and Aziz (2000) and Zhang et al. (2003) have proven to be more robust than previous models, although further investigations and testing of these models are needed with high-quality field and laboratory data in larger diameter, high-pressure systems. Readers are referred to the original references for a detailed treatment of these models.

3.4.2 Steady-State Three-Phase Flow

Compared to numerous investigations of two-phase flow in the literature, there are only limited works on three-phase flow of gas-liquid-liquid mixtures. In fact, the complex nature of such flows makes prediction very difficult. In an early study, Tek (1961) treated the two immiscible liquid phases as a single fluid with mixture properties; thus, a two-phase flow correlation could be used for pressure loss calculations. On the other hand, more recent studies by Pan (1996) showed that the classical two-phase flow correlations for gas-liquid two-phase flow can be used as the basis in the determination of three-phase flow parameters; however, the generality of such empirical approaches is obviously questionable. Hence, an appropriate model is required to describe the flow of one gas and two liquid phases. One of the most fundamental approaches used to model such systems is the two-fluid model, where the presented approach can be used by combining the two liquid phases as one pseudo-liquid phase and modeling the three-phase flow as a two-phase flow. However, for more accurate results, three-fluid models should be used to account for the effect of liquid-liquid interactions on flow characteristics, especially at low flow rates. Several three fluid models were found in the literature. All of these models are developed from the three-phase momentum equations with few changes from one model to the other. The most obvious model was developed by Barnea and Taitel (1996); however, such a model introduces much additional complexity and demands much more in computer resources compared with the two-fluid model for two-phase flow (Bonizzi and Issa, 2003).

3.4.3 Transient Multiphase Flow

Steady-state operation is the exception rather than the rule in multiphase systems due to a number of factors such as the nature of the flow regimes, interactions with the pipeline profile, and changes in supply or demand. As a result, detailed information of the transient flow behavior is necessary for the designer and the operator of the system to construct and operate the pipeline economically and safely.

Transient multiphase flow is traditionally modeled by one-dimensional averaged conservation laws, yielding a set of partial differential equations. In this section, two models of particular industrial interest are described:

- The two-fluid model (TFM), consisting of a separate momentum equation for each phase.
- The drift-flux model (DFM), consisting of a momentum equation and an algebraic slip relation for the phase velocities.

The two-fluid model is structurally simpler but involves an extra differential equation when compared to the drift-flux model. They do yield somewhat different transient results, though the differences are often small (Masella et al., 1998).

The following major assumptions have been made in the formulation of the differential equations:

1. Two immiscible liquid phases (oil and water) are assumed to be a single fluid with mixture properties.
2. Flow is one-dimensional in the axial direction of the pipeline.
3. Flow temperature is constant at wall, and no mass transfer occurs between the gas and liquid phase. Note that most commercial codes allow phase change.
4. The physical properties of multiphase flow are determined at the average temperature and pressure of flow in each segment of the pipeline.

3.4.3.1 Two-Fluid Model

The two-fluid model (TFM) is governed by a set of four partial differential equations, the first two of which express mass conservation for gas and liquid phases, respectively:

$$\frac{\partial}{\partial \mathrm{t}}[\rho_G H_G] + \frac{\partial}{\partial \mathrm{x}}[\rho_G H_G V_G] = 0 \tag{3-46}$$

$$\frac{\partial}{\partial \mathrm{t}}[\rho_L H_L] + \frac{\partial}{\partial \mathrm{x}}[\rho_L H_L V_L] = 0 \tag{3-47}$$

The last two equations represent momentum balance for the gas and liquid phases, respectively:

$$\frac{\partial}{\partial \mathrm{t}}[\rho_G H_G V_G] + \frac{\partial}{\partial \mathrm{x}}\left[\rho_G H_G V_G^2 + H_G \Delta P_G\right] + H_G \frac{\partial P}{\partial x} = \tau_G + \tau_i - \rho_G g \ \mathrm{Sin}\theta \tag{3-48}$$

$$\frac{\partial}{\partial \mathrm{t}}[\rho_L H_L V_L] + \frac{\partial}{\partial \mathrm{x}}\left[\rho_L H_L V_L^2 + H_L \Delta P_L\right] + H_L \frac{\partial \mathrm{P}}{\partial \mathrm{x}} = \tau_L - \tau_i - \rho_L g \ \mathrm{Sin}\theta \tag{3-49}$$

In Equations 3-48 and 3-49, parameter P denotes the interface pressure; while V_k, ρ_k, and H_k are the velocity, the density, and the volume fraction of phase $k \in \{G, L\}$, respectively. The variables τ_i and τ_k are the interfacial and wall momentum transfer terms. The quantities ΔP_G and ΔP_L correspond to the static head around the interface (De Henau and Raithby, 1995), defined as follows:

$$\Delta P_G = P_G - P = -\rho_G \left[\frac{1}{2}\ \mathrm{Cos}\left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_G}\ \mathrm{Sin}^3\left(\frac{\omega}{2}\right)\right] g\, D\, \mathrm{Cos}\theta \tag{3-50}$$

$$\Delta P_L = P_L - P = -\rho_L\left[\frac{1}{2}\mathrm{Cos}\left(\frac{\omega}{2}\right) + \frac{1}{3\pi H_L}\mathrm{Sin}^3\left(\frac{\omega}{2}\right)\right] g\, D\, \mathrm{Cos}\theta \quad (3\text{-}51)$$

where ω is the wetted angle.

The detailed description of the solution algorithm for the preceding equations is based on a finite volume method (Masella et al., 1998).

One major limitation of this type of model is the treatment of the interfacial coupling. While this is relatively easy for separated flows (stratified and annular), this treatment is intrinsically flawed for intermittent flows.

3.4.3.2 Drift-Flux Model

The drift-flux model (DFM) is derived from the two-fluid model by neglecting the static head terms ΔP_G and ΔP_L in Equations 3-48 and 3-49 and replacing the two momentum equations by their sum. In the DFM this leads to a new model that consists of three partial differential equations, i.e., Equations 3-46, 3-47, and 3-52.

$$\frac{\partial}{\partial t}[\rho_G H_G V_G + \rho_L H_L V_L] + \frac{\partial}{\partial x}\left[\rho_G H_G V_G^2 + \rho_L H_L V_L^2 + P\right] = \tau_G + \tau_L - (\rho_G H_G + \rho_L H_L)\,\mathrm{Sin}\theta \quad (3\text{-}52)$$

The detailed description of the numerical solution for the relevant equations can be found in Faille and Heintze (1996).

The main advantages of this three-equation model are as follows:

- The equations are in conservative form, which makes their solution by the finite volume methods less onerous.
- The interfacial shear term, τ_i, is canceled out in the momentum equations, although it appears in an additional algebraic relation called the slip law.
- The model is well posed and does not exhibit complex characteristics.

Note that the drift-flux approach is best applied to closely coupled flows such as bubbly flow. Its application to stratified flows is, at best, artificial (Banerjee, 1986).

3.4.4 Multiphase Gas and Condensate Flow

In order to achieve optimal design of gas/condensate pipelines and downstream processing facilities, one needs a description of the relative amount of condensate and the flow regime taking place along the pipelines, where fluid flowing in pipelines may traverse the fluid phase envelope such that the fluid phase changes from single-phase to two-phase or vice versa. Hence, compositional single-phase/multiphase hydrodynamic modeling, which couples the hydrodynamic model with the natural gas phase behavior model, is necessary to predict fluid dynamic behavior in gas/condensate transmission lines. The hydrodynamic model is required to obtain flow parameters along the pipeline, and the phase behavior model is required for determining the phase condition at any point in the pipe, the mass transfer between the flowing phases, and the fluid properties.

Despite the importance of gas flow with low liquid loading for the operation of gas pipelines, few attempts have been made to study flow parameters in gas/condensate transmission lines. While the single-phase flow approaches have been previously applied to gas/condensate systems, only a few attempts have been reported for the use of two-fluid model for this purpose. Some of them have attempted to make basic assumptions (e.g., no mass transfer between gas and liquid phases) in their formulation, and others simply assume one flow regime for the entire pipe length. However, a compositional hydrodynamic model that describes the steady-state behavior of multiphase flow in gas/condensate pipelines was presented by Ayala and Adewumi (2003). The model couples a phase behavior model, based on the Peng and Robinson (1976) equation of state, and a hydrodynamic model, based on the two-fluid model. The proposed model is a numerical approach that can be used as an appropriate tool for engineering design of multiphase pipelines transporting gas and condensate. However, the complexity of this model precludes further discussion here.

Note that the presence of liquid (condensates), besides reducing deliverability, creates several operational problems in gas/condensate transmission lines. Periodic removal of the liquid from the pipeline is thus desirable. To remove liquid accumulation in the lower portions of pipeline, pigging operations are performed. These operations keep the pipeline free of liquid, reducing the overall pressure drop and at the same time increase the pipeline flow efficiency. However, the pigging process associates with transient flow behavior in the pipeline. Thus, it is imperative to have a means of predicting transient behavior encountered in multiphase, gas/gas-condensate pipelines. Until recently, most available commercial codes were based on the two-fluid model. However, the model needs many modifications to be suitable for simulating multiphase transient flow in gas/gas-condensate transmission lines. For example, the liquid and gas continuity equations need to be modified to account for the mass transfer between phases. So far, several codes have been reported for this purpose, where three main commercial transient codes are OLGA (Bendiksen et al., 1987, 1991), PROFES (Black et al., 1990), and TACITE (Pauchon et al., 1993, 1994). Detailed discussion of these codes is beyond the scope of this book; readers are referred to the original papers for further information.

3.5 PREDICTING TEMPERATURE PROFILE OF MULTIPHASE PIPELINES

Predicting the flow temperature and pressure changes has become increasingly important for use in both the design and operation of flow transmission pipelines. It is therefore imperative to develop appropriate methods capable of predicting these parameters for multiphase pipelines. A simplified flow chart of a suitable computing algorithm is shown in Figure 3-6. This algorithm calculates pressure and temperature along the pipeline by iteratively converging on pressure and temperature for each sequential "segment" of the pipeline. The algorithm converges on temperature in the outer loop and pressure in the inner loop because robustness and computational speed are obtained when converging on the least sensitive variable first (Brill and Beggs, 1991).

The pipeline segment length should be chosen such that the fluid properties do not change significantly in the segment. More segments are recommended for accurate calculations for a system where fluid properties can change drastically over short distances. Often, best results are obtained when separate segments (with the maximum segment length less than about 10% of total line length) are used for up, down, and horizontal segments of the pipeline (Brill and Beggs, 1991).

Prediction of the pipeline temperature profile can be accomplished by coupling the pressure gradient and enthalpy gradient equations as follows (Brill and Beggs, 1991):

$$\Delta h = \frac{-V_M V_{SG} \Delta P}{(778)(32.17)(\bar{P})} + \frac{\Delta Z}{778} - \frac{U\pi D(\bar{T} - T_a)\Delta L}{3600(\dot{M}_M)} \tag{3-53}$$

where Δh is enthalpy change in the calculation segment, Btu/lb$_m$; V_M is velocity of the fluid, ft/sec; V_{SG} is superficial gas velocity, ft/sec; ΔP is estimated change in pressure, psi; $\bar{P}$ is average pressure in calculation segment, psia; ΔZ is change in elevation, ft; U is overall heat transfer coefficient, Btu/hr-ft^2-°F; D is reference diameter on which U is based, ft; $\bar{T}$ is estimated average temperature in calculation segment, °F; T_a is ambient temperature, °F; ΔL is change in segment length, ft; and $\dot{M}_M$ is gas-liquid mixture mass flow rate, lb$_m$/s.

As can be seen from Equation 3-53, temperature and pressure are mutually dependent variables, so generating a very precise temperature profile requires

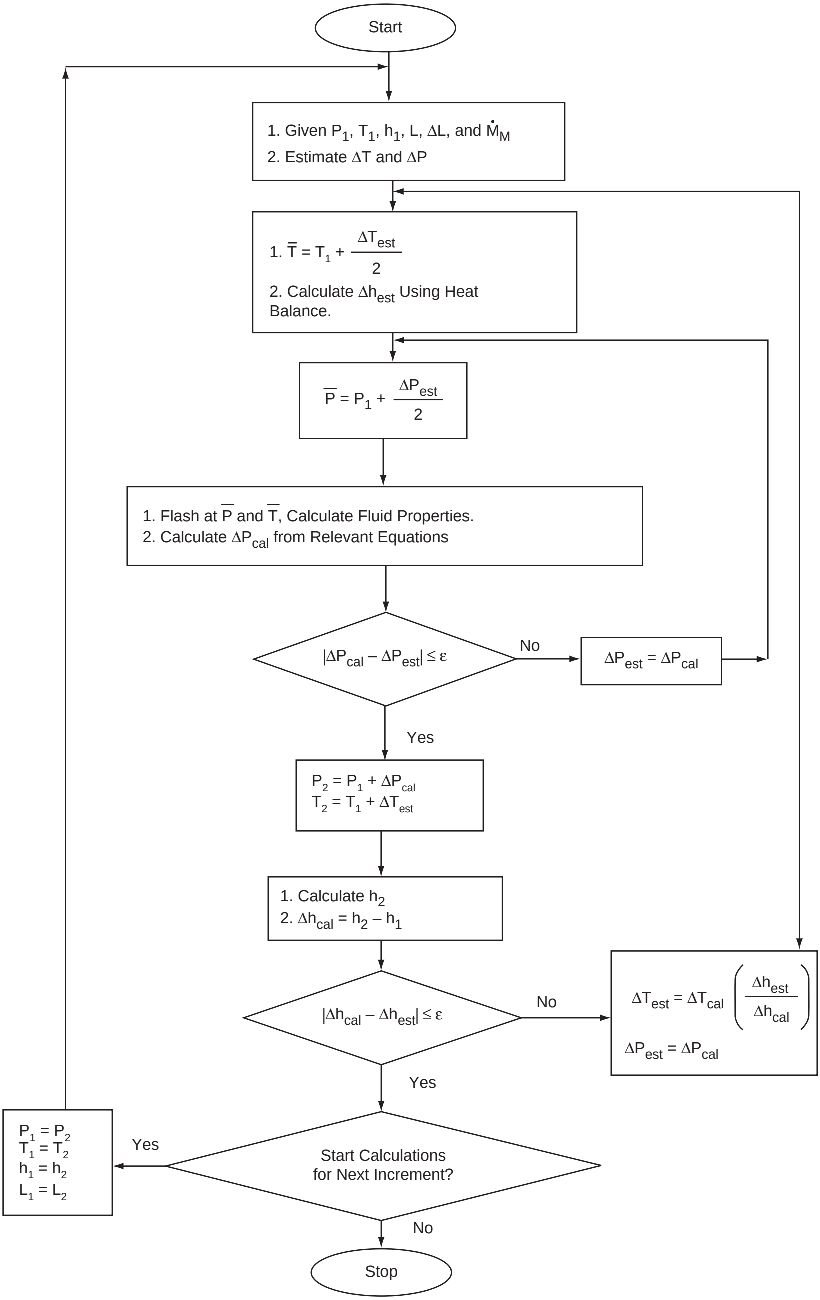

Figure 3-6 Pressure and temperature calculation procedure (Brill and Beggs, 1991).

numerous iterative calculations. The temperature and pressure of each pipe segment is calculated using a double-nested procedure in which for every downstream pressure iteration, convergence is obtained for the downstream temperature by property values evaluated at the average temperature and pressure of that section of pipe. Experience shows that it is essential to use a good pressure-drop model to assess the predominant parameter, i.e., pressure.

The overall heat transfer coefficient (*U*) can be determined by the combination of several coefficients, which depend on the method of heat transfer and pipe configuration. Figure 3-7 shows a cross-section of a pipe, including each "layer" through which heat must pass to be transferred from the fluid to the surroundings or vice versa. This series of layers has an overall resistance to heat transfer made up of the resistance of each layer.

In general, the overall heat transfer coefficient for a pipeline is the reciprocal of the sums of the individual resistances to heat transfer, where each resistance definition is given in Table 3-4.

The individual resistances are calculated from the given equations in Table 3-5.

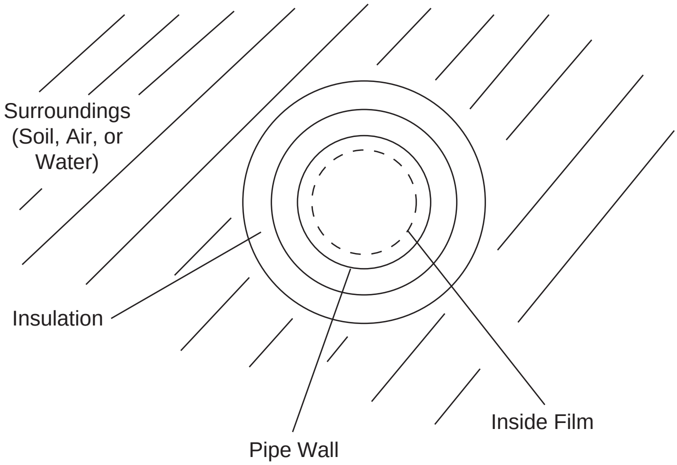

Figure 3-7 Cross-section of a pipe showing resistance layers.

Table 3-4 Types of Resistance Layers for a Pipe

Resistance	**Due to**
$R_{\text{inside, film}}$	Boundary layer on the inside of the pipe
R_{pipe}	Material from which the pipe is made
$R_{\text{insulation}}$	Insulation (up to five concentric layers)
R_{surr}	Surroundings (soil, air, water)

Table 3-5 Heat Transfer Resistances for Pipes (Hein, 1984; Brill and Beggs, 1991)

$R_{\text{inside, film}} = D/[0.027\ K_f\,\text{Re}^{0.8}\ \text{Pr}^{0.33}]$	For stationary surroundings like soil,
D is reference diameter on which U is based, inch	
K_f is thermal conductivity of the fluid, Btu/hr-ft-°F	$R_{\text{surr}} = \dfrac{D\,\text{Ln}\left[(2D' + (4D'^2 - D_t^2)^{0.5})/D_t\right]}{24k_{\text{surr}}}$
Re is Reynolds number $= \dfrac{(\rho_L\lambda_L + \rho_G\lambda_G)(V_{SL} + V_{SG})(D_i)(124.016)}{\lambda_L\mu_L + \lambda_G\mu_G}$	D is reference diameter on which U is based, inch
	k_{surr} is soil thermal conductivity, Btu/hr-ft-°F
	D' is depth from top of soil to pipe centerline, ft
	D_t is diameter of the pipe plus insulation, ft
ρ_L is density of the liquid, lb/ft^3	For fluid surroundings like air or water,
λ_L is no-slip liquid holdup	
ρ_G is density of the gas, lb/ft^3	$R_{\text{surr}} = \dfrac{D}{12\ k_{\text{surr}}10^{[0.26694(\text{LogRe}_{\text{surr}})^{1.3681}]}}$
λ_G is no-slip gas holdup $= 1-\lambda_L$	
V_{SL} is liquid superficial velocity, ft/sec	
V_{SG} is gas superficial velocity, ft/sec	D is reference diameter on which U is based, inch
D_i is inside diameter of the pipe, inch	k_{surr} is thermal conductivity of the surroundings, Btu/hr-ft-°F
μ_L is liquid viscosity, cp	

μ_G is gas viscosity, cp
Pr is Prandtl number $= 2.42\,(\mu_L\lambda_L + \mu_G\lambda_G)(C_{pL}\lambda_L + C_{p,G}\,\lambda_G)/\,K_f$

C_p is specific heat at constant pressure, Btu/lb-°F

$$R_{pipe} = \frac{D\,\mathrm{Ln}(D_o/D_i)}{24k_p}$$

k_p is thermal conductivity of the pipe, Btu/hr-ft-°F
D_o is outside diameter of the pipe, ft
D_i is inside diameter of the pipe, ft

$$R_{insulation} = \frac{D\sum_{j=1}^{n}\frac{1}{k_j}\mathrm{Ln}(D_j/D_{j-1})}{24}$$

D is reference diameter on which U is based, inch
k_j is conductivity of jth layer of insulation, Btu/hr-ft-°F
D_j is outer diameter of the jth layer, ft
D_{j-1} is inner diameter of the jth layer, ft

Re_{surr} is Reynolds number for the surroundings is $0.0344\,\rho_{surr}\,V_{surr}\,D_t/\mu_{surr}$
ρ_{surr} is density of ambient fluid, lb/ft^3
V_{surr} is velocity of ambient fluid, ft/hr

D_t is total diameter of pipe plus insulation, inch

μ_{surr} is viscosity of surrounding fluid, cp

Since fluid properties are key inputs into calculations such as pressure drop and heat transfer, the overall simulation accuracy depends on accurate property predictions of the flowing phases. Most of the required physical and thermodynamic properties of the fluids are derived from the equation of state. However, empirical correlations are used for calculation of viscosity and surface tension. When the physical properties of the fluids are calculated for two-phase flow, the physical properties of the gas and liquid mixture can be calculated by taking the mole fraction of these components into account. Similarly, when water is present in the system, the properties of oil and water are combined into those of a pseudo-liquid phase. To produce the phase split for a given composition, pressure, and temperature, one must use an equilibrium flash calculation utilizing an appropriate equation of state. A simple and stable three-phase flash calculation with significant accuracy for pipeline calculation is described in Appendix 1.

3.6 VELOCITY CRITERIA FOR SIZING MULTIPHASE PIPELINES

After a line size is selected to meet throughput and pressure drop constraints, it is important to check whether acceptable flowing conditions exist within the pipeline. However, a more useful means of assessing whether flowing conditions are acceptable is to check whether fluid velocities lie within certain limits.

At the low end of the normal throughput range, the actual (not superficial) liquid velocity should ideally be greater than 3 ft/s. This will ensure that sand and water are continuously transported with the liquid phase and not allowed to drop out and accumulate at the bottom of the pipe. However, at the maximum throughput conditions, the maximum mixture velocity in the multiphase pipeline must be calculated in order to check that its value does not exceed the erosional velocity limit.

The current industry standard method of determining the erosional velocity limit is through use of the relationship given in API RP 14E (1991):

$$V_e = \frac{C}{(\rho_M)^{0.5}} \tag{3-54}$$

where V_e is the maximum acceptable mixture velocity to avoid excessive erosion, (ft/s); C is empirical constant; and ρ_M is the no-slip mixture density at flowing conditions, lb/ft^3.

Industry experience to date indicates that for solid-free fluids, a C factor of 100 for continuous service and a C factor of 125 for intermittent service are conservative.

The preceding simple criterion is specified for clean service (noncorrosive and sand-free), and that the limits should be reduced if sand or corrosive conditions are present. However, no guidelines are provided for these reductions (McLaury and Shirazi, 2000; Salama, 2000).

If the erosional velocity limit is exceeded, either the line diameter must be increased or the production profile constrained to reduce the maximum mixture velocity. Alternatively, a higher grade of pipeline material could be used.

As well as the erosional velocity limit, API RP 14E (1991) also recommends that a line velocity of 60 ft/sec should not be exceeded to ensure that the level of noise emitted by the flow is not excessive. In addition, in some pipeline systems, flowing velocities need to be limited by a requirement to avoid removal of corrosion inhibitor film, ensuring a lubricating effect that shifts the erosion velocity limit.

3.7 MULTIPHASE PIPELINE OPERATIONS

After a pipeline is installed, efficient operating procedures must be provided to maintain safe, long-term operation of the pipeline in the face of unexpected upsets, and to improve efficiency and economics of operation. Leak detection, depressurization, and pigging are typically important procedures.

3.7.1 Leak Detection

Leaks in the pipelines originate from a variety of causes, which may include material-related damage, physical damage caused by construction in the right-of-way, etc. Because accidental product discharges cannot be entirely eliminated, one of the most effective methods of reducing the impact of spills is to quickly detect the leak and to act quickly to stop the discharge. Methods used to detect product leaks along the pipeline can be divided in two categories: external and internal leak detection systems (API, 1995a). Externally based methods include traditional procedures such as visual inspection, as well as technologies like hydrocarbon sensing using fiber-optic or dielectric cables (Sandberg et al., 1989). Internally based methods, also known as computational pipeline monitoring (CPM), use instruments to monitor internal pipeline parameters (i.e., flow, pressure, temperature and fluid properties), which are continuously input into computer simulation software linked to a Supervisory Control and Data Acquisition (SCADA) system that mathematically or statistically analyzes the information. Leaks result in unexpected variations or well-defined deviation patterns between simulated and measured values. These patterns can be detected and assessed to determine if a leak is present (API, 1995b). The method of leak detection selected for a pipeline is dependent on a variety of factors, including physical pipeline characteristics (length, diameter, thickness, etc.), product characteristics (density, viscosity, etc.), instrumentation and communications capabilities, and economics (Muhlbauer, 1996). Compared to other leak detection methods, the SCADA-based leak detection methods have the widest range of applicability to pipeline leak detection and are by far the most highly developed of the leak detection methods presented as they can rapidly detect large leaks and, over a period of time, detect smaller leaks as well (Jolly et al., 1992). However, for multiphase flow pipelines, the SCADA-based leak detection methods become much more difficult to apply, and the sensitivity to detecting leaks is reduced because metering multiphase lines is very difficult and inaccurate. Most condition the flow so as to create a measurable flow stream by making a uniform mist or separate the gas and liquid content. The big problem in multiphase leak detections is the varying percentage of gas and liquid content and, most notably, the formation and expulsion of liquid slugs in the pipeline. Hence, for multiphase SCADA-based leak detection systems, accurate and reliable flow measurement instrumentation and transient flow computer models will need to be developed. Moreover, the capabilities and advantages/disadvantages of new technologies should be analyzed with special consideration given to the possible application in subsea pipelines, for which the remoteness of these pipelines, coupled with a number of complex interactions between the released fluids and the subsea environment, makes detection much more difficult.

3.7.2 Pipeline Depressurization

Pipeline depressuring is generally used to refer to the controlled and relatively slow evacuation of a pipeline system. Depressuring is usually performed for process reasons/maintenance requirements. When pipelines transport a gas and liquid mixture, a rupture can cause rapid depressuring and critical flow at the break; additionally, with gas/condensate lines, the temperature can drop rapidly due to Joule-Thomson cooling. This can have implications on the brittle fracture of the pipeline and safety valves that may cool significantly before being actuated.

The simulation of the depressuring of a pipeline system is performed in order to determine the minimum temperature experienced or to calculate the time required to depressure a system such that the temperature does not drop below the minimum specification value.[2] The transient analysis of pipeline depressuring can become very complex depending on what is required from the analysis and to what accuracy. OLGA commercial software can simulate the depressuring of multiphase pipelines and is recommended for most cases.

3.7.3 Pigging

"Pigging" is a term used to describe a mechanical method for removing contaminants and deposits within the pipe or to clean accumulated liquids in the lower portions of hilly terrain pipelines, using a mechanized plunger or pigs (Tiratsoo, 1992). Because of the ability of the pig to remove both corrosion products and sludge from the line, it has been found to be a positive factor in the corrosion control of the offshore pipelines. Some specially instrumented pigs, known as "intelligent pigs," may also be used intermittently for the purpose of pipeline integrity monitoring (Cranswick, 2001), which includes detecting wall defects (e.g., corrosion, weld defects, and cracks). Pipeline pigs fit the inside diameter of the pipe and scrape the pipe walls as they are pushed along the pipeline by the flowing fluid. For offshore platforms the pig is launched from offshore and received in the onshore pig catcher. The receiver is in a direct line with the sealine and can be isolated from it to allow the pigs to be removed. Pigs are available in various shapes (see Figure 3-8) and are made of different materials, depending on the pigging task to be accomplished. Some have spring-loaded steel knives, wire brushes, or abrasive grit surfaces for removal of adhering contaminants. Others are semirigid, nonmetallic spheres.

During pigging operations there are often technical problems due to the lack of reliable tools for the prediction of the many variables related to the motion of the pig through the pipeline. Hence, the pigging operation requires careful control and coordination. For example, overfrequent pigging causes production downtime or higher operation expenses, and infrequent pigging results in less production reduction but increases the risk of pipeline blockages, including sticking a pig. The pipeline operator must therefore

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[2] The time taken to fully depressure a pipeline to atmospheric pressure will depend on several factors, not least of which will include the size and type of pipeline inventory, the operating pressure at time of rupture, the rate of flow escaping, and the maximum vent rate at the end section.

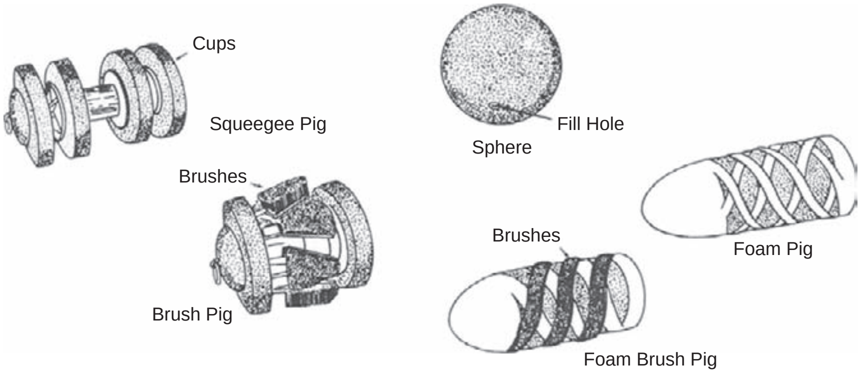

Figure 3-8 Typical types of pigs (Campbell, 1992).

give serious consideration to whether the pipeline really needs to be pigged and whether it is economical to do so. Since there is no commercially available tool to determine the optimum pigging frequency, operators must choose their pigging frequency using rules of thumb based on their field experience, which often involves a high degree of uncertainty.

In pigging operations where the pipe content is unloaded, the liquid holdup builds up as a slug ahead of the moving pig. The arrival of a slug at production or processing equipment is problematic. It causes both mechanical problems (due to high velocities and momentum) and process problems (increasing liquid levels, causing surges and trips). There are, in general, several means to reduce slugging in pipelines (Haandrikman et al., 1999; Havre et al., 2000). In some cases, operators can minimize liquid accumulation by managing fields and pipelines in such a way as to create a suitable fluid flow regime (i.e., mist flow regime) under which the gas velocity is high enough to keep liquids continuously dispersed. While it is desirable to design the lines to avoid slugging, in practice this can be difficult while maintaining the ability to turn down the pipeline flow rate. In these cases, consideration should be given to providing suitable process equipment to handle possible slugging (Xiao and Shoup, 1998).

The pigging operation in multiphase pipelines is a transient operation. Transient flow is observed not only during the pig running time, but also for a long period after the pig exits. This situation occurs even if the inlet gas and liquid flow rates and outlet pressures are kept constant (Minami, 1991). Analysis of such transient flow behavior in a pipeline is necessary not only for designing the downstream processing facilities, but also for establishing safe operating procedures. Hence, there is a definite need to develop reliable and comprehensive pigging models for better understanding of transient behavior of fluids during these operations. A simple model to simulate transient flow behavior in a two-phase flow pipeline under a pigging operation was presented by Minami (1991). In this model, Minami (1991) assumed that the gas phase can be considered to be flowing in a quasi-steady condition, and then coupled it with the Taitel et al. (1989) simplified transient two-fluid model. The model, however, needs significant modifications in order to be used for simulating transient flow in a pipeline-riser system, where the quasi-steady-state approach is not suitable for such systems due to high accumulation of gas upstream of the pig (Yeung and Lima, 2002). For this purpose, a new transient two-fluid model was developed by Yeung and Lima (2002), which is appropriate for estimating the two-phase flow pigging hydraulics, especially in pipeline-riser systems.

3.8 MULTIPHASE FLOW ASSURANCE

Flow assurance,[3] which is a critical component in the design and operation of multiphase production facilities, refers to ensuring the flow of produced hydrocarbons from the reservoir to the downstream processing facilities. Flow assurance encompasses the thermal-hydraulic design and assessment of multiphase production/transportation systems as well as the prediction, prevention, and remediation of flow problems such as gas hydrate formation, wax and asphaltene deposition on walls, corrosion, erosion, scaling, emulsions, foaming, and severe slugging. In all cases, flow assurance designs must consider the capabilities and requirements for all parts of the system throughout the entire production life of the system to reach a successful solution for securing the production operations, minimizing the downtime, and reducing the production/transportation costs.

The following sections address the major flow assurance issues for the raw gas transmission lines and summarizes the commonly used mitigation practices in industry for such purpose. They also provide innovative techniques and practical strategies to design the deepwater pipelines, where traditional approaches are inappropriate for deepwater development systems due to extreme distances, depths, temperature, or economic constraints (Wilkens, 2002).

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[3] “Flow assurance” is a term originally coined by Petrobras in the early 1990s. The term in Portuguese was “Garantia de Fluxo,” which translates literally to “Guarantee the Flow.”

3.8.1 Gas Hydrates

A gas hydrate is an ice-like crystalline solid called a clathrate, which occurs when water molecules form a cage-like structure around small guest molecules at certain temperatures and pressures. The most common guest molecules are methane, ethane, propane, isobutane, normal butane, nitrogen, carbon dioxide, and hydrogen sulfide, of which methane occurs most abundantly in natural hydrates. Several different hydrate structures are known. The two most common are structure I and structure II. Type I forms with smaller gas molecules such as methane, ethane, hydrogen sulfide, and carbon dioxide, whereas structure II is a diamond lattice, formed by large molecules like propane and isobutane (Sloan, 1998). However, nitrogen, a relatively small molecule, also forms a Type II hydrate (Carroll, 2003). In addition, in the presence of free water, the temperature and pressure can also govern the type of hydrate structure, where the hydrate structure may change from structure II hydrate at low temperatures and pressures to structure I hydrate at high pressures and temperatures (GPSA, 2004). It should be noted that n-butane does form a hydrate but is very unstable (Ng and Robinson, 1976; Kumar, 1987). However, it will form a stabilized hydrate in the presence of small "help" gases such as methane or nitrogen (Edmonds et al., 1998). It has been assumed that normal paraffin molecules larger than n-butane are nonhydrate formers (Kumar, 1987). Furthermore, the existence of another hydrate structure, Type H, was described by Ripmeester et al. (1987). Some isoparaffins and cycloalkanes larger than pentane are known to form structure H hydrates (Mehta and Sloan, 1996). However, little is known about type H structures.

While many factors influence hydrate formation, the two major conditions that promote hydrate formations are (1) the gas being at the appropriate temperature and pressure, and (2) the gas being at or below its water dew point (Sloan, 1998). Note that free water is not necessary for hydrate formation, but it certainly enhances hydrate formation (Carroll, 2003). Other factors that affect but are not necessary for hydrate formation include turbulence, nucleation sites, surface for crystal formation, agglomeration, and the salinity of the system. The exact temperature and pressure at which hydrates form depend on the composition of the gas and the water. For any particular composition of gas at a given pressure, there is a temperature below which hydrates will form and above which hydrates will not form. As the pressure increases, the hydrate formation temperature also increases. As a general rule, when the pressure of the gas stream increases or as the gas becomes colder, the tendency to form hydrates increases (GPSA, 2004). Hence, many gas-handling systems are at significant risk of forming hydrate plugs during shut-in and subsequent start-up (Hunt, 1996; Ramachandran et al., 2000; Wilkens, 2002).

Although gas hydrates may be of potential benefit both as an important source of hydrocarbon energy and as a means of storing and transmitting natural gas, they represent a severe operational problem as the hydrate crystals may deposit on the pipe wall and accumulate as large plugs that can completely block pipelines, shutting in production (Edmonds et al., 1998; Sloan, 1998). Acceleration of these plugs due to a pressure gradient can also cause considerable damage to production facilities. In addition, the remediation of hydrate blockages can present significant technical difficulties with major cost implications (Mehta et al., 2001). Because of these problems, methods of preventing hydrate solids development in gas production/transportation systems have been of considerable interest for a number of years (Bufton, 2003).

3.8.1.1 Hydrate Locus for Natural Gas Components

The thermodynamic stability of hydrates, with respect to temperature and pressure, may be represented by the hydrate curve. The hydrate curve represents the thermodynamic boundary between hydrate stability and dissociation. Conditions to the left of the curve represent situations in which hydrates are stable and “can” form. Operating under such conditions does not necessarily mean that hydrates will form, only that they are possible. Figure 3-9 shows the hydrate locus for natural gas components. The extension to mixtures is not obvious from this diagram. The hydrate curve for multicomponent gaseous mixtures may be generated by a series of laboratory experiments, or more commonly, is predicted using thermodynamic software based on the composition of the hydrocarbon and aqueous phases in the system.

The thermodynamic understanding of hydrates indicates the conditions of temperature, pressure, and composition that hydrates are stable. However, it does not indicate when hydrates will form and, more importantly, whether they will cause blockages in the system.

3.8.1.2 Prediction of Hydrate Formation Conditions

There are numerous methods available for predicting hydrate formation conditions. Three popular methods for rapid estimation of hydrate

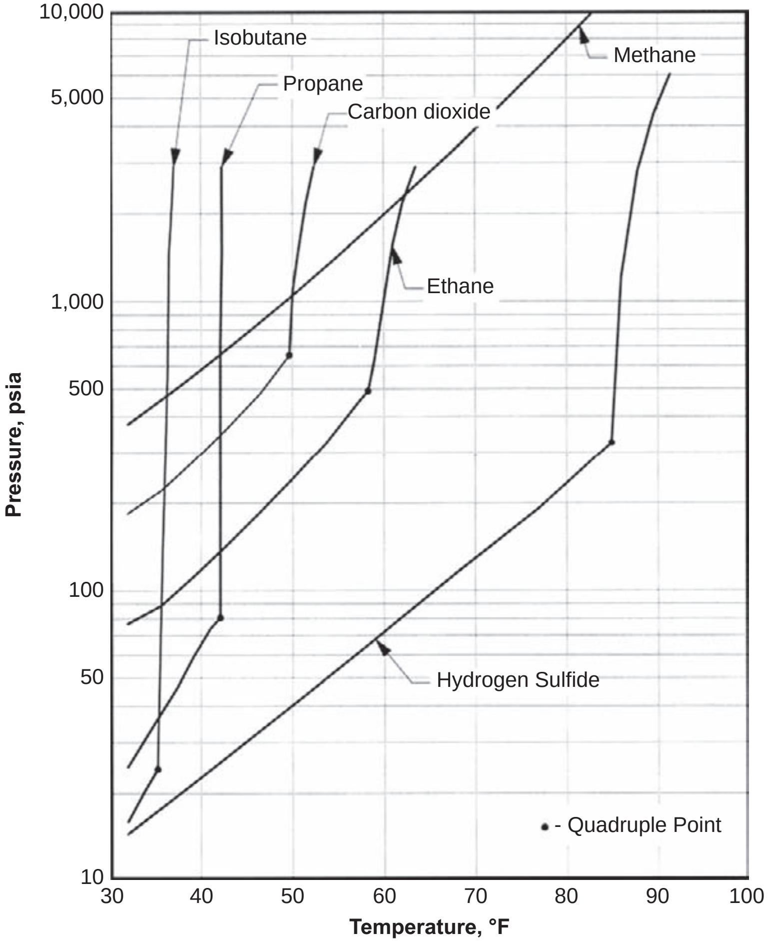


Figure 3-9 Hydrate-forming conditions for natural gas components (Carroll, 2003).

formation conditions are discussed in the following sections. A detailed discussion of other methods that are, perhaps, beyond the scope of the present discussion can be found in publications by Kumar (1987), Sloan (1998), and Carroll (2003).

3.8.1.2.1 *K*-Factor Method

The *K*-factor method was originally developed by Carson and Katz (1942), although additional data and charts have been reproduced since then. In this method, the hydrate temperature can be predicted using vapor–solid (hydrate) equilibrium constants. The basic equation for this prediction is (Carson and Katz, 1942)

$$\sum_{i=1}^{n}\left(\frac{y_\mathrm{i}}{K_\mathrm{i}}\right) = 1.0 \tag{3-55}$$

where y_i is mole fraction of component i in gas on a water-free basis, K_i is vapor-solid equilibrium constant for component i, and n is the number of components.

The calculation is iterative, and the incipient solid formation point will determine when the preceding equation is satisfied. This procedure is akin to a dew point calculation for multicomponent gas mixture.

The vapor-solid equilibrium constant is determined experimentally and is defined as the ratio of the mole fraction of the hydrocarbon component in gas on a water-free basis to the mole fraction of the hydrocarbon component in the solid on a water-free basis (Carson and Katz, 1942):

$$K_\mathrm{i} = \left(\frac{y_\mathrm{i}}{x_\mathrm{i}}\right) \tag{3-56}$$

where x_i is mole fraction of component i in solid on a water-free basis.

Figures 3-10 through 3-14 provide the vapor-solid equilibrium constants at various temperatures and pressures. For nitrogen and components heavier than butane, the equilibrium constant is taken as infinity. It should be stressed that in the original method of Carson and Katz (1942), it was assumed that nitrogen is a nonhydrate former and that n-butane, if present in mole fractions less than 5%, has the same equilibrium constant as ethane. Theoretically, this assumption is not correct, but from a practical viewpoint, even using an equilibrium constant equal to infinity for both nitrogen and n-butane provides acceptable engineering results (Campbell, 1992).

The Carson and Katz (1942) method gives reasonable results for sweet natural gases and has been proven to be appropriate up to about 1,000 psia (GPSA, 2004). However, Mann et al. (1989) presented new K-value charts that cover a wide range of pressures and temperatures. These charts can be an alternate to the Carson and Katz (1942) K-value charts, which are not a function of structure or composition.

Poettmann et al. (1989) also developed vapor-solid equilibrium ratio charts for various natural gas systems using the Colorado School of Mines

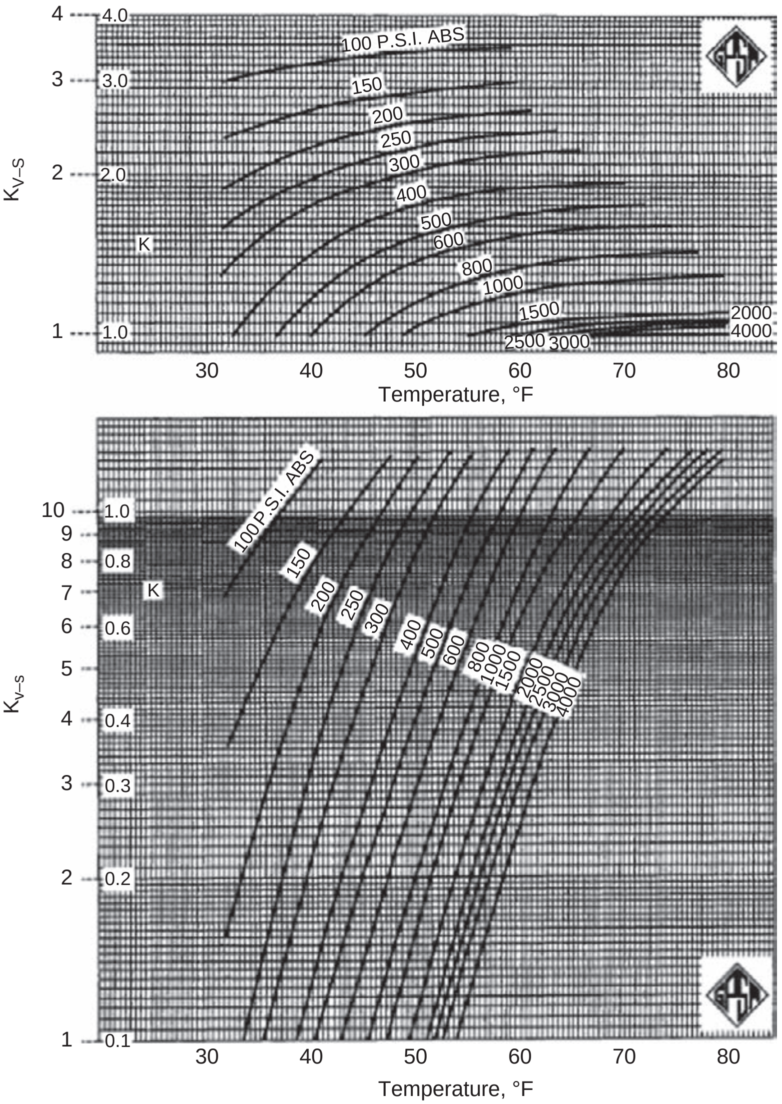


Figure 3-10 Vapor-solid equilibrium constants for methane (top) and ethane (bottom) (GPSA, 2004).

hydrate program. These charts covered a wide range of pressure, temperature, and composition. Polynomial regression analysis was used to curve fit the computer-generated *K* values for various systems so that the resulting equations could be programmed and used to predict the hydrate-forming conditions for natural gases.

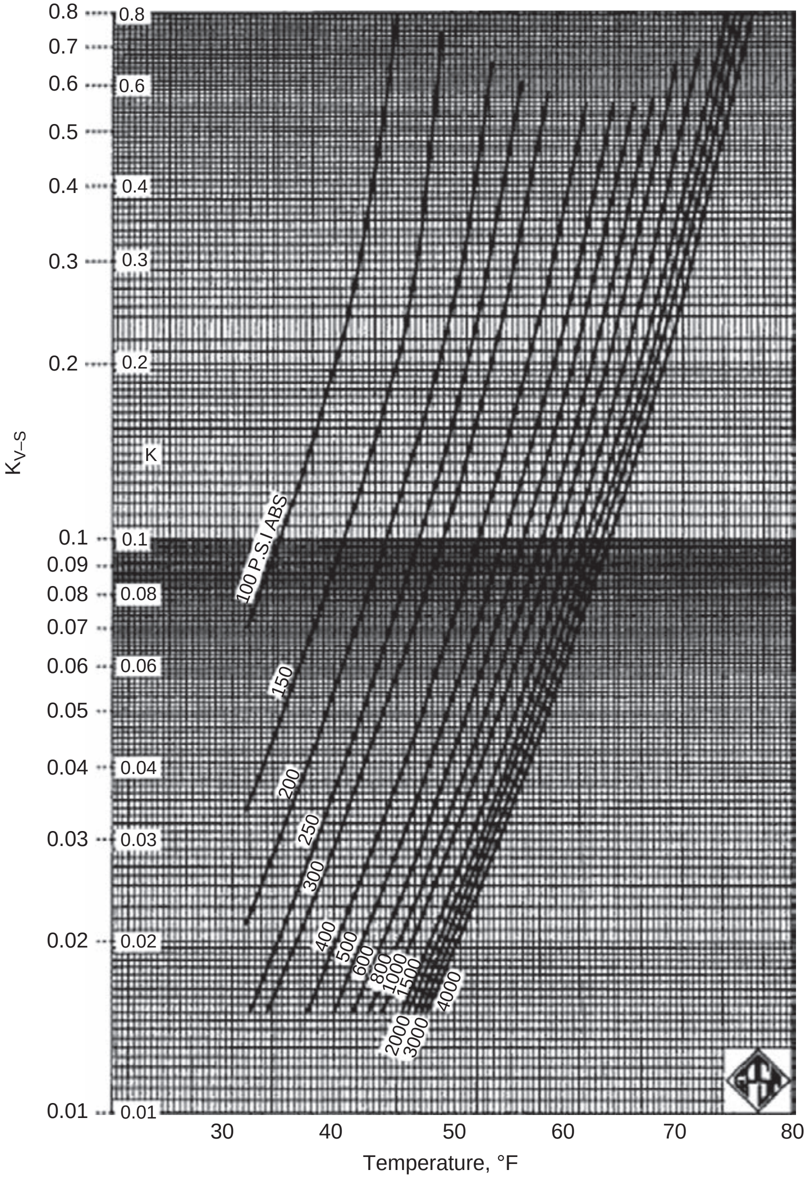


Figure 3-11 Vapor-solid equilibrium constant for propane (GPSA, 2004).

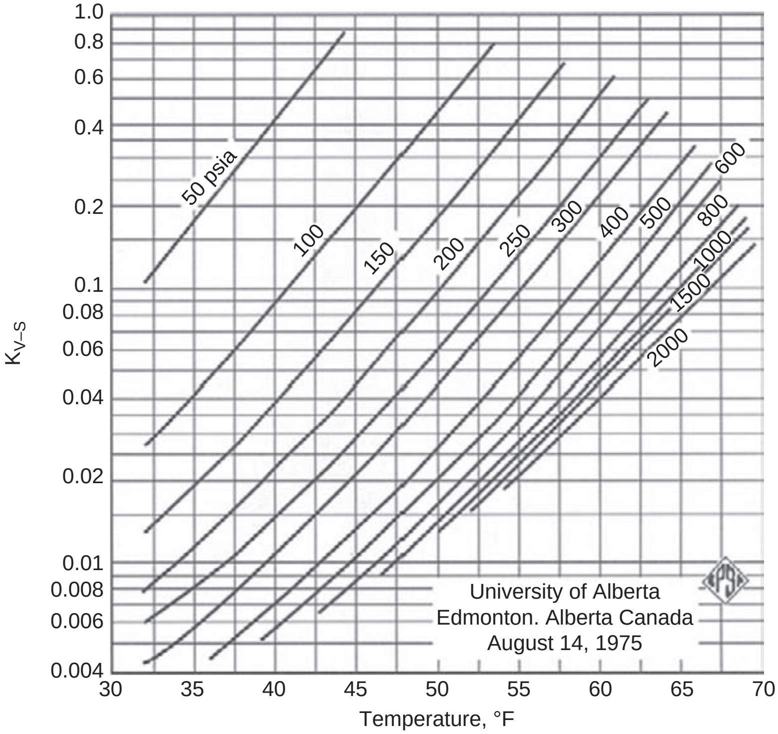


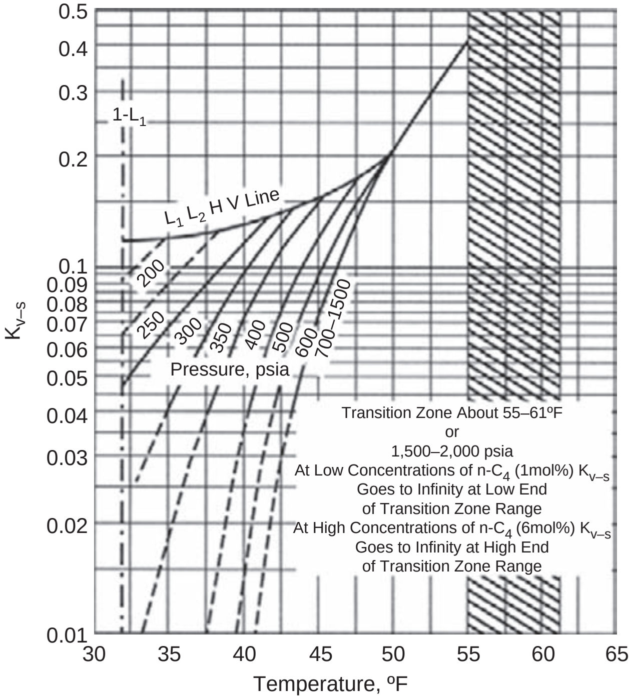
Figure 3-12 Vapor-solid equilibrium constants for isobutane (GPSA, 2004).

3.8.1.2.2 Baillie and Wichert Method

The method presented by Baillie and Wichert (1987) is a chart method (see Figure 3-15) that permits the estimation of the hydrate formation temperatures at pressure in the range of 100 to 4,000 psia for natural gas containing up to 50% hydrogen sulfide and up to 10% propane (Carroll, 2003). The method may not apply to a sweet gas mixture containing CO_2, but is considered fairly accurate if the CO_2 is less than about 5 mol% (Carroll, 2004).

3.8.1.2.3 Gas Gravity Method

Until now, several other methods have been proposed for predicting hydrate-forming conditions in natural gas systems. The most reliable of these require a gas analysis. However, if the gas composition is not known, even the previous methods cannot be used to predict the hydrate formation conditions. The Katz (1945) gravity chart (see Figure 3-16) can be used to predict the approximate pressure and temperature for hydrate formation, provided hydrates exist in the pressure-temperature region above



Transition Zone About 55–61ºF
or
1,500–2,000 psia
At Low Concentrations of n-C_4 (1mol%) K_{V-S}
Goes to Infinity at Low End
of Transition Zone Range
At High Concentrations of n-C_4 (6mol%) K_{V-S}
Goes to Infinity at High End
of Transition Zone Range

Figure 3-13 Vapor-solid equilibrium constants for n-butane (GPSA, 2004).

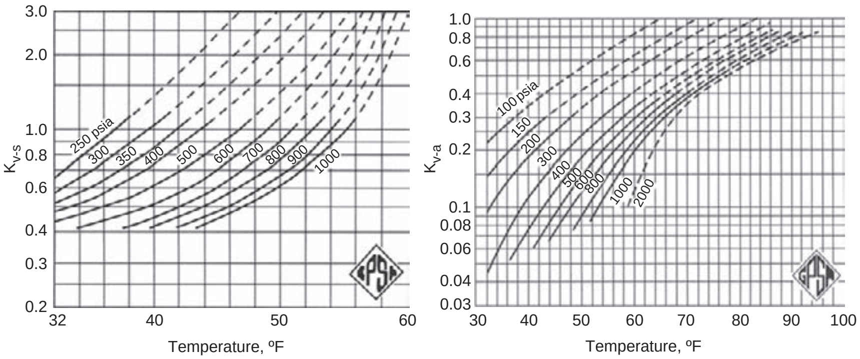


Figure 3-14 Vapor-solid equilibrium constants for carbon dioxide (left) and for hydrogen sulfide (right) (GPSA, 2004).

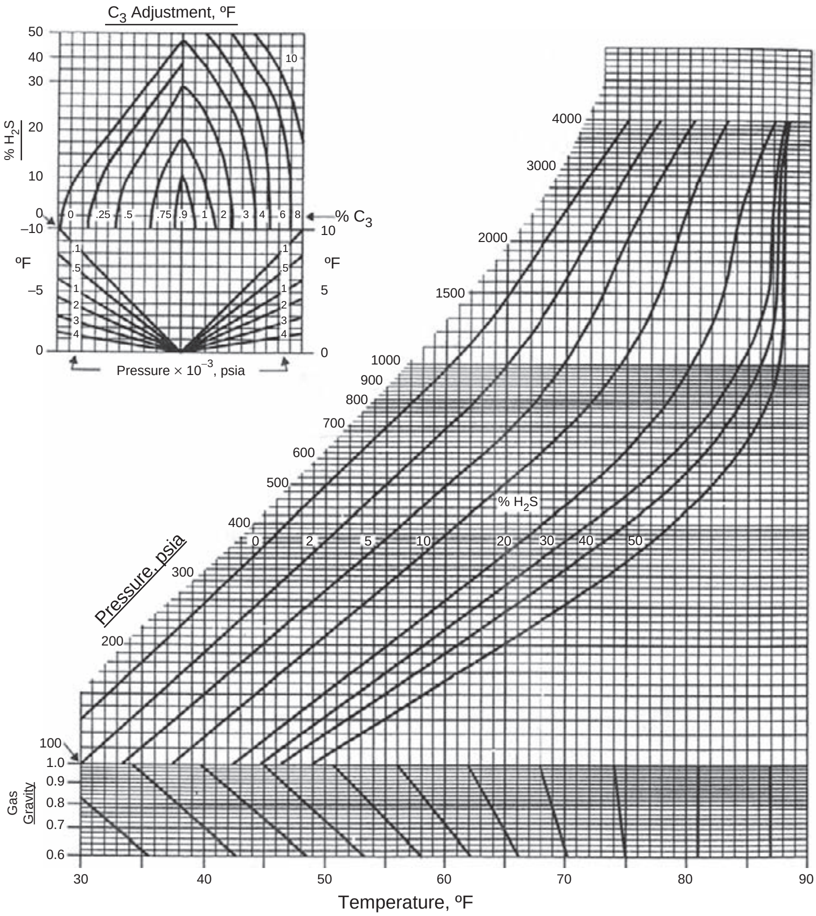


Figure 3-15 Baillie and Wichert chart for estimating hydrate formation conditions (GPSA, 2004).

the appropriate gravity curve. Therefore, as a first step to predict hydrate formation temperature, one can develop an appropriate equation representing the Katz (1945) gravity chart. Such a correlation uses two coefficients that correlate temperature, pressure, and specific gravity of gas (Towler and Mokhatab, 2005):

$$T_h = 13.47\,\mathrm{Ln}(P) + 34.27\,\mathrm{Ln}(\mathrm{SG}) - 1.675[\mathrm{Ln}(P) \times \mathrm{Ln}(\mathrm{SG})] - 20.35 \tag{3-57}$$

where T is gas flow temperature, °F; P is gas flow pressure, psia; and SG is specific gravity of gas (air = 1.0).

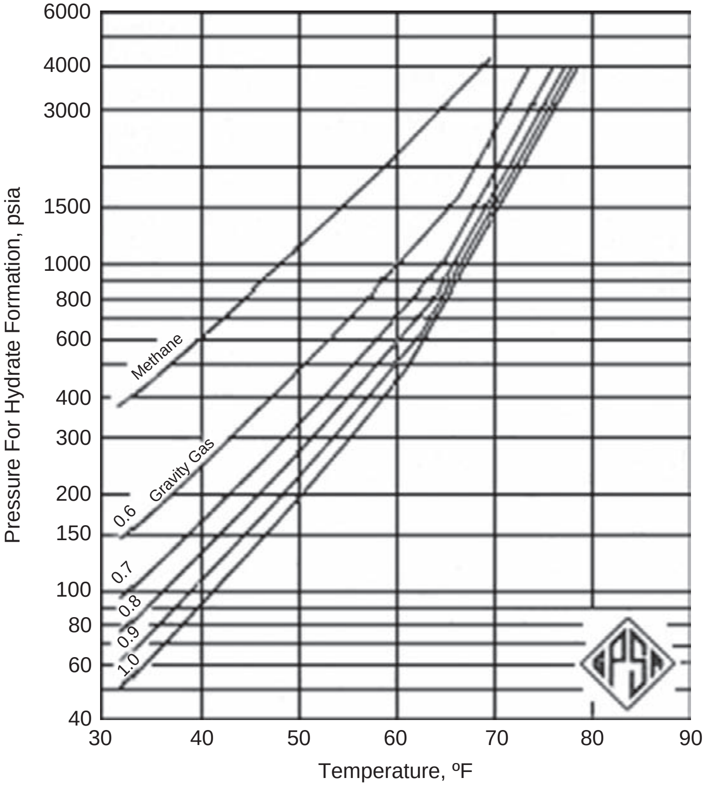


Figure 3-16 Katz's gravity chart for predicting hydrate formation conditions (GPSA, 2004).

Note that while Equation 3-57 is based on the GPSA chart, it is only accurate up to 65°F. Beyond that, it overestimates the temperature slightly.

The Katz (1945) gravity chart was generated from a limited amount of experimental data and a more substantial amount of calculations based on the *K*-value method. The components used for the construction of this chart are methane, ethane, propane, butane, and normal pentane, and therefore using this chart for compositions other than those used to derive these curves will produce erroneous results (Sloan, 1998). In fact, this method is an appropriate method of estimating hydrate formation conditions for sweet natural gas mixtures. However, the Baillie and Wichert (1987) method is better than this chart when applied to sweet gas, because of the inclusion of a correction factor for propane (Carroll, 2003).

3.8.1.2.4 Commercial Software Programs

Hydrate formation conditions based on fluid compositions are normally predicted with commercially available software programs. These programs are generally quite good and so simple to use that they often require less time than the simplified methods presented. The bases of these computer programs are the statistical thermodynamic models, which use a predictive algorithm with additional experimental data included to modify or "tune" the mathematical predictions. Most commercially available software uses algorithms developed by D.B. Robinson and Associates (EQUIPHASE), and by Infochem Computer Services Ltd (MULTIFLASH). A detailed discussion of the accuracy of these programs and other hydrate software programs can be found in publications by Sloan (1998) and Carroll (2003).

3.8.1.3 Hydrate Prevention Techniques

The multiphase fluid produced at the wellhead will normally be at a high pressure and a moderate temperature. As the fluid flows through the pipelines, it becomes colder, which means such pipelines could experience hydrates at some point in their operating envelope. For this reason, the hydrate formation in gas transmission pipelines should be prevented effectively and economically to guarantee the pipelines operate normally. Control of hydrates relies upon keeping the system conditions out of the region in which hydrates are stable. It may be possible to keep the fluid warmer than the hydrate formation temperature (with the inclusion of a suitable margin for safety) or operate at a pressure less than the hydrate formation pressure.

Although there are several methods to avoid hydrate formation, depending on the possible location of a gas hydrate plug, some of the techniques to remediate onshore hydrates (e.g., installation of line heaters and line depressurization) may not be practical in long and high pressure, subsea gas transmission pipelines (Son and Wallace, 2000; Wilkens, 2002).

There are few methods of preventing hydrate formation in offshore transmission systems. The permanent solution is removal of water prior to pipeline transportation, using a large offshore dehydration plant that is not often the most cost-effective solution. Pipeline system depressurization is also often used to melt the hydrate plug formed in a pipeline. However, the process is fairly slow and can take up to weeks or even months to completely melt a long hydrate plug. In general, at the wellsite, two methods are applicable, namely, thermal and chemical. These techniques are discussed in the following sections.

3.8.1.3.1 Thermal Methods

Thermal methods use either the conservation or introduction of heat in order to maintain the flowing mixture outside the hydrate formation range. Heat conservation is common practice and is accomplished through insulation[4] (Oram, 1995; Hunt, 1996). When insulation is used, it is easy to keep fluid flowing temperature everywhere along the pipeline above the hydrate formation temperature. But no matter how much insulation is put on the pipeline, after a long pipeline shutdown, the fluid temperature will fall below the hydrate formation temperature and eventually cool down to the ambient temperature. Therefore, thermal insulation itself is not enough for hydrate mitigation for long pipeline shutdowns, so other mitigation strategies will be needed (Guo et al., 2005).

A number of different concepts are available for introducing additional heat to a pipeline. The simplest is an external hot-water jacket, either for a pipe-in-pipe system or for a bundle. Other methods use either conductive or inductive heat tracing (Hunt, 1996). There is concern over the reliability of conductive systems. An electrical-resistance heating system may be desirable for long offset systems, where available insulation is insufficient, or for shut-in conditions (Langner et al., 1999). The ability to heat during production depends on the specific electrical heating implementation. Such systems provide environmentally friendly fluid temperature control without flaring for pipeline depressurization. The effect is also an increase in production as there is no time lost by unnecessary depressurization, pigging, heating-medium circulation, or removal of hydrate blockage. However, still it is difficult to persuade operators to install an acting heating system.

3.8.1.3.2 Chemical Inhibition

An alternative to the thermal processes is chemical inhibition. Chemical inhibitors are injected at the wellhead and prevent the hydrate formation by depressing the hydrate temperature below that of the pipeline operating temperature. Chemical injection systems for subsea lines have a rather high capital expenditure price tag associated with them, in addition to the often high operating cost of chemical treatment (Covington et al., 1999; Sloan, 2000; Frostman et al., 2003). However, hydrate inhibition using chemical inhibitors is still the most widely used method for unprocessed gas streams,

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[4] Burying the pipeline alone is not enough for thermal insulation, and some extra insulation will be needed.

and the development of alternative, cost-effective, and environmentally acceptable hydrate inhibitors is a technological challenge for the gas production industry.

3.8.1.3.2.1 Types of Inhibitors Traditionally, the most common chemical additives used to control hydrates in gas production systems have been methanol, ethylene glycol, or triethylene glycol at a high enough concentration (Sloan, 1991). These chemicals are called thermodynamic inhibitors and have the effect of shifting the hydrate curve to the left, which causes the hydrate stability point to be displaced to a lower temperature and/or a higher pressure. Figure 3-17 shows how the hydrate curve shifts with different amounts of methanol inhibition. Increasing salt content in the produced brine (by injecting electrolyte solutions such as sodium chloride, calcium chloride, and potassium chloride) can also provide some hydrate temperature suppression[5] (see Figure 3-18), but normally this alone is not sufficient to avoid hydrates in the subsea gas production systems. In some cases, blended inhibitors of methanol/glycols and electrolyte(s) are preferred for subsea applications (Zuo and Zhang, 1999).

The inhibitor selection process often involves comparison of many factors, including capital/operating cost, physical properties, safety, corrosion inhibition, and gas dehydration capacity. However, a primary factor in the selection process is whether or not the spent chemical will be recovered, regenerated, and reinjected. Usually, methanol is not regenerated because its

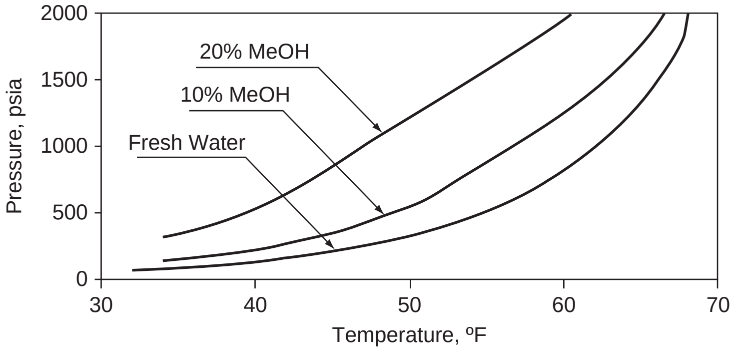

Figure 3-17 Hydrate curves with different amount of methanol inhibition (Guo et al., 2005).

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[5] Even though a salt solution can be used for hydrate inhibition, too much salt can cause salt deposition and scale deposits in the system facilities. Salt solution is also corrosive and can cause corrosion problems to equipment.

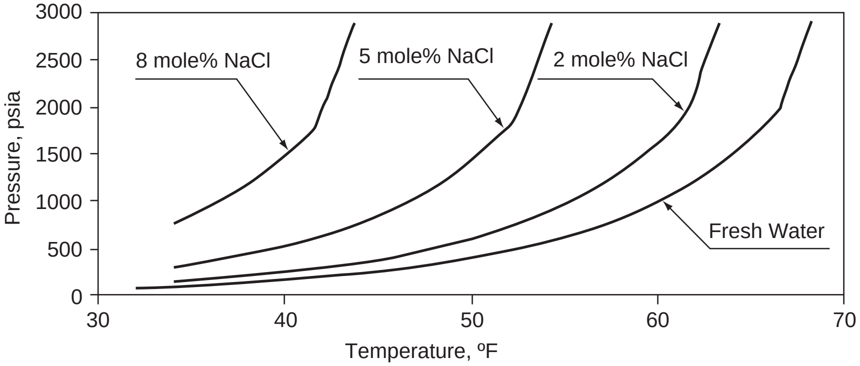

Figure 3-18 Impact of NaCl salt on hydrate formation curve (Guo et al., 2005).

use is intermittent (i.e., during start-up or shutdown). However, when it is injected continuously, as is often observed in gas systems, then it is sometimes regenerated. However, losses to the vapor phase can be prohibitive, in which case operators select monoethylene glycol (GPSA, 2004). Often when this inhibitor is applied, there is a significant expense associated with the cost of "lost" methanol. However, since methanol has lower viscosity and lower surface tension, it makes for an effective separation from the gas phase at cryogenic conditions (below −13°F), and it is usually preferred (Esteban et al., 2000). In many cases, hydrate plug formation is prevented through the addition of glycols (usually, ethylene glycol because of its lower cost, lower viscosity, and lower solubility in liquid hydrocarbons, and lower vapor pressure, giving reduced gas-phase losses) to depress the hydrate formation temperature. In order to be effective, glycols must be added at rates of up to 100% of the weight of water. Since glycols are expensive inhibitors, there is a definite need for extra, costly, and space-consuming onshore or offshore plants for their regeneration. Therefore, it would be useful to develop new hydrate depressants that can be used at much lower concentrations and therefore much lower cost. Recently, two new types of low-dosage inhibitors were developed, which will enable the subsea gas transmission pipelines to handle increased gas volumes without additional glycol injection or extra glycol recovery units (Bloys et al., 1995; Lederhos et al., 1996; Palermo et al., 2000). These new hydrate inhibitors can lead to very substantial cost savings, not only for the reduced cost of the new inhibitors but also in the size of the injection, pumping, and storage facilities (Mehta et al., 2002; Frostman et al., 2003). These new hydrate inhibitors, called low-dosage hydrate inhibitors (LDHIs), form the basis of a technique that does not operate by changing the thermodynamic conditions of the system. In fact, LDHIs act at the early stages of hydrate formation by modifying the rheological properties of the system (Sinquin et al., 2004). There are two types of LDHIs: the kinetic hydrate inhibitors (KHIs), and antiagglomerants (AAs). Most commercial kinetic inhibitors are high molecular weight polymeric chemicals (i.e., poly [N-vinyl pyrrolidone] or poly [vinylmethylacetamide/vinylcaprolactam]), which are effective at concentrations typically 10 to 100 times less than thermodynamic inhibitors concentrations. KHIs may prevent crystal nucleation or growth during a sufficient delay compared to the residence time in the pipeline. The deeper a system operates in the hydrate region, the shorter the time during which kinetic hydrate inhibitors can delay hydrate formation. The achievable delays range between weeks if the pipeline operates less than 42°F in the hydrate region to hours if the pipeline operates at 50°F in the hydrate region (Mokhatab et al., 2007b). Kinetic inhibitors are relatively insensitive to the hydrocarbon phase and may therefore turn out to be applicable to a wide range of hydrocarbon systems. However, the industrial application of kinetic inhibitors depends on the repeatability of multiphase pipeline testing results among laboratory, pilot plant, and field, and the transferability among different plants (Kelland et al., 2000).

In contrast to other types of inhibitors, AAs, which are surface-active chemicals (i.e., alkyl aromatic sulphonates or alkylphenyl ethoxylates), don't prevent the formation of hydrate crystals but keep the particles small and well dispersed so that fluid viscosity remains low, allowing the hydrates to be transported along with the produced fluids. AAs performance is relatively independent of time. In addition, AAs appear to be effective at more extreme conditions than KHIs, which makes these products of interest to operators looking for cost-effective hydrate control in deepwater fields. These additives are currently applied in the Gulf of Mexico, the North Sea, and West Africa (Mehta et al., 2002; Frostman, 2003). However, they mainly have limitations in terms of water cut, where they require a continuous oil phase and therefore are applicable only at lower water cuts. The maximum water cut is expected to be between 40% and 50%. This limitation is caused by the rheological properties of suspensions with high solid fraction and may depend on flow regime conditions (Kelland et al., 1995; Frostman, 2000; Mehta et al., 2002).

As stated earlier, the choice between inhibitor alternatives must be based on physical limitations as well as economics. However, operating conditions may also limit the number of available choices. For example, in a recent project carried out by Baker Petrolite, it was shown that under severe conditions, the required dosage of an antiagglomerator, unlike thermodynamic and kinetic inhibitors, does not increase as the degree of subcooling increases. Therefore, this method of treatment would be a cost-effective solution for the control of gas hydrates (Ramachandran et al. 2000).

3.8.1.3.2.2 Prediction of Inhibitor Requirements The inhibitor must be present in a minimum concentration to avoid hydrate formation. Accurate prediction of this minimum inhibitor concentration is required for cost-effective design and operation of multiphase pipelines. Various empirical methods, charts, and computer programs have been developed for this purpose, including the venerable Hammerschmidt's empirical equation (1934), which is a relatively simple method that has been used to calculate the amount of inhibitor required in the water phase to lower the hydrate formation temperature:

$$\Delta T = \frac{K.W}{100(MW) - (MW)W} \tag{3-58}$$

where ΔT is depression of hydrate formation temperature, °F; MW is molecular weight of inhibitor; W is weight percent of inhibitor in final water phase; and K is constant, depending on the type of inhibitor.

Experimentally determined values of K and molecular weights of inhibitors are given in Table 3-6. To use this equation, one must know the hydrate formation temperature in the gas without the inhibitor being present. In fact, Equation 3-58 only predicts the deviation from the hydrate formation temperature without an inhibitor present.

The Hammerschmidt equation is limited to inhibitor concentrations of about 20–25 wt% for methanol and 60–70 wt% for the glycols (GPSA,

Table 3-6 Physical Constants of Inhibitors (Hammerschmidt, 1934)

Inhibitor	*MW*	*K*
Methanol	32.04	2335
Ethanol	46.07	2335
Isopropanol	60.10	2335
Ethylene glycol	62.07	2200
Propylene glycol	76.10	3590
Diethylene glycol	106.10	4370
Triethylene glycol	150.17	5400

2004). However, for higher methanol concentrations, Nielsen and Bucklin (1983) recommended the following equation:

$$\Delta T = -129.6\ \mathrm{Ln}(X_{\mathrm{H2O}}) \tag{3-59}$$

where X_{H2O} is the mole fraction of water in the aqueous phase.

They claim that this equation is accurate up to 90 wt% methanol, which gives the maximum suppression because methanol freezes at concentrations above 90 wt%. However, recent studies of GPSA (2004) recommend the Nielsen and Bucklin (1983) equation only for methanol concentrations ranging from 25–50 wt%. Equation 3-59 was developed for use with methanol; however, this equation is actually independent of the choice of inhibitor, and therefore, theoretically, it can be used for any glycols.

Maddox et al. (1991) described a graphical procedure for estimating the required inhibitor concentration for both methanol and glycol. This method is a trial-and-error approach that can be used when the activity coefficients of water in methanol and glycol are available. Although this method provides better accuracy, no recommended applicable range is provided (Esteban et al., 2000).

If the produced water or seawater is in contact with the hydrocarbon fluid, the salinity of the water will itself inhibit hydrate formation. Therefore, it is important to be able to estimate the effect of the brine in the produced water on the hydrate formation temperature. For this purpose, McCain (1990) presented the following equation:

$$\Delta T = \mathrm{AS} + \mathrm{BS}^2 + \mathrm{CS}^3 \tag{3-60}$$

where ΔT is temperature depression, °F; S is water salinity, wt%; and

$$\mathrm{A} = 2.20919 - 10.5746\ (\mathrm{SG}) + 12.1601\left(\mathrm{SG}^2\right) \tag{3-60-1}$$

$$\mathrm{B} = -0.10605 + 0.72269\ (\mathrm{SG}) - 0.85093\ \left(\mathrm{SG}^2\right) \tag{3-60-2}$$

$$\mathrm{C} = 0.00347 - 0.01655\ (\mathrm{SG}) + 0.01976\ \left(\mathrm{SG}^2\right) \tag{3-60-3}$$

Equation 3-60 is limited to salt concentrations of less than 20 wt% and for gas-specific gravities (SG) ranging from 0.55 to 0.68.

All of these simple methods predict the depression of the hydrate formation temperature; they do not predict the actual hydrate formation conditions. However, several thermodynamic models have been proposed for predicting the hydrate formation conditions in aqueous solutions containing methanol/glycols and electrolytes (Edmonds et al., 1996; Nasrifar et al., 1998; Zuo and Zhang, 1999). These rigorous models can also account for the effect of pressure and the type of hydrate neglected in the simple mentioned methods. However, available models have limitations that include the types of liquid, compositions of fluids, and inhibitors used.

The amount of inhibitor required to treat the water phase, plus the amount of inhibitor lost to the vapor phase and the amount that is a soluble in the hydrocarbon liquid, equals the total amount required. Methanol vaporization loss can be estimated from Figure 3-19, while glycol vaporization losses are generally very small and typically can be ignored. More details on how to estimate the amount of methanol or glycols lost in vapor and condensate can be found in Sloan (2000).

In addition, prediction of inhibitor losses to the hydrocarbon liquid phase is difficult. However, many of the commercially available software programs include proper calculations to account for the loss of methanol and glycols to the hydrocarbon liquid phase.

3.8.1.3.2.3 Design of Injection Systems Proper design of an inhibitor injection system is a complex task that involves optimum inhibitor selection, determination of the necessary injection rates, pump sizing, and pipeline diameters. Inhibitors for a subsea gas transmission system are selected before

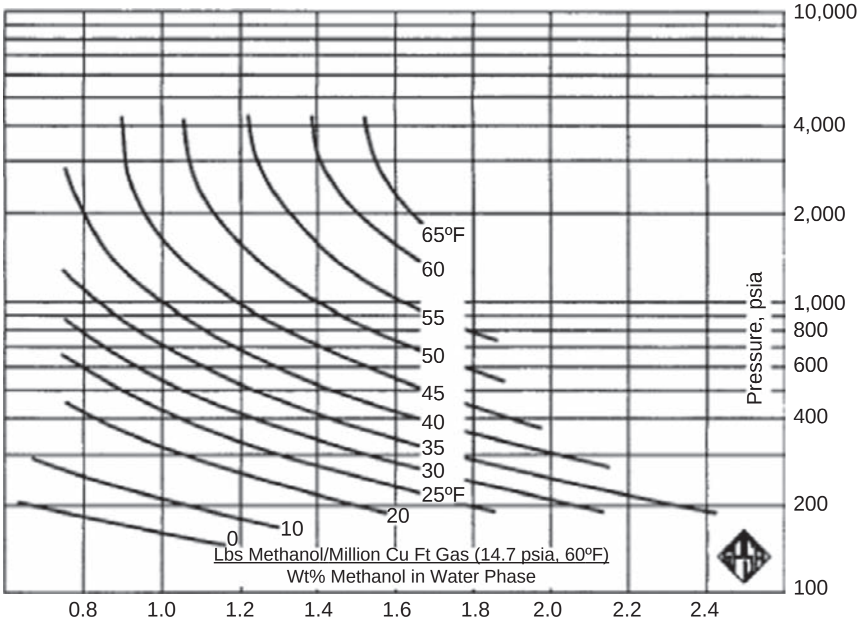

Figure 3-19 Prediction of methanol vaporization loss (GPSA, 2004).

gas production is started on the facility. This makes inhibitor selection difficult as a large number of factors, including brine composition, temperature, and pressure, that affect the performance of inhibitors, are unknown (Ramachandran et al., 2000). Therefore, at this stage, an appropriate multiphase flow simulation package must be used to calculate some of the unknown necessary variables, which are required for injection systems design.

To determine the appropriate injection rate, it is preferable to determine inhibitor requirements by field testing. The inhibitor dosing requirements are later used to determine the requirements for inhibitor storage, pumping capacities, and number of inhibitor lines in order to ensure that the inhibitor can be delivered at the required rates for treating wells and flow lines during start-up and shut-in operations. In addition, points of injection should be chosen to provide maximum benefit in the pipeline system. The most effective position for injection is generally at the center of the pipe in the direction of the product flow. However, the injection rate and location of injection points will be a function of flow geometry, fluid properties, pressure-temperature relationships, etc., that will be encountered in the actual field application. For instance, hydrate inhibitors for a deepwater production system are often pumped through long umbilicals that are ½ inch to ¼ inch in diameter. The injection pump is a positive-displacement metering pump (Kumar, 1987) capable of generating sufficient injection line pressure (normally between 3,000 and 4,000 psi) to overcome the line operating pressure. Ideally, the injection pressure should be 100 psi above line pressure. However, varied injection rates can be achieved by changing the pressure differential. Ramachandran et al. (2000) provide a discussion on proper design of deepwater injection systems and predicting systems behavior.

3.8.2 Corrosion

One of the common problems in the multiphase flow transmission pipelines is metal corrosion. Corrosion is defined as the deterioration of material, usually a metal, because of its reaction with the environment or handling media. This material degradation leads to the impairment of the intended function of the metal, environment, or the integrity of the system. This can be general corrosion (regular loss of metal on the exposed surface), or it can be localized corrosion where only a limited portion of the surface is in contact.

Corrosion can occur in different forms and can be caused by a variety of different reasons. The cause of corrosion can be directly attributed to the impurities found in the produced gas as well as the corrosive components that are by-produced. Because of the nonspecificity of the components produced from a production well, some or all of these components may be active to create a corrosive environment in the pipelines. Corrosion in multiphase systems is a complex phenomenon, including dependency on the partial pressure, temperature, flow regime/velocity, pH, and concentration of corrosive components. Consequently, corrosion prediction requires substantial understanding of the simultaneous interaction of many process variables that govern both the flow and corrosion conditions.

An important aspect of maintaining pipeline performance is adequate control of corrosion,[6] both internally—caused by the flow components and their byproducts—and externally—because of pipeline exposure to the soil and water (Hartt and Chu, 2004). Pipeline corrosion can be inhibited by several means:

- Choice of corrosion-resistant metals
- Injection of corrosion inhibitors
- Cathodic protection
- External and/or internal protective coatings

While corrosion control can be achieved by selection of an appropriate corrosion-resistant metal, operating considerations usually dictate that a high-efficiency corrosion control system be used. Protecting pipelines from corrosion is achieved internally by injection of inhibitors to mitigate internal corrosion and externally by use of cathodic protection and/or combination of coatings and cathodic protection (for buried or subsea pipelines).

3.8.2.1 Choice of Corrosion-Resistant Metals

Corrosion resistance is a basic property related to the ease with which materials react with a given environment. All metals have a tendency to return back to stable conditions. This tendency causes metals to be classified according to rising nobleness that again leads to classification of decreasing activity and increasing potential. When it is observed that the existing material of construction is prone to corrosive attack, generally a decision is made to change the materials of construction and select alternate material to suit the specific need. For specific recommendations concerning materials selection, refer to the ANSI B31.3 and B31.8, API RP 14E (1991), and NACE (1975) MR-01-75. It should be noted that no single material can serve as a cure for all the corrosion evils. Therefore, a detailed study of process and operating conditions has to be carried out before selection of a new material.

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[6] Corrosion can result in the loss of millions of dollars if a pipeline is not properly protected.

Corrosion resistance is not the only property to be considered in the material selection process, and the final selection will generally be the result of several compromises between corrosion resistance and economic factors. Historically, carbon steel has been the most economical material for construction of long-distance, large-diameter gas transmission pipelines.[7] However, the inherent lack of corrosion resistance of these materials in subsea application requires a corrosion control system with a high degree of reliability (Hartt and Chu, 2004).

3.8.2.2 Corrosion Inhibitors

Sometimes it is necessary to reduce the intensity of corrosive attack of the environment by adding certain chemicals known as "inhibitors" to reduce the aggressiveness of the media. Corrosion inhibition is necessary in gas/condensate pipelines when the required corrosion allowance for an uninhibited pipeline is excessive. If this is the case, a corrosion allowance for the inhibited pipeline system can be calculated from the allowance for the uninhibited system and the inhibitor efficiency.[8] If the required corrosion allowance is still excessive, a higher grade material must be used for the pipeline.

Corrosion inhibitors are cationic surfactant chemicals such as filming amines, which, when added in a small concentration (between 10 and 1,000 ppm), effectively reduce the corrosion rate of a metal.[9] The injected inhibitor forms a protective layer on the internal surface of the pipe, and the continuity and the quality of this layer are the controlling factor for effective corrosion protection. Corrosion inhibitors can be applied in batches, resulting in a protective film that may last for weeks or months. Enough inhibitor should be introduced to provide an appropriate wall coating thickness. Frequency of the treatment is governed by the remaining effectiveness of the inhibitor after a specified amount of commodity has been moved through the line. Inhibitors are often continuously injected in low concentrations where a thin film is laid down and maintained over time (Tiratsoo, 1992).

Corrosion inhibitors will only perform at their nominal efficiency if they remain in contact with the pipe wall at the correct concentration (Erickson et al., 1993). However, for some specific flow patterns in multiphase flows,

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[7] Corrosion-resistant alloy (CRA) is often used to replace carbon steel for corrosive applications. But CRAs are normally more expensive than carbon steel, and their use should depend upon overall economics.

[8] For smooth pipeline, the efficiency of corrosion inhibitors can be as high as 85%–95% but can drop if the shear stress increases drastically at locations such as fittings, valves, and bends (Guo et al., 2005).

[9] In gas/condensate pipeline, adding the thermodynamic hydrate inhibitors can also reduce the corrosion rate. The reason is that they absorb free water and make the water phase less corrosive (Strommen, 2002).

such as slug flow, inhibitor stripping can occur due to turbulence and high shear stress (Kang et al., 1996). But the slug can also help by distributing corrosion inhibitors and preventing top-of-line corrosion. As gas/condensate pipelines operate primarily in stratified flow regime, corrosion inhibitor is deposited on the pipe's top wall only if there is a substantial fraction of the liquid entrained as droplets in the gas phase. However, at some locations, corrosion will still occur due to the condensation of water that dilutes the inhibitor concentration.

There are numerous corrosion inhibitor types and compositions. Corrosion inhibitor selection is a complex process that needs to balance a number of competing demands. The inhibitor must both reduce corrosion rates to an acceptable level and be compatible with system materials and other production chemicals such as hydrate inhibitors and biocides. The selected corrosion inhibitor must also not adversely affect hydrocarbon production processes by formation of foams or emulsions. Finally, the selected chemical must meet environmental regulations.

Although inhibitors can be used to great advantage to suppress the corrosion of metals in many environments, there are certain limitations of this type of corrosion prevention that should be recognized. First, it may not be possible to add inhibitors to all corrosive systems because they may contaminate the environment. Further, inhibitors generally lose their effectiveness rapidly as the concentration and temperature of the environment increase. Finally, many inhibitors are toxic, and their application is limited. Therefore, new inhibitors are continuously being developed to handle more aggressive conditions and comply with more stringent environmental regulations.

3.8.2.3 Cathodic Protection

Cathodic protection is the most successful method for reducing or eliminating corrosion for buried or submerged metallic structures that involves using electric voltage to prevent corrosion (Colson and Moriber, 1997). When two metals are electrically connected to each other in an electrolyte (e.g., seawater), electrons will flow from the more active metal (anode) to the other (cathode) due to the difference in the electrochemical potential. The anode supplies current, and it will gradually dissolve into ions in the electrolyte and at same time produce electrons that the cathode will receive through the metallic connection with the anode. The result is that the cathode will be negatively polarized and hence be protected against corrosion. The two methods of achieving cathodic protection are (1) the use of sacrificial or reactive anodes with a corrosion potential lower than the metal to be protected, and (2) applying a direct current (Uhlig and Revie, 1985). The use of direct current systems is less costly than sacrificial anodes and provides a higher range of possible potential differences (Copp, 1970), although they may require greater maintenance during the lifetime of the operation.

3.8.2.4 *Protective Coatings*

While cathodic protection has historically been employed as the sole corrosion control methodology for subsea gas production systems, the nature of multiphase pipelines is such that the combined use of protective coatings with cathodic protection is necessary to achieve the effective protection[10] (Samant, 2003). In fact, protective coatings help control pipeline corrosion by providing a barrier against reactants such as oxygen and water. However, because all organic coatings are semipermeable to oxygen and water, coatings alone cannot prevent corrosion, so a combination of cathodic protection is often used (Varughese, 1993).

Pipelines are often internally coated to minimize corrosion when lying in pipeline dumps prior to construction, and to provide a smooth surface that reduces friction when fluids are in transit. It should be noted that the initial period in a production well's life can be its most corrosive time due to the high partial pressure of the carbon dioxide. Therefore, it may be more economical to protect the multiphase pipelines from a young field. The coating must be compatible with the commodity and should have suitable resistance to attack by the commodity as well as any contaminants, corrosives, or inhibitors associated with it. Coatings such as epoxies, plastics, or metallic compounds can be used for selected applications. The oldest and still-used approach consists of hot-applied bituminous material wrapped with an appropriate covering (Polignano, 1982). External coatings are used to reduce the value of the external current to an economic level by imposing a barrier of high electrical resistance between metal and its environment (Copp, 1970). Rhodes (1982) reported the use of many such coatings, including polyurethane, phenolic resin, phenolic epoxy, and fusion-bonded epoxy powder.

3.8.2.5 *Corrosion Monitoring*

Pipelines are susceptible to both internal and external corrosion. Therefore, both internal as well as external monitoring of pipelines is required for a complete assessment, thus providing the direction that will ensure proper

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[10] For some installations (e.g., deepwater), one might choose continuous inhibition over protective coating due to the implications of a coating failure.

utilization of materials and corrosion control methodologies. While traditional inspection and monitoring techniques (such as pigging and corrosion coupons) may represent effective solutions for assessment of the condition and integrity of a pipeline, the sensitivity and accuracy of these methods may be inadequate for monitoring inhibitor performance since they do not provide real-time information but rather focus on historical analysis. Therefore, traditional techniques of monitoring must be supplemented with online techniques that provide more timely real-time data. This allows operators to be more proactive in preventing corrosion-related pipeline failures.

3.8.3 Wax

Multiphase flow can be severely affected by deposition of organic solids, usually in the form of wax crystals, and their potential to disrupt production due to deposition in the production/transmission systems. The wax crystals reduce the effective cross-sectional area of the pipe and increase the pipeline roughness, which results in an increase in pressure drop. The deposits also cause subsurface and surface equipment plugging and malfunction, especially when oil mixtures are transported across Arctic regions or through cold oceans. Wax deposition leads to more frequent and risky pigging requirements in pipelines. If the wax deposits get too thick, they often reduce the capacity of the pipeline and cause the pigs to get stuck. Wax deposition in well tubings and process equipment may lead to more frequent shutdowns and operational problems.

3.8.3.1 Wax Deposition

Precipitation of wax from petroleum fluids is considered to be a thermodynamic molecular saturation phenomenon. Paraffin wax molecules are initially dissolved in a chaotic molecular state in the fluid. At some thermodynamic state, the fluid becomes saturated with the wax molecules, which then begin to precipitate. This thermodynamic state is called the onset of wax precipitation or solidification. It is analogous to the usual dew point or condensation phenomenon, except that in wax precipitation a solid is precipitating from a liquid, whereas in condensation a liquid is precipitating from a vapor. In wax precipitation, resin and asphaltene micelles behave like heavy molecules. When their kinetic energy is sufficiently reduced due to cooling, they precipitate out of solution, but they are not destroyed. If kinetic energy in the form of heat is supplied to the system, these micelles will desegregate and go back into stable suspension and Brownian motion.

3.8.3.1.1 Wax Deposition Envelope

Many reservoir fluids at some frequently encountered field conditions precipitate field waxes. It is very important to differentiate field waxes from paraffin waxes. Field waxes usually consist of a mixture of heavy hydrocarbons such as asphaltenes, resins, paraffins (or paraffin waxes), cyclo-paraffins, and heavy aromatics. Wax precipitation primarily depends on fluid temperature and composition, and it is dominated by the van der Waals or London dispersion type of molecular interactions. Pressure has a smaller effect on wax precipitation. As with asphaltenes, the fact that waxes precipitate at some and not at other thermodynamic states, for a given fluid, indicates that there is a portion of the thermodynamic space that is enclosed by some boundary within which waxes precipitate. This bounded thermodynamic space has been given the name wax deposition envelope, or WDE (Leontaritis, 1995). A typical WDE is shown in Figure 3-20.

The upper WDE boundary can have either a positive slope or a negative slope. In most cases where the WDE was obtained experimentally, the upper WDE boundary was very close to a vertical line.

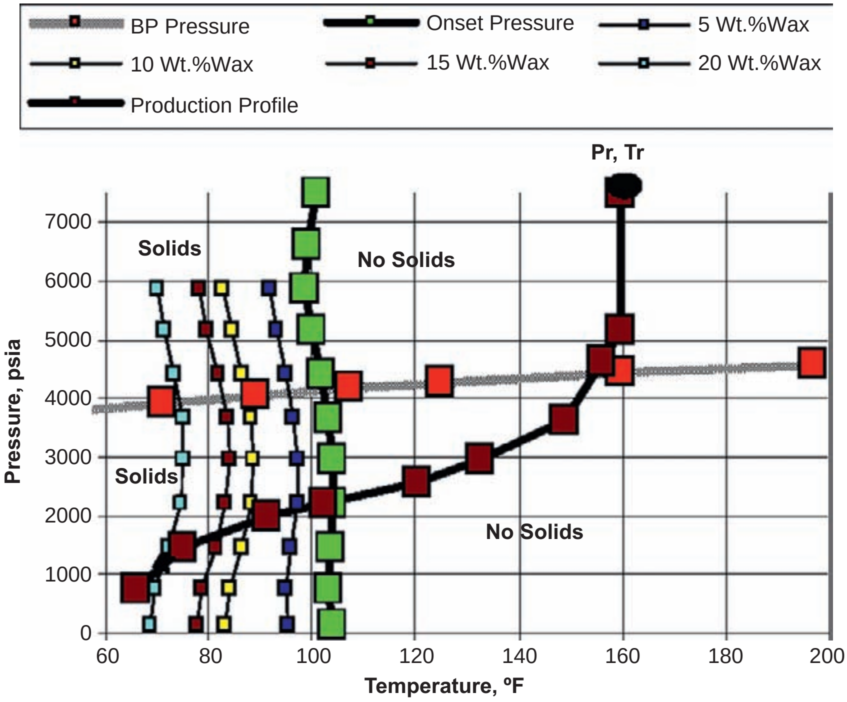

Figure 3-20 Typical wax deposition envelope (Leontaritis, 1995).

The intersection of the WDE boundary with the bubble-point line is generally expected to be to the left of the onset of wax crystallization (cloud point) of the stock tank oil. This has been the case with many reservoir fluids whose WDE has been measured (Leontaritis, 1995). This is due to the fact that light ends, when pressured into oil, usually cause a suppression of the onset of wax crystallization temperature. The actual shape of the lower WDE boundary is primarily a function of the compositions of the intermediates and light ends of the reservoir fluid. The technology for measuring WDEs only recently has been perfected. Because the technology is new and the cost of measurements is relatively high, most companies do not find it economical to obtain the complete WDE of their reservoir fluid through laboratory measurements. Hence, the preferred method is obtaining only a few experimental data points and using them to fine-tune phase behavior models that are then used to calculate the remaining WDE more economically. It should be noted that most of these models have very little predictive capacity at this time, and they are primarily used as combination correlational-predictive type of tools. One of the main reasons for the lack of pure predictive capability is improper or inadequate crude oil characterization (Leontaritis, 1997a).

3.8.3.1.2 Gas/Condensate Wax Deposition Envelope

Some gas/condensates, especially rich gas/condensates with yields in excess of 50 bbls/MMscf, are known to contain high carbon number paraffins that sometimes crystallize and deposit in the production facilities. The obvious question is what is the shape of the thermodynamic envelope (i.e., P and T surface) of these gas/condensates within which waxes crystallize? Or, in order to maintain the previous terminology, what is the WDE of gas/condensates typically?

The shapes of the WDEs of two gas/condensates in the Gulf of Mexico are presented here. The shapes of the previously mentioned WDEs indicate potential wax deposition in those cases where the gas/condensate contains very high carbon number paraffins that precipitate in solid state at reservoir temperature. In other words, the temperature of the reservoir may not be high enough to keep the precipitating waxes in liquid state. Hence, the gas/condensate, which is a supercritical fluid, enters the WDE at the "dew point" pressure. This casts new insight into the conventional explanation that the productivity loss in gas/condensate reservoirs, when the pressure near the wellbore reaches the dew point, is only due to relative permeability effects.

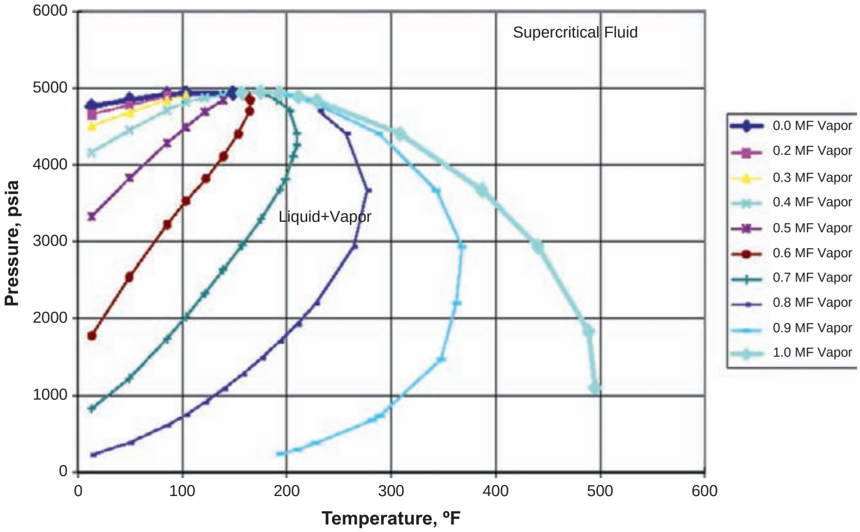

Figure 3-21 Vapor-liquid envelope, Gulf of Mexico gas/condensate "A" (Leontaritis, 1998).

Figure 3-21 shows the vapor-liquid envelope (V-L envelope) of what one might call a typical Gulf of Mexico gas/condensate. This gas/condensate (called gas/condensate "A" for our purposes here) was analyzed with paraffin-aromatic-resin-asphaltene (PARA) analysis (Leontaritis, 1997a) and found to contain normal paraffins with carbon numbers exceeding 45. The V-L envelope was simulated using the Peng and Robinson (1976) original equation of state (EOS) that had been fine-tuned to PVT data obtained in a standard gas/condensate PVT study. The first question that was addressed in a wax study involving this fluid was what happens as the fluid is cooled at some constant supercritical pressure? What actually happened is shown in Figure 3-22.

Figure 3-22 shows several onset of wax crystallization data points obtained with the near-infra-red (NIR) equipment (Leontaritis, 1997b) by cooling the gas/condensate "A" at different constant pressures. It was evident from the NIR data that there was a thermodynamic envelope, similar to the one defined and obtained experimentally for oils, to the left of which (i.e., at lower temperatures) wax crystallization occurred. The complete wax deposition envelope shown in Figure 3-22 was calculated with a previously tuned wax phase behavior model (Narayanan et al., 1993). Despite the clarity of the WDE obtained for gas/condensate "A" as shown in Figure 3-22, more data were needed to confirm the presence

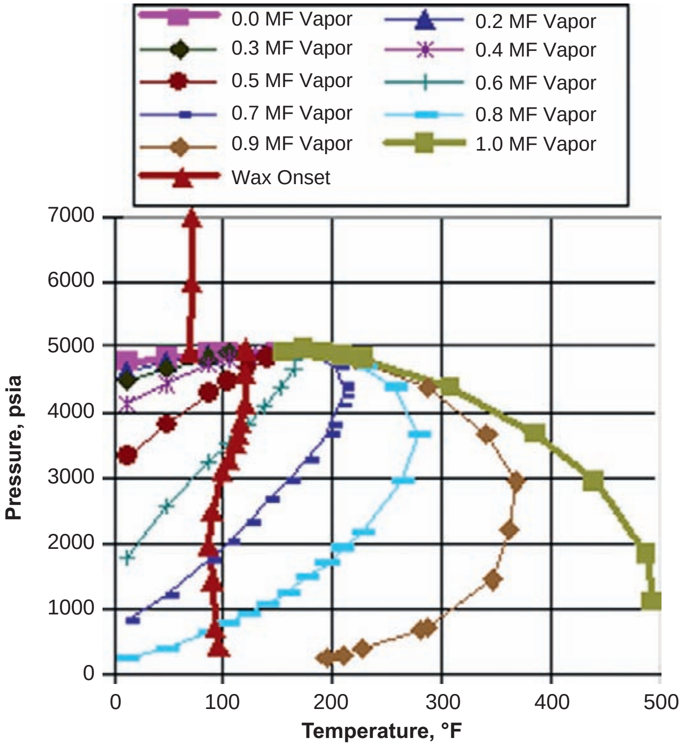

Figure 3-22 Wax deposition envelope, Gulf of Mexico gas/condensate "A" (Leontaritis, 1998).

of WDE in other condensates and establish its existence as a standard thermodynamic diagram.

Figure 3-23 shows the V-L envelope of another typical Gulf of Mexico condensate. This condensate (called gas/condensate "B" for our purposes here) also contains paraffins with carbon numbers exceeding 45, although the data show that gas/condensate "B" is lighter than gas/condensate "A." The V-L envelope was again simulated using the Peng and Robinson (1976) original EOS after it had been tuned to PVT data obtained in a standard gas/condensate PVT study.

Figure 3-24 shows the NIR onset data superimposed on the V-L envelope. It is evident again from the NIR data that there is a thermodynamic envelope to the left of which (i.e., at lower temperatures) wax crystallization occurs. Once again, the complete wax deposition envelope shown in Figure 3-24 was calculated with a previously tuned wax phase behavior model (Narayanan et al., 1993).

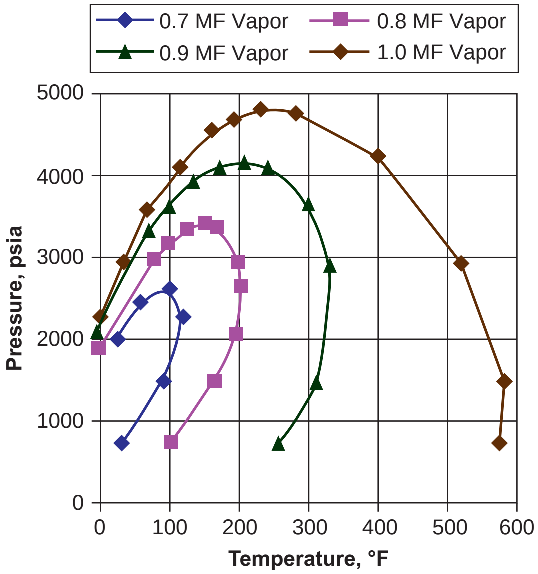

Figure 3-23 Vapor-liquid envelope, Gulf of Mexico gas/condensate "B" (Leontaritis, 1998).

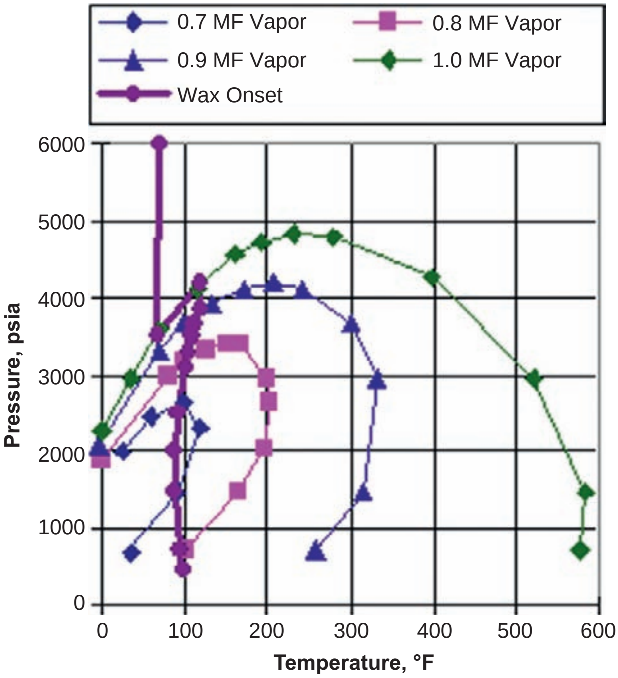

Figure 3-24 Wax deposition envelope, Gulf of Mexico gas/condensate "B" (Leontaritis, 1998).

Data presented here confirm the presence of a WDE in gas/condensates that contain high carbon number paraffin waxes (≥ 45). This WDE is similar to oil WDEs, and as a result, it should be considered a standard thermodynamic diagram. The shape of the WDE inside the V-L envelope seems to be consistent with existing information regarding the effect of light hydrocarbons on the onset of wax crystallization or wax appearance temperature. That is, as the pressure rises, the WDE tilts to the left (negative slope) due to the ability of light hydrocarbons to depress wax crystallization. However, at the pressure where retrograde condensation begins, the WDE turns forward, thus acquiring a positive slope. The reason is that the light ends begin to vaporize, and the waxes remaining in the liquid phase begin to concentrate. This is simply caused by the change in normal paraffin concentration, which, in turn, is caused by retrograde concentration. In most condensates, the V-L envelope is fairly horizontal at the saturation line (dew point or bubble point). Hence, when this general pressure is reached, the WDE seems to coincide with the V-L saturation line until the temperature becomes low enough for the waxes to begin crystallizing from the supercritical condensate. This is in agreement with prior observations that indicate a substantial increase in the solvent power of some fluids when they become supercritical (i.e., propane, CO_2, etc.). That is, supercritical hydrocarbon fluids are expected to require cooling to much lower temperatures before paraffin waxes begin to crystallize because of their increased solvent power.

3.8.3.2 Wax Formation in Multiphase Gas/Condensate Pipelines

When the production pressure and temperature profile of a gas/condensate crosses the WDE and the hydrate envelope (HE), waxes and hydrates may form. The formed hydrates can grow in mass and yield strength to the point that they restrict and finally stop the flow. If the waxes form while the liquid is in contact with the wall, some of them should attach to the wall, thus resulting in wax deposition. The P-T diagram of a Gulf of Mexico gas/condensate as shown in Figure 3-25 demonstrates the aforementioned situation.

The diagram indicates that there was not any live NIR done with this fluid. The only point on the WDE that was measured was the cloud point of the condensate, shown in the diagram at 95°F at the bottom of the orange vertical line. The clean tubing and flow line of this system would deliver about 14.5 MMscfd on 6/17/2002. From the light blue P and T profile line and dark blue hydrate line, it was very clear at the outset that the main flow assurance issues in this system would be hydrate and wax formation and

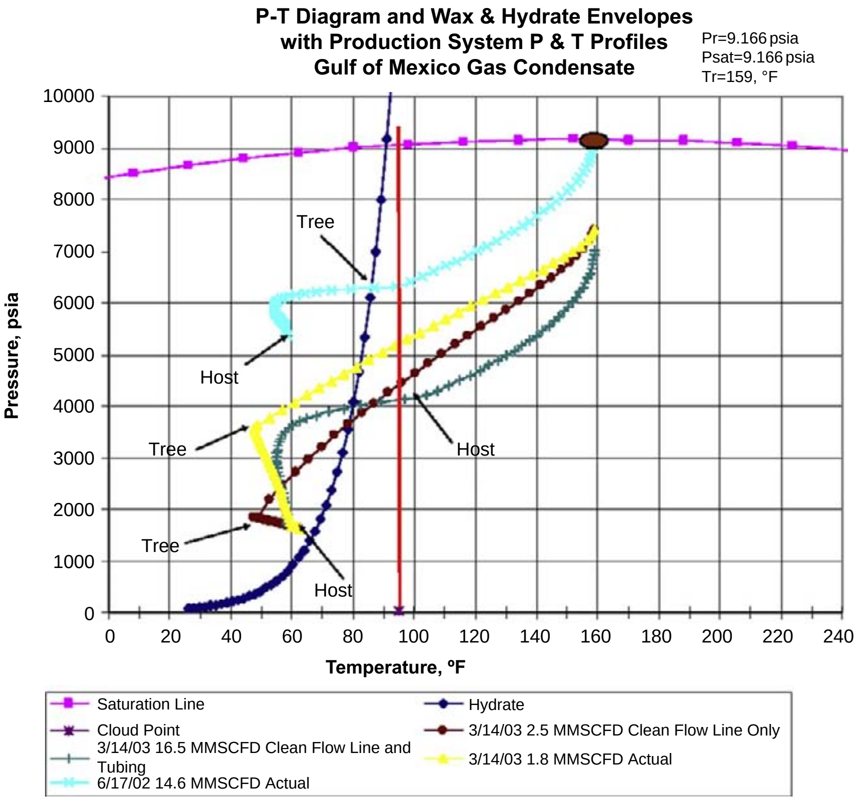

Figure 3-25 Gulf of Mexico gas/condensate wax and hydrate deposition envelopes (AsphWax, Inc., 2003).

deposition. Hence, because of this important illustrative information, the effort to obtain the aforementioned diagram is obviously very worthwhile.

Sometimes oil and gas operators decide to design the facilities such that their operation is to the left of the WDE and HE (such as the example in Figure 3-25). In these situations, wax and hydrate formation takes place if the fluids are left untreated with chemicals. In the previous example, methanol was injected to inhibit successfully the formation of hydrates formed due to production of reservoir equilibrium water. However, because the operator was not aware of the wax phase behavior of this condensate at start-up, no wax chemical was injected. Wax deposition was severe enough to cause the production rate to drop down to 1.8 MMscfd on 3/14/2003. The estimated via simulation maximum production rate on 3/14/2003 was 16.5 MMscfd. The data in the plot show that the maximum friction loss was in the upper part of the tubing. Indeed, the light P and T line with triangles shows that the wax deposition and hydrate envelopes were being crossed in the tubing at that time.

3.8.3.2.1 Identification of Wax Deposition Problems

A rather simple chart that allows daily monitoring of wax deposition problems in multiphase gas/condensate pipelines is shown in Figure 3-26.

This performance chart is for the gas/condensate shown in Figure 3-25. The chart was made by plotting the performance index (PI) versus time. The PI is calculated from the following equation:

$$\mathrm{PI} = \frac{\Delta P^2}{Q^{1/0.54}} \tag{3-61}$$

where ΔP is the pressure drop in the pipeline, psi; and Q is the flow rate, MMscfd.

The performance index (PI) should remain as a horizontal straight line during production if there is no restriction formed in the line. It is evident in Figure 3-26 that around 10/21/2002 a restriction was being formed in

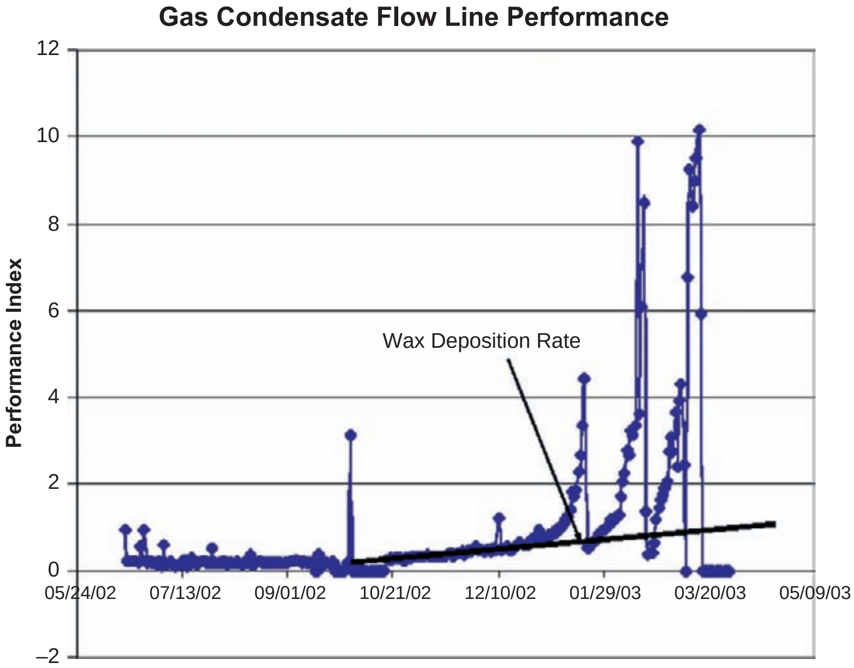

Figure 3-26 Gas/condensate flow line performance index (AsphWax, Inc., 2003).

the line. At the beginning of January, there was a very sharp loss of hydraulic capacity. This loss cannot be due to wax deposition because the rate of wax deposition does not change so sharply in a produced fluid. Indeed, the sharp rise in the PI was caused by hydrate formation that occurred on top of wax crystal formation and deposition. Several sharp rises in PI followed, caused by hydrates, until it was determined that the well started producing free water and required a much higher methanol injection rate. While all of this was happening with the advent of water and hydrate formation, the wax deposition rate was increasing steadily, as shown on the chart.

It should be noted that the engineer needs to be very careful in attributing a rise in the PI only to wax and hydrates. One needs to make sure that other culprits such as fines, salts, returning drilling, and completion fluids, are not present.

3.8.3.2.2 Wax Deposition Inhibition/Prevention

It is not always clear to the engineer what would be the wax deposition mechanism and what type of chemical treatment would be the most effective in gas/condensate situations. In oil flow lines, wax deposition occurs by diffusion of wax molecules and crystals toward and attachment at the wall. Prevention or inhibition of wax deposition is mainly accomplished by injection of a special class of molecules that interact with paraffin molecules at temperatures above the cloud point and influence their crystallization process in a way that diminishes the attraction of the formed crystals toward the wall. The inhibited formed wax crystals are removed from the system by the shear forces caused by the flowing oil.

The aforementioned mechanism occurs when the wax crystal forms at the wall. This is the case with most liquid-filled bare (uninsulated) lines. The two important requirements are fast heat transfer and fast diffusion of paraffin molecules toward the wall. This is not the case in many situations of flow lines carrying gas/condensates. There is fast cooling, but the majority of the gas cools down while in the main flow, thus forming wax crystals with the main flow that would have a tendency to deposit or sediment by gravity in the liquid holdup. Also, the liquid that forms near the inlet of the bare line that has not reached its cloud point cools with a velocity that is almost standstill compared to the main flow. Slugging in this case would be beneficial in removing wax slush from the line. Additionally, these flow lines are essentially soaked in methanol to prevent hydrate formation. It just so happens that most (if not all) chemicals with wax crystal modification properties, as discussed later, are incompatible with alcohols and glycols. Hence, the inhibited wax crystals accumulate in the flow line in the liquid holdup, thus adding to the wax accumulation. This accumulation forms what has been called wax slush. This wax slush is viscous, but it moves and can flow given enough shear force.

The preceding theory has been documented several times from field data. In the year 2000, enough production and laboratory data and simulations were available to make the case for one field, but there was no direct evidence of wax slush from the field. During a shutdown caused by a hurricane, the flow line was depressurized at the tree, in an effort to dissolve any existing hydrates. At start-up, the well was started very fast to try to initiate movement and flow of any liquids (via an induced slug) and the previously theorized existing "wax slush." The fast flow did mobilize a phase that was loaded with wax. The wax slush flowed into the separators and plugged lines and equipment, making them temporarily incapable of operation.

It is appropriate at this time to give a brief description of the main two chemical classes used to treat wax deposition. There are three types of wax crystals: plate crystals, needle crystals, and mal or amorphous crystals.

Paraffinic oils form plate or needle crystals. Asphaltenic oils form primarily mal or amorphous crystals. Asphaltenes act as nucleation sites for wax crystal growth into mal crystals.

Plate crystals look, as their name implies, like plates under the microscope. Needle crystals look like needles, and mal crystals are amorphous and generally like small round spheres. The interaction between crystals and the pipe wall increases from mal to needle to plates. Thus, maintaining newly formed wax crystals small and round, i.e., like mal crystals, is desirable.

The behavior and properties, e.g., cloud point and pour point, of paraffin crystals precipitating from a hydrocarbon can be affected in three ways:

1. Crystal size modification; that is, modification of the crystal from larger sizes to smaller sizes.
2. Nucleation inhibition; that is, inhibition of the growth rate of the crystal and its ultimate size.
3. Crystal type or structure modification; that is, modification of the crystal from one type to the other—for instance, modifying a crystal from needle to mal type.

A wax crystal modifier works primarily to modify the crystal size. The plate crystals of an n-paraffin look much smaller under the microscope when precipitating from hydrocarbons inhibited with a wax crystal modifier. Smaller crystals have lower molecular weights and thus higher solubility in oil. Furthermore, smaller particles have smaller energy of interaction among themselves and the pipe wall. A crystal modifier interrupts the normal crystal growth of the n-paraffins by inserting itself in the crystal and thus stopping its growth.

It is noted that another name for a wax crystal modifier is pour point depressant, or PPD. As the name implies, wax crystal modifiers are very effective at suppressing the pour points of crude oils because they suppress the wax crystal growth and thus minimize the strength of their interactions.

A wax dispersant may act to inhibit wax nucleation and change the type of wax crystal from plate or needle to mal or amorphous. Inhibited amorphous crystals are smaller and carried much easier by the hydrodynamic drag of the flowing fluid. The presence of asphaltenes and resins facilitates the effect of the dispersant. The dispersant interacts with the asphaltenes and resins and ties them up, thus removing nucleation sites required by wax crystals to grow. A dispersant, when added to the oil at a temperature above the cloud point, has occasionally an additional benefit of suppressing the cloud point by interacting with and tying up the asphaltenes and resins that tend to come out first. A dispersant tends usually to disperse the wax particles at the water oil interface.

Both wax crystal modifiers and wax dispersants are useful chemicals that have the ability to diminish wax formation and deposition, although through different mechanisms. Wax dispersants are usually much smaller in molecular weight and size than wax crystal modifiers. Hence, their viscosity and flow properties in general are more favorable in cold applications. Also, some wax dispersants are soluble with alcohols and glycols, thus making them compatible for simultaneous injection. The selection of the wax chemical should be made after careful consideration of the produced hydrocarbon and facilities.

3.8.3.2.3 Wax Deposit Remediation

On occasion, if a substantial amount of wax accumulates in the line, as evidenced by the PI chart such as the one shown in Figure 3-26, a temporary shutdown to do a chemical soak, with a potential modification to the chemical to give it more penetrating power at cold temperatures, and fast start-up might be necessary to cause the wax slush cough.

Pigging is an option, but only after very careful consideration of the system's performance to understand the dynamics of the moving pig and continuous removal of the wax cuttings ahead of it. This is a very difficult job. Controlling the bypass of flow around the pig for such a purpose is difficult. Hence, many pigging operations end up in failure with stuck pigs.

The pigging analysis and decision must be left to true experts. Starting a pigging program at the beginning of the life of the system has a better chance of success than at any other time. Even then, excellent monitoring of the system's PI is a must. It is recommended that a short shutdown, chemical soak, and fast start-up be considered first, because this is the safest option.

3.8.3.2.4 Controlled Production of Wax Deposits

From a technical standpoint, spending enough capital initially to design the facilities to operate outside the wax and hydrate forming conditions or to the right of the WDE and HE is obviously the best solution. However, very often in practice, controlling wax deposits during production is preferred because of the lower facilities cost. Economically marginal fields can only be produced under this scenario. Hence, in these cases, the following three options prevail:

1. Pigging only
2. Chemical injection only
3. Combination of pigging and chemical injection

While frequent pigging of the line clears up any wax deposits, it may be necessary to chemically inhibit wax deposition during times when pigging is unavailable. Pigging often is inadequate or uneconomical, unless used in conjunction with a chemical treatment program. This program is often performed into two stages: (1) removal of wax deposits in the production/transmission lines, and (2) continuous chemical injection or periodic treatment (such as batch treatments) in order to ensure pipeline integrity.

Chemical injection is the safest of the three, if good technical support and testing are available. The approach is discussed as follows:

1. Inject a strong chemical dispersant/inhibitor down the hole to keep the formed wax particles small and suspended in the flow line. The majority of them would be carried away with the gas/liquid flow.
2. The chemical dispersant/inhibitor must be compatible and soluble with methanol to prevent precipitation of the chemical itself in the line.
3. Monitor the hydraulics in the flow line and occasionally, if necessary, cause a "slug" or "cough" to cough-up any accumulated wax.
4. On occasion, if a substantial amount of wax accumulates in the line, a temporary shutdown to do a chemical soak, with a modification to the chemical to give it more penetrating power at cold temperatures, and fast start-up, might be necessary to cause the wax slush cough.

When properly implemented, the preceding approach should provide the lowest-cost wax deposition control in gas/condensate transmission lines.

3.8.4 Slugging

Slugging, which refers to varying or irregular flows of gas and liquids (or liquid surges) in pipelines, is a major flow assurance challenge. Slug flow can pose serious problems to the designer and operator of two-phase flow systems. Large and fluctuating rates of gas and liquid can induce severe mechanical vibrations in the pipe and severely reduce the production and, in the worst case, shut down or damage downstream equipment like separator vessels and compressors. As a result, prediction of slug characteristics is essential for the optimal, efficient, safe, and economically feasible design and operation of two-phase gas-liquid slug flow systems.

Slugging can take several forms (hydrodynamic slugging, terrain/riser induced slugging, and operationally induced slugging) that will be discussed here.

3.8.4.1 Hydrodynamic Slugging

Hydrodynamic slugs are formed by the growing of waves on the gas-liquid interface in the stratified flow under certain flowing conditions, where the amplitude of these waves becomes sufficiently large and can bridge the pipe. Hydrodynamic (normal) slugs tend to be longer, the larger the pipe is and can, unfortunately, be thousands of feet long. Brill et al. (1981) noted from field tests at Prudhoe Bay that not all slugs are of the same length and that a log-normal distribution was observed.

When hydrodynamic slugs are numerous, the pipeline is said to be operating in the slug flow regime. Hydrodynamic slugs travel through a pipeline and can cause significant damage to, or operational problems for, downstream equipment. A usual design practice is to find the volume of the large slugs and design a separator or slug catcher able to handle this. Two main methods are typically used for slug size prediction: the simple correlation of field data and transient one-dimensional slug-tracking simulators, which use empirical correlations to compensate for the fact that their one-dimensional models cannot model these mechanisms directly. The simple correlations were developed by correlating field slugging results to key parameters such as flow line diameter and fluid physical properties. However, these correlations ignore the important effect of line topography and therefore give misleading results. Slug-tracking simulators give reasonable predictions when they are tuned to field data. However, their ability to make *a priori* predictions of slug flow (including slug size and frequency) remains limited. Therefore, good design practice still requires the application of healthy design margins (FEESA, 2003).

3.8.4.2 *Terrain-Induced Slugging*

In pipes across undulating terrain, a major cause of slugging is the topography. Liquid tends to build up and sit at the lowest points of the pipeline, until it is forced onward through the rest of the pipe by the pressure of the gas caught behind. Terrain slugging is most likely to occur at low flow rates, with a low pipeline pressure. When terrain-induced slugging is predicted, slugs will only form in upwardly inclined sections of flow line. These slugs are unlikely to persist throughout the length of the line. Instead, they will steadily decay and then collapse in horizontal or downwardly inclined sections (Bendiksen et al., 1986).

Terrain-induced slugging is highly undesirable due to the long-duration instabilities and the related oscillating momentum that can damage process equipment and necessitate major slug catchers. A mechanistic model or a transient program can provide a more realistic estimate of the terrain slug characteristics (McGuinness and McKibbin, 2002). During production, the length and velocity of a slug can be also estimated by monitoring the pressure variation at the outlet of the pipeline.

3.8.4.3 *Riser-Induced (Severe) Slugging*

Terrain-induced slugging at a pipeline-riser system is denoted by severe slugging. For this system, liquid will accumulate in the riser and the pipeline, blocking the flow passage for gas flow. This results in a compression and pressure buildup in the gas phase that will eventually push the liquid slug up the riser, and a large liquid volume will be produced into the separator that might cause possible overflow and shutdown of the separator. The severe slugging phenomenon is very undesirable due to pressure and flow-rate fluctuations, resulting in unwanted flaring and reducing the operating capacity of the separation and compression units (Schmidt et al., 1980). Figures 3-27 and 3-28 show sample time traces for the riser outlet liquid and gas flow rates during severe slugging, respectively. These figures show the large surges in liquid and gas flow rates accompanying the severe slugging phenomenon. Clearly, such large transient variations could present difficulties for topside facilities unless they are designed to accommodate them.

Usually, the operators try not to operate in the severe slugging region. But the inlet conditions of a production pipeline are linked to the number and the capacity of the producing wells, the availability of wells and also to some undesirable operations such as shutdown or restart. The natural trend when dimensioning a production line is to do whatever is possible to avoid

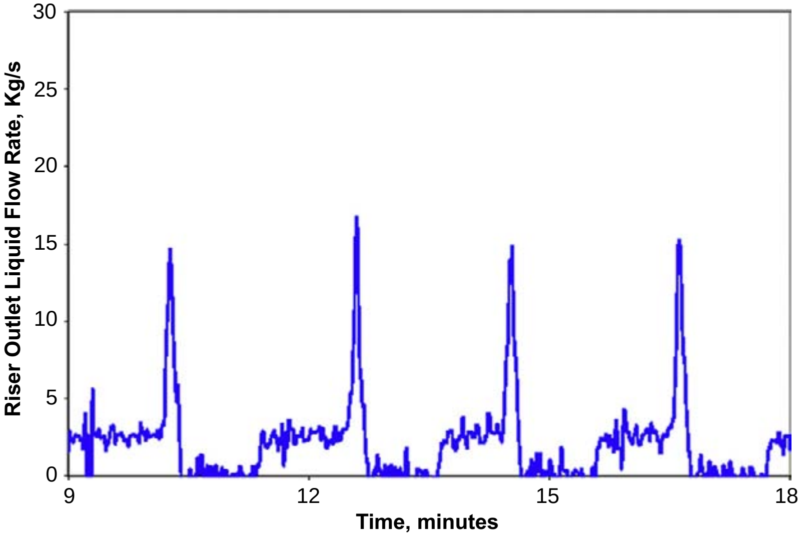

Figure 3-27 Sample time trace of the riser outlet liquid flow rate during severe slugging (Mokhatab, 2007).

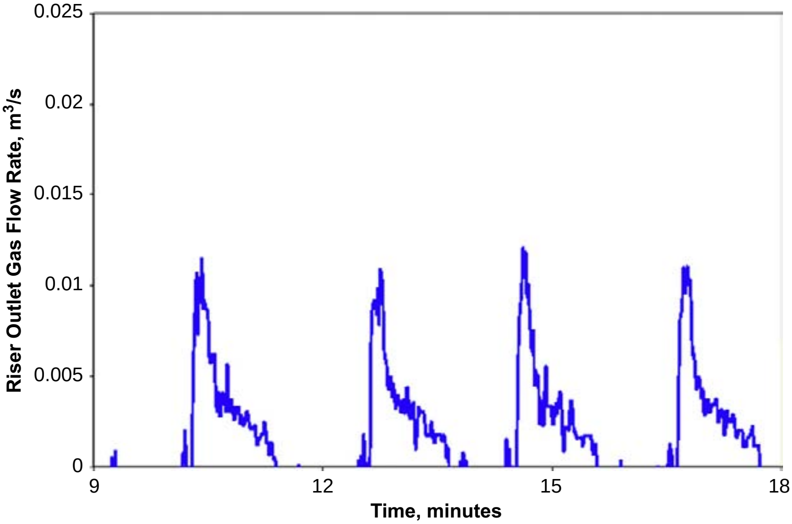

Figure 3-28 Sample time trace of the riser outlet gas flow rate during severe slugging (Mokhatab, 2007).

critical flooding of the separator and therefore to overdimension the separator unit. But in offshore production, overdimensioning the installation is very costly and not always possible. So, the design engineers require more accurate dynamic simulations to correctly design and dimension their production schemes (which are more sensitive to transients occurring when slug flow conditions build up and require high-performance control systems to maintain the plant within the preset operating ranges), and to be able to propose new concepts suitable to every situation they can face (Sagatun, 2004). The prediction of pipeline simulation with respect to slug dynamics will be strongly affected by the specification of boundary conditions. Hence, a tight integration of the hydrodynamic model of the pipeline-riser system with the dynamic model of the receiving plant is very important for achieving correct simulation results, especially for studying severe slugging dynamics (Mokhatab and Towler, 2007a).

Given these potential problems of severe slugging in such systems, understanding how and when severe slugging will form, as well as providing significant information on the best method to prevent and control this phenomenon, is a highly necessary undertaking.

Although the research community has undertaken a thorough exposition of severe slugging phenomenon in flexible risers, there is a lack of sufficiently wide-ranging and openly available transient code testing and full-scale data on flexible risers for better understanding the physics and characteristics of severe slugging and carrying out more vigorous verification (Mokhatab and Towler, 2007b). Unless such a vigorous link with reality is maintained, designers and operators will not have the confidence to expand the use of flexible risers in more critical applications.

3.8.4.3.1 Severe Slugging Mechanism

The process of severe slugging in a pipeline-riser system is considered to consist of four steps; (1) slug formation, (2) slug production, (3) bubble penetration, and (4) gas blow down. This phenomenon was previously identified by Schmidt et al. (1980) as a cyclic flow-rate variation, resulting in periods both of no flow and very high flow rates substantially greater than the time average. Figure 3-29 illustrates the stages of a severe slugging cycle.

The first step, slug formation, corresponds to an increase of pressure in the bottom of the riser. The liquid level does not reach the top of the riser. During this period, the liquid is no longer supported by the gas and begins to fall, resulting in riser entrance blockage and pipeline pressure buildup until

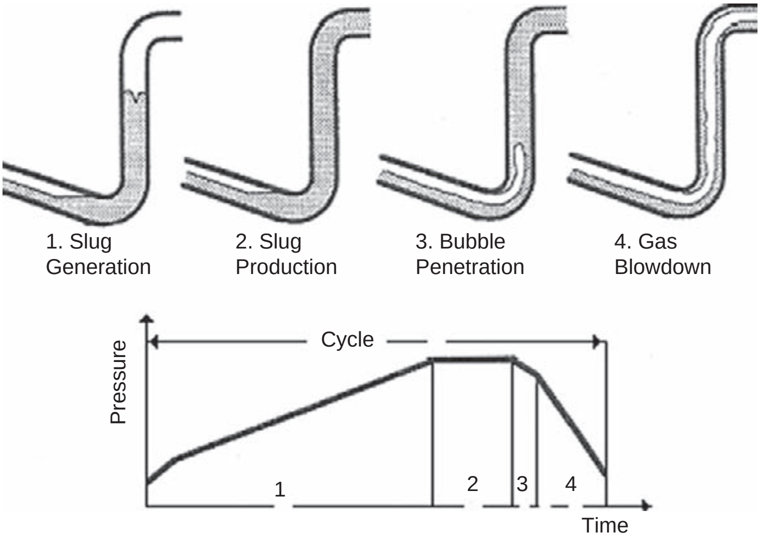

Figure 3-29 Description of severe slugging in pipeline-riser systems (Fabre et al., 1990).

the liquid level in the riser reaches to the top. During the second step, slug production, the liquid level reaches the riser outlet, and the liquid slug begins to be produced until the gas reaches the riser base. In the third step, bubble penetration, gas is again supplied to the riser, so the hydrostatic pressure decreases. As a result, the gas flow rate increases. The fourth step corresponds to gas blow down. When the gas produced at the riser bottom reaches the top, the pressure is minimal and the liquid is no longer gas-lifted. The liquid level falls and a new cycle begins (Fabre et al., 1990). This cyclic process becomes steady state when the rate of penetration of the gas into the riser is always positive. However, it is also possible that the penetration of the gas into the riser becomes zero. In this case, liquid blocks the bottom of the riser. This is followed by a movement of the liquid interface into the pipeline and blocking of the gas passage into the riser until the liquid interface reaches the bottom of the riser. At this point, penetration of gas into the riser starts and a new cycle begins again.

When liquid penetrates into the pipeline, the gas in the riser propagates to the top until all of the gas in the riser disappears. When the liquid input is very low, the propagation of the gas toward the top of the riser causes accumulation of all the gas at the top as the liquid falls back. This process is termed cyclic process with fallback, while the former case is termed cyclic process without fallback. In summary, three different possibilities that can occur as a result of penetration of gas into a liquid column in a quasi-steady severe slugging process are identified (Jansen et al., 1996):

1. Penetration of the gas that leads to oscillation, ending in a stable steady-state flow.
2. Penetration of the gas that leads to a cyclic operation without fallback of liquid.
3. Penetration of the gas that leads to a cyclic operation with fallback of liquid.

Severe slugging in a pipeline-riser system can be considered as a special case of flow in low-velocity hilly terrain pipelines, which are often encountered in offshore fields. This is a simple case of only one downward-inclined section (pipeline), one riser, and constant separator pressure. For this reason, severe slugging has been termed terrain-induced slugging (Fuchs, 1987). This phenomenon has also various names in the industry, including riser-base slugging, and riser-induced slugging.

3.8.4.3.2 Stability Analysis

The flow characteristics of multiphase flow in a pipeline-riser system are divided into two main regions: the stable (steady flow) region, and the unstable (pressure cycling) region, in which the stability line on the flow pattern map separates the two regions. The steady region includes acceptable slugging, annular mist, and bubble flows; while the pressure cycling region includes the severe slugging and transitional flows. Figure 3-30 shows a typical flow map for a pipeline-riser system developed by Griffith and Wallis (1961) featuring regions of stable and unstable behavior. In this figure, N_{Fr} and λ_G are the Froude number of two-phase gas/liquid flow and no-slip gas holdup, defined previously. As it can be seen from the figure, at low Froude

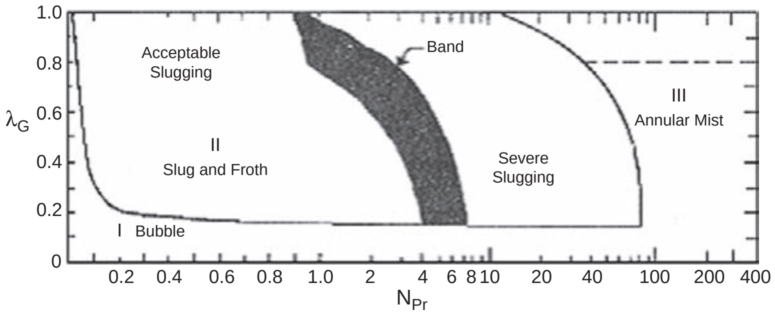

Figure 3-30 Griffith and Wallis flow pattern map with Yocum (1973) transition band (Brill and Beggs, 1991).

numbers, bubble flow prevails and fluids will flow through riser pipes without slug formation. However, as the Froude number increases, the slug flow range is entered.

The stability analysis predicts the boundary between stable and unstable regions, where the resultant stability map helps the engineer to design systems that operate well into the stable zone, thus offering an adequate margin of safety. The stability analysis seeks to model a particular process required for severe slugging and hence predicts the likelihood of severe slugging; as such, these stability models are termed criteria for severe slugging. The severe slugging process was first modeled by Schmidt et al. (1980); however, their model formed the basis of much of the early works for the stability of severe slugging. A review of the existing stability criteria for predicting severe slugging in a pipeline-riser system can be found in Mokhatab (2005).

3.8.4.3.3 Prevention and Control of Severe Slugging

The flow and pressure oscillations due to severe slugging phenomenon have several undesirable effects on the downstream topside facilities unless they are designed to accommodate them. However, designing the topside facilities to accept these transients may dictate large and expensive slug catchers with compression systems equipped with fast-responding control systems. This may not be cost effective, and it may be more prudent to design the system to operate in a stable manner (Sarica and Tengesdal, 2000). While lowering production rates (slowing fluid velocity) can minimize severe slugging, operators are investigating alternatives that would allow for maximum production rates without the interruptions caused by slugs (Furlow, 2000). Mokhatab et al. (2007a) referenced a combination of industrial experience and information from the literature to compile a list of methods of remediating the problems associated with severe slugging in pipeline-riser systems. In general, the severe slugging prevention and elimination strategies seek the three following approaches.

3.8.4.3.3.1 Riser Base Gas Injection The riser base gas injection method provides artificial lift for the liquids, moving them steadily through the riser. This technique can alleviate the problem of severe slugging by changing the flow regime from slug flow to annular or dispersed flow, but it does not help with transient slugging, in which the liquid column is already formed before it reaches the riser base. It is one of the most frequently used methods for the current applications. However, for deepwater systems, increased frictional pressure loss and Joule-Thomson cooling are potential problems resulting from high injection gas flow rates.

The riser base gas injection method was first used to control hydrodynamic slugging in vertical risers. However, Schmidt et al. (1980) dismissed it as not being economically feasible due to the cost of a compressor or pressurizing the gas for injection and piping required to transport the gas to the base of the riser. Pots et al. (1985) investigated the application of the method to control severe slugging. They concluded that the severity of the cycle was considerably lower for riser injection of about 50% inlet gas flow. Hill (1990) described the riser-base gas injection tests performed in the S.E. Forties field to eliminate severe slugging. The gas injection was shown to reduce the extent of the severe slugging.

The riser base gas lift method may cause additional problems due to Joule-Thomson cooling of the injected gas, where the lift gas will cause cooling and make the flow conditions more susceptible for the wax precipitation and hydrate formation. Hence, Johal et al. (1997) proposed an alternative technique “multiphase riser base lift” that requires nearby high-capacity multiphase lines diverted to the pipeline-riser system that alleviate the severe slugging problem without exposing the system to other potential problems.

Sarica and Tengesdal (2000) proposed a new technique for sourcing riser base gas lift. The principle of the proposed technique is to connect the riser to the downward-inclined segment of the pipeline with a small diameter conduit, where the conduit will transfer the gas from the downward-inclined segment to the riser at points above the riser base (multiposition gas injection). The transfer process reduces both the hydrostatic head in the riser and the pressure in the pipeline, consequently lessening or eliminating the severe slugging occurrence. This method can be considered as self-gas lifting. Sarica and Tengesdal (2000) claimed that the proposed technique is expected to increase production since it does not impose additional backpressure to the production system. The cost of implementing and operating the proposed systems in the field application is also expected to be low compared to other elimination methods.

3.8.4.3.3.2 Topside Choking The topside choking method induces bubble flow or normal slug flow in the riser by increasing the effective backpressure at the riser outlet. While a topside choke can keep liquids from overwhelming the system, it cannot provide required control of the gas surges that might be difficult for the downstream system to manage. This is a low-cost slug-mitigation option, but its application might be associated with considerable production deferment.

Topside choking was one of the first methods proposed for the control of the severe slugging phenomenon (Yocum, 1973). Yocum observed that increased backpressure could eliminate severe slugging but would severely reduce the flow capacity. Contrary to Yocum's claim, Schmidt et al. (1980) noted that the severe slugging in a pipeline-riser system could be eliminated or minimized by choking at the riser top, causing little or no changes in flow rates and pipeline pressure. Taitel (1986) provided a theoretical explanation for the success of choking to stabilize the flow as described by Schmidt et al. (1980).

Jansen et al., 1996 investigated different elimination methods such as gas lifting, choking, and gas-lifting and choking combination. They proposed the stability and the quasi-equilibrium models for the analysis of the aforementioned elimination methods. They experimentally made three observations: (1) large amounts of injected gas were needed to stabilize the flow with the gas-lifting technique; (2) careful choking was needed to stabilize the flow with minimal backpressure increase; (3) the gas-lifting and choking combination was the best elimination method, reducing the amount of injected gas and the degree of choking to stabilize the flow.

3.8.4.3.3.3 Control Methods Control methods (feed-forward control, slug choking, active feedback control) for slug handling are characterized by the use of process and/or pipeline information to adjust available degrees of freedom (pipeline chokes, pressure, and levels) to reduce or eliminate the effect of slugs in the downstream separation and compression unit. Control-based strategies are designed based on simulations using rigorous multiphase simulators, process knowledge, and iterative procedures. If one wishes to design efficient control systems, it is therefore advantageous to have an accurate model of the process (Storkaas et al., 2001; Bjune et al., 2002).

The feed-forwarded control aims to detect the buildup of slugs and, accordingly, prepares the separators to receive them, e.g., via feed-forwarded control to the separator level and pressure control loops. The aim of slug choking is to avoid overloading the process facilities with liquid or gas. This method makes use of a topside pipeline choke by reducing its opening in the presence of a slug and thereby protecting the downstream equipment (Courbot, 1996). Like slug choking, active feedback control makes use of a topside choke. However, with dynamic feedback control, the approach is to solve the slug problem by stabilizing the multiphase flow. Using feedback control to prevent severe slugging was proposed by Hedne and Linga (1990) and by other researchers (Henriot et al., 1999; Havre et al., 2000; Molyneux et al., 2000; Havre and Dalsmo, 2001; Bjune et al., 2002). The use of feedback control to stabilize an unstable operating point has several advantages. Most importantly, one is able to operate with even, nonoscillatory flow at a pressure drop that would otherwise give severe slugging. Figure 3-31 shows a typical application of an active feedback control approach on a production flow line/pipeline system and illustrates how the system uses pressure and temperature measurements (PT and TT) at the pipeline inlet and outlet to adjust the choke valve. If the pipeline flow measurements (FT) are also available, these can be used to adjust the nominal operating point and tuning parameters of the controller.

Note that the response times of large multiphase chokes are usually too long for such a system to be practical. The slug suppression system (S3) developed by Shell has avoided this problem by separating the fluids into a gas and liquid stream, controlling the liquid level in the separator by throttling the liquid stream, and controlling the total volumetric flow rate by throttling the gas stream. Hence, the gas control valve back pressures the separator to suppress surges, and as it is a gas choke, it is smaller and therefore more responsive than a multiphase choke.

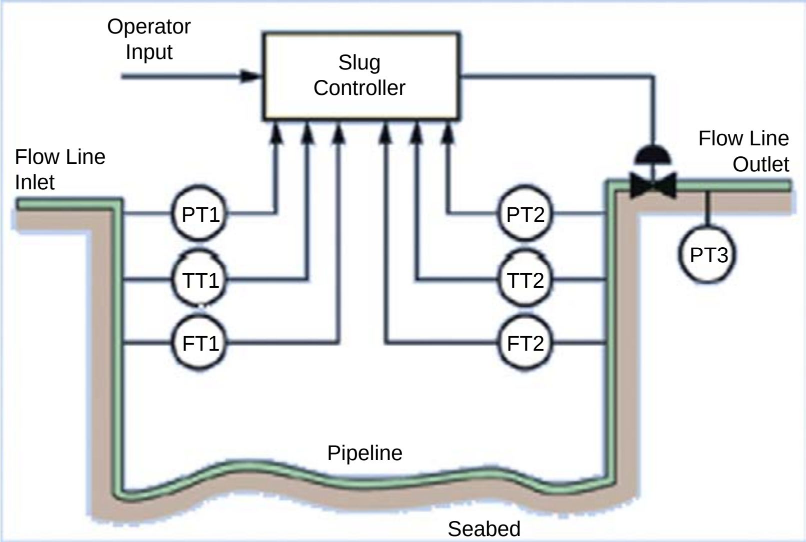

Figure 3-31 Typical configuration of feedback control technique in flow line/riser systems (Bjune et al., 2002).

The S3 is a small separator with dynamically controlled valves at the gas and liquid outlets, positioned between the pipeline outlet and the production separator. The outlet's valves are regulated by the control system using signals calculated from locally measured parameters, including pressure and liquid level in the S3 vessel and gas and liquid flow rates. The objective is to maintain constant total volumetric outflow. The system is designed to suppress severe slugging and decelerate transient slugs so that associated fluids can be produced at controlled rates. In fact, implementation of the S3 results in a stabilized gas and liquid production approximating the ideal production system. Installing S3 is a cost-effective modification and has lower capital costs than other slug catchers on production platforms. The slug suppression technology also has two advantages over other slug-mitigation solutions, where unlike a topside choke, the S3 does not cause production deferment and controls gas production, and the S3 controller uses locally measured variables as input variables and is independent of downstream facilities (Kovalev et al., 2003).

The design of stable pipeline-riser systems is particularly important in deepwater fields, since the propensity toward severe slugging is likely to be greater and the associated surges more pronounced at greater water depths. Therefore, system design and methodology used to control or eliminate the severe slugging phenomenon become very crucial when considering the safety of the operation and the limited available space on the platform. Currently, three basic elimination methods have been already proposed. However, the applicability of current elimination methods to deepwater systems is very much in question. In anticipation of this problem, different techniques should be developed to be suitable for different types of problems and production systems (Mokhatab et al., 2007a).

3.8.4.4 Operationally Induced Slugging

These types of slugging are initiated by pipeline operations, which may result in large production losses. Pigging of a pipeline causes most of the liquid inventory to be pushed from the line as a liquid slug ahead of the pig. When a pipeline is shut down, the liquid will drain to the low points in the line, and when the pipeline is restarted, the accumulated liquid can exit the pipeline as a slug. Slugs may also form during depressurization due to high gas velocities. In addition to these mechanisms, slugs may be produced as a result of flow-rate changes. Increasing or decreasing the flow rate of either gas or liquid leads to a change in liquid holdup. This can come out in the form of a slug, depending on the flow rate.

A transient multiphase flow program must be used to estimate the characteristics of these types of slugs. The volume of these transient slugs must be determined to ensure that they can be handled by the downstream separation and processing facilities. If there is insufficient capacity, operating procedures can be introduced to limit the size of any produced slugs.

3.8.5 Flow Assurance Risk Management

Flow assurance risk management for a new pipeline project is usually performed during the different project engineering and design phases. The better that these flow assurance risks are defined during the project phases, the fewer operational problems that are likely to be encountered. Similarly, the sooner the flow assurance risks are identified, the earlier their impacts on the project economics and the design of the pipeline system can be established.

Six phases exist for consideration of the flow assurance risk management for pipelines: assessing the risks, defining the mitigation strategies, defining flow operability, finalizing the operating procedures, optimizing system performance, and real-time monitoring. A brief description of each of these phases of flow assurance management follows (Song, 2008).

3.8.5.1 Phase I: Assessing Flow Assurance Risks

In Phase I, flow assurance risks are assessed from the sample analysis of the reservoir fluid, which is one of the most critical phases in flow assurance management. No matter how accurate the laboratory measurements and interpretations are, if the fluids do not represent the real production fluids, erroneous conclusions may be drawn.

From the collected samples, producers then consider which tests are critical to enabling the goal of properly evaluating the flow characteristics of the fluid and designing a pipeline system. Any inaccurate fluid properties that are reported are likely to lead to inappropriate flow assurance mitigation strategies that will not work properly and also lead to the risk of the pipeline system encountering severe operational impacts.

Water sampling to determine composition extent is also very critical in establishing flow assurance risks and may be very challenging to perform. Analysis of hydrate risk, corrosion prediction, and scaling tendency are dependent upon an assessment of the salinity and composition of the water produced with oil or gas. However, no water sample may be available because either the exploration wells never reach the aquifer zones or the water samples are contaminated by drilling muds (Guo et al., 2005).

Also many oil fields start production with little associated water initially present, but water content gradually increases over the production life. In such cases, the flow assurance specialist must often be content with estimates of salinity and compositions based upon analogs, such as samples taken from nearby fields or adjacent reservoir zones that have already been put into production. Consequently, all these potential water sampling issues may result in high uncertainty in the development of the flow assurance mitigation strategies (Wilkens, 2002).

3.8.5.2 Phase II: Defining Flow Assurance Mitigation Strategies

How all the flow assurance risks will be mitigated is studied in Phase II, and a high-level mitigation strategy is then developed. The flow assurance mitigation strategy is best developed with a good understanding of pipeline operating issues. The mitigation strategy details are also very closely related to the configuration of the pipeline system. The system configuration will dictate how many different ways the flow assurance must be mitigated and, similarly, critical flow assurance mitigation requirements will also drive some specific configurations being developed. For example, different hydrate mitigation strategies must be developed if the system configuration consists of a single flow line or dual flow lines. On the other hand, if chemical inhibitors will be required to mitigate the hydrate risks, then a separate, additional service line or a flow line could be required in the pipeline system to deliver the chemicals.

3.8.5.3 Phase III: Defining Flow Operability

Phase III involves defining flow operability. Operability is the set of design provisions and operating strategies that ensure the pipeline system can be started, operated, and shut down under all possible operating conditions (planned and unplanned) throughout the system life cycle. Operability is intrinsic to the flow assurance/pipeline system design process. In fact, after the high-level flow assurance mitigation plans are defined in phase II, good operating procedures (applying dynamic multiphase flow modeling of the system) will be defined that enable the pipeline system to perform with minimal flow assurance risks under a specified range of operating conditions.

3.8.5.4 Phase IV: Finalizing the Pipeline Operating Procedures

During Phase IV, the pipeline system components (e.g., flow line connectors, insulation system, flow line joints, and valves) are changed from the original design for various reasons, such as different vendors, different materials, and/or different properties. Due to these changes, the corresponding flow assurance mitigation techniques and operating procedures may need to be modified to reflect the changes. Also, even if the components are not modified; the actual manufactured ones may have different thermal-hydraulic properties from the designed ones based upon performance tests conducted on selected components. At the end of this phase, the flow assurance mitigation procedures and pipeline operating procedures will be finalized and used for system commissioning, start-up, and daily operations.

3.8.5.5 Phase V: Optimizing System Performance

In Phase V, the procedures will be modified based upon the actual recorded performance data from the pipeline system. Analysis of such data may identify some requirements that can be beneficially adjusted to optimize the system performance.

3.8.5.6 Phase VI: Real-Time Flow Assurance Monitoring

The last phase, VI, describes real-time flow assurance monitoring. In most cases, flow assurance problems cannot be completely eliminated due to unpredictable system component failures, unsuitability for operating conditions, faulty operational procedures for some situations, or operator failures/human errors that can occur in real-time operations. Significant efforts are therefore necessary to minimize the occurrence and impact of such failures. In this regard, real-time flow assurance monitoring systems can provide optimum asset management. Continuously monitoring the operating conditions of the pipeline for anomalous readings may indicate possible restrictions leading to potential blockage (i.e., observation of erratic pressure fluctuations is usually indicative of hydrate formation or wax deposition).

In fact, reliable, real-time data provided quickly and continuously throughout the life cycle of the pipeline can be tied into a software simulation program that can enable the data to be analyzed and modeled in a timely fashion. When a process interruption or operational upset occurs, the software can predict where problems are most likely to occur (some can recommend the best corrective action sequence). This feature allows faster detection and diagnosis of problems, decreasing the potential additional cost of multiphase production and substantially reducing the risk of environmental disasters due to the failure of unmonitored pipelines.

Continuous monitoring systems can provide wide bandwidth capability via fiber-optic-distributed sensors. Real-time data can pass at optimal transmission rates for analysis, ensuring that the data is moved reliably and appropriately referenced at each interval along the pipeline (Brower et al., 2005). However, for deepwater installations, the need for accurate and advanced distributed sensor systems is more critical. Such systems present greater opportunities for the occurrence of flow assurance problems with potentially significant adverse consequences.

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CHAPTER 4

Basic Concepts of Natural Gas Processing

4.1 INTRODUCTION

Natural gas coming from the well contains hydrocarbons, carbon dioxide, hydrogen sulfide, and water together with many other impurities. Raw natural gas after transmission through a network of gathering pipelines therefore must be processed in a safe manner and with minimal environmental effect before it can be moved into long-distance pipeline systems for use by consumers. While some of the needed processing can be accomplished at or near the wellhead (field processing), the complete processing of natural gas takes place at a processing plant, usually located in a natural-gas-producing region. The objective of a gas processing plant is to separate natural gas, associated hydrocarbons,[1] acid gases, and water from a gas-producing well and condition these fluids for sale or disposal. The processing philosophy depends on the type of project being considered and the level of treating required, i.e., the difference between the feed gas and product specifications. This determines what components will need to be removed or recovered from the gas stream.

This chapter describes the scope of natural gas processing and briefly reviews the function and purpose of each of the existing process units of the gas processing plants with greater details to follow in subsequent chapters.

4.2 SCOPE OF NATURAL GAS PROCESSING

The important factors that usually determine the extent of gas processing include the processing objectives, the type or source of the gas and the location and size of the gas fields (Manning and Thompson, 1991).

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[1] Associated hydrocarbons, known as natural gas liquids (NGLs), include ethane, propane, butane, iso-butane, and natural gasoline (pentanes plus). These NGLs are sold separately and have a variety of different uses.

4.2.1 Processing Objectives

Processing of a natural gas stream may have one of the following three principal objectives (Odello, 1981; Rojey et al., 1997):

- To produce a sales gas stream that meets typical specifications shown in Table 4-1. These specifications are mainly intended to meet pipeline requirements and the needs of industrial and domestic consumers.
- To maximize natural gas liquids (NGLs) production,[2] by producing a lean gas stripped of most of the hydrocarbons other than methane. NGLs are often extracted to enhance project economics.
- To deliver a commercial gas. Such gas must be distinguished by a certain range of gross heating value. In fact, customers on the transmission grid expect heating values to fall in a range compatible with their combustion equipment.

Note should be made that small changes in a sales gas specification can have a large effect on the cost and complexity of the gas processing plant. However, the negotiations of sales gas agreements between the seller and buyer frequently focus on gas quantity rather than gas quality issues, resulting in unclear gas quality specifications that are difficult or impossible to meet (Hubbard, 2009).

Table 4-1 Typical Natural Gas Specifications in the Salable Gas Stream

Characteristic	Specification
Water content	4–7 lbm H_2O/MMscf of gas
Hydrogen sulfide content	0.25–1.0 grain/100 scf
Gross heating value	950–1200 Btu/scf
Hydrocarbon dew point	14–40°F at specified pressure
Mercaptans content	0.25–1.0 grain/100 scf
Total sulfur content	0.5–20 grain/100 scf
Carbon dioxide content	3 mole%–4 mole%
Oxygen content	0.01 mole% (max)
Nitrogen content (often not specified)	4 mole%–5 mole%
Total inerts content (often not specified)	3 mole%–5 mole%
Sand, dust, gums, and free liquid	Commercially free
Typical delivery temperature[a]	120°F
Typical delivery pressure[b]	714.7 psia

[a] This temperature is set, sometimes, to have consistent metering and also to protect pipeline coatings.
[b] This pressure is set so that the pipeline company can properly design its system.

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[2] The required characteristics of NGLs are given in GPSA Engineering Data Book (2004).

4.2.2 Effect of Gas Type in Field Processing

The gas composition of the field is the most important issue in choosing a processing scheme. In other words, depending on the type of reservoir and the composition of the produced gas, the gas processing plant may contain extensive facilities to recover NGLs from natural gas for safety and economic reasons. Given the current level of gas market price volatility, gas producers are continuously evaluating the alternatives for gas processing. Obviously, producers will seek to maximize the value for their gas, whether it is processed for NGL removal or not. The gas processing scheme will also be dictated by the format of the sales contract and its specifications. The contract may be totally different for each customer, depending on the composition and amount of gas, plant recoveries, and the contractual preferences of the customer (Bullin and Hall, 2000; Spletter and Adair, 2001; Starr and Adair, 2004).

4.2.3 Location of the Gas Field

Since the location and output of the wells can vary widely, then not surprisingly, the systems that have been designed to collect and process this output also vary widely (Thorn et al., 1999). However, in pursuit of greater safety, lower environmental impact, and lower capital expenditure, it is desired to reduce (or preferably eliminate) offshore processing.

4.3 TYPICAL SETUP OF GAS PROCESSING PLANTS

Industrial gas plants can vary considerably in shape and in size—from a simple gas receiving and letdown terminal to a full gas processing facility with NGL extraction and sulfur recovery. In order to define the type of gas processing plant and its different treating units, the products to be obtained (e.g., gas for pipeline, NGL recovery, . . .) and its specifications should be fixed, where the quantity and type of feed impurities determine the necessary treatment steps.

A typical scheme for most gas processing plants designed to produce pipeline gas from a sour gas feed is shown in Figure 4-1. Each process operation module consists of a single piece or a group of equipment performing a specific function. All the modules shown will not necessarily be present in every gas plant. In some cases, little processing is needed; however, most natural gas requires all processing units at the gas processing plant to remove impurities, water, and excess hydrocarbon liquid and to control delivery pressure. Therefore, the actual processes used depend on the feed-gas

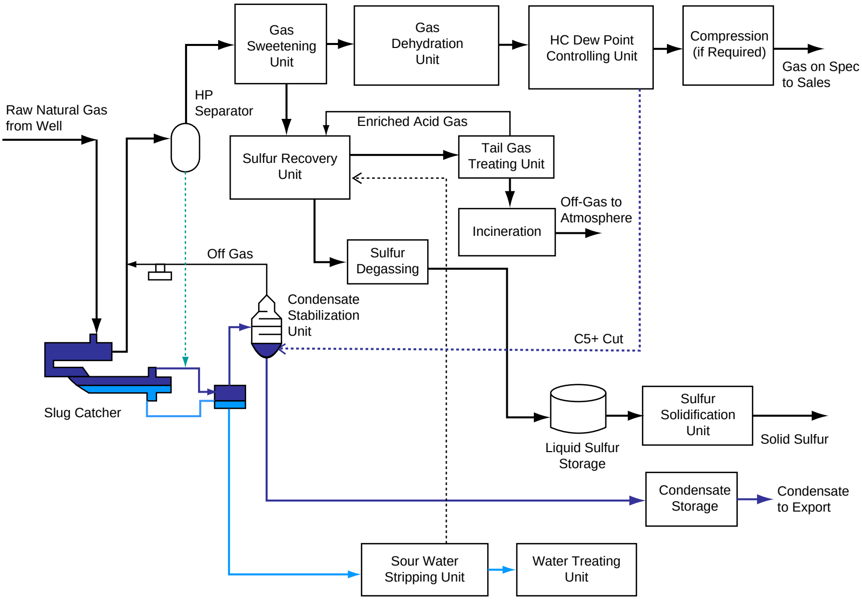

Figure 4-1 Typical setup of gas processing plant producing sales gas.

composition and the sales specifications for both the gas and liquid products. In the meantime, the unit operations used in a given application may not be arranged in the sequence shown in this figure, although this sequence is typical. The choice of process modules to be used and the arrangement of these modules are determined during the design stage of each gas-field development project (Mokhatab and Meyer, 2009; Mokhatab, 2010).

As can be seen from Figure 4-1, field production upon arrival at the processing plant will be processed in a slug catcher, which catches liquid slugs and then allows them to flow into downstream equipment and facilities at a rate at which the liquid can be properly handled. Produced gas from the outlet of the slug catcher is directed to a high-pressure (HP) separator, where final separation of liquid from gas takes place. More detailed information about phase separation is presented in Chapter 5. Note should be made that the interface between production and reception facilities requires due attention to ensure a stable trouble-free operation of the entire complex. Zomerman et al. (2009) identified the issues that could result in onshore gas reception facilities due to inadequate design caused by poor interface between production and reception facilities and presented design improvements as well as an effective approach in identifying a way forward for reception facilities with sour wet pipelines.

The HP raw gas flows through to the gas sweetening unit (GSU), in which acidic components like hydrogen sulfide (H_2S) and carbon dioxide (CO_2) are removed from the gas stream.[3] Acid gas removal processes are covered in Chapter 7. Simultaneous removal of the organic sulfur compounds such as mercaptans (R-SH), carbonyl sulfide (COS), and carbon disulfide (CS_2) in the GSU is also desired,[4] as it facilitates the downstream processing and purification steps and contributes to the reduction of the total sulfur content of the treated gas.[5] However, processes that remove hydrogen sulfide and carbon dioxide from natural gas are not effective in removing mercaptans, carbonyl sulfide, carbon disulfides, etc. The reason is that the these organic sulfur compounds do not readily hydrolyze to form anions that react with caustic, amine, or metal oxides. The preferred removal approach is physical absorption using a solvent with a high affinity for organic sulfur compounds, either neat or in concert with an amine (Seagraves, 2001; Bedell et al., 2008). The most widely used physical solvent-based processes are briefly described in Chapter 7. Note should be made that although mercaptans have acidic properties, they are not completely removed during the sweetening treatment. Taking into consideration the concentration values permitted for the sales gas, in some cases it may be necessary to proceed with further treatment. In this case, a significant reduction of mercaptans (usually heavy mercaptans) can be obtained with the cooling of the gas for deep condensate removal. As a consequence, in an overall gas processing system, which involves sweetening followed by deep condensate removal, both light and heavy mercaptans are removed by the combined effect of the two treatments. Therefore, there is normally no need to resort to specific treatment to remove mercaptans, since this is achieved by the treatment process as a whole.

The acid gas stream from the GSU is sent to the flare or to processing. If the acid gas stream is CO_2 and contains no sulfur compounds, it is either vented to atmosphere or reinjected into the subsurface for enhanced oil recovery or for CO_2 sequestration. When the recovered acid gas stream contains sulfur compounds, venting or flaring typically is not an option (Hubbard, 2009). In these cases, the acid gas stream can be processed to

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[3] H_2S and CO_2 are referred to as acid gas components because they dissolve in water to form a weak acid.

[4] There are two motives for removing mercaptans in the gas phase. The first is corrosivity. If the mercaptan content is high enough, then NACE metallurgy may be necessary. The second reason is that it will concentrate in the liquid hydrocarbon streams, and these products will not pass the copper strip test.

[5] This specification is set to control the foul-smelling compounds. These can be, besides mercaptans, thiophenes and disulphides. Also, sulfur compounds can be a catalyst poison in some reactions involving natural gas.

produce elemental sulfur in a sulfur recovery unit (SRU), consisting of a Claus unit and an associated tail gas treating unit (TGTU) if higher recovery rates are specified for the SRU itself. The acid gas stream from TGTU can be treated in an acid gas enrichment unit utilizing a specialty solvent that preferentially absorbs the H_2S and leaves most of the CO_2 behind. The H_2S stream is then recycled to the Claus unit. The final residual gas from the TGTU is incinerated before being vented to atmosphere. Sulfur recovery and handling options are covered in Chapter 8. Note should be made that the economics of recovering sulfur from sour natural gas have become unfavorable for small fields. Considering this and in accordance with this point that air emission standards and regulations are increasingly stringent, there is an essential need for an environmentally friendly, cost-effective method to deal with acid gas. Acid gas compression and reinjection into depleted reservoirs or disposal zones offer a viable alternative to traditional sulfur recovery processes with the added advantages of reducing greenhouse gas emissions and providing pressure support for producing reservoirs. Most injection schemes include dehydration facilities to remove the saturated water from the gas, preventing corrosion and hydrate formation. However, a less-costly approach keeps the water in the acid gas phase throughout the injection circuit, eliminating the need to dehydrate. More details about acid gas injection theory and practical applications, which are beyond the scope of this book, can be found in Carroll (2010).

Sweetened gas from the GSU then flows to the gas dehydration unit (GDU) to reduce and control the water content of gas, to ensure safe processing and transmission. Gas dehydration is discussed in detail in Chapter 9.

The dry sweet gas is then sent to the hydrocarbon dew point controlling unit (DPCU) to meet the required hydrocarbon dew point and heating value specifications of gas prior to entering transmission pipelines, which typically requires only minimal extraction of $C_5{}^+$ components. Natural gas processing for liquid recovery is discussed in Chapter 10.

The gas pressure (depending on the type of process used in the DPCU) may also need to be boosted by the high-pressure compressors before passing through the sales gas meter and into the export pipeline. Natural gas compression is discussed in Chapter 11. Transport of sales gas is done at high pressure in order to reduce pipeline diameter. More details are available in Chapter 12.

The liquids from the reception facilities are separated again into produced (associated) water and condensate.[6] Other sources of wastewater include

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[6] The hydrocarbon liquid fraction obtained from a gas stream containing essentially pentanes plus.

water used during start-up, purge water from amine sweetening units, and wash water used in the tail gas treating section of the sulfur recovery unit. Of these sources, associated water is the largest in total volume; however, the latter two are the largest sources of sour water[7] (Addington et al., 2011). Sour water must be processed, typically in a sour water stripping unit, before it is either used or disposed of in a variety of ways. While municipal or other sewage systems may be used when available, many plants contain onsite wastewater treatment facilities, where water is processed and discharged.

Note should be made that in recent years increasing amounts of saline formation water are typically present in produced hydrocarbon fluids. In these cases, the monoethylene glycol (MEG) used for the hydrate control in hydrocarbon production pipelines is often contaminated with the salts contained in the formation waters.[8] Since salt is nonvolatile, it will remain in the lean glycol during regeneration as the water is boiled off, and serious fouling and plugging of equipment and pipelines can result (Son and Wallace, 2000). MEG processing technology in the form of MEG regeneration/ reclamation packages has developed over the past few years to be designed specifically to remove salts and other solids while also removing the water to achieve the required outlet glycol purity. The processed MEG is then collected in a lean MEG tank for reinjection to the production field.

The condensate from the reception facilities is combined with the hydrocarbon liquids knocked out in the hydrocarbon dew point controlling unit. It is then processed in the condensate stabilization unit (CSU) to reduce the vapor pressure[9] (by the full removal of light hydrocarbon fractions from the condensate) to allow storage in atmospheric storage tanks. Flashed gas from the CSU typically is used as fuel gas or is recycled to the feed gas. Condensate stabilization is discussed in Chapter 6. In the case of sour condensate, if there is any treatment required for the heavier sulfur components, any of the liquid sweetening processes discussed in Chapter 10 can be used for such purpose. However, in many cases, the condensate volumes are low, and sweetening at the plant is not justified. Dehydration of condensate usually is also unnecessary. When in storage, free water drops out, and the remaining low-water-content condensate should meet customer requirements. However, emulsions of water and

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[7] In gas processing plants, the main contaminant of concern in sour water is H_2S.

[8] At land locations, formation brine is usually separated from the gas near the wellhead and prior to injection of glycol. This separation step reduces salt contamination of the glycol. However, for offshore production systems, separation of produced water on the sea floor is not normally possible.

[9] Generally, a specification of 12–14 psia for a Reid vapor pressure has to be met.

condensate occur, which require emulsion breakers to obtain a clean water-hydrocarbon separation (Kidnay and Parrish, 2006).

Note should be made that the illustrated arrangement for the gas processing units in Figure 4-1 is typical and shows the functional blocks required to achieve the objectives of the plant designed to produce pipeline gas from a sour gas feed. The function of each of these blocks (or combinations of blocks) can be achieved in a variety of ways; some technology options may result in one or more of these functions being achieved simultaneously (thus reducing the complexity of the design). As an example, molecular sieves can be used for both dehydration and mercaptans removal, reducing the two steps presented in the previous scheme to one. The water and mercaptans desorbed from the bed during the regeneration can be captured by a physical absorption process to concentrate the mercaptans in a stream being sent to the SRU. However, operational problems and upsets of the molecular sieve unit (MSU) can make the molecular sieve unit the bottleneck of the plant. Where a molecular sieve unit is a bottleneck, onsite extensive know-how of molecular sieves can improve the performance of the molecular sieve unit (Bruijn et al., 2002).

Figure 4-2 also illustrates a typical block flow diagram of a gas processing plant designed to produce pipeline gas with a full range of NGL products from a sour feed gas. In the conventional NGL recovery plants, feed gas

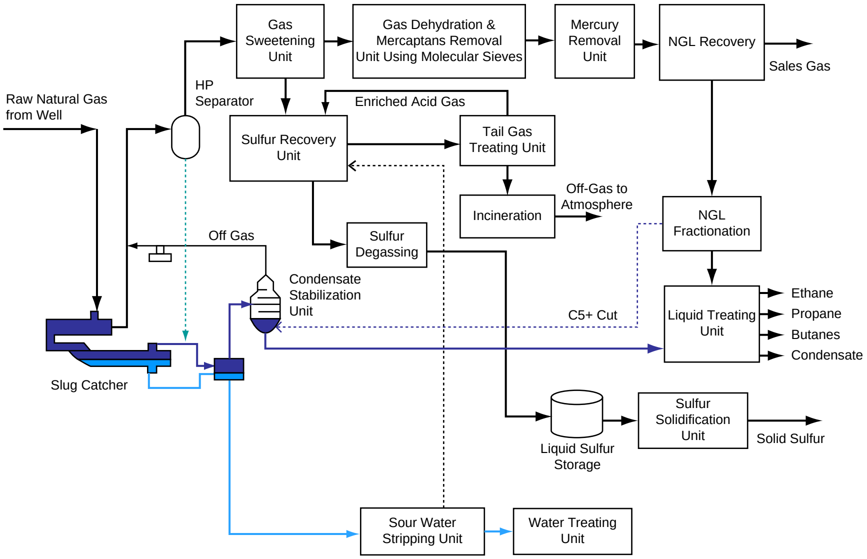

Figure 4-2 Typical setup of gas processing plant producing sales gas and NGLs.

must be pretreated for removal of acid gases, water, and mercury to produce a purified gas suitable for liquids recovery under cryogenic temperatures. The typical specification to be met is H_2S removal to less than 4 ppmv, CO_2 to less than 1%–2%, water to less than 10 ppmv, and mercury to levels of 0.01 $\mu g/Nm^3$.

The level of CO_2 removal depends on the whether the NGL recovery unit is on ethane recovery or propane recovery. For ethane recovery, the CO_2 content in the ethane product must be met since the relative volatility between CO_2 and ethane is close and CO_2 is condensed together with ethane. Typically, if ethane product is used for petrochemical feedstocks, it should contain no more than 2% volume of CO_2. Under this condition, CO_2 in the feed gas to the NGL recovery unit must be reduced to a low level. On the other hand, if high ethane recovery such as higher than 90% is required, the demethanizer is required to operate at a very low overhead temperature. Under this condition, the demethanizer column is prone to CO_2 freezing, and CO_2 must typically be removed to below 1% to stay away from the CO_2 freezing zone. The level of CO_2 also depends on the feed gas composition. A lean gas is more prone to CO_2 freezing. If only propane is required to be recovered, the deethanizer column operates at a warmer temperature, and generally stays away from the CO_2 freezing zone. For propane recovery, the treated gas needs only to meet the sales gas specification, which is typically 3 mole%.

The feed gas stream entering an NGL recovery unit must be also dehydrated in a molecular sieve unit to a relatively low level (with the value depending on whether the NGL recovery unit recovers ethane or propane and the resulting temperature required for desired recovery) to avoid hydrate formation. Any undesirable amounts of the light mercaptans that are not absorbed during the sweetening process will be removed completely by the molecular sieves used for dehydration. A detailed discussion of molecular sieve units is given in Chapter 9.

Almost every NGL plant will have a mercury removal unit installed. The reason is that the risks from mercury attack for the NGL plant are severe, and it is difficult to predict the possibility of mercury production from hydrocarbon reservoirs. Mercury can result in severe corrosion of the brazed-aluminum heat exchangers used in cryogenic section of the NGL plant, and potentially can present environmental and safety hazards. There is also increasing concern about the presence of mercury in the feed stocks (i.e. LPG) supplied to petrochemical plants due to corrosion of cryogenic equipment and poisoning of precious metal catalysts (Row and

Humphrys, 2011). However, corrosion is a particular concern for NGL plants and for this reason conservative design requires mercury removal to levels below 0.01 $\mu g/Nm^3$. The mercury removal step can be positioned upstream of the acid gas sweetening or downstream of the dehydration step. When upfront mercury removal is required on raw gas, the objective is to prevent mercury migration to various locations within the plant and to avoid any subsequent partitioning into processed natural gas and condensate streams (Edmonds et al., 1996). This approach uses large vessels of nonregenerative sorbents (metallic and carbon-based sorbents containing sulfur) to remove mercury from the raw gas upstream of the GSU. Although the non-regenerative methods are simple, however, eventual disposal of the used sorbent can be expensive since the sorbent not only picks up the mercury, but it will often contain other hazardous material such as benzene and other hydrocarbons (Markovs and Clark, 2005). The protection of aluminum heat exchangers can be also accomplished by using the silver-impregnated molecular sieve inside the MSU to chemisorb elemental mercury. The mercury is then regenerated off the adsorbent, where the mercury condenses and recovers as liquid mercury. If necessary, the regeneration gas can be further treated with a non-regenerative mercury removal process to remove the last traces of mercury. In this method of mercury removal, since the mercury does not accumulate on the adsorbent, it presents no spent adsorbent disposal issues. Also, this option will give good mercury removal performance with longer service life. However, this ignores the problem of mercury emissions from the GSU and MSU as well as the presence of mercury in liquid hydrocarbons and the contamination of process equipment (Markovs and Clark, 2005; Eckersley, 2010). Note should be made that the selecting the best mercury removal option will depend on many factors. The best method for mercury removal in some plants may be to use a combination of both non-regenerative and regenerative mercury removal methods.

The final treating step in an NGL plant would be an NGL recovery unit. This can include a range of options from simple to complex NGL recovery processes that produce the entire range of NGLs available from natural gas to a high recovery level. The level of NGL extraction from natural gas is somewhat discretionary. Safety issues dictate the minimum extraction level, whereas a balance between the technology and the relative value of the NGLs determines the maximum extraction level (Spletter and Adair, 2001). Fractionation of the NGLs to produce purity products such as petrochemical-grade ethane feedstock or fuel-grade propane may take place in a gas processing plant but more commonly occurs at another location, usually a regional market center. The NGLs products produced from gas processing technologies rarely achieve product specifications without further treatment. NGL products may therefore require additional treatment before they can be sold into the desired market. More detailed information about NGL recovery processes, fractionation, and products treating is presented in Chapter 10.

Occasionally, the feed gas has a nitrogen concentration that exceeds the sales gas specification (4 mole%–5 mole%). Removal of nitrogen is therefore necessary for conditioning of the natural gas that leads to reduced transportation volumes and an increased heating value. The most common method of removing nitrogen from natural gas is cryogenic distillation that will be done in the NGL extraction section of the plant. Cryogenic distillation is the most economical process and can provide higher nitrogen rejection at high gas throughput. However, at low gas flow rates, membranes and pressure swing adsorption are economically feasible technologies for nitrogen separation or rejection. More details on nitrogen removal from natural gas can be found in Kidnay and Parrish (2006). Many sales gas transmission pipelines also consider limiting maximum oxygen concentrations to 10 ppm vol (0.01 mol%). In fact, oxygen ingress in gas pipelines can cause corrosion in small concentrations and even combustion in larger amounts (Sridhar et al., 2001). It can also cause the sales gas to be subquality if the concentration exceeds the sales gas specification. Oxygen is removed in sales gas with nitrogen rejection, but this step follows the processing steps most susceptible to oxygen damage (Kidnay and Parrish, 2006).

4.4 FINDING THE BEST GAS PROCESSING ROUTE

For a given gas processing application, the different process configurations and choice of technologies can be vast (Mokhatab and Meyer, 2009; Mokhatab, 2010). The different options therefore should be evaluated individually to find the best solution for the given gas composition. However, for optimal treating-package design, process selection for the individual units must be made on the basis of an integrated approach that considers interactions between units. In fact, a process configuration cannot simply be taken for granted; it needs careful thought and evaluation—giving special consideration to consequences of selecting different technologies and any peculiar requirements of downstream processes (McMahon, 2004). Chapter 13 provides an awareness of pros and cons for the different process technologies/lineups and opportunities that may be encountered.

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CHAPTER 5

Phase Separation

5.1 INTRODUCTION

Separation of oil and gas is a critical field processing operation. As producing pressure is increased and lighter condensates are produced, efficient separation has become more critical than ever. Moreover, some of the new concepts in separation technology have been applied to advantage on old leases, producing oil at moderate or low pressures. As gas transmission lines raise their standards, separation becomes a part of the overall field processing necessary to condition the gas. Several technologies are available to remove liquids and solids from gases. However, selecting gas/liquid separation technologies requires not only knowledge of the process conditions, but a knowledge of the characteristics of the liquid contaminants. Selection should be made based on the droplet size, concentration, and whether the liquid has waxing or fouling tendencies (Brown et al., 1994). Before evaluating specific technologies, one needs to understand the mechanisms used to remove liquids and solids from gases. Three principles used to achieve physical separation of gas and liquids or solids are momentum, gravity settling, and coalescing. Any separator may employ one or more of these principles; however, the fluid phases must be immiscible and have different densities for separation to occur. Momentum force is utilized by changing the direction of flow and is usually employed for bulk separation of the fluid phases. The gravitational force is utilized by reducing velocity so the liquid droplets can settle out in the space provided. Gravity segregation is the main force that accomplishes the separation, which means the heaviest fluid settles to the bottom and the lightest fluid rises to the top. However, very small droplets such as mist cannot be separated practically by gravity. These droplets can be coalesced to form larger droplets that will settle by gravity. The purpose of this chapter is to review the principles governing the basic separation process and associated equipment design procedure.

5.2 GRAVITY SEPARATORS

Gravity separators are pressure vessels that separate a mixed-phase stream into gas and liquid phases that are relatively free of each other. In a gravity separator, gravitational forces control separation, and the efficiency of the gas/liquid separation is increased by lowering the gas velocity. Because of the large vessel size required to achieve settling, gravity separators are rarely designed to remove droplets smaller than 250 μm (Talavera, 1990). However, an analysis of this type of separator is given because it is useful to help understand the settling mechanism of other separators.

Gravity separators are often classified by their geometrical configuration (vertical, horizontal) and by their function (two-phase/three-phase separator). In other words, gravity separators are classified as "two-phase" if they separate gas from the total liquid stream and "three-phase" if they also separate the liquid stream into its crude oil and water-rich phases. Additionally, separators can be categorized according to their operating pressure. Low-pressure units handle pressures of 10 to 180 psi. Medium-pressure separators operate from 230 to 700 psi. High-pressure units handle pressures of 975 to 1,500 psi.

Separators are sometimes called "scrubbers" when the ratio of gas rate to liquid rate is very high. These vessels usually have a small liquid collection section and are recommended only for the following items:

- Secondary separation to remove carryover fluids from process equipment such as absorbers and liquid dust scrubbers.
- Gas line separation downstream from a separator and where flow lines are not long.
- Miscellaneous separation where the gas-liquid ratio is extremely high.

In any case, these vessels have the same configuration and are sized in accordance with the same procedure of separators.

5.2.1 General Description

All gravity separators normally have the following components or features (API Spec 12J, 1989):

- A primary gas/liquid separation section with an inlet diverter to remove the bulk of the liquid from the gas.
- A gravity-settling section providing adequate retention time so that proper settling may take place.
- A mist extractor at the gas outlet to capture entrained droplets or those too small to settle by gravity.
- Proper pressure and liquid-level controls.

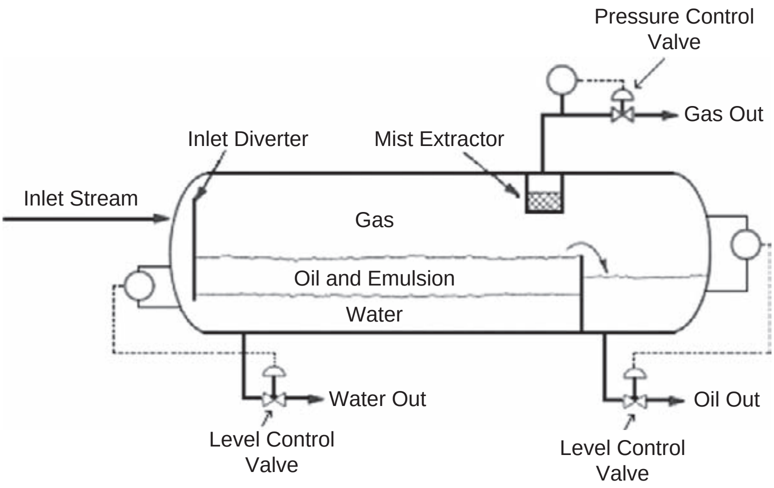

Figure 5-1 A typical scheme of horizontal three-phase separator.

Gravity separators are designed as either horizontal or vertical pressure vessels. Figure 5-1 shows a typical scheme for a three-phase horizontal separator. The fluid enters the separator and hits an inlet diverter. This sudden change in momentum generates the initial bulk separation of liquid and gas. In most designs, the inlet diverter contains a downcomer that directs the liquid flow below the oil/water interface. This forces the inlet mixture of oil and water to mix with the water continuous phase in the bottom of the vessel and rise through the oil/water interface. This process is called "water washing" and it promotes the coalescence of water droplets that are entrained in the oil continuous phase. The inlet diverter assures that little gas is carried with the liquid, and the water wash assures that the liquid does not fall on top of the gas/oil or oil/water interface, mixing the liquid retained in the vessel and making control of the oil/water interface difficult. The liquid-collecting section of the vessel provides sufficient time so that the oil and emulsion form a layer or "oil pad" at the top. The free water settles to the bottom. The produced water flows from a nozzle in the vessel located upstream of the oil weir. An interface-level controller senses the height of the oil/water interface. The controller sends a signal to the water dump valve, thus allowing the correct amount of water to leave the vessel so that the oil/water interface is maintained at the design height.

The gas flows horizontally and exits through a mist extractor (normally known as a demisting device) to a pressure control valve that maintains constant vessel pressure. The level of the gas/oil interface can vary from half the diameter to 75% of the diameter depending on the relative importance of liquid/gas separation and what purpose the separator has. For example, degassers and produced water flash drums have very high liquid/gas interfaces. However, the most common configuration is half full.

Figure 5-2 shows a typical configuration for a vertical three-phase separator. In the vertical separator, the flow enters the vessel through the side, as in the horizontal separator, and the inlet diverter separates the bulk of the gas. The gas moves upward, usually passing through a mist extractor to remove suspended mist, and then the dry gas flows out. A downcomer is required to transmit the liquid collected through the oil-gas interface so as not to disturb the oil-skimming action taking place. A chimney is needed to equalize gas pressure between the lower section and the gas section. The spreader or downcomer outlet is located at the oil-water interface. From this point as the oil rises, any free water trapped within the oil phase separates out. The water droplets flow countercurrent to the oil. Similarly, the water flows downward, and oil droplets trapped in the water phase tend to rise countercurrent to the water flow. It should be clear that the principles of operation (such as oil/water interface level controlling) of three-phase

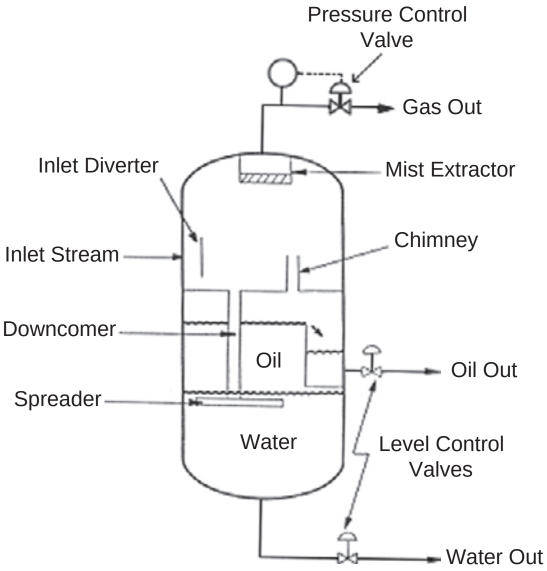

Figure 5-2 A typical scheme of vertical three-phase separator.

vertical separators are the same as the three-phase horizontal separators described previously. Essentially, the only difference is that horizontal separators have separation acting tangentially to flow, whereas vertical separators have separation acting parallel to flow. In the vertical separator, also, level control is not critical, where the liquid level can fluctuate several inches without affecting operating efficiency (GPSA, 2004). However, it can affect the pressure drop for downcomer pipe (from the demister) and can therefore affect demisting device drainage.

5.2.2 Separator Selection

There are no simple rules for separator selection. Sometimes, both configurations should be evaluated to decide which is more economical. The relative merits and common applications of vertical and horizontal separators are summarized by Manning and Thompson (1995).

5.2.2.1 Horizontal Separators

Horizontal separators are most commonly used in the following conditions:

- Large volumes of gas and/or liquids
- High-to-medium gas/oil ratio (GOR) streams
- Foaming crudes
- Three-phase separation

The advantages and disadvantages of these separators are as follows:

Advantages:

- Require smaller diameter for similar gas capacity as compared to vertical vessels.
- No counter flow (gas flow does not oppose drainage of mist extractor).
- Large liquid surface area for foam dispersion generally reduces turbulence.
- Larger surge volume capacity.

Disadvantages:

- Only part of shell available for passage of gas.
- Occupy more space unless "stack" mounted.
- Liquid-level control is more critical.
- More difficult to clean produced sand, mud, wax, paraffin, etc.

5.2.2.2 Vertical Separators

These separators are used in the following conditions:

- Small flow rates of gas and/or liquids.
- Very high GOR streams or when the total gas volumes are low.

- Plot space is limited.
- Ease of level control is desired.

The advantages and disadvantages of these separators are as follows:

Advantages:

- Liquid-level control is not so critical.
- Have good bottom-drain and clean-out facilities.
- Can handle more sand, mud, paraffin, and wax without plugging.
- Less tendency for reentrainment.
- Have full diameter for gas flow at top and oil flow at bottom.
- Occupy smaller plot area.

Disadvantages:

- Require larger diameter for a given gas capacity, therefore, most competitive for very low GOR or very high GOR or scrubber applications.
- They are not recommended when there is a large slug potential.
- More difficult to reach and service top-mounted instruments and safety devices.

5.2.3 Gravity Separation Theory

Vapor/liquid separation is usually accomplished in three stages. The first stage, primary separation, uses an inlet diverter to cause the largest droplets to impinge by momentum and then drop by gravity.[1] The next stage, secondary separation, is gravity separation of smaller droplets as the vapor flows through the disengagement area. Gravity separation can be aided by utilizing distribution baffles that create an even velocity distribution in the fluid, thus allowing enhanced separation. The final stage is mist elimination, where the smallest droplets are coalesced on an impingement device, such as a mist pad or vane pack, followed by gravity settling of the larger formed droplets.

In the gravity-settling section of the separators, the liquid drops will settle at a velocity determined by equating the gravity force (F_B) on the drop with the drag force (F_D) caused by its motion relative to the vapor continuous phase. When the drag force is equal to the buoyancy (gravity) force, the droplets acceleration is zero so that it moves at a constant velocity.

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[1] Inlet diverters are very old technology now and are very rarely used. Revamps in the North Sea replace these inlet devices all the time, especially where the asset is producing more throughput than originally expected. If incorrectly sized, inlet devices can cause serious separation issues due to droplet shatter. Schoepentoeters became the popular inlet device for a while, but these were designed for vertical separators and are therefore not always applicable. Cyclonic and distribution baffle inlet devices are more common nowadays.

This velocity is the terminal or free-settling velocity, which is determined with respect to the following equations:

$$F_B = \left(\frac{\pi}{6}\right) D_d^3(\rho_L - \rho_V)\left(\frac{g}{g_C}\right) \tag{5-1}$$

where D_d is the drop diameter, ft; ρ_L is liquid density, lbm/ft^3; ρ_V is vapor density, lbm/ft^3; g is the gravitational constant, 32.174 ft/sec^2; and g_c is the conversion factor, 32.174 lbm-ft/sec^2-lbf.

Also, the drag force on the droplet is given by

$$F_D = C_D(A_P)\left(\frac{\rho_V\ V_d^2}{2\,g_C}\right) \tag{5-2}$$

where C_D is the drag coefficient, dimensionless; A_p is the projected drop area, ft^2; and V_d is the drop velocity, ft/sec.

Therefore, the terminal-settling velocity of the liquid droplets (V_t) can be determined by equating Equations 5-1 and 5-2 as follows:

$$V_t = \sqrt{(4/3)D_d(\rho_L - \rho_V)g/(C_D\rho_V)} \tag{5-3}$$

The drag coefficient can also be calculated as follows (Svrcek and Monnery, 1993):

$$C_D = \frac{5.0074}{\mathrm{Ln}(x)} + \frac{40.927}{\sqrt{X}} + \frac{484.07}{X} \tag{5-4}$$

and

$$X = \frac{0.95 \times 10^8\ D_d^3\,\rho_V\,(\rho_L - \rho_V)}{{\mu_V}^2} \tag{5-5}$$

where D_d is in ft, densities are in lb/ft^3, and viscosity is in cP.

The droplet-settling velocity equation considers the escape of a drop from the continuous phase (e.g., the escape of an oil drop from the gas phase). For this purpose, the droplet-settling velocity must be greater than the superficial upward bulk vapor velocity, V_V. Typically, the allowable vapor velocity is set between 0.75 V_t and V_t (Svrcek and Monnery, 1993). Equation 5-3 can be rearranged as a Souders and Brown (1934) type equation as follows:

$$V_t = K_{SB}\sqrt{(\rho_L - \rho_V)/\rho_V} \tag{5-6}$$

where

$$K_{SB} = \sqrt{\frac{4gD_d}{3\ C_D}} \tag{5-7}$$

In practice, the Souders and Brown design coefficient (K_{SB}) depends primarily on the type of mist extractor present, separator geometry, flow rates, and fluid properties. Therefore, K_{SB} is usually determined from experiments. A well-known source of empirical K_{SB} factors for mist pads is the *GPSA Engineering Data Book* (2004). The GPSA's K_{SB} factors have been curve fitted and are given as

$$K_{SB} = 0.35 - 0.0001\ (P - 100) \tag{5-8}$$

Where K_{SB} is in ft/sec, and P is the separator pressure in psig.

Also, the factor K_{SB} should be adjusted as follows:

1. For most vapors under vacuum, $K_{SB} = 0.20$.
2. For glycols and amine solutions, multiply K_{SB} values by 0.6–0.8.
3. For compressor suction scrubbers and expander inlet separators, multiply K_{SB} by 0.7–0.8.

Maximum terminal velocities calculated using the K_{SB} factors are for separators normally having a wire-mesh mist extractor and should allow all liquid droplets larger than 10 μm to settle out of the gas. If no mist extractor is present, multiply K_{SB} by 0.5.

It is often necessary to separate two immiscible liquids, the light and heavy phases, and a vapor. Typical examples in the petroleum industry are the separation of water, and a hydrocarbon liquid and vapor. For this system, the flow of rising light droplets in the heavy liquid phase or settling heavy droplets in the light liquid phase is considered laminar and is governed by Stokes law (Monnery and Svrcek, 1994):

$$V_t = \frac{K_s(\rho_H - \rho_L)}{\mu} \tag{5-9}$$

$$K_s = 2.06151 \times 10^{-5}\ D_d^2 \tag{5-10}$$

where V_t is in inch/min; densities of light and heavy liquid phases (ρ_L, ρ_H) are in lb/ft^3; viscosity is in cP; and D_d is in microns (1 μm = 3.28084×10^{-6} ft).

As can be seen from Equation 5-9, the settling velocity of a droplet is inversely proportional to the viscosity of the continuous phase. Hence, it requires more time for the droplets to settle out of the continuous phase with greater viscosity. In practice, V_t is typically limited to 10 inch/min.

5.2.4 Design Considerations

The following factors must be determined before beginning separator design:

- Gas and liquid flow rates (minimum, average, and peak).
- Operating and design pressures and temperatures.
- Surging or slugging tendencies of the feed streams.
- Physical properties of the fluids such as density, viscosity, and compressibility.
- Designed degree of separation (e.g., removing 100% of particles greater than 10 μm).

Consideration for the future life of the field should also be included. For example, most North Sea separators were designed for high oil cuts, but are now high water cuts, which produces a lot of nozzle problems. In the separator design, it is also worthwhile to clarify two definitions: holdup and surge times. Holdup is defined as the time it takes to reduce the liquid level from normal (NLL) to low (LLL) while maintaining a normal outlet flow without feed makeup. Surge time is defined as the time it takes for the liquid level to rise from normal (NLL) to high (HLL) while maintaining a normal feed without any outlet flow. Holdup time is based on the reserve required to maintain good control and safe operation of downstream facilities, whereas surge time is usually based on requirements to accumulate liquid as a result of upstream or downstream variations or upsets, for example, slugs. In the absence of specific requirements, surge time may be taken as one-half of holdup time. Table 5-1 shows typical values of holdup time, t_H, and surge time, t_S (Svrcek and Monnery, 1993).

Table 5-1 Typical Values of Holdup and Surge Times

Service	t_H, min	t_S, min
A. Unit feed drum	10	5
B. Separators		
1. Feed to column	5	3
2. Feed to other drum or tankage		
with pump or through exchanger	5	2
without pump	2	1
3. Feed to fired heater	10	3

5.2.5 Design Procedure

The initial design and calculation of gravity separators are discussed in many books and basic references (Kumar, 1987; Campbell, 1992; Arnold and Stewart, 1998). However, a more accurate and detailed sizing of two- and three-phase separators can be performed by using the design methods developed by Svrcek and Monnery (1993) and Monnery and Svrcek (1994), which have been well received by the industry worldwide. These procedures are a result of a review of literature sources and accepted industrial design guidelines and allow the production facility engineer to choose the detailed sizing parameters of the two- and three-phase separators. The following sections attempt to address the basics of three-phase separator design and provide step-by-step procedures for three-phase vapor/liquid/liquid separator design. To add a degree of conservatism to the design, we ignored the volume available in the heads. It is reasonably common nowadays for vessel sizing to be subcontracted out either directly to the vessel supplier or sometimes to the internals supplier.

5.2.5.1 Vertical Separators

For a three-phase vertical separator, the total height can be broken into different sections, as shown in Figure 5-3. The separator height is then calculated by adding the heights of these sections. If a mist eliminator pad is used, additional height is added.

The calculations of diameter and height are detailed as follows:

1. Calculate the vertical terminal velocity using Equation 5-6 and set $V_V = 0.75\ V_t$ for a conservative design.
2. Calculate the vapor volumetric flow rate:

$$Q_V = \frac{\dot{m}_V}{3600\rho_V} \tag{5-11}$$

3. Calculate the vessel internal diameter, D_i;

$$D_i = \left(\frac{4\ Q_V}{\pi V_V}\right)^{0.5} \tag{5-12}$$

   If there is a mist eliminator, add 3–6 inches to D_i and round up to the next 6-inch increment to obtain D; if there is no mist eliminator, $D = D_i$.

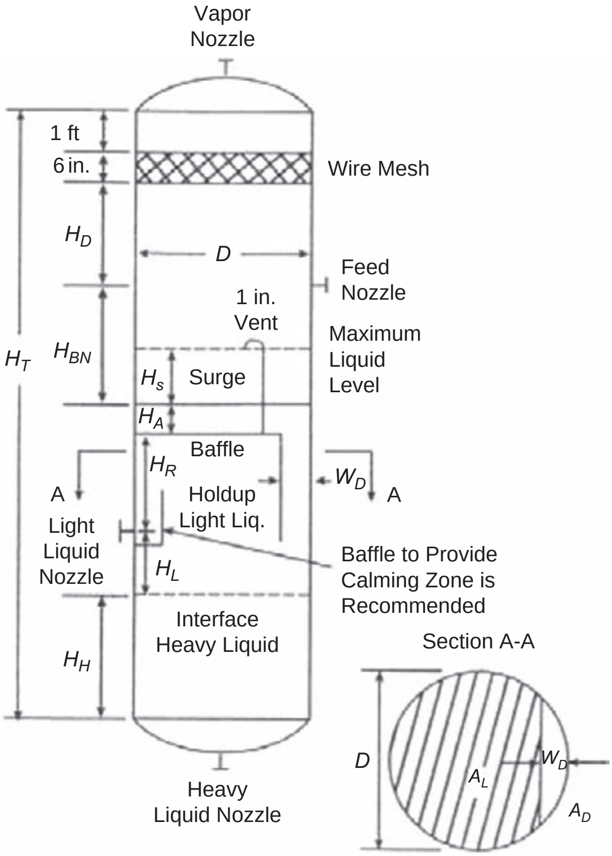

Figure 5-3 Basic design of three-phase vertical separator (Monnery and Svrcek, 1994).

4. Calculate the settling velocity of the heavy liquid out of the light liquid using Equation 5-9 (with $\mu = \mu_L$).
5. Calculate the rising velocity of the light liquid out of the heavy liquid phase using Equation 5-9 (with $\mu = \mu_H$).
6. Calculate the light and heavy liquid volumetric flow rates, Q_{LL} and Q_{HL}:

$$Q_{LL} = \frac{\dot{m}_{LL}}{60\ \rho_L} \tag{5-13}$$

$$Q_{HL} = \frac{\dot{m}_{HL}}{60\ \rho_H} \tag{5-14}$$

7. Assume $H_L = 1$ ft (minimum) and $H_H = 1$ ft (minimum) and calculate the settling times for the heavy liquid droplets to settle ($t_{s,HL}$) and for the light liquid droplets to rise ($t_{s,LL}$) through this section, respectively, as

$$t_{s,HL} = \frac{12\,H_L}{V_{HL}} \tag{5-15}$$

$$t_{s,LL} = \frac{12\,H_H}{V_{LL}} \tag{5-16}$$

where the settling velocity of heavy liquid droplets out of light liquid (V_{HL}) and the rising velocity of light liquid droplets out of heavy liquid (V_{LL}) are in inch/min.

8. If there is a baffle plate, calculate the area, A_L, as

$$A_L = A - A_D \tag{5-17}$$

where A is the vertical vessel cross-sectional area, ft^2; and A_D is the downcomer cross-sectional area, ft^2.

In the preceding equation, the larger value of A_D that can be calculated from the two following equations should be used:

$$A_D = \left(\frac{7.48\text{ gal}}{\text{ft}^3}\right)\left(\frac{60\text{ min}}{1\text{hr}}\right)\left(\frac{Q_{LL} + Q_{HL}}{G}\right) \tag{5-18}$$

where baffle liquid load (G) will be obtained from the chart in Figure 5-4.

$$\frac{A_D}{A} = \frac{-4.755930 \times 10^{-5} + 0.174875\left(\frac{W_D}{W}\right) + 5.668973\left(\frac{W_D}{W}\right)^2 - 4.916411\left(\frac{W_D}{W}\right)^3 - 0.145348\left(\frac{W_D}{W}\right)^4}{1.0 + 3.924091\left(\frac{W_D}{W}\right) - 6.358805\left(\frac{W_D}{W}\right)^2 + 4.018448\left(\frac{W_D}{W}\right)^3 - 1.801705\left(\frac{W_D}{W}\right)^4} \tag{5-19}$$

where downcomer chord width (W_D) is assumed to be 4 inch.

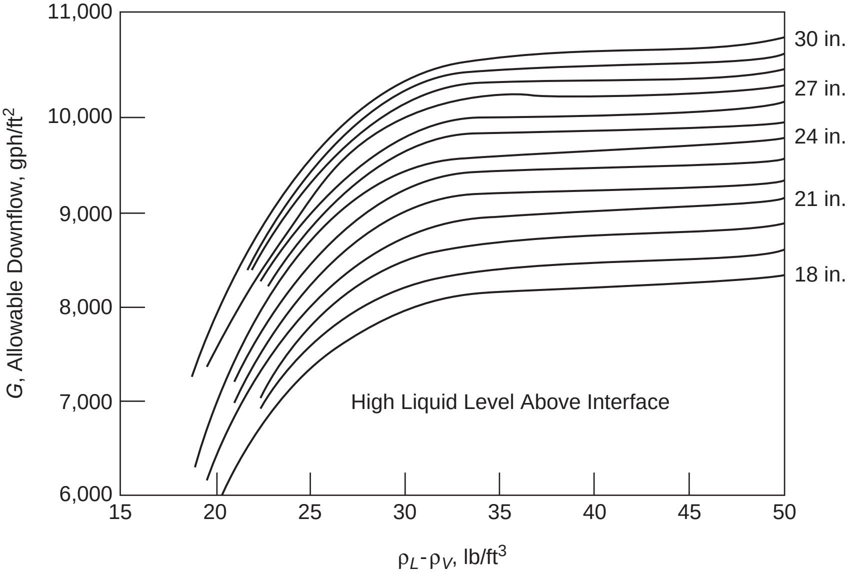

Figure 5-4 Determining the downcomer allowable flow (Monnery and Svrcek, 1994).

In Figure 5-4 the high liquid level above interface is equal to $H_L + H_R$, where the height from the light liquid nozzle to baffle (H_R) is assumed to be 9 inch as a minimum.

9. Calculate the residence time (t_r) of each phase based on the volumes occupied by the light and heavy phases as

$$t_{r,LL} = \frac{H_L\ A_L}{Q_{LL}},\quad \text{min} \tag{5-20}$$

$$t_{r,HL} = \frac{H_H\ A_H}{Q_{HL}},\quad \text{min} \tag{5-21}$$

If $t_{r,LL} \langle t_{s,HL}$ or $t_{r,HL} \langle t_{s,LL}$, increase the diameter and repeat the procedure from step 7 (liquid separation is controlling). Note that $A_H = A$.

10. Calculate H_R based on the required holdup time (t_H) as

$$H_R = \frac{Q_{LL}\ t_H}{A_L} \tag{5-22}$$

Check this value with that assumed in step 8 to ensure that the assumed value is reasonable. If surge is not specified, calculate the surge height (H_S) based on surge time (t_S):

$$H_S = \frac{t_S(Q_{LL} + Q_{HL})}{A} \tag{5-23}$$

where the minimum H_S is 6 inch.

11. Calculate the vessel total height (H_T) as

$$H_T = H_H + H_L + H_R + H_A + H_{BN} + H_D \tag{5-24}$$

where H_A is liquid level above baffle, which is 6 inch (minimum); and H_{BN} is liquid height from above baffle to feed nozzle, ft

$$H_{BN} = \frac{1}{2}(D_N, \text{ft}) + \text{greater of } (2\text{ft or } H_S + 0.5 \text{ ft})$$

where the nozzle diameter (D_N) is calculated using the following criterion:

$$D_N \geq \left[\frac{4\ Q_M}{\pi 60/\sqrt{\rho_M}}\right]^{0.5} \tag{5-25}$$

Q_M and ρ_M are volumetric flow rate and no-slip density of vapor/liquid mixture, respectively.

H_D is disengagement height (ft) = 0.5D or a minimum of

36 inch + ½(D_N); without mist eliminator
24 inch + ½(D_N); with mist eliminator

If a mist eliminator pad is used, additional height is added, as shown in Figure 5-3.

5.2.5.2 Horizontal Separator

There are different types of horizontal three-phase separators. However, the most common type includes a boot, as shown in Figure 5-5, which is a better design for the system of a small amount of heavy liquid (<15%–20% of total liquid by weight) and large amount of vapor. In this section, only design procedure for this type of separator is presented. Readers are referred to the original paper of Monnery and Svrcek (1994) for the detailed design of other types of horizontal three-phase separators.

The horizontal design procedure incorporates optimizing the diameter and length by minimizing the approximate weight of the shell and heads.

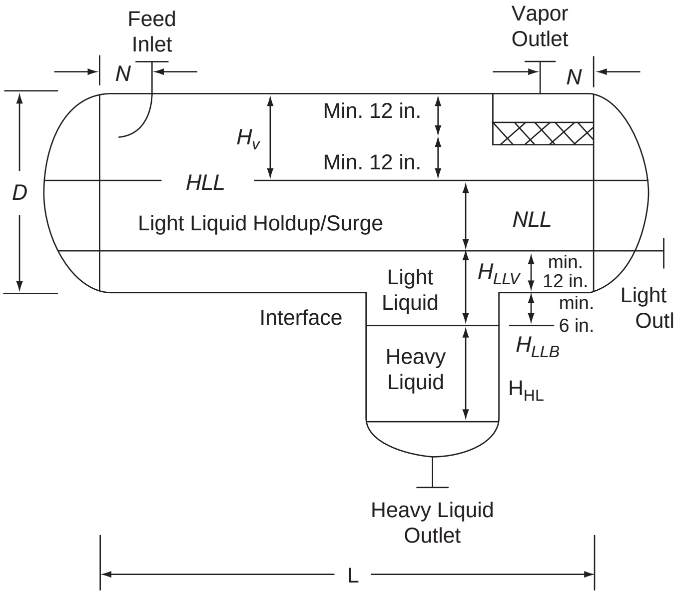

Figure 5-5 Basic design of horizontal three-phase separator with boot (Monnery and Svrcek, 1994).

A stepwise procedure for designing horizontal three-phase separators with a boot is as follows (Monnery and Svrcek, 1994):

1. Calculate the vapor volumetric flow rate, Q_V, using Equation 5-11.
2. Calculate the light and heavy liquid volumetric flow rates, Q_{LL} and Q_{HL}, per Equations 5-13 and 5-14.
3. Calculate the vertical terminal velocity, V_t, using Equation 5-6 and set $V_V = 0.75\ V_t$.
4. Select holdup and surge times from Table 5-1 and calculate the holdup and surge volumes, V_H and V_S, from the following equations (unless surge is otherwise specified, such as slug volume):

$$V_H = t_H \times Q_L \tag{5-26}$$

$$V_S = t_S \times Q_L \tag{5-27}$$

5. Obtain L/D from Table 5-2 and initially calculate the diameter according to

$$D = \left[\frac{4(V_H + V_S)}{0.6\pi\left(\frac{L}{D}\right)}\right]^{\frac{1}{3}},\ \text{ft} \tag{5-28}$$

Then calculate the total cross-sectional area, $A = \frac{\pi D^2}{4}$.

Table 5-2 *L/D* Ratio Guidelines

Vessel Operating Pressure, psig	***L/D***
$0 < P \leq 250$	1.5–3.0
$250 < P \leq 500$	3.0–4.0
$P > 500$	4.0–6.0

6. Set the vapor space height, H_V, to the larger of $0.2D$ or 2 ft (1 ft if there is no mist eliminator). Using $\frac{H_V}{D}$ in Equation 5-19 $\left(\text{with replacing } \frac{W_D}{W}\right)$, obtain $\frac{A_V}{A}$ and calculate A_V.
7. Set the light liquid heights in the vessel and boot, H_{LLV} and H_{LLB}.
8. Calculate the cross-sectional area of the light liquid above the bottom of the vessel, A_{LLV}, using $\frac{H_{LLV}}{D}$ $\left(\text{instead of } \frac{W_D}{W}\right)$ in Equation 5-19.
9. Calculate the minimum length to accommodate the liquid holdup/surge:

$$L_1 = \frac{V_H + V_S}{A - A_V - A_{LLV}} \tag{5-29}$$

10. Calculate the liquid dropout time, ϕ, using the following equation:

$$\phi = \frac{H_V}{V_V}\text{, sec} \tag{5-30}$$

11. Calculate the actual vapor velocity, V_{AV}, as

$$V_{AV} = \frac{Q_V}{A_V}\text{, ft/sec} \tag{5-31}$$

12. Calculate the minimum length required for vapor/liquid separation, L_2, as

$$L_2 = V_{AV} \times \phi\text{, ft} \tag{5-32}$$

13. If $L_1 < L_2$, then set $L_1 = L_2$ (vapor/liquid separation controls). This simply results in some extra holdup and residence time. If $L_1 << L_2$, then increase H_V and recalculate A_V; then repeat from step 9. If $L_1 > L_2$, the design is acceptable for vapor/liquid separation. If $L_1 >> L_2$, liquid holdup controls. L_1 can be reduced and L_2 increased only if H_V is reduced. H_V may only be reduced if it is greater than the minimum specified in step 6. With reduced H_V, recalculate A_V and repeat from step 9.
14. Calculate the settling velocity of the heavy liquid out of the light liquid phase, V_{HL}, as

$$V_{HL} = \frac{K_S(\rho_H - \rho_L)}{\mu_L} \tag{5-33}$$

where K_S is obtained from Equation 5-10.

15. Calculate the settling time of the heavy liquid out of the light liquid phase as

$$t_{s,\mathrm{HL}} = \frac{12(H_{\mathrm{LLB}} + D - H_{\mathrm{V}})}{V_{\mathrm{HL}}} \tag{5-34}$$

16. Calculate the residence time of the light liquid as

$$t_{r,\mathrm{LL}} = \frac{(A - A_{\mathrm{V}})\,L}{Q_{\mathrm{LL}}} \tag{5-35}$$

This volume of light liquid ignores the light liquid volume in the boot.

17. If $t_{r,\mathrm{LL}} < t_{s,\mathrm{HL}}$, then increase the vessel length (liquid separation controls) as

$$L = \frac{t_{s,\mathrm{HL}} \times Q_{\mathrm{LL}}}{A - A_{\mathrm{V}}} \tag{5-36}$$

18. Calculate L/D. If $L/D << 1.5$, then decrease D (unless it is already at a minimum), and if $L/D >> 6.0$, then increase D; repeat from step 5.
19. Calculate the wall thickness and surface area of the shell and heads, and approximate vessel weight according to Table 5-3.
20. Increase or decrease the vessel diameter by 6-inch increments and repeat the calculations until L/D ranges from 1.5 to 6.0.

Table 5-3 Wall Thickness, Surface Area, and Approximate Vessel Weight

Component	**Wall Thickness, inch**	**Surface Area, ft^2**
Shell	$\dfrac{PD}{2SE - 1.2P} + \sigma_C$	πDL
2:1 Elliptical Heads	$\dfrac{PD}{2SE - 0.2P} + \sigma_C$	$1.09\,D^2$
Hemispherical Heads	$\dfrac{PD}{4SE - 0.4P} + \sigma_C$	$1.571\,D^2$
Dished Heads	$\dfrac{0.885\,PD}{SE - 0.1P} + \sigma_C$	$0.842\,D^2$
Approximate Vessel Weight	$W = \left(\dfrac{490\,\mathrm{lb}}{\mathrm{ft}^3}\right)\left(\dfrac{\sigma}{12}\right)(A_{\mathrm{Shell}} + 2A_{\mathrm{Head}})$	

Notes: The design pressure (P, psig) is typically either the operating pressure with 15–30 psi added to it or the operating pressure +10%, whichever is greater. For the allowable stress, S, refer to ASME Pressure Vessel Code (1986). The joint efficiency, E, ranges from 0.6 to 1.0. The corrosion allowance, (σ_C), typically ranges from 1⁄16 to 1⁄8 inch.

21. With the optimum vessel size (minimum weight), calculate the normal and high liquid levels (H_{NLL} and H_{HLL}) as

$$H_{HLL} = D - H_V \tag{5-37}$$

$$A_{NLL} = A_{LLV} + \frac{V_H}{L} \tag{5-38}$$

Determine H_{NLL} using Equation 5-19 from $\frac{A_{NLL}}{A}$ (replacing W_D with H_{NLL} and A_{NLL} with A_D).

22. Design the heavy liquid boot. Set the height of the heavy liquid, H_{HL}; calculate the rising velocity of the light liquid out of the heavy liquid phase (V_{LL}) as

$$V_{LL} = \frac{K_s(\rho_H - \rho_L)}{\mu_H} \tag{5-39}$$

Set boot velocity (V_B) to 0.75 V_{LL} and calculate the heavy liquid boot diameter, D_B, as

$$D_B = \left(\frac{4 \times 12\ Q_{HL}}{\pi V_B}\right)^{0.5},\ \text{ft} \tag{5-40}$$

Then calculate the settling time of the light liquid ($t_{s,LL}$) out of the heavy liquid phase using Equation 5-16.

23. Calculate the residence time of the heavy liquid ($t_{r,HL}$) in the boot as

$$\left(t_{r,HL}\right)_B = \frac{\left(\pi D_B^2\right) H_{HL}}{4\ Q_{HL}} \tag{5-41}$$

If $(t_{r,HL})_B < t_{s,LL}$, then increase the boot diameter.

5.2.6 Practical Separator Design

The biggest development in recent years is the widespread recognition that the actual performance of a separator may fall far short of the theoretical performance, due to the actual flow patterns within the vessel being far from the ideal (although strictly this is a "rediscovery"). It has, however, been helped by two visualization techniques: computational fluid dynamics (CFD) and physical modeling, which vividly show what can go wrong and how to correct it.

The most important areas to ensure a separator performs to design are

- Correct inlet nozzle sizing and a good inlet device (momentum breaker). The two main sorts are the "vane" inlet (schoepentoeter, or the many proprietary versions of this) and the cyclonic inlet.

- Primary fluid distribution–distribution plates to translate the reduced but still high velocities from the inlet device into quiescent flows in a liquid–liquid separator body, or distribution plates either side of a vane pack (downstream is best as upstream ones shatter droplets unnecessarily) or other gas demister.
- Intermediate fluid distribution when necessary (e.g., to align flow before entry to a plate pack or where a separator has an unusual shape or aspect ratio).
- Exit devices: vortex breakers and anti-liquid-pickup details.

The methods of verifying performance "before" and "after" the introduction of good practice such as these internals are CFD and physical modeling, which uses dimensionless number similarity to model at a reduced scale, with air-water, air-oil-water, or gas-oil-water. The more advanced centers have pressure flow loops that can run on oil and gas at pressure but have the drawback that one cannot view what is happening inside the vessel in the way a perspective model allows.

5.3 MULTISTAGE SEPARATION

To achieve good separation between the gas and liquid phases and maximize the hydrocarbon liquid recovery, one needs to use several separation stages at decreasing pressures in which the well stream is passed through two or more separators arranged in series. The operating pressures are sequentially reduced; therefore, the highest pressure is found at the first separator and the lowest pressure at the final separator. In practice, the number of stages normally ranges between two and four, which depends on the gas/oil ratio (GOR) and the well stream pressure. Two stages of separation are usually used for low GOR and low well stream pressure; three stages of separation are used for medium to high GOR and intermediate inlet pressure; and four stages of separation are used for high GOR and high well stream pressure. Note that three-stage separation usually represents the economic optimum, where it allows 2%–12% higher liquid recovery in comparison with two-stage separation and, in some cases, recoveries up to 25% higher (Rojey et al., 1997). To recover the gases fractions produced in the separators operating at medium pressure and at low pressure, one needs to recompress them to the pressure of a high-pressure separator. However, for an associated gas, recompression is sometimes considered too costly; hence, the gas produced from the low pressure separator may be flared.

It should be noted that the main objective of stage separation is to provide maximum stabilization to the resultant phases (gas and liquid) leaving the final separator, which means the considerable amounts of gas or liquid will not evolve from the final liquid and gas phases, respectively. The quantities of gas and liquid recovered at a given pressure are determined by equilibrium flash calculations using an appropriate equation of state (EOS). This helps to optimize the value of pressure that is set for each separator. The pressures are often staged so that the ratio of the pressures in each stage is constant. Therefore, if the pressure in the first separator (which is normally fixed by specification or economics) and the pressure in the final separator (which will be near the atmospheric pressure) are known, the pressure in each stage can be determined.

5.4 CENTRIFUGAL SEPARATORS

In centrifugal or cyclone separators, centrifugal forces act on a droplet at forces several times greater than gravity as the droplet enters a cylindrical separator. This centrifugal force can range from 5 times gravity in a large, low-velocity unit to 2,000 times gravity in small, high-pressure units. Generally, centrifugal separators are used for removing droplets greater than 100 μm in diameter, and a properly sized centrifugal separator can have a reasonable removal efficiency of droplet sizes as low as 10 μm.

Centrifugal separators are also extremely useful for gas streams with high particulate loading (Talavera, 1990). Such equipment has already been studied; however, a simple, compact, and lightweight gas-liquid cylindrical cyclone (GLCC) separator, recently developed by Tulsa University, requires little maintenance and is easy to install and operate. The compact dimensions, smaller footprint, and lower weight of the GLCC have a potential for cost savings to the industry, especially in offshore applications. Also, the GLCC reduces the inventory of hydrocarbons significantly, which is critical to environmental and safety considerations. The GLCC separator, used mainly for bulk gas/liquid separation, can be designed for various levels of expected performance. Typical performance levels from the GLCC separator are 0.5–2.0 gallons of liquid per MMscf in the gas outlet and 0%–5% gas in the liquid outlet (NATCO, 2002). The design, control system studies, experimental investigations, and field applications of GLCC separators are discussed in greater detail by Gomez et al. (2000), Mohan and Shoham (2003), and Wang et al. (2003).

5.5 TWISTER SUPERSONIC SEPARATOR

The Twister supersonic separator is a unique combination of known physical processes, combining the expansion, cyclonic gas/liquid separation, and recompression process steps in a compact, tubular device to condense and separate water and heavy hydrocarbons from natural gas. Condensation and separation at supersonic velocity are key to achieving step-change reductions in both capital and operating costs. The residence time inside the Twister supersonic separator is only milliseconds, allowing hydrates no time to form and avoiding the requirement for hydrate inhibition chemicals. The elimination of the associated chemical regeneration systems avoids harmful benzene, toluene, and xylene (BTX) emissions to the environment or the expense of chemical recovery systems. The simplicity and reliability of a static device with no rotating parts that operates without chemicals ensures a simple facility with a high availability, suitable for unmanned operation in harsh and/or offshore environments. In addition, the compact and low-weight Twister system design enables debottlenecking of existing space and weight-constrained platforms. The first commercial application at the Shell-operated B11 platform offshore Malaysia proves the technical and commercial maturity of Twister supersonic separator, which is a major milestone toward industry acceptance of this innovative technology (Brouwer et al., 2004). However, the feasibility and benefits of this technology must be studied for a specific gas development. The system design of the Twister supersonic separator is discussed in greater detail by Okimoto and Brouwer (2002) and Brouwer and Epsom (2003).

5.6 SLUG CATCHERS

Slug catchers are used at the terminus of large gas/condensate transmission pipelines to catch large slugs of liquid in pipelines, to hold these slugs temporarily, and then to allow them to follow into downstream equipment and facilities at a rate at which the liquid can be properly handled. In fact, the slug catcher provides temporary storage for any surges (slugs) in liquid flows and roughly separates the gas from the liquids. The gas then exits the top of the slug catcher and flows to the plant inlet separator via a pressure control valve, which reduces the pressure of the gas and further condenses water and some of the heavier hydrocarbons.

Slug catchers may be either of the vessel or of the manifolded pipe type. A vessel-type slug catcher is essentially a knockout vessel. This type is simple in design and maintenance. A pipe-type slug catcher consists of several long pieces of pipe (fingers), which together form the buffer volume to store the largest slugs expected from the upstream system. Vessel-type slug catchers can be used only if the incoming liquid volume is small. When larger liquid volumes have to be accommodated, say of more than 100 m^3 (3,531 ft^3), the pipe-type slug catcher should be used (Shell DEP 31.40.10.12–Gen, 1998). Pipe-type slug catchers are frequently less expensive than vessel-type slug catchers of the same capacity due to thinner wall requirements of smaller pipe diameter. The manifold nature of multiple pipe-type slug catchers also makes possible the later addition of further capacity by laying more parallel pipes. Because the pipe-type slug catcher is defined as a piping configuration rather than a pressure vessel, it is not constrained to the same requirements as a normal vessel. However, due to its large size, it will contain the majority of high-pressure hydrocarbon gas on the site. It is therefore recommended that the slug catcher be automatically depressurized (for prevention of fire) as quickly as possible without imposing unusually high flow rates on the flare system.

A schematic of a pipe-type slug catcher is shown in Figure 5-6. The general configuration consists of the following parts:

- Fingers with dual slope and three distinct sections: gas/liquid separation, intermediate, and storage sections.
- Gas risers connected to each finger at the transition zone between the separation and intermediate sections.
- Gas equalization lines located on each finger. These lines are located within the slug storage section.
- Liquid header collecting liquid from each finger. This header will not be sloped and is configured perpendicular to the fingers.

Note that it has been assumed that all liquids (condensate and water) are collected and sent to an inlet three-phase separator although it is possible to separate condensate and water directly at the fingers. When doing condensate/water separation at the slug catcher itself, one has to allow separately for the maximum condensate slug and the maximum water slug in order to ensure continuous-level control.

Separation of gas and liquid phases is achieved in the first section of the fingers. The length of this section will promote a stratified flow pattern and permit primary separation to occur. Ideally, liquid droplets, 600 μm and below, will be removed from the gas disengaged into the gas risers, which are located at the end of this section. The length of the intermediate section is minimal such that there is no liquid level beneath the gas riser when the slug catcher is full; i.e., the storage section is completely full. This section is

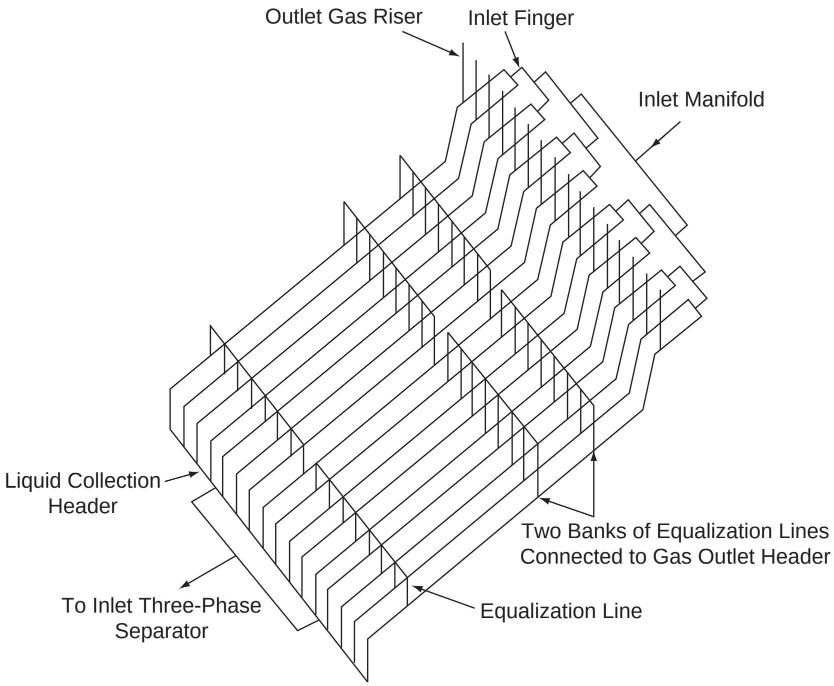

Figure 5-6 Three-dimensional rendering of finger-type slug catcher.

composed of a change in elevation between the gas risers and the storage section that allows a clear distinction between the liquid and gas phases. The length of the storage section ensures that the maximum slug volume can be retained without liquid carryover in the gas outlet. During normal operations, the normal liquid level is kept at around the top of the riser from each finger into the main liquid collection header, which is equivalent to approximately 5 minutes' operation of the condensate stabilization units at maximum capacity.

Because the slug catcher is the first element in the gas processing plant, determining its proper size is vital to the operation of the entire plant. In fact, if more liquid is brought in than the slug catcher can handle, the plant normally shuts down. Therefore, slug sizing results should always be treated with caution, and slug catchers should be designed with an ample design margin (Burke and Kashou, 1996; Xiao and Shoup, 1998). Shell DEP 31.40.10.12–Gen (1998) specifies requirements and gives recommendations for the design of multiple-pipe slug catchers. However, vendors should be requested to provide detailed design guidelines.

5.7 HIGH-EFFICIENCY LIQUID-GAS COALESCERS

Aerosols in gas streams can often be less than 5 μm in size and require the use of special separation equipment. High-efficiency liquid-gas coalescers have been applied effectively for the removal of fine aerosols in gas production, processing, and transmission. Coalescers are typically constructed as cartridges that use pleated-glass fiber media supported by a metal core. The coalescer cartridges are then placed in a housing that controls the inlet/outlet gas velocities to ensure good separation and prevent any reentrainment of liquids. The coalescer media contain a much finer pore structure and larger surface area as compared to traditional separators that often use mesh pads or vane pack internals.

5.7.1 Aerosols

Aerosols are formed by three mechanisms: condensation, atomization, and entrainment. The relative sizes produced by these formation mechanisms are shown in Figure 5-7.

Aerosols formed by condensation of a vapor into a liquid are the smallest and most difficult to remove contaminants having a size distribution in the range of 0.2 to 5 μm. Atomization creates aerosol drops by breaking up larger liquid drops through mechanical shear such as passing through a constriction in a valve under a high velocity. Atomization forms aerosols in the size range of 10 to 200 μm. Entrainment involves the movement of liquid slugs along pipelines, and here the liquid drop sizes are very large, from 500

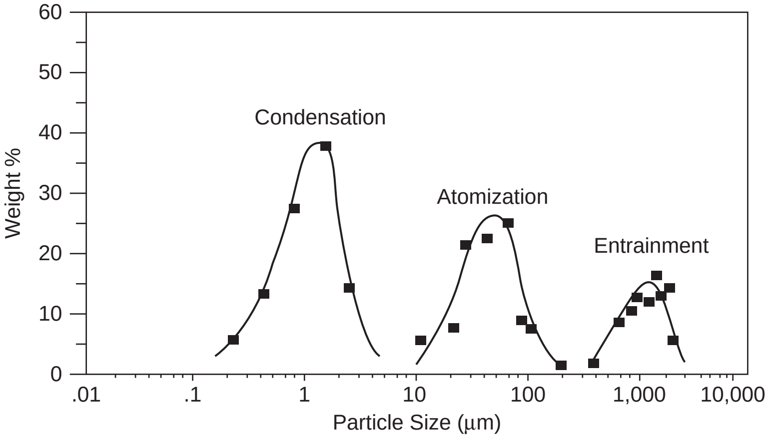

Figure 5-7 Aerosol types (Brown et al., 1994).

to 5,000 μm. All three types of aerosol liquids are commonly found in gas systems. High-efficiency liquid-gas coalescers can effectively remove the fine aerosols created by the condensation mechanism.

5.7.2 Coalescer Construction/Operation Principles

High-efficiency liquid-gas coalescers are generally constructed from glass fibers since this material allows for a fine porous structure with fiber diameters of a few microns. The small pore size is needed to achieve greater capture and separation of these small aerosols. The primary rationale for the use of high-efficiency coalescers is that significant aerosol contaminant exists in the plants that are in the submicron and low-micron size range (Brown et al., 1994).

This type of liquid-gas coalescer can operate at significantly lower flow rates than the initial design flow rate and therefore can tolerate a high turndown ratio. This is due to the separation mechanisms for coalescing that are primarily diffusion and direct interception, unlike vane separators and mesh pads that rely heavily on inertial separation principles. This allows the high-efficiency liquid-gas coalescer systems a greater degree of flexibility, and they can operate at peak performance even for high turndown ratios (reduced flow rates), which can occur during commonly encountered partial plant shutdowns and upset conditions.

The use of a surface treatment (Miller et al., 1988) on high-performance vertical liquid-gas coalescer cartridge systems has been proven to significantly enhance performance by allowing higher flow rates or smaller housing diameters compared to untreated coalescers. The surface treatment alters the properties of the coalescer medium so that it will not wet out with oil- or water-based fluids. The treatment has also been found to extend the service of the coalescer by reducing fouling and also to lower the saturated pressure drop. A Pall vertical high-efficiency liquid-gas coalescer system is depicted in Figure 5-8.

The inlet gas with liquid aerosol contamination first enters at the bottom of the housing into a first-stage knockout section. Here, any slugs or larger-size droplets (something less than 300 μm) are removed by gravitational settling. The gas then travels upward through a tube sheet and flows radially from the inside of the cartridges through the coalescer medium to the annulus. The inlet aerosol distribution is in the size range of 0.1–300 μm and after passing through the coalescer medium is transformed to enlarged coalesced droplets in the size range of 0.5–2.2 mm. The advantage of flowing from the inside to the outside of the coalescer cartridge is that the gas

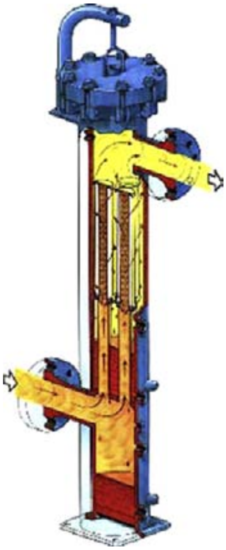

Figure 5-8 High-efficiency vertical liquid-gas coalescer system (Schlotthauer and Hashemi, 1991; Wines, 2004).

velocity can be more easily adjusted in the annulus by selecting the optimum housing diameter to prevent reentrainment of coalesced droplets.

As the gas leaves the coalescer cartridge and travels upward in the annulus, it contributes to the total flow, thereby increasing the annular velocity. The annular velocity is modeled as a linear function with vertical distance, and the annular velocity is zero at the bottom of the cartridge and increases to a maximum value at the top of the cartridge.

Once the coalesced droplets are formed, they immediately drain vertically downward in the coalescer medium pack. The surface treatment greatly enhances this drainage, and as a direct consequence of the treatment, the coalesced droplets are shielded from the upward gas flow in the annulus in the upper two-thirds section of the coalescer cartridge. The coalesced droplets are first exposed to the annular gas flow when they appear on the external face of the coalescer medium pack at the bottom third of the coalescer cartridge (see Figure 5-9). Once the coalesced droplets are released to the annular space, they are subject to the force of the upward-flowing gas. The trajectory of the coalesced droplets is modeled on a force balance between gravity settling and the drag force created by the gas flow past the droplets. This analysis leads to the calculation of a critical annular velocity for reentrainment.

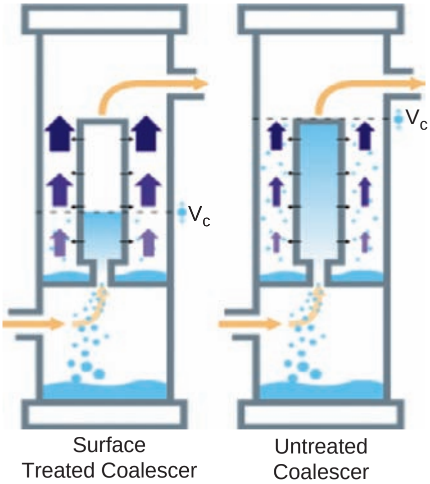

Figure 5-9 Effect of surface treatment on coalescer drainage (Wines, 2004).

Due to the surface treatment, there are minimal coalesced droplets present in the annulus above the drainage point at the bottom third of the coalescer cartridge. For a coalescer cartridge that is not specially surface treated, the coalesced liquids are present throughout the length of the coalescer in the annulus space, and the critical annular velocity for reentrainment is given for the top of the element (see Figure 5-9). For the treated coalescer, it is allowable to have annular velocities greater than the critical value for reentrainment in the portion of the annulus space where there are no liquids present. This allows the maximum annular velocity at the top of the coalescer cartridge to be about three times the critical reentrainment value needed at the vertical position of the lower one-third of the cartridge height where liquids are present.

Therefore, the maximum annular velocity at the top of the coalescer cartridge is found to be about three times greater than the value for an untreated coalescer. The annular area is determined using the maximum allowable annular velocity and designed to be of sufficient size to prevent reentrainment and as small as possible to minimize the housing diameter.

5.7.3 Modeling the Liquid-Gas Coalescer

The modeling of the liquid-gas coalescer system can be divided into two basic aspects for performance: media velocity and annular velocity. The other consideration to be taken into account is pressure drop. The pressure drop for a given system can be decreased by using more coalescer elements.

5.7.3.1 Media Velocity

The media velocity (V_{med}) is defined as the actual flow rate divided by the coalescer filter area (Wines, 2004):

$$V_{\mathrm{med}} = \frac{Q_a}{N\, A_{\mathrm{med}}} \tag{5-42}$$

where N is the number of coalescers; A_{med} is the media area for one coalescer; Q_a is the actual system flow rate at system conditions and is obtained from the standard system flow rate (Q_s) as

$$Q_a = \frac{Q_s(\mathrm{SG})\rho_{\mathrm{Air,stp}}}{\rho_G} \tag{5-43}$$

where SG is gas specific gravity; $\rho_{\mathrm{Air,stp}}$ is the density of air at standard temperature and pressure; and ρ_G is the density of gas at system conditions.

The media velocity is not the actual velocity through the open pores of the media, but rather an average by convention over the combined pore area and solid matrix area in the spatial plane normal to the flow direction. The maximum media velocity for a coalescer construction is related to a number of factors intrinsic to the particular coalescer design and to the physical properties of the system. Four steps have been identified with the mechanism of the formation and removal of droplets in the coalescer medium: (1) capture, (2) coalescing, (3) release, and (4) drainage.

The formation of the coalesced droplets first involves the capture of the small aerosols onto the fibers of the coalescer medium. The actual coalescing or merging of the fine droplets is believed to take place on the fibers and especially at fiber intersections. The coalesced droplets are then released from the fiber due to the drag force of the gas flow exceeding the adsorption energy. This process is repeated through the depth of the coalescer medium until the coalescing process is completed and the largest possible stable droplet size is achieved. During the coalescing stages, the growing droplets also drain downward inside the media pack due to the force of gravity.

The surface treatment allows the release and drainage process to proceed at a faster rate, which, in turn, frees up more coalescing sites on the fibers and allows the coalescer to process higher inlet liquid aerosol concentrations than the untreated coalescer medium.

5.7.3.2 Effect of System Conditions on Media Velocity

The ability of the coalescer medium to perform effectively will also depend on the system environment. While different coalescer constructions will exhibit quantitative differences, they will follow the same qualitative behavior. The media velocity has been determined to depend on system parameters such as inlet aerosol concentration, aerosol density, gas density, and gas viscosity.

At low aerosol concentrations, the maximum media velocity is constant and is unaffected by aerosol levels. Under these conditions, the media are limited by the capture mechanism and are not affected by drainage. At higher levels of aerosol concentration, the coalescer medium becomes limited by drainage and is inversely proportional to the aerosol concentration. The effect of the surface treatment on this process is to enhance the drainage and allow for higher maximum media velocities under the same aerosol loading when limited by drainage.

5.7.3.3 Annular Velocity

The annular velocity (V_{ann}) is defined as the actual flow rate divided by the annulus area (Wines, 2004):

$$V_{\mathrm{ann}} = \frac{Q_{\mathrm{a}}}{A_{\mathrm{ann}}} \tag{5-44}$$

where A_{ann} is the cross-sectional annular area defined as the cross-sectional area of the housing without coalescers minus the area of the coalescer end caps:

$$A_{\mathrm{ann}} = \pi R_{\mathrm{h}}^2 - N\pi R_{\mathrm{c}}^2 \tag{5-45}$$

where R_{h} is the radius of the housing; R_{c} is the radius of coalescer end cap; and N is the number of coalescers.

The enlarged droplets leaving the coalescer media pack can be assumed to be as large as possible for the given flow conditions when complete coalescence has occurred. Therefore, the coalesced droplet diameter will be the same for any specific design of the coalescer cartridge as long as complete coalescence has been achieved. If complete coalescence is not achieved, the calculation of the coalesced droplets must take into account the degree of coalescence.

In most industrial applications, the coalesced droplets will range in size from 0.5 to 2.2 mm and will be mostly influenced by the interfacial tension, which is significantly affected by the liquid density, system temperature, and system pressure. As the pressure is increased, the gas density will increase while the liquid density is only slightly affected. The solubility of the gas in the liquid is enhanced with increasing pressure. This leads to a substantial decrease in interfacial tension with increasing pressure and consequently to significantly smaller coalesced droplets at the higher pressures.

Once the coalesced droplet size has been estimated, the next step is to determine the maximum annular velocity that can be sustained without reentrainment. In general, the coalesced droplets will produce Reynolds numbers (Re) outside of the creeping flow regime (<0.1) and Stokes law. Instead, a force balance is used between the liquid droplets settling by gravity and the drag force of the gas flowing upward in the opposite direction.

5.7.3.4 Determination of Minimum Housing Diameter

The housing diameter is determined from the area of the annulus and the area of the coalescer end caps. The maximum annular velocity at the top of the coalescer cartridges is used to determine the annular area required. The value of the maximum annular velocity ($V_{\text{ann,max}}$) at the top of the coalescer cartridges is dependent on the critical annular velocity for reentrainment (V_c) and the vertical location at which the coalesced droplets are present in the free annulus space. This relationship can be described as follows (Wines, 2004):

$$V_{\text{ann,max}} = k_a V_c \tag{5-46}$$

where k_a is the annular velocity enhancement factor due to drainage.

For the untreated coalescer medium, the coalescer cartridge is completely wetted, and coalesced droplets are present in the annulus space up to the top of the annulus where the annular velocity is highest. There is no drainage enhancement and $k_a = 1$. The maximum annular velocity to prevent reentrainment is then equal to the critical value for reentrainment:

$$\text{Untreated Coalescer}: V_{\text{ann,max}} = V_c \tag{5-47}$$

The effect of the surface treatment is to greatly increase the drainage, and the annular velocity at the top of the coalescer cartridge can now be significantly higher than the critical value since there are no coalesced droplets present in the annulus except in the bottom third of the cartridge. The maximum annular velocity is now determined with $k_a = 3.1$ as follows:

$$\text{Surface Treated Coalescer}: V_{\text{ann,max}} = 3.1\,V_c \tag{5-48}$$

Convincing evidence for the enhanced maximum annular velocity given by Equation 5-46 has been demonstrated by laboratory tests (Williamson et al., 1988; Murphy, 1984) and is presented in Figure 5-9. Visual observations during these tests also confirmed that liquids are present on the outside of the coalescer pack only at the bottom third for the surface-treated coalescer and are present throughout the length of the wetted untreated coalescer.

5.7.4 Coalescer Performance/Operational Limits

Generally, the high-efficiency liquid-gas coalescers are used for inlet aerosol concentrations of a few thousand ppmw or less and are placed downstream of other bulk removal separators as the final purification stage. Under these conditions, typical service life for liquid-gas coalescers is 1–2 years. Coalescer systems are usually sized for a clean differential pressure (DP) of 2–5 psi, and when this DP reaches 15 psi, they are replaced with new elements.

Outlet concentrations for these high-efficiency liquid-gas coalescers are as low as 0.003 ppmw (Murphy, 1984; Williamson et al., 1988). The experimental apparatus for measuring coalescer efficiency is shown in Figure 5-10. This test was developed by Pall Corporation and is known as the liquid aerosol separation efficiency (LASE) test. At this time, there is no standard test for liquid-gas coalescers that the industry has universally adopted. Some of the important aspects of measuring coalescer performance that are addressed in this test are the following:

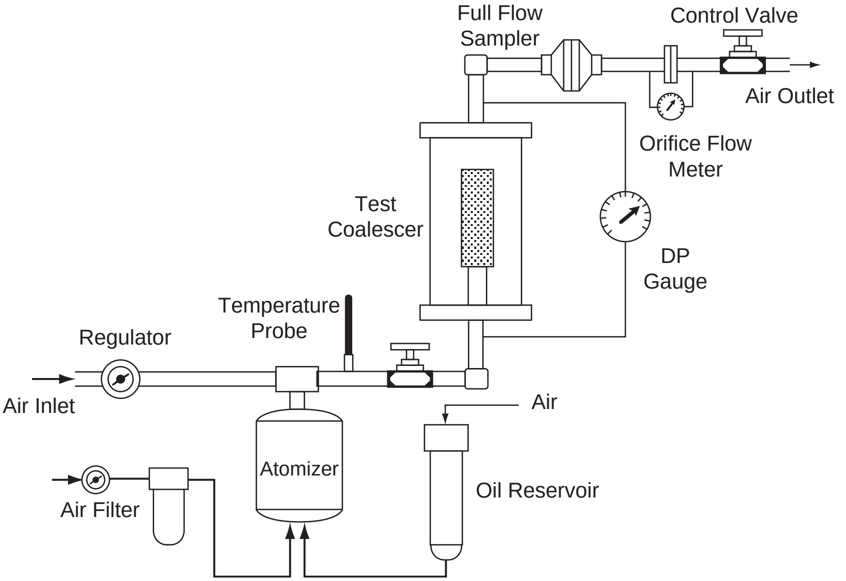

Figure 5-10 Liquid aerosol separation efficiency (LASE) test.

- Difficult aerosol challenge with drops ranging from 0.1 to 1 μm created in an atomizer—representative of condensation aerosols.
- Test operated under pressurized conditions—not vacuum.
- Coalescer evaluated after it is saturated with liquids—representative of field operation.
- Outlet contaminant collected by a full-flow sampler that collects any aerosols or liquids flowing along the walls of the pipe.

5.7.5 Liquid-Gas Coalescer Applications

The separation of liquid aerosol contamination with high-performance liquid-gas coalescer cartridge systems has found widespread acceptance in refinery and gas plants in recent years for a number of applications, including protection of compressors, turbo equipment, low NOx burner nozzles, amine and glycol contactors, molecular sieve beds, and wellhead hydrate inhibition. This has largely been the result of traditional separation approaches, including knockout vessels, centrifugal separators, mesh pads, or vane separators, not meeting the end user's requirements for aerosol reduction. A brief description of some of the main applications is given in the following sections.

5.7.5.1 Compressor Protection

Contaminants in the inlet gas can have a severe effect on compressor reliability. At least 20% of all reciprocating compressor failures can be attributed to inlet gas contaminants (Smith et al., 1997). Even when the compressors are not having catastrophic failure, a costly maintenance schedule may be followed where the compressor is shut down every 6 months or less for inspection and minor repairs. A preventative system using a high-performance liquid-gas coalescer system, however, was found to provide protection of the compressor for a 2-year period (Wines and Lorentzen, 1999).

Gas contaminants are made up of solid particulates and liquid aerosols, often including a high percentage in the submicron size range. Solids are usually corrosion products (iron oxides or iron sulfides), salts, or silt. Liquids can be hydrocarbons (refinery products, lube oil, condensate), aqueous (water, alcohol, dissolved salts, caustic, acid), or a combination of both. Contaminants will affect reciprocating and centrifugal compressors in different ways.

Reciprocating compressors are affected by aerosol contaminant primarily in the piston cylinder intake/exhaust. A buildup of contaminants on the valves can lead to sticking valves or valves with partial bypass, resulting in a reduced compression ratio and increased power consumption. The buildup of contaminants inside the cylinder can also damage the piston rings, pistons, and the cylinder wall.

Centrifugal compressors impart a dynamic head to the inlet gas by use of high-speed impellers that are confined in a casing containing stationary diffusers. While centrifugal compressors are less sensitive to inlet gas contaminants, they are still adversely affected. For centrifugal compressors, the primary contaminant-related problem is the buildup of foulants on the rotating blades (also known as salting). This can lead to a partial blockage in the flow path, causing increased power consumption, and can create an imbalance in the blades, leading to serious mechanical failures due to resultant vibrations if left unchecked.

5.7.5.2 Amine/Glycol Contactor Protection

Gas-treating processes that use liquid-gas contactors can be prone to foaming when contaminants build up in the solvent systems. The most common gas-treating solvent systems are alkanol amines used to remove H_2S or CO_2 and triethylene glycol used to remove water vapor. Activated carbon and antifoam chemical additives are often used to alleviate foaming issues. These approaches will not be effective for significant contaminant ingression as the activated carbon will become saturated past its capacity at a rapid rate. The use of antifoam treats a symptom rather than addresses the root cause by removing the contaminant. Upsets in contaminant levels can often trigger foaming as it is difficult to gauge the respective antifoam level required. High levels of antifoam additive can also have the opposite effect and lead to foam promotion in some instances. A total fluid-management approach of using high-efficiency liquid-gas coalescers on the incoming gas and 10 μm absolute filters on an amine loop was found to be an effective approach to minimizing foaming and amine losses (Pauley et al., 1988).

5.7.5.3 Well Head Hydrate Inhibition

Hydrates, solid water/methane complexes, can form in natural gas under high pressures even at ambient temperatures. Hydrates can wreak havoc on gas processing and transmission systems by plugging off pipelines and process equipment. Methanol and/or glycol injection are often used to inhibit hydrate formation. Removing water aerosols right at the wellhead is an even more effective way to prevent issues with hydrates forming in the downstream pipeline and will reduce chemical injection costs for methanol and glycol.

5.7.5.4 Molecular Sieve Protection

Molecular sieves (zeolites) are used to dehydrate gas to achieve dew points well below saturation and are effective at removing water vapor. When free water in the form of aerosols is present, the molecular sieves can reach maximum water capacity, rapidly leading to frequent regeneration. Regeneration is accomplished thermally and will have an associated cost for this heating value. Hydrocarbon and solid aerosols can lead to fouling of the molecular sieve bed and thereby reduce the number of active sites to bind to water vapor. The use of high-efficiency liquid-gas coalescers is an effective and economic means to protect molecular sieve beds.

5.7.5.5 Low and Ultra-Low NOx Burner Protection

Burners are used extensively in the refining and petrochemical industries. Applications include furnaces used for crude oil heating, cracking, reforming, and coking as well as boilers, gas turbines, driers, and incinerators. To meet environmental regulations, advanced burner technology including low and ultra-low NOx burner tips is being increasingly applied. These advanced burner designs are more intricate in design and have finer orifices that are more prone to fouling than older designs.

Contamination in the fuel gas can lead to process problems with the furnace operation. The nozzles can become plugged, leading to poor furnace performance, and in extreme cases, can damage the convective sections of the furnace. At some plants, furnace maintenance has become a costly task requiring burner tip replacement or cleaning every few days. Fuel gas can contain contaminants from various sources. Corrosion products can form in the process piping and lead to fouling by inorganic materials. Liquid hydrocarbons can result in coking at the burner tip, leading to fouling by organic materials. Some of the contaminants found in fuel gas include iron sulfides, iron oxides, amine, glycol, water, and hydrocarbon liquids.

High-efficiency liquid-gas coalescers have been applied successfully to protect low and ultra-low NOx burners (Wines, 2004). They can be installed either at the point of use directly upstream of a bank of furnaces or at a central location conditioning the feed to multiple banks of burners. The piping should be thoroughly cleaned prior to installing the coalescers, and any piping downstream of the coalescers should be either heat traced or insulated to prevent temperature drops that could cause liquids to condense.

5.8 HIGH-EFFICIENCY LIQUID–LIQUID COALESCERS

Liquid–liquid separations may require the use of special equipment when the drop sizes are small, typically in the range of 1 to 50 μm in size. These fluid systems are classified as stable emulsions, and often conventional bulk separators with mist pads or plate type internals will not be effective. High-efficiency liquid–liquid coalescers have been developed to break these emulsions and provide improved separation.

5.8.1 Emulsions

Emulsions consist of three components: oil (representing hydrocarbon or organic liquids), water (including any aqueous mixtures), and surfactants. Depending on the ratio of these components, oil-in-water emulsions or water-in-oil emulsions can exist. The structure of the oil-in-water or water-in-oil emulsions is well defined with spherical droplets of the dispersed phase surrounded by a bulk continuous phase and surfactant sheathing the droplets. Surfactants contain both hydrophilic (water loving) and hydrophobic (water fearing) portions in the same molecule. This unique structure allows them to associate at water-oil interfaces and helps them to stabilize the droplet shape. A spherical drop shape is formed to minimize the surface area between the oil and water, and this also minimizes the free energy required to make this surface. In order to make an emulsion, the system must be subjected to shear or mixing to allow the three components to break up into the droplet structure.

Emulsions are inherently unstable and will spontaneously separate out into two bulk phases. This process requires that the small droplets merge together or coalesce repeatedly until they form increasing drop sizes that eventually merge with a bulk phase until all of the drops are gone. Depending on the nature of the emulsion, the separation can occur in a matter of seconds or months. Many of the same factors that affect emulsion stability also influence coalescer performance.

5.8.1.1 Surfactants

Surfactants can be broadly classified into three groups: cationic, anionic, and nonionic. All surfactants consist of polar or hydrophilic groups joined to nonpolar or hydrophobic hydrocarbon chains. Cationic surfactants contain polar head groups that have a positive charge, whereas anionic surfactants have polar groups that have negative charges. Nonionic surfactants have polar groups that are neutral and are typically made up of ethylene oxide groups.

Surfactants can have a wide array of configurations including a polar head and nonpolar tail structure, branched tails, and random placement of polar groups within a nonpolar hydrocarbon chain. The ability of surfactants to aid in the emulsification process will depend on the ratio of the polar to nonpolar groups, the charge density and size of the polar group, and the volume occupied by the nonpolar groups (branching and length). Surfactants can also be influenced by other species such as cosurfactants (alcohols) or aqueous contaminants (salts).

The sources of surfactants found in industrial processes include intentional additives, surfactants found in nature, and surfactants created inadvertently through reaction processes. Some examples of surfactants classified this way are as follows:

- Additives: corrosion inhibitors, deemulsifiers, scale inhibitors, flocculents.
- Natural: petroleum naphtha sulfonates, naphthenic acids, and mercaptides in crude oil (Hughes, 1997).
- Reaction products: oxidation/caustic treatment of hydrocarbons in refineries leading to formation of cresylic acids and phenol homologs (Suarez, 1996).

5.8.2 Coalescer Principles and Materials of Construction

Coalescers are typically manufactured as either pads or cartridge filters that have been designed especially to take small droplets in an emulsion and grow them into large drops that are more easily separated. This process is accelerated over natural coalescing by the fibers present in coalescer media that force the contact of small droplets, thereby promoting the coalescing process. The pore gradient of coalescer medium is constructed so that the inlet medium has fine pore sizes that increase in size with the flow direction (see Figure 5-11).

Coalescers have been primarily constructed with glass fiber media until recently when polymer and fluoropolymer materials were adopted. Glass fiber works adequately for emulsions with interfacial tensions >20 dyne/cm. It is known to disarm and lose efficiency in the presence of surfactants (Hughes, 1997). These coalescers are widely used to dewater jet fuel for the aviation industry.

High-efficiency liquid–liquid coalescers are the newest generation of coalescers, incorporating the latest in coalescer technology. They are constructed from polymer and fluoropolymer materials that have been optimized to separate the most difficult emulsions with interfacial tensions

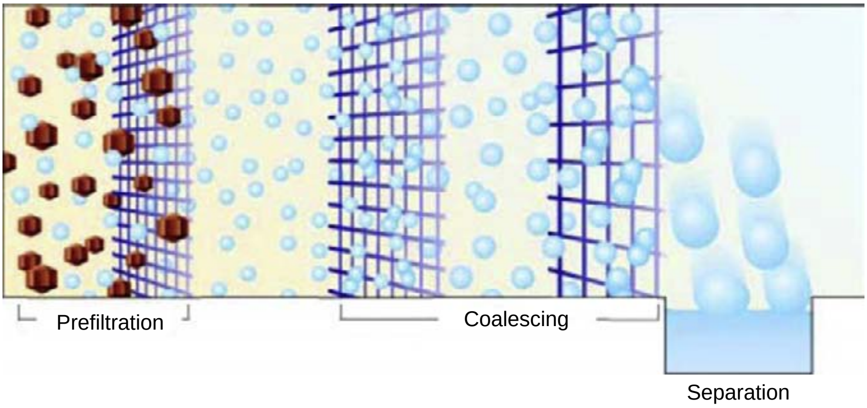

Figure 5-11 Coalescing in the media.

as low as 0.5 dyne/cm. This coalescer can be used with a broad range of applications. It can process aggressive chemicals and handle demanding operating conditions while providing the highest level of performance.

5.8.3 Coalescer Mechanism of Operation

The liquid–liquid coalescing system operates in three stages: separation of solids/preconditioning, coalescence, and separation of coalesced drops.

5.8.3.1 Separation of Solids/Preconditioning of the Fluid

Solids can increase the stability of an emulsion, and removing solids can make coalescing easier. Generally, this step can be achieved by a separate cartridge filter system or by a regenerable backwash filter system for high levels of solids. In addition, the filtration stage protects the coalescer and increases service life. This step also initiates the coalescence of the hydrocarbon droplets, thereby enhancing the separation capabilities of the system.

5.8.3.2 Coalescence

The next step in the process is the primary coalescence. In this stage, the pore dimensions begin with a very fine structure and then become more open to allow for void space for the coalescing droplets. In the primary coalescence zone, the inlet dispersion containing fine droplets in the size range of 0.2 to 50 μm is transformed into a suspension of enlarged droplets in the size range of 500 to 5,000 μm.

The coalescence mechanism can be described by the following steps:

1. Droplet adsorption to fiber.
2. Translation of droplets to fiber intersections by bulk flow.
3. Coalescence of two droplets to form one larger droplet.
4. Repeated coalescence of small droplets into larger droplets at fiber intersections.
5. Release of droplets from fiber intersections due to increased drag on adsorbed droplets caused by bulk flow.
6. Repeat of steps 1–5 with progressively larger droplet sizes and more open media porosity.

Based on this mechanism, we can predict that a number of factors will influence the coalescence performance. The specific surface properties of the coalescer fibers are critical in influencing the adsorption of droplets as well as the ultimate release after coalescing. There is a balancing act between increasing the attraction or adsorption characteristics of the fibers against the release mechanism, which strong adsorption would inhibit. The necessary condition that droplet-fiber adsorption occur for coalescing has been supported by a number of sources (Jeater et al., 1980; Basu, 1993).

5.8.3.3 Separation of Coalesced Droplets

Once the droplets have been coalesced, they are now as large as possible for the given flow conditions. The separation stage can be achieved in one of two ways:

- Horizontal configuration: The coalescer housing contains a settling zone that relies on the difference in densities between the coalesced droplets and the bulk fluid (see Figure 5-12). This configuration can be used for

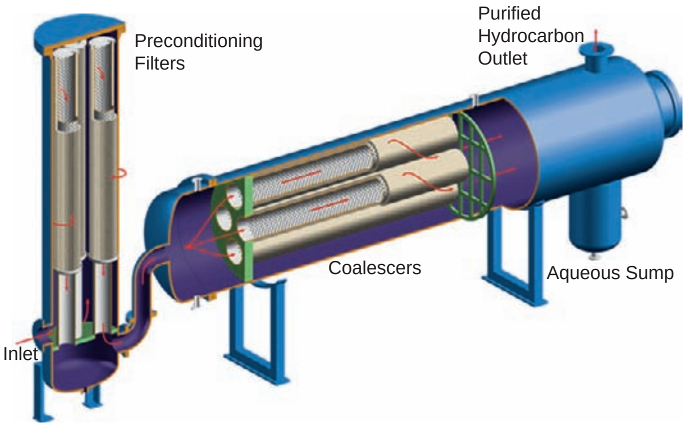

Figure 5-12 Horizontal liquid–liquid coalescer configuration (Katona et al., 2001).

both hydrocarbon from water and water from hydrocarbon separation, but the location of the collection sump and outlet nozzle will need to be reversed. For the case of the removal of hydrocarbon from water, a collection sump is located at the top of the housing, and the purified water leaves at the bottom outlet nozzle. The sump can be manually drained on a periodic basis or equipped with an automatic level control and drain system. Estimation of the coalesced drop size and required settling zone are best determined through pilot scale tests at field conditions.

- Vertical configuration: Once the droplets have been coalesced, they are now as large as possible for the given flow conditions, in the range of 0.5 mm to 2 mm in diameter. The separation stage is achieved using hydrophobic separator cartridges that provide an effective barrier to aqueous coalesced drops, but allow hydrocarbon to pass through them. The separator cartridges can be stacked below the coalescers for the most efficient utilization of the separator medium. This configuration applies only to the separation of water or aqueous contaminants from hydrocarbons (see Figure 5-13). After leaving the coalescing stage, the large aqueous coalesced drops and hydrocarbon then flow axially in a downward direction, and the flow direction is from the outside of the separator to the inside. The large coalesced drops are repelled by the separators and are collected in the bottom sump. The purified hydrocarbon passes through the separators and exits at the bottom of the housing. The aqueous phase in the collection sump can be manually drained on a periodic basis or equipped with an automatic level control and drain system.

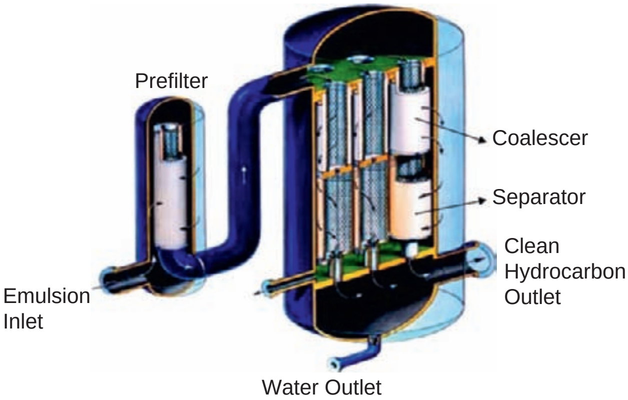

Figure 5-13 Vertical liquid–liquid coalescer configuration (Hampton et al., 2001).

5.8.4 Liquid–Liquid Coalescer Performance

Properly designed and sized high-efficiency coalescer systems can process inlet discontinuous-phase concentrations up to 10% and reduce them to ppm levels in the outlet for interfacial tensions as low as 0.5 dyne/cm. For water from hydrocarbon separations, coalescer outlets below 15 ppmv per the AquaGlo (*registered trademark of Gammon Corp.*) method (ASTM D3240) can be achieved, and for hydrocarbons from water, concentrations below 20 ppmw per the oil and grease method have been demonstrated.

The use of polymers and fluoropolymers in the coalescer materials of construction allows for expanded use of coalescers over earlier conventional types so that they can withstand an array of aggressive chemical applications over a wide range of temperatures from –40°F up to 300°F.

5.8.5 Limitations of Using Coalescers

While liquid–liquid coalescers have many benefits in breaking tough emulsions, there are some limitations to consider. Solids can become problematic at higher concentrations and lead to excessive change out of disposable prefilters. Generally, the solids range that liquid–liquid coalescers can operate economically with disposable filters is less than 10 ppm. Above this level of solids, further pretreatment will be required, such as backwash cartridge filters, mixed media packed beds, or hydrocyclones for solids removal.

The operational limits of the coalescer for removal of free liquids must also be understood. If the clarified stream leaving the coalescer is then cooled, condensation of previously dissolved contaminant can occur, leading to a hazy fluid at the lower temperature. The coalescer will not be able to remove contaminants that are dissolved in solution, and therefore, the location to place the coalescer and any subsequent change in process conditions after the coalescer must be considered carefully.

The coalescers typically will have a service life of 1–2 years when protected adequately by prefiltration. Despite the long life, the coalescers will eventually require disposal and replacement; however, this will be minimal given the low frequency of replacement.

For liquid–liquid coalescers constructed from glass fiber medium, the problem of surfactant disarming must also be considered, and for low IFT emulsion systems (<20 dyne/cm), they will not operate efficiently for separation. For these conditions, nondisarming fluoropolymer or polymer coalescers should be considered. These types of materials also have wider compatibility for chemical streams and a wider temperature range.

5.8.6 Applications

High-efficiency liquid–liquid coalescers are finding increasing applications in industry where problematic emulsions exist. They are used to protect equipment, recover valuable streams, and meet environmental discharge limits. Some examples are given in the following sections.

5.8.6.1 Pipeline Condensate

In gas production, associated light-end hydrocarbon condensates and water are produced along with the gas. In many instances, chemical additives such as methanol and glycol are used to prevent hydrate formation. These chemicals can greatly increase the tendency for the water-hydrocarbon condensate to form stable emulsions. High-efficiency liquid–liquid coalescers have been applied to separate these difficult emulsions to purify both the water and hydrocarbon phases.

5.8.6.2 Produced Water

Discharge limits on produced water are becoming more stringent with current levels in many areas at 29 ppm oil and grease. To treat produced water to achieve this target often requires multiple separation options, including mixed sand beds, dissolved air flotation, and hydrocyclones. Despite these efforts, emulsified oil can still penetrate through to the effluent, so the use of high-efficiency liquid–liquid coalescers offers a reliable way to meet environmental limits.

5.8.6.3 Petrochemical Final Products

In refinery operations, steam stripping is often used to remove volatile components (hydrogen sulfide and light hydrocarbon fractions) from petroleum products, including gasoline, diesel, and kerosene. This can lead to water condensation and off-specification hazy final products.

5.8.6.4 Caustic Treating

In the production of gasoline, organic sulfur compounds such as mercaptans often need to be removed. One of the more well-known processes for removal of mercaptans is known as the Merox *(registered trademark of UOP)* process licensed by UOP. In the reactor, mercaptans are extracted into the caustic and are then converted to disulfide oils by oxidation and catalytic action. Phenols are also extracted into the caustic phase. The reactor effluent flows to a three-phase separator where air, disulfides, and regenerated caustic are separated. The regenerated caustic is then recycled to the reactor.

The reactor effluent typically contains a quantity of carried-over caustic that results in hazy gasoline product, high costs of caustic makeup, and corrosion of downstream piping. The hazy gasoline is problematic and has to be blended off or reprocessed; otherwise, it would cause the product to be off-specification in terms of sodium. Installing a high-efficiency liquid–liquid coalescer system was found to be an effective solution to recover the caustic carryover (Katona et al., 2001).

5.8.6.5 Recover Liquid Catalysts

Liquid or homogeneous catalysts are used in a number of processes where the catalyst and the reactants are insoluble and form emulsions. Recovery of the catalyst is important to reduce the need for makeup catalyst and to protect downstream equipment from fouling by the catalyst.

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CHAPTER 6

Condensate Stabilization

6.1 INTRODUCTION

Hydrocarbon condensate recovered from natural gas must be treated to make it safe and environmentally acceptable for storage, processing, and export. Therefore, removing water and salt is mandatory to avoid corrosion. Separation of any dissolved gases, which belong to the light hydrocarbon components (methane and ethane in particular), along with hydrogen sulfide, mercaptans, and other sulfur compounds, will make condensate safe and environmentally acceptable to handle.

The process of increasing the amount of intermediate (C_3 to C_5) and heavy (${C_6}^+$) components in the condensate is called condensate stabilization. This process is performed primarily in order to reduce the vapor pressure of the condensate liquids so that a vapor phase is not produced upon flashing the liquid to atmospheric storage tanks. Stabilized liquid generally has a vapor pressure specification, since the product will be injected into a pipeline or transport pressure vessel, which has definite pressure limitations.

Apart from stabilization problems of sweet condensate, sour condensates also present unusual processing problems. Along with stabilization, condensate sweetening permits easier and safe downstream handling as well as improving and upgrading the condensate marketability. The most economical place for the sweetening process is usually in the refinery. However, when marketing or pipeline requirements are imposed, gas processing plants are needed for processing the sour condensate prior to shipping.

The purpose of this chapter is to describe the basic processes for stabilizing the condensate against two important constraints imposed by its vapor pressure and the allowable hydrogen sulfide content, and the associated equipment design procedure. Condensate sweetening will be discussed in Chapter 10.

6.2 STABILIZATION PROCESSES

Stabilization of condensate streams can be accomplished through either flash vaporization or fractionation.

6.2.1 Flash Vaporization

Stabilization by flash vaporization is a simple operation employing only two or three flash tanks. This process is similar to stage separation utilizing the equilibrium principles between the vapor and condensate phases. Equilibrium vaporization occurs when the vapor and condensate phases are in equilibrium at the temperature and pressure of separation. Figure 6-1 shows a typical scheme of the condensate stabilization through the flash vaporization process. As it can be seen, condensate from the inlet separator, after passing through the exchanger, enters into the high-pressure (HP) flash tank, where the pressure is maintained at 600 psia. A pressure drop of 300 psia is obtained here, which assists flashing of large amounts of lighter ends that join the sour vapor stream after recompression. The vapor can either be further processed and put into the sales gas or recycled into the reservoir and used as gas lift to produce more crude oils. The bottom liquid from the high-pressure tank flows to the middle-pressure (MP) flash tank operated at 300 psia. Additional methane and ethane are released in this tank. The bottom product is withdrawn again to the low-pressure (LP) tank operated at 65 psia. To ensure efficient separation, condensate is degassed in the stripper vessel at the lowest possible pressure prior to storage. This reduces excess flashing of condensate in the storage tank and reduces the inert gas blanket pressure required in it.

Note that flash vaporization as a condensate stabilization method is old technology and is not used in a modern gas plant. However, variations of the flash technology might also be found on oil production facilities stabilizing crude oil.

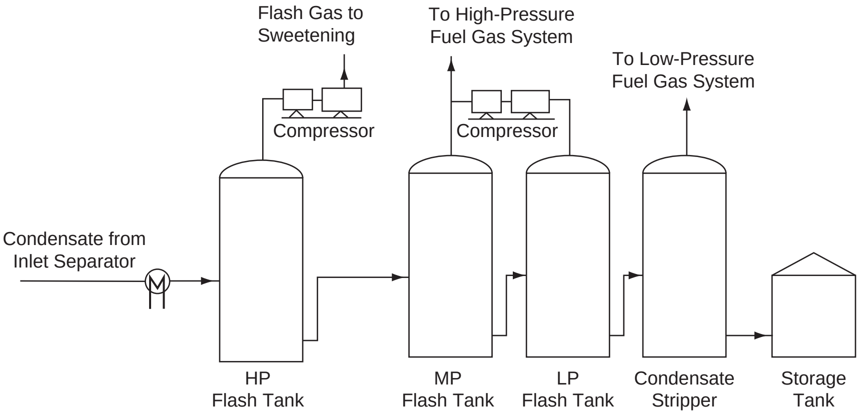

Figure 6-1 Schematic of condensate stabilization through flash vaporization process.

6.2.2 Stabilization by Fractionation

Stabilization by fractionation is a detailed process, very popular in the industry and precise enough to produce liquids of suitable vapor pressure. During the operation, the undesirable components (low-boiling-point hydrocarbons and hydrogen-sulfide gas) are removed. The finished product from the bottom of the column is composed mainly of pentanes and heavier hydrocarbons, with small amounts of butane. The process actually makes a cut between the lightest liquid component (pentane) and the heaviest gas (butane). The bottom product is thus a liquid free of all gaseous components able to be stored safely at atmospheric pressure. Stabilization by fractionation is a modern operation and economically attractive next to flash vaporization. It is a single tower process since only one specification product is required. The bottom product of the column is capable of meeting any kind of rigid specification with the proper operating conditions.

6.2.2.1 Process Description

Figure 6-2 shows a schematic condensate stabilization process. The liquid hydrocarbon (condensate) is brought into the system from the inlet separator and preheated in the stabilizer feed/bottoms exchanger, before entering the stabilizer feed drum.[1] Liquid from the feed drum is fed to the stabilization tower at approximately 50 to 200 psi, depending on whether it is sour (sour stabilization is carried out at the low end of the range and sweet stabilization at the high end of the range). The condensate stabilizer reduces vapor pressure of the condensate by removing the lighter components. The stabilization is typically carried out in a reboiled absorber, with tray-type internals. However, if a better separation is required, typically the column is changed from a top-feed reboiled absorber to a refluxed distillation tower. As the liquid falls into the column, it becomes leaner in light components and richer in heavy ends. At the bottom of the tower, some of the liquid is circulated through a reboiler to add heat to the tower. As the gas goes up from tray to tray, more and more of the heavy ends get stripped out of the gas at each tray, and the gas becomes richer in the light components and leaner in the heavy ends. Overhead gas from the stabilizer, which would seldom meet market specifications for the natural gas market, is then sent to the low-pressure fuel gas system through a back-pressure control valve that maintains

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[1] Sometimes the liquid is flashed down to a feed drum at pressure slightly above the tower pressure. This flashes off vapor so that the stabilization tower can often be a smaller diameter.

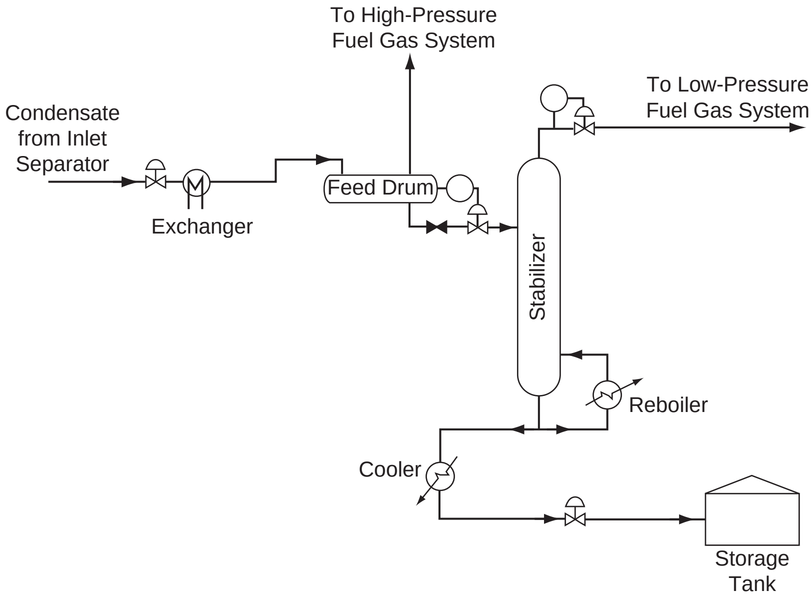

Figure 6-2 A schematic of condensate stabilization system.

the tower pressure to setpoint. The liquids leaving the bottom of the tower have undergone a series of stage flashes at ever-increasing temperatures, driving off the light components. These liquids therefore must be cooled to a sufficiently low temperature to keep vapors from flashing to atmosphere in the condensate storage tank.

Most gas processing plants are implementing advanced process control (APC) on their condensate production systems in order to maximize condensate yields, improve stability of the condensate stabilization process, and ensure that product quality limits are adhered to at all times. However, most often, the main reason to implement APC on this unit is product quality and control, where the gas suppliers sell the condensate product with a Reid vapor pressure (RVP) defined by the customer.[2] In this instance, a distributed control system is used by the APC system to maintain RVP within quality limits and push this to a higher specification, therefore, generating greater revenues as condensate throughput is increased (Hotblack, 2004).

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[2] The vapor pressure of a petroleum product is determined by a standard laboratory method called RVP, which measures its inherent tendency to evaporate at 100°F with vapor/liquid ratio of 4/1. For details of the equipment and procedures, refer to ASTM D323 (Vapor pressure of petroleum products Reid method).

From a control standpoint, some optimization schemes include

- Nonlinear level control concepts on the feed drum can smooth the flow to the stabilizer column. This allows the feed drum to be truly used as a capacitance. Feed rate to the stabilizer is not changed if the level in the feed drum is within a dead band and not changing too quickly. As the level approaches the dead-band limits, flow is gradually changed. As the level moves outside the dead band, then the feed rate is moved more aggressively.
- Predictive models can be employed to anticipate the effects of feed rate and compositional effects on the bottom composition.
- An inferential property for bottom RVP can be determined based on trays with sensitive pressure-compensated temperatures (PCT) to improve the control of the product quality. Alternatively, the most sensitive tray temperature or PCT can be cascaded to the flow of heat input to the reboiler.
- Separation efficiency is improved as tower pressure is lowered. This can reduce heat requirements for reboiling while making a better separation between light and heavy key components in the tower. The pressure can be lowered subject to maximum valve opening of the pressure control valve or column flooding as indicated by differential pressure measurement across the tower. The tower is more susceptible to flooding at lower pressures. If an overhead compressor is employed to boost the tower overhead up to fuel system pressure, then the pressure can be lowered subject to the maximum speed of the compressor driver, rod load limitations for a reciprocating compressor, or surge considerations for a centrifugal compressor.

6.2.2.2 Design Considerations of Stabilization Column

In most cases of lease operation, the stabilization column will operate as a nonrefluxed tower. This type of operation is simpler but less efficient than the refluxed tower operation. The nonrefluxed tower requires no external cooling source, so it is particularly applicable to remote locations. A condensate stabilization column with reflux will recover more intermediate components from the gas than a cold-feed stabilizer. However, it requires more equipment to purchase, install, and operate. This additional cost must be justified by the net benefit of the incremental liquid recovery, less the cost of natural gas shrinkage and loss of heating value, over that obtained from a cold-feed stabilizer. When a condenser is used in a stabilization column, it will always be a partial condenser because of the quantities of methane and ethane that must be removed from the tower feed.

The stabilization tower pressure depends on the amount of liquid to be stabilized and whether it is sweet or sour. For sweet stabilization, the pressure should be as high as possible to minimize overhead vapor recompression, since this gas is remixed with the separator vapor. This also tends to decrease the cost of reflux cooling, if it is used. However, relative volatility of the components also decreases with pressure, and as previously stated, driving H_2S overhead requires a relatively low pressure.

Figure 6-3 shows the maximum recommended feed temperature to a stabilizer as a function of operating pressure of the stabilizer. An exception to this may be the case where either sour or small quantities of liquid are being handled and where first cost is therefore very critical. In these cases, many times a 40 to 70 psia working pressure for nonrefluxed tower may show an economic advantage (Campbell, 2000).

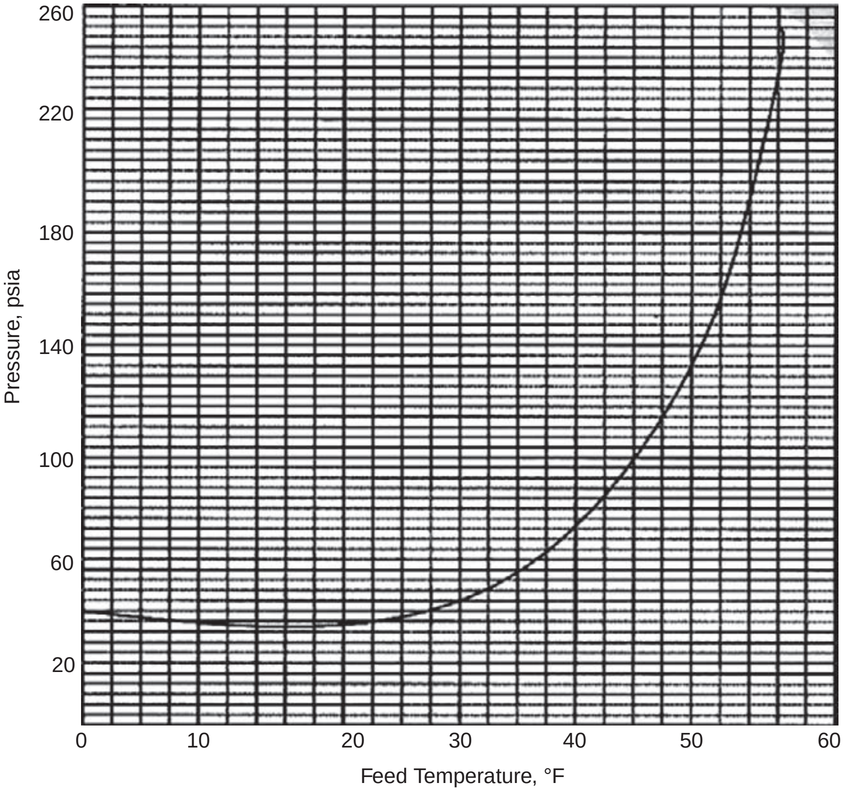

Figure 6-3 Maximum recommended feed temperature to cold-feed stabilizer (Campbell, 2000).

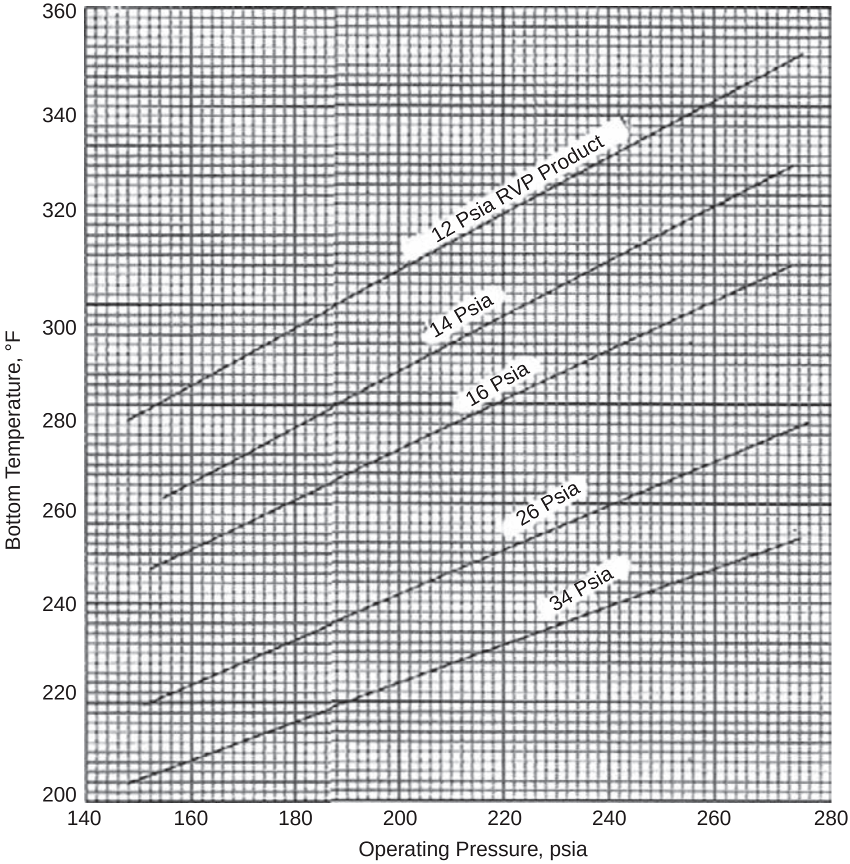

Figure 6-4 Estimation of proper bottom temperature of nonrefluxed stabilizer (Campbell, 2000).

In some cases, the reboiler for the stabilizer will be an indirect salt bath heater or a steam-fired heat exchanger. Figure 6-4 shows suggested bottom (reboiler) temperatures for producing a specified Reid vapor pressure (RVP) product. In fact, the temperature used on the bottom is limited by the thermal breakdown characteristics of the condensate and, of course, product specs.

After the pressure has been chosen and the operating temperatures have been established through use of the charts shown in Figures 6-3 and 6-4, the split in the tower must be predicted. There are several methods in which this can be done, but one of the most convenient manual methods involves utilization of pseudo-equilibrium constant (*K*) values for each component between the top and the bottom of the tower. Using this concept, one can estimate the separation that can be achieved across a nonrefluxed stabilizer by use of the pseudo K values and a simple flash calculation. The vapor from the flash calculation will be the composition of the overhead product, and the liquid from the flash calculation will be the composition of the bottom liquid. Today most often distillation and absorption/stripping calculations are done with process simulation computer software.

Note that for estimating the desired composition of the bottom liquid if a split of nC_4 (normal butane) is assumed, the mole fraction of each component in the liquid can be estimated from the following equations:

$$L_i = \frac{F_i(\mathrm{nC_4\ split})}{RV_i} \tag{6-1}$$

$$X_i = \frac{L_i}{\sum_{i=1}^{n} L_i} \tag{6-2}$$

where X_i is mole fraction of component i in the liquid; F_i is total number of moles of component i in the feed; L_i is total number of moles of component i in the bottom liquid; nC_4 split is assumed moles of component nC_4 in the bottom liquid divided by moles of nC_4 in the feed; RV_i is relative volatility of component i from Table 6-1; and n is the number of components in the bottom liquid.

Table 6-1 Reid Vapor Pressure (RVP) and Relative Volatility of Various Components (Reid et al., 1977)

Component	RVP Psia	Relative Volatility
C_1	5000	96.9
C_2	800	15.5
C_3	190	3.68
i-C_4	72.2	1.40
n-C_4	51.6	1.00
i-C_5	20.4	0.40
n-C_5	15.6	0.30
C_6	5.0	0.10
C_7^+	≈0.1	0.00
CO_2	—	Infinite
N_2	—	Infinite
H_2S	394	7.64

The vapor pressure is the primary property used to make this split. It is assumed that the mole fraction of each component times its vapor pressure represents the contribution of that component to the total mixture vapor pressure. The total mixture vapor pressure can then be computed from Equation 6-3:

$$P_{\mathrm{V}} = \sum_{i=1}^{n} (P_{\mathrm{Vi}} \times X_{\mathrm{i}}) \tag{6-3}$$

where P_{V} is the vapor pressure of the mixture, psia; and P_{Vi} is vapor pressure of component i, psia.

If the vapor pressure of the mixture is higher than the desired RVP of the bottom liquid, choose a lower number for the nC_4 split. If the calculated vapor pressure is lower than the desired RVP, choose a higher number for the nC_4 split. Iterate until the calculated vapor pressure equals the desired RVP of the bottom liquid.

The bottom liquid temperature can also be determined by calculating the bubble point of the liquid described by the previous iteration at the chosen operating pressure in the tower. This is done by choosing a temperature, determining pseudo K values from the chart in Figure 6-5 and computing parameter C by using the following equation:

$$C = \sum_{i=1}^{n} (L_{\mathrm{i}} \times K_{\mathrm{i}}) \tag{6-4}$$

If C is greater than 1.0, the assumed temperature is too high. If C is lower than 1.0, the assumed temperature is too low. By iteration, a temperature can be determined where $C = 1.0$. Typically, bottoms temperatures will range from 200°F to 400°F, depending on operating pressure, bottoms composition, and vapor pressure requirements. Temperatures should be kept to a minimum to decrease the heat requirements, limit salt buildup, and prevent corrosion problems.

6.3 CONDENSATE STORAGE

Condensate is stored between production and shipping operations in condensate storage tanks,[3] which are usually of floating-roof type (external and internal). If the condensate does not meet the specifications, the

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[3] Condensate storage tanks are needed in order to receive the stabilized condensate, before pumping to the pipelines, as well as to allow for measuring condensate properties, sampling, and gauging.

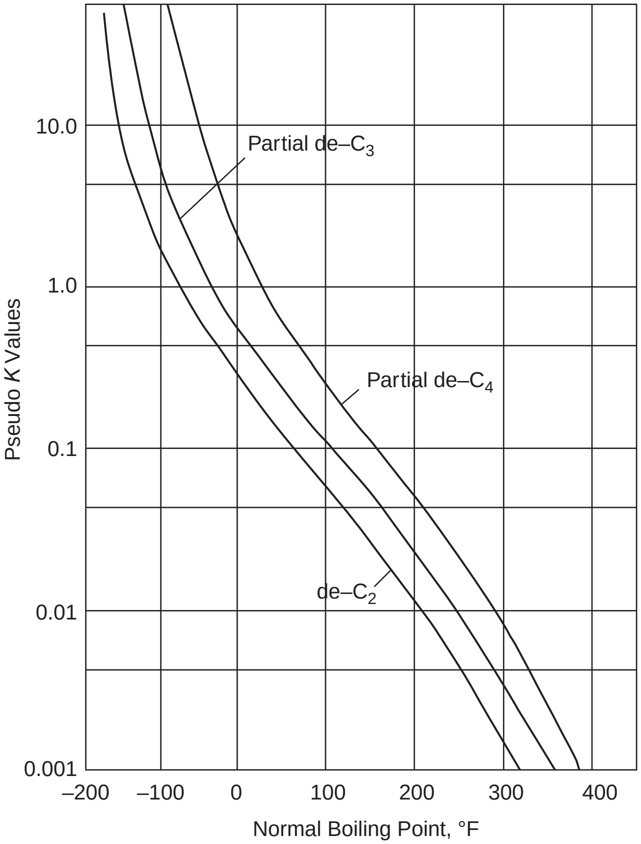

Figure 6-5 Pseudo *K* values for cold-feed stabilizers (Campbell, 2000).

off-specification condensate may be routed to an off-specification condensate storage fixed-roof tank (vertical and horizontal) until it is recycled to the condensate stabilization unit by the relevant recycle pump if the latter is available at the plant. The primary quality criterion for the condensate is its RVP, which is affected by atmospheric pressure (plant elevation) and maximum ambient temperature. The Reid vapor pressure is usually set in the range of 10–12 psia. To store the condensate in floating roof storage tanks, it is very crucial to control the RVP at the desired level (especially in warm seasons).

6.3.1 Types of Storage Loss

Hydrocarbon losses from condensate storage tanks are normally categorized as occurring from breathing losses, standing storage losses, or working losses. The term "breathing loss" refers to those losses that result without any corresponding change in the liquid level within the tank. Most likely, these types of emissions result from hydrocarbon vapors that are released from the tank by expansion or contraction caused by changes in either temperature or pressure. Sources of standing losses are vapor escape from hatches or other openings and from glands, valves, and fittings. Working loss represents those emissions that occur due to changes in the liquid level caused by either filling or emptying the tank itself (U.S. EPA's AP-42 Manual, 1995).

The working pressure required to prevent breathing and thereby save standing storage losses depends on the vapor pressure of the product, the temperature variations of the liquid surface and the vapor space, and the setting of the vacuum vent. When these factors are known, the storage pressure required to eliminate venting can be computed by Equation 6-5 (GPSA, 2004):

$$P_S = P_{V,\max} + (P_T - P_{V,\min})\left[\frac{T_{\max} + 460}{T_{\min} + 460}\right] - P_{\text{atm}} \tag{6-5}$$

where P_S is required storage pressure, psia; $P_{V,\max}$ is true vapor pressure of liquid at maximum surface temperature, psia; $P_{V,\min}$ is true vapor pressure of liquid at minimum surface temperature, psia; $T_{\max}$ is maximum average temperature of vapor, °F; $T_{\min}$ is minimum average temperature of vapor, °F; P_{atm} is atmospheric pressure, psia; P_T is absolute internal tank pressure at which vacuum vent opens, psia.

For the condition where $P_{V,\min}$ is greater than P_T, Equation 6-5 can be represented as

$$P_S = P_{V,\max} - P_{\text{atm}} \tag{6-6}$$

With the situation represented by Equation 6-5, gas is admitted to the vapor space through the vacuum vent, where a gas vent system should be employed for preventing safety problems. However, for the situation represented by Equation 6-6, gas may be purged and kept out of the tank (Campbell, 2000).

Maximum liquid surface temperatures vary from 85°F to 115°F. Sufficient accuracy will generally result from the assumption that it is 10°F higher than the maximum temperature of the body of the liquid in a tank at that location.

Note that true vapor pressure (TVP) may be the most difficult term to calculate in the preceding equation. An API nomograph has been devised that relates TVP to both the Reid vapor pressure (RVP) and the storage temperature (T_S). Numerically, the relationship between TVP, RVP, and storage temperature can be expressed as follows (TRW Environmental, Inc., 1981):

$$\text{TVP} = (\text{RVP})\ \text{EXP}\left\{C_0\left(C_1 - \frac{1}{559.69}\right)\right\} \tag{6-7}$$

where C_0 is constant depending on the value of RVP; $C_1 = 1/(T_S + 460)$; and T_S is the temperature of the stored liquid, °F.

In the preceding equation, the term C_0 is dependent on the given value of RVP, as shown in Table 6-2 (TRW Environmental, Inc., 1981).

Once the parameters C_0, T_S, and RVP have been determined, one is technically able to calculate a value for TVP. It should be noted, however, that an error was discovered in the API nomograph calculated values of TVP, such that the RVP was not equal to the TVP at 100°F as should be expected, given the general definition of RVP. Through use of linear regression methods, correction factor (C_F) was developed; it should be added to the calculated values of TVP in order to obtain correct TVP values as follows (TRW Environmental, Inc., 1981):

$$\text{Corrected TVP} = \text{Calculated TVP} + C_F \tag{6-8}$$

The correction factor was found to be dependent on RVP according to the following equations:

$$\text{RVP} < 3 : C_F = 0.04(\text{RVP}) + 0.1 \tag{6-9}$$

$$\text{RVP} > 3 : C_F = \text{EXP}\{2.345206\ \text{Log}(\text{RVP}) - 4.132622\} \tag{6-10}$$

Table 6-2 C_0 for Different RVP Values

RVP	C_0	RVP	C_0
2<RVP<3	−6439.2	RVP=7	−6109.8
RVP=3	−6255.9	7<RVP<8	−6238.9
3<RVP<4	−6212.1	RVP=8	−6367.9
RVP=4	−6169.2	8<RVP<9	−6477.5
4<RVP<5	−6177.9	RVP=9	−6587.0
RVP=5	−6186.5	9<RVP<10	−6910.5
5<RVP<6	−6220.4	RVP=10	−7334.0
RVP=6	−6254.3	10<RVP<15	−8178.0
6<RVP<7	−6182.1	RVP>15	−9123.2

6.3.2 Vapor Recovery

The loss of hydrocarbon vapors from condensate storage tanks could be minimized using what is called a vapor recovery unit (VRU). These vapors, if allowed to escape to the atmosphere, will not only cause a loss of hydrocarbon volume but also lead to pollution and fire hazards. The main functions for the vapor recovery unit are to collect vapor from storage/loading facilities, to reliquefy vapors, and to return liquid hydrocarbons to storage tank.

Various methods can be recommended to recover vapors generated in storage tanks and from other sources. These methods usually involve one or a combination of the different schemes implemented through what is referred to as the VRU. However, carbon adsorption vapor recovery is recognized by the U.S. Environmental Protection Agency as both the best-demonstrated technology and maximum-achievable control technology. A typical carbon adsorption vapor recovery unit is composed of two adsorption vessels filled with activated carbon, a vacuum system used to strip hydrocarbon vapor from the carbon bed during the carbon regeneration process, and a recovery device (absorber or condenser), which turns the regenerated vapors into a liquid product. These systems are usually designed to meet the most stringent emission control requirements (99+%) providing valuable recovered products with the lowest operating costs.

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CHAPTER 7

Natural Gas Sweetening

7.1 INTRODUCTION

Natural gas, while ostensibly being hydrocarbon in nature, contains large amounts of acid gases such as hydrogen sulfide (H_2S) and carbon dioxide (CO_2).[1] In addition to acid gases, natural gas may contain other contaminants such as mercaptans (R–SH) and carbonyl sulfide (COS) and carbon disulfide (CS_2). Natural gas with H_2S or other sulfur compounds is called *sour gas,* whereas gas with only CO_2 is called *sweet gas*. Sour gas can cause extensive damage to natural gas pipelines if not properly processed. The combustion of sulfur compounds produces serious air pollutants and eventually produces acid rain when combined with water. These sulfur compounds are poisonous and lethal to humans and animals, and are corrosive to metals and other materials used for the handling and transporting natural gas. In addition, carbon dioxide is nonflammable and, consequently, large quantities are undesirable in a fuel.[2] Like H_2S, it forms a weak, corrosive acid in the presence of water. It is therefore obvious that removal of acid gases along with other sulfur species is a major concern in gas processing.

The principal challenge of the natural gas sweetening processes is to remove the acid gases and sulfur compounds from the natural gas to as low a level as possible, consistent with the prevailing emission regulations, and as economically as possible. A number of methods are available for removal of acid gases from product gas streams. However, processes that remove hydrogen sulfide and carbon dioxide from natural gas are not effective in removing mercaptans, disulfides, carbonyl sulfide, etc. The reason is that the aforementioned organic sulfur compounds do not readily ionize (hydrolyze) to form anions that react with caustic, amine, or metal oxides. The focus of this chapter is the treatment of natural gas to remove sour compounds, mainly H_2S and CO_2 but also on occasion mercaptans, carbonyl sulfide, and carbon disulphide, in order to meet specific treated gas specifications or permit additional processing in the plant without corrosion and plugging problems.

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[1] H_2S and CO_2 are referred to as acid gas components because they dissolve in water to form a weak acid.

[2] Carbon dioxide reduces the heating value and thus the sale value of the natural gas.

7.2 GAS SWEETENING PROCESSES

Acid gas removal processes are primarily of two types: adsorption and absorption. Adsorption is a physical-chemical phenomenon in which the gas is concentrated on the surface of a solid to remove impurities. Absorption differs from adsorption in that it is not a physical-chemical surface phenomenon. Absorption is achieved by dissolution (a physical phenomenon) or by reaction (a chemical phenomenon).

There are several processes for natural gas sweetening. Because the concentrations of CO_2 and H_2S in the raw gas to be processed and the allowable acid gas levels in the final product vary substantially, no single process is markedly superior in all circumstances, and, consequently, many processes are presently in use. Figure 7-1 illustrates the more important processes and groups them into the generally accepted categories. Details of processes not discussed in this chapter may be found in Kohl and Nielsen (1997).

If the sulfur recovery is an option, then direct conversion processes can be used. These processes are based on the oxidization of H_2S to elemental sulfur, where H_2S is absorbed in an alkine solution containing an oxidizing agent that converts it to sulfur. The solution is then regenerated by air in an oxidizer. H_2S conversion to elemental sulfur will be discussed in Chapter 8.

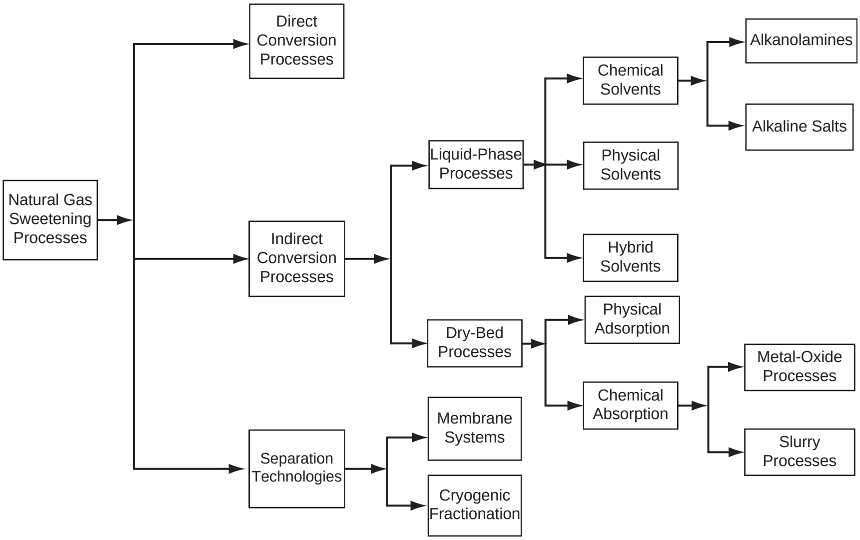

Figure 7-1 Alternatives for natural gas sweetening.

7.3 LIQUID-PHASE PROCESSES

Liquid-phase absorption processes are classified into three categories: chemical solvents, physical solvents, and hybrid solvents that contain both amine and a physical solvent.

In chemical solvent processes, absorption of the acid gases is achieved mainly by use of alkanolamines or alkaline salts of various weak acids such as sodium and potassium salts of carbonate[3] (Kohl and Nielsen, 1997). Regeneration (desorption) can be brought about by use of reduced pressures and/or high temperatures, whereby the acid gases are stripped from the solvent. Chemical absorption processes chemically absorb the H_2S, CO_2, and to some extent COS. Organic sulfur components do not chemically react with the solvent. Chemical solvents are specifically suitable when contaminants at relatively low partial pressure have to be removed to very low concentrations. Chemical solvents will not remove mercaptans down to low levels due to the low solubility of these components. An advantage, however, is that there is minimum coabsorption of hydrocarbons.

In physical solvent processes, which use an organic solvent, no chemical reaction occurs, and acid gas/organic sulfur components removal depends entirely on physical absorption. Physical solvents tend to be favored over chemical solvents when the partial pressure of the contaminants is high, the treated gas specification is moderate, and large gas volumes have to be purified. Unlike chemical solvents, physical solvents are noncorrosive, requiring only carbon steel construction. In addition, physical solvents can usually be stripped of impurities by reducing the pressure without the addition of heat (Carlsson et al., 2007). However, if the concentration of heavy hydrocarbons in the feed gas is high, a physical solvent may not be the best option due to higher coabsorption of hydrocarbons, particularly pentanes plus.

Hybrid solvent processes, which use a mixture of a chemical and a physical solvent, combine to a large extent the advantages of a chemical with those of a physical solvent. Hybrid solvents are more efficient in removing organic sulfur compounds and COS, but have the disadvantage of relatively poor hydrocarbon selectivity, resulting in hydrocarbon losses in the separated acid gases (Lallemand et al., 2008).

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[3] The most popular salt in the industry has been potassium carbonate.

7.3.1 Amine Processes

Chemical absorption processes with aqueous alkanolamine solutions are used for treating gas streams containing hydrogen sulfide and carbon dioxide. However, depending on the composition and operating conditions of the feed gas, different amines can be selected to meet the product gas specification. Amines are compounds formed from ammonia (NH_3) by replacing one or more of the hydrogen atoms with another hydrocarbon group. Amines are categorized as being primary, secondary, and tertiary, depending on the degree of substitution of the hydrogen atoms by organic groups. Replacement of a single hydrogen atom produces a primary amine (monoethanolamine, diglycolamine), replacement of two hydrogen atoms produces a secondary amine (diethanolamine, diisopropylamine), and replacement of all three of the hydrogen atoms produces a tertiary amine (e.g., methyl-diethanolamine). Physical properties of the most common amines are given in Appendix 3.

Different amines have different reaction rates with respect to the various acid gases. In addition, different amines vary in their equilibrium absorption characteristics for the various acid gases and have different sensitivities with respect to solvent stability and corrosion factors. Specific details for each amine follow.

7.3.1.1 Monoethanolamine

Monoethanolamine (MEA) is a primary amine and the strongest amine among others. MEA is a stable compound and, in the absence of other chemicals, suffers no degradation or decomposition at temperatures up to its normal boiling point. MEA reacts with H_2S and CO_2 as follows:

$$2(\mathrm{RNH_2}) + \mathrm{H_2S} \leftrightarrow (\mathrm{RNH_3})_2\mathrm{S} \tag{7-1}$$

$$(\mathrm{RNH_3})_2\,\mathrm{S} + \mathrm{H_2S} \leftrightarrow 2(\mathrm{RNH_3})\mathrm{HS} \tag{7-2}$$

$$2(\mathrm{RNH_2}) + \mathrm{CO_2} \leftrightarrow \mathrm{RNHCOONH_3R} \tag{7-3}$$

These reactions are reversible by changing the system temperature.

MEA can remove both H_2S and CO_2 from gas streams to meet sales gas specifications. In practice, however, acid gas loadings and solution concentrations are limited because of corrosion problems. MEA also reacts irreversibly with carbonyl sulfide (COS) and carbon disulfide (CS_2), which can result in degradation of the solvent and a buildup of heat-stable salts (HSS) that cannot be regenerated. MEA has a higher vapor pressure than the other amines. This can result in significant solution losses through vaporization. The problem usually can be overcome by a simple water wash of the sweet gas stream.

7.3.1.2 Diglycolamine

Diglycolamine (DGA) is the trade name for 2-(2-aminoethoxy) ethanol, which is a primary amine. The reactions of DGA with acid gases are the same as for MEA. This highly reactive primary amine removes COS, CS_2, and (partially) mercaptans, where degradation products from reactions with COS and CS_2 can be easily regenerated.

Compared with MEA, low vapor pressure allows DGA to be used in relatively high concentrations (45%–50%), which results in lower circulation rates. A disadvantage is the relatively high steam consumption. In addition, DGA absorbs aromatics and is more expensive than MEA.

7.3.1.3 Diethanolamine

Diethanolamine (DEA) is a secondary amine with lower reactivity and corrosivity than MEA. It has a lower vapor pressure (lower solvent losses) and lower heat of reaction (easier to generate) than MEA. DEA is nonselective and reacts with hydrogen sulfide and carbon dioxide as follows:

$$2R_2NH + H_2S \leftrightarrow (R_2NH_2)_2S \tag{7-4}$$

$$(R_2NH_2)_2S + H_2S \leftrightarrow 2(R_2NH_2)HS \tag{7-5}$$

$$2R_2NH + CO_2 \leftrightarrow R_2NCOONH_2R_2 \tag{7-6}$$

Difficulties are sometimes encountered with low-pressure DEA systems being able to reduce H_2S levels to pipeline specifications.[4] DEA also partially removes carbonyl sulfide and carbon disulfide. However, the reaction of DEA with COS and CS_2 is lower than those of MEA, and results in different products that can be regenerated. Consequently, there are minimum DEA losses caused by reaction with these sulfur components.

7.3.1.4 Diisopropanolamine

Diisopropanolamine (DIPA) is a secondary amine, and its reactions with acid gases are the same as for MEA. However, degradation products from reactions with COS and CS_2 are easily regenerated. It is also noncorrosive and requires less heat for rich amine regeneration. DIPA systems can be used either selectively to remove H_2S (at low pressures) or to remove both CO_2 and H_2S (higher pressures).

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[4] The SNPA–DEA process is a modification of the DEA process, which has been used with considerable success for sweetening high-pressure gas streams with high acid gas concentrations, down to the level of approximately 1.6 ppm H_2S.

7.3.1.5 Methyldiethanolamine

Methyldiethanolamine (MDEA) is a tertiary amine and commonly used for removal of acid gases from gas streams. MDEA does not react with COS and CS_2, which avoids loss of solution. COS can be physically removed by MDEA only with very high solvent circulation rates, at which point the CO_2 is also removed quantitatively (Korens et al., 2002).

The various reactions between MDEA and acid gases have been described in the literature. In general, the H_2S is thought to react almost instantaneously with MDEA by proton transfer as is the case with primary and secondary amines:

$$H_2S + R_2NCH_3 \leftrightarrow R_2NCH_4^+ + HS^- \tag{7-7}$$

Since MDEA is a tertiary amine and does not have a hydrogen atom attached to the nitrogen, the CO_2 reaction can occur only after the CO_2 dissolves in the water to form a bicarbonate ion. The bicarbonate ion then undergoes an acid-base reaction with the amine to yield the overall CO_2 reaction:

$$CO_2 + H_2O + R_2NCH_3 \leftrightarrow R_2NCH_4^+ + HCO_3^- \tag{7-8}$$

Since the MDEA reaction with CO_2 is slow, the MDEA system selectively removes H_2S to pipeline specifications while “slipping” some of the CO_2 in the gas (Blanc et al., 1981). This high selectivity allows for a reduced solvent circulation rate, as well as a richer H_2S feed to the sulfur recovery unit. Other advantages include lower vapor pressure and solution losses, lower energy demand for regeneration, low corrosiveness, and higher resistance to degradation. However, in generic MDEA processes, amine regeneration of H_2S down to trace levels becomes progressively more difficult. In this regard, MDEA acidified with phosphoric, sulfuric, or other acids will be regenerated more easily. However, the level of acidification has a certain limit, as there is an impact on the pH value of the amine solution and the equilibrium curve of the acid gas solubility in the solvent. Alternatively, the new proprietary promoter system developed by BASF (sMDEA$_+$) shows a different behavior: a substantial increase in regeneration ability while hardly affecting absorption capacity. The solvent provides an option not only for grassroots designs but also for revamps in order to fulfill new H_2S regulations or even a capacity increase (Vorberg et al., 2010).

7.3.1.6 Sterically Hindered Amines

Sterically hindered amines are compounds in which the nitrogen atom of the amine molecule is partially shielded by neighboring groups so that larger molecules cannot easily approach and react with the nitrogen (Sartori et al., 1987).

The hindered amine concept is based on the reaction rates of the acid gases with different amine molecules. In the case of CO_2 removal, the capacity of the solvent can be greatly enhanced if one of the intermediate reactions, i.e., the carbamate formation reaction, can be slowed down by providing steric hindrance to the reacting CO_2. In addition to slowing down the overall reaction, bulkier substitutes give rise to less stable carbamates. By making the amine carbamate unstable, one can theoretically double the capacity of the solvent (Chakma, 1994).

ExxonMobil developed the FLEXSORB SE process, which is based on a family of proprietary sterically hindered amines in aqueous solutions or other physical solvents. The FLEXSORB SE process removes H_2S selectively or removes a group of acidic impurities (H_2S, CO_2, COS, CS_2, and mercaptans) from a variety of gas streams, depending on the solvent used. One version of the solvent, FLEXSORB SE Plus, is used to selectively remove H_2S, as well as organic sulfur impurities commonly found in natural gas. It is claimed that FLEXSORB SE Plus use in such services results in lower circulation rates than for other amines (Garrison et al., 2002).

7.3.1.7 Mixed Amines

The selectivity of MDEA can be reduced by addition of various amounts of primary or secondary amines as so-called promoters, which enhance CO_2 removal while retaining desirable characteristics of MDEA (Mandal et al., 2001). Promoters work by a shuttle mechanism and affect thermodynamics, but more importantly, they allow the reactivity of the mixture as a whole toward CO_2 to be closely controlled. Finding an optimum concentration for mixed amines (also called blended amines) strongly depends on the H_2S and CO_2 content of the sour gas, operating pressures, and sale gas specifications. After the blend formula is established, normal control schemes will be used to ensure the H_2S specification. However, the treated gas CO_2 content will not be controllable except by adjusting the blend composition from time to time (Weiland et al., 2003).

Amine mixtures are particularly useful for low-pressure applications because MDEA becomes less capable of CO_2 pickup sufficient enough to meet pipeline specifications. At higher pressures, amine mixtures appear to have little or no advantage over MDEA (Polasek et al., 1992).

7.3.1.8 Formulated or Specialty Amines

Among all amine solvent processes, mixed amines have been extensively used because they offer many advantages, such as their ability to meet the most stringent H_2S and CO_2 specifications. However, their performance as regards mercaptan/COS removal is extremely limited. A series of chemical activators used with methyl diethanolamine offers the most cost-effective answer to complete or controlled removal of acid gases as well as bulk removal of mercaptans and COS from sour natural gases. The BASF-activated MDEA (aMDEA) process is probably the most cost-effective solution today developed for such purposes.[5] The BASF-formulated MDEA solvent achieves a high degree of COS removal and retains appreciable selectivity for H_2S over CO_2. H_2S can be virtually completely removed while the COS removal level can be targeted to meet the overall sulfur removal regulations. This strategy would limit CO_2 coabsorption and may preclude the need for a COS hydrolysis unit upstream of the amine unit (Hugo and Wagner, 2000). The BASF aMDEA process is highly energy efficient due to the elevated acid gas loadings achievable with the solvent; this enables using low circulation rates and reduced energy consumption, as well as reducing required equipment size. Additional advantages include very low hydrocarbon coabsorption, no degradation products, no corrosion (mainly carbon steel equipment can be used), and low foaming tendency; also no reclaimer operation is necessary, and the solvent is nontoxic and biodegradable.

7.3.2 Process Description

The general process flow diagram for an amine sweetening plant varies little, regardless of the aqueous amine solution used as the sweetening agent (see Figure 7-2). The sour gas containing H_2S and/or CO_2 will nearly always enter the plant through an inlet separator (scrubber) to remove any free liquids and/or entrained solids. The sour gas then enters the bottom of the absorber and flows upward through the absorber in intimate countercurrent contact with the aqueous amine solution, where the amine absorbs acid gas constituents from the gas stream. Sweetened gas leaving the top of the absorber passes through an outlet separator (to recover entrained amines) and then flows to a dehydration unit (and compression unit, if necessary) before being considered ready for sale.

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[5] BASF has used piperazine in its activated MDEA formulations. Piperazine, even at low levels (about 5%), enhances the rate of CO_2 absorption an order of magnitude over non-activated MDEA (Rochelle et al., 2001).

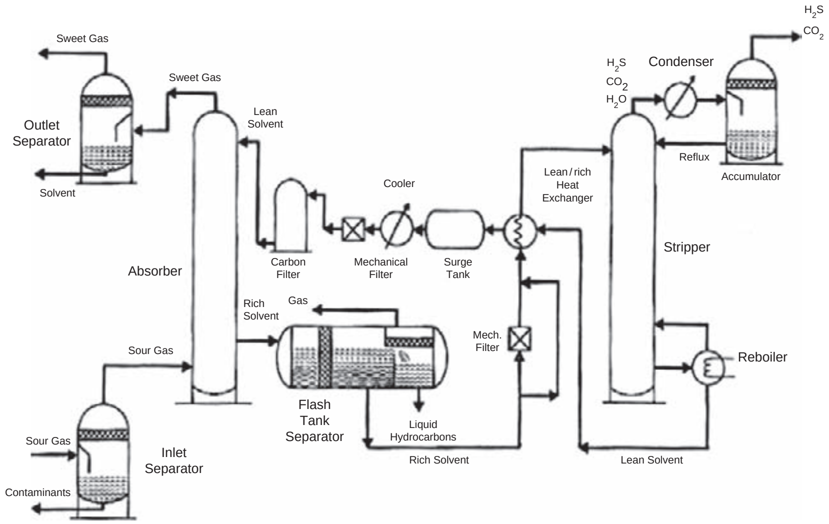

Figure 7-2 Schematic of amine gas sweetening process flow.

In many units, the rich amine solution is sent from the bottom of the absorber to a flash tank (a three-phase separator operating at a lower pressure than the absorber) to recover hydrocarbons that may have dissolved or condensed in the amine solution in the absorber. The flashed gas is often used locally as fuel. A small, packed tower with a lean amine wash may be installed on top of the flash drum to remove H_2S from the flashed gas if sweet fuel gas is required. The rich solvent from the flash tank is then preheated before entering the top of the stripper operating at low pressure (i.e., 18 psig). The amine-amine heat exchanger serves as a heat conservation device and lowers total heat requirements for the process. A part of the absorbed acid gases will be flashed from the heated rich solution on the top tray of the stripper. The remainder of the rich solution flows downward through the stripper in countercurrent contact with vapor generated in the reboiler. The reboiler vapor (primarily steam) strips the acid gases from the rich solution. The acid gases and the steam leave the top of the stripper and pass overhead through a condenser, where the major portion of the steam is condensed and cooled. The condensed steam is returned to the top of the stripper as reflux. The acid gases are separated in the reflux drum (accumulator) and sent to the flare or to processing. If the acid gas stream is CO_2 and contains no sulfur compounds, it is either vented to atmosphere or reinjected into the subsurface for enhanced oil recovery or for CO_2 sequestration. When the recovered acid gas stream contains sulfur compounds, venting or flaring typically is not an option (Hubbard, 2009). In these cases, if the acid gas stream is substantially H_2S, it frequently is sent to a sulfur recovery unit (SRU) utilizing the Claus process, where sulfur species are converted to elemental sulfur. If the feed gas to the Claus unit contains less than some 40 mole% H_2S, there will be too many inerts present in the feed that increase the size of the unit and reduce the sulfur recovery efficiency.[6] To avoid such problems, one can enrich the feed gas by selectively removing the H_2S from the regenerator off-gas in a second amine absorption step.[7] If the solvent absorbs large quantities of heavy hydrocarbons, an additional process may be required on the acid gas leaving the unit, or a high-efficiency burner must be installed in the Claus process to destroy these components. Hindered amines are used as an attractive approach for low-pressure acid gas enrichment. One disadvantage of this technology is the solvent cost of these proprietary amines, when compared to classic acid gas enrichment based on selective MDEA.

Lean amine from the bottom of the regenerator after precooling in the lean/rich amine exchanger passes to the amine surge tank. The amine surge tank has the capacity to hold the total amine inventory and is used to take system surges during operation. The tank also serves as a convenient location for amine makeup. Lean amine from the surge drum is further cooled in the lean amine cooler. The amine cooler serves to lower the lean amine temperature to the 100°F range. Higher temperatures of the lean amine solution will result in excessive amine losses through vaporization and also lower acid gas carrying capacity in the solution because of temperature effects. The cooled lean amine is pumped through a lean amine circulation pump and recirculated to the absorber. A slipstream of lean amine flows through a series of filters to remove particulate matters that can promote erosion/corrosion in the system, degradation of amine, and foaming.

There are many variations of the process flow scheme shown in Figure 7-2, including several flash regeneration steps, several absorbers in series, and several absorbers in parallel and split-flow configuration. These variations depend on the solvent used; the composition, pressure, and temperature of the feed gas; and the treating requirements. In addition, the operating conditions of the amine process depend on the type of the amine used. When one is considering generic amines (see Table 7-1), there is a

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[6] For low H_2S levels, there are also other sulfur recovery processes available that are not restricted by a minimum H_2S content in the feed gas.

[7] When an enrichment step is selected, this can be included in the acid gas removal step, where the H_2S/CO_2 ratio and process conditions will determine the applicability of this option. Alternatively, acid gas enrichment is achieved by a dedicated unit selectively absorbing H_2S from the acid gas produced by the main CO_2 removal unit. Due to the better selectivity, this option may be more economically attractive.

Table 7-1 Typical Operating Parameters for Commonly Used Generic Amines

Solvent	MEA	DGA	DEA	DIPA	MDEA
Typical concentration, wt%	15–20	45–50	25–30	30–40	35–50
Typical lean loading, mole/mole	0.1–0.15	0.05–0.1	0.05–0.07	0.02–0.05	0.004–0.01
Typical rich loading, mole/mole	0.30–0.35	0.35–0.40	0.35–0.40	0.30–0.40	0.45–0.55
Typical steam use, lb/gal	1.0–1.2	1.1–1.3	0.9–1.1	0.8–1.1	0.9–1.1
Heat of reaction with CO_2, Btu/lb	825	850	653	550	475
Heat of reaction with H_2S, Btu/lb	820	674	511	475	455

marked difference in solution concentration due to the physical-chemical properties and potential corrosion of the solution.

Proper amine selection, which can have a major impact on the performance and cost of a gas sweetening unit, depends on the compounds to be removed, the process conditions, the level of treatment necessary, process metallurgy, and other considerations such as process economics, reliability, and versatility factors of the design, while taking environmental constraints into account (Polasek and Bullin, 1994). While many of the recent published papers concerning amine selection or amine conversions have focused on the utilization of formulated amines over the generic amines, there are many recent cases in which these generic amines have been the best and even, perhaps, the only choice for recent new plant design (Jenkins and Haws, 2002). The principal advantage of a generic amine process design is the avoidance of a licensing fee and the flexibility and the economic advantage it gives the plant owner in being able to buy solvents and other services. The main disadvantage is the potential lack of technical support if the unit does not perform as designed. Obviously, experienced process licensors will be able to offer the most know-how based on actual operating data. However, physical and chemical property data and process performance data for proprietary MDEA formulations must be purchased from the solvent or process developers or licensors (Korens et al., 2002).

7.3.3 Operational Issues

The operating problems encountered in amine systems are usually linked to design, where trouble-free operation results from an appropriate design of the unit as well as a proper solution control.

Amine solutions frequently become contaminated with amine degradation products, heat-stable salts (HSS), heavy hydrocarbons, and particulates. Due to the closed-loop nature of amine processes, nonregenerable contaminants tend to accumulate and can cause operational problems such as capacity reduction, corrosion, and excessive foaming, resulting in unscheduled upsets and outages. Amine plant operational problems are often attributed to the accumulation of amine heat-stable salts. These heat-stable salts lead to costly maintenance problems such as corrosion, frequent filter replacement, foaming in the absorber, absorber plugging, heat exchanger fouling, and a reduction in the amount of amine available for gas treatment, thereby reducing the unit's productivity. In order to prevent the HSS from building up beyond critical limits, amine plant operators have been making conscious attempts to control impurities, especially HSS. There are basically four types of remediation action: partially purging the contaminated solution and replacing it with fresh amine, complete replacement of the contaminated amine, caustic addition, and reclaiming of the amine solutions. Continuous amine reclaiming, which ensures high solution purity necessary for reliable unit operation, is being increasingly recognized as the most effective solution for HSS-related problems. The method of reclaiming depends on the amine involved. MEA systems usually include a reclaimer, which is a kettle-type reboiler operating on a small side stream of lean amine solution. The temperature in the reclaimer is maintained such that the water and MEA boil to the overhead and are piped back to the stripper. The heat-stable salts remain in the reclaimer until the reclaimer is full. The reclaimer is then shut in and dumped to a waste disposal. Therefore, the impurities are removed, but the MEA bonded to the salts is also lost. For DEA systems, a reclaimer is not required because the reactions with COS and CS_2 are reversed in the stripper. The small amount of degradation products from CO_2 can be removed by a carbon filter on a side stream of lean solution. Because MDEA is more difficult to recover by conventional distillation methods, ion exchange methods are often used to reclaim MDEA. The HSS degradation products are removed by ion exchange resins, and the reclaimed MDEA is returned to the process (Bacon et al., 1988).

A comprehensive discussion of the operating issues and their troubleshooting methods, which are beyond the scope of this book, is given by Seagraves et al. (2010). However, as mentioned previously, proper solution control leads to more reliable and trouble-free operation. It is therefore important to test the amine behavior with appropriate analytical techniques. General analysis includes amine tests, acid gas loading, water content, solids and ash content, heat-stable salt content, and anion analysis. Moreover, a specific test for foaming ability could be of great interest to monitor the operation of the amine unit.

7.3.4 Design Considerations

Designing amine units for different solvents is similar. However, a reclaimer must be added in the case of more reactive amines (such as MEA) to recover these amines. In the following sections, some design aspects for the major equipments used in amine units are described.

7.3.4.1 Absorber

Amine absorber (contactor) uses countercurrent flow through a trayed or packed tower to provide intimate contact between the lean amine solvent and the sour gas so that the H_2S and CO_2 molecules can transfer from the gas phase to the solvent-liquid phase. The temperature of the amine solution entering the absorber is very critical to the operation of the absorber. If the amine inlet temperature is below the sour gas inlet temperature, some gas condensate may form in the amine solution and cause foaming of the amine. This can lead to carryover of amine from the system and other contacting problems, giving rise to a poor sweetening process. In order to prevent the formation of liquid hydrocarbons, the lean amine solution entering the absorber should be 10°F–15°F above the absorber inlet sour gas temperature.

In amine absorber, the degree of sweetening achieved is largely dependent on the number of trays or the height of packing available in the absorber. The absorber is typically designed with 20 float-valve trays (or an equivalent number of packing for smaller vessels). Tray spacing should be sufficient for liquid foam disengaging space.

In most cases, a mist eliminator pad is installed near the gas outlet of the absorber (the distance between the top tray and the mist pad is 3–4 feet) to trap entrained solvent, and an outlet knockout drum is provided to collect solvent carryover. This requirement is a function of contactor design and the perceived likelihood of entrainment or foaming. If foaming is expected, a water spray to collapse the foam may be appropriate.

Some absorbers have a water wash consisting of two to five trays at the top of the absorber to minimize vaporization losses of amine, which is often found in low-pressure monoethanolamine systems. Selective solvents typically have multiple feed points on the absorber to allow slippage of some acid gas, usually CO_2 (Bullin, 2003). If carbon dioxide absorption is desired, all of the lean amine should, in general, be fed on the top tray, thus utilizing all available stages.

7.3.4.2 Pumps

The amine reflux and booster pumps are normally centrifugal, inline, or horizontal pumps. Seal systems are an important consideration on the reflux pumps when treating H_2S gases, as seal failure will result in H_2S emissions. Selection of the circulation pump depends on the contactor operating pressure and the amine circulation rate. Positive displacement pumps should be considered for 250 gpm and below and for high discharge pressures. Centrifugals are considered for low head cases and large-volume, high head cases.

The solvent circulation rate should always be reduced to the minimum required to meet the treating needs for the unit subject to the constraints such as maximum loading of the rich amine. Since the reboiler duty is almost always tied directly to the circulation rate, lower circulation rates reduce the overall energy requirements. Lower circulation rates also tend to increase the CO_2 slip and can improve the quality of the feed to the sulfur recovery unit (Bullin, 2003).

7.3.4.3 Flash Tank

The rich amine solution from the high-pressure absorber enters a flash tank (operating at a pressure of 15 psig or less) that allows the lightest of the hydrocarbons to flash. A small percentage of acid gases will also flash when the pressure is reduced. Flash gas can be disposed of to the vapor recovery system, flare, incinerator, or fuel gas system.[8] The heavier hydrocarbons separate from the aqueous amine, forming the upper liquid layer (due to a lower density than the aqueous amine) and can be skimmed off the top. Flash tanks therefore must be provided with skimming nozzles at locations around the normal liquid operating level to allow removal of liquid hydrocarbon buildup or be built for three-phase operation. A good rule of thumb for hydrocarbon vapor pickup is 2 scf/gal, and typical residence time is between 5 and 30 minutes.

7.3.4.4 Lean/Rich Amine Exchanger

The lean/rich amine heat exchanger is a heat conservation device used to decrease energy consumption. It preheats the rich amine solution and reduces the duty of the reboiler. It also cools the lean amine and reduces the duty of the aerial cooler.

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[8] When flash gas containing H_2S is used as fuel, it should be treated with the lean amine. The sulfur specification for treated fuel gas is typically 50–160 ppmv H_2S, depending on the environmental regulatory requirement.

Shell-and-tube and plate-frame exchangers are typical choices for exchangers. The rich solution is passed through the tubes, which are usually made of stainless steel, with a low inlet velocity (2–3.5 ft/sec) to minimize corrosion. Temperature of the rich outlet should be approximately 200°F,[9] and it should lean down to 160°F to maximize heat recovery. Using two or more exchangers in series is common practice to accommodate a temperature cross (Manning and Thompson, 1991). Pressure drop on the cold side should be kept at 5 psig or less, and depending on whether booster pumps are in the design, hot-side pressure drop should kept at or below 2 psig. Strainers should be considered on both sides of the exchanger to keep solids out of the exchanger.

7.3.4.5 Stripper

The amine stripper (regenerator) is usually a trayed tower (with approximately 20 trays), although packed towers are occasionally used. To minimize amine vaporization loss, there may be a water wash section at the top of the column with an additional 4 to 6 trays (Kohl and Nielsen, 1997). Trays are normally on 24-inch spacing, and liquid and jet floods in the 65%–75% range with a 75% foam factor. It is highly recommended that trays be stainless steel due to the corrosivity of the environment.

The critical parameter for the amine stripper is the column bottom's pressure. Selecting a high pressure would raise the temperature in the column bottom and lead to amine degradation. Selecting a very low pressure would result in insufficient acid gas pressure to the SRU.[10] In practical situations, it is generally recommended to operate amine regenerators at approximately 2.1–2.2 bar (Kohl and Nielsen, 1997).

7.3.4.6 Reboiler

The amine reboiler provides the heat input to an amine stripper, which reverses the chemical reactions and drives off the acid gases. Steam is normally used as the heat medium;[11] there are alternatives, but tube wall temperature must be kept below some maximum for each amine. The reboiler duty should be maintained as low as possible but must be adequate to regenerate

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[9] A set point of 210°F has been well established over the years in both operation and the literature to avoid acid gas breakout and corrosion problems in the equipment.

[10] The SRU needs about 12 psig acid gas pressure at the inlet to the unit. This is required to account for the pressure drop across the SRU and TGTU sections with reasonably sized equipment.

[11] The steam ratio, which is defined as the mass flow of steam per volume of amine circulation, plays a major role not only in the quality of the lean amine but also in its composition. Common industry practice is to set the 1 lb/gal specification for the reboiler steam (Addington and Ness, 2009).

the amine solution sufficiently to meet the sweet gas requirements and to ensure that the CO_2 loadings in the reboiler do not cause excessive corrosion. Higher reboiler duties do not reduce circulation rates to any degree and just consume energy (Bullin, 2003). Typically, heat flux rate should be kept in the 7,500–10,000 Btu/hr/ft^2 range to assure no surface burning of the solvent.

7.3.4.7 Reflux Condenser and Accumulator

The amine reflux condenser is required to cool the overhead gases and condense the overhead steam to water. The overhead condenser may be either an aerial cooler or shell-and-tube. The inlet temperature to the cooler can be found using the partial pressure of the overhead steam to determine the temperature from steam tables. The cooler outlet temperature is typically 130°F–145°F depending on the ambient temperature. Condenser tubes should be made of stainless steel, as this is a wet, acid gas environment and sloped to the outlet side.

The reflux accumulator is a two-phase separator used to separate the acid gases from the condensed water.[12] The water is accumulated and pumped back to the top of the stripper as reflux. The acid gases are directed to a vent, incinerator, or sulfur recovery unit. The reflux drum pressure is usually determined by the pressure required to deliver the acid gas to a sulfur recovery unit and/or the pressure required to provide the desired amine regenerator bottom temperature. The minimum reflux drum pressure is usually about 5 psig, with an upper pressure limit of about 15 or 20 psig depending on the amine and the maximum allowed operating temperature for that amine (Kohl and Nielsen, 1997).

7.3.4.8 Reclaimer

On primary amines, a reclaimer is used to permit cleanup of amine contaminated with heat-stable salts. The reclaimer works by converting the amine salts to sodium salts and boiling the amine away from the resulting salt solution. Neutralization with caustic has long been practiced as an economical means of displacing bound amine from heat-stable salts.

Reclaimer is usually a kettle-type reboiler with 316 SS tubes and provision for easy addition of caustic. It frequently contains spargers to pass steam directly through the residue sludge to assist in reacting caustic. Steam supply must be hot enough to vaporize the amine and is usually in the 150–250 psig range.

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[12] The reflux accumulator is sized to provide for vapor-liquid separation and for 5 minutes of reflux pump capacity holdup.

7.3.4.9 Lean Amine Surge Tank

The lean amine from the heat exchanger passes to the amine surge tank, which provides a steady positive suction head pressure for the amine solution booster pumps. The surge tank is also used for the addition of makeup solution to maintain the correct amount and concentration of amine solution in the system.

An amine surge tank is usually an API tank with fuel gas or a nitrogen blanket. It may also be a pressure vessel. The tank may need steam coils to assure that the solution viscosity does not increase in cold weather. Great care must be taken in venting such tanks; usually a vent to an incinerator or heater firebox is acceptable.

7.3.4.10 Lean Amine Cooler

The amine cooler is typically an air-cooled, forced draft with automatic louvers for temperature control, which lowers the lean amine temperature before it enters the absorber. If cooling water is available, a plate heat exchanger (PHE) is an excellent choice, provided the operating pressure of the lean system is within the limitations of a PHE. The lean amine entering the absorber should be approximately 10°F warmer than the sour gas entering the absorber. Lower amine temperatures may cause the gas to cool in the absorber and thus condense hydrocarbon liquids. Higher temperatures would increase the amine vapor pressure and thus increase amine losses to the gas. The duty for the cooler can be calculated from the lean-amine flow rate, the lean-amine temperature leaving the rich/lean exchanger, and the sour-gas inlet temperature.

7.3.4.11 Amine Filter

An amine filter may be full flow or slip stream (10%–20%), on the lean or the rich side of the system and may be cartridge, bag, precoat, sand, etc. Most systems apply lean slipstream cartridge or bag filters. A reasonable maximum is 10 μm, and a reasonable minimum for filter element porosity is 5 μm. A carbon filter, which removes entrained hydrocarbons but not dissolved acids or salts, must be immediately downstream of the lean amine filter to ensure it does not become clogged with particulates.

7.3.5 Hot Potassium Carbonate Process

The hot potassium carbonate process employs an aqueous solution of potassium carbonate (K_2CO_3) to remove both CO_2 and H_2S. It also removes some

COS and CS_2 through hydrolysis. The following reactions, which are reversible based on the partial pressures of the acid gases, occur in this process:

$$K_2CO_3 + CO_2 + H_2O \leftrightarrow 2KHCO_3 \tag{7-9}$$

$$K_2CO_3 + H_2S \leftrightarrow KHS + KHCO_3 \tag{7-10}$$

It has been demonstrated that the process works best near the temperature of reversibility of the reactions (Benson et al., 1956). However, as can be seen from those reactions, a high partial pressure of CO_2 is required to keep $KHCO_3$ in solution, and H_2S will not react if the CO_2 pressure is not high. For this reason, this process cannot achieve a low concentration of acid gases in the exit stream. Special designs or a two-stage system are therefore required for removing H_2S to pipeline specifications or to reduce CO_2 to low levels.

The hot carbonate process, which is shown in Figure 7-3, is referred to as the "hot" process because both the absorber and the stripper operate at elevated temperatures in the range of 230°F–240°F. In this process, the sour gas enters at the bottom of the absorber and flows countercurrently to the carbonate liquid stream. The sweet gas exits at the top of the absorber. The rich carbonate solution exits from the bottom of the absorber and is flashed in the stripper, where acid gases are driven off. The lean carbonate solution is pumped back to the absorber. The lean solution may or may not be cooled slightly before entering the absorber. The main disadvantage of this solution is linked to the need to replace the water that saturates the treated gas at high temperatures. To reduce this phenomenon while

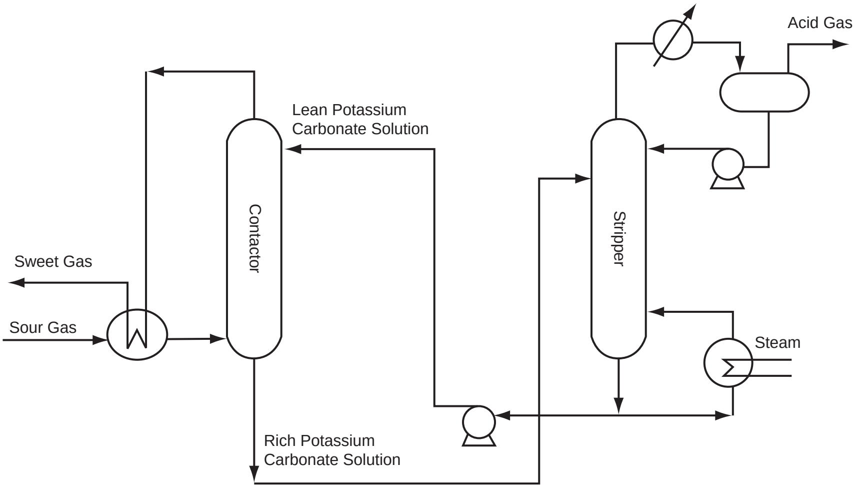

Figure 7-3 Typical hot potassium carbonate process flow diagram.

simultaneously improving the unit's energy balance, a common practice is to install a heat exchanger upstream of the absorber, between the sour gas and the sweet gas, with the recovery of condensed water from the latter.

In this process, the entire system is operated at high temperature to increase potassium carbonate solubility. Thus, a dead spot where the solution is likely to cool and precipitate should be avoided. If solids do precipitate, the system may suffer from plugging, erosion, or foaming. Potassium carbonate causes general stress corrosion of the unit. Therefore, all carbon steel must be stress-relieved to limit corrosion. Various corrosion inhibitors are available to decrease corrosion. However, the solvents react with some corrosion inhibitors and cause erosion of the unit.

The hot potassium carbonate process is particularly useful for removing large quantities of CO_2. The main advantages of carbonate solutions for CO_2 removal are the high chemical solubility of CO_2 in the carbonate/bicarbonate system and low solvent costs. The major difficulty is a relatively slow reaction in the liquid phase, causing low mass transfer rates and therefore requiring a large contact surface. Process improvements have been obtained through the use of catalytic additions and promoters to the solution, such as DEA, arsenic trioxide, selenous acid, and tellurous acid. In applications for the removal of hydrogen sulfide, tripotassium phosphate (K_3PO_4) may be used. These activators increase the performance of the hot potassium carbonate system by increasing the reaction rates both in absorber and stripper. In general, these processes also decrease corrosion in the system. Kohl and Nielsen (1997) thoroughly discussed the basic process and its many modifications.

7.3.6 Physical Solvent Processes

Various organic liquids (solvents) are used in these processes to remove acid gases from gas streams by physical absorption without chemical reaction. The driving force of this process is the high solubility of acid gases in the organic solvent. Solubility depends first and foremost on partial pressure and secondarily on temperature. In most cases, solubility increases as the temperature decreases and the pressure increases. The solvents are then regenerated by either heating or pressure reduction. If H_2S is present at only very low concentrations or is entirely absent, the regeneration of the solvent by flashing to atmospheric pressure or vacuum is usually applicable since CO_2 concentrations as high as 2 mole% or 3 mole% can often be tolerated in the product gas. When H_2S is present in significant amounts, thermal regeneration is usually necessary to accomplish the thorough stripping of the solvent needed to reach stringent H_2S purity requirements

(Burr and Lyddon, 2008). Heat requirements are usually much less for physical solvents than for chemical solvents since the heat of desorption of the acid gas for the physical solvent is only a fraction of that for chemical solvents. The circulation rate of the physical solvent may also be less, particularly when the acid gas partial pressure is high. However, physical solvent processes have a high affinity for heavy hydrocarbons. If the natural gas stream is rich in heavy hydrocarbons, then the use of a physical solvent may result in a significant loss of heavy hydrocarbons.

For physical solvents to be economically usable, they must meet certain criteria such as high selectivity for acid gases, very low vapor pressures, low viscosity, and heat stability, and they must be noncorrosive to metals. Only a few solvents meet all these criteria (Kohl and Nielsen, 1997). Among the ones that have been used commercially include dimethyl ether of polyethylene glycol (DEPG), propylene carbonate (PC), N-methyl-2-pyrrolidone (NMP), and methanol (MeOH). However, a detailed analysis must be performed to determine the most economical choice of solvent based on the feed gas composition, process objectives, and characteristics of the solvents. Designers can easily perform comparisons for these physical solvents using verified process simulation programs such as ProMax (Burr and Lyddon, 2008). It should be noted that some physical solvent-based process licensors claim that physical solubility properties of their solvents in some commercial process simulation packages are inaccurate and could produce misleading results. Therefore, the process simulation, design, and optimization tasks should involve the collaboration of the selected vendor.

The following sections give brief descriptions of the processes utilizing those solvents.

7.3.6.1 Selexol Process

The Selexol process uses Selexol solvent, a physical solvent made of dimethyl ethers of polyethylene glycol (DEPG),[13] to selectively remove H_2S/COS plus remove CO_2 in bulk from gas streams. This process can also be used to remove and control mercaptan sulfur. The Selexol process was first commercially used in the late 1960s. The technology is currently owned and licensed by Dow, with commercial packages available from a variety of companies, including UOP, Udhe, and Coastal.

There is no universal or standard flow sheet for DEPG units. In all the process flowsheets, the sour gas enters the bottom of the absorber, contacts

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[13] $CH_3O(C_2H_4O)_nCH_3$, where n is between 3 and 9.

with lean DEPG solvent from the top countercurrently, and leaves at the top as sweet gas, while the solvent flows out of the absorber at the bottom as the rich solvent. Desorption of the rich Selexol solvent can be accomplished either thermally, by flashing, or by stripping gas. Selective H_2S removal with deep CO_2 removal usually requires a two-stage process with two absorption and regeneration columns. H_2S is selectively removed in the first column by a lean solvent that has been thoroughly stripped with steam, while CO_2 is removed in the second absorber. The second-stage solvent can be regenerated with air or nitrogen for deep CO_2 removal, or using a series of flashes if bulk CO_2 removal is required (Burr and Lyddon, 2008). DIPA can be added to this solvent to remove CO_2 down to pipeline specifications. This addition will increase the stripper heat duty; however, this duty is relatively low.

In the Selexol process, it is sometimes necessary to reduce the operating temperature to increase acid gas solubility and reduce solvent circulation rate. However, Selexol solvent has a high viscosity, which can significantly affect mass and heat transfer as the solvent temperature is reduced. The Selexol solvent is suitable for operation at temperatures up to 347°F, while, for stability reasons, the minimum operating temperature is usually 0°F (Kohl and Nielsen, 1997).

7.3.6.2 Fluor Process

The Fluor process (licensed by Fluor Daniel, Inc.) uses propylene carbonate to remove CO_2, H_2S, C_2^+ hydrocarbons, COS, CS_2, and H_2O from natural gas. Thus, in one step, the natural gas can be sweetened and dehydrated.

The Fluor process is used to remove the bulk CO_2 down to 3%. It can be also used for treating feed gases containing low levels of H_2S, typically less than 20 ppmv. However, improved stripping allows treatment to meet H_2S sales gas requirement for gases containing up to 200 ppmv H_2S. This new improvement in stripping uses medium pressure (125 psia) flash gas as the stripping medium in a vacuum stripper (Mak et al., 2007). It should be noted that hydrocarbon losses increase as the amount of stripping gas increases.

The Fluor solvent can operate at lower temperatures without becoming too viscous for good mass transfer. The operating temperature for propylene carbonate is limited to greater than 0°F and a maximum operating temperature of 149°F.

Propylene carbonate has low solubility for light hydrocarbons, which results in lower hydrocarbon losses in the vent gas stream. A new Fluor process improvement involves feed chilling to 0°F in order to reduce absorption of hydrocarbons. Chilling also increases the solvent's acid gas holding capacity, resulting in a lower overall solvent circulation rate and lower plant cost.

7.3.6.3 Purisol

The Purisol process (licensed by Lurgi GmbH) uses N-methyl-2-pyrrolidone (NMP) as a solvent to remove H_2S, CO_2, RSH, and H_2O. The feature of this solvent is that it is highly selective for H_2S. COS is not as soluble as H_2S, but it is hydrolyzed by the NMP solvent.

The flow schemes used for this solvent are similar to those used for DEPG. However, regeneration is accomplished by two strippers, where dissolved hydrocarbons are stripped off as fuel gas nitrogen in the first drum and acid gases are stripped in the second stripper; regenerated NMP is recycled back to the absorber. The process can be operated either at ambient temperature or with refrigeration down to about 5°F.

7.3.6.4 Methanol-Based Processes

Historically, methanol was the first commercial organic physical solvent and has been used for hydrate inhibition, dehydration, gas sweetening, and liquids recovery (Kohl and Nielsen, 1997; Font-Freide, 2004). Most of these applications involve low temperature where methanol's physical properties are advantageous compared with other solvents that exhibit high viscosity problems or even solids formation. Operation at low temperatures tends to suppress methanol's most significant disadvantage: high solvent loss. Furthermore, methanol is relatively inexpensive and easy to produce, making the solvent a very attractive alternative for gas processing applications.

The use of methanol has been further exploited in the development of the Rectisol process (licensed by Lurgi GmbH), which is used for H_2S, COS, and bulk CO_2 removal, whereby organic and inorganic impurities are also removed. The Rectisol process uses chilled methanol at a temperature of about –40°F to –80°F. Unfortunately, at cold temperatures, methanol also has a high affinity for hydrocarbon constituents of the gas streams. Chilled methanol will also remove many metallic trace components from the gas, including mercury. If mercury is present in the gas, then a carbon bed may be required ahead of the Rectisol unit to prevent the mercury from forming metal amalgams in the low-temperature sections of the process (Korens et al., 2002).

There are two versions of the Rectisol process: the two-stage and the once-through. The first step of the two-stage process is desulfurization before shift conversion; the concentrations of hydrogen sulfide and carbon dioxide are about 1 and 5 vol%, respectively. Regeneration of the methanol following the desulfurization of the feed gas produces high sulfur feed for sulfur recovery. The once-through process is applicable only for high-pressure partial oxidation products. The once-through process is also applicable when the hydrogen sulfide to carbon dioxide content is unfavorable, in the neighborhood of 1:50 (Esteban et al., 2000). The main advantage of the Rectisol process over other processes is the use of a cheap, stable, and easily available solvent; a very flexible process; and low utilities. However, Rectisol's complex scheme and the need to refrigerate the solvent are its main disadvantages, resulting in high capital and operating costs.

The Institut Français du Pétrole (IFP) has also developed a process using methanol that has the simultaneous capability to dehydrate, remove acid gas, and control hydrocarbon dew point (Rojey et al., 1990). The IFPEXOL-1 is used for water removal and hydrocarbon dew point control; the IFPEXOL-2 process is used for acid gas removal. The IFPEXOL-2 process is very similar to an amine-type process except for the operating temperatures. The absorber operates below −20°F to minimize methanol losses, and the regenerator operates at about 90 psi. Cooling is required on the regenerator condenser to recover the methanol. This process usually follows the IFPEXOL-1 process, so excessive hydrocarbon absorption is not as great a problem (Minkkinen and Jonchere, 1997).

7.3.7 Hybrid Solvents

Hybrid solvent processes use a mixture of amines and physical solvents to take advantage of the best characteristics of both. Depending on the physical solvent-amine combination, nearly complete removal of H_2S, CO_2, and organic sulfur compounds is possible. A few other advantages are higher acid gas loading, lower energy requirements for regeneration, lower corrosion rates, and lower foaming tendency. The most widely known hybrid solvent process is the Shell Sulfinol process, which applies a mixture of sulfolane, water, and diisopropanolamine (DIPA) or methyldiethanolamine (MDEA), Sulfinol-D, and Sulfinol-M, respectively. This process is used to selectively remove H_2S, COS, RSH, and other organic sulfur compounds for pipeline specification, while co-absorbing only part of the CO_2 (Huffmaster, 1997; Klinkenbijl et al., 1999; Rajani and Bowerbank, 2004).

Today, hybrid solvents are often used for increased mercaptan removal efficiencies (Bedell et al., 2008). Like the physical solvent processes, the hybrid systems may absorb more hydrocarbons (including BTEX), resulting in hydrocarbon losses in the separated acid gases. An effective solution for minimizing hydrocarbon losses when using a hybrid solvent in the gas sweetening unit (GSU) is to remove only part of the mercaptans in the GSU, thus mitigating hydrocarbon coabsorption. The remaining amount of mercaptans not removed by the GSU can be removed from the gas stream using a downstream molecular sieve unit (Rajani and Bowerbank, 2004; Carlsson et al., 2007). It should be noted that Total has identified and developed a new hybrid solvent formulation offering both good affinity toward mercaptans and high selectivity toward hydrocarbons. The newly developed process using the hybrid solvent takes advantage of Total's extensive know-how and experience in acid gas removal with amine mixtures and is a good contender for the development of sour gas fields with increasingly stringent commercial gas specifications (Lallemand et al., 2008).

7.4 DRY-BED PROCESSES

The dry-bed processes use a fixed bed of solid materials to remove acid gases either through ionic bonding or chemical reactions. These processes are limited to sulfur-removal applications and in general are applied when the amount of sulfur to be removed is small, less than 100 lbm/day (Hubbard, 2009). Typically, in solid-bed processes, the gas stream will flow through a fixed bed of solid materials that remove the acid gases and hold them in the bed. When the bed is saturated with acid gases, it will be regenerated or replaced. The dry-bed processes can be divided into two categories: physical adsorption (i.e., molecular sieve process) and chemical absorption.

7.4.1 Physical Adsorption

Adsorption is the selective concentration of one or more components of a gas at the surface of a microporous solid called "adsorbent." The attractive forces holding the adsorbate on the adsorbent are weaker than those of chemical bonds, and the adsorbate can generally be released (desorbed) by raising the temperature or reducing the partial pressure of the component in the gas phase. When an adsorbed component reacts chemically with the adsorbent, the process is called chemisorption and desorption is generally not possible. The main advantage of physical adsorption over chemical or physical absorption is its simple and energy-efficient operation and regeneration, which can be achieved with a pressure swing or temperature swing cycle. However, the preferred method to regenerate adsorbents is with a hot gas such as N_2 or contaminant-free light hydrocarbon gas. The heating regeneration gas flow is upward through the bed, and it is recommended to maintain the cooling gas flow also in the same upward direction to take advantage of "thermal pulsing" in the bed and minimize the regeneration heat requirement, therefore reducing the energy consumption. The hot regeneration gas is then cooled and processed by an amine/physical solvent unit to remove H_2S and organic sulfur compounds from the regeneration gas, which can be sent to a sulfur recovery unit. The sweet regeneration gas rejoins the main gas stream downstream of the solid beds.

Solid bed adsorption systems for sweetening and dehydrating natural gases can be optimized by understanding the inherent advantages and disadvantages of commonly used adsorbents. Many types of adsorbents exist, of which the most important ones are molecular sieves, activated alumina, silica gel, and activated carbon. Molecular-sieve adsorbents have become widespread for removing contaminant from natural gas. The sieve bed can be designed to sweeten and dehydrate simultaneously. Molecular sieves are zeolite-based adsorbents consisting of crystalline aluminosilicates and clay. They possess highly localized polar charges on their surface that act as adsorption sites for polar materials at even very low concentrations. The typical continuous molecular sieve configuration consists of two or more fixed-bed adsorbers and regeneration facilities. With two adsorption beds, one bed is treating the feed gas while the other is being regenerated. Figure 7-4 shows a process scheme where two adsorption beds are used in series, and the sour vent gas is flared, simultaneously sweetening and drying the natural gas.

During adsorption, there are four major zones in a molecular sieve bed. as shown in Figure 7-5. Water occupies the position closest to the inlet followed by RSH, H_2S, and CO_2. During operation, these zones progress toward the outlet, and when the key contaminant (usually H_2S) reaches the end, the bed must be regenerated.

Molecular sieves are available in four different types (3A, 4A, 5A, and 13X). These types are unique and different from each other based on its pore size and chemical composition.[14] Type 4A molecular sieves are very effective for H_2S removal and can produce gas containing extremely low levels of this impurity. Small quantities of CO_2, which are about the same size as H_2S molecules, will become trapped in the pores. In this way, small portions of CO_2 are removed. However, CO_2 will obstruct the access of H_2S to active sites and decrease the effectiveness of the pores.[15] Light and heavier mercaptans removal duties are also done with the larger-pored 5A and 13X types,

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[14] Molecular sieves are classified by their nominal pore diameter in Angstroms.

[15] CO_2 removal with molecular sieves will be most attractive when the feed gas contains less than 1.5% CO_2.

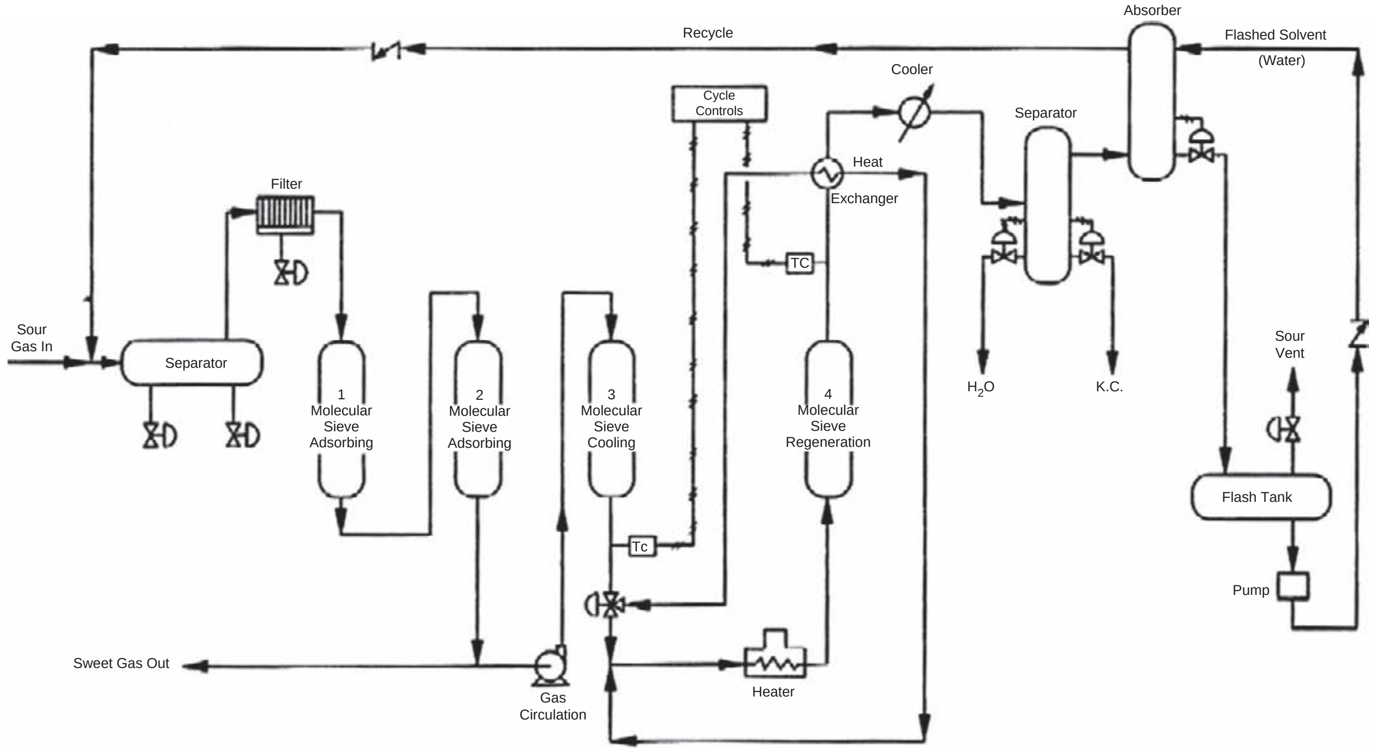

Figure 7-4 The molecular sieve process flow diagram (Thomas and Clark, 1967).

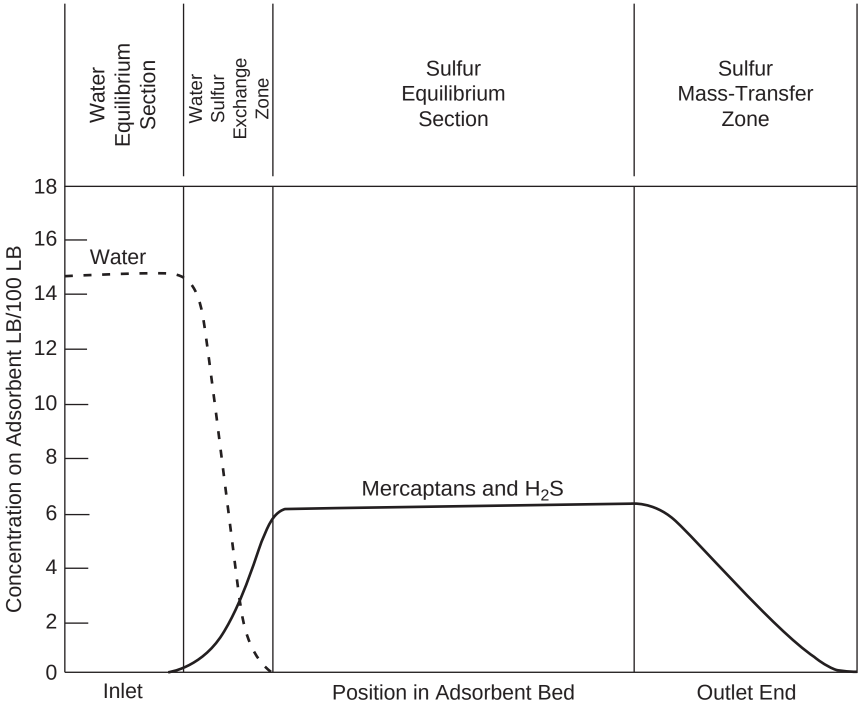

Figure 7-5 Adsorption zone in molecular sieve bed (Kohl and Nielsen, 1997).

respectively. However, heavy hydrocarbons (like BTX) and other impurities will coadsorb with mercaptans. Consequently, it is possible for these compounds to crack and form coke if the thermal wave "outruns" them during regeneration. Therefore, if the impurities are not desorbed and flushed out before the thermal wave arrives, they may react in some fashion, possibly catalyzed by the solid itself. Again, reduced regeneration temperatures will tend to reduce these undesirable side reactions.

Standard molecular sieves tend to display poorer chemical resistance (especially to acids), are more prone to coking, and require more energy for their effective regeneration. However, molecular sieve suppliers offer "acid resistant" molecular sieves. This special grade of molecular sieve is manufactured with a binder that is susceptible to acid attack and breakdown in the presence of H_2S/CO_2 and water.

Molecular sieves can dehydrate natural gas to 0.1 ppmw and reduce the sulfur content to 4 ppmv. However, molecular sieves' application for acid gas removal is generally limited to small gas streams operating at moderate pressures. Due to these operating limitations, molecular sieve units have seen limited use for gas sweetening operations. They are generally used for dehydration of sweet gas streams and for polishing applications following one of the other treating processes.

7.4.2 Chemical Absorption

For sour gas wells where the total production is small and the location remote from a gas processing plant, the conventional sweetening processes discussed previously may be uneconomical due to the large investment required. In such cases, the option of using other cheaper sweetening processes, such as chemical absorption processes, must be considered.

Chemical absorption processes have the common requirement that the process be operated as a batch system where, at the end of the cycle, the chemical agent must be changed in order to continue treating. Although not as widely used as liquid-phase continuous processes, these processes offer advantages such as simplicity and process efficiency (almost independent of pressure). However, these processes are limited to sulfur-removal applications (only H_2S is generally removed) and in general are applied when the amount of sulfur to be removed is small, less than 400 lbm/day. Currently marketed processes are described in the following sections.

7.4.2.1 Metal Oxide Processes

The metal oxide processes scavenge hydrogen sulfide from gas streams through reactions with solid-based media. These dry sorption processes are governed by the reaction of a metal oxide with H_2S to form a metal sulfide compound. For regenerable reactions, the metal sulfide compound can then react with oxygen to produce elemental sulfur and a regenerated metal oxide.

The primary metal oxides used for dry sorption processes are iron oxide and zinc oxide.

7.4.2.1.1 Iron Sponge Process

The iron sponge fixed-bed chemical absorption is the oldest and still the most widely used batch process. This process tends to be highly selective for H_2S and does not normally remove significant quantities of carbon dioxide. As a result, the hydrogen sulfide stream from the process is usually high purity.

In this process, the inlet gas is fed at top of the fixed-bed reactor filled with hydrated ferric oxide (Fe_2O_3). The basic reaction is

$$2Fe_2O_3 + 6H_2S \rightarrow 2Fe_2S_3 + 6H_2O \qquad (7\text{-}11)$$

The reaction requires the presence of slightly alkaline water and a temperature below 110°F. A pH level on the order of 8–10 should be maintained through the injection of caustic soda with the water.

The bed is regenerated by controlled oxidation as

$$2Fe_2S_3 + 3O_2 \rightarrow 2Fe_2O_3 + 6S \tag{7-12}$$

Some of the sulfur produced might cake in the bed, and oxygen should be introduced slowly to oxide this sulfur. If oxygen is introduced quickly, the heat of reaction may ignite the bed:

$$S_2 + 2O_2 \rightarrow 2SO_2 \tag{7-13}$$

Generally, the iron oxide process is suitable only for small to moderate quantities of hydrogen sulfide (300 ppm) operating at low to moderate pressures (50–500 psig). Removal of larger amounts of hydrogen sulfide from gas streams requires a continuous process, such as the ferrox process or the Stretford process. The ferrox process is based on the same chemistry as the iron oxide process except that it is fluid and continuous. The Stretford process employs a solution containing vanadium salts and anthraquinone disulfonic acid (Maddox, 1982).

7.4.2.1.2 Zinc Oxide Process

The zinc oxide process is also used for hydrogen sulfide removal from the gas stream. It uses a solid bed of granular zinc oxide to react with hydrogen sulfide as

$$ZnO + H_2S \rightarrow ZnS + H_2O \tag{7-14}$$

In this process, the exit H_2S concentration can be as low as 1 ppm at a temperature of about 572°F. The process has been decreasing in use due to the difficulty of disposing zinc sulfide, which is considered a heavy metal.

7.4.2.2 Slurry Processes

In the slurry processes, slurries of iron oxide have been used to selectively absorb hydrogen sulfide. Two different slurry processes are described next.

7.4.2.2.1 Chemsweet Process

The Chemsweet process (developed by NATCO) is a zinc oxide–based process for the removal of hydrogen sulfide from natural gas. The Chemsweet white powder, which is a mixture of zinc oxide, zinc acetate, and a dispersant to keep the zinc oxide particles in suspension, is mixed one part in five parts of water. The presence of CO_2 is not important as the pH is low enough to prevent significant absorption of CO_2, even when the ratio of CO_2 to H_2S is high (Manning and Thompson, 1991).

The following reactions are performed in this process:

$$\text{Absorption: } ZnAc_2 + H_2S \leftrightarrow ZnS + 2HAc \tag{7-15}$$

$$\text{Regeneration: } ZnO + 2HAc \leftrightarrow ZnAc_2 + H_2O \tag{7-16}$$

$$\text{Overall: } ZnO + H_2S \leftrightarrow ZnS + H_2O \tag{7-17}$$

The Chemsweet process can treat gas streams with high hydrogen sulfide concentration and has been operated between pressures of 90 and 1400 psia. Mercaptans concentrations in excess of 10% of the hydrogen sulfide concentration in the gas stream can be a problem. Some of the mercaptans will react with the zinc oxide and be removed from the gas. The resulting zinc mercaptides [Zn(OH)RH] will form a sludge and possibly cause foaming problems (GPSA, 2004).

7.4.2.2.2 Sulfa-Check Process

The Sulfa-Check process (marketed by NALCO) selectively removes hydrogen sulfide from natural gas, in the presence of carbon dioxide (Dobbs, 1986). This process is accomplished in a one-step single-vessel design using an aqueous solution of sodium nitrite ($NaNO_2$) buffered to stabilize the pH above 8.

This process is generally operated at ambient temperature and produces a byproduct slurry of sulfur and sodium salts:

$$NaNO_2 + 3H_2S \leftrightarrow NaOH + NH_3 + 3S + H_2O \tag{7-18}$$

Carbon dioxide in the gas also reacts with the sodium hydroxide to form sodium carbonate and sodium bicarbonate. Sodium nitrite and sodium hydroxide are consumables in the processes and cannot be regenerated. When the chemical is spent, the system is shut down to remove the spent chemical and recharged with a fresh solution to resume the operation.

7.5 MEMBRANES

Membrane systems, which consist of semipermeable elements (polymeric membranes), separate gases by selective permeation of the gas constituents in contact with the membrane. The gases dissolve in the membrane material and move across the membrane barrier under an imposed partial pressure gradient, which is established by feeding high-pressure gas to one side of the membrane while maintaining the permeate side at much lower pressure.

Many different types of membranes have been developed or are under development for industrial separations, but for natural gas separations, the industry standard is presently cellulose acetate. These membranes are of the solution-diffusion type, in which a thin layer of cellulose acetate is on top of a thicker layer of a porous support material. The membranes are thin to maximize mass transfer and, thus, minimize surface area and cost, so the support layer is necessary to provide the needed mechanical strength (Baker, 2002).

Membranes are used widely for the removal of CO_2 from natural gas.[16] These systems can handle higher CO_2 concentration (around 85%) compared to amine systems at high pressures in the range of 200 psia to 1,200 psia (Bhatia, 2008).

Membranes generally are not effective for the removal of H_2S and other sulfur species because of the very low outlet concentrations required. However, UOP has developed polymeric membranes that can be used for bulk removal of H_2S from natural gas even at very high H_2S concentrations and at high operating pressures. This membrane system can be used to make a large bulk cut of the acid gases, typically in the 70%–90% reduction range. A traditional amine process is then used to meet final pipeline specifications. Significant reductions in capital and operating costs can be achieved through this new approach (Cnop, 2008).

Membrane systems have major advantages, as a smaller footprint is needed. These systems are cleaner, need less maintenance, have lower capital and operating costs, require lower energy, and are environmentally friendly. The main limitation of their use is linked to the significant loss of hydrocarbons in the effluent (permeated) stream. However, the hydrocarbon recovery of a two-stage system is significantly better than that for a single-stage system.

Membrane systems perform well at reduced feed flow rates, but their performance drops when design flow rates are exceeded. Additional modules must be therefore added in parallel to accept higher flow rates. As a result, the process does not realize the economies of scale as the flow rate is increased (Kohl and Nielsen, 1997).

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[16] Most membranes used for CO_2 removal have an inherent ability to dehydrate a gas stream and because of this, certain problems can arise when using membranes for dehydration.

Some early installations of membranes highlighted the need for suitable pretreatment prior to the membrane process, where improper pretreatment generally leads to performance decline rather than complete nonperformance. The development of a robust and comprehensive pretreatment scheme, in combination with the continuing development of advanced membranes, has even further enhanced the reliability and performance of membrane technology in a variety of processing conditions (Dortmundt and Doshi, 1998).

7.6 CRYOGENIC FRACTIONATION

For many years, the low-temperature distillation (cryogenic separation) has been a commercial process used to liquefy and purify CO_2 from relatively high purity (>90%) sources. It involves cooling the gases to a very low temperature so that the CO_2 can be liquefied and separated. This technology requires substantial energy to provide necessary refrigeration. It also needs pretreatment of feed gas to remove components that have a freezing point above the operating temperature to avoid freezing of lines and blockages of process equipment.

Cryogenic fractionation seems to have a good prospect for removing CO_2 and H_2S from natural gas because the vapor pressures of the principal components are considerably different. However, problems are associated with the separation of CO_2 from methane, CO_2 from ethane, and CO_2 from H_2S (GPSA, 2004). A number of techniques are available for solving these problems. CFZ technology, developed at ExxonMobil Upstream Research Company, is a cryogenic distillation process for the single-step removal of CO_2 and H_2S from natural gas involving the controlled freezing and remelting of CO_2. It provides the ability to more economically process natural gas without imposing limitations on the amount of CO_2 or H_2S present in the feed gas. Further, the acid gas components are discharged as a high-pressure liquid stream that can be easily pumped for geosequestration or for use in enhanced oil recovery operations, while yielding a high-quality methane product (Northrop and Valencia, 2009).

7.7 OTHER PROCESSES

Many methods have been developed over the years for the removal of H_2S from sour natural gases. However, they require relatively large investment and operational costs as well as special operational safety and health procedures (Gadre, 1989). In addition, they may produce waste products. Continual search for more economical, safe, and simple removal methods has led to investigations into microbiological solutions for H_2S removal (Sublette and Sylvester, 1987). One of the most common technologies for biological treatment of sour gas is the THIOPAQ process (jointly developed by Paques B.V., Shell Global Solution Int. BV, and UOP), which removes H_2S from gaseous streams by absorption into a mild alkaline solution (under a pressure of up to 75 barg) followed by oxidation of the absorbed sulfide to elemental sulfur by use of a consortium of naturally occurring colorless sulfur bacteria called *Thiobacilli*. For the past few years, the process has been marketed as a very attractive alternative to traditional gas sweetening technologies, which resulted in the start-up of the first commercial unit in 2002. Cline et al. (2003) explained the underlying working principle of the THIOPAC technology and provided some information on the special and potentially very valuable characteristics of the THIOPAC sulfur.

7.8 PROCESS SELECTION

Facing the vast range of applicable technologies, the process designer's challenge lies in selecting the optimal technology for the project at issue. It may become useful for a first preselection, to discard technologies with application ranges that do not fit the project requirements. Table 7-2 summarizes the main characteristics of the aforementioned gas sweetening technologies.

Selecting an appropriate gas sweetening process is not easy, and a number of variables must be weighed prior to making a process selection. These include the following:

- Sour gas flow rate, composition, pressure, and temperature
- Specifications for the sweet gas and the acid gas
- The depth of acid gases and total sulfur removal required, plus the selectivity desired
- Impact of the composition of the concentrated acid gas on the design and costs of the downstream sulfur recovery and tail gas treating processes
- Process complexity and transient behavior (e.g., load-following capability)
- Process flexibility with regard to feedstock changes and potential future tightening of emission limits
- Process train philosophy
- Capital, operating, and royalty costs for the process

Decisions in selecting a gas sweetening process can many times be simplified by feed gas composition and operating conditions. Dry-bed processes are more economical when the sulfur content of the feed gas is below 20 lb sulfur per day (Manning and Thompson, 1991). If the operating conditions

Table 7-2 Main Characteristics of Natural Gas Sweetening Technologies (Bergel and Tierno, 2009)

Item		Amines	Membranes	Hot Potassium Carbonate	Physical Solvents	Hybrid Solvents	Physical Adsorption	Nonregenerative Solvents
Acid Gas Content at Inlet		Up to 70% V	Up to 90% V	5% V to 50% V	$P_{CO2} > 3.5$ bara	$P_{CO2} > 7.6$ bara	0.1% V to 2% V	$< 0.1\%$ V
Typical Acid Gas Content at Outlet		From 2% V down to deep removal	1% V	≥1.5% V (single-stage scheme)	1% V	$< 0.5\%$ V	50 ppmV	5 to 300 ppmV
Typical Gas Flow Rate (MMSCMD)		From low to more than 10	From very low to more than 10	From low to 7	From 3 to 11	From low to more than 10	From low to 3	Low
Typical Operating Conditions	Pressure	Absorber: 5 to 120 bara Regenerator: 1.5 bara	27 bara to 100 bara	Absorber: 69 bara Regenerator: 1.5 bara	Absorber: 69 bara	Absorber: 69 bara Regenerator: 3 bara or lower	Adsorption: 14 to 69 bara	As required
	Temperature	30°C to 60°C (absorption)	< 60°C	110°C to 120°C	–18°C to Ambient	Ambient to 60°C (absorption)	Adsorption: 30°C to 60°C Regeneration: 350°C	Ambient
Typical Hydrocarbon Losses		Less than 1%	1 stage: 8%–15% 2 stages: 2%	Very low	Absorbs heavy hydrocarbons and aromatics	Absorbs heavy hydrocarbons and aromatics	None	None
Process Turndown (Gas Flow Rate)		30%	20%	30%	Approximately 30%	Approximately 30%	Low	No limitation
Water Content in Treated Gas		Saturated with water	Dehydrated	Saturated with water	Dehydrated	Close to saturation	Dehydrated	Saturated with water
Contaminants		Oxygen, Heavy HC (liquid state), Solid particles, Organic acids	Heavy HC, BTEX, Amines, Liquid water	Solid particles, Heavy HC (liquid state)	Solid particles, Heavy HC (liquid state)	Solid particles, Heavy HC (liquid state), Oxygen	Heavy HC (liquid state), Liquid water	Heavy HC (liquid state)
Materials Requirements		SS for certain parts	Pretreatment: CS or SS (high acid gas content) Membrane Skid: CS	SS for certain parts	Carbon steel	SS for certain parts (Lean/Rich Heat Exchanger)	Carbon steel	Carbon steel
Ease of Operation		High complexity	Low complexity	Very high complexity	High complexity	High complexity	Medium complexity	Low complexity
Cost	Investment	High	Medium	High	Medium	High	Medium	Low
	Operation	Medium	1 stage: Low 2 stages: Medium	Low	Low	Medium	Low	High

require the removal of a large amount of H_2S and/or CO_2, liquid-phase processes are used. If it is required to selectively remove H_2S, then physical solvents will be suitable; they also remove mercaptans, COS, and CS_2. High partial pressures of acid gases enhance the probability of using a physical solvent. However, the presence of significant quantities of heavy hydrocarbons in the feed discourages using physical solvents due to a significant loss of these heavy hydrocarbons because they are released from the solvent with the acid gases and cannot be recovered economically. On the other hand, if the feed contains a high amount of H_2S and CO_2 and it is required to remove both of them, then chemical solvents such as carbonates or amines could be used. Carbonates also can be used if removing some COS and CS_2 from the feed is required. However, carbonates cannot be used in sweetening gas streams with only H_2S because CO_2 is necessary to regenerate the potassium bisulfide. The carbonate process works best when the CO_2 partial pressure is in the range of 30–90 psia. Low partial pressures of acid gases and low outlet specifications generally require the use of amines for adequate sweetening. The amine processes offer good reactivity at low cost and good flexibility in design and operation. However, the selection of an amine for gas sweetening is complex and must be based on several process considerations. In the design of the amine process, the primary concern is that the sweetened gas meets the required purity specifications with respect to H_2S and CO_2. The secondary objective is to select the amine that optimizes equipment size and minimizes plant operating costs. Hybrid solvent processes, which combine the advantages of both amine and physical solvents, are used to remove H_2S, CO_2, and organic sulfur components from natural gases at high partial pressure. The selected amine depends on the acid gases in the feed and whether CO_2 removal is required. Like the physical solvent processes, heavy hydrocarbon absorption remains one of the disadvantages of the hybrid solvent process.

Selection criteria for the solvent-based processes were discussed by Tennyson and Schaaf (1977). However, their guidelines should be used only for an initial selection of potential process choices and should never replace good engineering judgment. Kurimura et al. (1993) also developed a fuzzy-logic-based expert system to select near-optimum acid gas treating processes in natural gas processing plants using heuristic knowledge from experts and literature. The developed expert system provides more rigorous and realistic results. However, for selecting a certain natural gas sweetening technology, it is important to take into account the interaction between the gas sweetening unit and the other upstream or downstream units in a gas processing plant. Depending on the technology chosen, different schemes might be developed that need to be evaluated, taking into account some characteristics: CAPEX and OPEX (life-cycle cost), license availability and fee, environmental and operational aspects, process safety, maintenance, utility needs, etc. (McMahon, 2004; Bergel and Tierno, 2009). Customer preferences and biases, as well as the experience of plant designers and constructors, may also influence the selection of the gas sweetening process.

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CHAPTER 8

Sulfur Recovery and Handling

8.1 INTRODUCTION

Roughly 25% of the natural gas brought into production from new sources requires some degree of treatment to remove hydrogen sulfide (H_2S) and recover elemental sulfur from acid gas streams containing H_2S in high concentrations. In fact, due to the latest legislation plus environmental and safety considerations, venting or flaring H_2S to the surroundings is now an unacceptable option,[1] so conversion to elemental sulfur is necessary. Elemental sulfur is easy to store, handle, and transport in bulk. Ease of storage is an important advantage as it enables sulfur to be stockpiled economically in periods of reduced demand. Chemical conversion of H_2S for disposal as solid waste (such as calcium sulfate) is technically feasible but uneconomical. Converting H_2S to sulfuric acid is also a feasible option. However, sulfuric acid is both toxic and corrosive and, as a liquid, more expensive to store and transport than solid sulfur. Where environmental regulations permit and production of elemental sulfur is not economically attractive, injection wells provide a safe means for H_2S disposal.

This chapter discusses the properties of elemental sulfur and then describes the most common methods available for dealing with large quantities of H_2S.

8.2 SULFUR PROPERTIES

Elemental sulfur, which usually occurs in the form of eight-membered rings, is a relatively nontoxic and chemically inert substance and insoluble in water and most other liquids. It is a solid at ambient temperatures and exists in two crystalline and amorphous forms. Sulfur in crystalline form has different shapes. Rhombic and monoclinic are the most common forms of crystalline sulfur. These two forms differ in the way in which the rings are stacked. Rhombic sulfur is the stable form of the element at room temperature. If rhombic sulfur is heated to about 203°F, it changes into monoclinic crystals. Above 203°F and

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[1] Carbon dioxide is usually vented to the atmosphere but sometimes is recovered for CO_2 floods.

up to 239°F, monoclinic sulfur becomes the more dominant structural form. Amorphous sulfur is formed when liquid sulfur, which has been heated to elevated temperatures, is allowed to cool rapidly. Amorphous sulfur slowly changes to the rhombic crystalline form at ambient temperatures.

When melted, sulfur becomes a brownish-yellow transparent liquid whose molecular structure also consists of eight-member rings. As liquid sulfur is heated to about 320°F, the structure of the liquid undergoes a change (forming long-chain sulfur polymers in conjunction with the eight-member sulfur rings); its color becomes a dark reddish brown, and its viscosity increases significantly.

Sulfur vapor exists as S_x (where x can have values from 1 through 8), but the principal sulfur species are S_2, S_6, and S_8. The distribution of sulfur-vapor species varies significantly with changes in temperature, where at lower temperatures S_8 dominates, but as the temperature rises S_8 converts to S_6 and finally to S_2.

The physical and chemical properties of sulfur are unusual. Therefore, any attempt to understand sulfur recovery processes requires some study of the behavior of the various sulfur species. The properties of elemental sulfur are well described in various literatures (Meyer, 1976; Chao, 1980; Shuai and Meisen, 1995).

8.3 SULFUR RECOVERY

Sulfur recovery refers to the conversion of H_2S to elemental sulfur. The most common method of converting H_2S into elemental sulfur is the Claus process or one of its modifications. Other sulfur recovery processes can replace the Claus process where it is uneconomic or cannot meet the required specifications. Usually, such processes are used in small-scale units, or where the H_2S content of the acid gas is too low for a Claus unit or one of its modified versions. The following sections cover the conventional Claus process and its various modifications and extensions, as well as some other lesser-utilized sulfur recovery processes.

8.3.1 Standard Claus Process

The Claus sulfur recovery process is the most widely used technology for recovering elemental sulfur from acid gas feed streams.

In the original Claus process, elemental sulfur was produced by the partial oxidation of H_2S in a single step over a preheated catalyst bed, according to the following exothermic reaction:

$$H_2S + 1/2\,O_2 \rightarrow 1/8\,S_8 + H_2O \tag{8-1}$$

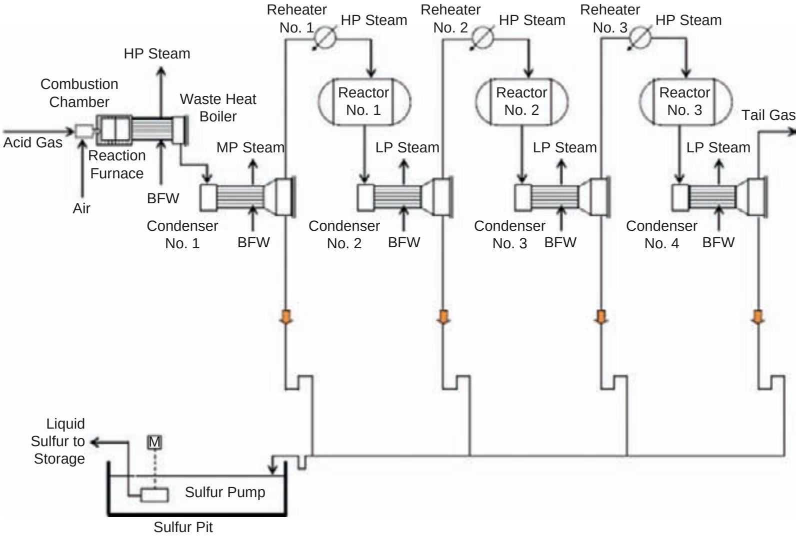

Figure 8-1 Typical illustration of a three-stage Claus sulfur recovery unit.

However, since the heat of the reaction was dissipated only by direct radiation, high sulfur recovery was very difficult to obtain. A very important modification to the Claus process was made in 1940, which allowed increased sulfur recoveries. In this modified-Claus process (see Figure 8-1), the overall reaction is carried out in two stages.

In the first stage (thermal section), enough air is added to oxidize only one-third of the incoming H_2S to SO_2, shown by reaction 8-2:

$$H_2S + 3/2\,O_2 \rightarrow SO_2 + H_2O \tag{8-2}$$

This reaction is highly exothermic and is not limited by equilibrium.

In the reaction furnace, the unburned H_2S in the acid gas reacts with the produced SO_2 to yield elemental sulfur vapor. This reaction is referred to as the Claus reaction and is shown by reaction 8-3. This reaction is endothermic and is limited by equilibrium.

$$SO_2 + 2H_2S \leftrightarrow 3/2\,S_2 + 2H_2O \tag{8-3}$$

Usually, 60%–70% of the total conversion of H_2S to elemental sulfur is achieved in the thermal stage (Gamson and Elkins, 1953; Opekar and Goar, 1966). During the thermal stage, side reactions also occur in the presence of significant amounts of impurities (hydrocarbons, CO_2, COS, NH_3, etc.[2])

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[2] Ammonia (NH_3) in the Claus unit feed can usually be traced to an upstream sour water stripper.

that might be in the feed acid gas stream (Borsboom and Clark, 2002). The reactions that have COS or CS_2 as a product typically result in a reduction in overall sulfur recovery. Ammonia (NH_3) can react with H_2S, SO_2, CO_2, and other compounds, which are found in the SRU, to form ammonium salts that would cause plugging in the downstream catalyst beds and equipment and eventually result in unplanned shutdowns. Therefore, when the feed gas contains contaminants, a higher flame temperature is often required to successfully destroy them in the reaction furnace.

The hot combustion products from the reaction furnace flow through the tubes of a waste heat boiler (WHB) and are partially cooled to 600°F–800°F by generating high-pressure (HP) steam (up to 600 psig) on the shell side of the WHB. As the gases from the reaction furnace are cooled in the WHB, the S_2 sulfur vapor species shift through a number of exothermic reactions to other sulfur species, primarily S_6 and S_8 (Goar and Fenderson, 1996). The cooled gases from the WHB are further cooled (to 340°F–375°F typically) in the first sulfur condenser by generating medium-pressure (MP) steam (40–65 psig typically),[3] where most of the sulfur vapor formed in the reaction furnace is condensed. The condensed sulfur is separated from the gas and drained from the condenser through a hydraulic seal and rundown line to the sulfur collection pit.

In the second stage (catalytic section), the overall conversion of H_2S to elemental sulfur is increased in a series of catalytic reactors (1 to 3) by reaction of the generated SO_2 and the unreacted H_2S over fixed beds of Claus catalysts at much lower temperatures (400°F–650°F). The Claus reactions in the catalytic section are slightly different from the thermal section reaction. These reactions are both exothermic and are represented as

$$2H_2S + SO_2 \leftrightarrow 3/6\,S_6 + 2H_2O \tag{8-4}$$

$$2H_2S + SO_2 \leftrightarrow 3/8\,S_8 + 2H_2O \tag{8-5}$$

The use of appropriate catalysts at selected temperatures optimizes the Claus reaction yield and also allows COS and CS_2 produced in the thermal stage to be eliminated.[4] There are basically five types of industrial Claus catalysts.

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[3] While high-pressure steam can be generated in the waste heat boiler, the sulfur condenser is normally limited to medium-pressure steam generation because of the low temperature required to obtain maximum sulfur condensation.

[4] COS and CS_2 conversion to H_2S is achieved by hydrolysis at the relatively high temperatures found in the first catalytic converter.

However, specially promoted activated alumina catalysts and TiO_2 catalysts can only improve the hydrolysis of COS and CS_2.

In the catalytic conversion stage, the first step is reheating the process gas from the upstream sulfur condenser to a temperature of 420°F–450°F[5] (often by using some of the high-pressure steam generated in the WHB) before passing to the reactor to prevent sulfur deposition on the catalyst, which leads to rapid catalyst deactivation. The gas from the catalytic reactor flows through the tubes of the sulfur condenser to condense the sulfur vapor and separate the elemental sulfur formed after each catalytic reactor, where the liquid sulfur goes to the collection pit. Removing sulfur from the process flow allows additional reactions to proceed in the next conversion stage. The heat released by the Claus reaction is recovered as low-pressure (LP) steam in the sulfur condenser. The process of reheating, catalytically reacting, and sulfur condensing may be repeated, in one, two, or even three additional catalytic stages. After leaving the last sulfur condenser,[6] the effluent (tail) gas, which still contains appreciable amounts of sulfur compounds and a small amount of sulfur as vapor and entrained liquid, is routed to either a treating unit for further processing (depending on the local air pollution control regulations) or to a thermal oxidizer to incinerate all of the sulfur compounds in the tail gas to sulfur dioxide before dispersing the effluent to the atmosphere.

The sulfur recovery depends on such things as feed composition, age of the catalyst, and number of reactor stages. Typical sulfur recovery efficiencies for Claus units are 90%–96% for a two-stage catalytic conversion and 95%–98% for a three-stage catalytic conversion. A fourth catalytic converter is normally not economical as it increases conversion by less than 1% (Kohl and Nielsen, 1997). Because of equilibrium limitations and other sulfur losses, overall sulfur recovery efficiency in a Claus unit usually does not exceed 98% (Lagas et al., 1989).

The straight-through configuration of modified-Claus is the technology of choice for acid gases containing high concentrations of hydrogen sulfide (typically above about 50 mole% H_2S). However, concentrations as low as 40 mole% are acceptable if the air or acid gas is preheated[7] (Parnell, 1985).

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[5] The reheat temperatures are kept to a practical minimum because the Claus reaction equilibrium is favored by lower temperatures.

[6] The final condenser is often operated with the outlet stream temperature as low as 265°F–270°F to minimize sulfur vapor losses in the vent gas and improve the overall sulfur recovery (Goar and Fenderson, 1996).

[7] Acid gas preheating is more difficult to apply when the acid gas is recovered from the amine regenerator at low pressure. Moreover, corrosion must be properly checked as thermal cracking of the acid gas constituents may occur.

When H_2S content is lower than 40 mole%, the flame in the reaction furnace becomes unstable and cannot be maintained.[8] This results in dramatic operating problems in the downstream Claus reactors, which may be exposed to process gas containing unreacted oxygen from the reaction furnace (Goar, 1974). Therefore, the straight-through design is no longer a good choice, and the following possibilities must be chosen.

8.3.1.1 Split-Flow Thermal Stage Operation

In the split-flow mode, only about 35%–45% of the total acid gas stream goes to the reaction furnace, and the remainder joins the main process stream just upstream or downstream of the first sulfur condenser. When part of the feed gas is bypassed, the combustion air is adjusted to oxidize all the H_2S to SO_2 and, consequently, the necessary flame temperature can be maintained (Kohl and Nielsen, 1997). In the split-flow process, sufficient gas must be bypassed so that the flame temperature is greater than approximately 1,700°F. In addition, the upper limit for acid gas bypass is two-thirds of feed gas, as one-third of the total hydrogen sulfide must be combusted to form SO_2. However, if contaminants are present in the feed, they will remain in part and enter directly into the catalytic section of the Claus unit. This results in deactivation and plugging of the catalytic reactors. Nevertheless, when this solution is applicable, it is the simplest and most economical way to treat medium acid gas composition. If air preheating is used with the split-flow configuration, gases with as little as 7 mole% H_2S can be processed (Parnell, 1985). However, note should be made that when the acid gas stream contains less than 10 mole% H_2S, it may be too difficult to operate the facilities and achieve the necessary recovery efficiency in a split-flow modified Claus unit. In such a situation, the acid gas stream can be treated in an acid gas enrichment unit utilizing a specialty solvent that preferentially absorbs the H_2S and leaves most of the CO_2 behind.

8.3.1.2 Oxygen Enrichment

Lean acid gases can also be processed in a straight-through mode of Claus unit using pure oxygen or highly enriched air to replace a portion or all of the combustion air.[9] In fact, oxygen enrichment raises the flame temperature by limiting the nitrogen effect of air. Moreover, this results in

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[8] A flame temperature of approximately 1,800°F–2,000°F would provide for stable combustion in the reaction furnace.

[9] For lean acid gas stream with less than about 20 mole% H_2S, supplemental fuel firing will also be required even with 100% oxygen (Goar and Fenderson, 1996).

reductions of the equipment size and investment cost, since the global flow processed in the sulfur recovery unit is lower (Lee and Moore, 1997). SRU operators have also used oxygen enrichment as a means of increasing unit capacity and sulfur recovery,[10] as well as tail gas cleanup efficiency (Best et al., 1999). The application of the process has now been extended to grass-roots installations, even for rich H_2S feed streams, to provide operating flexibility at lower costs than would be the case for conventional Claus units. Nevertheless, in order to be applied, this technology requires an available and economical source of oxygen.

There are three levels of oxygen enrichment: low (less than 28 vol% oxygen), medium (28 to 45 vol% oxygen), and high (greater than 45 vol% oxygen). Figure 8-2 shows capacity increase for SRU as a function of oxygen concentration in the process air. However, for now, only low-level enrichments are addressed, since they lead to minor changes at the Claus unit and provide up to a 25% increase in capacity. Higher oxygen-enrichment levels require the use of special oxygen piping and component cleaning and metallurgy, and special burner technology (Goar and Fenderson, 1996; Kohl and Nielsen, 1997).

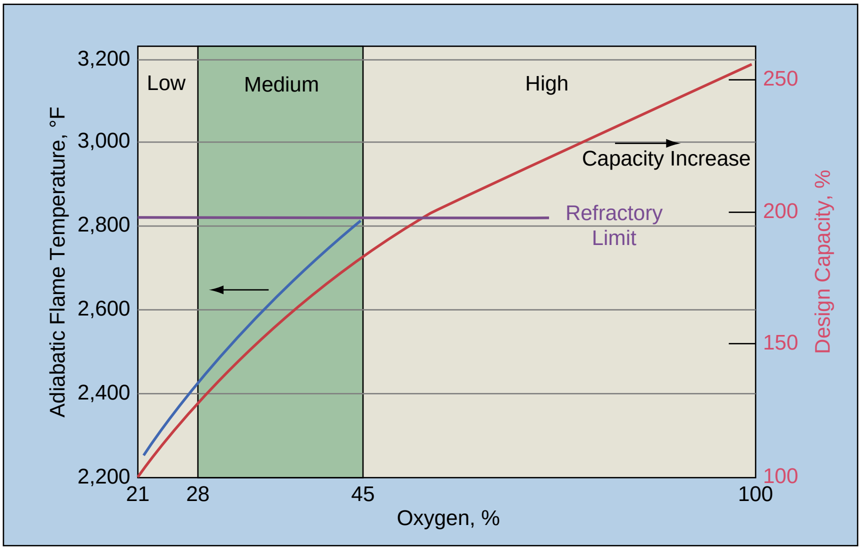

Figure 8-2 Effect of oxygen enrichment on the calculated reaction furnace temperature and % Claus unit capacity increase (Gandhi et al., 2010).

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[10] While oxygen enrichment doesn't improve sulfur recovery to a significant extent, it allows for complete combustion of all hydrocarbons that might be in the feed, thus preventing carbon or soot formation and resultant poisoning of catalyst beds.

A challenge for early oxygen-blown Claus process developers was finding a means to control high flame temperatures in the front-end furnace (FEF) to avoid overheating and damaging the reaction furnace and burner refractory.[11] Two principal means have been used to achieve temperature control, recycle of cooled gas to the FEF (COPE process developed by Air Products and Goar, Allison & Associates) and special burners that control the combustion process itself (Lurgi OxyClaus process). Temperature control via staged combustion is also employed (SURE process developed by British Oxygen Corp. and Parsons Corp.). Detailed descriptions of each of these processes are provided by Kohl and Nielsen (1997). Besides these processes, several other oxygen suppliers (i.e., Air Liquide and Praxair) offer their own versions of the oxygen-blown Claus processes.

8.3.1.3 Selectox Process

For acid gases that are 5–30 mole% H_2S, the Claus process is less competitive because it is difficult to ensure a stable flame at a low operating cost; consequently, some other technologies are of interest. One method is to replace the thermal section of the modified-Claus process with a catalytic converter bed containing special Selectox catalyst developed by UOP. Over the catalyst, air oxidizes H_2S to SO_2 (catalytic oxidation without the presence of a flame) that reacts with additional H_2S to produce elemental sulfur. Overall sulfur recoveries of 95%–97% are possible with one Selectox and two Claus stages (Goar and Fenderson, 1996).

Two versions of the process are offered by UOP. If the H_2S content in the acid gas feed is 5 mole% or less, then the once-through mode can be used, as the temperature rise in the Selectox catalytic bed is within an acceptable range. In this scheme, the gas stream out of the Selectox catalytic converter is cooled and then reheated and treated in one or two additional conventional Claus catalytic stages. If the H_2S content is above 5 mole% in the acid gas feed, then the recycle Selectox process can be used. In the recycle version, gas is recycled from the downstream of the first condenser to cool the outlet temperature of Selectox catalytic converter not to exceed 700°F. The temperature limit is set so that carbon steel can be used for the reactor vessel. The gas out of the Selectox catalytic stage is cooled to condense and separate the formed sulfur and may be reheated for further treating in Claus catalytic stages (Gowdy and Bertram, 1998).

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[11] These flame-temperature-controlling methods are not required for lean acid gas streams as their flame temperatures, even at 100% oxygen, are not high enough to damage high-quality refractory (Goar and Fenderson, 1996).

The Selectox process appears to be particularly applicable for acid gases with very lean H_2S feeds, but at a relatively small scale. The absence of a reaction furnace allows some contaminants, such as various hydrocarbons, to react on both Selectox and Claus catalysts, causing hot spots (Jones and Bertram, 2001). Thus, the process may not be as forgiving as conventional Claus.

8.3.1.4 Clinsulf Process

To overcome the problems of traditional Claus units with low H_2S feed concentrations, Linde A.G. developed the Clinsulf process in which H_2S is directly oxidized to sulfur and H_2O ($2H_2S + SO_2 \leftrightarrow 3/XS_x + 2H_2O + \text{energy}$) in a reactor filled with a suitable catalyst. In this process, the heat of reaction is removed directly from the Claus reactor with a coil-type tubular heat exchanger arranged within the catalyst bed. The possibility of internal cooling in this type of reactor, which is called the Linde isothermal reactor, keeps the reaction temperature as close as possible to the sulfur dew point, making the process capable of attaining maximum sulfur yield.

In the Clinsulf process, the lean acid gas (containing 20% or less H_2S) is preheated and mixed with air before being fed into the isothermal reactor. After leaving the reactor, sulfur is condensed and removed while the tail gas is incinerated. A single-stage Clinsulf unit is capable of attaining up to 92% sulfur recovery for very lean acid gases (<5% H_2S) and up to 95% for gases with 10%–20% H_2S content. However, with the Linde isothermal reactor, it is also possible to lower the reaction temperature below the sulfur dew point or even below the sulfur solidification point, which results in a high overall sulfur recovery efficiency of typically 99.5%–99.8% (Gemmingen and Lahne, 1994; Heisel et al., 1999).

8.3.2 SuperClaus Process

The SuperClaus process was developed by Comprimo B.V. (now Jacobs Comprimo Sulfur Solutions) to increase the overall sulfur recovery of the Claus process by reducing its inherent thermodynamic limitations (Lagas et al., 1989, 1994). The process uses special SuperClaus catalyst (Fe_2O_3 on silica support for the last generation) in the last conversion stage to selectively oxidize H_2S contained in the process gas directly to elemental sulfur. Unlike the conventional Claus reaction, the oxidation reaction is not equilibrium limited, and most of the H_2S (over 85%) reacts to elemental sulfur. However, the other sulfur species pass over the special catalyst unreacted. Therefore, prior reduction of non-H_2S species results in greater conversion in the direct oxidation bed and greater overall recovery for the unit. Two versions of the process (the SuperClaus 99 and SuperClaus 99.5) have been developed for such purposes. The SuperClaus 99 process consists of a conventional Claus unit thermal stage, followed by two reactors filled with standard Claus catalyst, and a final reactor filled with the new SuperClaus catalyst. In the thermal stage, the acid gas is burned with a substoichiometric amount of controlled combustion air such that the tail gas leaving the second reactor contains typically 0.8–1.0 vol% of H_2S. In cases where substoichiometric combustion in the reaction furnace is not possible due to minimum flame temperature requirements, the 0.8–1.0 vol% H_2S concentration is maintained by routing a portion of the acid gas feed downstream of the first condenser. A total sulfur recovery efficiency of up to 99% can be obtained with the SuperClaus 99 process. Attaining over 99% requires a hydrogenation step prior to the selective oxidation step. This step converts any remaining SO_2 to H_2S over a cobalt/molybdenum catalyst (so there is no longer a requirement to operate the Claus unit with excess hydrogen sulfide) and potentially raises sulfur recovery to about 99.5% for a rich H_2S acid gas feed. This version of the process is called SuperClaus 99.5.

8.3.3 Small- and Medium-Scale Processes

While the Claus process is still the main process for recovering sulfur from H_2S, several new processes are vying for acceptance for the recovery of sulfur from low-quality acid gas streams (less than 40 mole% H_2S), or if recovery is required for very low mass flow rates of sulfur production (i.e., less than 10 tons of sulfur per day). For small amounts of sulfur (less than 0.2 long tons/day), nonregenerable H_2S scavengers such as activated carbon, iron oxide, and caustic solution are frequently used with good success (Fisher et al., 1999). However, for the midsize niche (between 0.2 and 25 long tons/day), the high operating costs associated with scavenging and the high capital costs associated with Claus processes (which may be combined with amine pretreatment and tail gas treatment) have made cost-effective treating challenging (Rueter, 2002). Often, gas in this niche has been processed using "liquid redox" technology.

Liquid redox sulfur recovery processes are based on reduction-oxidation (redox) chemistry to oxidize the H_2S to elemental sulfur in an alkaline solution containing a redox catalyst. In a first step, the aqueous solution absorbs the H_2S, which then reacts with the oxidized form of the catalyst to yield elemental sulfur as the catalyst is converted into its reduced form. In a second step, the reduced catalyst is regenerated by contact with air in the oxidizer prior to reuse. Vanadium and iron are the two redox catalysts used. Stretford and Unisulf are two examples of redox processes utilizing the vanadium catalyst. However, vanadium-based processes find little use now because of the toxic nature of the vanadium solution. The most prominent examples of the processes using iron as a catalyst are LO-CAT (developed by ARI Technologies in the early 1970s) and SulFerox (developed by Shell Oil and Dow Chemical in the early 1980s). Both the LO-CAT and the SulFerox processes are essentially the same in principle. In the LO-CAT process, the acid gas stream is fed to the LO-CAT absorber, which employs a dilute water solution of ferric iron, held in solution by organic chelating agents, which are proprietary compounds that maintain iron ions in solution and prevent the precipitation of $Fe(OH)_3$ and FeS (Hardison, 1984). The SulFerox process differs from the LO-CAT in that SulFerox uses a different chelating agent, which permits concentrations of up to 4 wt.% of the iron in solution. This has the advantage of lower circulation rates and smaller process equipment (Fong et al., 1987).

Liquid redox sulfur recovery processes are extremely efficient, removing over 99% of the sulfur in the feed. The main limitations of current redox processes are their relatively high chemical costs for makeup and catalyst replacement, and the quality of the sulfur produced, which is generally not as pure as Claus sulfur. Furthermore, no liquid redox technology has been demonstrated to reliably recover sulfur in a single step if the pressure is greater than around 500 psig, unless special and generally uneconomic design fixes were implemented. This is due to technological problems linked to the presence of solid sulfur particles in the aqueous redox solution used to treat the gas (McIntush and Petrinec, 1995; Holloway, 1996). As a result of the problems encountered with the aqueous liquid redox sulfur recovery systems, most processes that appeared to be targeted to this niche have made little penetration into the market for medium-size sulfur recovery from high-pressure natural gas (McIntush et al., 2001). Recognizing the problems with existing liquid redox processes, Gas Research Institute (GRI) and an industry partner developed the CrystaSulf process, a new liquid redox process that removes H_2S from gas streams containing small amounts of sulfur using novel, nonaqueous chemistry and solvents. CrystaSulf solves the problems of the aqueous-based liquid redox processes by eliminating the presence of sulfur solids in the high-pressure equipment (Rueter, 2002).

8.3.4 Other Processes

The Bio-SR process was developed in Japan in 1984 to recover sulfur from sour gas (Satoh et al., 1988). In this process, the sour gas is contacted with a solution containing ferric sulfate [$Fe_2(SO_4)_3$] in an absorber. A chemical reaction occurs between the ferric sulfate and the H_2S, yielding sulfur and ferrous sulfate ($FeSO_4$). The elemental sulfur formed is conglomerated to a large particle size in the slurry tank. Sulfur is separated from solution through a filter and settler. The ferrous sulfate solution is then introduced into the bioreactor where *Thiobacillus ferrooxidans* bacteria are employed for the regeneration of the solution. The bacteria oxidize ferrous sulfate into ferric sulfate. The regenerated solution is then recycled to the absorber to repeat the cycle. In this closed loop system, there is no degradation of the solution, no waste is generated, and essentially neither catalyst nor any chemicals need to be added.

The Thiopaq biological gas desulfurization process (developed jointly by UOP, Shell, and Paques) can be also used to remove H_2S from high-pressure sour gas or from low-pressure acid gas (Kijlstra et al., 2001). In this process, the sour gas or acid gas is contacted in an absorber with a mild alkaline solution (pH 8.0 to 9.0) that irreversibly reacts with the H_2S to form hydrosulfide ions (HS^-). The solution containing the HS^- ions leaves the absorber and flows through a low-pressure flash tank to allow dissolved gases to evolve. It then flows into a proprietary bioreactor operating at atmospheric pressure, where hydrosulfide (HS^-) is oxidized under controlled conditions to elemental sulfur in the presence of microorganisms. These aerobic (oxygen-consuming) organisms use the released energy from the sulfide oxidation for metabolic processes. The elemental sulfur produced has a hydrophilic nature and is separated from the aqueous effluent in a proprietary three-phase separator. The regenerated solution is then returned to the absorber to repeat the preceding reaction steps.

8.4 TAIL GAS TREATING

Tail gas from a typical Claus process, whether a conventional Claus or one of the extended versions of the process, usually contains small but varying quantities of sulfur compounds. Historically, the removal of these trace quantities was not considered a problem since there were no regulations that the quantity of sulfur compounds emitted to the atmosphere be reduced. However, as air pollution control regulations became more stringent, the unrecovered sulfur compounds in Claus unit tail gas streams became the target of further regulation. This led to the development of a large number of Claus tail gas treating (TGT) processes in order to remove the last remaining sulfur species[12] and to meet tail gas effluent specifications. Tail gas specifications can vary considerably, depending on the local environmental regulatory agency. For the United States, the requirement for an oxidation control system or a reduction control system followed by incineration is 250 ppm by volume (dry basis) of SO_2 at 0% excess air. Emissions requirements, however, can be significantly more stringent, say between 10 ppmv and 100 ppmv SO_2, in certain areas.

The TGT processes, which have achieved commercial status, can be categorized into three general groups:

- Sub-dew point processes
- Oxidation to SO_2 followed by SO_2 scrubbing
- SO_2 reduction and recovery of H_2S

Ballaguet and Barrere-Tricca (2006) provided an overview of tail gas treating processes. The following sections provide a sample process in each category.

8.4.1 Sub-Dew Point Processes

In the sub-dew point processes, a sub-dew point reaction stage is added to a conventional Claus unit to increase its recovery. This stage extends the capability of the Claus process by operating the Claus reaction at a temperature below the sulfur dew point, resulting in a favorable shift in the reaction equilibrium and higher sulfur conversion. The sulfur deposited in the sub-dew point catalyst bed is removed cyclically by passing hot incinerator off-gas first through the bed and then through a sulfur condenser to recover sulfur. The catalyst bed is then cooled to the operating temperature before going online for adsorption. Cyclical operation requires two beds: one operating on conversion and the other on regeneration. Processes of this type include cold bed absorption, or CBA (Goddin et al., 1974); MCRC (Heigold and Berkeley, 1983); and Sulfreen technology (Willing and Linder, 1994). The principal differences between the processes lie in the way regeneration gas is handled.

In the Sulfreen process, regeneration gas is self-contained within the sub-dew point portion of the process (recycled back from the condenser, through a heater, and back to the bed being regenerated). In the other

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[12] With new environmental requirements, sulfur recovery efficiencies of 98.5%–99.95% are now required.

two processes (CBA and MCRC), regeneration gas comes from the last-stage Claus condenser. The Sulfreen process (developed by Elf and Lurgi), which has the most installations,[13] is the only one discussed here. The Sulfreen process basically consists of two (occasionally three for large capacities) sub-dew point reactors (operating at low temperatures in the range of 260°F–320°F) in series with the conventional Claus reactors. The catalyst originally used in the Sulfreen process was a specially prepared activated carbon. This, although highly efficient, requires high temperatures (about 900°F) to vaporize the adsorbed sulfur during regeneration. A special activated alumina catalyst, which requires lower regeneration temperatures about 570°F, was subsequently developed, and all Sulfreen units since 1972 have been designed to operate with alumina catalyst. As the sulfur accumulates on the catalyst, its activity decreases and the catalyst beds have to be regenerated thermally at 480°F–570°F. During the regeneration step, sulfur is desorbed, and the catalyst is restored to full activity. Sulfur from the hot regeneration stream is then condensed. In the Sulfreen process, sulfur recoveries will typically vary from about 98.5% to 99.5%, depending on the richness of the acid gas and the number of sub-dew point reactors.

Two enhanced versions of the Sulfreen process (Hydrosulfreen and Doxosulfreen) are also available and enable higher sulfur recoveries to be reached. The Hydrosulfreen process is based on the pretreatment of the Claus tail gas by adding a hydrogenation reactor, which employs a catalyst such as CRS31, before the sub-dew point reactors. This eliminates any excess COS and CS_2 that may be present, and could potentially raise sulfur recovery up to 99.7%. The Doxosulfreen process is based on a post-treatment after a Sulfreen unit, where the remaining H_2S at the outlet of the Sulfreen reactor is oxidized with air (over the special alumina catalyst) to elemental sulfur at a low temperature (212°F–266°F). Both Sulfreen and oxidation reactors are regenerated through a common regeneration loop. The Doxosulfreen process is capable of reaching an overall sulfur recovery of up to 99.9%.

Note should be made that IFP has also developed the Clauspol process, which is based on the continuation of the Claus reaction at low temperatures (below the sulfur dew point) in an organic solvent containing a proprietary catalyst. The solvent first proposed was tributyl orthophosphate containing an alkaline substance as the catalyst. However, polyethylene glycol soon

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[13] About 50 commercial Sulfreen units have been built. However, the cyclical nature of the unit operation and their sulfur recovery limitations may discourage their application in the future.

became the solvent of choice because of its good thermal and chemical stability, low vapor pressure, low cost, and availability, as well as the low solubility of sulfur in the solvent and of the solvent in sulfur (Barthel et al., 1971). Due to very simple phase separation of the two main reaction products (water eliminated as a vapor in the gas phase and sulfur, which settles as a separate liquid phase), the process is continuous and does not need any regeneration of the solvent. The latest version of the Clauspol process allows sulfur recoveries up to 99.9+%, provided that COS and CS_2 have been hydrolyzed down to 300 ppmv of sulfur or less in the process, by using a TiO_2–based catalyst (Barrere-Tricca et al., 2000).

8.4.2 Oxidation to SO_2 Plus SO_2 Scrubbing

The basic principle behind these processes lies in the incineration or catalytic oxidation of Claus tail gas streams to convert all sulfur species to SO_2. If the resulting SO_2 concentration is below legal limits, the gas can be discharged to atmosphere. If the SO_2 concentration following oxidation is higher than permissible, the gas must be treated to remove SO_2 and then recycle the SO_2 stream to the Claus unit for conversion to elemental sulfur. In this case, the total capacity of the Claus process is thus increased. These processes can achieve an extremely high sulfur recovery of over 99.9%. The main difference between the processes resides in the different scrubbing mediums employed to remove SO_2. The simplest SO_2 removal process is caustic wash, but it is nonregenerative and creates waste disposal problems. Regenerative processes such as the Wellman–Lord process are preferred for TGT applications. The Wellman–Lord process is designed to absorb SO_2 and decrease the concentration of SO_2 in the tail gas to or below 200 ppm. The absorption is accomplished through the use of a sodium sulfite–sodium bisulfite solution. The product SO_2 released from the absorbent in the evaporator-crystallizer is cooled and recycled to the front end of the Claus unit (Potter and Earl, 1973).

In recent years, there has been strong interest to develop new processes of this type, but with some exceptions such as the Linde A.G. Clintox process, these have not been commercial successes because of the complexity of SO_2 recovery processes and their comparatively low SO_2 removal efficiencies. The Clintox process also involves burning the tail gas so as to convert the sulfur species to SO_2, which is absorbed in a physical solvent, stripped from the solvent, and returned to the Claus unit feed. With such a tail gas clean-up process, sulfur recovery rates of nearly 100% are attainable with approximately 80 ppmv residual SO_2 in the flue gas (Heisel and Marold, 1992).

8.4.3 SO_2 Reduction and Recovery of H_2S

Reduction processes convert other sulfur compounds to H_2S by hydrogenation followed by H_2S scrubbing by one of the amine-type processes[14] so that H_2S–rich gas can be recycled to the inlet of the Claus Unit. Processes of this type include Shell Claus Off-gas Treating (SCOT) and the Parsons/Unocal's Beavon Sulfur Removal (BSR) series of processes. The SCOT process was the first to be developed and is the most widespread, with more than 200 units constructed all over the world since 1973. The early SCOT units consisted of a hydrogenation reactor and a conventional amine unit. Figure 8-3 illustrates the concept involved in SCOT units. As can be seen from Figure 8-3, Claus tail gas enters an inline burner where fuel gas is burned with deficient air to form hydrogen and heat the gas stream to reaction temperature (about 570°F) before it enters the catalytic hydrogenation reactor (containing a Co-Mo catalyst), which converts all sulfur compounds to H_2S. The reactor effluent is cooled first in a heat recovery system to about 320°F and then in a water quench tower to approximately 100°F (Naber et al., 1973). The water from the tower is condensed, and the excess water is sent to a sour water stripper (SWS). Gas from the quench tower then

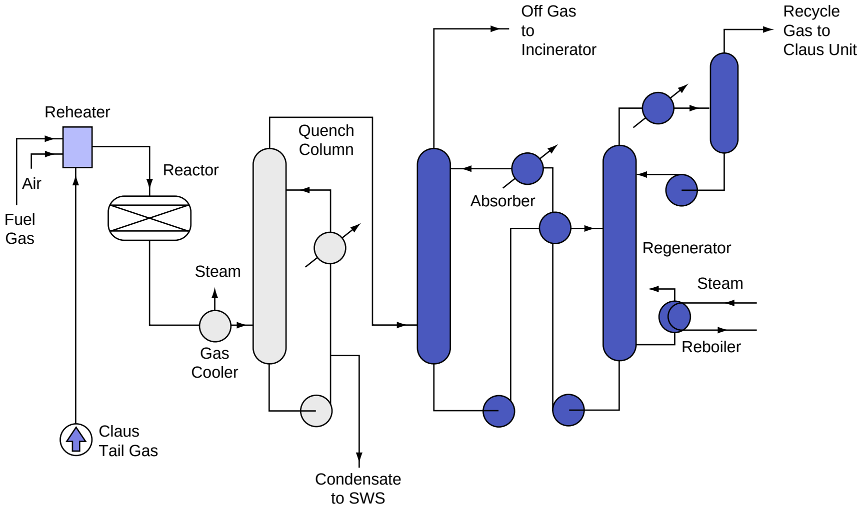

Figure 8-3 Typical illustration of the SCOT process (Brand, 2002).

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[14] The primary criterion for selecting TGTU solvent is high selectivity for H_2S absorption over CO_2 absorption. Formulated MDEA and/or sterically hindered amine such as Flexsorb SE Plus solvent is typically selected to meet the desired absorber performance at the minimum amine circulation rate.

contacts an aqueous amine solution in the absorber.[15] H_2S in the tail gas is removed in the absorber from which treated gas (containing small enough quantities of H_2S) passes to an incinerator before discharge to atmosphere.[16] The rich amine that leaves the bottom of the absorber flows to the regenerator, where heat is applied to strip the H_2S from the amine solution. H_2S separated from the rich solvent in the amine regenerator is recycled back to the Claus unit for conversion to elemental sulfur. Lean amine is cooled and returned to the absorber.

The SCOT process can produce an exit gas that contains 10 to 400 ppmv of total sulfur (Harvey and Verloop, 1976), while increasing total sulfur recovery to 99.7% or higher. Other licensors offer similar processes to SCOT, using their proprietary solvents and catalytic reduction reactors. However, the SCOT process remains by far the most widely used commercially. The SCOT process is reported to be quite flexible and not very sensitive to upsets in the upstream Claus unit. Relatively wide variations in the H_2S/SO_2 ratio in the Claus plant, as well as fluctuations in the feed gas volume, can be tolerated in a properly designed SCOT unit (Kohl and Nielsen, 1997).

Note should be made that TGT units are quite expensive, often costing an order of magnitude of Claus units. The proper selection of TGT process is therefore of some importance to operators, whereas the choice is made complicated by the different performance levels and life-cycle costs. A selection of the most popular TGT processes based on their sulfur recovery efficiency and, most importantly, in terms of the recovery cost per tonne of SO_2 was compared by Gall and Gadelle (2003).

8.5 SULFUR DEGASSING

Sulfur pits are normally used to collect raw liquid sulfur from the sulfur condensers under gravity. Raw liquid sulfur produced from the Claus sulfur condensers typically contains about 300 ppmw of soluble H_2S as hydrogen polysulfide (H_2S_X) and about 20–30 ppmw SO_2. H_2S_X is a weakly bound polymeric sulfur compound formed by the equilibrium reaction between sulfur and H_2S [$H_2S + (X\text{-}1)\ S \leftrightarrow H_2S_X$]. This reaction proceeds to the right

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[15] Early SCOT units used Sulfinol-D solution (Harvey and Verloop, 1976). As environmental regulations tightened, the use of Sulfinol-M, together with a different unit configuration, allowed H_2S levels to be reduced to below 10 ppmv in the treated tail gas.

[16] Because the amine solvent is not highly loaded with acid gases in the absorber (operating at near atmospheric pressure), there is no need for an intermediate flash vessel, and the loaded solution goes directly to a stripper.

with increasing temperature conditions. During storage or transport of the sulfur, the H_2S_X compounds will decompose (via procession of the preceding equation to the left) as equilibrium is achieved at the operating temperature. This results in formation of dissolved H_2S in the liquid sulfur, which will pass to the gaseous phase by physical desorption. H_2S and SO_2 can potentially accumulate to dangerous levels in the headspace of storage pits. Therefore, the contaminated degassing off-gas is typically vented to the thermal oxidizer for oxidation of all sulfur compounds to SO_2, but can also be directed to the front end of the SRU to eliminate these sulfur emissions and improve total sulfur recovery.[17] The net result is a sulfur product with an H_2S concentration of 10 ppmwt or less, suitable for transport.[18]

To meet the generally accepted guideline of 10 ppmw total H_2S for sulfur transport, sulfur degasification processes employ a combination of residence time, agitation,[19] and sometimes catalysts to reduce the residence time required to meet the degassed sulfur specification. The most prominent commercial degassing processes capable of meeting a 10 ppmw total H_2S spec are D'GAASS (Goar, Allison & Associates/Air Products), WorleyParsons (former ExxonMobil), Shell (Jacobs Engineering), Aquisulf (Lurgi), Amoco (BP—Ortloff, Black & Veatch), and HySpec (Enersul). Among these degassing processes, the first widely implemented is the Aquisulf process (Nougayrede and Voirin, 1989), with more than 80 references from 15–1,200 t S/d (single train). This process is based on two principles: mechanical degassing by agitation and pulverization to favor gas-liquid contact; and chemical degassing by catalyst injection to speed up the decomposition of H_2S_X. The Aquisulf process is available in batch and continuous versions. In the continuous version (see Figure 8-4), the liquid sulfur pit consists of two or more compartments. The sulfur in the first compartment is pumped and mixed with the catalyst and sprayed back into the compartment. The liquid sulfur overflows from a weir to the second compartment. The liquid sulfur is again pumped and sprayed in this tank to provide more agitation.

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[17] For improving emissions control from sulfur storage tanks, proper design methods have been developed and proven to provide this assurance. However, concerns for reducing emissions should be directed toward improvements in the primary SRU/TGT systems where the largest reductions can be made (Johnson, 2005).

[18] If liquid sulfur is not degassed, H_2S will be released during storage, handling, loading, and transport. Un-degassed sulfur can create an explosive mixture of H_2S in air and poses a toxicity hazard as well as a noxious odor problem when H_2S is released from the un-degassed liquid sulfur.

[19] All degassing processes involve agitation of the liquid sulfur and removal of the evolved H_2S with a sweep gas. Generally, air is used as the sweep gas since oxygen helps to release the H_2S from the polysulfide molecule.

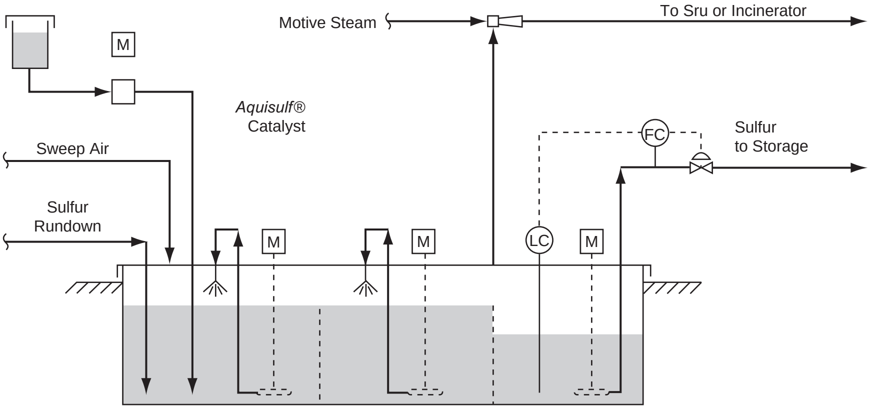

Figure 8-4 Continuous Aquisulf scheme (Nougayrede and Voirin, 1989).

The original catalyst was ammonia. However, problems associated with ammonium salts have resulted in the development of a proprietary Aquisulf liquid catalyst,[20] which does a better job than ammonia of decomposing the H_2S_X and does not result in solid salt formation.[21]

Currently, all the sulfur recovery units are designed to have the produced sulfur flows by gravity to the underground sulfur pit, or the entire unit has to be elevated if the sulfur pit is located above ground. Also, the sulfur pit should be very near sulfur condensers to meet the gravity flow constraints and hydraulic limitations. Considering this, WorleyParsons recently developed a new process called RSC-D^{TM} that allows the sulfur pit above or below ground and provides the flexibility of the sulfur pit location with more distance from the SRU using the new invention of liquid jet or eductor for collecting sulfur. The process also provides entrainment and enough agitation in the liquid sulfur where degassing occurs simultaneously in the sulfur collection system (Rameshni, 2009).

8.6 SULFUR STORAGE AND HANDLING

The degassed liquid sulfur, containing 10 ppmw total H_2S plus H_2S_X (maximum), and some traces, perhaps 5–10 ppmw SO_2,[22] is directed to an insulated and heated carbon steel storage tank to provide a buffer between the

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[20] Although proprietary, it can be purchased on the open market.

[21] With 10 ppm of Aquisulf, 9 hours are necessary to meet 10 ppmw H_2S.

[22] Recent research and operating data has demonstrated that liquid sulfur, degassed to < 10 ppmw H_2S, has a strong equilibrium driving force to react with air (oxygen) to form SO_2 in air-purged storage tanks (Johnson, 2005).

sulfur recovery unit and the downstream handling system. Even with sulfur degassed down to 10 ppmw total H_2S plus H_2S_X, some additional H_2S will evolve due to the residence time in the tank, the cooling that generally occurs, and the favorable air/sulfur/H_2S equilibrium concentration driving force. Despite the fact that both H_2S and sulfur are flammable in air, the current safe and proven industry practice is to use an air sweep of the sulfur tank vapor space to maintain the H_2S level to well below the lower explosive limit, which is 3.4% volume at a storage temperature of 330°F (Iyengar et al., 1995).

The stored liquid sulfur can be shipped to the market by tank trucks or by railcars. However, in most cases, it will be converted into a solid form suitable for marketing and transportation purposes. In emergency situations, the liquid sulfur may be poured to a block for short-term storage. The solid sulfur reclaimed from the block is normally melted to provide liquid feed to liquid shipments and/or sulfur-forming facilities.

8.6.1 Molten Sulfur Handling System

Safety, environmental, and operational concerns associated with transporting molten sulfur by trucks or railcars are causing sulfur producers to review other transportation alternatives in an effort to determine a better transportation method.[23] Molten sulfur transmission pipelines have been a continual challenge to the sulfur industry because of both the upper and lower temperature limits that are necessary to handle molten sulfur properly and prevent solidification. In fact, a minimum temperature (255°F) is needed to pump the sulfur, but if it is heated to an excessive temperature (320°F), the sulfur degrades and thickens. Therefore, all lines and equipment must be adequately heated to maintain the sulfur at the optimum temperature. A number of heating systems have been successfully employed, including steam jacketing, steam tracing, thermal oil or glycol (jacket or tracing), pressurized hot water, and skin effect electrical heat tracing (SEEHT). It is widely accepted by pipeline operators that the SEEHT system, monitored and controlled by fiber-optic technology, is cost effective, reliable, and safe for long sulfur transmission pipelines when compared with other heating options applied to molten sulfur pipelines (Beres et al., 2004). A SEEHT

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[23] H_2S will be released during this transfer and evolve into the vessel vapor space. Most often, the vapor mixture (H_2S and air) in the vapor space is displaced to the atmosphere through the vessel's loading and vent nozzles. Hydrogen sulfide, in addition to being a combustible gas, is also a significant safety hazard, because of its toxic nature, and presents dangers to workers in the area.

system must be designed to maintain the sulfur at an average fluid temperature of 275°F–284°F over the entire length of the pipeline. The fiber-optics temperature measurement should be accurate to ±1°F, to allow for accurate identification of cold spots throughout the pipelines network,-thus serving as backup to the SEEHT temperature monitors (Johnson, 2008).

Note should be made that in the case of supplying the molten sulfur to a forming facility, it is typically cooled with air or water to about its freezing point (250°F) prior to forming in order to increase the quantity and quality of the product produced from any of the available forming processes. It is then filtered to remove any particulates that may plug sulfur spray nozzles or tiny perforations, resulting in reduced production, deteriorating product quality, and eventually, a shutdown to remove the obstructions (Johnson, 2009).

8.6.2 Sulfur Forming

For several decades, the standard industry practice was to ship sulfur as a crushed bulk solid, which was obtained directly from the sulfur block by breaking up the sulfur and crushing the larger lumps to a maximum size of about 0.8 inch. However, crushed bulk contained many fines,[24] and so was very dusty to be handled. This led to the installation of sulfur-forming facilities at the larger plants, where only a small amount of solid sulfur is still shipped as bulk sulfur to markets[25] (Riegel and Kent, 2003).

Bulk solid sulfur is produced in many different forms, which vary in their friability and mechanical strength, using different proprietary technologies and processes. Slating was the sulfur industry's first attempt at developing a formed product that would generate less dust than the traditional crushed bulk sulfur commonly used until 1960. Slate form of sulfur is produced by pouring and spreading liquid sulfur onto a slow-moving belt while simultaneously cooling the belt by air from above and with water from below. By the time it reaches the end of the belt, it has solidified into a solid sheet with a thickness of about 0.1 to 0.2 inch. The sulfur begins to break into smaller pieces (having an irregular shape with many sharp edges) when it is separated from the belt and when sulfur is discharged from the belt at the head pulley. When the slate form of sulfur is handled further, there is a tendency for it to

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[24] Sulfur particles or dust with maximum dimensions not exceeding 0.012 inch are generally considered fines.

[25] The solid sulfur reclaimed from the block is normally melted to provide liquid feed to sulfur-forming facilities and/or liquid shipments.

break into smaller pieces and fine dust. Therefore, slated (flake) sulfur is a less desirable form of material that now faces commercial resistance.

Currently available technologies for solidifying sulfur create prilled, pastilled, or granulated products to minimize the creation of sulfur dust during transport operations. Prilled sulfur is created when sulfur droplets are cooled by a surrounding fluid, which may be air or water. Air-prilled sulfur is created when spraying liquid sulfur from the top of a tower against an upward flow of air. As the sulfur falls, the sulfur droplets break into small droplets and cool into prills. No air-prilling installations have been built recently due the cost of the tower and requisite high-volume air scrubbing system as well as sulfur emissions from the top of the tower into the atmosphere. Wet-prilled sulfur is created when sulfur droplets fall into an agitated water bath, which solidifies and cools the sulfur into uniformly sized prills. The prills are then separated from the water in high-frequency vibrating screens. Pastilled sulfur is created when a feeding system specially developed for sulfur deposits a regular array of liquid sulfur in the form of droplets onto the steel running belt, which is cooled from the underside by water sprayed through nozzles. The drops are cooled along the length of the belt and are discharged in the form of solid hemispherical pastilles onto a conveyor system. Granulated sulfur is created through the gradual buildup of successive layers of sulfur around a central core. As a result, the ultimate product size can be controlled. In the granulation process, small seed particles of sulfur are introduced at the feed end of a long horizontal rotating drum where the particles are spray coated with molten sulfur. Each layer of sulfur is allowed to dry before additional sulfur is sprayed onto the particles. Repeated application of sulfur layers allows the particles to grow until they are discharged at the far end of the drum.

There are a number of factors that may point the sulfur producer to one technology or another. Therefore, a detailed assessment of forming alternatives is recommended prior to selection of the preferred process for each potential project application.

8.6.3 Conveying Formed Sulfur

From the forming facility, the bulk sulfur must then be transported a short distance to the storage area. Rubber belt conveyors have proven to be the best means of transferring large quantities of sulfur from production to storage.[26] During the normal course of handling solid sulfur by a straight-run

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[26] Front-end loaders are often used to transfer bulk sulfur from a stockpile to a hopper feeding a conveyor.

conveyor, some sulfur dust will be generated. To keep this fugitive dust from accumulating within the conveyors, the conveyor enclosures are swept with air. This dust-laden air system, if vented directly, would be a major source of particulate emissions. Therefore, a dust-suppression system must be installed at conveyor transfer points and at load-out stations. This system typically uses a combination of water, compressed air, and a surfactant mixed together in a proprietary foam generator where the foam is applied to the formed sulfur at the abovementioned points and helps suppress dust (Riegel and Kent, 2003).

8.6.4 Storage of Formed Sulfur

Sulfur storage has been incorporated into the design of many sulfur recovery units to allow for accumulation of sulfur during interruptions in the sulfur shipping system or during periods of low sulfur prices. The sulfur is often stored in the open in huge stockpiles, where it is exposed to wind, rain, dust, etc. In some cases, sulfur is stored indoors, where some protection from the elements is available. Although this refinement may be expensive, it is the preferred modern method of storing formed sulfur product awaiting delivery to end users. However, if indoor storage cannot be provided, a location shielded from prevailing winds should be selected to reduce the migration of fugitive dust. Also, the capability to collect and dispose of water effluent off the pad will be required.

8.7 SULFUR DISPOSAL BY ACID GAS INJECTION

The unfavorable economics of constructing small Claus sulfur recovery units, due to their inherent operating difficulties when the H_2S feed concentration is low and the decreased demand and oversupply of elemental sulfur,[27] necessitate an alternative technology. As the issue of greenhouse gases heats up, acid gas compression and reinjection will become viable for even moderate to large operations.[28] Generally, this is found to be a more economical choice if there is a host reservoir. In the acid gas injection (AGI) project, acid gas mixture is compressed through several stages, dehydrated and injected into a reservoir. In many injection schemes, compression

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[27] In the past, the recovered elemental sulfur had considerable value and was sold in the commercial marketplace. However, as the hydrocarbon extraction industry recovers more sulfur from oil sands, sour crude, and subquality natural gas, the supply of sulfur far exceeds the demand.

[28] The injection pressure depends mainly on the pressure in the disposal zone. If the zone is a depleted gas reservoir, then the injection pressure can be quite low initially.

and cooling alone are sufficient to dehydrate the gas to a point where neither free water nor hydrates are a problem. However, in some cases additional dehydration may be required to ensure safe, cost-effective operation (Carroll, 2009).

AGI is a mature technology on a small (less than 10 MMscfd of injected fluid) and medium scale (less than 100 MMscfd). Some of the newer projects are now suggesting the injection on a larger scale (greater than 100 MMscfd). In this case, many of the principles used to design the small- and medium-sized injection schemes can be transferred to the larger schemes. However, the larger injection schemes pose new challenges. Carroll (2009) discusses some of the differences between the small- and large-scale injection schemes. Further details on design and operations of acid gas injection process, which is beyond the scope of this book, can also be found in Carroll (2010).

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CHAPTER 9

Natural Gas Dehydration

9.1 INTRODUCTION

Natural, associated, or tail gas usually contains water, in liquid and/or vapor form, at source and/or as a result of sweetening with an aqueous solution. Operating experience and thorough engineering have proved that it is necessary to reduce and control the water content of gas to ensure safe processing and transmission. The major reasons for removing the water from natural gas are

1. Natural gas in the right conditions can combine with liquid or free water to form solid hydrates that can plug valve fittings or even pipelines.
2. Water can condense in the pipeline, causing slug flow and possible erosion and corrosion.
3. Water vapor increases the volume and decreases the heating value of the gas.
4. Sales gas contracts and/or pipeline specifications often have to meet the maximum water content of 7 lb H_2O per MMscf. This ensures that water-based problems will not hamper downstream operations.[1]

Pipeline drips installed near wellheads and at strategic locations along gathering and trunk lines will eliminate most of the free water lifted from the wells in the gas stream. Multistage separators can also be deployed to ensure the reduction of free water that may be present. However, the removal of the water vapor that exists in solution in natural gas requires a more complex treatment. This treatment consists of "dehydrating" the natural gas, which is accomplished by lowering the dew point temperature of the gas at which water vapor will condense from the gas.

There are several methods of dehydrating natural gas. The most common of these are liquid desiccant (glycol) dehydration, solid desiccant dehydration, and cooling the gas. The first two methods utilize mass transfer of the water molecule into a liquid solvent (glycol solution) or a crystalline structure (dry desiccant). The third method employs cooling below the initial dew point, by expansion and/or refrigeration, to condense the water molecule to the liquid

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[1] The water content in a natural gas stream entering a liquefaction plant must be reduced to less than 0.1 ppmv to prevent hydrate formation and freezing in the cryogenic section of the plant.

phase with the subsequent injection of inhibitor to prevent hydrate formation. Any of these methods may be used to dry gas to a specific water content. Usually, the combination of the water content specification, initial water content, process character, operational nature, and economic factors determine the dehydration method to be utilized. However, the choice of dehydration method is usually between glycol and solid desiccants. These are presented in depth in subsequent portions of this chapter.

Several other dehydration technologies (i.e., membranes, vortex tube and supersonic processes) are less commonly used, and they are not discussed here. There are also many commercially available processes for customized dehydration systems. These types of processes (i.e., IFPEXOL) or solvents (i.e., Selexol) are often designed to enhance conventional equipment performance. However, the suitability of these processes should be evaluated on a case-specific basis.

9.2 WATER CONTENT DETERMINATION

The first step in evaluating and/or designing a gas dehydration system is to determine the water content of the gas. This datum is most important when one designs sour-gas dehydration facilities and estimates water production with sour gas in the plant inlet separator.

Determining the saturation water content of a gas is a standard but complex problem in thermodynamics. There are numerous methods available for determining water contents of natural gases. A detailed discussion of those methods that are perhaps beyond the scope of the present discussion can be found in Carroll (2002, 2009). In general, for acid gas concentrations less than about 30%, existing methods are satisfactory. For higher acid gas concentrations (above 50%), particularly at higher pressures, existing methods can lead to serious errors in estimating water contents (Hubbard, 1993). An appropriate method was introduced by Wichert and Wichert (2003). It is chart based and provides good estimates of the equilibrium water vapor content of sour natural gas for a range of conditions, including H_2S contents of 3–38 mole% with CO_2 contents of 3–43 mole%, pressures from 290 psi to 10,153 psi, and temperatures from 50°F to 347°F. The overall average error of this method is less than 1%. However, a few points showed a discrepancy of more than ± 10%. In the method developed by Wichert and Wichert (2003), the water content of the sour gas is calculated by multiplication of correction factor and the water content of sweet gas from the McKetta and Wehe (1958) chart, as follows:

1. Determine the equilibrium water vapor content of sweet gas at the operating temperature and pressure conditions using the McKetta and Wehe (1958) chart (see Figure 9-1).

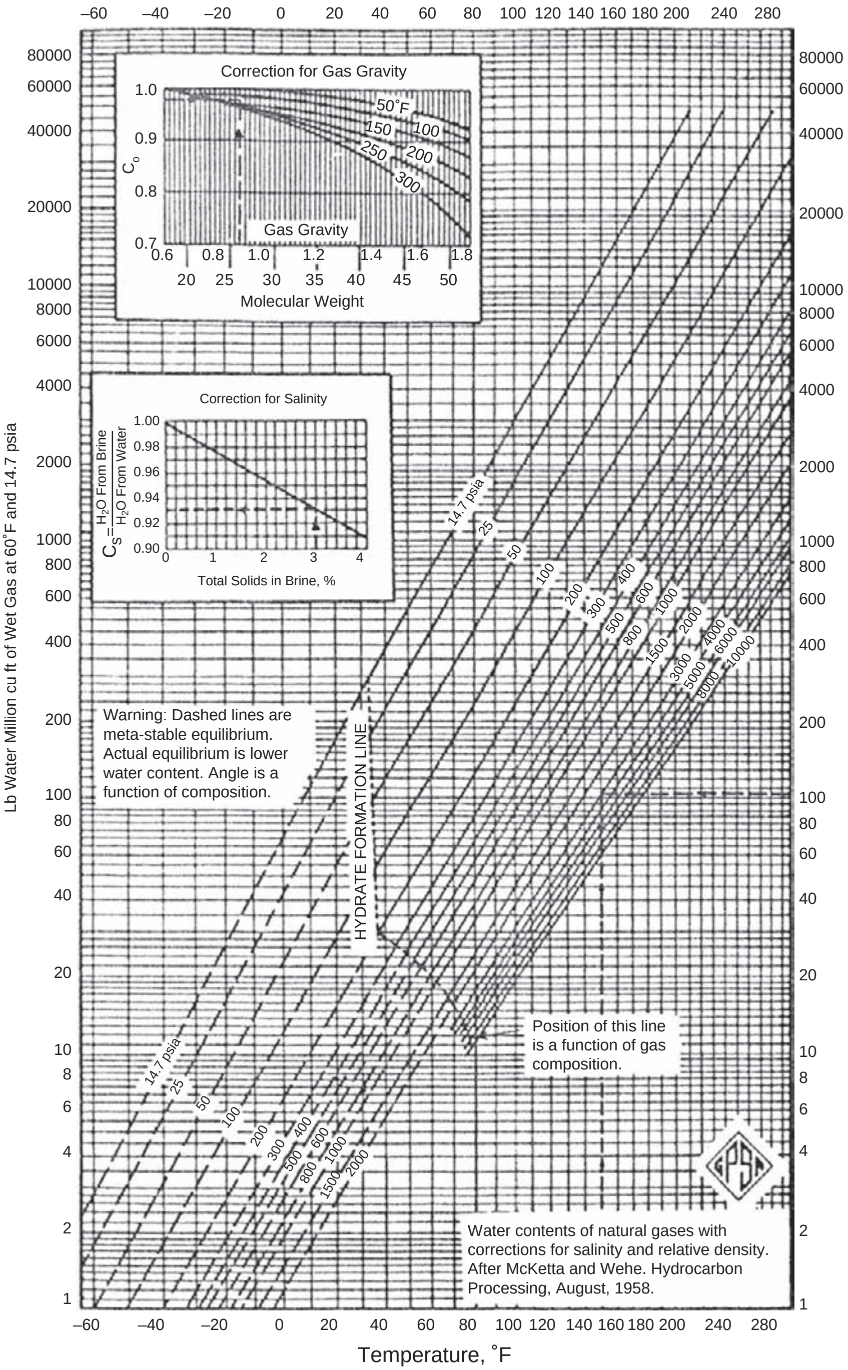

Figure 9-1 McKetta and Wehe (1958) pressure-temperature correlation (GPSA, 2004).

2. Determine the mole% H_2S equivalent concentration of the sour gas as

$$\text{Mole\% } H_2S \text{ equivalent} = \text{mole\% } H_2S + 0.7 \times \text{mole\% } CO_2 \qquad (9\text{–}1)$$

3. From the bottom-left temperature scale in Figure 9-2, move to the right to the mole% H_2S equivalent (interpolate between the lines if necessary).
4. From this point, move to the upper chart, to the pressure of interest. From the pressure point, move to the left, to the ratio scale.
5. Multiply the value from step 4 by the water content determined for sweet gas in step 1. The result is the estimate of the saturated water content of the sour gas at the pressure and temperature of interest.

Bahadori et al. (2009) also developed an easy-to-use correlation-based method, based on the reported experimental data by Lukacs and Robinson (1963), that provides good estimates of the equilibrium water vapor content

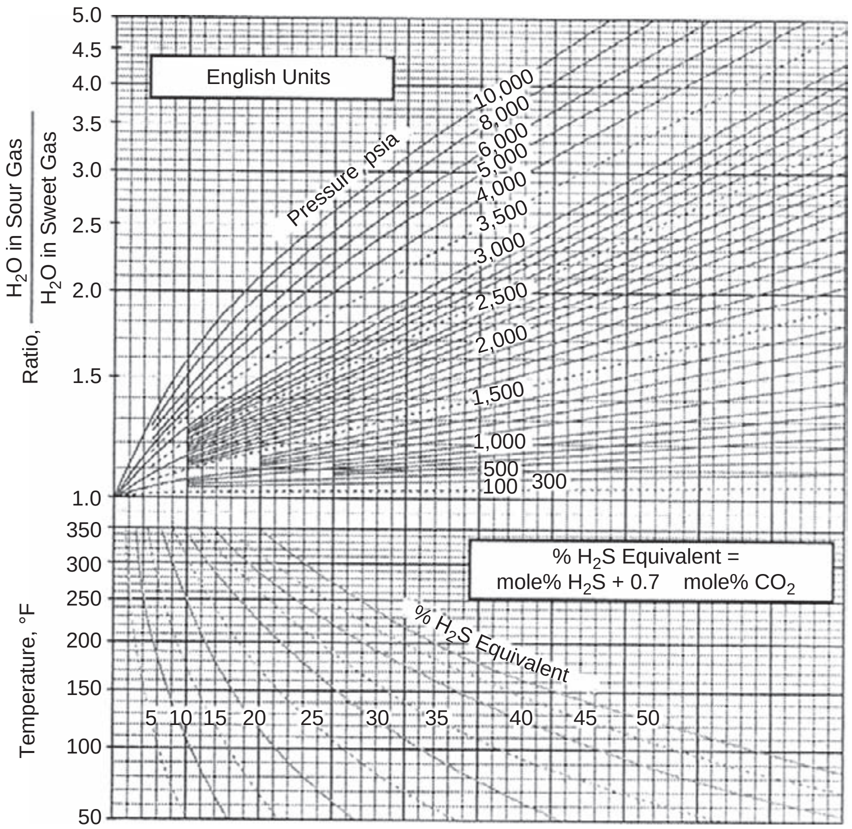

Figure 9-2 Water-content ratio chart (Wichert and Wichert, 2003).

of sour natural gas for a wide range of conditions, including H_2S contents of 0–50 mole% with CO_2 contents of 0–40 mole%, pressures from atmospheric to 14,504 psi for sweet gases and 10,153 psi for sour gases, and temperatures from 50°F to 392°F for sweet gases and from 50°F to 302°F for sour gases. The obtained results of the method developed by Bahadori et al. (2009) show good agreements with the observed values, where the overall average error is about 2.4125%. Therefore, their method can be of immense practical value for designers/engineers who want to quickly check on the water content of sour natural gases at various temperatures and pressures without performing any experimental measurements.

9.3 GLYCOL DEHYDRATION

Among the different gas drying processes, absorption is the most common technique, where the water vapor in the gas stream becomes absorbed in a liquid solvent stream. Although many liquids possess the ability to absorb water from gas, the liquid that is most desirable to use for commercial dehydration purposes should possess the following properties:

1. High absorption efficiency
2. Easy and economic regeneration
3. Noncorrosivity and nontoxicity
4. No operational problems when used in high concentrations
5. No interaction with the hydrocarbon portion of the gas and no contamination by acid gases

Glycols are the most widely used absorption liquids as they approximate the properties that meet the commercial application criteria. Several glycols have been found suitable for commercial application. The commonly available glycols and their uses are described as follows[2] (Katz et al., 1959):

1. Monoethylene glycol (MEG); high vapor equilibrium with gas to tend to lose to gas phase in contactor. Use as hydrate inhibitor where it can be recovered from gas by separation at temperatures below 50°F.
2. Diethylene glycol (DEG); high vapor pressure leads to high losses in contactor. Low decomposition temperature requires low reconcentrator temperature (315°F–340°F) and thus cannot get pure enough for most applications.

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[2] Physical properties of the most common glycols are given in Appendix 3.

3. Triethylene glycol (TEG); Reconcentrate at 340°F–400°F for high purity. At contactor temperatures in excess of 120°F tends to have high vapor losses to gas. Dew point depressions up to 150°F are possible with stripping gas.
4. Tetraethylene glycol (TREG); more expensive than TEG but less loss at high gas contact temperatures. Reconcentrate at 400°F–430°F.

TEG is by far the most common liquid desiccant used in natural gas dehydration. It exhibits most of the desirable criteria of commercial suitability listed here (Manning and Thompson, 1991; Hubbard, 1993):

1. TEG is more easily regenerated to a concentration of 98%–99% in an atmospheric stripper because of its high boiling point and decomposition temperature.
2. TEG has an initial theoretical decomposition temperature of 404°F, while that of diethylene glycol is only 328°F.
3. Vaporization losses are lower than monoethylene glycol or diethylene glycol. Therefore, the TEG can be easily regenerated to the high concentrations needed to meet pipeline water dew point specifications.
4. Capital and operating costs are lower.

Natural gas dehydration with TEG is discussed in the following sections on process description and design considerations.

9.3.1 Process Description

Figure 9-3 shows the scheme of typical TEG dehydration unit. As can be seen, wet natural gas first typically enters an inlet separator to remove all liquid hydrocarbons from the gas stream. Then the gas flows to an absorber (contactor) where it is contacted countercurrently and dried by the lean TEG. TEG also absorbs volatile organic compounds (VOCs) that vaporize with the water in the reboiler.[3] Dry natural gas exiting the absorber passes through a filter coalescer in order to minimize TEG carryover to downstream units and then into the sales line. The wet, or "rich," glycol leaves the absorber and enters a cooling coil that controls the water reflux rate at the top of the stripper. This temperature control ensures that the water vapor leaving the still does not carry over excess glycol. Rich glycol is then flashed to remove dissolved hydrocarbons. The latter can be used for fuel and/or stripping gas. Downstream of the flash drum, the rich TEG is filtered before being heated in the reconcentrator. This prevents impurities such as

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[3] VOC emissions are an environmental challenge for the natural gas industry; hence, glycol dehydration systems require monitoring and control of VOC emissions.

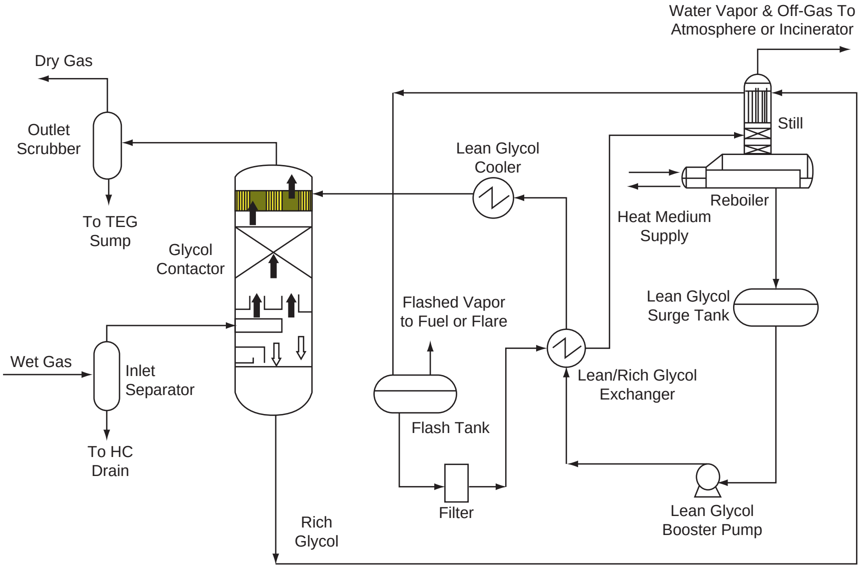

Figure 9-3 Typical flow diagram for TEG dehydration system.

solids from plugging the packed column and fouling the reboiler fire tube. After the glycol/glycol heat exchanger, the rich glycol enters the stripping column and flows down the packed-bed section into the reboiler. Steam generated in the reboiler strips absorbed water and VOCs out of the glycol as it rises up the packed bed. The water vapor and desorbed natural gas are vented from the top of the stripper. The hot, regenerated, lean glycol flows out of the reboiler into the accumulator (surge tank), where it is cooled via cross-exchange with returning rich glycol; it is pumped to a cooler and back to the top of the absorber.

In order to ensure that the unit functions well, a surge vessel for the circulation pumps is also needed; this is connected to the kettle reboiler, which feeds it by gravity.

9.3.2 Design Considerations

The following sections provide more detailed information about each equipment operation and design used in a TEG dehydration unit.

9.3.2.1 Absorber (Contactor)

The incoming wet gas and the lean TEG are contacted countercurrently in the absorber to reduce the water content of the gas to the required specifications. (In an ethylene glycol system, glycol is injected directly into the natural gas stream; therefore, an absorber is not used.) The key design parameters for the absorber are

- Gas flow rate and specific gravity
- Gas temperature
- Operating pressure (gas pressure)
- Outlet dew point or water content required

The amount of water to be removed in a TEG system is calculated from the gas flow rate, the water content of incoming gas, and the desired water content of outgoing gas. The water removal rate, assuming the inlet gas is water saturated, can be determined as

$$W_r = \frac{Q_G(W_i - W_o)}{24} \tag{9-2}$$

where W_r is water removed, lb/hr; W_i is water content of inlet gas, lb/MMscf; W_o is water content of outlet gas, lb/MMscf; and Q_G is gas flow rate, MMscfd.

The glycol circulation rate is determined on the basis of the amount of water to be removed and is usually between 2 and 6 gallons of TEG per pound of water removed, with 3 gallons TEG/lb water being typical. Higher circulation rates provide little additional dehydration while increasing reboiler fuel and pumping requirements. Problems can arise if the TEG circulation rate is too low; therefore, a certain amount of overcirculation is desired. An excessive circulation rate may overload the reboiler and prevent good glycol regeneration. The heat required by the reboiler is directly proportional to the circulation rate. Thus, an increase in circulation rate may decrease reboiler temperature, decreasing lean glycol concentration, and actually decrease the amount of water that is removed by the glycol from the gas. Only if the reboiler temperature remains constant will an increase in circulation rate lower the dew point of the gas. An overly restricted circulation rate can also cause problems with tray hydraulics, contactor performance, and fouling of glycol-to-glycol heat exchangers. Therefore, operators should include a margin of safety, or "comfort zone," when calculating reductions in circulation rates. An optimal circulation rate for each dehydration unit typically ranges from 10% to 30% above the minimum circulation rate (EPA430-B-03-013, 2003). The minimum glycol circulation rate can then be calculated as

$$Q_{\mathrm{TEG,min}} = G \times W_r \tag{9-3}$$

where $Q_{TEG,min}$ is the minimum TEG circulation rate (gal TEG/hr); and G is the glycol-to-water ratio (gal TEG/lb water removed). The industry accepted rule of thumb is 3 gallons of TEG per pound of water removed.

Figure 9-4 shows the effect of TEG concentration and circulation rate on dew point depression for a fixed amount of absorber contact. Notice that the curves become relatively flat at high circulation rates.

The diameter of the absorber and the number of absorber stages are selected on the basis of the gas and glycol flow rates and gas-specific gravity. The diameter of the contactor (absorber) can be estimated from the Souders and Brown (1932) correlation as follows:

$$V_{max} = K_{SB}\left[\frac{\rho_L - \rho_G}{\rho_G}\right] = \left[\frac{4\ Q_G}{\pi\ D^2}\right] \tag{9-4}$$

where D is internal diameter of glycol contactor, ft; Q_G is gas volumetric flow rate, ft^3/hr; V_{max} is maximum superficial gas velocity, ft/hr; K_{SB} is the Souders and Brown coefficient, ft/hr; ρ_L is glycol density, lb/ft^3; and ρ_G is gas density at column condition, lb/ft^3.

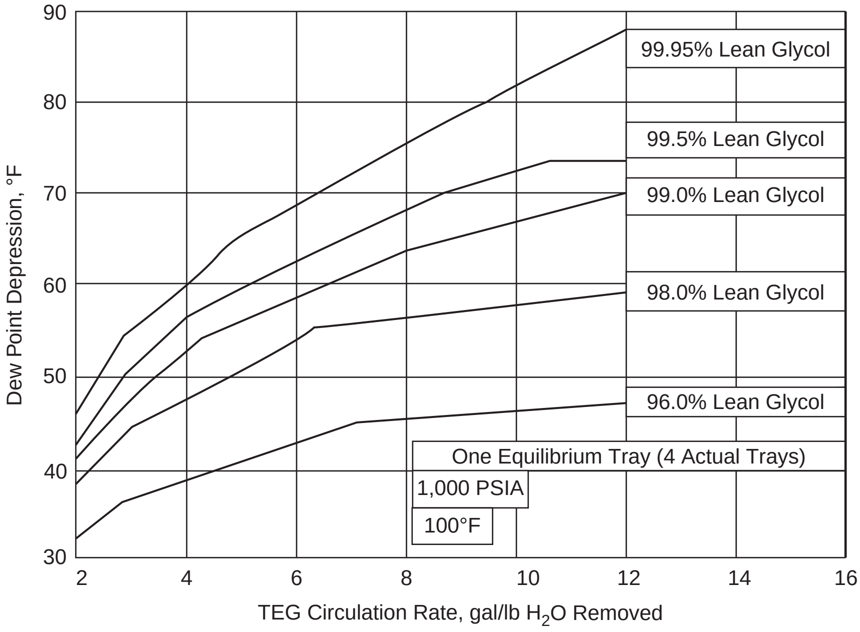

Figure 9-4 Effect of TEG concentration and circulation rate on dew point depression (GPSA, 2004).

Traditionally, the glycol absorber contains 6–12 trays that provide an adequate contact area between the gas and the glycol (Kean et al., 1991). The more trays, the greater the dew point depression for a constant glycol circulation rate and lean glycol concentration. Conversely, if more trays with the same TEG concentration are specified, a lower circulation rate is required. When more trays are specified, fuel savings can be realized because the heat duty of the reboiler is directly related to the glycol circulation rate.

Although either bubble cap trays or valve trays may be used, some operators prefer bubble cap trays because they are suitable for viscous liquids, handle high turndown ratios and low liquid/gas ratios well, and are not subject to weeping. Calculated tray efficiency values are dependent on the TEG/water equilibrium data used. To achieve an accurate design method, column efficiencies consistent with the accurate equilibrium data must be therefore recommended. There are still uncertainties in equilibrium data for the TEG/water/natural gas system. However, when accurate equilibrium data are used, an overall bubble cap tray efficiency of 40%–50% and a Murphree efficiency of 55%–70% can be expected at normal absorption conditions, 86°F and 99%–99.5% wt TEG[4] (Oi, 1999). Earlier, overall tray efficiencies between 25% and 40% were recommended for design. This is regarded as too conservative. It has been suggested that using 50% Murphree efficiency based on accurate phase equilibrium data should give a conservative design at normal high-pressure dehydration conditions (Oi, 2003).

The standard tray spacing in glycol contactors is 24 inches; closer spacing increases glycol losses if foaming occurs because of greater entrainment. The total height of the contactor column will be based on the number of trays required plus an additional 6–10 feet to allow space for a vapor disengagement area above the top tray and an inlet gas area at the bottom of the column.

One option to the trayed TEG contactor is the use of structured packing. Structured packing was developed as an alternative to random packing to improve mass transfer control by the use of a fixed orientation of the transfer surface.[5] The combination of high gas capacity and reduced height of an equilibrium stage, compared with trayed contactors, makes the application of structured packing desirable for both new contactor designs and existing

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[4] The Murphree tray efficiency (for tray number *n*) is defined as (the mole fraction in the gas from tray *n* – the mole fraction from the tray below)/(the mole fraction in equilibrium with the liquid at tray *n* – the mole fraction from the tray below).

[5] Random packings are mainly used in small columns and are not recommended in glycol contactors due to low glycol flow rates versus gas flow.

trayed-contactor capacity upgrades. Hence, the structured packing may offer potential cost savings over trays (Kean et al., 1991). A detailed discussion on calculating column diameter and the required packing height can be found in Ghoshal and Mukhopadhyay (1993).

The absorber is usually vertical to allow proper glycol flow with sufficient gas/liquid contact and operates at the pressure of the incoming gas. A demister pad or mist eliminator at the top of the absorber or a separator vessel following the absorber can reduce glycol losses by preventing glycol from being carried out with the dry gas. Egger and Breu (2010) explained design characteristics for the different mass transfer and demisting internals, and compared capacity, efficiency, pressure drop, glycol loss, sensitivity to fouling and moving conditions, operating pressure, and costs.

9.3.2.2 Still (Stripper)

The still or stripper column is used in conjunction with the reboiler to regenerate the glycol. On many dehydrators, the still is placed vertically on top of the reboiler so that vapor from the reboiler directly enters the bottom of the distillation column. A given lean TEG concentration is produced in the reboiler and still column (regenerator) section by control of reboiler temperature, pressure, and the possible use of a stripping gas. As shown in Figure 9-5, the reboiler temperature controls the concentration of the water in the lean glycol. Reboiler temperatures for TEG are limited to 400°F, which limits the maximum lean glycol concentration without stripping gas. Some operators limit the reboiler temperature to between 370°F and 390°F to minimize degradation of the glycol. This effectively limits the lean glycol concentration to between 98.5% and 98.9%.

There are improved regeneration techniques that have higher glycol concentration and, therefore, lower treated gas dew points. When dry (stripping) gas is injected into the base of the glycol reboiler in order to (1) strip off water vapor from the glycol by reducing the vapor partial pressure and (2) agitate the glycol to accelerate the release of water vapor, TEG concentration increases from 99.1% to 99.6% by weight. A dry gas injection process can be enhanced using a packed column of countercurrent gas stripping, which increases the capability of TEG for gas dehydration by reconcentrating the glycol to as high as 99.6%. The Drizo process, developed by Dow Chemical Co., also uses hydrocarbon solvents as azeotropic stripping agents for improving trace quantities of water in the TEG. The process performs a tertiary azeotropic distillation resulting in glycol concentrations to 99.9% (Smith and Tsai, 1987). This process avoids flaring stripping gas and reduces

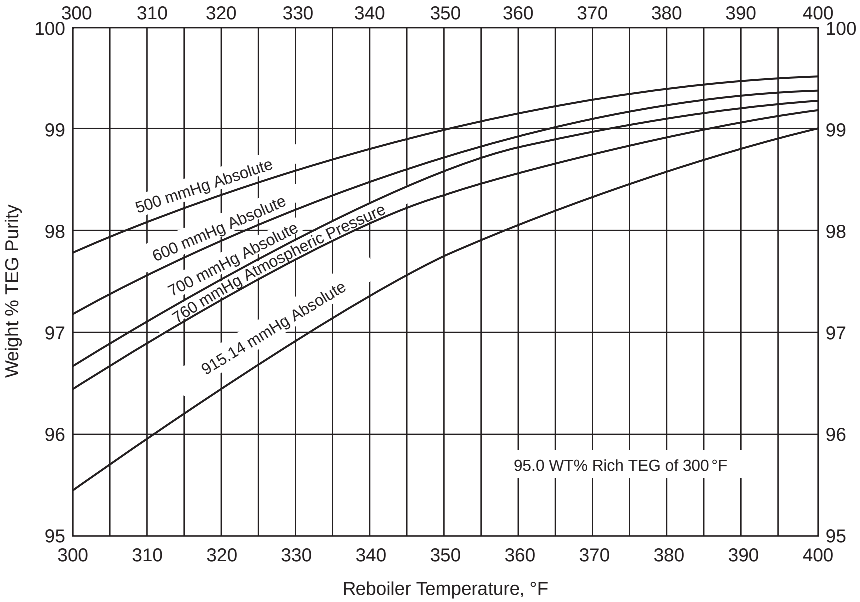

Figure 9-5 Glycol purity versus reboiler temperature at different levels of vacuum (GPSA, 2004).

operating costs. Other patented processes in use to enhance the glycol purity and thereby achieve a more stringent water dew point depression (CLEANOL+, COLDFINGER, PROGLY, and ECOTEG) are described in GPSA (2004).

The diameter of the still is based on the liquid load (rich glycol and reflux) and the vapor load (water vapor and stripping gas). Manufacturers' chart or standard sizes based on the required reboiler heat load may be used to determine the column diameter (Caldwell, 1976; Sivalls, 1976). Alternatively, the following approximate equation is used:

$$D = 9(Q_{\mathrm{TEG}})^{0.5} \tag{9-5}$$

where Q_{TEG} is TEG circulation rate, gal/min; and D is inside diameter of stripping column, inch.

Smaller-diameter towers (less than 2 feet in diameter) are often packed with ceramic Intalox saddles or stainless steel Pall rings instead of trays. Larger-diameter towers may be designed with 10–20 bubble cap trays or structured packing.

To prevent excessive glycol losses from vaporization at the top of the still column, reflux is controlled by a condenser maintained at about 215°F or lower if stripping gas is used. A few larger gas dehydration units may use a tubular water-cooled condenser with temperature control. Temperature control can also be obtained by circulating cool rich glycol through a reflux coil inside the top of the stripping column. This system normally includes a bypass valve that allows the operator to better control the temperature at the top of the column. Many smaller field dehydrators employ a finned atmospheric condenser at the outlet at the top of the still.

9.3.2.3 Reboiler

The reboiler and still are typically a single piece of equipment. The reboiler supplies heat to regenerate the rich glycol in the still by simple distillation. The separation is relatively easy because of the wide difference in boiling points of water and glycol. Most remote field locations use a direct-fired firebox to provide the heat for vaporization. Larger dehydrator systems may use indirect heat sources such as dowtherm heat transfer fluid (hot oil), electricity, or medium pressure steams.

Reboiler duty can be estimated using the following equation (Sivalls, 1976):

$$Q_R = 900 + 966(G) \tag{9-6}$$

where Q_R is regenerator duty, Btu/lb H_2O removed; and G is glycol-to-water ratio, gal TEG/lb H_2O removed.

This estimate does not include stripping gas and makes no allowance for combustion efficiency.

For a TEG system, the reboiler normally operates at a temperature of 350°F–400°F, where this temperature controls the lean glycol water concentration. The purity of the lean glycol can be increased by raising the reboiler temperature. However, the reboiler should not be operated above 400°F to allow a safety margin to prevent decomposition (TEG starts to decompose at 404°F), and the burner should have a high temperature shutdown for safety.

9.3.2.4 Surge Tank (Accumulator)

Lean glycol from the reboiler is routed through an overflow pipe or weir to a surge tank or accumulator. This vessel is not insulated in many cases, so the lean glycol is cooled to some extent via heat loss from the shell. The surge tank may also contain a glycol/glycol heat exchanger. If the surge tank contains the glycol/glycol heat exchanger, the tank should be sized to allow a

30-minute retention time for the lean glycol. Some designs also provide for a separate lean glycol storage tank. The storage tank, when used as a surge tank, may be vented to allow accumulated gases to escape, or it is sometimes fitted with an inert gas blanket to prevent the oxygen from contacting the glycol and causing oxidation. Venting from the surge tank is only a minor emission source because the lean glycol has already been stripped of most VOCs in the still.

9.3.2.5 Heat Exchanger

Two types of heat exchangers are found in glycol dehydration systems: glycol/glycol exchangers and gas/glycol exchangers.

A glycol/glycol exchanger cools the lean glycol while preheating the rich glycol. It may be an external exchanger or may be located within the surge tank (an integral exchanger). External heat exchangers can preheat the rich glycol to about 300°F. However, for small standard designs, the integral exchanger is economical to fabricate but may not heat the rich glycol above 200°F.

A dry gas/lean glycol exchanger uses the exiting dry natural gas to control the lean glycol temperature to the absorber. High glycol temperatures relative to the gas temperature reduce TEG's moisture-absorption capacity. Conversely, temperatures that are too low promote glycol loss due to foaming and increase the glycol's hydrocarbon uptake and potential still-vent emissions (Kirchgessner et al., 2004).

Heat exchangers can be sized using conventional design procedures. Manning and Thompson (1991) described typical guidelines.

9.3.2.6 Flash Tank (Phase Separator)

Many glycol dehydration units contain a flash tank (phase separator), operating at 50–100 psia and 100°F–150°F, that removes dissolved gases/liquid hydrocarbons from the warm rich glycol and reduces VOC emissions from the still. The flash gas from the separator can be used as supplemental fuel gas or as stripping gas on the reboiler.

If the gas has a negligible heavy hydrocarbon content, the separator (flash drum) only needs to degas the glycol. In this case, it can be a vertical two-phase separator with modest holdup time (5–10 minutes). When the gas is rich in light ends, the rich TEG may also contain liquid hydrocarbons. This could result in emissions or cause excess glycol losses from the still vent and must therefore be removed as efficiently as possible. For this purpose, a three-phase horizontal separator is used, with long liquid holdup times (20–30 minutes).

Because the hydrocarbons are collected under a vacuum, they are stable, and no vapor losses or weathering occur (Kirchgessner et al., 2004).

9.3.2.7 Glycol Circulation Pumps

A circulation pump is used to move the glycol through the unit. Various pump and driver types are used in glycol systems, including gas/glycol-powered positive displacement or glycol balance pumps (e.g., Kimray pumps) and electric motor-driven reciprocating or centrifugal pumps. The gas/glycol pump is common in field TEG dehydrators where electricity is typically not available. This type of pump uses the high-pressure glycol, leaving the absorber to provide part of its required driving energy. Gas, taken under pressure from the absorber, is used to supply the remaining driving energy. This additional pump gas accounts for up to 8 scf/gal (depending on the absorber pressure) and should be recovered in the phase separator and used to the extent possible as fuel or stripping gas in the reboiler. Larger dehydrators in plants may use motor-driven pumps for higher efficiency. Where feasible, using electric pumps as alternatives to glycol balance pumps can yield significant economic and environmental benefits, including a financial return through reduced gas losses, increased operational efficiency, and reduced maintenance costs (EPA430-B-03-014, 2004).

9.3.2.8 Filters

Two types of filters are commonly used in glycol systems: fabric (particulate) and activated carbon filters.

Fabric filters (e.g., sock) are used to remove particulate matter. In fact, the suspended solids content of glycol should be kept below 0.01 wt% to minimize pump wear, plugging of exchangers, fouling of absorber trays and stripper packing, solids deposition on the firetube in the reboiler, and glycol foaming. Solids filters are selected to remove particles with a diameter of 5 μm and larger. A common design is a 3-inch diameter by 36-inch-long cylindrical element in housing, sized for a flow rate of 1–2 gallons per minute per element. The filters are sized for a 12–15 psi pressure drop when plugged. The preferred location for solids filtration is on the high-pressure side at the bottom of the absorber. This filter will remove any foreign solid particles picked up in the absorber before entering the glycol pump. A low-pressure glycol filter can also be installed between the glycol/glycol exchanger and the reboiler for added filtration. An advantage of this placement is that the viscosity of the glycol is lower after it has been warmed in the glycol/glycol exchanger.

Activated carbon filters are used to remove dissolved impurities in the glycol, such as high-boiling hydrocarbons, surfactants, well-treating chemicals, compressor lubricants, and TEG degradation products. The preferred type of carbon is the hard, dense, coal-based carbon. The carbon filter should be located downstream of the sock filter to prevent the carbon filter from becoming plugged with particles. The most common arrangement for a carbon filter is as a slipstream filter, with a minimum flow rate of 20% of the total flow. A preferred arrangement is to use a full-flow carbon filter vessel. Such a vessel is sized for a residence time of 15–20 minutes, with a superficial velocity of 2–3 gal/min/ft^2.

9.3.3 Operational Problems

The operating problems associated with each piece of equipment in the TEG dehydration unit are described individually in the following sections (Ghoshal and Mukhopadhyay, 1993; Hubbard, 1993; Gas Technology Institute, 2001; Chakraborty and Bagde, 2004).

9.3.3.1 Absorber

The main operating problems associated with the absorber are insufficient dehydration, foaming, and hydrocarbon solubility in glycol, which are discussed next.

9.3.3.1.1 Insufficient Dehydration

Causes of insufficient dehydration (i.e., wet sales gas) include excessive water content in the lean glycol, inadequate absorber design, high inlet gas temperature, low lean glycol temperature, and overcirculation/undercirculation of glycol. Lean glycol purity (i.e., glycol concentration) plays a vital role in the rate of absorption of moisture. A minimum lean glycol concentration is therefore needed to achieve a specified dew point depression. Higher water concentrations in the lean glycol will result in poor dehydration.

The outlet gas moisture dew point will indicate the performance of the absorber. Inadequate absorber design most often occurs when a glycol unit is moved from its original field. In fact, gases with the same flow rate can contain very different quantities of water, depending on the field temperatures, pressures, and gas composition. Therefore, when a glycol unit is moved to a different facility, the water load on the unit should be checked.

Temperature of the inlet gas dictates the amount of water fed into the unit, where lower inlet gas temperature will require less water to be removed by the glycol. Hence, performance of the sweet gas cooler as well as lean amine cooler of the gas sweetening unit needs to be monitored.

Lean glycol temperature at the top of the absorber will affect the water partial pressure at the top stage, where high TEG temperature may cause high moisture content of the outlet gas. Reboiler temperature can therefore be increased up to 400°F, above which glycol degradation starts.

9.3.3.1.2 Foaming

Foaming causes glycol to be carried out of the absorber top with the gas stream, resulting in large glycol losses and decreased glycol unit efficiency. Foaming can normally be traced to mechanical or chemical causes. High gas velocity is usually the source of mechanical entrainment. At excessive velocities, glycol can be lifted off the trays and out of the vessel with the gas. High velocity can be caused by poor design, operation at gas flow rates above design levels, or damaged/plugged trays/packing. Although an efficient demister pad is normally fitted at the top of the absorber, extremely high gas velocities may carry glycol through the demister.

Chemical foaming is caused by contaminants in the glycol, liquid hydrocarbons, well-treating chemicals, salts, and solids. Adequate inlet separation and filtration systems (cartridge filter and activated carbon bed) are therefore needed to prevent foaming due to chemical contamination. The filters are generally effective until they become plugged by particulate matter (indicated by a high pressure drop across the filter) or saturated with hydrocarbons; thus, the only operational issue is the filter replacement frequency.

9.3.3.1.3 Hydrocarbon Solubility in TEG Solution

Aromatic hydrocarbon solubility in glycols is a significant issue in the gas dehydration technology due to the potential release of aromatics to the atmosphere at the regenerator. In fact, in the absorber, TEG can absorb significant amounts of aromatic components in the gas [benzene, toluene, ethyl benzene, and xylene (BTEX)], which are often released to the atmosphere at the regenerator. While these emissions are generally small on a mass basis, they have received a great deal of attention from environmental and safety regulatory agencies. Several BTEX emission mitigation methods have been proposed. By far, the most common method implemented to date is condensation of the BTEX components in the regenerator overhead and subsequent separation from the condensed water. This scheme is simple and relatively low cost although it does complicate water disposal due to the high solubility of BTEX in water.

9.3.3.2 Still (Stripper)

The major operational problem with the still is excessive glycol losses due to vaporization. The TEG concentration in the vapor (and thus the glycol vaporization losses) increases significantly above 250°F. The appearance of the plume leaving the glycol still can identify excess glycol vaporization. Since glycol is heavier than air, a plume sinking to the ground instead of rising indicates that glycol is being vaporized. Excessive glycol vaporization is more of a problem for finned atmospheric condensers than for water-cooled or glycol-cooled condensers. Although finned atmospheric condensers are simple and inexpensive, they are sensitive to extremes in ambient temperature. For example, during cold winter periods, a low temperature at the top of the still column causes excessive condensation and floods the reboiler. This prevents adequate regeneration of the glycol and reduces the potential dew point depression of the glycol, causing insufficient dehydration in the absorber. Also, excessive glycol losses may occur as the reboiler pressure increases and blows the liquids out the top of the column. During the summer months, inadequate cooling may allow excessive glycol vaporization losses.

9.3.3.3 Reboiler

Operational problems associated with the reboiler include salt contamination, glycol degradation, and acid gas-related problems.

9.3.3.3.1 Salt Contamination

Carryover of brine solutions from the field can lead to salt contamination in the glycol system. Sodium salts (typically sodium chloride, NaCl) are a source of problems in the reboiler since NaCl is less soluble in hot TEG than in cool TEG; NaCl will precipitate from the solution at typical reboiler temperatures of 350°F–400°F. The salt can deposit on the firetube, restricting heat transfer. If this occurs, the surface temperature of the fire tube will increase, causing hot spots and increased thermal degradation of the glycol. The deposition of salt may also result in corrosion of the fire tube. Dissolved salts cannot be removed by filtration. As a general rule, when the salt content reaches 1%, the glycol should be drained and reclaimed. If the level of salts is allowed to increase beyond 1%, both severe corrosion and thermal degradation threaten the system.

9.3.3.3.2 Glycol Degradation

Glycol degradation is caused primarily by either oxidation or thermal degradation. Glycol readily oxidizes to form corrosive acids. Oxygen can enter the system with incoming gas, from unblanketed storage tanks or sumps, and through packing glands. Although oxidation inhibitors (such as a 50-50 blend of monoethanolamine (MEA) and 33% hydrazine solution) can be used to minimize corrosion, a better approach to controlling oxidation is blanketing the glycol with natural gas, which can be applied to the headspace in storage tanks and any other area where glycol may contact oxygen. Thermal degradation of the glycol results from the following conditions: high reboiler temperature, high heat flux rate, and localized overheating. The reboiler temperature should always be kept below 402°F to prevent degradation, and a good firetube design should inherently prevent a high heat flux rate. Localized overheating can be caused by deposits of salts or hydrocarbons. In addition, thermal degradation of the glycol produces acidic degradation products that lower the pH and increase the rate of degradation, creating a destructive cycle.

9.3.3.3.3 Acid Gas

Some natural gas contains H_2S and/or CO_2, and these acid gases may be absorbed in the glycol. Acid gases can be stripped in the reboiler and still. Mono-, di-, or triethanolamine may be added to the glycol to provide corrosion protection from the acid gases.

9.3.3.4 Surge Tank

When surge tanks also serve as glycol/glycol heat exchangers, the level must be monitored to ensure that the lean glycol covers the rich glycol coil. Otherwise, inadequate heat exchange will occur, and the lean glycol will enter the absorber at an excessively high temperature.

9.3.3.5 Heat Exchanger

The primary operational problem with the heat exchangers is poor heat transfer, which results in lean glycol that is too warm. When this occurs, poor dehydration and insufficient dew point depression can result. Also, glycol vaporization losses to the product gas may be higher with increased lean glycol temperature. Poor heat transfer and the resulting high lean glycol temperature are usually caused by fouled heat exchangers, undersized heat exchangers, or overcirculation. Exchangers may be fouled by deposits such as salt, solids, coke, or gum. In the case of undersized exchangers, additional heat exchangers may be required. Corrosion of the coil in surge tank heat exchangers can also present operating problems since it can lead to cross-contamination of rich and lean glycol.

9.3.3.6 Phase Separator (Flash Tank)

Inadequate residence time in the phase separator may result in a large quantity of glycol being included in the hydrocarbon stream and vice versa. This is most likely a result of overcirculation of the glycol. The effects of hydrocarbons and overcirculation are discussed next.

9.3.3.7 Glycol Circulation Pump

The major problems associated with the circulation pump and rates are related to reliability, pump wear, and overcirculation or undercirculation.

9.3.3.7.1 Reliability

Pump reliability is important because pumps are the only moving parts in the entire dehydrator system. It is good design practice to include a strainer or sock filter in the pump suction line to prevent damage by foreign objects. Pump reliability is also enhanced by limiting the lean glycol temperature to 180°F–200°F, and ensuring good filtration. Pump wear, leakage, and failures increase if the glycol becomes dirty or hot. In severe cases, glycol losses of as much as 35 gal/day from seal leakage have been observed.

9.3.3.7.2 Pump Wear

As the O-rings and seals on the glycol balance pumps wear out, there is the potential for contamination of the lean glycol by the rich glycol. This increases the water content of the lean glycol and may cause (1) the gas to no longer be dried to pipeline specifications; and/or (2) the operator to increase the glycol circulation rate (and therefore the emissions) in an effort to compensate for the wetter lean glycol. Because of the leakage, it may also be difficult to determine an accurate glycol circulation rate.

9.3.3.7.3 Overcirculation/Undercirculation

Excessively high glycol circulation rates can lead to many problems. If the unit is overcirculating the glycol, the lean glycol may have insufficient heat exchangers to be properly cooled, and the resulting hot lean glycol may not achieve the desired water removal rate. A high circulation rate may not allow adequate residence time in the phase separator for the hydrocarbons to be removed, which may lead to hydrocarbon deposits, glycol losses, foaming, and emissions. Excessive glycol circulation rates can also result in increased sensible heat requirements in the reboiler. Also, because emissions are proportional to the circulation rate, overcirculation results in greater VOCs emissions.

Undercirculating the glycol provides an insufficient quantity of glycol in the absorber for the quantity of water to be removed and results in wet sales gas.

Considering the aforementioned matters, glycol flow rate should be optimized, where it can be done by checking the treated gas moisture dew point.

9.3.4 Future Technology Developments

Glycol dehydration will continue to be a workhorse in gas processing applications due to its proven and robust performance. However, there are a few items utilizing this conventional method that will require slightly different design and operating considerations. Over the past few years, most of the research and development work has focused on environmental issues (in particular, aromatic hydrocarbon emissions) or cost reduction, such as more effective equipment design (high-capacity internals in glycol contactors) or improved operating efficiency. Future developments will therefore dramatically increase the range and efficiency of operation and undoubtedly decrease the footprint of glycol units, both from the standpoint of space and environmental emissions (Hubbard and Clinton, 2008).

9.4 SOLID DESICCANT DEHYDRATION

Adsorption (or solid bed) dehydration is the process in which a solid desiccant is used for the removal of water vapor from a gas stream. The mechanisms of adsorption on a surface are of two types: physical and chemical. In physical adsorption, the bonding between the adsorbed species and the solid phase holds liquids and solids together and gives them their structure. In chemical adsorption, involving a chemical reaction that is termed "chemisorption," a much stronger chemical bonding occurs between the surface and the adsorbed molecules. Chemical adsorption processes find very limited application in gas processing. This section considers only physical adsorption, and all references to adsorption mean physical adsorption.

Physical adsorption is an equilibrium process, where for a given vapor-phase concentration (partial pressure) and temperature, an equilibrium concentration exists on the adsorbent surface that is the maximum concentration of the adsorbate on the surface. The measurement of the amount of gas adsorbed over a range of partial pressures at a single temperature results in a graph known as an adsorption isotherm, which can have very different shapes depending on the type of adsorbent, the type of adsorbate, and intermolecular interactions between the gas and the surface. In addition to concentration (i.e., partial pressure for gases), two properties of the adsorbate (polarity and size) dictate its concentration on the absorbent surface (Kidnay and Parrish, 2006).

9.4.1 Desiccant Capacity

The capacity of a desiccant for water is expressed normally in mass of water adsorbed per mass of desiccant. The dynamic moisture-sorption capacity of a desiccant will depend on a number of factors, such as the relative humidity of the inlet gas, the gas flow rate, the temperature of the adsorption zone, the mesh size of the granule, and the length of service and degree of contamination of the desiccant (Huntington, 1950) and not the least on the desiccant itself. Moisture-sorption capacity is not affected by variations in pressure, except where pressure may affect the other variables listed here. There are three capacity terms used as follows (Campbell, 2000):

- *Static equilibrium capacity:* The water capacity of new, virgin desiccant as determined in an equilibrium cell with no fluid flow (corresponding to the adsorption isotherm).
- *Dynamic equilibrium capacity:* The water capacity of desiccant where the fluid is flowing through the desiccant at a commercial rate.
- *Useful capacity:* The design capacity that recognizes loss of desiccant capacity with time as determined by experience and economic consideration and the fact that all of the desiccant bed can never be fully utilized.

9.4.2 Desiccant Selection

Various solid desiccants are available in the market for specific applications. Some are good only for dehydrating the gas, whereas others are capable of performing both dehydration and removal of heavy hydrocarbon components. The selection of proper desiccant for a given application is a complex problem. For solid desiccants used in gas dehydration, the following properties are desirable (Campbell, 2000; Daiminger and Lind, 2004):

1. High adsorption capacity at equilibrium. This lowers the required adsorbent volume, allowing for the use of smaller vessels with reduced capital expenditures and reduced heat input for regeneration.
2. High selectivity. This minimizes the undesirable removal of valuable components and reduces overall operating expenses.
3. Easy regeneration. The relatively low regeneration temperature minimizes overall energy requirements and operating expenses.
4. Low pressure drop.

5. Good mechanical properties (such as high crush strength, low attrition, low dust formation, and high stability against aging). These factors lower overall maintenance requirements by reducing the frequency of adsorbent change out and minimizing downtime-related losses in production.
6. Cheap, noncorrosive, nontoxic, chemically inert, high bulk density and no significant volume changes upon adsorption and desorption of water.

The most common commercial desiccants used in solid bed dehydrators are molecular sieves, silica gels (i.e., Sorbead), and activated alumina.

Molecular sieves are crystalline alkali metal alumino-silicates comprising a three-dimensional interconnecting network of silica and alumina tetrahedral.[6] The structure is an array of cavities connected by uniform pores with diameters ranging from about 3°A–10°A, depending on the sieve type. Of these, 4A is the most common, but the smaller pore 3A is sometimes preferred to standard 4A for minimizing coadsorption of CO_2 and H_2S if they are present in the feed. If both oxygen and H_2S are present, 3A reduces the production of elemental sulfur that can block adsorbent pores. However, plant operators usually have little incentive to use 3A for dehydrating gas going to hydrocarbon recovery. Bruijn et al. (2002) and Meyer (2005) discussed in detail the different types of molecular sieves and their applications.

A molecular sieve is the most versatile adsorbent because it can be manufactured for a specific pore size, depending on the application. This sieve

- Is capable of dehydration to less than 0.1 ppm water content.
- Is the overwhelming choice for dehydration prior to cryogenic processes.
- Is the most expensive adsorbent but offers greater dehydration.
- Requires higher temperatures for regeneration, thus higher operating cost.

A molecular sieve dehydration system is also an alternative to the Drizo process. But because of multiple high-pressure and high-temperature vessels, the installed cost of a molecular sieve system is two or three times more than an equivalent Drizo system (Dow Chemical Co., 1985). Molecular sieves are alkaline and subject to attack by acids. However, special acid-resistant sieves are available for very sour gases.

Silica gel (a generic name for a gel manufactured from sulfuric acid and sodium silicate) is a widely used desiccant that can be used for gas and liquid dehydration and hydrocarbon recovery from natural gas. It has the following characteristics:

- It is best suited for normal dehydration of natural gas.
- It is more easily regenerated than molecular sieves.

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[6] Through the use of a binder (clay, silica, alumina, etc.), molecular sieves are produced in different aggregates (beads, pellets, mesh).

- It has a high water capacity, where it can adsorb up to 45% of its own weight in water.
- It costs less than a molecular sieve.
- It is capable of dew points to −140°F.

Silica gel used for natural gas drying should be of Sorbead type because this is the water-stable silica gel type. Most other silica gel types will produce fines in contact with water. Engelhard Sorbead is a high-performance, extremely robust silica gel adsorbent primarily used for control of hydrocarbon dew point in natural gas. However, it can also be used for dehydration only; the main benefit, then, is a longer lifetime. High adsorption capacity, drying performance, and low dew points (−158°F) are characteristic of Sorbead. Sorbead adsorbents come in a range of sizes and physical characteristics to fit any manufacturing environment. Their longer life reduces operating costs while their high performance enhances the operating safety of natural gas treatment plants, among others. Since silica gels are acidic, they can handle sour gases, but not alkaline materials such as caustic or ammonia. Although there is no reaction with H_2S, sulfur can deposit and block their surface. Therefore, gels are useful if the H_2S content is less than 5%–6%.

A hydrated form of aluminum oxide (Al_2O_3), alumina is the least expensive adsorbent for gas dehydration. It is alkaline and cannot be used in the presence of acid gases. There are several types of alumina available for use as a solid desiccant. Activated alumina is a manufactured or naturally occurring form of aluminum oxide that is activated by heating. It is widely used for gas and liquid dehydration,[7] and it will produce a dew point below −158°F if applied properly. Less heat is required to regenerate alumina than for molecular sieve, and the regeneration temperature is lower. However, molecular sieves give lower outlet water dew points (GPSA, 2004).

Table 9-1 lists the more important properties of three adsorbents compiled primarily from commercial literature. The properties are representative and vary between manufacturers.

It should be noted that no desiccant is perfect or best for all applications. In some applications, the desiccant choice is determined primarily by economics. Sometimes the process conditions control the desiccant choice. For example, silica gels are used mostly where a high concentration of water vapor is present in the feed, and low levels of water in the dehydrated gas are

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[7] It has good resistance to liquids but little resistance to disintegration due to mechanical agitation by the flowing gas.

Table 9-1 Properties of Solid Desiccants (Campbell, 2000)

Desiccant	Silica Gel	Activated Alumina	Molecular Sieves
Pore diameter, Å	10–90	15	3, 4, 5, 10
Bulk density, lb/ft^3	45	44–48	43–47
Heat capacity, Btu/lb°F	0.22	0.24	0.23
Minimum dew point, °F	–60 to –90	–60 to –90	–100 to –300
Design capacity, wt%	4–20	11–15	8–16
Regeneration stream temperature, °F	300–500	350–500	425–550
Heat of adsorption, Btu/lb	—	—	1800

not needed. Aluminas are used for moderate levels of water in the feed when low levels of water in the product are not required. However, for gas going into cryogenic processing, the only adsorbent that can obtain the required dehydration is a molecular sieve. They can guarantee ppm impurities levels using regenerable, standalone, and flexible units during several years of continuous service. However, depending on the type of feed and species to remove, a careful selection among different grades of molecular sieves is made according to their various properties (4A for water, 5A for light sulfur, 13X for heavy and branched sulfur species). Another important aspect is that it is sometimes better to use a compound bed with successive layers of molecular sieves in order to adsorb the different impurities. This helps reducing both column size and undesired side effects like coadsorption (CECA, 2008). Blachman and McHugh (2000) discussed the use of multiple adsorbents in the same bed for applications in which both higher water concentrations and acid gases are present. However, to ensure an efficient and successful operation of an adsorption unit, vessel layout plays a very important part in extending the lifetime of the installation. CECA's experience and know-how have highlighted some key points for designing the standard vessel layout (CECA, 2009a).

9.4.3.1 Adsorption Technology

The theory of solid bed adsorption has been thoroughly reviewed in a number of publications. In commercial practice, adsorption is carried out in a vertical, fixed bed of adsorbent, with the feed gas flowing down through the bed. The feed gas, in entering the bed from the top and the upper zone, becomes saturated first, where an equilibrium between the water partial pressure in the gas and the water adsorbed on the desiccant will be established and no additional adsorption occurs. This is why this zone is called the equilibrium zone (EZ). With the adsorption going on, the EZ will grow, and more and more water will be adsorbed. The depth of bed from saturation to initial adsorption is known as the mass transfer zone (MTZ). This is simply a zone or section of the bed where a component is transferring its mass from the gas stream to the surface of the desiccant. In the mass transfer zone, the water content of the gas is reduced from saturation to less than 1 ppm (GPSA, 2004). As the flow of gas continues, the mass transfer zone moves downward through the bed, and water displaces the previously adsorbed gases until finally the entire bed is saturated with water vapor. When the leading edge of the MTZ reaches the end of the bed, breakthrough occurs. In principle, beds can be run until the first sign of breakthrough. This practice maximizes cycle time, which extends bed life because temperature cycling is a major source of bed degeneration, and minimizes regeneration costs. However, most plants operate on a set time cycle to ensure no adsorbate breakthrough.

The optimization of desiccant bed size is a key factor for the overall cost of a project, as it allows Capex and Opex savings. For a given performance, having shorter beds mean smaller vessels and lower utilities requirements. However, adsorbent requirements must be sufficient to accommodate the equilibrium section as well as the mass transfer zone's length. The MTZ is usually assumed to form quickly in the adsorption bed and to have a constant length as it moves through the bed, unless particle size or shape is changed. The length of the MTZ is usually 0.5–6 feet, and the gas is in the zone for 0.5–2 seconds (Trent, 2004). To maximize bed capacity, the MTZ needs to be as small as possible because the zone nominally holds only 50% of the adsorbate held by a comparable length of adsorbent at equilibrium. Both tall, slender beds, which reduce the percentage of the bed in the MTZ, and smaller particles in the MTZ that offer a larger specific area (CECA, 2009b), make more of the bed effective. However, smaller particle size, deeper beds, and increased gas velocity will increase pressure drop.

9.4.4 Operation of Solid Bed Dehydrator

Optimizing the performance of the solid bed dehydration unit requires a detailed understanding of its operation. The flow scheme of a typical solid adsorbents-based dehydration process is shown in Figure 9-6. The process is conducted alternately and periodically, with each bed going through successive steps of adsorption and desorption. During the adsorption step, the gas

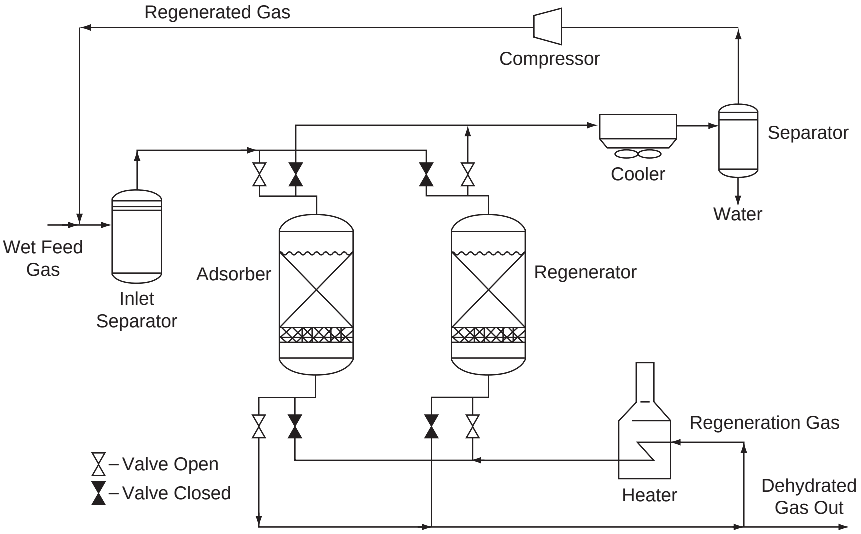

Figure 9-6 Typical flow scheme for a solid desiccant dehydration plant (GPSA, 2004).

to be processed is sent on the adsorbent bed, which selectively retains the water. When the bed is saturated, hot gas is sent to regenerate the adsorbent. After regeneration and before the adsorption step, the bed must be cooled. This is achieved by passing through cold gas. After heating, the same gas can be used for regeneration. In these conditions, four beds are needed in practice in cyclic operation to dry the gas on a continuous basis,[8] two beds operating simultaneously in adsorption or the gas drying cycle, one bed in the cooling cycle, and one bed in the regeneration cycle. In the simplest case (as shown in Figure 9-6), one bed operates in adsorption, while the second operates in desorption, and both beds are periodically switched (Rojey et al., 1997). For huge amounts of gas to be treated, several adsorbers can be parallel in adsorption and up to two in regeneration, with cooling and heating in series if regeneration gas flow rate should be reduced.

In the gas drying cycle, the wet inlet gas first passes through an inlet separator where free liquids, entrained mist, and solid particles are removed. This is a very important part of the system because free liquids can cause the adsorbent materials to break down. This leads to higher pressure drop and channeling, reducing the overall performance of the unit. If the adsorption unit is downstream from an amine unit, glycol unit, or compressors, a filter separator is required.

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[8] Adsorbents are normally unsuitable for continuous circulation, owing to mechanical problems and also due to the risks of attrition.

In addition to the use of an inlet separator to maximize water-droplet removal, a guard layer (equal to about 10%–20% of bed volume) of specialized, water-stable adsorbent can be added on top of the main adsorbent bed. This water stability can be imparted using tempering (i.e., heating to high temperatures over a long time); however, this process greatly reduces water adsorption capacity (Daiminger and Lind, 2004).

In the adsorption cycle, the wet inlet gas usually flows downward through the tower in order to prevent fluidization of the beds. The adsorbable components are adsorbed at rates dependent on their chemical nature, the size of their molecules, and the size of the pores in the solid material. The water molecules are adsorbed first in the top layers of the desiccant bed. Dry hydrocarbon gases are adsorbed throughout the bed. As the upper layers of desiccant become saturated with water, the water in the wet gas stream begins displacing the previously adsorbed hydrocarbons in the lower desiccant layers. Liquid hydrocarbons will also be absorbed and will fill pore spaces that would otherwise be available for water molecules. The dried gas then goes through a dust filter that will catch fines before the gas exits the unit.[9]

Obviously, the towers must be switched from the adsorption cycle to the regeneration cycle (heating and cooling) before the desiccant bed is completely saturated with water. When a tower is switched to the regeneration cycle, a slip stream of dry gas[10] (usually about 5%–10% of gas throughput) is heated to temperatures of 450°F–600°F[11] in the high-temperature heater and routed to the tower in countercurrent direction from bottom to top in order not to push the adsorbed water through the whole adsorbent bed, thus keeping the bottom section, the mass transfer zone, as dry as possible. As the temperature within the tower is increased, the water captured within the pores of the desiccant turns to steam and is absorbed by the natural gas. The hot, wet regeneration gas then goes through a cooler and inlet separator to remove the water before being recompressed and mixed with incoming wet feed. Once the bed has been "dried" in this manner, it is necessary to flow cool gas through the tower to return it to normal operating temperatures (about 100°F–120°F) before placing it back in service to dehydrate gas. The cooling gas could either be wet gas or gas that has already

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[9] If the gas goes on to a cryogenic section with plate fin heat exchangers, then dust can be collected in the exchangers and reduce heat transfer and dramatically increase pressure drop.

[10] Sales gas is sometimes used instead of a slip stream. The sales gas stream has the advantage of being free of heavier hydrocarbons that can cause coking.

[11] If COS formation is a problem, it can be mitigated by lowering regeneration temperatures to 400°F–450°F or lower.

been dehydrated. If wet gas is used, it must be dehydrated after being used as cooling gas, where a hot tower will not sufficiently dehydrate the gas.

The switching of the beds is controlled by a time controller that performs switching operations at specified times in the cycle by using opening and closing valves. The length of the different phases can vary considerably. Longer cycle times will require larger beds and additional capital investment but will increase the bed life. A typical two-bed cycle might have an 8-hour adsorption period with 6 hours of heating and 2 hours of cooling for regeneration. The 16-hour adsorption time for an adsorption unit with three beds, two beds in adsorption and one bed in regeneration, also makes a full cycle time of 24 hours, which gives a good 3-year guarantee.

9.4.5 Design Considerations

A few references (Campbell, 2000; GPSA, 2004; Kidnay and Parrish, 2006) provide some details and rules of thumb for designing the adsorber systems. However, this information serves only as the basis for performing preliminary design calculations based on a given cycle length, number of vessels and their configuration, and a given desiccant. Therefore, it is highly recommended to refer to desiccant vendors for designing solid-desiccant dehydration unit, as desiccant's useful capacity is highly dependent on its aging behavior. To take aging into account, experience is very important, and no published correlations exist as all desiccant vendors regard that as their own intellectual property and know-how. To use only literature data will result in either uneconomical units or potentially nonworking units.

9.4.6 Operational Problems

The following sections review a number of problems that can affect the proper operation of the solid bed dehydration unit and the procedures that can be used to avoid or deal with those problems (Manning and Thompson, 1991; GPSA, 2004; Meyer, 2010a,b).

9.4.6.1 Bed Contamination

The most frequent cause is incomplete removal of contaminants in the inlet gas separator. Also, if the regeneration gas leaving the separator is commingled with the feed gas to the dehydrators, then a separator malfunction can dump liquid hydrocarbons and water onto the adsorbent. Regeneration separators should usually be equipped with filtration levels similar to the inlet gas to prevent recontamination.

9.4.6.2 Wrong Bed Loading

In terms of bed loading, this is for gas dryers very often translated in a mixed bed of large and small particles, with the small particles being at the bottom of the bed. Conservative designs often involve only large particles. However, it is possible to decrease significantly the quantity of product by the use of smaller particles in the MTZ area. In the MTZ, contrary to EZ, the species have not reached the adsorption equilibrium due to the diffusion kinetics; they are not adsorbed in the whole volume of the product, but mostly on the particles' surface. Therefore, as smaller particles offer a larger specific area, the overall capacity of a given volume is higher. If desiccant beds are wrongly loaded with much fewer small particles, more desiccant mass is needed to meet the specs since the mass transfer zone for large particles is longer. A shorter adsorption time is the direct consequence.

9.4.6.3 Liquid Carryover

Liquid carryover in the solid bed has a negative impact on the drying process (i.e., poor gas flow distribution due to cake or fines formation in consequence of chemical attack, causing an increase in the pressure drop and a decrease in the adsorption time). Therefore, liquid carryover to solid beds should be prevented in general.

A major source of free water in the desiccant bed is a potential water carryover from the upstream wet cold feed gas separator. Water carryover could occur due to several causes, such as the mechanical design of the separator itself, which may shorten the residence time required for separating the water in the vessel. Process upsets due to malfunctioning in the separator-level controllers is another reason for water carryover to the downstream gas dehydrators. In order to reduce the liquid carryover in the adsorbers, one must modify the separators to improve their efficiency. A protective layer of activated alumina may be added on top of the solid bed to minimize the effect of free water on the desiccant. It is also important to avoid the formation of water at the top of the adsorbent bed through condensation of the desorbed water vapor from the bottom section during the regeneration step. Hence, the regeneration procedure should be changed so as to have a moderate temperature increase to avoid water recondensation.

Excessive water content in the wet feed gas can cause operating trouble. It is therefore very important to respect the inlet feed temperature of the adsorbers in case of saturated gas, where small variations in temperature will lead to significant increases in the water content.

9.4.6.4 Incomplete Desiccant Regeneration

Insufficient or incomplete regeneration of the solid beds will lead to a sudden loss in adsorption capacity and a significantly premature breakthrough. Insufficient desiccant regeneration is a result of one or all of the following main factors:

- Very low regeneration gas flow rate
- Insufficient regeneration time available
- Change of regeneration gas
- Very low regeneration gas temperature

For the heating phase, a certain quantity of heat at a minimum temperature level should be brought in the adsorber in order to desorb the adsorbed products well. If the regeneration gas flow quantity (flow and/or duration) is too small, there will be accumulation on the top layers of the desiccants because of the lack of desorption energy. In the case of laminar flow or channeling, the gas does not pass through the entire bed but only a part. Local accumulation may happen a little bit everywhere in the bed. While channeling can be detected by calculation or low pressure drop through the adsorber, the too-low flow rate or insufficient heating time can be seen by analyzing the temperature profiles during the adsorption. For the cooling, laminar flow (channeling) can be accepted. If the flow or the duration is too short, the result will be a temperature peak of the gas in adsorption when switching from regeneration to adsorption. Depending on the upstream equipment, this may cause eventual alarms or shutdown due to high temperature.

Even if a significant composition change is unusual in an existing plant, the temperature curves may give an indication for its influence (e.g., incomplete regeneration). As soon as the character of the gas (e.g., hydrocarbons only) does not change (e.g., more ethane instead of methane), it is important to keep the mass flow constant and not the volume flow.

To be sure to regenerate the adsorbents well, one should analyze the inlet and outlet temperatures of the adsorber in regeneration. At the end of the heating step, the outlet temperature should be almost constant during a certain time, 30 minutes to 2 hours, depending on the design of the adsorber; and the temperature difference between the inlet and outlet should not be more than 59°F–68°F, depending on the quality of the heat insulation.

9.4.6.5 Unequal Flow Distribution

Sometimes a unit may face premature breakthrough on only one vessel, and this is true particularly if working with several vessels in adsorption in parallel. If there is a problem on one vessel only, it is, in general, an issue of bed configuration and not of the molecular sieves. Unequal flow can be detected by observing the pressure drop through the vessels in adsorption.

9.4.6.6 Bottom Support

Experience shows that more and more problems of solid bed failure are caused due to leakage of product through the support grid. Nevertheless, very often the reason is not the support grid itself but the way it was installed. The important point here is the good mechanical design of the support bed, putting three wire meshes on the support grid and installing the correct quantity and size of ceramic balls.

9.4.7 Maximizing Molecular Sieve Unit Performance

The use of molecular sieve adsorbents has become widespread for natural gas dehydration and desulfurization. However, with increasing treatment requirements, greater plant size, and increasingly challenged feedstocks, there is considerable incentive to improve molecular sieve performance. Simultaneously, decades-long experience from operating plants has begun to reveal limitations in the conventional use of these materials, where molecular sieve manufacturers are developing effective strategies for enhancing the use of these materials and maximizing their potential.

Bruijn et al. (2002) presented state of the art of water adsorption on molecular sieves, starting with the basic understanding of molecular sieves, and discussed the know-how of Shell Global solutions regarding the proper design and operations of molecular sieve units that can sustain or improve gas processing plant competitiveness and profitability.

Northrop and Sundaram (2009) addressed key limitations, which are recognized in conventional molecular sieve operating practice, using enhanced adsorbent combinations, modified thermal and/or pressure cycles, and modified partial desorption regeneration, or combinations thereof. They also explored more selective adsorbents to reduce hydrocarbon loss and regeneration energy requirements.

Rastelli and Shadden (2009) described a computer model of the adsorption and regeneration process developed by UOP, which can help processors understand the severity of conditions in molecular sieve units. The model, which is adaptable to different regeneration schemes, has improved design capabilities and can be used to optimize the adsorbent configuration for maximum performance and life.

Meyer (2010a) described CECA's good design practices for molecular sieve beds, with details and design rules of thumb for pressure drop, adsorption capacity, and regeneration procedure. Meyer (2010a,b) also discussed several ways of debottlenecking molecular sieve units by using real-case modifications of existing plants.

Note should be made that molecular sieve units are reliable units but, in a certain way, are black boxes for the operators. Regarding lack of performance, the problem is not systematically the failure of the molecular sieves, and urgent replacement is not necessary. When the plant data, such as pressure drop and regeneration temperature curves, are analyzed well, very often the lower performance can be explained, and the unit stabilized to prepare a secure replacement. Nevertheless, it is recommended to have experienced specialists analyze the plant data (Meyer, 2010b).

9.5 PROCESS SELECTION

A number of factors should be considered in the evaluation of a dehydration process or combination of processes. If dehydration is required only to avoid free water formation or hydrate formation or to meet the pipeline specification of 4 to 7 lb/MMscf, any of the aforementioned processes may be viable. Traditionally, TEG dehydration has been the process of choice.

Solid desiccant dehydrators are typically more effective than glycol dehydrators because they can dry a gas to less than 0.1 ppmv (0.05 lb/MMscf). However, in order to reduce the size of the solid desiccant dehydrator, a glycol dehydration unit is often used for bulk water removal. The glycol unit would reduce the water content to around 60 ppmv, and this would help to reduce the mass of solid desiccant necessary for final drying. Using desiccant dehydrators as alternatives to glycol dehydrators can yield significant economic and environmental benefits, including reduced capital cost, reduced operation and maintenance cost, and minimal VOC and hazardous air pollutants (BTEX). A detailed discussion on determining their economics and environmental benefits can be found in EPA430-B-03-016 (2003).

Selection of the drying process must consider mercaptans removal. If drying and mercaptans removal are necessary, another step or processing unit for the removal of mercaptans is often necessary. The only process capable of simultaneously drying and removing mercaptans is molecular sieves. The molecular sieves are able to concentrate the water and the mercaptans in the relatively small quantity of regeneration gas used to regenerate the sieves. All the impurities will be desorbed from the sieves in peaks, the mercaptans first and then the water. Unless the water can be isolated by cooling the regenerated gas, it is not easy to concentrate the mercaptans in a stream to be able to treat them. The hydrocarbons possibly coabsorbed in the physical absorption must be separated prior to sending the mercaptans to a sulfur recovery unit (SRU) unless the full regeneration gas quantity is burned for heat recovery or flared. This is often not possible because the regeneration flow is too high for the fuel gas balance and the SO_2 emissions limits are more and more stringent (Mokhatab and Meyer, 2009).

Silica gel allows the treated gas to meet the dew point pipeline specs for water and hydrocarbons by requiring less regeneration gas compared to molecular sieves due to the lower adsorption enthalpy of the contaminants on the silica gel. The adsorbed products are more loosely bound to the adsorbent. Untreated gas is used for the heating and cooling and recycled to the inlet of the unit after partially condensing the water and the heavy hydrocarbons. The adsorption capacity is typically exhausted in the range of dozen of minutes up to 2–3 hours, which increases the number of absorption cycles and reduces the life of the silica gel. Mercaptans can be partially adsorbed (Kane et al., 2004).

When considering susceptibility to inlet feed contamination, one should keep in mind that replacing a solvent is much easier and cheaper than changing out an adsorbent bed. However, prevention of contamination by use of properly designed inlet scrubbers and coalescing filters, if required, is the best solution.

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CHAPTER 10

Natural Gas Liquids Recovery

10.1 INTRODUCTION

Most natural gas is processed to remove the heavier hydrocarbon liquids from the natural gas stream. These heavier hydrocarbon liquids, commonly referred to as natural gas liquids (NGLs), include ethane, propane, butanes, and natural gasoline (condensate). Recovery of NGL components in gas not only may be required for hydrocarbon dew point control in a natural gas stream (to avoid the unsafe formation of a liquid phase during transport), but also yields a source of revenue, as NGLs normally have significantly greater value as separate marketable products than as part of the natural gas stream. Lighter NGL fractions, such as ethane, propane, and butanes, can be sold as fuel or feedstock to refineries and petrochemical plants, while the heavier portion can be used as gasoline-blending stock. The price difference between selling NGL as a liquid and as fuel, commonly referred to as the "shrinkage value," often dictates the recovery level desired by the gas processors. Regardless of the economic incentive, however, gas usually must be processed to meet the specification for safe delivery and combustion. Hence, NGL recovery profitability is not the only factor in determining the degree of NGL extraction. The removal of natural gas liquids usually takes place in a relatively centralized processing plant, where the recovered NGLs are then treated to meet commercial specifications before moving into the NGL transportation infrastructure.

This chapter presents the basic processes used to separate natural gas liquids from the gas, fractionating them into their various components, and briefly describes the processing required to produce marketable liquid products.

10.2 NGL RECOVERY PROCESSES

Figure 10-1 shows the phase behavior of a natural gas as a function of pressure and temperature. Obviously, any cooling outside the retrograde condensation zone will induce condensation and yield NGL. The retrograde condensation phenomenon has important applications in NGL production.

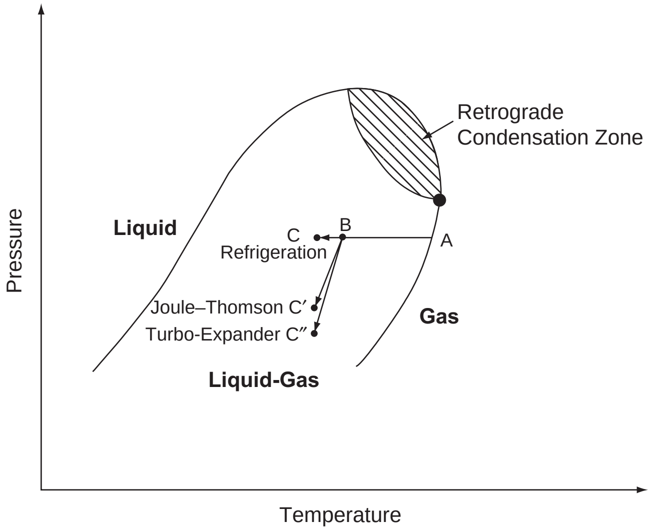

Figure 10-1 Thermodynamic pathways of different NGL recovery technologies.

Some plants operate at inlet pressures above the critical point and thus revaporize NGLs when the temperature drops below the retrograde temperature. It's therefore important to know where we are on the phase envelope.

The basic NGL recovery processes are related to Figure 10-1, where possible. The following sections provide a review of the commercially available technology options that are available for the NGL extraction, while also looking at the future trends in technology development.

10.2.1 Refrigeration Processes

Refrigeration processes are used at many different temperature levels to condense or cool gases, vapor, or liquids. There are several refrigeration processes for NGL recovery; these processes are discussed in more detail next.

10.2.1.1 Mechanical Refrigeration

Mechanical refrigeration is a simple process used widely in gas conditioning applications (e.g., hydrocarbon dew point controlling), but it is also used in NGL recovery applications as the primary refrigeration option or in conjunction with another refrigeration option. Mechanical or external refrigeration, as shown in Figure 10-2, is supplied by a vapor-compression refrigeration cycle that usually uses propane as the refrigerant and reciprocating or centrifugal types of compressors to move the refrigerants from the low- to high-pressure operating conditions. As shown in Figure 10-2,

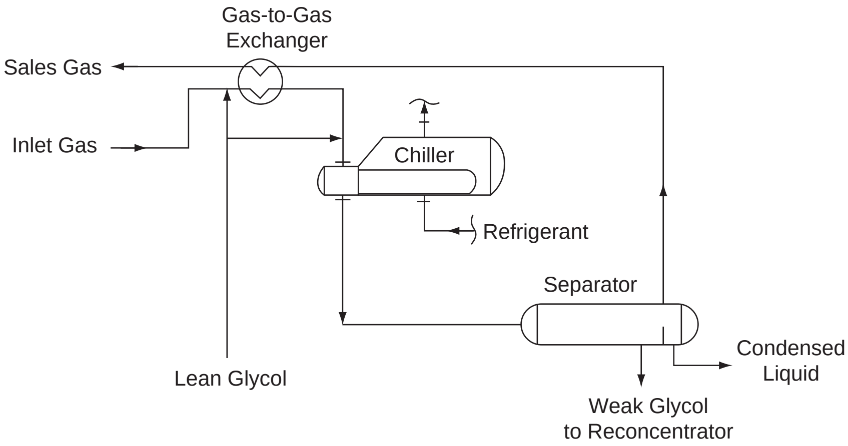

Figure 10-2 Flow sheet of a mechanical refrigeration process (Geist, 1985).

the gas-to-gas heat exchanger recovers additional refrigeration by passing the gas leaving the cold separator countercurrent to the warm inlet gas. The temperature of the cold gas stream leaving this exchanger "approaches" that of the warm inlet gas. Economically, this approach can be as close as 5°F. The chiller in Figure 10-2 is typically a shell and tube, kettle-type unit. The process gas flows inside the tubes and gives up its energy to the liquid refrigerant surrounding the tubes. The refrigerant (often propane) boils off and leaves the chiller vapor space essentially as a saturated vapor.

The thermodynamic path followed by the gas in an external refrigeration process is shown as line ABC in Figure 10-1. From A to B indicates gas-to-gas heat exchange; from B to C, chilling.

When water is present in a natural gas processed by refrigeration, hydrate formation is prevented either by dehydration of the gas or by injection of a hydrate inhibitor. If the processing temperature is relatively low, the gas is usually dehydrated before the refrigeration step. If not, the injection of an inhibitor (usually methanol or glycols) upstream of the gas-to-gas heat exchanger is often the simplest and most economical solution (Rojey et al., 1997).

In this case, ethylene glycol is injected at the inlet of the gas-to-gas exchanger and/or chiller to prevent hydrate formation, or freeze-up, in these exchangers with the latter being more common. Freeze-up will partially block exchanger tubes, thus increasing pressure drop and decreasing heat exchange. The weak glycol solution, containing absorbed water, is separated in the cold separator, reconcentrated, and recycled.

A mechanical refrigeration process is adopted when sizeable amounts of condensate are expected. This process may also lead to the recovery of liquefied petroleum gas (LPG),[1] where for LPG recovery up to 90%, a simple propane refrigeration system provides refrigeration at temperatures to −40°F (Lee et al., 1999). There are many straight refrigeration process schemes, which vary according to all the design variables such as gas composition, process pressure, and LPG recovery objectives. A detailed discussion can be found in Russell (2001).

10.2.1.1.1 Choice of Refrigerant

Any material could be used as a refrigerant. The ideal refrigerant is nontoxic, noncorrosive, has PVT and physical properties compatible with the system needs, and has a high latent heat of vaporization. The practical choice reduces to one, which has desirable physical properties, and will vaporize and condense at reasonable pressures, at the temperature levels desired. In general, the lower practical limit of any refrigerant is its atmospheric pressure boiling point. It is desirable to carry some positive pressure on the chiller to obtain better efficiency in the compressor, reduce equipment size, and avoid air induction into the system. Propane is by far the most popular refrigerant in the gas processing applications. It is readily available (often manufactured on-site), is inexpensive, and has a "good" vapor pressure curve. It is flammable, but this is not a significant problem if proper consideration is given to the design and operation of the facility (Campbell, 2000).

10.2.1.1.2 Cascade Refrigeration

Cascade refrigeration refers to two refrigeration circuits thermally connected by a cascade condenser, which is the condenser of the low-temperature circuit and the evaporator of the high-temperature circuit. A cascade system utilizes one refrigerant to condense the other primary refrigerant, which is operating at the desired evaporator temperature. This approach is usually used for temperature levels below −90°F, when light hydrocarbon gases or other low boiling gases and vapors are being cooled (Lee et al., 1999). To obtain the highest overall efficiency for the system, the refrigerants for the two superimposed systems are different. Cascade refrigeration systems are not common in gas processing. Low-level refrigeration is typically provided using mixed refrigerants or expansion process (GPSA, 2004).

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[1] Liquefied petroleum gas (LPG) is the general expression for propane, butane, and a mixture of the two, which are produced from two distinct sources.

10.2.1.1.3 Mixed Refrigerants

An alternative to cascade refrigeration is to use a mixed refrigerant. A mixed refrigerant is a mixture of two or more components. The light components lower the evaporation temperature, and the heavier components allow condensation at ambient temperature. The evaporation process takes place over a temperature range rather than at a constant temperature as with pure component refrigerants. The mixed refrigerant is blended so that its evaporation curve matches the cooling curve for the process fluid. Heat transfer occurs in a countercurrent exchanger, probably aluminum plate-fin, rather than a kettle-type chiller. Mixed refrigerants have the advantage of better thermal efficiency because refrigeration is always being provided at the warmest possible temperature (Mackenzie and Donnelly, 1985). The amount of equipment is also reduced to a cascade system. Disadvantages include a more complex design and the tendency for the heavier components to concentrate in the chiller unless the refrigerant is totally vaporized (Campbell, 2000).

10.2.1.2 Self-Refrigeration

As opposed to external refrigeration, in the self-refrigeration process (see Figure 10-3), the inlet gas is precooled against the treated gas (through the gas-to-gas exchanger) and subsequently further cooled by isenthalpic

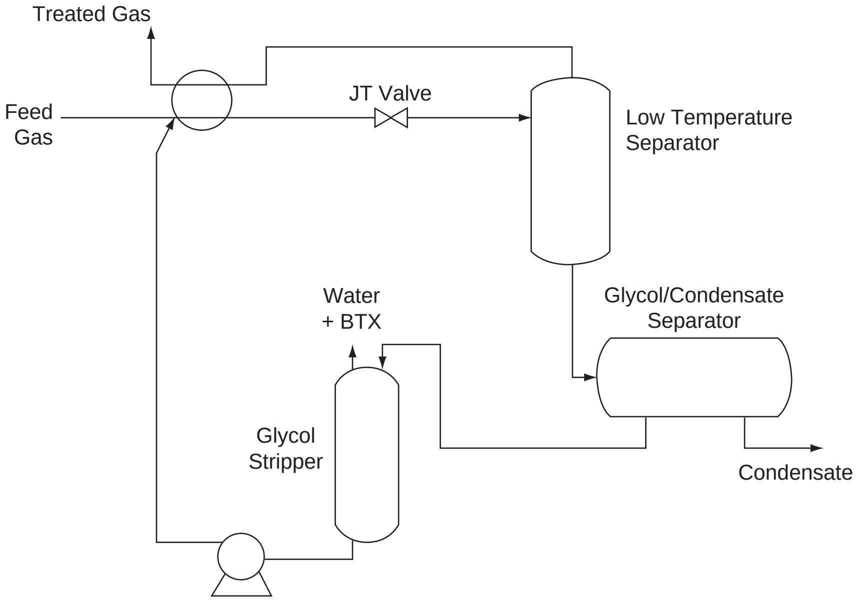

Figure 10-3 Flow sheet of a self-refrigeration system (Brands and Rajani, 2001).

expansion (i.e., Joule–Thomson expansion) through a valve resulting in heavy hydrocarbons and water to condense.[2] Another term applied to the process is low temperature separation (LTS). In this process, nonideal behavior of the inlet gas causes the gas temperature to fall with the pressure reduction, as shown by line ABC′ in Figure 10-1. The temperature change depends primarily on the pressure drop. The condensed liquids are then removed in one or more separators to meet the vapor pressure and composition specifications. The gas from the low temperature separation (LTS), now satisfying the sales gas specification, is reheated against the incoming feed. Generally, treated gas must be recompressed to sales pipeline pressure if it has been expanded to a lower pressure.

Note that the extent of cooling in this process is often limited by the hydrate formation temperature at the prevailing pressure, unless a hydrate inhibitor such as methanol or glycol is injected upstream of the gas-to-gas heat exchanger. In this case, the wet glycol and hydrocarbon condensate are heated and separated in a three-phase separator. Glycol can be regenerated in a stripper. Since glycol shows some affinity for hydrocarbons, the off-gas from the glycol stripper often contains BTX (Brands and Rajani, 2001).

This process is usually used in hydrocarbon dew point controlling applications, but is also used in NGL recovery applications. If the objective is to recover ethane or more propane than obtainable by mechanical refrigeration, a good process can be self-refrigeration, which is particularly applicable for smaller gas volumes of 5–10 MMcfd (Russell, 2001). The advantages of this process are simplicity and low cost. The primary drawback is the pressure drop that occurs across the valve, which often ranges from 500 to 1,000 psia. This process may be attractive when the plant inlet pressure is sufficiently high to eliminate the need for compression.

10.2.1.3 Cryogenic Refrigeration

Cryogenic refrigeration processes are commonly used for deep NGL recovery, with minimum temperatures below −150°F achieved routinely. In the cryogenic or turbo-expander plant, the chiller or Joule–Thomson (JT) valve used in the two previous processes is replaced by an expansion turbine. As the entering gas expands, it supplies work to the turbine shaft, thus reducing the gas enthalpy. This decrease in enthalpy causes a much larger temperature drop than that found in the simple JT (constant enthalpy) process. The

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[2] Due to this chilling, the self-refrigeration process achieves high ethane recoveries; typical ethane recoveries are about 70% (Mehra, 2004).

expansion process is indicated as line ABC″ in Figure 10-1. The turbine can be connected to a compressor that recompresses the gas with only a small loss in overall pressure. This results in a higher treated gas pressure, which can be increased to the pipeline specification by a second compression step.

Although there are variations in the design of expander plants, most expander plants have the same basic process flow as shown in Figure 10-4. The inlet gas is first cooled in the high-temperature, gas-to-gas heat exchanger and second in the propane chiller. The partially condensed feed gas is sent to a separator. The liquid from the separator is fed to the demethanizer, and the gas is cooled further in the low-temperature gas-to-gas exchanger and fed into a second cold separator. Gas from the cold separator expands through the expansion turbine to the demethanizer pressure, which varies from 100 to 450 psia. The turboexpander simultaneously produces cooling/condensing of the gas and useful work, which may be used to recompress the sales gas. Typically, 10%–15% of the feed gas is condensed in the cold separator, which is usually at −30°F to −60°F. The expander lowers the pressure from the inlet gas value (600–900 psia) to the demethanizer pressure of 100–450 psia. Typical inlet-gas temperatures to the demethanizer are −130°F to −150°F, sufficiently low that a great deal of the ethane is liquefied. The demethanizer is a low-temperature distillation column that makes a separation between methane and ethane. Methane and components lighter than methane, such as nitrogen, are the principal products in the vapor near the top of the column; whereas the ethane and heavier components, such as propane, butanes, and heavier hydrocarbons, comprise the principal components in the bottom product of the column. The molar ratio of methane to ethane in the bottom product is typically 0.01 to 0.03. Since the outlet of the expander is usually two-phase flow, the liquid produced in the expander serves as reflux for the demethanizer (Elliot et al., 1996). The bottom product

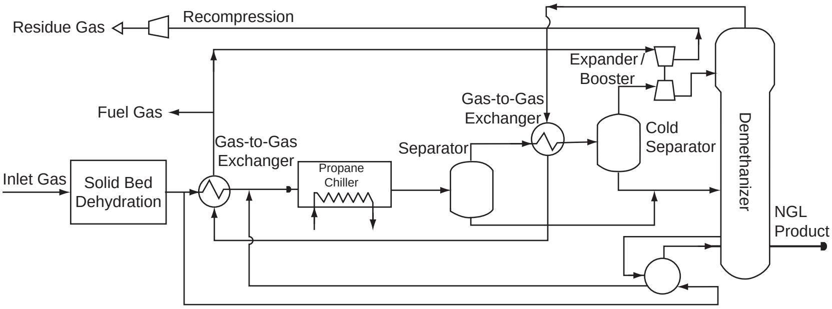

Figure 10-4 Typical flow sheet of a cryogenic refrigeration plant (Ewan et al., 1975).

from the demethanizer can be further fractionated to produce pure product streams of ethane, propane, butanes, and natural gasoline. The bottom product temperature is often below ambient, so feed gas may be used as the heat-transfer medium for the reboiler. This provides additional refrigeration to the feed and yields higher ethane recovery, typically 80% (Holm, 1986). The top product from the demethanizer, after heat exchange with the inlet gas, is recompressed to pipeline pressure and delivered as the sales gas.

Cryogenic processes can be applied only if the gas pressure after expansion is sufficiently high for condensation of the heavier components to take place. However, if the gas arrives at low pressure (say, less than 725 psi), an external mechanical refrigeration must be used to cool it to the specified temperature (Cranmore and Stanton, 2000). In addition, if the NGL content of the feed gas is relatively low (less than 2.5–3 gallons per Mcf, GPM),[3] self-refrigeration (gas-to-gas heat exchanger) usually suffices. However, for moderately rich feeds (> 3 GPM), mechanical refrigeration should be considered to obtain high ethane recovery most economically (Ewan et al., 1975).

Cryogenic refrigeration processes are also used for hydrocarbon dew point controlling applications in which the required pressure drop across the expander is substantially less than that required across a JT valve, resulting in lower compression costs over the life of the field. Note that, in this application, similar to the self-refrigeration process, upstream dehydration of the gas is required to prevent hydrate formation. Solid bed dehydration upstream of the plant is a proven technique for preventing hydrate formation. Sometimes, small quantities of methanol or glycol can also be injected upstream of the expander.

The cryogenic refrigeration process is generally the most technically advanced type of NGL recovery used today. This combines high recovery levels (typically allowing full recovery of all of the propane and heavier NGLs and recovery of 50% to more than 90% of the ethane) with low capital cost and easy operation (Lee et al., 1999). This is less attractive on very rich gas streams or where the light NGL product (C_2 and C_3) are not marketable, whereas for gases very rich in NGL, simple refrigeration is probably the best choice.

Within the liquids recovery section of the gas processing plant, there are both operating cost and operating flexibility issues that directly impact the processing cost. While it is easily recognized that the efficiency of the selected liquids recovery process is an important factor in the processing cost,

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[3] GPM is an indication of the potential recoverable NGLs contained in the gas stream. The NGL content of produced gas can vary from 0.2 GPM for very dry gas to 20 GPM for very rich gas.

the flexibility of operating the process to either recover or reject ethane without sacrificing efficiency or propane recovery is often the critical factor (Pitman et al., 1998). As the industry has matured and the demand for more efficient ethane recovery has increased, several new and very clever designs have been developed (Gahier et al., 2010). Up to now, Ortloff's gas subcooled process (GSP) and residue split-vapor (RSV) process have been the state of the art for efficient NGL/LPG recovery from natural gas, particularly for those gases containing significant concentrations of carbon dioxide,[4] which are discussed next.

10.2.1.3.1 Ortloff's Gas Subcooled Process

Increasing ethane recovery beyond the 80% achievable with the conventional design requires that a source of reflux must be developed for the demethanizer. Ortloff's Gas Subcooled Process (GSP) was developed to overcome this problem and others encountered with the conventional expander scheme. GSP configuration allows recovering ethane from around 50% to 99%, with propane recovery at 99%. In this process, as shown in Figure 10-5, a portion of the gas from the cold separator is sent to a heat exchanger, where it is totally condensed and subcooled with the overhead stream. This stream is then flashed to the top of the demethanizer, providing reflux to the demethanizer. The expander feed is sent to the tower several stages below the top of the column. Because of this modification, the cold separator operates at much warmer conditions well away from the system critical. Additionally, the residue recompression is less than with the conventional expander process.

The GSP design has several modifications. One is to take a portion of the liquid from the cold separator along with the gas to the overhead exchanger. Generally, this can help to further reduce the horsepower required for recompression. Also, the process can be designed to just use a portion of the cold separator liquid for reflux. This modification is typically used for gases richer than 3 GPM (C_2^+).[5] The GSP design is very CO_2 tolerant; many designs require no upfront CO_2 removal to achieve high recovery. CO_2 levels are very composition and operating pressure dependent, but levels up to 2% can usually be tolerated with the GSP design. When

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[4] Many NGL recovery processes require removal of the CO_2 to avoid solids formation (freezing) in the cold sections of the processing plant. Since CO_2 removal unit can add significantly to both the investment and operating costs of the pretreatment section of the gas-processing plant, there is considerable advantage to use a CO_2-tolerant NGL recovery process (Pitman et al., 1998).

[5] C_2^+ or ethane plus refers to an NGL stream of heavier hydrocarbons of ethane, propane, butane, and pentanes.

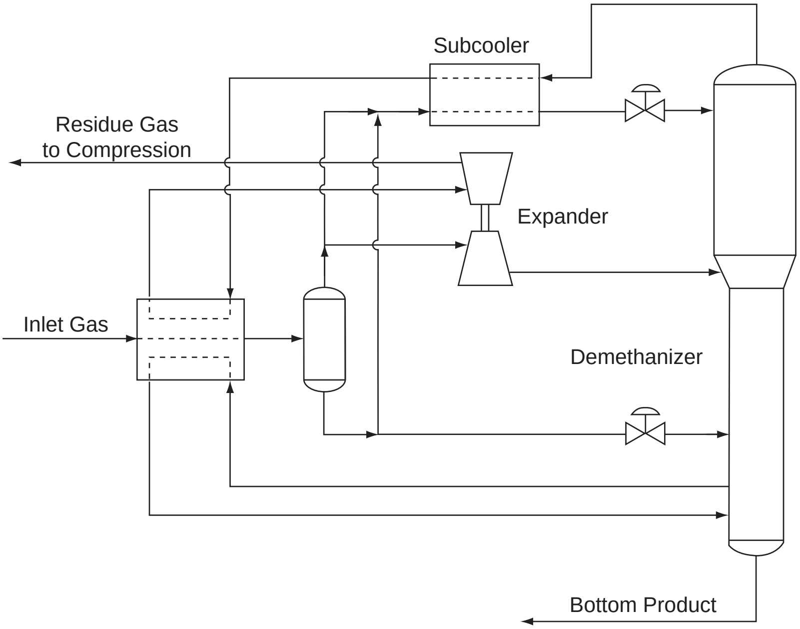

Figure 10-5 Schematic of Ortloff's gas subcooled process (Pitman et al., 1998).

CO_2 is present in the feed gas, the higher concentrations of C_2^+ components in the cold liquids help reduce the amount of CO_2 concentrating in the upper, colder sections of the tower, allowing higher ethane recovery levels without CO_2 freezing. This same process can be operated to reject ethane, but propane recovery efficiency suffers significantly when operated in this mode due mainly to the higher concentration of propane present in the top feed (Pitman et al., 1998).

Since the early 1980s, the GSP has been the pioneer process for high ethane recovery NGL plants, where all of the current "state-of-the-art" NGL recovery technologies are basically enhancements of the GSP concept (Nasir et al., 2003).

10.2.1.3.2 Ortloff's Residue Split-Vapor Process

Another method of producing reflux is to recycle a portion of the residue gas, after recompression, back to the top of the column using the residue split-vapor (RSV) process. As shown in Figure 10-6, the process flow is similar to the GSP design except that a portion of the residue gas is brought back through the inlet heat exchanger. At this point, the stream is totally condensed and is at the residue gas pipeline pressure. The stream is then flashed

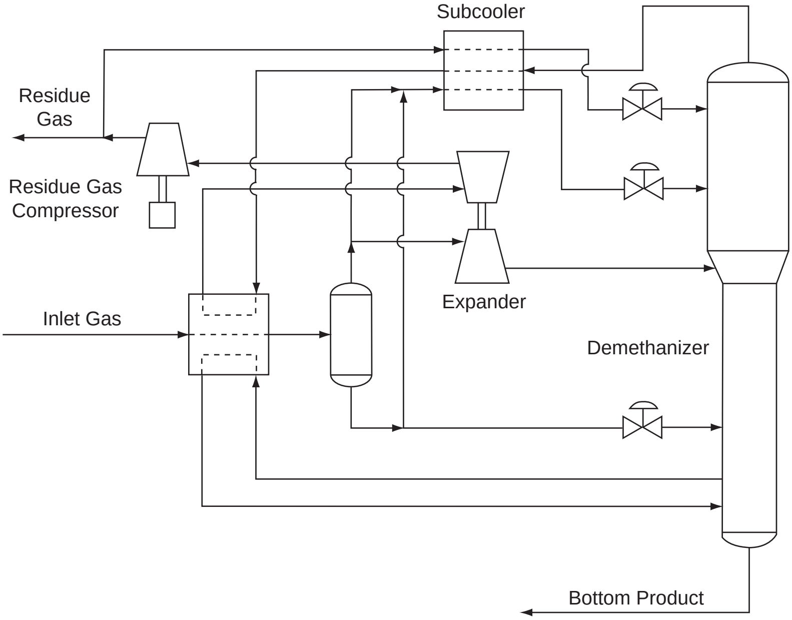

Figure 10-6 Schematic of Ortloff's residue split-vapor process (Pitman et al., 1998).

to the top of the demethanizer to provide reflux. The subcooled inlet gas split and the expander outlet stream are sent lower down in the tower rather than to the top of the column. The reflux provides more refrigeration to the system and allows very high ethane recovery to be realized. The recovery level is a function of the quantity of recycle in the design.

The RSV process has been used successfully in numerous facilities. It is CO_2 tolerant, and the recovery can be adjusted by the quantity of recycle used. The RSV process can be used for very high ethane recoveries limited only by the quantity of horsepower provided.

10.2.2 Lean Oil Absorption

The absorption method of NGLs recovery is very similar to using absorption for gas dehydration. The main difference is that, in NGL absorption, an absorbing oil is used as opposed to glycol.[6] This absorbing oil has an "affinity" for NGLs in much the same manner as glycol has an affinity for water.

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[6] Other absorptive processes using different absorbents can be used. One example is the use of methanol in the IFPEXOL process, which provides not only NGL recovery but also simultaneous acid gas removal and dehydration.

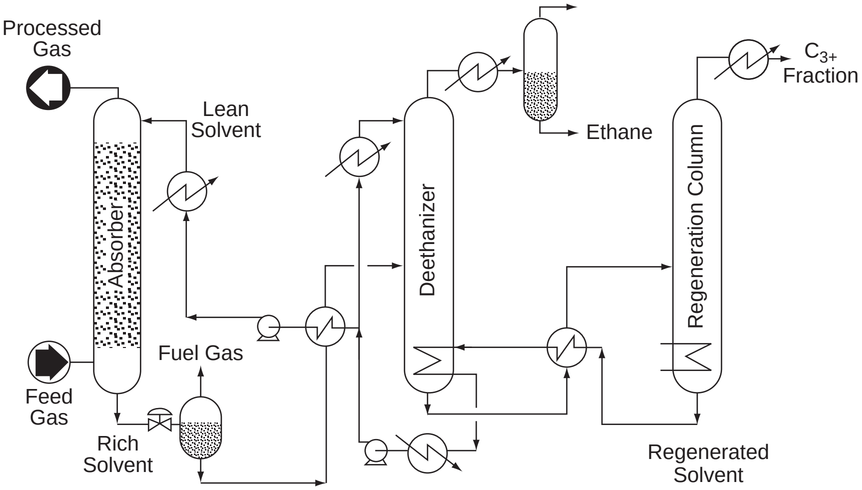

Figure 10-7 Simplified flow diagram of an oil absorption plant (Rojey et al., 1997).

Lean oil absorption is the oldest and least efficient process to recover NGLs.[7] In this process, as shown in Figure 10-7, the gas to be processed is contacted in a packed or tray absorption column (typically operated at the ambient temperature and a pressure close to the sales gas pressure) with an absorption oil (lean oil),[8] which absorbs preferentially the most heavy hydrocarbons ($C_3-C_7^+$) from natural gas. The gas leaves the top of the absorber while the absorber oil, now rich in heavy hydrocarbons from the gas, leaves the bottom of the absorber and is expanded to liberate most of the absorbed methane. Rich oil is then sent to a deethanizer to reject all the methane and part of the ethane absorbed. At the top of this column, additional cold oil is injected to limit the desorption of ethane and to prevent the desorption of C_3^+. Rich oil then flows to a regeneration column, where it is heated to a high enough temperature to drive the propanes, butanes, pentanes, and other natural gas liquid components to the overhead, and the regenerated solvent (lean oil) is recycled.

Note that the oil absorption plant cannot recover ethane and propane effectively when it requires circulating large amounts of absorption oil, demands attendant maintenance, and consumes too much fuel. However, an oil absorption plant can be modified to improve its propane recovery

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[7] Lean oil absorption process was the predominant design prior to the development of turboexpander processes in the 1970s.

[8] Since reducing the molecular weight of lean oil enhances the lighter component absorption, 100–110 molecular weight lean oils are generally used in this process (Mehra, 2004).

by adding a propane refrigeration cycle for cooling. The refrigerated lean oil absorption process improves the recovery of propane to the 90% level, and depending on the gas composition, up to 40% of ethane may be recovered (Elliot, 1996). Nowadays, Advanced Extraction Technologies (AET) has licensed a new process that allows recovering up to 96% of ethane and 99% of propane. However, for reaching such a high rate of recovery, high circulation rates are required that also induce an increased cost (Mehra, 2004).

Lean oil absorption plants are not as popular as they once were and are rarely, if ever, constructed anymore.[9] They are expensive and more complex to operate, and it is difficult to predict their efficiency at removing liquids from the gas as the lean oil deteriorates with time (Arnold and Stewart, 1999). Lean oil processes have traditionally lost out since the development of turboexpander processes but should still be evaluated early on in the project as they may provide the optimal solution in certain applications such as rich low-pressure feedstocks and low NGL recovery targets (McMahon, 2004).

10.2.3 Solid Bed Adsorption

The solid bed adsorption method uses adsorbents that have the capability to adsorb heavy hydrocarbons from natural gas. The adsorbent may be silica gel or activated charcoal. Activated alumina cannot be used in the presence of heavy hydrocarbons, which foul the adsorbent (Rojey et al., 1997). Note that the design of an adsorbent-based system for heavy hydrocarbon removal is more complicated than that of a system for the removal of water only. For instance, different grades of adsorbent may be required, and the system must also be designed to accommodate the adsorption of more than one component (Daiminger and Lind, 2004).

The adsorption process required for removing heavy hydrocarbons is schematically shown in Figure 10-8. The process is continuous with respect to the gas but cyclical with respect to the adsorbent bed because the latter must be regenerated when it becomes saturated with condensate. Regeneration is accomplished by passing heated recycle gas through the bed. The condensate is recovered from the regeneration gas by cooling, condensation, and phase separation. To recover a large fraction of the hydrocarbons, while limiting the volume of adsorbent, it is preferable to use a relatively short cycle time, about 1 hour. In practice, this cycle time may vary within a fairly

[9] A few lean oil plants are still used in the United States and Canada (Hubbard, 2009).

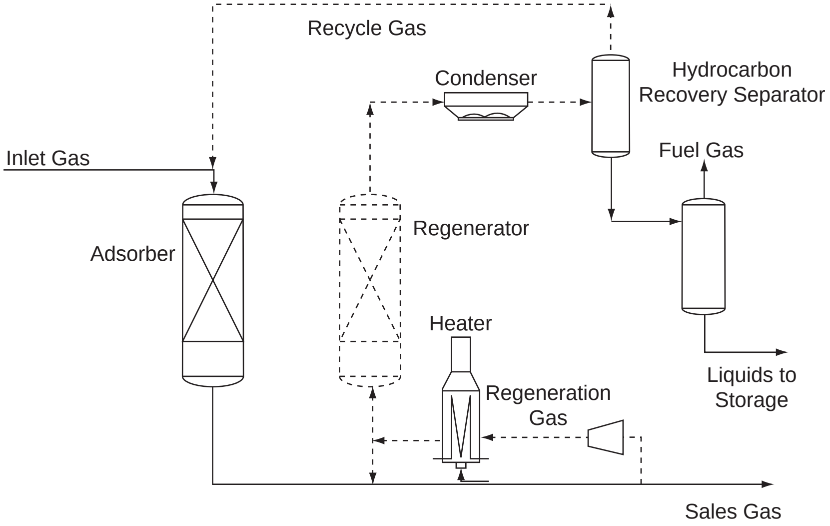

Figure 10-8 Schematic of a solid bed adsorption plant (Foglietta, 2004).

wide interval, between 20 minutes and several hours, depending on how hydrocarbon rich the gas is (Ballard, 1965).

This process is appropriate for relatively low concentrations of heavy hydrocarbons. It can be also appropriate if the gas is at a high pressure, close to the cricondenbar. In this case, refrigeration processes become ineffective and separation by adsorption may offer the only way to obtain the required specifications (Parsons and Templeman, 1990).

The adsorption processes are easy to start up and to operate at high turndown (changes in throughput), and so are useful for variable and "on-off" operation. However, the adsorption beds are heavy in weight and expensive. This process option is not often used but may be considered in special applications, such as hydrocarbon dew point control in remote locations.

Silica gel can achieve a simultaneous reduction of the hydrocarbon and the water dew point. Shell's SORDECO process is an example of this type of technology. The SORDECO process applies Sorbead (a special grade of silica gel) adsorbent beads packed into an adsorber column. It selectively removes water and hydrocarbons from natural gas. When the adsorbent is saturated, stripping with hot regeneration gas regenerates it. Condensing the stream when leaving the regenerating adsorber separates water and heavy hydrocarbons. More than 200 silica gel units are installed in natural gas applications under different conditions worldwide, on- and offshore. Usually, an adsorptive hydrocarbon recovery unit consists of three or more vessels, where adsorption and regeneration take place in parallel. The installation costs depend on the case itself, as they vary with the composition and the flow rate. They are often competitive to other technologies or redeem quickly due to long lifetimes and low operation costs. One of the most interesting points concerning adsorption of hydrocarbons and water with silica gel is the flexibility of these units in view of operating parameters and gas compositions. If the amount of water or hydrocarbons changes in the feed gas, the cycle time of the adsorption plant can be varied easily in order to achieve continuously low dew points (Schultz and Laukart, 1995; Brands and Rajani, 2001; Daiminger and Lind, 2004).

10.2.4 Membrane Separation Process

Refrigeration and cryogenic plants traditionally have been used for NGL recovery. These plants have high capital and operating costs. Moreover, they contain numerous rotating parts and are complicated to operate. The membrane separation process offers a simple and low-cost solution for removal and recovery of heavy hydrocarbons from natural gas. Figure 10-9 shows a schematic of a membrane separation process for NGL recovery (Foglietta, 2004). As can be seen, the separation process is based on a

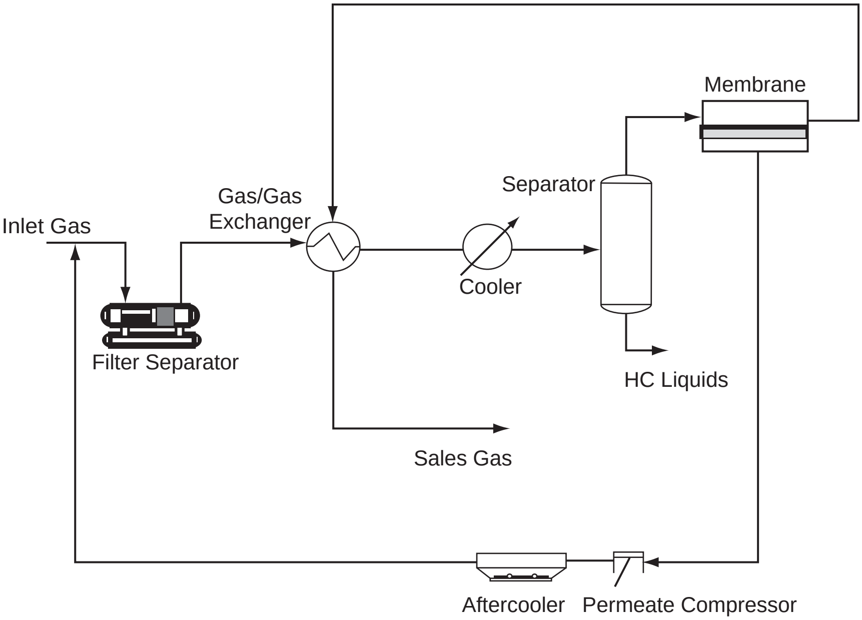

Figure 10-9 Schematic flow chart of the membrane separation process (Foglietta, 2004).

high-flux membrane that selectively permeates heavy hydrocarbons compared to methane. These hydrocarbons permeate the membrane and are recovered as a liquid after recompression and condensation. The residue stream from the membrane is partially depleted of heavy hydrocarbons and is then sent to the sales gas stream.

Gas permeation membranes are usually made with vitreous polymers that exhibit good diffusional selectivity. However, for separation to be effective, the membrane must be very permeable with respect to the contamination to be separated, which passes through the membrane driven by pressure difference, and it must be relatively impermeable to methane (Rojey et al., 1997). Membrane Technology & Research Inc. (MTR) recently developed and commercialized a new membrane-based process for the gas processing industry. The enabling technology of this process is a unique type of rubbery membrane, which is now being applied to the separation of C_3^+ hydrocarbons from methane in gas processing. The new membrane-based process is well suited for NGL recovery and dew point control for associated gas and may be also used to debottleneck existing gas processing plants. A detailed discussion on this novel, membrane-based process and its applications in the gas processing industry can be found in Lokhandwala and Jacobs (2000).

Membrane systems are very versatile and are designed to process a wide range of feed conditions. With very compact footprint and low weight, these systems are well suited for offshore applications. Membranes could potentially remove water and heavier hydrocarbons simultaneously, thus making these systems an attractive alternative to replace the conventional dehydration and hydrocarbon dew pointing design (Baker, 2002). However, more commercial testing is needed to gain industrial acceptance. The other issue with this technology compared to more standard hydrocarbon recovery methods is that when something goes wrong, the user is totally dependent on the vendor for troubleshooting, and this is typically expensive, time consuming, and sometimes futile.

10.2.5 Twister Supersonic Separation

The invention of supersonic separation technology in 1998 brought about another milestone in the hydrocarbon dew pointing and NGL extraction, as it offers solutions to certain limitations encountered by the other technologies previously described. Twister supersonic technology uses the concept that feed gas passing through a nozzle accelerates to supersonic speed, suffering a pressure and temperature drop, where the temperature drop causes condensation of the heavier hydrocarbons. Condensation and separation at supersonic velocity are key to achieving a significant reduction in both capital and operating costs.

Twister supersonic technology shares similar benefits of simplicity, robustness, and ease of operation as the LTS (JT valve). Two studies showed that Twister can recover more hydrocarbons than the JT valve for the same pressure drop (Genesis Oil & Gas Ltd., 2007; Schinkelshoek and Epsom, 2008). Therefore, it could potentially be operated at a reduced pressure drop for the same performance as a JT valve. This reduces the sales gas compression power and cost. It can be particularly interesting for debottlenecking or upgrading existing gas plants. An additional benefit of Twister is the ability to remove water and hydrocarbons simultaneously in its tubes. Twister technology also offers environmentally friendly, chemical-free operation within a small footprint. Recently, Twister BV introduced the Twister SWIRL valve, which improves HC dew pointing performance of existing LTS plants by improving the separation of two-phase flow across a pressure reduction valve, such as a choke valve, JT valve, or control valve. This, in turn, significantly improves the liquid separation efficiency of downstream separators. This improved separation can be used to either increase flow capacity of existing LTS plants or to reduce the pressure drop required for JT cooling or to lower the HC dew point and also reduce glycol carryover.

10.2.6 Future Trends in NGL Technology Development

Many NGL recovery processes are identified in the preceding sections. However, the number of processes continues to proliferate as evidenced by the number of patents issued in the past few years. Traditional technologies for NGL extraction are being developed to reduce equipment size, improve process intensification, reduce energy requirements, and lower capital expenditure. Focus has also been on the development of supersonic separation, membrane technologies, and new generations of integrated cryogenic plants to optimize production and reduce costs.

10.2.7 Selection of NGL Recovery Processes

Choosing a cost-effective NGL recovery technology requires consideration of a broad range of factors (Mehra and Gaskin, 1999). The main variables that affect the choice of the most cost-effective process for a given application include inlet conditions (gas pressure, richness, and contaminants), downstream conditions (residue-gas pressure, level of NGL recovery desired, product specifications, and liquid fractionation infrastructure), and overall conditions (utility costs and fuel value, plant location, existing location infrastructure, and market stability). Besides the feed gas composition and operation mode, the most decisive technical characteristics of any process are the feed gas pressure and permissible unit pressure drop.[10] In fact, a widely accepted means of comparing process design is the total compression power required by the process, which provides a rough measure of the plant's capital and operating costs.

The following guidelines have been suggested for the selection of an NGL recovery process (Brands and Rajani, 2001):

1. In case of sufficiently high inlet pressure, the self-refrigeration process requires the lowest capital investment. However, if the pressure differential between the feed gas and treated gas is insufficient, additional compression is required.
2. When the feed gas pressure is close to the treated gas pressure, over a large pressure drop range, it may be more economical to employ a cryogenic refrigeration process.
3. When the feed gas pressure is clearly below the required pipeline pressure, it is usually most economical to apply mechanical refrigeration with additional compression to remove heavy hydrocarbons rather than compression followed by the self-refrigeration process. This is due to the fact that compressors are capital-intensive equipment.
4. When the feed gas pressure is equal to or lower than the required pipeline pressure, solid bed adsorption seems a good option, as it is quick to start up and robust against changes in the feed gas composition and flow rate. Generally, the solid bed process is practical only for gas that has small amounts of heavy hydrocarbons. Richer gases require refrigeration. It is clear that the solid bed adsorption process will usually be competing against the self-refrigeration process. Specially, a solid bed adsorption unit is operated at lower differential pressure compared to self-refrigeration, and thus, no additional compression is required. In fact, at low feed gas pressure and for strict dew point specifications, the economical analysis favors the solid bed adsorption process.

Quick realizations regarding selection of an NGL recovery process are difficult. In some cases, an economic comparison between viable alternatives will be required. However, for selecting an optimum NGL recovery

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[10] Pressure loss across the hydrocarbon recovery unit affects the sales gas delivery pressure. Higher pressure loss leads to increased sales gas recompression horsepower and may even affect the compressor's number of stages.

technology, it is important to identify the interaction between this technology and the other processing steps, taking into account the following characteristics: life-cycle cost, license availability and fee, environmental and operational aspects, process safety, maintenance, utility needs, etc.

10.3 NGL FRACTIONATION

The bottom liquid from the NGL recovery plant may be sold as a mixed product. This is common for small, isolated plants where there is insufficient local demand. The mixed product is transported by truck, rail, barge, or pipeline to a central location for further processing. Often it is more economical to fractionate the liquid into its various components (ethane, propane, isobutane, normal butane, and gasoline), which have a market value as purity products. However, as the relative prices of natural gas and NGLs fluctuate, the relative incentive to extract the NGLs from the gas changes. The level of NGL extraction from natural gas is somewhat discretionary. Safety issues dictate the minimum extraction level, whereas a balance between the technology and the relative market value of the NGLs determines the maximum extraction level.

Fractionation may either be undertaken at the same location as the NGL extraction unit, or the recovered NGL mixture can be transported to a separate site for fractionation. The decision to fractionate the NGL stream at a particular location will be based on economic factors such as the location of the main market for the products, the price of individual components, the volume produced, and the transportation costs (Spletter and Adair, 2001).

NGLs are fractionated by heating mixed NGL streams and passing them through a series of distillation towers (fractionators),[11] where fractionation takes advantage of the differing boiling points of the various NGL products. Each fractionator is named for the hydrocarbon that is boiled off. Therefore, a deethanizer implies that the top products is ethane; a depropanizer indicates that the top product is propane, etc.

A conventional process for fractionating the NGLs is shown in Figure 10-10. As can be seen, natural gas liquids are normally fractionated by boiling off the various fractions in sequence from the lighter to the heavier hydrocarbons. The process starts with the deethanizer, which separates ethane from the NGL stream. Next, the depropanizer separates the propane; then the debutanizer boils off the butane, leaving natural

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[11] Number of distillation towers = Number of components − 1.

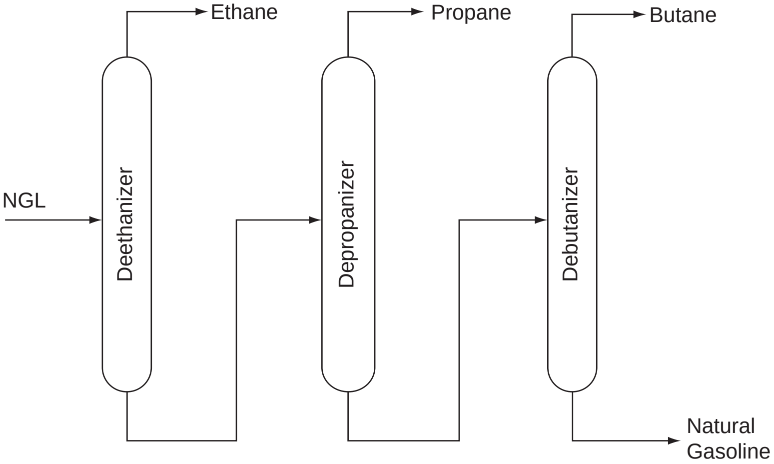

Figure 10-10 Typical fractionation process scheme.

gasoline. In some units, a butane splitter (or deisobutanizer) then further separates normal butane and isobutane. The isobutane goes overhead, and the normal butane is drawn from the bottom of the tower.

The system shown in Figure 10-10 most commonly produces ethane, commercial propane, butane, and condensate (natural gasoline) products. Ethane can be exported as a pressurized gas if there is a local market such as an ethylene cracker. Propane and butane can be stored and exported in the totally refrigerated state at atmospheric pressure or as a pressurized fluid at ambient temperature. Natural gasoline is normally stabilized so that it can be stored and exported at ambient temperature and atmospheric pressure.

There are many options in designing the order, or sequence,[12] for removing the individual products, with some schemes inherently more efficient than others (Manley, 1996; Shah, 2002). However, it is required to find the optimal arrangement of the fractionation steps that will be both technically and economically feasible.

10.3.1 Fractionator Operation

The fractionator operation takes place in a vertical column where vapor and liquid mixtures flow countercurrent and are brought into repeated contact. During each contact, part of the liquid vaporizes and part of the vapor condenses. As the vapor rises through the column, it becomes enriched in the

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[12] The number of possible technical sequences (S) is related to the number of components (C) by the relationship $S = [2(C - 1)]!/[C!\ (C - 1)!]$.

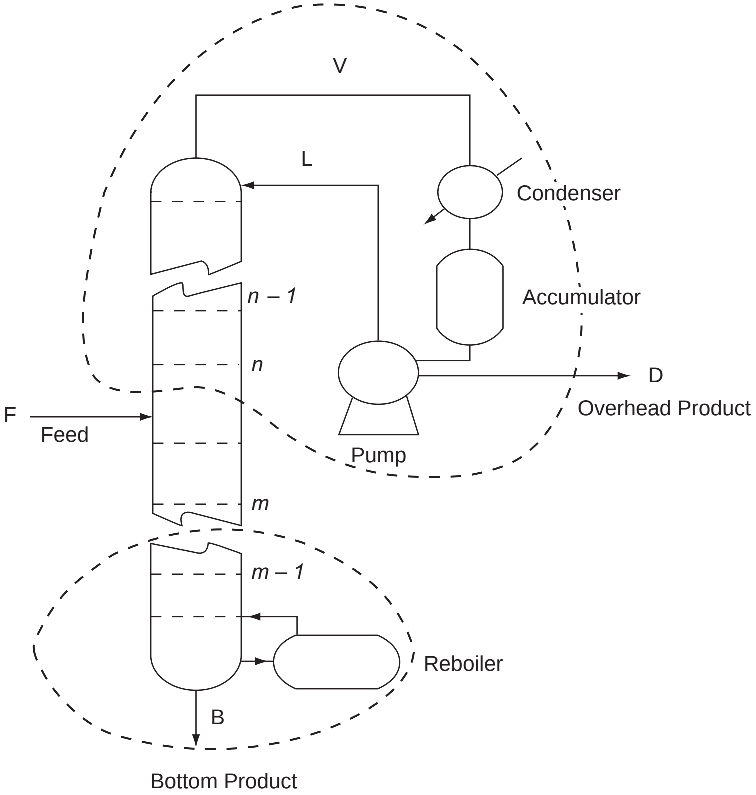

Figure 10-11 Schematic view of fractionating column (Campbell, 2000).

lighter or lower boiling components. Conversely, the downward-flowing liquid becomes richer in heavier, higher boiling components. Figure 10-11 provides a schematic view of a typical fractionating column. The liquid mixture that is to be processed is known as the feed, and this is introduced usually somewhere near the middle of the column to a tray known as the feed tray. The feed tray divides the column into a top (enriching or rectification) section and a bottom (stripping) section. The feed flows down the column, where it is collected at the bottom in the reboiler. Heat is supplied to the reboiler to generate vapor, which is reintroduced into the column at its bottom. The liquid removed from the reboiler is known as the bottom product, where the reboiler temperature sets the amount of light hydrocarbons in the bottom product. The vapor moves up the column, and as it exits the top of the column, it is cooled by a condenser. For columns using total condensers, such as depropanizers and debutanizers, all vapors are condensed. The condensed liquid is stored in a holding vessel known as the reflux drum (accumulator). Some of this liquid is recycled back to the top of the column, and this is called the reflux. The condensed liquid that is removed from the system is known as the distillate or overhead product. For columns employing partial condensers (such as deethanizers) the overhead product is produced as vapor.

To avoid reducing the operating pressure of the fractionator, which necessitates recompression, the fractionation process should be carried out with minimal loss of pressure and ideally at pipeline pressure if possible. The only limitation is that high-pressure fractionation must occur at a pressure safely below the critical pressure at every stage of the column to assure that both liquid and vapor phases will be present. The designer must also be concerned about the effect of pressure on the relative volatility of the critical components in the fractionating column.[13] As the operating pressure of the fractionating column increases, the relative volatility decreases, making it more difficult to get a clean separation between the two critical components.

The operating pressure of a fractionating column is normally set by a desired component separation and the temperature of condensing medium. This pressure is the minimum pressure at which the column can operate at the chosen condenser temperature. The pressure will correspond to either the bubble point or the dew point of the column overhead product. The overhead product will be at bubble point conditions for a liquid product or at dew point conditions for a vapor product.

10.3.2 Types of Fractionators

There are two types of fractionating columns: (1) tray column, where trays of various designs are used to hold up the liquid to provide better contact between vapor and liquid, hence better separation; and (2) packed column, where instead of trays, "packings" are used to enhance contact between vapor and liquid phases.

There are many types of tray designs, but the most common ones are bubble cap trays, valve trays, and sieve trays. However, valve and sieve trays—because of their efficiency, wide operating range, ease of maintenance, lower cost, and increased capacity (for a given diameter)—have replaced the once highly thought of bubble cap trays in many applications.

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[13] The two components in a multicomponent mixture, between which a separation is desired, are spoken of as the "key" or "critical" components. The light-boiling key component will appear concentrated in the overhead product, and the heavy key component will appear concentrated in the bottom product. The choice of these components is a decision made by the designer.

Many kinds of packing have been also invented, and several types are in common use. Packings are divided into those that are dumped at random into the column and those that must be stacked by hand. Dumped packings are used in the smaller column, whereas stacked packings are used only in the larger columns.

Traditionally, the majority of fractionating columns in gas processing plants were equipped with trays; however, packed columns have become more common in recent years. Advantages of packed columns as compared to trayed columns include lower pressure drop (0.2–0.6 inches of water per foot of packed depth) and more capacity (at a given diameter) for high liquid-to-vapor ratio systems. Packed columns also allow the use of smaller equipment and, thus, lower capital cost. Offsetting these advantages are (1) column plugging, where packed towers will be more susceptible to plugging from dirt and other foreign materials, and (2) packing's more limited turndown (about 50%) and possible problems with liquid maldistribution (which has a serious effect on packed columns having a large number of theoretical stages) and channeling (which is the chief reason for the poor performance of large, packed columns). A detailed discussion on design of packed columns, which is beyond the scope of this book, can be found in McCabe et al. (2001) and GPSA (2004).

10.3.3 Fractionator Design

Important factors in the design and operation of fractionating columns are the number of trays required to obtain the desired separation, the diameter of the column, the heat input to the reboiler, and the heat output from the condenser. In accordance with general principles, the analysis of the performance of fractionating columns is based on material and energy balances.

10.3.3.1 Overall Material Balance

In fractionator design, material balances around the column (as shown in Figure 10-11) should be made first. In steady state, two independent overall material balances can be written as follows:

$$F = D + B \tag{10-1}$$

$$F.X_{\mathrm{F}} = D.X_{\mathrm{D}} + B.X_{\mathrm{B}} \tag{10-2}$$

where F is feed molar flow rate; D is distillate molar flow rate; B is bottoms molar flow rate; X_{F} is mole fraction of a component in the feed; X_{D} is mole fraction of a component in the distillate; and X_{B} is mole fraction of a component in the bottoms.

It is obvious from Equations 10-1 and 10-2 that at a given feed composition, the distillate (overhead) and bottoms product specifications set the overall material balance for the column. In other words, in order to exactly meet the distillate and bottoms specification, there is one (and only one) distillate or bottoms flow rate, which can be drawn from the column.

10.3.3.2 Shortcut Models

The main objective of the shortcut models is to obtain estimates for the vapor/liquid traffic (i.e., reflux and reboil ratios) and the number of trays in each section of the column, which serve as the basis for sizing the equipment and estimating its capital and operating costs. The shortcut models allow the design engineer to screen and scope all of the design options, without going into minute detail, and provide a means to evaluate the effects of the operating parameters and perform a sensitivity analysis.

10.3.3.2.1 Minimum Theoretical Trays

The number of trays required to meet a given specification at the outlet is generally determined by first considering theoretical stages. A theoretical stage (tray) is an ideal stage from which the phases exist in equilibrium. However, an actual tray will not achieve equilibrium due to limitations of vapor-liquid contact time. Once the number of theoretical trays required is known, the number of actual trays is obtained by taking account of the overall efficiency of each of the trays.[14] Typical values for tray efficiency range from 0.5 to 0.7 and depend on a number of factors, such as the types of trays being used, and internal liquid and vapor flow conditions. Detailed information of this subject can be found in Perry (1997) and McCabe et al. (2001).

The minimum theoretical trays for a given separation will occur when the column is operating at a total reflux. The most rigorous method for determining minimum trays is a tray-to-tray calculation. However, the Fenske (1932) equation for minimum trays offers a rapid and fairly accurate method of computing minimum trays. It can be applied to any two components i and j in a column at infinite reflux ratio. When the keys have been chosen, components i and j are the light and heavy keys, respectively.

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[14] Sometimes, additional trays are added (up to 10%) to accommodate the possibility that the column may be under-designed.

The Fenske (1932) equation can be written in several forms (Kister, 1992). The most convenient form is as follows:

$$S_m = \frac{\log\left[\left(\frac{X_{LK}}{X_{HK}}\right)_D\left(\frac{X_{HK}}{X_{LK}}\right)_B\right]}{\log(\bar{\alpha})} - 1 \tag{10-3}$$

where S_m is minimum number of theoretical trays; X_{LK} is the mole fraction of the light key component; X_{HK} is the mole fraction of the heavy key component; and $\bar{\alpha}$ is relative volatility at average column temperature.

Subscripts D and B refer to the distillate and bottoms, respectively. Relative volatility is also defined as the K value of the light key divided by the K value of the heavy key at a given condition.[15] If the change in the value of relative volatility from the bottom of the column to the top is moderate, a geometric mean of the extreme values is recommended for relative volatility (McCabe et al., 2001).

10.3.3.2.2 Feed Tray Location

The feed tray location should be determined so that feed enters at a temperature such that no sharp change occurs in the column temperature gradient at the feed tray. There is no exact way to locate a feed tray. There are several calculation techniques that estimate feed tray location. A convenient empirical correlation is as follows (Kirkbride, 1944):

$$\log\left(\frac{N}{M}\right) = 0.206 \log\left[\left(\frac{B.X_{HK_F}}{D.X_{LK_F}}\right)\left(\frac{X_{LK_B}}{X_{HK_D}}\right)^2\right] \tag{10-4}$$

where N is the number of theoretical stages in the rectifying section; M is the number of theoretical stages in the stripping section; B is bottoms molar flow rate; D is distillate molar flow rate; X_{HK_F} is the composition of the heavy key in the feed; X_{LK_F} is the composition of the light key in the feed; X_{LK_B} is the composition of the light key in the bottoms; and X_{HK_D} is the composition of the heavy key in the distillate.

In Table 10-1, the actual number of trays is shown. The reason is that complete equilibrium between vapor and liquid is normally not reached on each tray. For calculation purposes, the number of theoretical trays may be quite a bit less than the number of actual trays.

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[15] The distribution coefficient or the K value is defined as the ratio of mole fraction of a component in the gas phase to the mole fraction of that component in the liquid phase.

Table 10-1 Typical Fractionator Parameters/Number of Trays (Arnold and Stewart, 1999)

	Approximate Ranges		
Column	*Pressure Range (psig)*	*Number of Actual Trays Above Main Feed*	*Number of Actual Trays Below Main Feed*
Lean Oil Plant			
Absorber	200–1,100	24–30	20–50
Rich Oil Demethanizer	450–600	20–30	20–50
Rich Oil Deethanizer	175–300	24–30	20–50
Rich Oil Still	85–160	12–60	16–60
Refrigeration Plant			
Demethanizer	550–650	14–30	26–30
Deethanizer	350–500	10–70	20–70
Depropanizer	200–300	17–70	18–70
Debutanizer	70–100	18–70	15–70

A McCabe–Thiele (1925) diagram may be used to estimate the feed tray location and is sometimes used since it is common practice to manifold a few trays on each side of the estimated location. Usually, as the feed stage is moved lower down the column, the top composition becomes less rich in the more volatile component, whereas the bottoms contain more of the more volatile component. However, the changes in top composition are not as marked as the bottoms composition.

10.3.3.2.3 Minimum Reflux Ratio

The minimum reflux ratio for a given separation will occur when the column is operating with an infinite number of trays,[16] and it corresponds to the minimum reboiler heat load and condenser cooling load for the separation. The most rigorous method for determining minimum trays is a tray-to-tray calculation. However, such a procedure is a laborious trial-and-error solution as exemplified by Holcomb and Brown (1942). Of the several methods of calculating the minimum reflux ratio, the most used is that of Underwood (1948), which involves use of the two following equations:

$$\sum_{i=1}^{i=n} \frac{\alpha_i X_{Fi}}{\alpha_i - \varphi} = 1 - q \tag{10-5}$$

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[16] The analysis of fractionating column is facilitated by the use of a quantity called the reflux ratio, which is the ratio of the reflux to the overhead product.

$$\sum_{i=1}^{i=n} \frac{\alpha_i X_{Di}}{\alpha_i - \varphi} = R_m + 1 \tag{10-6}$$

where α_i is relative volatility of component i at average column temperature; X_{Fi} is the mole fraction of component i in feed; X_{Di} is the mole fraction of component i in distillate; q is thermal condition of the feed, which is the total heat needed to convert one mole of feed into a saturated vapor divided by the molal latent heat of the feed (for boiling point feed, $q = 1.0$; for dew point feed, $q = 0$; and for two-phase feed, $0 < q < 1.0$); R_m is minimum reflux ratio; n is the number of components; and φ is constant.

Use of Equations 10-5 and 10-6 to calculate the minimum reflux ratio involves determination of the value of φ that will satisfy Equation 10-5. This is a trial-and-error solution, by the limitation that the only roots required are those in numerical value between the relative volatilities of the light and heavy keys. Since the relative volatilities for all components are based on the heavy key, this means that $1.0 < \varphi < \alpha_{LK}$. Thus, the value of φ that satisfies Equation 10-5 is used in Equation 10-6 to calculate the minimum reflux ratio. In this calculation, the distillate composition is obtained from the splits that were made for the column. Occasionally, the minimum reflux calculated by this method comes out a negative number. That, of course, is a signal that some other methods should be tried, or it may mean that the separation between feed and overhead can be accomplished in less than one equilibrium stage (Walas, 1990).

With the simplifying assumptions of constancy of molal overflow and temperature independence of relative volatilities, the Underwood (1948) method is exact. Further corrections can also be made for variable overflow and temperature variations of relative volatilities when the simplifying assumptions are not applicable.

The Underwood (1948) method gives only an approximate value for minimum reflux ratio. However, its results are 5%–20% high, and for this reason it can be used for column design (Kister, 1992).

10.3.3.2.4 Actual Reflux Ratio

As the reflux ratio is increased, more and more liquid that is rich in the more volatile components is being recycled back into the column. Separation then becomes better, and thus, fewer trays are needed to achieve the same degree of separation. The boundary conditions for a given separation are represented by minimum stages (total or infinite reflux) and minimum reflux (infinite stages). Obviously, neither of these conditions represents realistic operating conditions but simply shows the limit between which a column can be operated.

The Gilliland (1940) correlation is widely used for calculation of actual reflux ratio. Once the minimum theoretical stages and minimum reflux ratio have been calculated, the Gilliland correlation can be used to establish an actual reflux ratio for a fixed number of stages or, conversely, the number of stages for a fixed reflux ratio. An approximate correlation of Gilliland for tray-type columns may be represented by the following equation (Eduljee, 1975):

$$Y = 0.75\left(1 - X^{0.5668}\right) \tag{10-7}$$

where $Y = (S - S_m)/(S + 1)$; $X = (R - R_m)/(R + 1)$; R is actual reflux ratio; R_m is minimum reflux ratio; S is actual number of theoretical trays; and S_m is minimum number of theoretical trays.

The operating reflux is an amount in excess of the minimum that ultimately should be established by an economic balance between operating and capital costs for the operation. Most columns are designed to operate between 1.2 and 1.5 times the minimum reflux ratio because this is approximately the region of minimum operating costs (more reflux means higher reboiler duty). However, the total cost is not very sensitive to reflux ratio in this range, and better operating flexibility is obtained if a reflux greater than the optimum is used (McCabe et al., 2001).

Typical values of reflux ratio and tray efficiencies for various fractionating columns are given in Table 10-2. These are not design values; rather, they are guidelines for typical values in previous applications. The actual selection depends on many factors, such as feed composition, energy cost, and capital cost.

Table 10-2 Typical Fractionator Conditions (GPSA, 2004)

Tower	Reflux Ratio[1]	Reflux Ratio[2]	Tray Efficiency %
Demethanizer	Top Feed	Top Feed	45–60
Deethanizer	0.9–2.0	0.6–1.0	50–70
Depropanizer	1.8–3.5	0.9–1.1	80–90
Debutanizer	1.2–1.5	0.8–0.9	85–95
Butane Splitter	6.0–14.0	3.0–3.5	90–110
Rich Oil Fractionator (Still)	1.75–2.0	0.35–0.40	Top 67 Bottom 50
Rich Oil Deethanizer	——	——	Top 25–40 Bottom 40–60

[1] Reflux ratio relative to overhead product, lbmole/lbmole.
[2] Reflux ratio relative to feed, ft^3/ft^3.

Note that the actual number of trays required for a particular separation duty is determined by the efficiency of the tray and the packings if packings are used. Thus, any factors that cause a decrease in tray efficiency will also change the performance of the column. Tray efficiencies are affected by fouling, wear and tear, and corrosion, and the rates at which these occur depend on the properties of the liquids being processed. Therefore, appropriate materials should be specified for tray construction.

Determining the number of stages required for the desired degree of separation and the location of the feed tray is merely the first step in designing a fractionator. Other things that need to be considered are tray spacing, column diameter, internal configurations, and heating and cooling duties. All of these can lead to conflicting design parameters. Thus, fractionating column design is often an iterative procedure. If the conflicts are not resolved at the design stage, then the column will not perform well in practice.

10.3.3.2.5 Column Capacity

The capacity of a column is expressed in terms of the rate of either vapor or liquid flow per unit plate area. Depending on column diameter and the types of internals used, there are hydraulic limits to vapor and liquid handling capacity, both on the high and low sides. At very high vapor or liquid loads, the column can flood due to excessive liquid entrainment or insufficient downcomer capacity. At very low throughputs, problems such as liquid weeping and vapor pulsation can occur. Loss of desired separation efficiency can also occur before column hydraulic limits are reached, resulting in product specifications not being met. Hence, the column needs to operate within the boundaries of the region designated "satisfactory performance."

To evaluate the column hydraulics, one calculates the diameter required for vapor flow for those stages, where there is a significant change in the vapor and liquid flows. Such stages include the top and bottom trays, pump-around stages, and the feed stage. The diameter profile along the column is then obtained by plotting the diameters for various stages, where this profile allows identification of the column bottlenecks that limit throughput enhancement (Gadalla et al., 2003).

The column diameter required to handle a given vapor load can be found by applying the following equation at the point of greatest vapor load, generally just below the top tray (Souders and Brown, 1932):

$$V_{\max} = C\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \tag{10-8}$$

where $V_{\max}$ is maximum vapor velocity, ft/sec; ρ_L is liquid density, lb_m/ft^3; ρ_V is vapor density, lb_m/ft^3; and C is sizing constant, ft/sec.

The sizing constant (C) has been evaluated from the plant data, and various correlations of C against operating variables have been made (Perry, 1997). The values in Table 10-3 are indicative of those used commonly.

The column capacity as determined by the Souders and Brown (1932) equation is from 20% to 40% conservative, as pointed out by Brown and Martin (1939).

The general method of column design is to choose a plate spacing, estimate the permissible vapor load, and calculate the column diameter as follows (GPSA, 2004):

$$D = \sqrt{\frac{\dot{Q}_V}{V_{\max} \times 0.7854}} \tag{10-9}$$

where D is column diameter, ft; $\dot{Q}_V$ is volumetric vapor flow rate, ft^3/sec; and $V_{\max}$ is maximum vapor velocity, ft/sec.

This method was originally developed for bubble cap trays and gives a conservative diameter, especially for other types of trays.

In cases where a large liquid load is handled, the column diameter may be determined by the liquid load rather than the vapor load, and to obtain a safe liquid load, the final design may be underloaded with respect to vapor flow (McCabe et al., 2001).

If the diameter of a fractionator is 2 feet or less, it is more economical in many cases to use a packed column in place of a tray column. Manufacturers supply data for their packing material, which indicates the number of feet of packing required to provide the same mass transfer as a standard bubble cap tray.

Some recent advances in structured packing are being used by some operators in larger diameter towers where they normally would have used trays. The structured packing is said to allow both smaller diameter and less height of tower. In addition, packing provides extra interfacial area for liquid-vapor contact, and the efficiency of separation is therefore increased for the same column height.

Table 10-3 Value of Sizing Constant Versus Tray Spacing (Souders and Brown, 1932)

C, ft/sec	Tray Spacing, inch
0.122	18
0.150	21
0.167	24

The height of the column is a function of the number of theoretical stages and the efficiency of the actual stages. Once the operating conditions are established for a column, its diameter and height can be chosen using data available from tray and/or packing manufacturers.

10.3.3.3 Overall Heat Balance

The last step in the basic calculation scheme of the process design for a fractionating column is to make an overall heat balance around the fractionator. This is a very important step in determining the desirability and economic feasibility of some of the arbitrary assumptions that were made in the design of the tower.[17] The overall heat balance also has value in evaluation or troubleshooting existing fractionators. The overall energy balance may point out a control problem that exists or some other reason that a tower is operating improperly.

The overall heat balance around the fractionator can be written as follows:

$$Q_R + Q_C = h_D D + h_B B - h_F F \tag{10-10}$$

where Q_R is reboiler heat duty; Q_C is condenser heat duty; h_D is enthalpy of distillate product; h_B is enthalpy of bottom product; h_F is enthalpy of feed; and D, B, and F are rate of flow of distillate, bottom, and feed streams, respectively.

The heat duty of the reboiler given here should be increased by the heat loss from the fractionator. The heat loss is about 3% of the reboiler duty for light hydrocarbon fractionators but can be estimated more accurately by using a heat-transfer coefficient of 1.0 $\mathrm{Btu/hr/ft^2.°F}$ and the temperature difference between the column insulation and the atmospheric temperature. At this point, the results of all calculations in the process design of the fractionator should be carefully reviewed and studied. If all results appear satisfactory, then final calculation on column sizing may be complete.

10.3.3.4 Design Procedure

In order to determine the design parameters for a fractionating column, the following procedure is recommended (Campbell, 2000; GPSA, 2004):

1. Establish feed composition, flow rate, temperature, and pressure.
2. Make product splits for the column based on feed and desired product needs.

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[17] The actual variation in the vapor and liquid streams in a fractionating column is determined by an enthalpy balance, and the limitation imposed by the assumption of constant molal overflow can be removed by a rigorous enthalpy balance used in conjunction with material balances and phase equilibria.

3. Establish a condenser temperature (for the overhead stream) and column pressure, which will be used for calculation of the reboiler temperature.
4. Calculate the minimum number of theoretical trays and the minimum reflux ratio needed to produce the desired products.
5. Calculate the combination of actual theoretical trays and actual reflux ratio needed.
6. Make a heat balance around the condenser to determine condenser duty and around the column to determine reboiler duty.
7. Size the column.

Traditionally, fractionating columns have been designed using the equilibrium stage approach. However, the correct and general approach to column design is not to use the equilibrium stage but to change, head-on, the complete nonequilibrium problem by taking into account the heat and mass transfer processes on trays. The nonequilibrium stage approach must incorporate the proper description of interphase mass transfer. Also, simultaneous heat transfer effects need to be incorporated into the model formulation. The incentive to adopt the nonequilibrium stage approach is that column profile predictions using this approach and the traditional approaches can be markedly different. Such differences could have a significant effect on column design. However, the major bottleneck in the use of this approach is the lack of generally applicable mass transfer correlations for trays and packings. Hence, new design methods are still on the conservative side.

10.4 LIQUIDS PROCESSING

The liquid products produced from the NGL recovery unit rarely achieve product specifications without further treatment. The following sections discuss the processing required to produce marketable liquid products.

10.4.1 Sweetening

If acidic compounds are present in the feed gas and have not been removed prior to NGL recovery, then they will end up in the LPG and condensate products. Especially objectionable contaminants are H_2S, mercaptans, and elemental sulfur. These contaminants not only can lead to odor problems but also can form objectionable oxides on combustion and cause pollution.[18] Each of these contaminants can also cause corrosion problem unless the sour liquid stream is dehydrated properly[19] (Maddox, 1982). The presence of significant quantities of CO_2 can also cause problems by raising the vapor pressure of the light hydrocarbon liquids and lowering the heating value. Carbonyl sulfide (COS) and carbon disulfide (CS2), although not corrosive in LPG, will hydrolyze slowly to H_2S in the presence of free water and cause the liquid product to become corrosive (Bullin et al., 1995).

A liquid product containing the objectionable materials can be treated to remove the H_2S, COS, and elemental sulfur and to either remove the mercaptans or convert them to less-objectionable compounds. Condensate stream may or may not require sweetening depending on the intended markets. If sweetening is required, and depending on the specifics of the contaminants, this can be achieved in a variety of ways. The following discussion will emphasize removal of sulfur compounds (as well as residual CO2) and the removal or conversion of the mercaptans from light hydrocarbon liquids.

10.4.1.1 Caustic Treating

The hydrocarbon processing industry has historically used caustic solutions to extract or treat acidic impurities in liquid hydrocarbon streams. A number of caustic processes, both regenerative and nonregenerative, are used to remove sulfur compounds from hydrocarbon liquids. The simplest process is the use of a nonregenerative solid KOH bed, effective for removal of H2S but not other sulfur compounds. However, by injection of methanol into the bed with the propane, COS, mercaptans, and CS2 can be removed (Mick, 1976). One of the most common processes for treating hydrocarbon liquids is the use of regenerative caustic wash with sodium hydroxide (downstream of an amine extraction system for removing bulk acidic components and reducing the load on the caustic wash that removes mercaptans). When acid gas quantities are small, a simple caustic wash is both effective and economical. However, as the quantity of contaminants rises, a caustic wash process can be quite expensive due to the caustic cost and

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[18] While small amounts of mercaptans are used as odorizers for liquid products in the propane-butane range, larger amounts are objectionable.

[19] In general, most liquid hydrocarbon products must meet a 1A copper strip corrosion test that corresponds to less than 4 ppm H_2S. According to Perry (1977), the copper strip test is not sensitive to COS. However, in the presence of water, COS can react to form H_2S. Therefore, liquids containing COS will often not meet the copper strip test after storage.

disposal problems. Operating companies often choose a treating process based on spent caustic disposal rather than the favorable economics of caustic treating.

A large number of organic materials may be added to caustic solutions (as promoters) to markedly increase their solubility for mercaptans. More details on various promoter processes, which are effective in removing mercaptans from light hydrocarbon liquids, can be found in Maddox (1982). Mercaptans can be converted to disulfides by several methods. These disulphides will remain in the sweetened hydrocarbon product. The overall sulfur content, therefore, remains the same. However, the sulfur leaves as disulfide (no odor) rather than mercaptans. The method or combination of methods that can be used depends on the mercaptan content of the product to be treated and the specification that must be met (Fischer et al., 1993). Among the processes that convert (oxidize) mercaptans to disulfides, the "Doctor sweetening" process is the oldest. Doctor treatment ordinarily will leave 0.0004% mercaptan, at which level there is no or negligible effect on the tetraethyl lead susceptibility of the gasoline. In this process, an alkaline solution of lead oxide (usually sodium plumbite) contacts the hydrocarbon stream forming lead mercaptides (soluble in the oil) with the mercaptans. The mixture is then treated with powdered sulfur (which has a high affinity for lead), and a conversion of the mercaptide into a so-called disulfide (which remains in solution in the gasoline stream) occurs. The reactions of the "Doctor treating" process are considered to be

$$2RSH + Na_2PbO_2 \rightarrow (RS)_2Pb + 2NaOH \tag{10-11}$$

$$(RS)_2Pb + S \rightarrow R_2S_2 + PbS \tag{10-12}$$

Note that sulfur should be added in stoichiometric excess in order to get maximum conversion of the mercaptides to disulfides, where too much excess sulfur will cause formation of polysulfides (Maddox, 1982). Sometimes with no sulfur added but in the presence of atmospheric oxygen and sodium hydroxide solution, the same conversion (Eq. 10-12) occurs, but only slowly and not completely (McBryde, 1991).

The Merox™ process,[20] developed and commercialized over 50 years ago by UOP, is used to treat end-product streams by rendering any mercaptan sulfur compounds inactive. This process can be used for treating LPG, gasoline, and heavier fractions. The method of treatment is the extraction reaction of the sour feedstock containing mercaptans (RSH) with caustic soda (NaOH)

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[20] Merox stands for mercaptan oxidation.

in a single, multistage extraction column using high-efficiency trays. The extraction reaction is shown by the following equation:

$$RSH + NaOH \leftrightarrow NaSR + H_2O \tag{10-13}$$

After extraction, the extracted mercaptans in the form of sodium mercaptides (NaSR) are catalytically oxidized (by Merox WS catalyst, which is dispersed in the aqueous caustic solution) to water-insoluble disulfide oils (RSSR), as shown in the following equation:

$$4\ NaSR + O_2 + 2\ H_2O \rightarrow 2\ RSSR + 4\ NaOH \tag{10-14}$$

The disulfide oil is decanted and sent to fuel or to further processing in a hydrotreater. The regenerated caustic is then recirculated to the extraction column.

Typical product mercaptan levels can be controlled to less than 10 ppmw (UOP, 2003).

The Sulfrex™ process (developed by Axens) is another technology similar to the Merox process, which has demonstrated its industrial performance, robustness, simple design, and ease in operation since 1975.

10.4.1.2 Adsorption

Adsorption is a commonly used method for sweetening natural gas liquids. Molecular sieves have the advantage of drying the product as it is being sweetened, but their drawbacks are large capital and operating costs as well as catalyzing the formation of COS if both H_2S and CO_2 are present (Fleming et al., 1988). More details on this process can be found in Maddox (1982), GPSA (2004), and Kidnay and Parrish (2006).

10.4.1.3 Amine Treating

Amine treating is an attractive alternative, especially when an amine gas sweetening unit is already on-site. In such cases, the liquid treating unit can often be operated using a slipstream of amine from the main sweetening unit (Nielsen et al., 1997).

Amine treating is often used upstream of caustic treaters to minimize caustic consumption caused by irreversible reactions with CO_2. In this process, H_2S and CO_2 from the sour liquid feed are absorbed by liquid-liquid contacting the sour liquid with lean amine solvent. The design options for the absorber include the selection of an amine and the method of contact. Any of the commonly used ethanolamines (including DGA and DIPA, and the MDEA-based specialty solvents) will usually perform satisfactorily. The liquid-liquid contacting devices include packed towers, trayed towers, jet eductor-mixers, and static mixers. Amine treating is usually conducted in a packed tower, although sieve trays are effective as well (Sargent and Seagraves, 2003). In this process, liquid hydrocarbon enters the bottom of a packed absorber, and lean amine enters the top of the absorber. Sweet liquid leaves the absorber from the top, and rich amine leaves the absorber from the bottom. The sweet liquid is washed using a recirculating water wash to both recover amine and protect downstream caustic treaters (Stewart and Lanning, 1991). The treated liquid and the wash water are mixed in the water-wash static mixer, which is then coalesced into two liquid phases and separated in the water-wash separator (Nielsen, 1995). Makeup water is continuously added to the circulating water-wash circuit to control the buildup of amine in the wash water and to help maintain the water content of the amine system. Water is also continuously withdrawn from the water-wash circuit and mixed with the rich amine solution.

10.4.1.4 Other Processes

Not one of the aforementioned processes addresses the problem of elemental sulfur and carbonyl sulfide removal. Elemental sulfur is removed from the natural gasoline by contacting it with a polysulfide wash solution (GPSA, 2004). Carbonyl sulfide is a stable, unreactive compound that is very difficult to reduce to concentration levels below 1 ppmv using conventional amine and molecular sieve processes. The ADIP process is a regenerative amine process developed by Jacobs to selectively reduce COS to very low levels (5 ppm wt as S) in liquid hydrocarbons such as LPG and NGL. Numerous nonregenerative metallic oxide processes are also available to remove COS from liquid products. Some of these processes remove the COS directly, and others require water to hydrolyze the COS to H_2S before it is reacted (Maddox, 1982; Kidnay and Parrish, 2006).

10.4.2 Dehydration

Natural gas liquids must be dehydrated to meet requirements of a handling chain to a direct consumer. The acceptable water content in light hydrocarbon liquid streams varies from no free water present to very low levels of moisture in liquid products. Numerous processes are available for drying liquids, and many of the technologies discussed in Chapter 9 for dehydrating gases may be used for liquids.

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CHAPTER 11

Natural Gas Compression

11.1 INTRODUCTION

"Compression" is used in all aspects of the natural gas industry, including gas lift, reinjection of gas for pressure maintenance, gas gathering, gas processing operations (gas loading and discharge), transmission and distribution systems, and boil-off systems (in gas storage and tankers for vapor control and to avoid releasing gas to the atmosphere). In recent years, there has been a trend toward increasing pipeline-operating pressures. The benefits of operating at higher pressures include the ability to transmit larger volumes of gas (referred at base conditions) through a given size of pipeline, lower transmission losses due to friction, and the capability to transmit gas over long distances without requiring or even reducing additional compressor stations. In gas transmission, two basic types of compressors are used: reciprocating and centrifugal compressors. Reciprocating compressors are usually driven by either electric motors or gas engines, whereas centrifugal compressors use gas turbines or electric motors as drivers. The key variables for equipment selections are life-cycle cost, capital cost, and maintenance costs including overhaul and spare parts, fuel, or energy costs. The units level of utilization, as well as demand fluctuations, plays an important role. Both gas engines and gas turbines can use pipeline gas as a fuel, but electric motors have to rely on the availability of electric power. Due to the number of variables involved, the task of choosing the optimum driver can be quite involved, and a comparison between the different types of drivers should be done before a final selection is made (Lubomirsky et al., 2010). An economic feasibility study is of fundamental importance to determine the best selection for the economic life of a project. Furthermore, it must be decided whether the compression task should be divided into multiple compressor trains, operating in series or in parallel (Santos, 1997, 2004).

This chapter presents a brief overview of the two major types of compressors, a procedure for calculation of the required compression power, as well as additional and useful considerations for the design of compressor stations. All performance calculations should be based on compressor suction and discharge flange conditions. For reciprocating compressors, pressure losses at the cylinder valves as well as the pressure losses in pulsation dampeners have to be included in the calculation. Additional losses for process equipment such as suction scrubbers, intercoolers, and aftercoolers have to be accounted to define compressor design conditions.

11.2 RECIPROCATING COMPRESSORS

A reciprocating compressor is a positive displacement machine in which the compressing and displacing element is a piston moving linearly within a cylinder. The reciprocating compressor uses automatic spring-loaded valves that open when the proper differential pressure exists across the valve. Figure 11-1 illustrates the action of a reciprocating compressor using

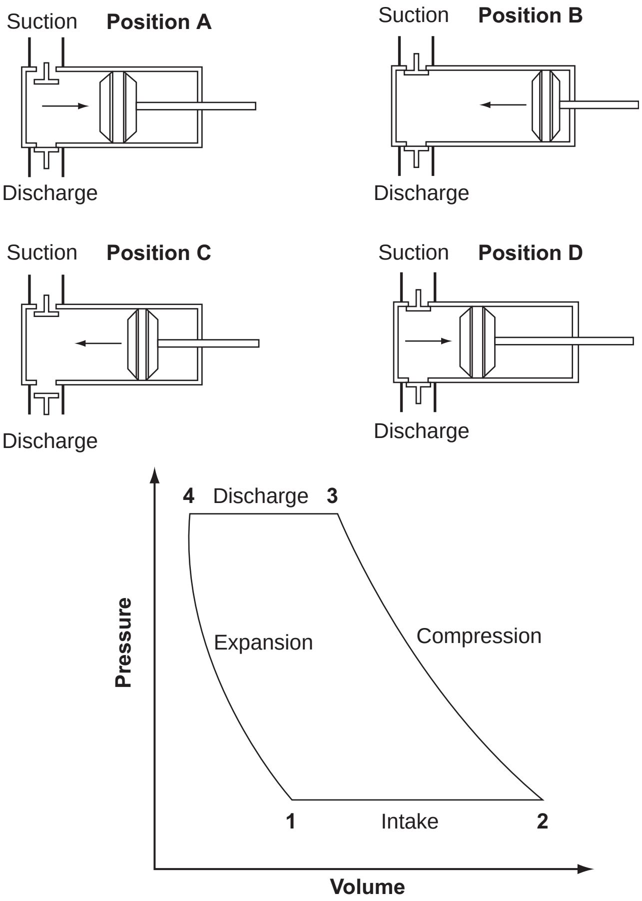

Figure 11-1 Reciprocating compressor compression cycle.

a theoretical pressure-volume (PV) diagram. In position A, the suction valve is open, and gas will flow into the cylinder (from point 1 to point 2 on the PV diagram) until the end of the reverse stroke at point 2, which is the start of compression. At position B, the piston has traveled the full stroke within the cylinder, and the cylinder is full of gas at suction pressure. Valves remain closed. The piston begins to move to the left, closing the suction valve. In moving from position B to position C, the piston moves toward the cylinder head, reducing the volume of gas with an accompanying rise in pressure. The PV diagram shows compression from point 2 to point 3. The piston continues to move to the end of the stroke (near the cylinder head) until the cylinder pressure is equal to the discharge pressure and the discharge valve opens (just beyond point 3). After the piston reaches point 4, the discharge valve will close, leaving the clearance space filled with gas at discharge pressure (moving from position C to position D). As the piston reverses its travel, the gas remaining within the cylinder expands (from point 4 to point 1) until it equals suction pressure and the piston is again in position A.

The flow to and from reciprocating compressors is subject to significant pressure fluctuations, due to the reciprocating compression process. Therefore, pulsation dampeners have to be installed upstream and downstream of the compressor to avoid damages to other equipment. The pressure losses (several percent of the static flow pressure) in these dampeners have to be accounted for in the station design.

Reciprocating compressors are widely utilized in the gas processing industries because they are flexible in throughput and discharge pressure range. Reciprocating compressors are classified as either “high speed” or “slow speed.” Typically, high-speed compressors operate at speeds of 900 to 1,200 rpm and slow-speed units at speeds of 200 to 600 rpm. High-speed units are normally “separable.” That is, the compressor frame and driver are separated by a coupling or gearbox. For an “integral” unit, power cylinders are mounted on the same frame as the compressor cylinders, and the power pistons are attached to the same drive shaft as the compressor cylinders. Low-speed units are typically integral in design.

11.3 CENTRIFUGAL COMPRESSORS

We want to introduce now the essential components of a centrifugal compressor that accomplish the tasks specified in the preceding section (see Figure 11-2). The gas entering the inlet nozzle of the compressor is guided to the inlet of the impeller. An impeller consists of a number of rotating vanes

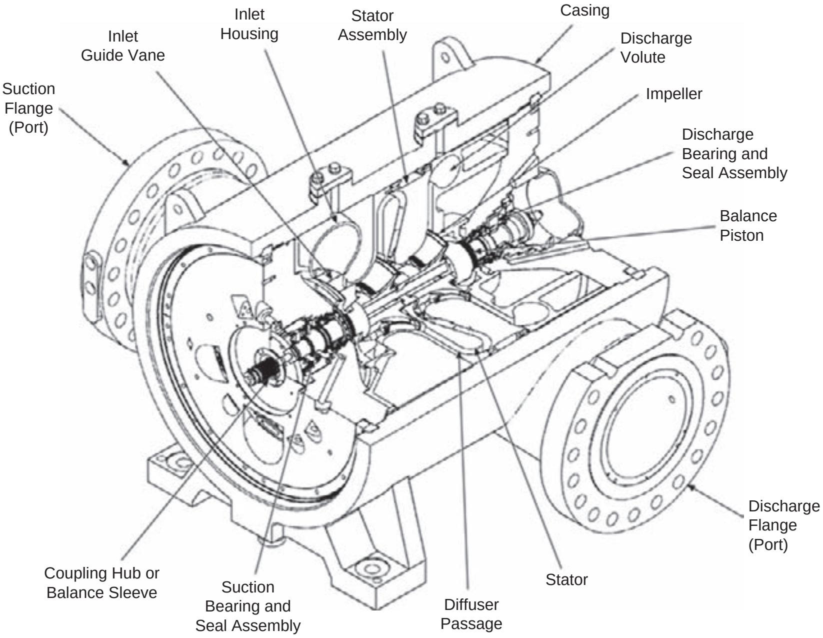

Figure 11-2 Typical centrifugal compressor cutaway (GPSA, 2004).

that impart the mechanical energy to the gas. The gas will leave the impeller with an increased velocity and increased static pressure. In the diffuser, part of the velocity is converted into static pressure. Diffusers can be vaned, vaneless, or volute type. If the compressor has more than one impeller, the gas will be again brought in front of the next impeller through the return channel and the return vanes. If the compressor has only one impeller, or after the diffuser of the last impeller in a multistage compressor, the gas enters the discharge system. The discharge system can make use either of a volute, which can further convert velocity into static pressure, or a simple cavity that collects the gas before it exits the compressor through the discharge flange.

The rotating part of the compressor consists of all the impellers. It runs on two radial bearings (on all modern compressors, these are hydrodynamic tilt pad bearings), while the axial thrust generated by the impellers is balanced by a balance piston, and the residual force is balanced by a hydrodynamic tilt pad thrust bearing. To keep the gas from escaping at the shaft ends, dry gas seals are used. The entire assembly is contained in a casing (usually barrel type).

A compressor stage is defined as one impeller, with the subsequent diffuser and (if applicable) return channel. A compressor body may hold one or several (up to 8 or 10) stages. A compressor train may consist of one or multiple compressor bodies. It sometimes also includes a gearbox. Pipeline compressors are typically single body trains, with one or two stages.

The different working principles cause differences in the operating characteristics of the centrifugal compressors compared to those of the reciprocating unit. Centrifugal compressors are used in a wide variety of applications in chemical plants, refineries, onshore and offshore gas lift and gas injection applications, gas gathering, and in the transmission and storage of natural gas. Centrifugal compressors can be used for outlet pressures as high as 10,000 psia, thus overlapping with reciprocating compressors over a portion of the flow-rate/pressure domain. Centrifugal compressors are usually either turbine or electric motor driven. Typical operating speeds for centrifugal compressors in gas transmission applications are about 14,000 rpm for 5,000 hp units and 8,000 rpm for 20,000 hp units.[1]

11.4 COMPARISON BETWEEN COMPRESSORS

The differences between reciprocating and centrifugal compressors are summarized as follows (GPSA, 2004).

The advantages of a reciprocating compressor over a centrifugal machine include

- Ideal for low volume flow and high-pressure ratios
- High efficiency at high-pressure ratios
- Relatively low capital cost in small units (less than 3,000 hp)
- Less sensitive to changes in composition and density

The advantages of a centrifugal compressor over a reciprocating machine include

- Ideal for high volume flow and low head
- Simple construction with only one moving part
- High efficiency over normal operating range
- Low maintenance cost and high availability
- Greater volume capacity per unit of plot area
- No vibrations and pulsations generated

11.5 COMPRESSOR SELECTION

The design philosophy for choosing a compressor should include the following considerations:

- Good efficiency over a wide range of operating conditions
- Maximum flexibility of configuration

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[1] HP or hp stands for horsepower, a unit of power.

- Low maintenance cost
- Low life-cycle cost
- Acceptable capital cost
- High availability

However, additional requirements and features will depend on each project and on specific experiences of the pipeline operator. In fact, compressor selection consists of the purchaser defining the operating parameters for which the machine will be designed. The "process design parameters" that specify a selection are as follows (Akhtar, 2002):

1. Flow rate
2. Gas composition
3. Inlet pressure and temperature
4. Outlet pressure
5. Train arrangement
    - **5.1** For centrifugal compressors: series, parallel, multiple bodies, multiple sections, intercooling, etc.
    - **5.2** For reciprocating compressors: number of cylinders, cooling, and flow control strategy
6. Number of units

In many cases, the decision whether to use a reciprocating compressor or a centrifugal compressor, as well as the type of driver, will already have been made based on operator strategy, emissions restrictions, general life-cycle cost assumptions, etc. However, a hydraulic analysis should be made for each compressor selection to ensure the best choice. In fact, compressor selection can be made for an operating point that will be the most likely or most frequent operating point of the machine. Selections based on a single operating point have to be carefully evaluated to provide sufficient speed margin (typically 3%–10%) and surge margin to cover other potentially important operating conditions. A compressor performance map (head versus flow map) can be generated based on the selection and is used to evaluate the compressor for other operating conditions by determining the head and flow required for those other operating conditions. In many applications, multiple operating points are available, based on hydraulic pipeline studies or reservoir studies. Some of these points may be frequent operating points, whereas some may occur during upset conditions. With this knowledge, the selection can be optimized for a desired target, such as lowest fuel consumption.

Selections can also be made based on a "rated" point, which happens to be the most onerous operating conditions (highest volumetric flow rate; lowest molecular weight; highest head or pressure ratio; highest inlet temperature). In this situation, however, the result may be an oversized machine that does not perform well at the usual operating conditions.

Once a selection is made, the manufacturer is able to provide parameters such as efficiency, speed, power requirements and, based on these and the knowledge of the ambient conditions (prevailing Summer and Winter temperatures and site elevation), can size the drivers. At this point, the casing arrangement, the number of required units based on desirable flexibility, growth scenarios, and sparing considerations, will play an important role in this decision.

11.6 THERMODYNAMICS OF GAS COMPRESSION

The task of gas compression is to bring gas from a certain suction pressure to a higher discharge pressure by means of mechanical work. The actual compression process is often compared to one of three ideal processes; isothermal, isentropic, or polytropic compression.

Isothermal compression occurs when the temperature is kept constant during the compression process. It is not adiabatic because the heat generated in the compression process has to be removed from the system.

The compression process is isentropic or adiabatic reversible if no heat is added to or removed from the gas during compression and the process is frictionless. With these assumptions, the entropy of the gas does not change during the compression process.

The polytropic compression process, like the isentropic cycle, is reversible, but it is not adiabatic. It can be described as an infinite number of isentropic steps, each interrupted by isobaric heat transfer. This heat addition guarantees the process will yield the same discharge temperature as the real process.

It is important to clarify certain properties at this time and, in particular, find their connection to the first and second law of thermodynamics written for steady-state fluid flows. The first law (defining the conservation of energy) becomes

$$\left(h_2 + \frac{u_2^2}{2} + gz_2\right) - \left(h_1 + \frac{u_1^2}{2} + gz_1\right) = q_{12} + W_{t,12} \qquad (11\text{-}1)$$

where h is enthalpy, u is velocity, g is gravitational acceleration, z is elevation coordinate, q is heat, and W_t is work done by the compressor on the gas.

Neglecting the changes in potential energy (because the contribution due to changes in elevation is not significant for gas compressors), we can write the energy balance equation for adiabatic processes ($q_{12} = 0$) as

$$\left(h_2 + \frac{u_2^2}{2}\right) - \left(h_1 + \frac{u_1^2}{2}\right) = W_{t,12} \tag{11-2}$$

where $W_{t,12}$ is the amount of work we have to apply to affect the change in enthalpy in the gas.[2] The work $W_{t,12}$ is related to the required power, P, by multiplying it with the mass flow:

$$P = \dot{m} W_{t,12} \tag{11-3}$$

Combining enthalpy and velocity into a total enthalpy ($h_t = h + \frac{u^2}{2}$), power and total enthalpy difference are thus related by

$$P = \dot{m}\left(h_{t,2} - h_{t,1}\right) \tag{11-4}$$

If we can find a relationship that combines enthalpy with the pressure and temperature of a gas, we have found the necessary tools to describe the gas compression process. For a perfect gas, with constant heat capacity, the relationship between enthalpy, pressures, and temperatures is

$$\Delta h = C_p(T_2 - T_1) \tag{11-5}$$

where T_1 is suction temperature, T_2 is discharge temperature, and C_P is heat capacity at constant pressure.

For an isentropic compression, the discharge temperature (T_{2s}) is determined by the pressure ratio as

$$T_{2s} = T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] + T_1 \tag{11-6}$$

where $k = C_p/C_v$; p_1 is suction pressure; and p_2 is discharge pressure.

Note that the specific heat at constant pressure (C_P) and the specific heat at constant volume (C_V) are functions of temperature only for ideal gases and can be related together with $C_P - C_V = R$, where R is the universal gas

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[2] Physically, there is no difference between work, head, and change in enthalpy. In systems with consistent units (such as the SI system), work, head, and enthalpy difference have the same unit (e.g., kJ/kg in SI units). Only in inconsistent systems (such as U.S. customary units), we need to consider that the enthalpy difference (e.g., in Btu/lb_m) is related to head and work (e.g., in ft.lb_f/lb_m) by the mechanical equivalent of heat (e.g., in ft.lb_f/Btu).

constant. The isentropic exponent (k) for ideal gas mixtures can also be determined as

$$k = \frac{\sum y_\mathrm{i}\ C_\mathrm{Pi}}{\left[\sum y_\mathrm{i}\ C_\mathrm{Pi}\right] - R} \tag{11-7}$$

where C_Pi is the molar heat capacity of the individual component, and y_i is the molar concentration of the component.

The heat capacities of real gases are a function of the pressure and temperature and thus may differ from the ideal gas case. For hand calculations, the ideal gas k is sufficiently accurate.

If the gas composition is not known, and the gas is made up of alkanes (methane, ethane, etc.) with no substantial quantities of contaminants and whose specific gravity (SG) does not exceed unity, the following empirical correlation can be used (Campbell, 1992):

$$k = 1.3 - 0.31(\mathrm{SG} - 0.55) \tag{11-8}$$

Combining Equations 11-5 and 11-6, we can thus determine the isentropic head (Δh_s) for the isentropic compression of a perfect gas as

$$\Delta h_\mathrm{s} = C_\mathrm{p} T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \tag{11-9}$$

For real gases (where k and C_p in the preceding equation become functions of temperature and pressure), the enthalpy of a gas, h, is calculated in a more complicated way using equations of state. These represent relationships that allow us to calculate the enthalpy of gas of known composition, if any two of its pressure, its temperature, or its entropy are known.

We therefore can calculate the actual head for the compression by

$$\Delta h = h(p_2, T_2) - h(p_1, T_1) \tag{11-10}$$

and the isentropic head by

$$\Delta h_\mathrm{s} = h(p_2, s_1) - h(p_1, T_1) \tag{11-11}$$

where entropy of the gas at suction condition (s_1) is

$$s_1 = s(p_1, T_1) \tag{11-12}$$

The relationships described can be easily seen in a Mollier diagram (see Figure 11-3).

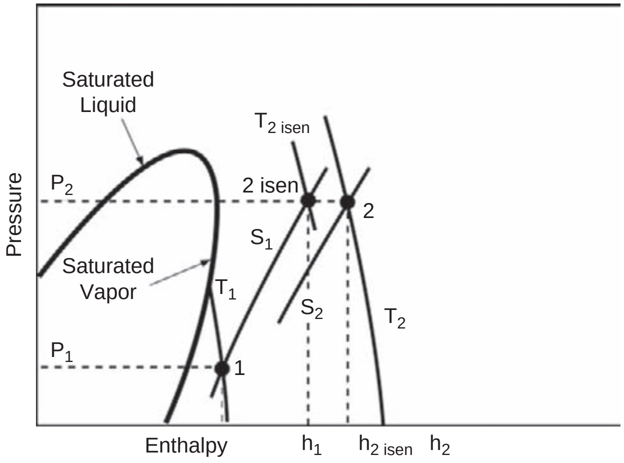

Figure 11-3 Compression process in a Mollier (Pressure-Enthalpy) diagram with lines of constant temperature and constant entropy. *(Courtesy of Solar Turbines Inc.)*

The performance of a compressor can be assessed by comparing the actual head (which directly relates to the amount of required compression power) with the calculated head for an ideal, isentropic compression. This defines the isentropic efficiency (η_s) as

$$\eta_s = \frac{\Delta h_s}{\Delta h} \tag{11-13}$$

For ideal gases, the actual head can be calculated from

$$\Delta h = \frac{1}{\eta_s} C_p T_1 \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] \tag{11-14}$$

and further, the actual discharge temperature (T_2) becomes

$$T_2 = \frac{T_1}{\eta_s} \left[\left(\frac{p_2}{p_1} \right)^{\frac{k-1}{k}} - 1 \right] + T_1 \tag{11-15}$$

The second law of thermodynamics tells us

$$\dot{m}(s_2 - s_1) = \int_1^2 \frac{dq}{T} + s_{\mathrm{irr}} \tag{11-16}$$

For adiabatic flows, where no heat enters or leaves, the change in entropy simply describes the losses generated in the compression process. These losses come from the friction of gas with internal surfaces and the mixing of gas of different energy levels. An adiabatic, reversible compression process therefore does not change the entropy of the system; it is isentropic. Our equation for the actual head implicitly includes the entropy rise Δs because

$$\Delta h = h(p_2, T_2) - h(p_1, T_1) = h(p_2, s_1 + \Delta s) - h(p_1, s_1) \quad (11\text{-}17)$$

If cooling is applied during the compression process (for example with intercoolers between two compressors in series), then the increase in entropy is smaller than an uncooled process. Therefore, the power requirement will be reduced.

Using the polytropic process for comparison reasons works fundamentally the same way as using the isentropic process for comparison reasons (Beinecke and Luedtke, 1983). The difference lies in the fact that the polytropic process uses the same discharge temperature as the actual process, whereas the isentropic process has a lower discharge temperature than the actual process for the same compression task. In particular, both the isentropic and the polytropic process are reversible processes. The isentropic process is also adiabatic, whereas the polytropic process assumes a specific amount of heat transfer. In order to fully define the isentropic compression process for a given gas, we have to know suction pressure, suction temperature, and discharge pressure. Additionally, to define the polytropic process, we have to know either the polytropic compression efficiency or the discharge temperature. The polytropic efficiency (η_p) is constant for any infinitesimally small compression step, which then allows us to write

$$\Delta h = \frac{1}{\eta_p} \int_{p_1}^{p_2} v\,dp = \frac{\Delta h_p}{\eta_p} \quad (11\text{-}18)$$

where v is specific volume and the polytropic head (Δh_p) can be calculated from

$$\Delta h_p = Z_1 R T_1 \frac{n}{n-1} \left[\left(\frac{p_2}{p_1} \right)^{\frac{n-1}{n}} - 1 \right] \quad (11\text{-}19)$$

This determines the polytropic efficiency as

$$\eta_p = \frac{\Delta h_p}{\Delta h} \quad (11\text{-}20)$$

For compressor designers, the polytropic efficiency has an important advantage. If a compressor has five stages, and each stage has the same isentropic efficiency η_s, then the overall isentropic compressor efficiency will be lower than η_s. If, for the same example, we assume that each stage has the same polytropic efficiency η_p, then the polytropic efficiency of the entire machine is also η_p. As far as performance calculations are concerned, either the approach using a polytropic head and efficiency or using isentropic head and efficiency will lead to the same result:

$$P = \dot{m} \cdot \Delta h = \dot{m}\left(\frac{\Delta h_s}{\eta_s}\right) = \dot{m}\left(\frac{\Delta h_p}{\eta_p}\right) \tag{11-21}$$

We also encounter energy conservation on a different level in turbomachines: The aerodynamic function of a turbomachine relies on the capability to trade two forms of energy: kinetic energy (velocity energy) and potential energy (pressure energy), as was outlined earlier.

11.7 REAL GAS BEHAVIOR AND EQUATIONS OF STATE

Understanding gas compression requires an understanding of the relationship between pressure, temperature, and density of a gas. An ideal gas exhibits the following behavior:

$$\frac{p}{\rho} = RT \tag{11-22}$$

where ρ is density of gas, and R is gas constant (as long as the gas composition remains unchanged). Any gas at very low pressures ($p \rightarrow 0$) and high temperatures (ambient and higher) can be described by this equation.

For the elevated pressures we see in natural gas compression, this equation becomes inaccurate, and an additional variable, the compressibility factor (Z), has to be added:

$$\frac{p}{\rho} = ZRT \tag{11-23}$$

Consequently, the compressibility factor itself is a function of pressure, temperature, and gas composition.

A similar situation arises when the enthalpy has to be calculated. For an ideal gas, we find

$$\Delta h = \int_{T_1}^{T_2} C_p(T)\,dT \tag{11-24}$$

where C_p is only a function of temperature. This is a better approximation of the reality than the assumption of a perfect gas used in Equation 11-5.

In a real gas, we get additional terms for the deviation between real gas behavior and ideal gas behavior (Poling et al., 2001):

$$\Delta h = (h^0 - h(p_1))_{T_1} + \int_{T_1}^{T_2} C_\mathrm{p} dT - (h^0 - h(p_2))_{T_2} \tag{11-25}$$

The terms $(h^0 - h(p_1))_{\mathrm{T1}}$ and $(h^0 - H(p_2))_{\mathrm{T2}}$ are called departure functions because they describe the deviation of the real gas behavior from the ideal gas behavior. They relate the enthalpy at some pressure and temperature to a reference state at low pressure but at the same temperature. The departure functions can be calculated solely from an equation of state, while the term $\int C_\mathrm{p} dT$ is evaluated in the ideal gas state.

Equations of state are semiempirical relationships that allow us to calculate the compressibility factor as well as the departure functions. For gas compression applications, the most frequently used equations of state are Redlich–Kwong, Soave–Redlich–Kwong, Benedict–Webb–Rubin, Benedict–Webb–Rubin–Starling and Lee–Kessler–Ploecker (Poling et al., 2001).

Kumar et al. (1999) and Beinecke and Luedtke (1983) compared these equations of state regarding their accuracy for compression applications. In general, all of these equations provide accurate results for typical applications in pipelines, i.e., for gases with a high methane content and at pressures below 3,500 psia.

11.8 COMPRESSION RATIO

Compression ratio (CR) is the ratio of absolute discharge pressure to the absolute suction pressure. Mathematically,

$$\mathrm{CR} = \frac{p_2}{p_1} \tag{11-26}$$

By definition, the compression ratio is always greater than one. If there are n stages of compression and the compression ratio is equal on each stage, then the compression ratio per stage is given by

$$\mathrm{CR}_{\mathrm{stage}} = \left(\frac{p_2}{p_1}\right)^{1/n} \tag{11-27}$$

If the compression ratio is not equal on each stage, then Equation 11-26 should be applied to each stage.

The term "compression ratio" can be applied to a single stage of compression and multistage compression. When applied to a single compressor or a single stage of compression, it is defined as the stage or unit compression ratio; when applied to a multistage compressor, it is defined as the overall compression ratio. The compression ratio for typical gas pipeline compressors is rather low (usually below 2). Low-pressure ratios can be covered in a single compression stage for reciprocating compressor and in a single body (with one or two impellers) in a centrifugal compressor.

While the pressure ratio is a valuable indicator for reciprocating compressors, the pressure ratio that a given centrifugal compressor can achieve depends primarily on gas composition and gas temperature. The centrifugal compressor is better characterized by its capability to achieve a certain amount of head (and a certain amount of head per stage). From Equation 11-9 follows that the compressor head translates into a pressure ratio depending on gas composition and suction temperature. For natural gas (SG = 0.58–0.70), a single centrifugal stage can provide a pressure ratio of 1.4. The same stage would yield a pressure ratio of about 1.6 if it would compress air (SG = 1.0). The pressure ratio per stage is usually lower than the preceding values for multistage machines.

For reciprocating compressors, the pressure ratio per compressor is usually limited by mechanical considerations (rod load) and temperature limitations. Reciprocating compressors can achieve cylinder pressure ratios of 3 to 6. The actual flange-to-flange ratio will be lower (due to the losses in valves and bottles). For lighter gases (i.e., natural gas), the temperature limit will often limit the pressure ratio before the mechanical limits do. Centrifugal compressors are also limited by mechanical considerations (rotordynamics, maximum speed) and temperature limits. Whenever any limitation is involved, it becomes necessary to use multiple compression stages in series and intercooling. Furthermore, multistage compression may be required from a purely optimization standpoint. For example, with increasing compression ratio, compression efficiency decreases, and mechanical stress and temperature problems become more severe. For reference, if we assume natural gas at 100°F suction temperature, a pressure ratio of 3 will—depending on the compressor efficiency—lead to about 275°F discharge temperature.

For pressure ratios higher than 3, it may be advantageous to install intercoolers between the compressors. Intercoolers are generally used between the stages to reduce the power requirements as well as to lower the gas temperature that may become undesirably high.[3] Theoretically, minimum

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[3] After the cooling, liquids may form. These liquids are removed in interstage scrubbers or knockout drums.

power requirement is obtained with ideal intercooling and no pressure loss between stages by making the ratio of compression the same in all stages. However, intercoolers invariably cause pressure losses (typically between 5 and 15 psi), which is a function of the cooler design. For preliminary design considerations, a value of 10 psi can be used (coolers, especially gas-to-air coolers for lower pressure drop, tend to become expensive).

Note that an actual compressor with an infinite number of compression stages and intercoolers would approach isothermal conditions (where the power requirement of compression cycle is the absolutely minimum power necessary to compress the gas) if the gas were cooled to the initial temperature in the intercoolers.

Interstage cooling is usually achieved using gas-to-air coolers. The gas outlet temperature depends on the ambient air temperature. The intercooler exit temperature is determined by the cooling media. If ambient air is used, the cooler exit temperature and thus the suction temperature to the second stage will be about 20°F–30°F above ambient dry bulb temperature. Water coolers can achieve exit temperatures about 20°F above the water supply temperature but require a constant supply of cooling water. Cooling towers can provide water supply temperatures of about wet bulb temperature plus 25°F.

For applications where the compressor discharge temperature is above some temperature limit of downstream equipment (a typical example is pipe coatings that limit gas temperatures to about 125°F–140°F), or has to be limited for other reasons (for example, not to disrupt or damage the permafrost), an aftercooler has to be installed.

11.9 COMPRESSION DESIGN

Compressor design involves several steps. They include selection of the correct type of compressor, as well as the number of stages required. In addition, depending on the capacity, there is also a need to determine the horsepower requirement for the compression.

11.9.1 Determining Number of Stages

For reciprocating compressors, the number of stages is determined from the overall compression ratio as follows (Arnold and Stewart, 1999):

1. Calculate the overall compression ratio. If the compression ratio is under 4, consider using one stage. If it is not, select an initial number of stages so that $CR<4$. For initial calculations, it can be assumed that the

compression ratio per stage is equal for each stage. Compression ratios of 6 can be achieved for low-pressure applications, however, at the cost of higher mechanical stress levels and lower volumetric efficiency.

2. Calculate the discharge gas temperature for the first stage. If the discharge temperature is too high (more than 300°F), we will either have to increase the number of stages or reduce the suction temperature through precooling. It is recommended that the compressors be sized so that the discharge temperatures for all stages of compression be below 300°F. It is also suggested that the aerial gas coolers be designed to have a maximum of 20°F approach to ambient, provided the design reduces the suction temperature for the second stage, conserving horsepower and reducing power demand. If the suction gas temperature to each stage cannot be decreased, increase the number of stages by one and recalculate the discharge temperature.

For centrifugal compressors, the number of compressor bodies is determined in a similar fashion:

1. Calculate the overall isentropic head using Equation 11-9 or 11-11. Typically, a centrifugal compressor casing can provide about 70,000 ft.lb/lb of head, due to limitations in speed, number of impellers, and discharge temperature limits (although 100,000 ft.lb/lb might be possible in some cases).
2. If the overall head exceeds 70,000 ft.lb/lb, take the square root of the overall pressure ratio (Equation 11-27) and calculate the isentropic head for the new pressure ratio. The compression train will consist of two compressors.
3. If the required isentropic head still exceeds 70,000 ft.lb/lb, take the cube root of the overall pressure ratio and calculate the isentropic head for the new pressure ratio. The compression train will consist of three compressors. More than three compressors per train are usually not practical, so if the required head per compressor still exceeds 70,000 ft.lb/lb, a second train may have to be considered.
4. For cases 2 and 3, i.e., multiple compressors per train, the head requirement for the second or third body must be recalculated considering about 20°F approach to ambient for gas-air-coolers, and about 10 psi pressure drop. Unless the gas has a very high inlet temperature, or the compressor is low, the 70,000 ft.lb/lb limit will keep discharge temperatures at an acceptable level below 350°F. In any case, the discharge temperature should be checked using Equation 11-10 or 11-15.

The 300°F temperature limit is used for reciprocating compressors because the packing life gets shortened above about 250°F and the lube oil, being directly involved in the compression process, will degrade faster at higher temperatures. The 350°F temperature limit pertains to centrifugals and is really a limit for the seals (although special seals can go from 400°F to 450°F) or the pressure rating of casings and flanges. Since the lube oil in a centrifugal compressor does not come into direct contact with the process gas, lube oil degradation is not a factor.

If oxygen is present in the process gas in the amount that it can support combustion (i.e., the gas-to-oxygen ratio is above the lower explosive limit), much lower gas temperatures than mentioned previously are required. In reciprocating machines, oil-free compression may be required (no lube oil can come into contact with the process gas). This requires special piston designs that can run dry. Also, special precaution has to be taken to avoid hot spots generated by local friction.

11.9.2 Inlet Flow Rate

The compressor capacity is a critical component in determining the suitability of a particular compressor. We can calculate the actual gas flow rate at suction conditions using[4]

$$Q_G = 0.0283 \frac{Z_1 T_1}{p_1} Q_{G,SC} \tag{11-28}$$

where Q_G represents an actual cubic feet per minute flow rate of gas; T_1 represents the suction temperature in °R; p_1 represents suction pressure in psia; and $Q_{G,SC}$ represents the standard volumetric flow rate of gas in MMscfd.

Note that using the value of actual gas volumetric flow rate and discharge pressure, we can roughly determine the type of compressor appropriate for a particular application. Although there is a significant overlap, some of the secondary considerations, such as reliability, availability of maintenance, reputation of vendor, and price, will allow us to choose one of the acceptable compressors.

11.9.3 Compression Power Calculation

Once we have an idea about the type of compressor we will select, we also need to know the power requirements so that an appropriate prime mover can be designed for the job. After the gas horsepower (GHP) has been

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[4] While the sizing of the compressor is driven by the actual volumetric flow rate (Q_G), the flow in many applications is often defined as standard flow. Standard flow is volumetric flow at certain, defined conditions of temperature and pressure (60°F or 519.7°R and 14.696 psia) that are usually not the pressures and temperatures of the gas as it enters the compressor.

determined by either method, horsepower losses due to friction in bearings, seals, and speed increasing gears must be added. Bearings and seal losses can be estimated from Scheel's equation (GPSA, 2004). For reciprocating compressors, the mechanical and internal friction losses can range from about 3% to 8% of the design gas horsepower. For centrifugal compressors, a good estimate is to use 1% to 2% of the design GHP as mechanical loss.

To calculate brake horsepower (BHP), we can use the following equation:

$$\mathrm{BHP} = \mathrm{GHP} + \text{Mechanical losses} \tag{11-29}$$

The detailed calculation of brake horsepower depends on the choice of type of compressor and number of stages. The brake horsepower per stage can be determined from Equation 11-30 (GPSA, 2004):

$$\mathrm{BHP} = 0.0854 \cdot Z_{\mathrm{ave}} \left[\frac{(Q_{\mathrm{G,SC}})(T_1)}{E \cdot \eta}\right] \left[\frac{k}{k-1}\right] \left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right] \tag{11-30}$$

where BHP is brake horsepower per stage; Z_{ave} is average compressibility factor; $Q_{\mathrm{G,SC}}$ is standard volumetric flow rate of gas, MMscfd; T_1 is suction temperature, °R; p_1,p_2 are pressure at suction and discharge flanges, respectively, psia; E is parasitic efficiency (for high-speed reciprocating units, use 0.72 to 0.82; for low-speed reciprocating units, use 0.72 to 0.85; and for centrifugal units, use 0.99); and η is compression efficiency (1.0 for reciprocating and 0.80 to 0.87 for centrifugal units).

In Equation 11-30, the parasitic efficiency (E) accounts for mechanical losses, and the pressure losses incurred in the valves and pulsation dampeners of reciprocating compressors (the lower efficiencies are usually associated with low-pressure ratio applications typical for pipeline compression).[5] Hence, suction and discharge pressures may have to be adjusted for the pressure losses incurred in the pulsation dampeners for reciprocation compressors. The compression efficiency accounts for the actual compression process. For centrifugal compressors, the lower efficiency is usually associated with pressure ratios of 3 and higher. Very low flow compressors (below 1000 acfm) may have lower efficiencies.

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[5] Many calculation procedures for reciprocating compressors use numbers for E that are higher than the ones referenced here. These calculations require, however, that the flange-to-flange pressure ratio (which is used in Equation 11-30) is increased by the pressure losses in the compressor suction and discharge valves and pulsation dampeners. These pressure losses are significant, especially for low head high flow applications.

The total horsepower for the compressor is the sum of the horsepower required for each stage. Reciprocating compressors require an allowance for interstage pressure losses. It can be assumed that there is a 3% loss of pressure in going through the cooler, scrubbers, piping, etc., between the actual discharge of the cylinder and the actual suction of the next cylinder. For a centrifugal compressor, any losses incurred between the stages are already included in the stage efficiency. However, the exit temperature from the previous stage becomes the inlet temperature in the next stage. If multiple bodies are used, the losses for coolers and piping have to be included as described previously.

Example

Given the following information for a centrifugal compressor, answer the following questions.

Operating conditions:

$$P_s = 750 \text{ psia} \quad P_d = 1046.4 \text{ psia} \quad T_s = 529.7\ ^\circ\text{R} \quad T_d = 582.6\ ^\circ\text{R}$$
$$Q_{G,SC} = 349 \text{ MMscfd}$$

Gas properties:

$$\text{SG} = 0.6, k = 1.3, Z_{ave} = 0.95$$

Questions:

1. What is the isentropic efficiency?
2. What is the actual volumetric flow rate?
3. What is the isentropic head?
4. What is the power requirement (assume a 98% mechanical efficiency)?

Solution:

1. With rearranging Equation 11-15, we find

$$\eta_s = \frac{T_1}{T_2 - T_1}\left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right]$$

$$\eta_s = \frac{529.7}{582.6 - 529.7}\left[\left(\frac{1046.4}{750}\right)^{\frac{0.3}{1.3}} - 1\right] = 0.80$$

2. Mass flow is

$$\dot{m} = 0.884 \times Q_{G,SC} \times \text{SG} = 0.884\frac{\text{lbm}}{\text{MMSCFD}}$$
$$\times\ 349 \text{ MMSCFD} \times 0.6 = 185.1 \text{ lb}_m/\text{s}$$

and thus the volumetric flow becomes

$$Q_1 = \frac{53.35}{144} \cdot \frac{\dot{m} \cdot Z_1 \cdot T_1}{p_1 \cdot SG} = \frac{53.35}{144} \cdot \frac{185.1 \cdot 0.95 \cdot 529.7}{750 \cdot 0.6}$$
$$= 76.68 \text{ cfs} = 4601 \text{cfm}$$

3. The isentropic head follows from Equation 11-9 with $C_p = (53.35/SG)$ $Z\,k/(k\text{-}1)$

$$\Delta h_s = \frac{53.35}{0.6} \times 0.95 \times \frac{1.3}{0.3} \times 529.7 \left[\left(\frac{1046.4}{750} \right)^{\frac{0.3}{1.3}} - 1 \right]$$
$$= 14482 \text{ ft.lb}_f/\text{lb}_m$$

4. The power can be calculated from Equation 11-30:

$$\text{BHP} = 0.0854 \times 0.95 \left[\frac{349 \cdot 529.7}{0.98 \cdot 0.8} \right] \left[\frac{1.3}{0.3} \right] \left[\left(\frac{1046.4}{750} \right)^{\frac{0.3}{1.3}} - 1 \right]$$
$$= 6622 \text{ hp}$$

11.10 COMPRESSOR CONTROL

To a large extent, the compressor operating point will be the result of the pressure conditions imposed by the system. However, the pressures imposed by the system may, in turn, be dependent on the flow. Only if the conditions fall outside the operating limits of the compressor (e.g., frame loads, discharge temperature, available driver power, surge, choke, speed), control mechanisms have to be in place. On the other hand, the compressor output may have to be controlled to match the system demand. The type of application often determines the system behavior. In a pipeline application, suction and discharge pressure are connected with the flow by the fact that the more flow is pushed through a pipeline, the more pressure ratio is required at the compressor station to compensate for the pipeline pressure losses. In process-related applications, the suction pressure may be fixed by a back pressure controlled production separator. In boost applications, the discharge pressure is determined by the pressure level of the pipeline the compressor feeds into, while the suction pressure is fixed by the process. In oil and gas field applications, the suction pressure may depend on the flow because the more gas is moved out of the gas reservoir, the lower the suction pressure has to be. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure, or speed may have to be controlled. The type of control also depends on the compressor driver. Both reciprocating compressors and centrifugal compressors can be controlled by suction throttling or recirculating of gas. However, either method is very inefficient for process control (but may be used to protect the compressor) because the reduction in flow or head is not accompanied by a significant reduction in the power requirement.

11.10.1 Reciprocating Compressors

The following mechanisms may be used to control the capacity of reciprocating compressors: suction pressure, variation of clearance, speed, valve unloading, and recycle. Reciprocating compressors tend to have a rather steep head versus flow characteristic. This means that changes in pressure ratio have a very small effect on the actual flow through the machine.

Reciprocating compressors generate flow pulsations in the suction and discharge lines that have to be controlled to prevent over- and underloading of the compressors, avoid vibration problems in the piping or other machinery at the station, and provide a smooth flow of gas. The flow pulsations can be greatly reduced through the use of properly sized pulsation bottles or pulsation dampeners in the suction and discharge lines.

The flow through the compressor can be controlled by varying the operating speed of the compressor. This method can be used if the compressor is driven by an internal combustion engine or a variable speed electric motor. Especially internal combustion engines but also variable speed electric motors produce less power if they operate at a speed different from their optimum speed. Internal combustion engines allow for speed control in the range of 70% to 100% of maximum speed. Since reciprocating compressors generate pulsations in the piping system, pulsation dampeners have to be installed. Since these have to be optimized based on the frequencies of the pulsations, and these frequencies are speed dependent, the allowable speed range may be limited.

If the driver is a constant speed electric motor, the capacity control consists of either inlet valve unloaders or clearance unloaders. Inlet valve unloaders can hold open the inlet valve into the compressor and thereby prevent compression. Clearance unloaders consist of pockets that are opened when unloading is desired. The gas is compressed into them at the compression stroke and expands back into the cylinder on the return stroke, thus reducing the intake of additional gas and, subsequently, the compressor capacity. Additional flexibility is achieved by using several steps of clearance

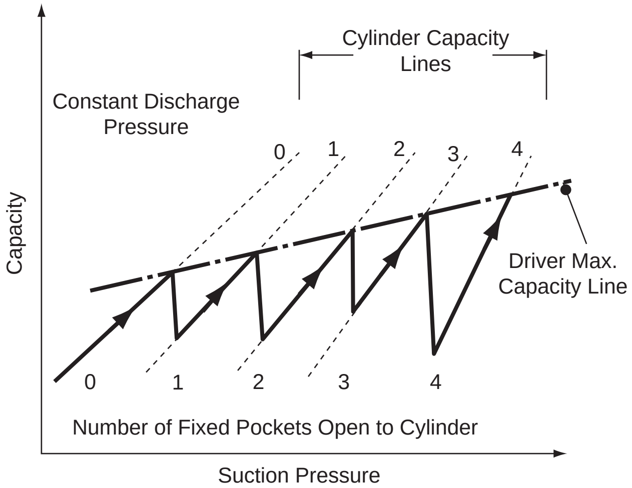

Figure 11-4 Control characteristic of a reciprocating compressor with constant speed driver and pockets (GPSA, 2004).

control and combinations of clearance control and inlet valve control. Figure 11-4 shows the control characteristic of such a compressor.

11.10.2 Centrifugal Compressors

As with reciprocating compressors, the compressor output must be controlled to match the system demand. The operation may require constant flow despite changes in suction or discharge pressure. Compressor flow, pressure, or speed may have to be controlled. The type of control also depends on the compressor driver. Centrifugal compressors tend to have a rather flat head versus flow characteristic. This means that changes in pressure ratio have significant effect on the actual flow through the machine.

Compressor control is usually accomplished by speed control, variable guide vanes, suction throttling, and recycling of gas. Only in rare cases are adjustable diffuser vanes used. To protect the compressor from surge, recycling is used. The flow through the compressor can be controlled by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 50% to 100% of maximum speed or more). Virtually any centrifugal compressor installed since the early 1990s in pipeline service is driven by variable speed drivers, usually a two-shaft gas turbine or a variable speed electric motor. Older installations and installations in other than pipeline services sometimes use single-shaft gas turbines (which allow a speed variation from about 90% to 100% speed) and constant speed electric motors. In these installations, suction throttling or variable inlet guide vanes are used to provide means of control.

The operating envelope of a centrifugal compressor is limited by the maximum allowable speed, the minimum allowable speed, the minimum flow (surge flow), and the maximum flow (choke or stonewall); see Figure 11-5. Another limiting factor may be the available driver power. Only the minimum flow requires special attention because it is defined by an aerodynamic stability limit of the compressor. Crossing this limit to lower flows will cause pulsating intermittent flow reversals in the compressor (surge), which eventually can damage the compressor. Modern control systems can detect this situation and shut down the machine or prevent it entirely by automatically opening a recycle valve. For this reason, virtually all modern compressor installations use a recycle line (see Figure 11-6) with a control valve that allows the flow to increase through the compressor if it comes near the stability limit. Modern control systems constantly monitor the operating point of the compressor in relation to its surge line and automatically open or close the recycle valve if necessary. The control system is designed to compare the measured operating point of the compressor with

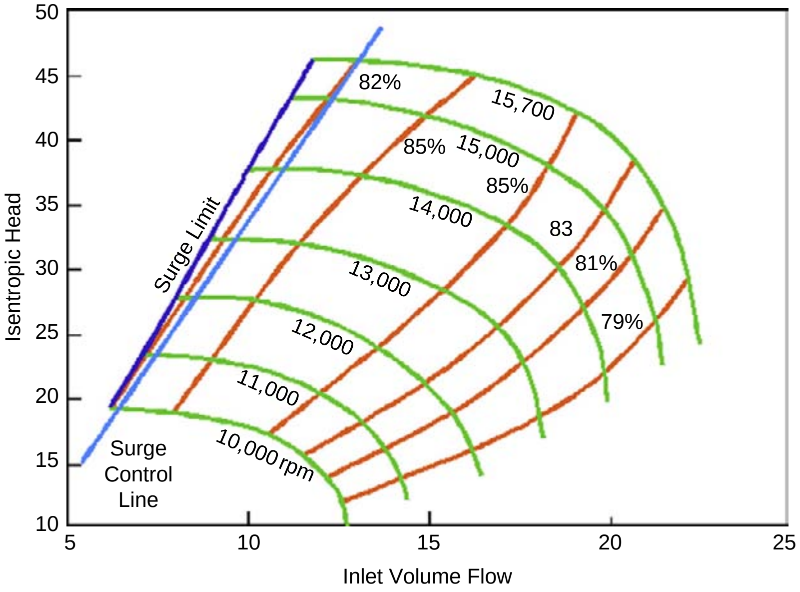

Figure 11-5 Typical performance map for a centrifugal compressor (Kurz and Brun, 2001).

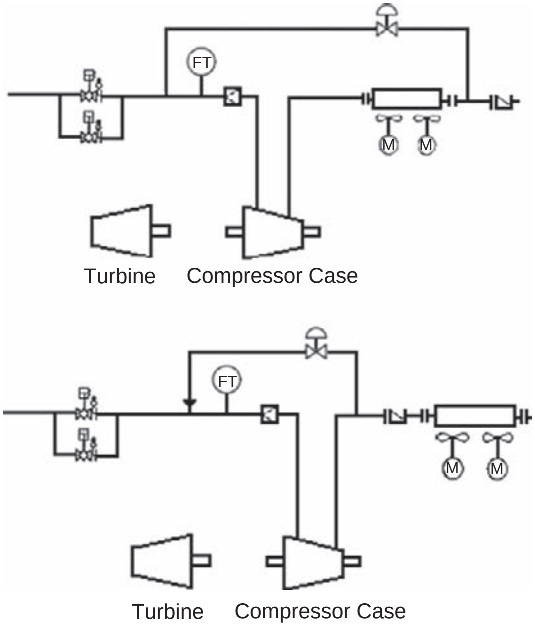

Figure 11-6 Recycle lines for centrifugal compressors: cooled (top) and hot (bottom) recycle lines. *(Courtesy of Solar Turbines Inc.)*

the position of the surge line (refer to Figure 11-5). To that end, flow, suction pressure, discharge pressure, and suction temperature, as well as compressor speed have to be measured.

11.11 COMPRESSOR PERFORMANCE MAPS

11.11.1 Reciprocating Compressors

Figure 11-7 shows some typical performance maps for reciprocating compressors. Since the operating limitations of a reciprocating compressor are often defined by mechanical limits (especially maximum rod load), and the pressure ratio of the machine is very insensitive to changes in suction conditions and gas composition, we usually find maps depicting suction and discharge pressures and actual flow. Maps account for the effect of opening or closing pockets and for variations in speed.

11.11.2 Centrifugal Compressors

For a centrifugal compressor, the isentropic or polytropic head (rather than the pressure ratio) is relatively invariant with the change in suction conditions and gas composition. As with the reciprocating compressor, the flow

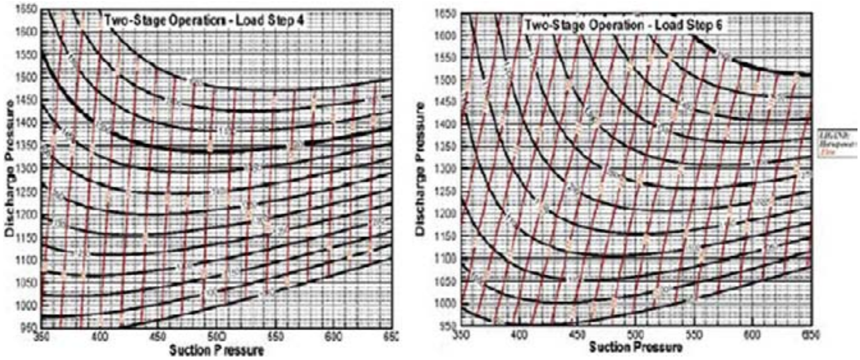

Figure 11-7 Typical maps for a speed-controlled reciprocating compressor for two different load steps. *(Courtesy of Solar Turbines Inc.)*

that determines the operating point is the actual flow as opposed to mass flow or standard flow.

Head versus actual flow maps (refer to Figure 11-5) are therefore the usual way to describe the operating range of a centrifugal compressor. These maps change very little even if the inlet conditions or the gas composition changes. They depict the effect of changing the operating speed and define the operating limits of the compressor such as surge limit, maximum and minimum speed, and maximum flow at choke conditions. Every set of operating conditions, given as suction pressure, discharge pressure, suction temperature, flow, and gas composition, can be converted into isentropic head and actual flow, using the relationships described previously. Once the operating point is located on a head-flow map, the efficiency of the compressor, the required operating speed as well as the surge margin can be determined.

11.12 EXAMPLE FOR OPERATING A COMPRESSOR IN A PIPELINE SYSTEM

To illustrate how the considerations regarding operating characteristics, system behavior, and control work together, we can look into the operation of a typical compressor station in a pipeline (Kurz et al., 2010).

For a situation in which a compressor operates in a system with pipe of the length L_u upstream and a pipe of the length L_d downstream, and further where the pressure at the beginning of the upstream pipe, p_u, and the end of the downstream pipe, p_e, are known and constant, we have a simple model of a compressor station operating in a pipeline system (see Figure 11-8).

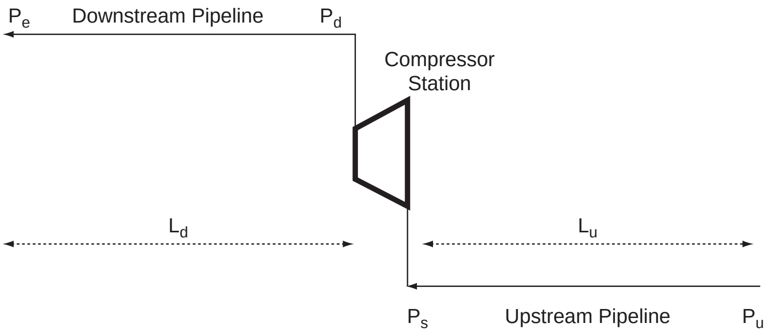

Figure 11-8 Conceptual model of a pipeline segment.

For a given, constant flow capacity, Q_{std}, the pipeline will then impose a pressure p_s at the suction and p_d at the discharge side of the compressor. For a given pipeline, the head (H_s) – flow (Q) relationship at the compressor station can be approximated by

$$H_s = C_p T_s \left[\left(\sqrt{\frac{1}{1 - \frac{C_3 + C_4 \cdot Q^2}{p_d^2}}} \right)^{\frac{k-1}{k}} - 1 \right] \tag{11-31}$$

where C_3 and C_4 are constants (for a given pipeline geometry) describing the pressure at either ends of the pipeline and the friction losses, respectively.

Among other issues, this means that for a compressor station within a pipeline system, the head for a required flow is prescribed by the pipeline system (see Figure 11-9). In particular, this characteristic requires the capability for the compressors to allow a reduction in head with reduced flow, and vice versa, in a prescribed fashion. The pipeline will therefore not require a change in flow at constant head (or pressure ratio).

In transient situations (for example, during line packing), the operating conditions initially follow a constant power distribution; i.e., the head-flow relationship follows Equation 11-32 and will asymptotically approach the steady-state relationship. This is universally valid for speed-controlled centrifugal compressors. For reciprocating compressors, there might be additional constraints due to rod load and rod load reversal concerns, limitations in clearance control, and others.

$$\begin{gathered} P = \dot{m} \cdot \frac{H_s}{\eta_s} = \mathrm{const} \\ \mathrm{H}_s = \frac{\eta_s \cdot \mathrm{const}}{\rho} \frac{1}{Q} \end{gathered} \tag{11-32}$$

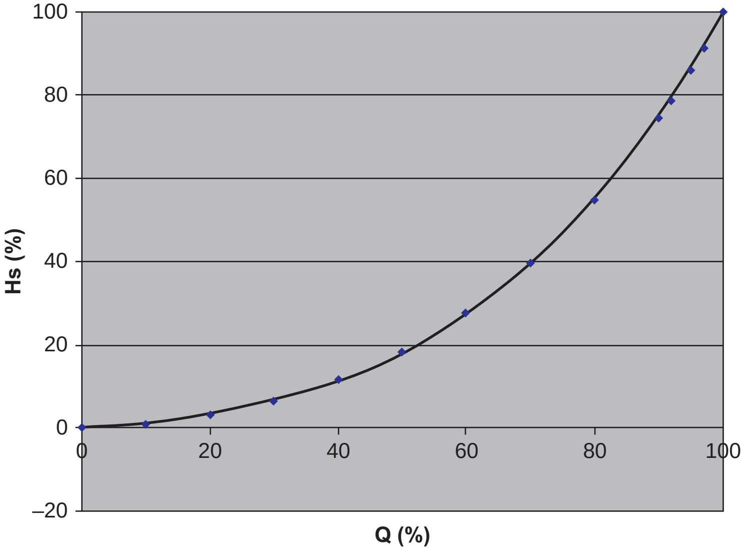

Figure 11-9 Station head-flow relationship based on Equation 11-31.

Based on the preceding requirements, the compressor output must be controlled to match the system demand. This system demand is characterized by a strong relationship between system flow and system head or pressure ratio (refer to Figure 11-9). Given the large variations in operating conditions experienced by pipeline compressors, important questions are how to adjust the compressor to the varying conditions and, in particular, how this influences the efficiency.

Centrifugal compressors tend to have rather flat head versus flow characteristic. This means that changes in pressure ratio have a significant effect on the actual flow through the machine. For a centrifugal compressor operating at a constant speed, the head or pressure ratio is reduced with increasing flow. The flow through the compressor can be controlled by varying the operating speed of the compressor. This is the preferred method of controlling centrifugal compressors. Two shaft gas turbines and variable speed electric motors allow for speed variations over a wide range (usually from 40% or 50% to 100% of maximum speed or more). It should be noted that the controlled value is usually not speed, but the speed is indirectly the result of balancing the power generated by the power turbine (which is controlled by the fuel flow into the gas turbine) and the absorbed power of the compressor.

The operating envelope of a centrifugal compressor is limited by the maximum allowable speed, the minimum flow (surge flow), and the

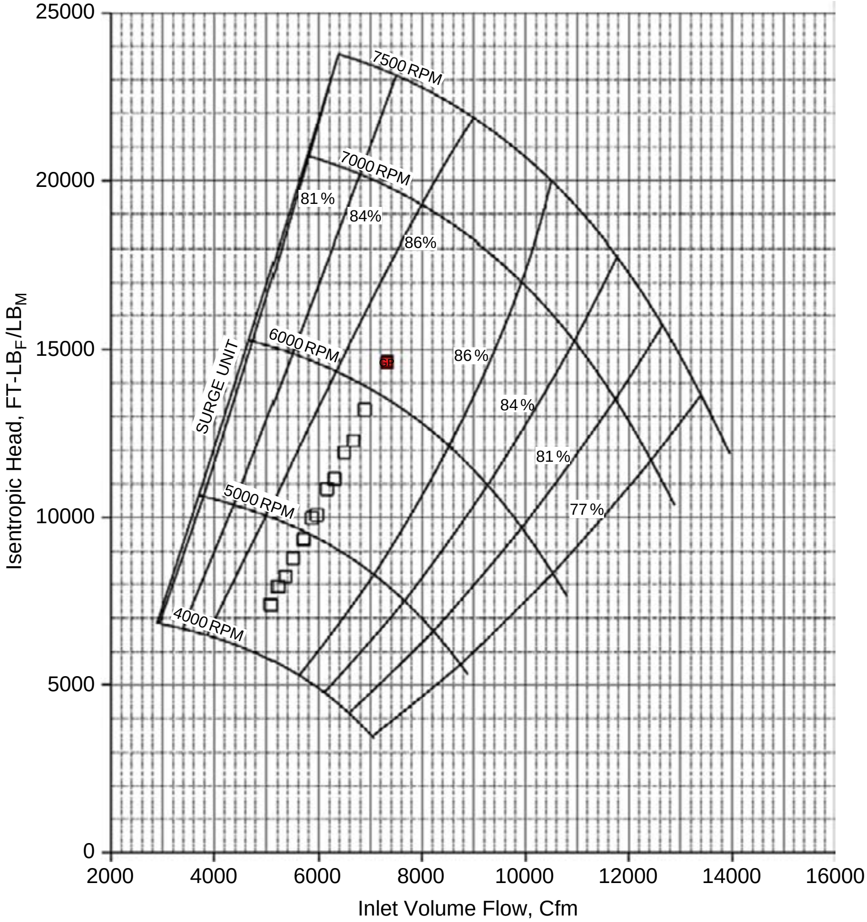

Figure 11-10 Typical pipeline operating points plotted into a typical centrifugal compressor performance map.

maximum flow (choke or stonewall); see Figure 11-10. Another limiting factor may be the available driver power. As mentioned earlier, only the minimum flow requires special attention because it is defined by an aerodynamic stability limit of the compressor.

Assuming the pipeline characteristic derived in Equation 11-31, the compressor impellers will be selected to operate at or near their best efficiency for the entire range of head and flow conditions imposed by the pipeline. This is possible with a speed-controlled compressor because the best efficiency points of a compressor are connected by a relationship that requires approximately (fan law)

$$H_s/N^2 = C_5 \qquad Q/N = C_6$$

$$H_s = Q^2 \cdot \frac{C_5}{{C_6}^2} \tag{11-33}$$

For operating points that meet the preceding relationship, the absorbed gas power P_g is (due to the fact that the efficiency stays approximately constant)

$$P_g = C_7 \cdot H_s \cdot Q = C_7 \cdot \frac{C_5}{{C_6}^2} \cdot Q^3 = C_5 \cdot C_6 \cdot C_7 \cdot N^3 \tag{11-34}$$

As it is, this power-speed relationship allows the power turbine to operate at or very close to its optimum speed for the entire range. The typical operating scenarios in pipelines therefore allow the compressor and the power turbine to operate at their best efficiency for most of the time. The gas producer of the gas turbine will, however, lose some thermal efficiency when operated in part load.

Figure 11-10 shows a typical real-world example: pipeline operating points for different flow requirements are plotted into the performance map of the speed-controlled centrifugal compressor used in the compressor station.

Reciprocating compressors will automatically comply with the system pressure ratio demands, as long as no mechanical limits (rod load, power) are exceeded. Changes in system suction or discharge pressure will simply cause the valves to open earlier or later. The head is lowered automatically because the valves see lower pipeline pressures on the discharge side and/or higher pipeline pressures on the suction side. Therefore, without additional measures, the flow would stay roughly the same—except for the impact of changed volumetric efficiency that would increase, thus increasing the flow with reduced pressure ratio. The control challenge lies in the adjustment of the flow to the system demands. Without additional adjustments, the flow throughput of the compressor changes very little with changed pressure ratio. Historically, pipelines installed many small compressors and adjusted flow rate by changing the number of machines activated. This capacity and load could be fine-tuned by speed or by a number of small adjustments (load steps) made in the cylinder clearance of a single unit. As compressors have grown, the burden for capacity control has shifted to the individual compressors.

Load control is a critical component to compressor operation. From a pipeline operation perspective, variation in station flow is required to meet pipeline delivery commitments, as well as implement company strategies for optimal operation (i.e., line packing, load anticipation). From a unit perspective, load control involves reducing unit flow (through unloaders or speed) to operate as close as possible to the design torque limit without overloading the compressor or driver. Critical limits on any load map curve are rod load limits and HP/torque limits for any given station suction and discharge pressure. Gas control generally will establish the units within a station that must be operated to achieve pipeline flow targets. Local unit control will establish load step or speed requirements to limit rod loads or achieve torque control.

The common methods of changing flow rate are to change speed, change clearance, or deactivate a cylinder end (hold the suction valve open). Another method is an infinite-step unloader, which delays suction valve closure to reduce volumetric efficiency. Further, part of the flow can be recycled or the suction pressure can be throttled, thus reducing the mass flow while keeping the volumetric flow into the compressor approximately constant.

Control strategies for compressors should allow automation and be adjusted easily during the operation of the compressor. In particular, strategies that require design modifications to the compressor (for example, rewheeling of a centrifugal compressor, changing cylinder bore, or adding fixed clearances for a reciprocating compressor) are not considered here. It should be noted that with reciprocating compressors, a key control requirement is not to overload the driver or to exceed mechanical limits.

As mentioned previously, we are describing the situation for a pipeline operating at or near steady-state conditions. For conditions where the compressor operates at low load in steady state, and the power supply is increased instantly, the compressor operating point will move to an increased flow, without initially seeing a change in pressure ratio (because the gas volume in the pipeline has to be increased first). Then the pressure ratio dictated by the pipeline will cause the compressor operating point to drift back to the steady-state line, following a line of more or less constant absorbed power (see Equation 11-32).

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CHAPTER 12

Sales Gas Transmission

12.1 INTRODUCTION

Natural gas continues to play a great role as a worldwide energy supply. In fact, major projects are being planned to move massive amounts of high-pressure sales gas from processing plants to distribution systems and large industrial users through large-diameter buried pipelines. These pipelines utilize a series of compressor stations along the pipeline to move the gas over long distances. In addition, gas coolers are used downstream of the compressor stations to maintain a specified temperature of the compressed gas for pipeline pressure drop reduction and to protect gas pipeline internal and external coating against deterioration due to high temperatures. This chapter covers all the important concepts of sales gas transmission from a fundamental perspective.

12.2 GAS FLOW FUNDAMENTALS

Optimum design of a gas transmission pipeline requires accurate methods for predicting pressure drop for a given flow rate or predicting flow rate for a specified pressure drop in conjunction with installed compression power and energy requirements, e.g., fuel gas, as part of a technical and economic evaluation. In other words, there is a need for using analytical methods to relate the flow of gas through a pipeline to the properties of both the pipeline and gas, and to the operating conditions such as pressure and temperature. Isothermal steady-state pressure drop or flow-rate calculation methods for single-phase dry gas pipelines are the most widely used and the most basic relationships in the engineering of gas delivery systems (Beggs, 1984; Smith, 1990; Aziz and Ouyang, 1995). They also form the basis of other more complex transient flow calculations and network designs.

12.2.1 General Flow Equation

Based on the assumptions that there is no elevation change in the pipeline and that the condition of flow is isothermal, the integrated

Bernoulli's equation is expressed by Equation 12-1 (Uhl, 1965; Schroeder, 2001):

$$Q_{sc} = C\left(\frac{T_b}{P_b}\right)D^{2.5}\left(\frac{P_1^2 - P_2^2}{f\,\gamma_G T_a Z_a L}\right)^{0.5} E \tag{12-1}$$

where Q_{sc} is standard gas flow rate measured at base temperature and pressure, ft^3/day; T_b is gas temperature at base condition, 519.6°R; P_b is gas pressure at base condition, 14.7 psia; P_1 is inlet-gas pressure, psia; P_2 is outlet-gas pressure, psia; D is inside diameter of pipe, inches; f is Moody friction factor; E is flow efficiency factor; γ_G is gas-specific gravity; T_a is average absolute temperature of pipeline, °R; Z_a is average compressibility factor; L is pipe length, miles; and C is 77.54 (a constant for the specific units used).

Although the assumptions used to develop Equation 12-1 are usually satisfactory for a long pipeline, the equation contains an efficiency factor, E, to correct for these assumptions. Most experts recommend using efficiency factor values close to unity when dry gas flows through a new pipeline. However, as the pipeline ages and is subjected to varying degrees of corrosion, this factor will decrease (Campbell et al., 1992). In practice, and even for single-phase gas flow, some water or condensate may be present if the necessary drying procedure for gas pipeline commissioning is not adopted or scrubbers are not installed. This puts compression equipment at risk of damage and also allows localized corrosion due to water spots (wetting of the pipe surface). The presence of liquid products in the gas transmission lines can also cause drastic reduction in the flow efficiency factor. Typically, efficiency factors may vary between 0.6 and 0.92 depending on the pipelines' liquid contents (Ikoku, 1984). As the amount of liquid content in the gas phase increases, the pipeline efficiency factor can no longer account for the two-phase flow behavior, so two-phase flow equations must be used (Brill and Beggs, 1991; Asante, 2002).

Pipelines are usually not horizontal; however, as long as the slope is not too great, a correction for the static head of fluid (H_c) may be incorporated into Equation 12-1 as follows (Schroeder, 2001):

$$Q_{sc} = C\left(\frac{T_b}{P_b}\right)D^{2.5}\left(\frac{P_1^2 - P_2^2 - H_c}{L\,\gamma_G T_a Z_a f}\right)^{0.5} E \tag{12-2}$$

where

$$H_c = \frac{0.0375\,g\,(H_2 - H_1)\,P_a^2}{Z_a T_a} \tag{12-3}$$

and H_1 is inlet elevation, ft; H_2 is outlet elevation, ft; and g is gravitational constant, ft/sec^2.

The average compressibility factor, Z_a, is determined from the average pressure (P_a) and average temperature (T_a), where P_a is calculated from Equation 12-4 (Campbell et al., 1992):

$$P_a = \frac{2}{3}\left[(P_1 + P_2) - \left(\frac{P_1 P_2}{P_1 + P_2}\right)\right] \tag{12-4}$$

where P_1 and P_2 are the upstream and downstream absolute pressures, respectively. The average temperature is determined by Equation 12-5:

$$T_a = \left[\frac{T_1 - T_2}{\ln\left(\frac{T_1 - T_S}{T_2 - T_S}\right)}\right] + T_S \tag{12-5}$$

In the preceding equation, parameter T_S is the soil temperature, and T_1 and T_2 are the upstream and downstream temperatures, respectively.

Having obtained P_a and T_a for the gas, one can obtain the average compressibility factor using Kay's rule and gas compressibility factor charts (Campbell et al., 1992).

12.2.2 Friction Factor Correlations

The fundamental flow equation for calculating pressure drop requires a numerical value for the friction factor. However, because the friction factor, f, is a function of flow rate, the whole flow equation becomes implicit. To determine the friction factor, one characterizes the fluid flow by a dimensionless value known as the Reynolds number (Equation 12-6):

$$N_{Re} = \frac{\rho V D}{\mu} \tag{12-6}$$

where N_{Re} is Reynolds number, dimensionless; D is pipe diameter, ft; V is fluid velocity, ft/sec; ρ is fluid density, lb_m/ft^3; and μ is fluid viscosity, lb_m/ft.sec.

For Reynolds numbers less than 2,000, the flow is considered laminar. When the Reynolds number exceeds 2,000, the flow is characterized as turbulent. Note that in high-pressure gas transmission pipelines with moderate to high flow rates, only two types of flow regimes are observed: partially turbulent flow (smooth pipe flow) and fully turbulent flow (rough pipe flow). For gases, the Reynolds number is given by Equation 12-7 (Kennedy, 1993):

$$N_{Re} = \frac{0.7105\, P_b \gamma_G Q_{sc}}{T_b \mu_G D} \tag{12-7}$$

where D is pipe diameter, inches; Q_{sc} is gas flow rate, standard ft^3/day; μ_G is gas viscosity, cp; P_b is base pressure, psia; T_b is base temperature, °R; and γ_G is gas-specific gravity, dimensionless.

For the gas industry, Equation 12-7 is a more convenient way to express the Reynolds number since it displays the value proportionally in terms of the gas flow rate.

The other parameter in the friction factor correlation is pipe roughness (ε), which is often correlated as a function of the Reynolds number and the pipe relative roughness (absolute roughness divided by inside diameter). Pipe roughness varies considerably from pipe to pipe; Table 12-1 shows the roughness for various types of new (clean) pipes. These values should be increased by a factor ranging between 2 and 4 to account for age and use.

The Moody (1944) friction factor, f, in Equation 12-1 is determined from the Moody diagram. The Moody correlation is shown in Figure 12-1. The Moody diagram consists of four zones: laminar, transition, partially turbulent, and fully turbulent zones.

The laminar zone, the left side, is the zone of extremely low flow rate in which the fluid flows strictly in one direction and the friction factor shows a sharp dependency on flow rate. The friction factor in the laminar regime is defined by the Hagen–Poiseuille equation (Streeter and Wylie, 1979):

$$f = \frac{64}{N_{Re}} \tag{12-8}$$

The fully turbulent zone, the right side, describes fluid flow that is completely turbulent (back mixing) laterally as well as in the primary direction. The turbulent friction factor shows no dependency on flow rate and is only a function of pipe roughness because an ideally smooth pipe never really exists in this zone. The friction factor to use is given by the rough pipe law of Nikuradse (1933):

$$\frac{1}{\sqrt{f}} = 2\,\text{Log}\,\frac{D}{\varepsilon} + 1.14 \tag{12-9}$$

Table 12-1 Pipe Roughness Value (Norsok Standard, 1996)

Type of Pipe (New, Clean Condition)	**ε (Inches)**
Carbon steel corroded	0.019685
Carbon steel noncorroded	0.001968
Glass fiber reinforced pipe (GRP)	0.0007874
Steel internally coated with epoxy	0.00018 to 0.00035

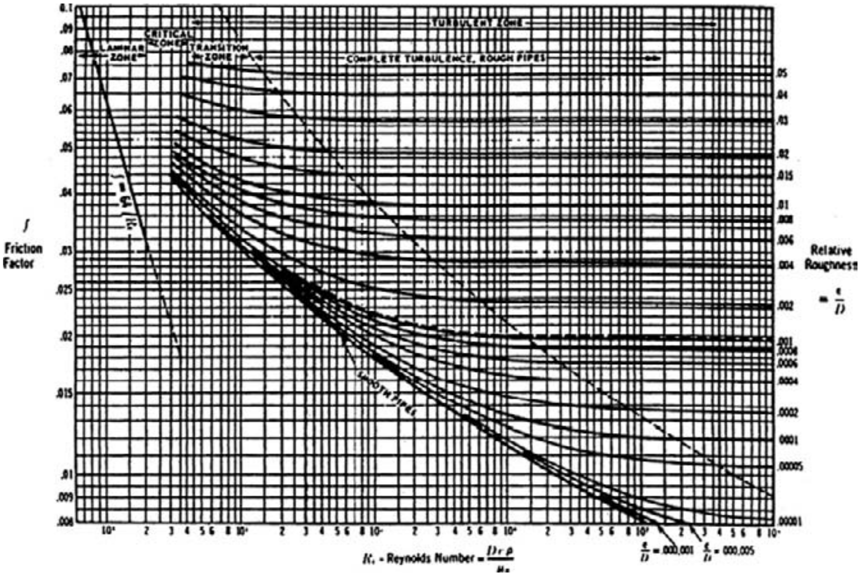

Figure 12-1 Moody friction factor diagram (Streeter and Wylie, 1979).

Equation 12-9 shows that if the roughness of the pipeline is increased, the friction factor increases and results in higher-pressure drops. Conversely, if the pipe roughness is decreased, lower friction factor or lower pressure drops are obtained. Note that most pipes cannot be considered ideally smooth at high Reynolds numbers (Schlichting, 1979); therefore, the investigations of Nikuradse (1933) on flow through rough pipes has been of significant interest to engineers.

The partially turbulent (transition) zone is the zone of moderately high flow rate in which the fluid flows laterally within the pipe as well as in the primary direction, although some laminar boundary layer outside the zone of roughness still exists. Partially turbulent flow is governed by the smooth pipe law of Karman and Prandtl (Uhl, 1965):

$$\frac{1}{\sqrt{f}} = 2\ \mathrm{Log}\left(N_{\mathrm{Re}}\sqrt{f}\right) - 0.8 \qquad (12\text{-}10)$$

The correlation in Equation 12-10 has received wide acceptance as a true representative of experimental results. However, a study by Zagarola (1996) on the flow at high Reynolds numbers in smooth pipes showed that the relevant correlation was not accurate for high Reynolds numbers, where the correlation was shown to predict too low values of the friction factor.

Consensus on how the friction factor varies across the transition region from an ideal smooth pipe to a rough pipe has not been reached. However, Colebrook (1939) presented additional experimental results and developed a correlation for the friction factor valid in the transition region between smooth and rough flow. The correlation is as follows:

$$\frac{1}{\sqrt{f}} = -2\ \mathrm{Log}\left(\frac{\varepsilon/D}{3.7} + \frac{2.51}{N_{\mathrm{Re}}\sqrt{f}}\right) \tag{12-11}$$

Equation 12-11 is universally accepted as standard for computing the friction factor of rough pipes and is incorporated in most thermohydraulic simulation software available on the market. Moody (1944) concluded that the Colebrook (1939) equation was adequate for friction factor calculations for any Reynolds number greater than 2,000. Certainly, the accuracy of the equation was well within the experimental error (about ± 5% for smooth pipes and ±10% for rough pipes).

The friction factor is sometimes expressed in terms of the Fanning friction factor, which is one-fourth of the Moody friction factor. Care should be taken to avoid inadvertent use of the wrong friction factor.

12.2.3 Simplified Flow Equations

The Moody friction factor, f, is an integral part of the general gas flow equation. Since it is a highly nonlinear function, it must be either read from a chart or determined iteratively from a nonlinear equation. Approximations to the Moody friction factor have been widely used because they allow the gas flow equation to be solved directly instead of iteratively. The four most widely published friction factor approximations are Weymouth, Panhandle A, Panhandle B, and IGT (Beggs, 1984; Ikoku, 1984). The Weymouth equation approximates the Moody friction factor using Equation 12-12, and the remaining three equations approximate the friction factor using Equation 12-13, where "m" and "n" are constants. These constants are given in Table 12-2.

$$f = \mathrm{m}\ (D)^{-\mathrm{n}} \tag{12-12}$$

$$f = \mathrm{m}\ (N_{\mathrm{Re}})^{-\mathrm{n}} \tag{12-13}$$

The Reynolds number, N_{Re}, can be approximated using Equation 12-7. In addition to the Reynolds number, the pipe roughness also affects the friction factor for turbulent flow in rough pipes. Hence, the efficiency factor is chosen to correctly account for pipe roughness (Ouyang and Aziz, 1996).

Table 12-2 Constants in Equations 12-12 and 12-13

Equation	m	n
Weymouth	0.032	0.333
Panhandle A	0.085	0.147
Panhandle B	0.015	0.039
IGT	0.187	0.200

These approximations can then be substituted into the flow equation for *f*, and the resulting equation is given using Equation 12-14 (Kennedy, 1993; Towler and Pope, 1994):

$$Q_{sc} = \frac{a_1 \left(\frac{T_b}{P_b}\right) E \left(P_1^2 - P_2^2\right)^{0.5} D^{a_2}}{(\gamma_G)^{a_3} (T_a Z_a L)^{a_4} (\mu_G)^{a_5}} \tag{12-14}$$

In Equation 12-14 the values for a_1 through a_5 are constants that are functions of the friction factor approximations and the gas flow equation. These constants are given in Table 12-3.

Inspection of Table 12-3 shows that the gas flow rate is not a strong function of the gas viscosity at high Reynolds numbers because viscosity is of importance in laminar flow, and gas pipelines are normally operated in the partially/fully turbulent flow region. However, under normal conditions, the viscosity term has little effect because a 30% change in absolute value of the viscosity will result in only approximately a 2.7% change in the computed quantity of gas flowing. Thus, once the gas viscosity is determined for an operating pipeline, small variations from the conditions under which it was determined will have little effect on the flow predicted by Equation 12-14 (Huntington, 1950).

Note that all of the equations noted previously were developed from the fundamental gas flow equation; however, each has a special approximation of the friction factor to allow for an analytical solution. For instance, the Weymouth (1912) equation uses a straight line for *f*, and thus, its approximation has been shown to be a poor estimation for the friction factor for

Table 12-3 Constants in Equation 12-14

Equation	a_1	a_2	a_3	a_4	a_5
Weymouth	433.46	2.667	0.5000	0.5000	0.0000
Panhandle A	403.09	2.619	0.4603	0.5397	0.0793
Panhandle B	715.35	2.530	0.4900	0.5100	0.0200
IGT	307.26	2.667	0.4444	0.5556	0.1111

most flow conditions (Kennedy, 1993). This equation tends to overpredict the pressure drop and thus provides a poor estimate relative to the other gas flow equations. The Weymouth equation, however, is of use in designing gas distribution systems in that there is an inherent safety in overpredicting pressure drop (Maddox and Erbar, 1982). In practice, the Panhandle equations are commonly used for large-diameter, long pipelines where the Reynolds number is on the flat portion of the Moody diagram. The Panhandle A equation is most applicable for medium- to relatively large-diameter pipelines (12–60-inch diameter) with moderate gas flow rate, operating under medium to high pressure (800–1,500 psia). The Panhandle B equation is normally appropriate for high flow rate, large-diameter (>36 inches), and high-pressure (> 1,000 psi) transmission pipelines (Maddox and Erbar, 1982; Kennedy, 1993).

Since friction factors vary over a wide range of Reynolds number and pipe roughness, none of the gas flow equations is universally applicable. However, in most cases, pipeline operators customize the flow equations to their particular pipelines by taking measurements of flow, pressure, and temperature and back-calculating pipeline efficiency or an effective pipe roughness.

12.3 PREDICTING GAS TEMPERATURE PROFILE

Predicting the pipeline temperature profile has become increasingly important in both the design and operation of pipelines and related facilities. Flowing gas temperature at any point in a pipeline may be calculated from known data in order to determine (1) location of line heaters for hydrate prevention (Towler and Mokhatab, 2004), (2) inlet-gas temperature at each compressor station, and (3) minimum gas flow rate required to maintain a specific gas temperature at a downstream point. To predict the temperature profile and to accurately calculate pressure drop, one needs to divide the pipeline into smaller segments. The temperature change calculations are iterative since the temperature (and pressure) at each point must be known to calculate the energy balance. Similarly, the pressure loss calculations are iterative since the pressure (and temperature) at each point along the pipeline must be known in order to determine the phase physical properties from which the pressure drop is calculated. Thus, generating a usable temperature profile requires a series of complex, interactive type calculations for which even the amount of data available in most cases is insufficient. Additionally, the pipeline outer environment properties such as soil data and temperatures vary along the pipeline route and therefore play a very important role and require consistent modeling to provide a reliable temperature profile evaluation. A simple and reasonable approach is to divide the pipeline into sections with defined soil characteristics and prevailing soil temperatures for summer and winter time and then to calculate the overall heat transfer that will be highly influenced by the outer conditions. The complexity of this method has led to the development of approximate analytical methods for prediction of temperature profile that, in most situations, are satisfactory for engineering applications.

Basic relationships needed for these calculation methods are thermal, mechanical energy balances and mass balance for the gas flow in the pipelines. The general, or thermal, energy balance can be written as follows (Buthod et al., 1971):

$$C_P\left(\frac{dT}{dx}\right)-\eta C_P\left(\frac{dP}{dx}\right)+\left(\frac{VdV}{g_C}\right)\left(\frac{dV}{dx}\right)+\left(\frac{g}{g_C}\right)\left(\frac{dH}{dx}\right)=-\frac{dq}{dx} \tag{12-15}$$

where T is gas temperature; P is absolute pressure of gas; V is gas linear velocity in the pipeline; q is heat loss per unit mass of flowing fluid; C_p is constant pressure specific heat; η is Joule–Thomson coefficient; H is height above datum; x is distance along the pipeline; g_c is conversion factor; and g is gravitational acceleration.

The major assumption in the development of Equation 12-15 is that the work term is zero between the compressor stations.

To calculate the heat transfer from the pipe to the ground (soil), per unit of pipeline length, one uses the Kennelly equation (Neher, 1949):

$$\frac{dq}{dx}=[2\,\pi\,K\,(T-T_S)]/m_G\,\mathrm{Ln}\left[(2H^{'}+\sqrt{4H^{'2}-D_0^2})/D_0\right] \tag{12-16}$$

where K is thermal conductivity of soil; T_S is undisturbed soil temperature at pipe centerline depth; m_G is mass flow rate of gas; $H^{'}$ is depth of burial of pipe (to centerline); and D_0 is outside diameter of pipe.

A basic assumption in Equation 12-16 is that the temperature of the gas is the same as the temperature of the pipe wall (the resistances to heat transfer in the fluid film and pipe wall are negligible).

The mechanical energy balance is given by Equation 12-17 (Streeter and Wylie, 1979):

$$\frac{1}{\rho}\frac{dP}{dx}+\frac{VdV}{g_c dx}+\frac{2f\;V^2}{g_c D_i}=0 \tag{12-17}$$

where ρ is density of gas; f is Fanning friction factor; and D_i is inside diameter of pipe.

The continuity equation, which follows, relates velocity to the pressure and temperature (Buthod et al., 1971):

$$\frac{dV}{dx} = \frac{-m_G}{\rho^2 A}\left[\frac{\rho}{P} - \frac{\rho}{Z}\left(\frac{\partial Z}{\partial P}\right)_T\right]\frac{dP}{dx} - \left[\frac{\rho}{T} + \frac{\rho}{Z}\left(\frac{\partial Z}{\partial T}\right)_P\right]\frac{dT}{dx} \qquad (12\text{-}18)$$

where A is inside cross-sectional area of pipe; and Z is gas compressibility factor.

Equations 12-15 to 12-18 are the basic equations that must be solved simultaneously for the calculation of the gas temperature and pressure profiles in pipelines. Details for solving this set of equations can be found in textbooks on numerical analysis, such as Constantinides and Mostoufi (1999).

The typical equations used to determine pipeline temperature loss are the integrated forms of the general equations. However, assumptions and simplifications must be made to obtain the integrated equations, even though the effects of these assumptions or simplifications are not always known. A major advantage of numerical integration of the differential equation is that fewer assumptions are necessary. Considering this fact, Coulter and Bardon (1979) presented an integrated equation as follows:

$$T_x = \left\{T_1 - \left[T_S + \left(\frac{\eta}{a}\right)\left(\frac{dP}{dx}\right)\right]\right\}e^{-ax} + \left[T_S + \left(\frac{\eta}{a}\right)\frac{dP}{dx}\right] \qquad (12\text{-}19)$$

where T_1 is the inlet-gas temperature; and the term "a" is defined as follows:

$$a = \frac{2\,\pi\,R\,U}{m_G\,C_P} \qquad (12\text{-}20)$$

where R is pipe radius; and U is overall-heat transfer coefficient.

Equation 12-19 can be used to determine the temperature distribution along the pipeline, neglecting kinetic and potential energy, and assuming that heat capacity at constant pressure, C_P, and Joule–Thomson coefficient remain constants along the pipeline. For most practical purposes, these assumptions are close to reality and generally do give quite good results. Moreover, for a long gas transmission pipeline with a moderate to small pressure drop, the temperature drop due to expansion is small (Buthod et al., 1971), and Equation 12-19 simplifies to Equation 12-21:

$$T_x = T_S + (T_1 - T_S)\ \mathrm{Exp}\ (-ax) \qquad (12\text{-}21)$$

Equation 12-21 does not account for the Joule–Thomson effect, which describes the cooling of an expanding gas in a transmission pipeline. Hence, it is expected that the fluid in the pipeline will reach soil temperature later than that predicted by Equation 12-19.

While there has been extensive effort in the development of such pipelines, little attention has been paid to the fact that the Joule–Thomson coefficient and heat capacity at constant pressure are not constants. However, an analytical technique for the prediction of temperature profile of buried gas pipelines was developed by Edalat and Mansoori (1988), while considering the fact that η and C_P are functions of both temperature and pressure. Readers are referred to the original reference for a detailed treatment of this method.

Example 12-1

A 0.827 specific gravity gas flowing at 180 MMscfd is transferred through a 104.4-mile horizontal pipeline with a 19-inch internal diameter. The inlet pressure and temperature are 1,165 psia and 95°F, respectively. The desired exiting pressure is 735 psia. Assuming an overall heat transfer coefficient of 0.25 $\text{Btu/hr.ft}^2.^\circ\text{F}$, Joule–Thomson coefficient of 0.1093°F/psi, gas heat capacity (C_P) of 0.56 $\text{Btu/lb}_m{}^\circ\text{F}$, and soil temperature of 60°F, how far will the gas travel before its temperature reaches the hydrate formation temperature? (Note that these conditions are very rare, as all export gas is dried to avoid hydrates. Hydrates could form in an export gas pipeline only when the dehydration unit breaks or the inhibitor pump breaks).

Solution

Using the given data:

$$\rho_G = \frac{P \times MW}{Z\,RT} = \frac{1165 \times 0.827 \times 28.96}{0.776 \times 10.7316 \times 554.6} = 6.041\ \frac{\text{lb}_m}{\text{ft}^3}$$

$$Q = Q_{SC}\left(\frac{P_b}{P}\right)\left(\frac{T}{T_b}\right)$$

$$Z = (180 \times 10^6)\left(\frac{14.7}{1165}\right)\left(\frac{554.6}{519.6}\right)(0.776)\left(\frac{1}{86400}\right) = 21.773\ \frac{\text{ft}^3}{\text{sec}}$$

$$m_G = (6.041)\,(21.773) = 131.530\ \frac{\text{lb}_m}{\text{sec}}$$

$$a = \frac{(\pi \times 1.5833 \times 0.25) \times 5280}{(131.530 \times 0.56 \times 3600)} = 0.0247\ \text{miles}^{-1}$$

Solving Equation 12-19 gives $T_x = 41.82 + 53.18\ \text{Exp}\,[-0.0247\,x]$.

By simultaneously using the preceding equation and the Katz gravity chart (refer to Figure 3-16) for prediction of hydrate formation temperature, one can determine the location of hydrate formation. In this case, the hydrate formation temperature is 70.5°F. The first location where hydrate formation occurs is 24.94 miles along the pipeline. However, Equation 12-21 is often used for most practical gas transmission pipelines. When the solution is used with these parameters, the equation becomes $T_x = 60 + 35 \text{ Exp } [-0.0247\ x]$. Based on this equation, the first location of hydrate formation is 48.62 miles along the line, which means Equation 12-21 predicts that the fluid in the pipeline will form hydrate much later than that predicted by Equation 12-19.

It should be noted that in this example the temperature profile is strongly affected by both the heat transfer coefficient, U, and the Joule–Thomson coefficient, η. According to Equation 12-19, heating would be required from 24.94 miles until 104.4 miles to prevent hydrate formation. This result depends strongly on U, η, and the soil temperature T_S. If the pipe was more strongly insulated and U was reduced to 0.1 Btu/hr.ft^2, the point where line heating is required would be extended out to 36.67 miles, but heating would still be required. It is also of note that the outlet temperature would drop to 45.87°F when $U = 0.25$ Btu/hr.ft^2.°F and to 43.27°F when $U = 0.1$ Btu/hr.ft^2.°F, both of which are below the outside soil temperature. This result may seem surprising to the reader, but it occurs due to the large pressure drop in the pipeline. In this case, line insulation is exacerbating the problem because it does not allow the soil to retard the Joule–Thomson cooling effect. Equation 12-21 is not able to account for this effect at all.

Note: In order to prevent the formation of hydrates, the temperature must be such that no point in the pipeline is in the region where a hydrate will form. A common technique to avoid hydrate formation in onshore gas transmission pipelines is thermal stimulation. Thermal stimulation involves the use of a source of heat applied directly in the form of injected steam or hot liquids, or indirectly via electric means. The addition of heat raises the temperature of the pipeline and forces hydrates to decompose. The direct approach works well during steady-state conditions but is of no benefit in certain transient or shut-in scenarios (Lervik et al., 1998). However, indirect heat such as the installation of line heaters for onshore gas transmission systems is feasible for these transient conditions. This technology is particularly applicable to transmission and distribution systems that operate in cold climates. The other method to avoid or prevent hydrate formation is insulation. In fact, the proper use of insulation may, in some cases, negate the requirement for a heater altogether (Carroll, 2003).

12.4 TRANSIENT FLOW IN GAS TRANSMISSION PIPELINES

Pipeline operation is such that transmission pipeline gas flow exists in the unsteady state, primarily due to variations in demand, inlet and outlet flow change, compressor start and stop, control setpoints, etc. In fact, steady-state operation is a rarity in practice. The unsteady nature of the gas flow indicates the need for a useful transient flow pipeline model to represent such conditions. In other words, the model should solve the time-dependent flow equations. When lines are modeled, however, it is sometimes convenient to make the simplifying assumption that flow is isothermal and steady state as long as we incorporate a load or swing factor contingent to a latter transient design, checking to prevent inadequate pipeline sizing with potential detrimental impact on the feasibility of pipeline projects (Santos, 1997). Steady-state models are widely used to design pipelines and to estimate flow and line pack. However, there are many situations in which an assumption of steady-state flow and its attendant ramifications produce unacceptable engineering results. Unsteady-state flow of gas in transmission lines can be described by a one-dimensional approach and by using an equation of state, continuity equation, momentum and energy equations. In practice, the form of the mathematical relationships depends on the assumptions made based on the operating conditions of the pipelines. For the case of slow transient flows due to fluctuations in demand, it is assumed that the gas in the line has sufficient time to reach thermal equilibrium with its constant-temperature surroundings. Similarly, for the case of rapid transient flows, it is assumed that the pressure changes occur instantaneously, allowing no time for heat transfer between the gas in the pipeline and the surroundings. However, for this case, heat conduction effects cannot be neglected. Streeter and Wylie (1970) presented different methods that provide an accurate means of simulating unsteady-state flow in gas transmission lines. For a detailed description of these method variations, which are beyond the scope of the present discussion, readers are referred to the original paper.

12.5 COMPRESSOR STATIONS AND ASSOCIATED PIPELINE INSTALLATIONS

Compressor stations are installed along the pipeline to boost the gas pressure in the pipeline to increase the pipeline capacity in order to meet the gas demand made by the users. Compressor stations are composed of one or more compressor units, each of which will include a compressor and driver together with valves, control systems and exhaust ducts, and noise attenuation systems. Each compressor station will have inlet filters or knockout vessels to protect the compressors from damage due to liquids and entrained particles. In addition to compressor stations, there may be gas injection and delivery points along the line where the pressure and flow will have to be monitored and controlled. Each of these locations will include pressure control facilities and flow measurement.

12.5.1 Compressor Drivers

Transmission systems have a high volume flow, and the compressor stations generally have a low head or low compression ration. Centrifugal compressors are the ideal for these low-head and high-volume applications. Centrifugal compressors are also high-speed machines and ideally should have high-speed drivers. Choices for drivers can be gas turbines, gas engines, and electric motors. The selection is usually based on considerations of cost, both capital and maintenance, fuel or energy cost, reliability, and availability. Dealing with each of these in turn, gas engines are low speed and therefore will require a gearbox to connect to the compressor. Gas engines may not be competitive with other drivers in terms of installed cost at the power levels demanded by large-diameter, high-pressure pipeline transmission systems. Gas turbines are high-speed machines and can be directly coupled to the compressor but require specialized operation and maintenance. Electric motors can be of several types with both fixed- and variable-speed options. Variable-speed drivers (VSDs) with electric motors present an overall performance much better than gas turbines, and their selection depends on site logistics and availability of reliable electric energy at a competitive price. They are very competitive in installed cost and maintenance cost against gas turbines. VSDs offer very low maintenance costs, quick starting, lower noise levels, and no emissions whatsoever.

The decision to use gas engine, gas turbine, or electric motor drivers is almost always based on feasibility studies that include site logistics, costs, availability, and reliability of the energy source. Overall system reliability and availability analysis will avoid the common notion that “there will always be gas in the pipeline, so the question of reliability and availability of the energy source for the gas turbine does not enter into the equation.” For the electric drive, there has to be a reliable electric grid within a certain distance range from the compressor station. Given that these conditions are satisfied, the decision then comes from the feasibility analysis that takes into account capital investment, fuel gas, electricity costs, overall performance, maintenance, and operation costs over the economic life cycle of the project.

The life-cycle cost must examine and test the results for sensitivity to cost escalation in power prices and gas prices, taking into account the correlation between these two commodities. A long-term power supply agreement would be required to mitigate risk. The question of using a diesel engine as a power source has not been considered because it offers no advantages over a similar gas engine, and it introduces another fuel, which invokes additional costs for transportation and storage.

12.5.2 Compressors Configurations

Since gas pipeline projects demand high amounts of capital expenditure and therefore are involved with investments risks, the project sponsors will try to maximize capacity usage and minimize investment so as to have a competitive transportation rate to offer to the market. At the same time, they will avoid the pipeline to operate with spare capacity. The decision on the compressor arrangement, whether series or parallel, is mostly based on economics and on simulation of failure analysis (Santos, 2000).

12.5.3 Reduction and Metering Stations

Each reduction and metering station branches off the pipeline and is used to reduce pressure and meter the gas to the various users. For the reduction and metering stations, the main equipment includes filters, heaters, pressure reduction and regulators, and flow-metering skids. In addition, each station is generally equipped with drain collection and disposal, an instrument gas system, and storage tanks.

12.5.3.1 Filters

Natural gas filter units are installed at each station to remove any entrained liquids and solids from the gas stream. The filters may be composed of cyclonic elements to centrifuge particles and liquids to the sides of the enclosing pressure vessel. These particles and liquids will then drop down for collection in a sump, which can be drained periodically.

12.5.3.2 Heaters

Natural gas heaters are installed to avoid the formation of hydrates, liquid hydrocarbons, and water as a result of pressure reduction. The gas heater is designed to raise the temperature of the gas such that after pressure reduction, the temperature of the gas will be above the dew point temperature at operating conditions and maximum flow. The heater is a water bath natural circulation type maintained at a temperature between 158°F and 176°F.

Where gas cost is high, an alternative is to use high-efficiency or condensing furnaces for the purpose of preheating the gas rather than the water bath heater.

12.5.3.3 Pressure Reduction and Regulation System

A pressure reduction system controls the supply pressure to the gas users at a regulated value. Each system consists of at least two trains of pressure reduction: one operating and the other standby. Each train will normally be composed of two valves in series: one being the "active valve" and the other the "monitor valve." Each valve will be equipped with a controller to operate the valve to maintain the preset discharge pressure.

12.5.3.4 Metering System

The flow rate of the gas has to be measured at a number of locations for the purpose of monitoring the performance of the pipeline system and, more particularly, at places where "custody transfer" takes place, that is, where gas is received from the supply source and gas is sold to the customer for distribution. Depending on the purpose for metering, whether for performance monitoring or for sales, the measuring techniques used may vary according to the accuracy demanded. Typically, a custody transfer metering station will be composed of one or two runs of pipe with a calibrated metering orifice in each run.

12.6 DESIGN CONSIDERATIONS OF SALES GAS PIPELINES

The typical design of a gas transmission pipeline involves a compromise between the pipe diameter, compressor station spacing, fuel usage, and maximum operating pressure. Each of these variables influences the overall construction and operating cost to some degree; hence, an optimized design improves the economics of the construction, operation of the system, and the competitiveness of the project.

12.6.1 Line Sizing Criteria

The pipe size generally is based on the acceptable pressure drop, compression ratio, and allowable gas velocities. Acceptable pressure drop in gas transmission pipelines must be one that minimizes the size of the required facilities and operating expenses such as the pipe itself, the installed compression power, the size and number of compressors, and fuel consumption. In fact, a large pressure drop between stations will result in a large compression ratio and might introduce poor compressor station performance. Experience has shown that the most cost-effective pipeline should have a pressure drop in the range between 3.50 and 5.83 psi/mile (Hughes, 1993). However, for those pipelines (short ones) in which pressure drop is of secondary importance, the pipe could be sized based on fluid velocity only. The flow velocity must be kept below maximum allowable velocity in order to prevent pipe erosion, noise, or vibration problems, especially for gases that may have a velocity exceeding 70 ft/sec. In systems with CO_2 fractions of as low as 1%–2%, field experiences indicate that the flow velocity should be limited to less than 50 ft/sec because it is difficult to inhibit CO_2 corrosion at higher gas velocities (Kumar, 1987).

In most pipelines, the recommended value for the gas velocity in the transmission pipelines is normally 40%–50% of the erosional velocity (Mohitpour et al., 2002). As a rule of thumb, pipe erosion begins to occur when the velocity of flow exceeds the value given by Equation 12-22 (Beggs, 1984):

$$V_e = \frac{C}{\rho_G^{0.5}} \tag{12-22}$$

where V_e is erosional flow velocity, ft/sec; ρ_G is density of the gas, lb/ft^3; and C is the empirical constant.

In most cases, C is taken to be 100. However, API RP 14E (1984) suggested a value of $C=100$ for a continuous service and 125 for a noncontinuous service. In addition, it suggested that values of C from 150 to 200 may be used for continuous, noncorrosive, or corrosion-controlled services, if no solid particles are present.

After the appropriate inside diameter for a pipe is selected, it is necessary to determine the pipe's outside diameter (wall thickness), which would result in the minimum possible fabrication cost while maintaining pipeline integrity.

Example 12-2

Given the following data for a pipe segment of a gas transmission line, calculate the erosional velocity, neglecting the gas viscosity effect.

$Q_{sc} = 25.7$ MMscfd
$T_a = 90°F$
$\gamma_G = 0.7$
$Z_a = 0.925$
$P_1 = 425$ psia
$L = 8{,}280$ ft
$D = 12$ inch
$E = 1.0$

Solution:

1. The outlet gas pressure is computed using the Panhandle A equation:

$$25.7\times10^{6}=\frac{\left[403.09\left(\frac{519.6}{14.7}\right)\left(P_1^2-P_2^2\right)^{0.5}(12)^{2.619}\right]}{\left(0.7^{0.4603}\right)\left(549.6\times0.925\times\frac{8{,}280}{5{,}280}\right)^{0.5397}}$$

$$P_1^2-P_2^2=7062.765;\ \text{therefore, } P_2=416.608\ \text{psia.}$$

2. Calculate the average pressure and gas density:

$$P_a=\frac{2}{3}\left[(425+416.608)-\frac{425\times416.608}{(425+416.608)}\right]=420.817\ \text{psia}$$

$$\rho_G=\frac{P_a\times MW}{Z_aRT_a}=\frac{420.817\,(0.7\times28.96)}{0.925\times10.731\times549.6}=1.563\ \frac{\text{lb}_m}{\text{ft}^3}$$

3. Compute the erosional velocity using the API RP 14E equation (Equation 12-22). Continuous service ($C=100$) is assumed.

$$V_e=\frac{100}{(1.563)^{0.5}}=79.971\ \frac{\text{ft}}{\text{sec}}$$

4. Check the gas velocities through the line to ensure that excessive erosion will not occur.

   The actual gas flow rate is calculated as

$$Q=Q_{SC}\left(\frac{P_b}{P_a}\right)\left(\frac{T_a}{T_b}\right)Z_a$$
$$=25.7\times10^{6}\left(\frac{14.7}{420.817}\right)\left(\frac{549.6}{519.6}\right)(0.925)=10.166\ \frac{\text{ft}^3}{\text{sec}}$$

   Therefore, the gas velocity is

$$V_G=\frac{10.166}{\left(\frac{\pi\,D^2}{4}\right)}=12.950\ \text{ft/sec}$$

   The gas velocity is below the erosional velocity, so erosion should not be expected. However, it is high enough to prevent solids from settling.

12.6.2 Compressor Station Spacing

In long-distance gas transmission systems with a number of operating compressors, there is a definite need to optimize the spacing between compressor stations. Compressor station spacing is fundamentally a matter of balancing capital and operating costs at conditions that represent the planned operating conditions of the transmission system. The process can become somewhat involved and lengthy, particularly as the selection of spacing needs to be designed in such a way to address a capacity ramp-up scenario that will cover not only the initial condition but also the future years associated with the economics of the pipeline. In case of unexpected growth opportunities, we can also rely on loop lines that may be a better additional choice to increase capacity even more.

For a given pipe diameter, the distance between compressor stations may be computed from the gas flow equation, assuming a value of pipeline operating pressure (station discharge pressure) and a next compressor station suction pressure limited to the maximum compression ratio adopted for the project. Ideally, the pipeline should operate as close to maximum allowable operating pressure (MAOP) as possible, as high density in the line of the flowing gas gives best efficiency. This would point to the selection of close compressor station spacing, but this approach would not be the best economical decision. A decision based on the pipeline economics is the recommended one. Based on the required gas flow, an initial diameter is assumed that results in a reasonable compression ratio (usually around 1.25 to 1.6 for transmission lines) and gas velocity, and the compressor station spacing is established by setting the maximum discharge pressure at the MAOP. Other diameters are tested, and compressor station spacing calculations are performed again. The optimum diameter is determined based on minimizing capital cost and operating cost, resulting in a chart (the so-called J curves because of their shape) that will plot transportation ratio in US$/MMBtu against transportation capacity based on predefined economic assumptions and risks (Santos and Saliby, 2003). Such assumptions include the design life of the facility, the required rate of return on capital employed, and the discount factor used to express the annual operating expenses incurred over time to a present value. The total cost is then plotted against compressor-discharge pressure, and a point of discharge pressure corresponding to the minimum total cost is picked as the best operating pressure. A good approach is to design for maximum capacity and define the required number of compressor stations and their spacing, and after that, going down on capacity on each operating year and taking off stations as required. This approach allows a better design and helps in defining equipment that would be equal. This also allows the installation schedule for the compressor stations and compressor units to be defined.

Capital expenditures (CAPEX) include costs such as pipes, valves, fittings, compressors, turbines (or electric motors), control and construction, and assembly costs. Operating expenditures (OPEX) include all maintenance and supervision and all fuel or energy costs. CAPEX can be derived from past experience and databases. OPEX has to be estimated based on the specific project and past experience. The most significant part of operating costs is fuel or energy and equipment overhaul. Fuel cost is directly related to compressor horsepower. In order to illustrate how compressor station spacing influences the economics of pipeline operation, we can set up a simple model. This hypothetical pipeline model is based on a system 1,000 miles long, operating at a maximum pressure of 1,000 psig and flowing 1,000 MMscfd. Assuming a uniform pressure loss per unit length of the pipe and station spacing, the inlet pressure at the first station downstream can be calculated, as can the horsepower needed to bring the pressure up to the discharge pressure setpoint. The process is repeated, and the total power needed is the sum of all the stations. Figure 12-2 shows the manner in which total power required increases with spacing of stations. Figure 12-3, based on the some data, shows horsepower in relation to number of stations. We need to keep in mind that even if we have an increase in required power for a pipeline with fewer compressor stations, the overall cost tends to be lower than many compressor stations with lower horsepower requirements. The installed cost per horsepower will be lower for larger compressor units, and thermodynamic performance will be much better, pointing to the

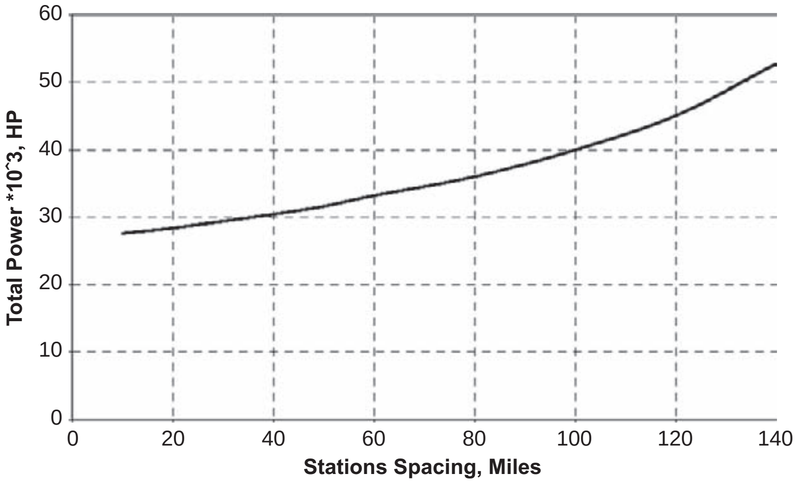

Figure 12-2 Effect of the compressor station spacing on total power (Cleveland and Mokhatab, 2005).

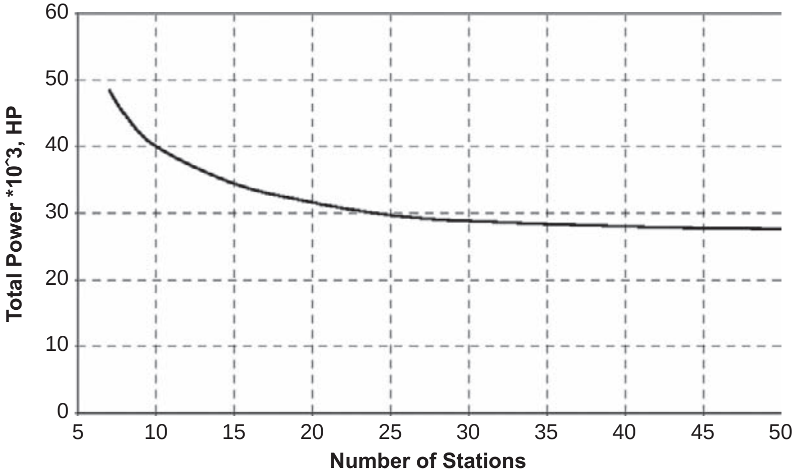

Figure 12-3 Effect of the number of compressor stations on total power (Cleveland and Mokhatab, 2005).

direction of having fewer compressor stations with larger compressor units. This explains why an economic evaluation has to be done for each project configuration, taking into consideration all related information in terms of CAPEX and OPEX.

Since fuel use is related to horsepower, minimum operating cost is associated with close compressor station spacing, which is logical because maximum transmission efficiency is obtained at the highest mean line pressure although the pipeline will have a larger diameter requiring higher CAPEX. However, the optimum is influenced by two factors: the first that small turbine compressor units are less efficient than large ones (have a higher specific fuel consumption), though this effect is small and almost negligible at unit powers above 20,000 horsepower. A much greater impact, though, lies in the cost of the stations, and it is this capital cost that declines as the number of stations is reduced (not linear, since large stations are proportionately more costly than small ones). This tends to move the optimum spacing away from the minimum distance shown on Figure 12-2.

Every project has to be considered individually because of specific factors and the relationship between CAPEX and OPEX that will differ, but the general conclusion that close spacing results in greater transmission efficiency may not be the best economic selection.

Experience shows that large units (compressor stations) are more efficient than smaller ones since larger centrifugal compressors and gas turbines have better efficiency. However, the impact of unit outage or failure must be simulated under a transient analysis (Santos, 1997) so that we can define the remaining capacity and therefore establish a maintenance criterion in terms of having standby units or spare equipment to allow quick replacing without affecting the contractual obligation in terms of transportation capacity.

When the preferred solution is found, it must be tested for robustness not only over the range of throughput anticipated but also for all credible upset conditions. After the optimum solution is developed, the result should then be applied to the practical case with elevation changes and other local factors, including the availability of sites, all of which may result in adjustments and minor changes.

12.6.3 Compression Power

The next step in the design of a pipeline system is to calculate the maximum power required at the stations and set the design point(s). Typically, a new pipeline system will grow from a low-flow condition to the maximum over a period of several years, and the decisions on compressors and drivers have to take these changing conditions into account. Growth of flow can be accommodated in several ways. Initially, compressors may be installed only at alternate stations and the intermediate stations built as the growth of flow dictates. Another option is to install one unit at each station location and then add units at the stations as flow increases. In the design phase, the capacity ramp-up will determine the installation schedule for the compressor station and also the additional units that will be necessary at the stations. Hydraulic simulators, both in steady and transient states, will help make an accurate design and will guarantee that the project will have good performance through the operating years without any unexpected situations. A preselection of the compressor units can also be performed while in the design phase (Santos, 1997). Another important job that can be checked early in transient analysis is the pipeline operating points inside the performance maps of the compressors so as to allow a proper selection of impellers and number of compressor units as the capacity increases yearly. Different compressor manufacturers can be modeled to check performance and fuel usage. Operations close to the surge line or operations that would require recycling would also be identified during transient analysis, underlying the importance of doing this kind of simulation in the design phase of a pipeline.

When the compressor design point is decided, the power required from the driver can be calculated using Equation 12-30.

12.7 PIPELINE OPERATIONS

In the industry supply chain, pipeline operations is an integral part of the transportation between exploration and production (E&P), or the "upstream end" that precedes it and distribution and the "downstream end" that follows it. Pipeline operations evolved from being prescriptive (i.e., defined by mandatory requirements) to its current stance of being performance based driven by risk management principles.

These trends stemmed from competitive forces that decrease operating costs; they also have evolved because of the experience gained from several decades of pipeline operations, along with the technologies and applications that developed along the way. These evolutions have given pipeline operators the tools they need to survive under these conditions. The pipeline facilities are mature to the point that many of them have exceeded their originally intended design life of approximately 25 years—at the time of conception. Today, most of these facilities continue to operate, partly for economic reasons because they are too costly to replace and also partly because these facilities still remain worthy of continued use (i.e., they are still deemed to be safe). Recognizing this, operating companies continue to extract value from these facilities, but under tremendous scrutiny and heightened awareness of their existence and vulnerabilities.

Current pipeline operation activities have taken on a new dimension of performance. While the basic activities continue (such as mechanical operations and maintenance of the facilities, including line pipe, valves, valve actuators, etc.; corrosion prevention and control; pipeline monitoring, as well as the focus on safety), today optimization of resources is being considered while still achieving safety, reliability, and efficiency. These challenges become more daunting given the fact that these pipeline systems expanded and merged, often acquiring systems built by others under different design, construction, and operating philosophies. Further, through staff reorganization and attrition, much of the corporate knowledge and information was misfiled or discarded. Some companies remained free-standing, whereas many became a part of a larger corporate entity. To overcome these developments, pipeline operators now strive for standardization in their procedures and compare their performance to industry benchmarks to gauge their performance and identify areas for improvement.

Certain time-dependent defects such as corrosion and environmental concerns started to manifest themselves in unplanned incidents. Development of other infrastructure at or near pipeline right-of-way saw an increase in third-party incidents and close calls. Pipeline regulators, too, evolved over these times and increased their vigil over the industry but allowed them to formulate their own facilities management programs. The industry sponsored research programs to better understand the consequence effects of pipeline incidents for risk evaluation.

Consequently, pipeline operation has been transformed toward the following areas of focus:

a) Making effective choices among risk-reduction measures.
b) Supporting specific operating and maintenance practices for pipelines subject to integrity threats.
c) Assigning priorities among inspection, monitoring, detection, and maintenance activities.
d) Supporting decisions associated with modifications to pipelines, such as rehabilitation or changes in service.

These focus points require that pipeline operations activities include the following elements (Mokhatab and Santos, 2005):

1. Baseline assessment and hazard identification
2. Integrity assessment by
    a. Inline inspection
    b. Hydrostatic testing
    c. Direct assessment
    d. Defect management and fitness for service
    e. Information management and data integration
    f. Risk management
3. Integrity management programs
4. Operator qualification and training
5. Operating procedures, including handling abnormal operating conditions
6. Change management
7. Operating excellence

These elements constitute a broad makeup of pipeline operations. Operators not only must be aware of them, but also must be well versed in their application, continuously improve them, and incorporate them into a comprehensive and systematic integrity management plan. Combined, these elements form the basis for directing a prevention, detection, and mitigation strategy for their system.

Pipeline operations will be the longest phase of the life cycle of a pipeline when cost management becomes a high priority. This priority will see pipeline operators performing many scenarios regarding life extension of the existing assets for enhancing value to the pipeline's stakeholders.

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CHAPTER 13

Selecting the Best Gas Processing Route

13.1 INTRODUCTION

The extreme utilization of natural gas represents a worldwide increase in consumption against the background of dwindling natural resources, which ultimately leads to an increase in the significance of natural gas treatment technologies. Various gas processing systems are available to provide products that meet specifications defined by plant owners. This variety places a huge burden and challenges on owners to select the right technologies for the project circumstances so that they can fulfill an optimized scheme meeting technological and economical targets. Given the magnitude of the investment in a gas processing plant, it is appropriate to carry out a rigorous treating process selection study to identify the most cost-effective and fit-for-purpose treatment package that removes contaminants in an environmentally friendly way. This chapter describes the most commonly used process technologies for designing the gas processing units and shows how integration of process technologies and expert process know-how make a difference. This chapter also discusses several integration aspects of the gas treating processes involved in order to establish the optimum treating lineup for designing gas processing plants, taking all the process limitations into account within a flexible, operable, and economically justified window.

13.2 PROPOSED TECHNOLOGIES FOR DESIGNING GAS PROCESSING PLANTS

In a typical example of the natural gas processing setup designed to produce pipeline gas from a sour gas feed, there are four main blocks, which usually means four licensors:

- Gas Sweetening Unit (GSU)
- Gas Dehydration Unit (GDU)
- Mercaptans Removal Unit (MRU)
- Sulfur Recovery Unit (SRU) + Tail Gas Treating Unit (TGTU)

Up to four license contracts are needed for each of the preceding steps, including the guarantees and well-defined border limits for the different interfaces. The large number of package vendors may give good reasons why the product specifications are not met. If one specification is not met in one unit, the other licensors cannot necessarily meet their liabilities. The other aspect is unit optimization. As each licensor is responsible for his part, he takes a certain security margin. The overall security margin is higher than the margin taking into account this concept as a whole with the strength and weaknesses of every single unit.

The following sections discuss several integration aspects of the main gas treating processes involved. The technical options considered may be a combination of both open-art technology and licensed processes.

13.2.1 Integrated Solutions

To eliminate all the constraints discussed in the preceding section, Lurgi (Germany), Shell (Netherlands), Prosernat (France), and UOP (USA) have developed four integrated solutions for improving, but not limited to, the overall sulfur recovery rate and curbing capital investment and operating costs. These integrated approaches, which are tailored to customer needs, will be the only comprehensive technology options where these integrated concepts offer the following advantages:

- One license contract/fee and one overall performance guarantee and liability.
- Tailor-made solution to avoid multiplying internal design margins, which can result in significant capital expenditure (CAPEX) and operating expenditure (OPEX) savings.
- These integrated concepts are based on Lurgi, Shell, Prosernat, and UOP proprietary know-how as leading technology licensors and engineering contractors that take the feed stream and deliver the required end product in an optimal manner, resulting in minimizing project execution time and a technically comprehensive plant.
- The scope of engineering services continues to detailed engineering as well as technical assistance at commissioning, start-up, and after-sales services (operator training and revamp studies).

13.2.1.1 Lurgi OmniSulf Package

Lurgi has developed a new concept, OmniSulf, based on its proprietary know-how as a leading technology licensor and engineering contractor;

it encompasses the following key processes, as shown in Figure 13-1 (Weiss, 2002):

- Gas sweetening with BASF-activated MDEA (aMDEA) process
- Sulfur recovery with Lurgi Claus, Lurgi Tail Gas Treating (LTGT), and Aquisulf technologies
- Gas dehydration and mercaptans removal with special Zeolite technology of Zeochem or CECA, typically a multi-layer bed of specialized 4A, 5A, and 13X molecular sieves
- Mercaptans recovery with Purisol technology

The activated MDEA process removes carbon dioxide (CO_2) and hydrogen sulfide (H_2S) to customer specifications, as well as mercaptans (RSH) and carbonyl sulfide (COS) partially. With regard to its characteristics—very low hydrocarbon co-absorption rate, energy-efficient solvent regeneration, and nondegrading/noncorrosive qualities—activated MDEA acts as a benchmark for competing technologies. This triggered an ongoing development of similar solvents (Hugo and Wagner, 2000). The sweetened gas is cooled in a chiller to remove the bulk of the moisture from the sweet gas. The cleaned gas is further routed to a dehydration and mercaptans removal unit (DMRU), wherein the moisture and mercaptans are removed by molecular sieve technology by Zeochem or CECA. If necessary, the cleaned gas can be further routed to a mercury (Hg) removal unit to remove traces of mercury using, e.g., impregnated activated carbon or activated alumina.

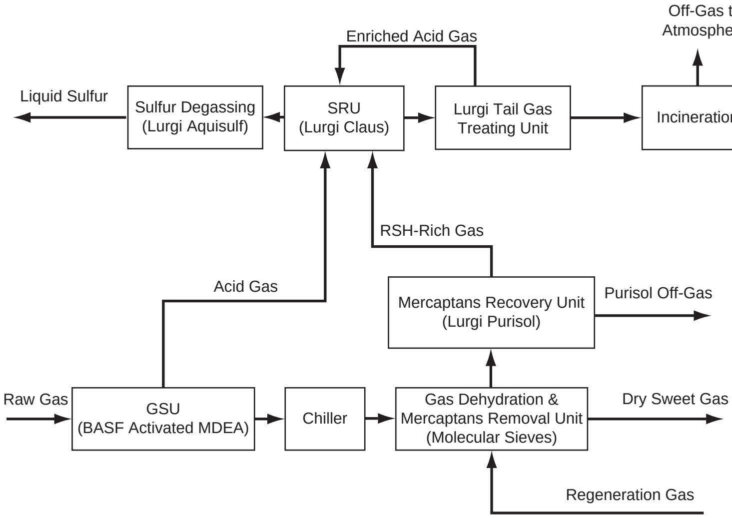

Figure 13-1 Overall arrangement of Lurgi OmniSulf package (Weiss et al., 2002).

Regeneration of the molecular sieve adsorbers is performed on a cyclic basis by means of regeneration gas coming from battery limits and/or dry-sweet gas from DMRU. The molecular sieve regeneration off-gas, rich in mercaptans, is routed to the mercaptans recovery unit utilizing Purisol technology. This is a physical gas purification technology, which uses an N-methyl-pyrrolidone (NMP) solution for the selective removal of sulfur components, specifically mercaptans. Other properties that make NMP a favorable choice are its high boiling point and extremely low vapor pressure, chemical and thermal stability, as well as its low viscosity, with the latter ensuring good heat and mass transfer. Mercaptans from the regeneration off-gas stream are absorbed by the circulating NMP solution; thus, it could be fed to the plant fuel gas system or used elsewhere. The separated mercaptan stream is processed in the sulfur recovery unit to recover the contained sulfur. The SRU consists of a modified Claus unit, a Lurgi tail gas treating unit (LTGTU), and a sulfur degassing system (using the Aquisulf process), all well-known, conventional, and proven technologies.

The novelty in the Lurgi OmniSulf concept is the combination of the Purisol unit for recovery of the mercaptans and the processing to elemental sulfur in the Lurgi modified Claus unit with its unique Multi-Purpose Burner system. Also, in the OmniSulf concept, the Lurgi tail gas treating unit includes an acid gas enrichment (AGE) step (utilizing generic MDEA solvent for the acid gas coming from GSU) that allows an overall sulfur recovery rate higher than 99% of all sulfur in the feed gas, indicated by a very low sulfur content in the tail gas treating unit's off-gas (and therefore low, environmentally friendly emissions).

The OmniSulf concept, which is an ideal combination of proprietary technologies contributed by Lurgi, BASF, Zeochem, or CECA, ensures a high on-stream factor and high reliability to provide products to specifications (Weiss et al., 2002).

13.2.1.2 Shell Integrated Gas Treatment Technologies

The Shell Global Solutions processes team offers customers the benefits of the knowledge and experience that the Royal Dutch/Shell Group has accumulated in over 100 years of operation. Some Shell technologies are protected by patent; other technologies are unique to Shell. The processes teams have developed innovative processes to meet emerging needs for designing a gas processing plant:

- One such innovation is the Sulfinol solvent, which is a mixture of Sulfolane, water, and DIPA (diisopropanolamine) or MDEA (methyldiethanolamine), Sulfinol-D, and Sulfinol-M, respectively. These remove hydrogen sulfide, carbon dioxide, carbonyl sulfide, mercaptans, and organic sulfur components from natural gas. Sulfinol is known to have a significantly higher removal capability for organic sulfur because of the higher physical solubilities for these components, due to the presence of the Sulfolane. However, the removal of all organic sulfur in only one absorber requires high Sulfinol circulation rates compared to cases in which only H_2S and CO_2 are removed (Bruijn et al., 2002). When mercaptans are present in the feed gas, the Shell Sulfinol process is strongly preferred as the acid gas removal step. Formulated MDEA solvents have a comparable capital cost to Sulfinol, but lack the mercaptans removal capabilities, one exception being the Flexsorb formulation (from Exxon) also containing Sulfolane. Although Sulfinol is more efficient in removing mercaptans, it has the disadvantage of relatively poor hydrocarbon selectivity, resulting in hydrocarbon losses in the separated acid gases, where combustion of the hydrocarbons in the SRU will result in an increase in CO_2 emissions (Rajani, 2003). However, this higher hydrocarbon solubility is still far less than that of a purely physical solvent and is normally considered acceptable (Klinkenbijl et al., 1999).

- Shell no longer possesses proprietary technology for the familiar Claus process, but in other areas of sulfur processing, Shell does have a number of successful processes such as Shell Claus Off-gas Treatment (SCOT), sulfur degassing, and catalytic incineration (Hoksberg et al., 1999). The SCOT process enhances sulfur recovery from Claus units (typical traditional recovery is about 95%) to levels over 99.8% and can help to meet the most stringent environmental limits (Brand, 2002; Rajani, 2003). The SCOT tail gas technology can readily be integrated with the main gas treatment, with or without an acid gas enrichment step, and if necessary, treatment of sour gas from molecular sieve regeneration (Klinkenbijl et al., 1999). Sulfur degassing (licensed with Jacobs Netherlands) removes hydrogen sulfide and polysulfides from liquid sulfur produced in Claus sulfur recovery units. The process eliminates potential explosion and odor hazards, and the risk to human health. It provides a high-quality sulfur product.
- Shell's molecular sieve package (e.g., with molecular sieves of CECA or UOP) helps gas plants improve operating cost, bed life, reliability, and availability of molecular sieve units (MSUs). As an example, Shell Global Solutions can translate the aging know-how of molecular sieves into operational procedures that maximize the lifetime of the molecular sieve, thus reducing the OPEX on adsorbents. The know-how may also be used to increase natural gas plant revenue by optimizing the amount of regeneration gas (to a minimum) while preventing cake

formation. For some plants, a reduction of the amount of regeneration gas also saves OPEX on fuel gas. The MSU regeneration gas is treated to remove the mercaptans from this stream by a second, smaller Sulfinol absorber before recycling the gas to the inlet of the main absorber. To reduce the hydrocarbon co-absorption from this process, the plant can use a source of regeneration gas that is very low in heavy hydrocarbons. The bed configuration selected for the MSU design offering by Shell minimizes the risk of hydrocarbon coadsorption on the molecular sieves without affecting the mercaptan-removal capacity of the beds (Grootjans, 2005). Although Shell has its own knowledge for molecular sieve design, the company gets its designs confirmed by molecular sieve suppliers to guarantee the performance of the adsorption process. Nevertheless, taking into account the differences between the different adsorbent product, the end customer should be aware that optimized designs are possible depending on the specific strengths of the adsorbents. As an example, a high-density adsorbent would allow small vessel designs with smaller regeneration gas quantity and flow rate, leading to significant investment and operating savings. CECA proposes this with its Optisieve concept.

It should be noted that the presented know-how of Shell Global Solutions has been shown to improve the performance of the MSU (Bruijn et al., 2002; Carlsson et al., 2005).

As stated previously, when the GSU is based on Sulfinol solvent, the solvent can be designed to remove all mercaptans. This specification is also easily met when the CO_2 has to be removed to deep specification and the solvent circulation is high. Subsequently, MSU has only to meet the water specification. Generally, this option will result in an expensive GSU design with a hydrocarbon co-absorption that is too large to be acceptable. In the aqueous-amine-solvent treating option, which removes only H_2S and part of the CO_2, although the co-absorption effect will be very low, the complexities of all mercaptans removal using molecular sieves can affect the net present value of an investment due to operational difficulties, resulting in unplanned downtime or a failure to meet design capacity for a certain period (Carlsson et al, 2007). The optimization of the lineup between the solvent-based step and the molecular sieves is very dependent on the level of mercaptans in the feed gas (Klinkenbijl et al., 1999). If the mercaptan content of the feed gas is high, use of an aqueous solvent for the GSU will require construction of a very large MSU, which would make this option uneconomical. However, if the mercaptan content is less than approximately 400 to 500 ppm, both processes offer advantages and disadvantages (Grootjans, 2005). The optimum solution in many cases is the distribution of the mercaptans removal capabilities over the mixed solvent in the GSU as well as the MSU. In this design, as shown in Figure 13-2, the regeneration of the MSU gas can be integrated with the GSU, using shared regeneration. The treated molsieve regeneration gas can be recycled either to the inlet of the molsieve unit or the inlet of the main absorber (Bruijn et al., 2002). An advantage of a dedicated acid gas enrichment unit (AGEU) is the possibility to separate the H_2S/CO_2 stream from the mercaptan stream, where the AGEU-treated gas is directly routed to the incinerator to ensure that small traces of H_2S and hydrocarbons are properly combusted. However, this stream contains mercaptans and heavy hydrocarbons removed in the GSU by the mixed solvent, and hence the mercaptans are lost from the sulfur recovery. The enrichment with this heated flash mechanism is not as efficient but can in some cases be sufficient. An aqueous amine unit, which is the best choice for economic reasons, will be used for increasing the

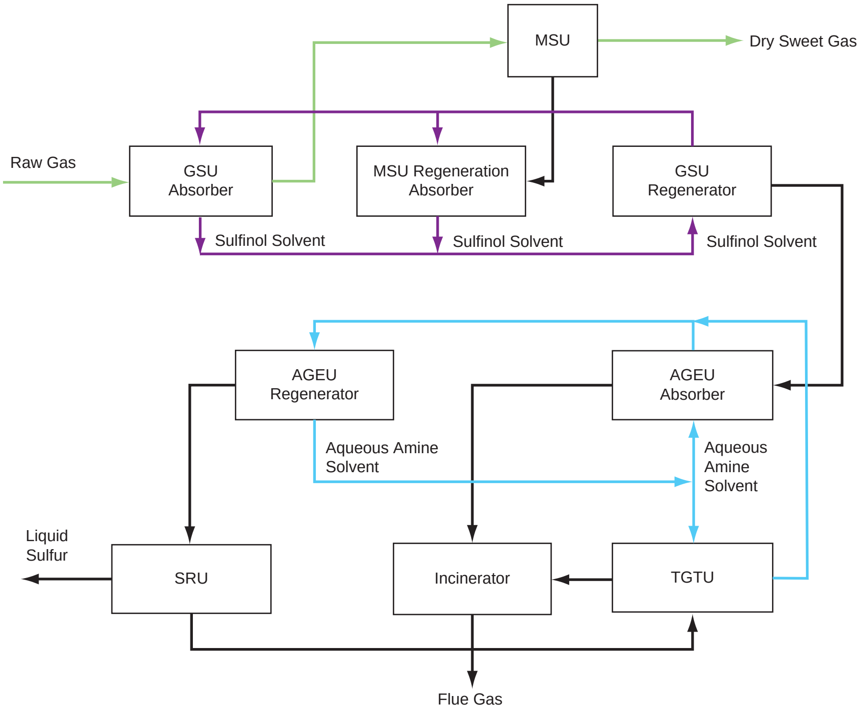

Figure 13-2 Overall arrangement of the Shell optimized solution with respect to maximization of treating train operability and flexibility (Klinkenbijl et al., 2005).

H_2S/CO_2 ratio in the SRU feed gas. The selective enrichment absorber can be integrated with the tail gas treating unit (TGTU) absorber with a common regenerator for optimization of capital and operating expenses (Klinkenbijl et al., 2005). The configuration described will ensure a sulfur recovery system with a recovery greater than 99.5% and also will ensure that the system will perform efficiently during feed gas composition variations.

The optimization is based on reducing the MSU size by addition of mercaptans removal capacity in the GSU. This option therefore represents the most flexible, operable, and cost-efficient solution at the expense of increased hydrocarbon co-absorption (van de Graaf and Klinkenbijl, 2003; Rajani and Bowerbank, 2004). Shell's optimized lineup is usually considered a cost-effective solution, requiring the lowest capital investment, for an existing plant to be converted to one that meets tight sulfur specifications for gas and liquid products, while offering some flexibility in the feed gas composition (Klinkenbijl et al., 2005).

13.2.1.3 Prosernat Integrated Solution

Prosernat offers a range of proprietary technologies that answer the issues of cost reduction, efficiency improvement, and environmental protection. These proprietary technologies are discussed next.

The AdvAmine technology portfolio developed by Total and licensed by Prosernat offers a wide range of solvents and configurations, all based on the use of widely available open market chemicals for gas sweetening applications (Streicher et al., 2004):

- **Hi**Load**DEA** for complete CO_2/H_2S removal, a process based on the use of the well-known DEA, which differentiates itself from open-art DEA technology by its advantageous proprietary features:
    - Use of a higher DEA concentration (up to 40 wt %) and very sour gases of high acid gas loadings (mol absorbed acid gas/mol amine) to reduce solvent flow rates
    - Special process configurations to minimize energy consumption
    - Proprietary process configuration to achieve up to 95% COS hydrolysis
- **MDEA**max for selective H_2S removal, a process to get the maximum benefit of the selectivity properties of aqueous solutions of MDEA. **MDEA**max selectively removes H_2S down to very low levels and maximizes CO_2 slippage to produce H_2S-rich gas for direct Claus treatment, for which the advantage is reducing energy consumption.
- Energized**MDEA** for controlled/complete CO_2/H_2S removal, a process using MDEA formulated with the addition of molecules, called

"energizers," that are able to accelerate the CO_2 capture by MDEA-based solutions. This process achieves complete removal of CO_2, while getting the benefits of using MDEA, a solvent easier to regenerate and less sensitive to degradation. A key benefit of Energized**MDEA** is the removal of CO_2, where some flash-procured regeneration can be obtained, leading to very low regeneration energy consumption. The type and amount of energizer can also be tailored to meet specific requirements and achieve, for instance, controlled CO_2 removal.

The AdvAmine portfolio is based on impressive know-how as well as the extensive industrial and operational experience of Total, as developer and user of these technologies (Streicher et al., 2004). Accumulated industrial experience has made AdvAmine a versatile gas sweetening technology portfolio meeting the industrial needs for

- All types of natural gas sweetening applications
- Reduced cost through optimal solvent use and improved process configurations (mainly high loading concept)
- Tight-treated gas specifications on H_2S, CO_2, but also COS (in order to reach very low levels of total sulfur) and some mercaptans
- Ease of operation (avoiding corrosion/foaming)
- Molecular sieve technology (Prosernat has had very good experience in this area with Axens or CECA) for gas dehydration and mercaptans removal unit

The proposed unit consists of adsorption on molecular sieves, enabling simultaneous removal of water and mercaptans. The regeneration off-gas from the molecular sieve unit is loaded with mercaptans, which in turn must be treated to separate the mercaptans. The amount of mercaptans in the regeneration off-gas varies according to the molecular sieve regeneration cycle, which then requires overdesign of some items of the Claus unit equipment and may impair the overall sulfur recovery efficiency when these mercaptans are recovered to be sent to a Claus unit (Lallemand et al., 2008). In some cases, incineration of the off-gas with peak RSH content can be proposed and the off-gas used as fuel gas when RSH levels are acceptable. It should be noted that Prosernat proposes a combination approach for removal of mercaptans from regeneration gas, in which the regeneration gas is first washed with aqueous DEA to remove the CO_2 (in order to minimize caustic consumption through irreversible reactions with CO_2). The acid gas obtained after regeneration of the DEA contains small amounts of H_2S and can be sent directly either to the thermal incinerator or at the inlet of the Claus. The regeneration gas is then washed with caustic solution to remove the mercaptans. The caustic solution loaded with mercaptans could then possibly be regenerated, for instance, with Axens's (Prosernat's sister company) Sulfrex technology. If this option is selected, it is possible to send the disulfide oil (DSO) obtained after regeneration of the caustic to a refining plant or to use other means to convert the DSO in order to minimize the SOx emission of the whole plant. Producing DSO to export to an oil refinery also prevents sending a stream of mercaptans with a high hydrocarbon content to the Claus unit (Streicher, 2009).

- AdvaSulf technological toolbox covering all SRU steps and options, and including (but not limited to) proprietary know-how and processes with an impressive track record of industrial experiences
- Claus technology including Total's experience, IFP's R&D support, and Axens's Catalysts' know-how
- Claus tail gas treatment (TGT) technologies including IFP's Clauspol II, Total's/Lurgi's Sulfreen (for some applications), and Prosernat's Sultimate. Basically, two main concepts can be used for tail gas treating:
  - Continuation of the Claus reaction at such lower temperatures that liquid sulfur will condense (sub-dew-point technologies): Clauspol II (the cheapest TGT technology available on the market from 99.4% to 99.8% sulfur recovery) and Sulfreen (best fit for large-capacity sulfur plants)
  - Conversion of all sulphur species present in the Claus tail gas into H_2S followed by selective H_2S absorption and recycle at Claus inlet: Sultimate (can achieve the highest levels of sulfur recovery up to 99.9+%)
- Acid gas enrichment (AGE) process for Claus feed stream quality upgrade: Total's AdvAmine (MDEAmax)
- Liquid sulfur degassing noncatalytic or catalytic with Total's Aquisulf

AdvaSulf is therefore Prosernat's answer to the key challenges of today's sulfur plants: reduced costs, higher sulfur recovery, higher flexibilities and reliabilities, and the possibility to treat diluted H_2S gas streams. Its applications include

- Complete H_2S extraction (with AdvAmine) and conversion (with AdvaSulf) treatment chain
- All types of H_2S-containing gas streams
- High conversions (99.9+%) achievable with a TGT unit, contributing to keeping AdvaSulf technologies at the forefront of today's environmental protection technologies

The Prosernat integrated solution is illustrated in Figure 13-3, which is a simplified diagram showing mainly the technology architecture.

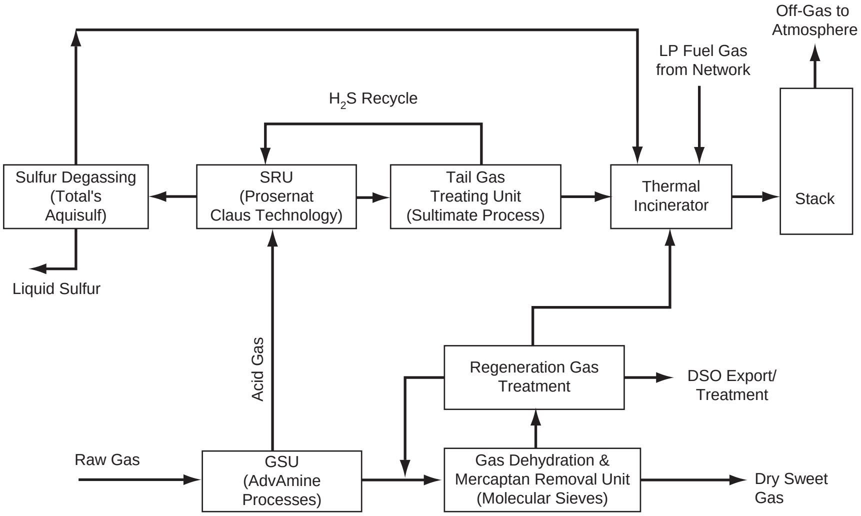

Figure 13-3 Technology architecture of Prosernat's integrated solution (Streicher, 2009).

For all its technologies, Prosernat provides high-quality fit-for-purpose designs, comprehensive training, and effective troubleshooting for customers in order to achieve successful long-term operation as well as single-source supply of fully guaranteed process-equipment packages.

13.2.1.4 UOP Integrated Approach

UOP has designed and implemented integrated natural gas treating technologies to effectively remove and control all acid gas components, water, and mercury (if necessary) prior to liquids recovery.

UOP's proven suite of solutions includes

- Amine Guard FS technology for gas sweetening
- Ortloff/Amoco's Claus sulfur recovery process
- MOLSIV adsorbents for dehydration and mercaptans removal
- Selexol technology to remove mercaptans, COS, and H_2S from molecular sieve regeneration gas
- Regenerative and nonregenerative fixed-bed technologies for mercury removal (if necessary)

UOP Amine Guard Formulated Solvent (FS) technology combines the Dow high-performance UCARSOL family of solvents with UOP's reliable Amine Guard system process technology. The process can be tailored for either bulk, selective, or trace acid gas removal for CO_2 and

H_2S to meet sales gas specifications. In this process, regeneration energy is minimized by choosing the optimum UCARSOL solvent for the situation, using high-solvent concentrations and proper selection of flow scheme. UOP has also developed Separex membrane systems, which are modular and skid-mounted units containing spiral-wound membrane elements, primarily for bulk CO_2 removal from natural gas streams (UOP, 1999). UOP's latest innovation, the Separex Multi-Tube Membrane, is a novel compact membrane system design that simplifies process stream connections, thus increasing membrane packing density and reducing the footprint and weight. The Multi-Tube system offers a 50% footprint reduction in most applications, making it even more attractive for offshore projects (Cnop, 2010).

UOP established a cooperative marketing agreement with Ortloff in 2002 for marketing and licensing Ortloff/Amoco sulfur recovery technologies. Ortloff offers Claus sulfur recovery processes based on the combination of Amoco's patented technology with the Ortloff know-how, developed from the design and operation of these plants. Amoco's cold bed adsorption (CBA) sulfur recovery process is the most widely used sub-dewpoint Claus process, normally employed downstream of a conventional modified Claus process section. The CBA process is generally capable of sulfur recovery efficiencies in the range of 97.5%–99.5%, depending on the H_2S concentration in the acid gas and the number of catalytic stages used. For recoveries in excess of 99.9%, the modified Claus process with tailgas cleanup is normally used. Due to the cyclic nature of the CBA process, the CBA switching valves are subjected to very demanding sulfur vapor service that has caused significant operation and maintenance problems in many of the CBA plants designed by others. In vivid contrast, Ortloff's proprietary sulfur vapor valve assemblies have functioned flawlessly without operational or leakage problems in the CBA plants designed by Ortloff. Ortloff sulfur recovery systems operate for longer periods of time between shutdowns, with reduced maintenance costs, and more operating ease and flexibility.

UOP uses MOLSIV adsorbents for natural gas dehydration and mercaptans removal. These adsorbents, composed of aluminosilicate crystalline polymers (zeolites), efficiently remove low concentrations of contaminants down to trace concentrations. Typical products used are UI-94, 4A-DG, and in special cases, AW-500 molecular sieves. UOP's UI-94 adsorbent is now a proven product that offers natural gas processors the opportunity to increase the life of molecular sieves by minimizing pressure drop increase over the life of the adsorbent (Rastelli and Shadden, 2007).

UOP Selexol technology is used to remove mercaptans, COS, and H_2S from molecular sieve regeneration gas. This process uses Dow's Selexol solvent—a physical solvent made of a dimethyl ether of polyethylene glycol, which is chemically inert and not subject to degradation. Since no chemical reactions are involved, Selexol usually requires less energy than the amine-based processes. However, at feed gas pressures below about 300 psia (2.07 MPa), the Selexol solvent capacity is reduced and the amine-based processes will usually be superior. Therefore, Selexol solvent is particularly effective in high-pressure, low-temperature, high-acid gas systems (Kohl and Nielsen, 1997).

UOP has developed a number of approaches to remove mercury from various locations within a plant. The first approach uses a large vessel of non-regenerative metal-sulfide adsorbent (UOP GB-562S) to remove mercury from the raw gas as it enters the plant. Another approach uses a silver-impregnated molecular sieve (HgSIV), designed to remove mercury from the raw gas prior to the treated gas passing to the cryogenic system. The UOP HgSIV adsorbent can be loaded into an existing molecular sieve dehydration unit to simultaneously remove mercury, water, and other impurities. HgSIV adsorbents are regenerated with a clean gas stream, whereby installing a small vessel of advanced nonregenerable mercury adsorbent (i.e., UOP GB-562) on the regeneration stream from the molecular sieve unit, mercury is effectively removed and captured. Applying such systems can provide cost-effective (lower capital cost) options with longer service life (Eckersley, 2010).

Note should be made that the types and quantities of the different acid gas components as well as the specific end-product requirements will determine the most cost-effective integrated gas treatment block configuration. As an example, for a sour feed gas containing mercaptans to be entered in a gas processing plant, the UOP's natural gas treatment technologies can be integrated as shown in Figure 13-4.

With the UOP integrated natural gas treatment solutions, the plant designer can expect lower utility requirements, lower waste disposal costs, higher hydrocarbon recovery, and greater sulfur recovery.

13.2.2 Combining Different Technology Licensors

The value-added contributions of competing process technologies are sometimes compromised by not properly selecting the boundary conditions. Each technology possesses its unique set of operating conditions where its performance excels. Likewise, each project is different and offers its own challenges to the available process technologies whereby a previously

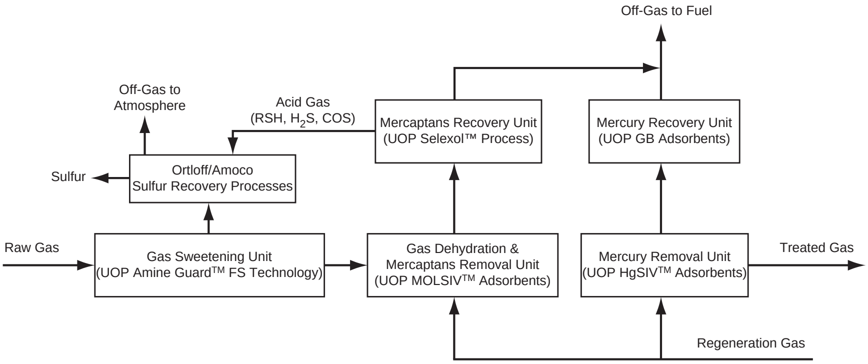

Figure 13-4 Typical illustration of UOP's integrated solutions for treatment of sour natural gas containing mercaptans (Modified after UOP, 2009).

utilized technology may not be cost effective. However, in this case, there are a few license contracts, guarantees and liabilities, and license fees. There are also multiple internal design margins, which can result in significant CAPEX and OPEX increases.

13.2.3 Integration of Open-Art Processes

Typically, contractors use equipment standards and process simulations, and, together with equipment vendor consultations, they are able to design gas processing units without the need to use licensed technologies. In this case, the plant owner may wish to accept the risks for design and operation of the plant and possible off-specification products. Licensed technologies, however, are mainly for specific unit processes where design has been optimized or proprietary materials (treating chemicals and solvents) are used. Work on process optimization has included increased control over the gas specification, reduction in energy consumption and waste generation, and reduction of capital costs through improved technologies and design. In fact, the investment in units utilizing proprietary/licensed "state-of-the-art" technology pays off in a very short time, resulting in a cost-effective and efficient unit.

13.2.4 Technology Selection Criteria

The earlier sections provided an overview of the different gas treatment technologies with their advantages and disadvantages. To design a full gas processing plant, one needs this information, together with well-defined selection criteria for the different technologies and their interfaces. The requirements of a strategic important technology with respect to competitive forces for a gas processing plant dictate that the most-effective treatment process technology should (1) be able to comply with sulfur specifications for gas and liquid products, (2) be able to process different feeds, and (3) be cost effective compared to other alternatives. Added to this are other considerations, such as reliability record, corrosion performance, quality of process guarantees, and availability of professional after-sales technical service (Mokhatab and Meyer, 2009).

Among the proposed alternatives, integrated solutions developed by Lurgi, Shell, Prosernat, and UOP will be the only comprehensive technology options. However, the results of an economic analysis will clearly indicate which alternative would be economically preferred. Determining the best process is very much dependent on the initial feed gas conditions, the treated gas specifications, and environmental requirements. Before the treatment technology is selected, it is strongly recommended that an optimization study is carried out to obtain the lowest CAPEX/OPEX and largest operating window with respect to feed gas composition (Klinkenbijl et al., 1999). In fact, the right technology, which results in a cleaner environment, improved reliability, and higher margins, has to be chosen on the basis of the environment of each individual project on a case-by-case basis, addressing the drawbacks and advantages of each option.

13.3 PROPOSED PROCESS TREATING LINEUPS

Natural gas processing is often thought of as being a mature technology with little opportunity for improvements or innovations. However, changes in requirements from customers continue to drive improvements in technology. The plant designer is therefore faced with applying existing knowledge to extend the design envelopes of the process with focus on improved solvent offerings and new applications of technologies. The following sections discuss five proposed integration aspects of the treating processes involved in a sour gas processing plant in order to establish the best treating lineup for designing the relevant gas processing circuit. A typical scheme of each lineup is illustrated in Figures 13-5 to 13-9.

Lineups 1 to 3, as illustrated in Figures 13-5 to 13-7, have the same sequence of gas treating with different technologies for hydrocarbon dew point controlling. In fact, in Lineups 1 to 3 the sweetened gas from GSU is first routed to the gas dehydration and mercaptans removal unit utilizing molecular sieve technology and then to the hydrocarbon dew point controlling unit (DPCU) utilizing propane refrigeration system, silica gel technology, and Joule–Thomson expansion technology, respectively.

Lineups 4 and 5, as illustrated in Figures 13-8 and 13-9, have the same sequence of gas treating with different technologies for gas dehydration and

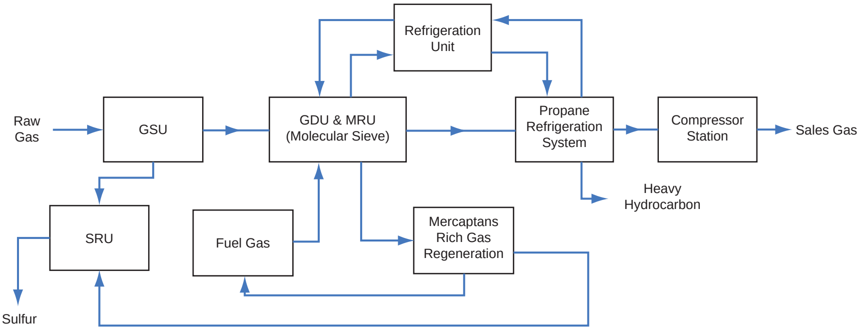

Figure 13-5 Typical illustration of Treating Lineup 1.

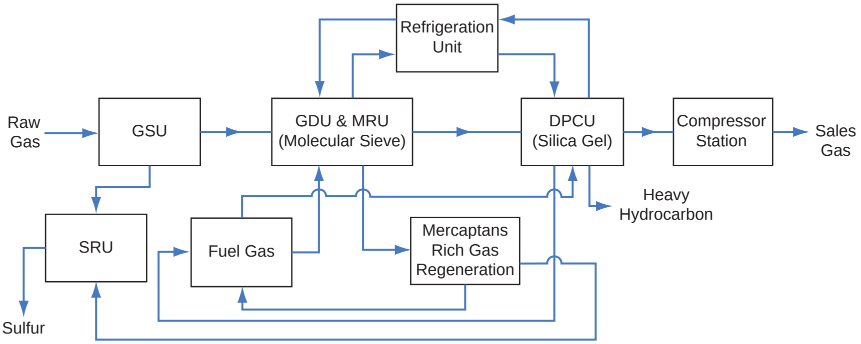

Figure 13-6 Typical illustration of Treating Lineup 2.

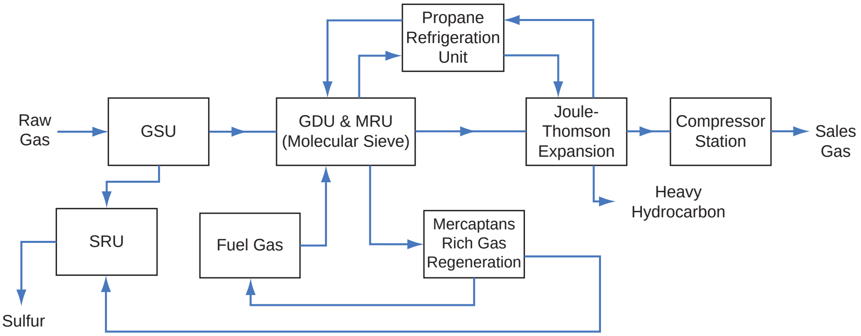

Figure 13-7 Typical illustration of Treating Lineup 3.

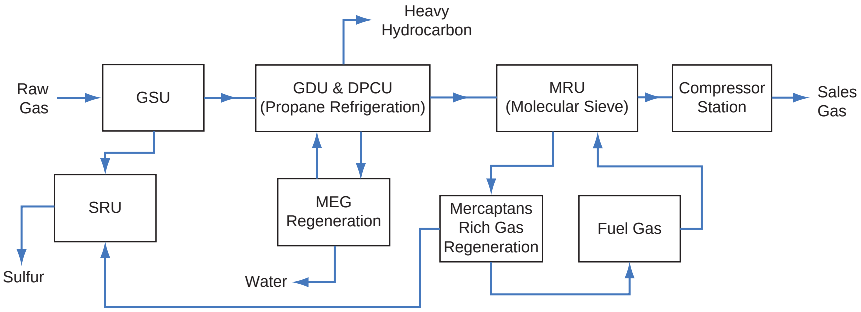

Figure 13-8 Typical illustration of Treating Lineup 4.

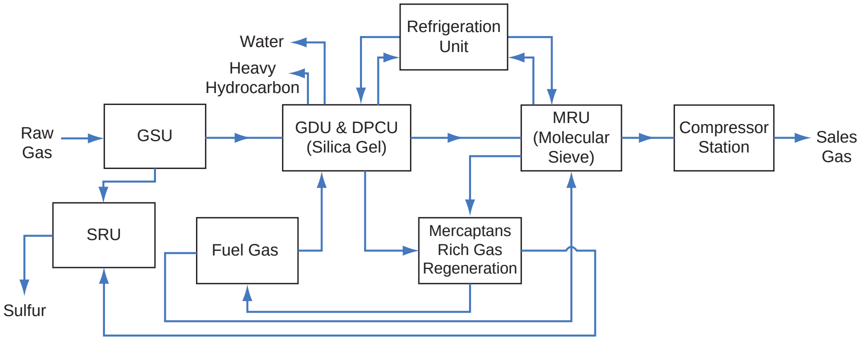

Figure 13-9 Typical illustration of Treating Lineup 5.

hydrocarbon dew point controlling. In Lineup 4 the sweetened gas from GSU is first routed to the gas dehydration and hydrocarbon dew point controlling unit using a propane refrigeration system accompanying monoethylene glycol (MEG) injection and then to the mercaptans removal unit utilizing molecular sieve technology. Lineup 5, as shown in Figure 13-9, is the same as Lineup 4, wherein silica gel technology has been applied for both gas dehydration and hydrocarbon dew point controlling.

13.3.1 Technical Analysis

Five options for integration between treating units in a sour gas processing plant have been proposed. However, due to this complexity, identification of the optimum scheme requires careful technical and economic evaluations of the alternatives on a case-to-case basis.

Selecting the right technology and tailoring the right process design for a given application require technical analysis that clearly indicates which alternative would be technically acceptable. Selecting the best process is very much dependent on the initial feed gas conditions, the treated gas specifications, and environmental requirements.

13.3.1.1 BTX Issue

In accordance with any given feed gas composition, there would be some technical issues with benzene, toluene, and xylene (BTX) in the gas dehydration and mercaptans removal unit utilizing molecular sieve technology (i.e., BTX co-adsorption on the sieve as well as their presence in the regeneration gas) in proposed Lineups 1 to 3, but this can be handled. Considering this, two options would be possible for a particular design of the gas dehydration and mercaptans removal unit with molecular sieves (MS):

- Option I: Complete water and mercaptans removal with 4A, 5A, and 13X molecular sieves
- Option II: Complete water removal with 4A, but bulk removal of light mercaptans (C1/C2 and n-C3SH) with 5A molecular sieve

Solution I would allow a complete sweetening of the gas, including the sweetening of the C_5^+ cut recovered from the DPCU. On the other hand, the BTX competition with the heavy mercaptans on the 13X should be taken into account. The overall design (bed size) would be higher than Solution II, resulting in a higher regeneration gas flow rate. Assuming the correct design, the BTX will not make impossible the RSH removal by MS, but the co-adsorbed BTX will be in the regeneration gas. As the most common treatment of the regeneration gas is a physisorption, it will be captured, too, and will have to be separated from the mercaptans before treatment in the Claus unit. Solution II would have the advantage of no BTX co-adsorption with additionally a lower regeneration gas flow rate. Most of the mercaptans will be removed from the gas; the heavier mercaptans will be knocked down with the C_5^+ cut, which is commonly treated with the liquid from the condensate stabilizer containing, in any case, other sulfur species that should be addressed by further treatment in a refinery.

13.3.1.2 Design Flexibilities and Issues

Drying cannot be separated from mercaptans removal. If both drying and mercaptans removal are necessary, one of the other drying options except molecular sieve technology very often needs another processing unit for the removal of mercaptans. When the beds are regenerated, the water and mercaptans desorb into the regeneration gas, which can then be treated using a physical solvent process to sweeten the regeneration gas and provide a concentrated mercaptans stream that can be sent to the SRU. Removing mercaptans in this way reduces the quantity of sulfur species feeding forward into the hydrocarbon recovery unit and thus into the liquid products. This results in a significant reduction in liquid treating requirements and can open up new technology opportunities in this area (e.g., the use of molecular sieves instead of caustic-based processes). Considering this, process Lineups 1 to 3, which utilize molecular sieve technology for gas dehydration and mercaptans removal simultaneously, would result in an appropriate design that takes into account the operating flexibility over the feed gas range. However, Lineup 3 would not be attractive unless the outlet pressure requirement is much lower compared to the inlet, which does not require gas compression, or the inlet pressure is too high, putting the operating range in a critical region.

In Lineup 1, although using propane refrigeration systems for hydrocarbon dew point controlling does require greater operator attention and maintenance than the silica gel system used in Lineup 2, there is a great flexibility in treating Lineup 1, where a branch of cold propane stream also can be used in order to keep the feed gas temperature of both gas dehydration and mercaptans removal units at 40°C or below, which is required by most solid bed vendors. The disadvantages of a silica gel adsorption system for hydrocarbon DPCU arise from the need to regenerate, by heating not only the silica gel bed to a suitable temperature (approximately 250°C) to drive off contaminants but also vessel and pipe work, too. This requires a relatively larger heater, usually a fired heater, which incurs large capital and operating costs. Internal insulation of the adsorber vessels may be justified to reduce heater and fuel gas costs. The need for at least two and possibly more large, high-pressure adsorber vessels also makes for a relatively high capital cost. Therefore, while silica gel provides a simple, reliable, and flexible process facility that requires little operator attention, it can come at a relatively high cost (Finn and Tomlinson, 2007).

Although the amount of molecular-sieve regeneration gas, rich in mercaptans, in Lineup 1 is more than its value into Lineup 4, which increases the difficulty of operation as well as the unit cost, there are some issues in processing Lineup 4 as follows:

- The MEG and condensed water/hydrocarbon mixture has high viscosity and needs to be heated to ensure good separation. Even then, uncertainty in composition and the extent of hydrocarbon absorption in the MEG may cause operational difficulties.
- Glycol co-adsorbs some BTX that ends up in the regeneration vapor stream. In fact, because glycol shows some affinity for hydrocarbons, the off-gas from the glycol stripper often contains BTX.

- Lineup 4 is completely dependent on propane refrigeration. When the propane system goes down, there is no dehydration. In fact, the purpose of the injected MEG is not to "dehydrate" the gas, but to prevent formation of hydrates.
- There is considerable loss of glycol caused by its solubility in hydrocarbon condensate.
- Rich MEG must be sent to the MEG regeneration system, which will add to the system cost.
- Determining the appropriate glycol injection rate and successive injection points is a difficult task that requires more operator attention in order to prevent hydrate formation in the hydrocarbon dew point controlling unit.

Silica gel technology, as indicated in Lineup 5, allows the single-step removal of both water and heavy hydrocarbons from natural gas in order to reach the dew point pipeline specs for water and hydrocarbons, by needing a lower regeneration gas compared to molecular sieves due to the short cycle operation. Untreated gas is used for the heating and cooling and recycled to the inlet of the unit after condensing partially the water and the heavy hydrocarbons. However, the adsorption times are usually in the range of dozens of minutes up to 2–3 hours, which, on the other hand, reduces lifetime expectations. This would also increase the required instrument air as well as the air compressor power consumption, which results in a relatively high operating cost. The other disadvantages of silica gel technology would be (1) fuel gas consumption resulting in emissions of CO_2 and NOx and (2) exhausted adsorbent would need to be disposed of with an approved environmental method.

13.3.1.3 Process Guarantees

Leading technology licensors commonly take the licensed units approach presented in Lineups 1 to 3 for optimal design of the gas processing plants. With this integrated concept, single-line project responsibility (one license contract, one overall guarantee and liability, and one transparent license fee), optimized tailor-made design, and customer-oriented supply from a single source result in considerable savings in investment as well as operating costs and enhanced flexibility of the whole plant.

13.3.2 Lineup Selection Criteria

In configuring the optimum flow scheme, the plant designer must understand the process technology options available for the individual units, their integration opportunities, and their limitations. Technical risk, licensor experience, degree of commercialization, safety, health, and environmental aspects all need to be weighed, along with the process and economic performance of the technologies concerned. Considering these and in accordance with all of the aforementioned technical items, process Lineups 1 to 3 should be selected for the design of a sour gas processing plant; otherwise, the overall concept will not be optimized, and the plant owner may wish to accept the risks for design and operation of the plant and possible off-specification products. However, in order to select the optimum treating lineup, a technical-economic analysis between proposed approaches for design of hydrocarbon DPCU should be done, such that the obtained result clearly indicates which alternative would be technically and economically acceptable (Mokhatab, 2010).

Selecting the right technology and tailoring the right process design for a given application requires both extensive industrial experience and the possibility to choose among various technologies and process options. This means, the investment in a high quality engineering consultant to provide process selection study will be well rewarded with a reliable fit-for-purpose and cost-effective treatment package. It is worth mentioning that selecting the best-fit alternative is a complex affair, extremely sensitive to a number of parameters that merit careful attention. As an example, plant owner preferences may also come into play, which may dictate the final solution or provide restrictions on the process configuration.

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CHAPTER 14

Process Control Fundamentals

14.1 INTRODUCTION

Processing natural gas to profitably produce end products requires precision and is potentially hazardous. Small changes in a process can have a large impact on the end result. Conditions such as pressure, temperature, flow, composition, and many other factors must be controlled carefully and consistently within specified limits to ensure quality and safety. Process control enables the safe and reliable production of natural gas and its derivatives while reducing manufacturing costs such as raw materials and energy.

This chapter introduces readers to a better understanding of the fundamental concepts in dynamics and process control, which include the ability to select a process control configuration for specific applications, decide on suitable instruments for process monitoring, develop input-output relationships, understand the dynamic open-loop and closed-loop characteristics, and design and tune feedback controllers.

14.2 DYNAMIC PROCESS CHARACTERISTICS

The dynamic process characteristics of most processes are described by three elements: resistance, capacitance, and dead time contributions. These elements will determine how the process responds to changes.

14.2.1 Resistance-Type Processes

A characteristic of a resistance element is the ability to transfer material or energy. Flow through a pipe is the most common example of a resistance-type process. A strictly resistance-type process has a proportional-only response. Any change in the resistance—for example, a control valve opening—will result in an immediate proportional change in the flow. The amount of change is a function of the process gain. Any change in the load—for example, the upstream and downstream pressure in this case of flow through a pipe—will also result in an immediate proportional change in the flow. The amount of change is again a function of the process gain.

14.2.2 Capacitance-Type Processes

A characteristic of a capacitance element is the ability to store energy and mass. Thermal capacitance is simply the product of the mass of an object and its specific heat. Energy can be stored in a heat exchanger. Mass capacitance is the accumulation of mass. In other words, input does not always equal output. Separators and surge dampers are common examples of strictly mass capacitance processes.

The gas capacitance of a tank is constant. The liquid or solid capacitance equals the cross-sectional area of the tank at the surface. If the cross-sectional area is constant, then the capacitance is constant at any head.

An example of a strictly capacitive process is a tank containing liquid with liquid removed at a constant rate—for instance, a positive displacement pump. The change in level of the tank is the product of the difference between the inlet and outlet flow rate and the time that the tank has been filling divided by the capacitance, or cross-sectional area, of the tank. Since the capacitance of the tank can change with level and therefore time—for instance, a sphere or horizontal tank with elliptical heads—the capacitance is a differential.

The larger the vessel is in relation to the flows, the larger the capacitance and hence the more slowly the controlled variable changes for a given change in the manipulated variable. The capacitance of the process tends to attenuate disturbances and hence makes control less difficult. The capacitance of the process is its time constant. The time constant is calculated from the differential equations used to model the process, but as an approximation is roughly equal to the process residence time.

14.2.3 Process Dead Time

Another contributing factor to the dynamics of many processes is a transportation lag or dead time. Dead time is the delay in time for an output to respond to an input. For example, if the flow into a pipeline is increased, a period of time elapses before a change in the outlet flow is detected. This elapsed time is a function of the line length and fluid velocity. Since dead time is the time required to move material from one point to another, it is frequently referred to as "transportation lag." Some examples of dead time include static mixers and conveyor belts. Dead time is also encountered when energy or mass sources must flow to the destination where temperature, composition, flow, or pressure is sensed. Delay also arises when a control signal requires time to travel between two points.

Dead time interferes with good control inasmuch as it represents an interval during which the controller has no information about the effect of a load change or control action already taken. In the design of a control system, every attempt should be made to minimize this delay by properly locating instruments and sampling points, ensuring sufficient mixing, and minimizing transmission lags.

Dead time is particularly difficult for feedback controllers for a variety of reasons. Feedback controllers are designed for immediate reaction, and dead time varies with throughput.

14.2.4 Inertia-Type Processes

The motion of matter as described by Newton's second law typifies inertia-type processes. These effects are important when fluids are accelerated or decelerated as well as in mechanical systems with moving components.

14.2.5 Combinations of Dynamic Characteristics

Few processes are strictly resistance or capacitance types. Systems that contain a resistance element and a capacitance element result in a single time-constant process. As noninteracting resistance and capacitance elements are added to the system, then multiple time constants result. When inertial effects or interactions between first-order resistance and capacitance elements are introduced, then processes exhibit a second-order response.

Single time-constant processes are described with a first-order differential equation containing a constant gain and a time constant. The gain is independent of the time characteristics and relates to the process amplification or ultimate output divided by input. The time constant describes the time required for the system to adjust to an input. This time constant is simply the product of the resistance and the capacitance.

The following equation describes a first-order process response without dead time:

$$\text{Change in Output} = \text{Gain} \times \text{Change in Input}\left(1 - e^{-\mathrm{t}/\text{time constant}}\right) \tag{14-1}$$

Two distinct principles apply to all first-order processes without dead time. First, the initial rate of change is the maximum, and, if unchanged, the system would reach the ultimate output in a period equal to the time constant. Second, the actual response is exponential with the output reaching 63.2% of the ultimate value at a time lapse equal to the time constant of the system. Dead time just shifts this response to begin after the delay has elapsed.

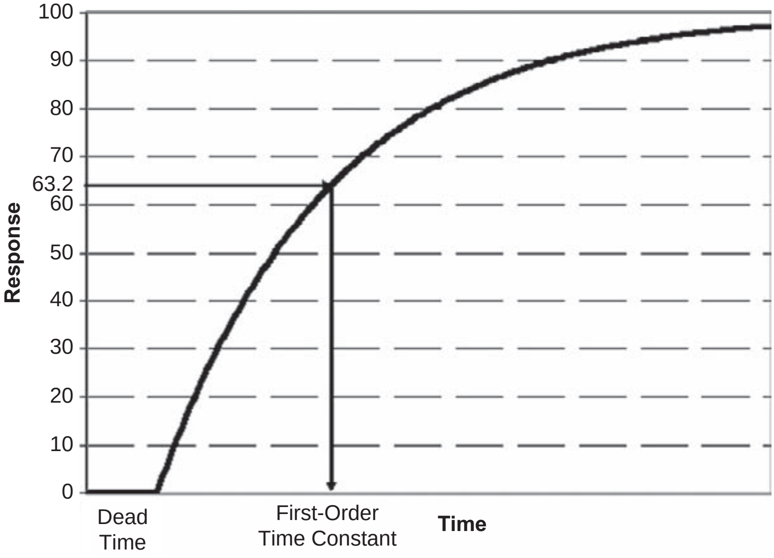


Figure 14-1 First-order open-loop process response.

Combinations of first-order lag with dead time, as shown in Figure 14-1, adequately estimate the response of most processes for purposes of simulation and control.

14.2.6 Examples: Simple Systems

14.2.6.1 Pipelines

Pipelines are controlled to provide a target pressure or flow at the end of the line. Pressure at the beginning of the pipeline is independent, as determined by a gas source or a compressor. The source pressure and the controlled pressure determine the flow rate.

The static gain (G) of flow to upstream pressure (P_1) is

$$G = F * P_1/\left(P_1{}^2 - P_2{}^2\right) \tag{14-2}$$

where P_2 is the downstream pressure.

The sum of the dead time and time constant can be determined empirically as (Liptak, 1995)

$$K * V * MW * (P_1 - P_2)/T * Z * F \tag{14-3}$$

where K is 0.015 for English units; V is the volume of the pipeline in cubic feet; MW is the molecular weight of the gas; P is the pressure in psia; T is the gas temperature in °K; Z is the compressibility factor; and F is the flow in lb/sec.

For the conditions of $F = 1{,}500$ lb/sec, $P_1 = 2{,}000$ psia, $P_2 = 1{,}000$ psia, $V = 3{,}226{,}200$ cubic feet (approximately 48 inches diameter by 50 miles pipeline), $MW = 20$, $T = 50°F$ (10°C or 283°K), and $Z = 0.95$, the process gain will be 1 lb/sec per psi, and the dead time plus first-order time constant will be 2,400 seconds, or 40 minutes.

Pumps, compressors, and turbines have static relationships between the process variables, so constant-speed rotating equipment introduces no dynamics to the process. Variable-speed equipment usually introduces a very small, usually negligible, lag.

14.2.6.2 Vessels and Piping

Systems of vessels with connecting piping are simple examples of combined capacitance and resistance processes. The piping contributes a resistance element, and the vessels contribute a capacitance element to the system. A first-order system accurately describes these combinations.

For the example of a vessel filled with gas and incoming piping, the process gain between vessel pressure and incoming flow is a function of the pressure drop in the line to the vessel and the flow rate. The time constant is calculated rigorously from the following (Liptak, 1995):

$$2 * V * MW * (P_1 - P_2)/(R * T * F) \tag{14-4}$$

where V is volume of the vessel; MW is molecular weight of the gas; $P_1 - P_2$ is the pressure drop; R is ideal gas constant; T is temperature; and F is flow rate.

For a vessel with volume of 100 cubic feet, gas with molecular weight of 29, a vessel pressure of 100 psia (P_2), a source pressure of 105 psia (P_1), gas constant (10.732 psia-ft^3/lbmole-°R), gas temperature of 62.33°F (522°R), and gas inlet flow rate of 10 lb mass per minute, the time constant will be about 0.5 minute, or 30 seconds. For fluid temperature and composition variations, the length of pipe introduces a dead time.

14.2.6.3 Concurrent Heat Exchanger

Gas processing strategies commonly employ concurrent heat exchangers. There are six process variables involved: cold stream flow, hot stream flow, cold stream inlet temperature, cold stream outlet temperature, hot stream inlet temperature, and hot stream outlet temperature. Four of the variables are independent, whereas two are dependent. The flow rate of one stream and the inlet temperatures are based on other system considerations. This leaves the outlet temperatures and one flow rate as potential independent variables. Most often, an outlet temperature is the controlled variable, and an inlet flow rate or bypass flow is the manipulated variable.

Assuming that the temperature change of the tube wall dominates the transfer function, the time constant is approximated by the mass of the tubes and the specific heat of the tubes divided by the product of the flow rates and specific heats of the two process streams. The following empirical formula for the time constant can be derived (Liptak, 1995):

$$K * M_t * Cp_t/(F_w * Cp_w * F_c * Cp_c) \tag{14-5}$$

where $K = 100$ for English units.

For the case in which M_t (mass of tubes) is 4,400 lbm, Cp_t (specific heat of tubes) is 0.112 Btu/lb °F, F_w (flow of warm stream) is 55 lb/sec, Cp_w (specific heat of warm stream) is 0.70 Btu/lb °F, F_c (flow of cold stream) is 80 lb/sec, and Cp_c (specific heat of cold stream) is 0.80 Btu/lb °F, the time constant will be 20 seconds.

Note should be made that the dynamic considerations for furnaces and boilers are similar to heat exchangers.

14.2.7 Effects of Variable Conditions

In the preceding examples, the time constant depends on flow rate, which varies. Therefore, the time constant will vary with flow. The process gain and dead time will change with flow as well. Flow control valves can also introduce nonlinear responses due to changes in inlet or outlet pressure as well as the characteristics of the valve itself.

Instrumentation can also introduce gain variations. Temperature measurements typically introduce a time lag while composition analyzers normally introduce a dead time. Pneumatic control valves without positioners can introduce delays. Although digital control systems significantly reduce the delays experienced with analog systems, delays can occur due to scan frequency.

14.3 CLOSED-LOOP CONTROL

The most important aspect of control is the dynamic relationship between the measured, dependent, or controlled variable and the correcting, independent, or manipulated variable. In order to control the desired outcome of the measured variable, the response of the corrective action must be expressed as a function of time. In other words, the control dynamics must be known. The dynamic behavior of a closed-loop system depends on the gain or transfer function of the process and the gain or transfer function of the control system, including the influences of valve movements and control signal delays. The size and configuration of the equipment such as piping, vessels, heat exchangers, distillation columns, compressors, and pumps as well as the laws of chemistry and physics establish the process gain.

Knowledge of the process gain is necessary in order to provide stable control of the loop. Systems of differential equations can accurately calculate the process gain. The solution of the differential equation system with properly defined initial conditions provides the values of the process variables over time.

In the absence of a dynamic simulation, we can use the first-order lag with a dead-time model to approximate the process gain.

14.3.1 Degrees of Freedom

In the development of a control system for a process, it is desirable to know precisely the number of process variables, which the system designer is entitled to attempt to regulate, commonly known as the degrees of freedom of the process. In fact, an analysis of the dynamics of a system must begin with an analysis of the degrees of freedom. The degrees of freedom are simply the number of variables that describe the system less the independent relationships that exist among the variables. The degrees of freedom dictate the number of variables to specify when properly simulating a system. Once the plant is specified, the design of a control structure requires that the control degrees of freedom be known. This is the number of variables that are controllable. It is very easy to calculate this number, even for quite complex processes, because it is equal to the number of manipulated variables (the number of control valves in the process). For control system design, the degrees of freedom limit the number of automatic controllers. However, these variables are different from the design optimization variables. When optimizing, the degrees of freedom determine the number of variables that must be included in the objective function to yield a single solution.

Once a variable is determined to be a manipulated or specified variable, then it becomes an independent variable and uses a degree of freedom. Other variables that have a relationship to this variable are dependent variables.

14.3.2 Controllers

In order to choose controls properly, one must know the requirements of the process and the corresponding characteristics of the controller. The controller characteristics vary with the control mode. The controller mode ultimately chosen depends on the process dynamics, control objectives, and controller cost.

Feedback or closed-loop controllers compare a measurement (controlled variable) to its desired value (setpoint) and generate a correction signal (change in the controller output) to a control element (manipulated variable) based on the difference between the measurement and the desired value. The difference between the setpoint and measurement is the error. Negative feedback controllers act to eliminate the error. A controller can operate on the error in a variety of ways.

Control action is determined by the way the controller output responds to the error. In "direct" action, the controller output signal increases when the controller input exceeds the setpoint. Conversely, with "reverse" action, the controller output signal decreases when the controller input rises. Many controllers have a switch that selects direct or reverse action.

Controller, valve, and process must match with the correct choice of action. The control loop and the valve failure position must be determined first because these two decisions dictate the correct controller action.

There are two basic modes of control:

- Discrete (On/Off): The valve is either fully open or fully closed, with no intermediate state. Alternatively, a manipulated variable such as a motor is "on" or "off."
- Continuous: The valve can move between fully open or fully closed, or held at any intermediate position.

Some common modes of continuous control responses include proportional (throttling), integral (reset), derivative (rate), and combinations of these modes.

14.3.3 On/Off Control

The most simple control strategy, and consequently the least expensive in terms of initial cost and maintenance, is the on/off control. A room thermostat is an excellent example of on/off control. In the field, an on/off control is known as "snap-acting." Level controls on low-pressure separators and temperature controls on indirect fired heaters are often snap-acting. The controller will turn the manipulated variable on when the measurement deviates from setpoint and will turn the manipulated variable off when the measurement is at setpoint. In practical applications, a small error, or dead band, is allowed to eliminate constant cycling of the manipulated variable.

In many cases, the irregular value of the controlled or manipulated variable resulting from a snap-acting control creates serious operating and control problems downstream of the processes. In these cases, a proportional control is used.

14.3.4 Proportional Control

Proportional (P) control is the basic action employed in all controllers not using snap-action. Proportional control is a condition when the change in the controller output (*m*) is proportional to the change in the error (*e*). The constant magnitude of the change in the controller output to the change in error is the controller proportional gain (K_c):

$$m = K_c \cdot e + b \tag{14-6}$$

A bias (*b*) adjusts the output—for instance, 4 milliamps for electronic loops. If the controller output is 0%–100%, then there is no bias.

The proportionality constant is sometimes expressed in terms of percent proportional band (P_b) where

$$P_b = 100/K_c \tag{14-7}$$

In some control systems, proportional action is adequate to meet the control objectives. Proportional control will result in deviations from setpoint except when the setpoint is at 50% of span. When the gain is increased, the error is reduced but is not eliminated. Most processes become unstable as the error is eliminated unless the time constant is extremely high (very slow processes).

14.3.5 Integral Control

Integral (I) action is added to the controller to minimize or eliminate error. Integral control is the condition when the change in the controller output is proportional to the integral of the error. The prime purpose of the integral control is to prevent error and keep the controlled variable at the control point even as the process load changes. The integral algorithm constantly calculates the accumulated error and corrects for past over- or undercompensation with an integral time setting (T_i):

$$m = 1/T_i \cdot \int e\, dt + b \tag{14-8}$$

The integral time expressed in repeats per minute or minutes per repeat is the reset time, representing the time (in minutes) for the integral action to repeat.

Integral-only control is sometimes used in flow control systems and is suitable for pure time-delay processes. There are conditions when an integral controller will continue to accumulate error even when it is not desirable and the output will go to the maximum or minimum value. Some examples of undesirable action are (1) the process is not active, (2) a surge is not encountered for a surge controller, and (3) a cascade master is switched off cascade. An external reset or antireset windup feature is necessary to protect against saturation in the idle state.

Integral mode introduces a lag into the system and is most often used in conjunction with proportional mode. A single algorithm combines these modes where the proportional gain is applied to the integral error as well as the current error. The resulting algorithm is

$$m = K_c * (e + 1/T_i * \int e\,dt) + b \tag{14-9}$$

The integral rate requires careful adjustment. When integral rate is set correctly, valve movement occurs at a rate that the process can respond stably and robustly. If set too fast, cycling will result because the valve moves faster than the process and the measurement. If set too slow, the process will not recover quickly enough, resulting in sluggish control.

Integral action lags proportional action. The proportional action provides the quick response to correct for the upset, whereas the integral action provides the gradual correction to bring the controlled variable back to the setpoint.

Proportional-integral control has the advantage of reducing the initial error that is encountered with integral-only control while eliminating error.

14.3.6 Derivative Control

Proportional plus integral control does not provide correction that is rapid enough for certain processes. A derivative (D) response anticipates a change in process load and transmits a corrective signal to minimize the lag. This action corrects based on the rate of change of the deviation from the setpoint.

Derivative action leads proportional in that it causes the valve to move faster and further than it ordinarily would with proportional action alone. Temperature control systems normally require derivative action. These systems incur large process and measurement lags. Special applications such as antisurge systems also use derivative action because a rapid valve response is imperative.

Derivative control has the property that the change in the controller output (m) is proportional to the rate of change in the error (e). The derivative mode predicts errors and takes corrective action prior to occurrence of the error proportional to the derivative time (T_d):

$$m = T_d\, de/dt \tag{14-10}$$

The derivative time is the duration of time that the algorithm will look into the future. Larger derivative times will contribute larger corrective action. This corrective action will also protect against overshoot of the setpoint.

As the size of processing equipment increases, the momentum makes it difficult to control without derivative action. However, there are several limitations of derivative control. When the change in the error is constant, derivative mode takes no action. In addition, derivative control will react to setpoint changes and step changes in discontinuous measurements, such as chromatographs. For these reasons, derivative control is never used alone. In a few instances, proportional-derivative control may be used, such as special instances of slave controllers in a temperature cascade system and batch pH control.

14.3.7 Proportional-Integral-Derivative Control

While proportional and integral (PI) control can eliminate error and provide stability, considerable time may occur before the error returns to zero. Proportional, integral, and derivative (PID) controllers are used in processes with large capacitance or long time constants or slowly changing process outputs. Temperature and concentration loops are the most common examples.

One algorithm describes PID modes:

$$m = K_c * (e + 1/T_i \cdot \int e\, dt + T_d\, de/dt) + b \qquad (14\text{–}11)$$

This algorithm can simultaneously respond to current error, eliminate error, and anticipate error. Much research has been conducted to determine optimum settings for proportional gain, integral time, and derivative time.

Sample and hold algorithms are used when the dead time of a loop is greater than the time constant. Conventional PID algorithms are not adequate in these instances. This algorithm utilizes the standard PID algorithm part of the time by switching the controller between automatic and manual. The period for which the controller is switched to manual is set by a timer to exceed the dead time and then alternates to automatic for an output update. In order to affect the required magnitude of change, the integral setting must be increased.

14.4 CONTROL LOOP TUNING

Controller tuning, which is the adjustment of the controller parameters to match the dynamic characteristics (or personality) of the entire control loop, has been referred to as the most important, least understood, and most poorly practiced aspect of process control. The tuning of feedback controllers is part of the overall commissioning of plants and machines. Often "trial and error" is used to achieve an acceptable combination of the tuning parameters for a particular process. "Good" control is a matter of definition and depends on such factors as individual preference, process disturbances and interactions, and product specifications.

The following sections discuss control quality and the performance criteria to consider when tuning a controller. They also examine methods of tuning PID controllers and the considerations for selection of the PID settings.

14.4.1 Quality of Control

If the PID controller parameters (the proportional, integral, and derivative terms) are chosen incorrectly, the controlled process input can be unstable; i.e., its output diverges, with or without oscillation, and is limited only by saturation or mechanical breakage. Tuning a control loop is the adjustment of its control parameters (gain/proportional band, integral gain/reset, derivative gain/rate) to the best values for the desired control response.

Good process control begins in the field, not in the control room. Sensors and measurements must be in appropriate locations, and valves must be sized correctly with appropriate trim. The final control elements, such as control valves, execute the changes required to manipulate the preferred process parameters such as flow, temperature, pressure, level, and ratio. If the instruments in the field do not function as required, then one cannot expect the overall process control to perform optimally. Tuning should be modified as the process and equipment change or degrade.

The controllability of a process depends on the gain that can be used. Higher gain yields greater rejection of disturbance and a faster response to setpoint changes. The predominant lag is based on the largest lag in the system. The subordinate lag is based on the dead time and all other lags. The maximum gain that can be used depends on the ratio of the predominant lag to the subordinate lag. From this ratio, we can draw two conclusions: (1) decreasing the dead time increases the maximum gain and the controllability, and (2) increasing the ratio of the longest to the second longest lag increases the controllability. In general, for the tightest loop control, the dynamic controller gain should be as high as possible without causing the loop to be unstable.

Controller tuning requires setting the three constants in the PID controller algorithm to provide control action designed for specific process requirements. The performance of the controller is evaluated in terms of the responsiveness of the controller to an error, the degree to which the controller overshoots the setpoint, and the degree of system oscillation. Note that the use of the PID algorithm for control does not guarantee optimal control of the system.

The best response to a process change or setpoint change varies depending on the application. Some processes must not allow an overshoot of the process variable beyond the setpoint if, for example, this would be unsafe. Other processes must minimize the energy expended in reaching a new setpoint. Generally, stability of response (the reverse of instability) is required, and the process must not oscillate for any combination of process conditions and setpoints. Some processes have a degree of nonlinearity, so parameters that work well at full-load conditions do not work when the process is starting up from no load.

14.4.2 Controller Response

Depending on the process to be controlled, the first consideration is to decide what type of response is best, or at least acceptable. The three possible general extremes of response that exist are overdamped, critically damped, and underdamped. When one is examining the response, there are several common performance criteria used for controller tuning that are based on characteristics of the system's closed-loop response. Some of the more common criteria include overshoot, error, rise time, and decay ratio. Of these simple performance criteria, control practitioners most often use decay ratio.

The decay ratio is the ratio of the amplitude of an oscillation to the amplitude of the preceding oscillation, as shown in Figure 14-2.

Most control engineers believe that for many control loops the optimum tuning is a 1/4 wave decay. This response provides robust disturbance rejection. In fact, the quarter decay ratio typically provides a good trade-off between minimum deviation from the setpoint after an upset and the fastest return to the setpoint.

14.4.3 Controller Performance from Plot Data

Through examination of the response plots, the relative performance of a specific controller can be compared and relative performance can be judged. Some specific performance metrics include rise time, peak overshoot ratio, settling time, and decay ratio. These and similar terms permit comparisons among the range of performance available.

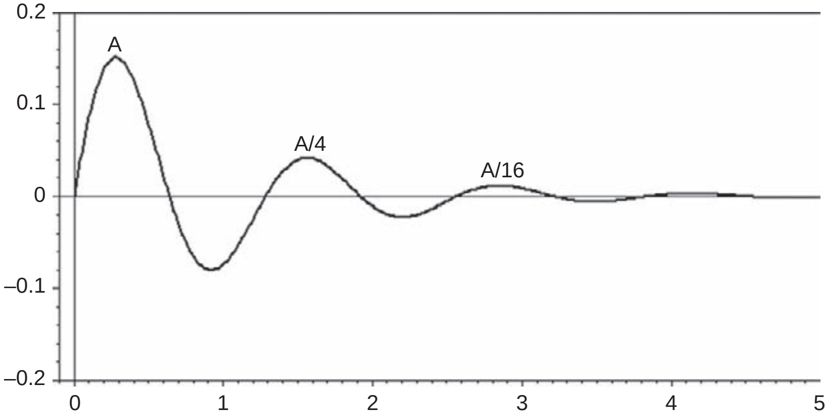


Figure 14-2 Second- or higher-order response to a setpoint change.

14.4.3.1 Peak-Related Criteria

Analysis of the magnitude of difference between setpoint and process variable without regard to time is one method of determining controller performance. These criteria for performance are the peak-related criteria. The popular peak related criteria include:

Peak overshoot ratio (POR) = Height of the first peak (B)/Size of the setpoint step (A)

Decay ratio = Height of the second peak (C)/Height of the first peak (B)

Figure 14-3 shows a setpoint step response plot with labels indicating peak features. In Figure 14-3, the process variable (PV) was initially at 20%, and a setpoint step moves it to 30%. Applying the peak-related criteria by reading off the PV axis: A = (30 − 20) = 10%, B = (34.5 − 30) = 4.5%, and C = (31 − 30) = 1%. For this response, POR = 4.5/10 = 0.45 or 45%, and decay ratio = 1/4.5 = 0.22 or 22%.

An old rule of thumb is that a 10% POR and 25% decay ratio (sometimes called a quarter decay) are popular values. Yet in today's industrial practice, many plants require a "fast response but no overshoot" control performance.

No overshoot means no peaks, and thus, B = C = 0. This increasingly common definition of "good" performance means the peak-related criteria discussed previously are not useful or sufficient as performance-comparison measures.

14.4.3.2 Time-Related Criteria

An additional set of measures focuses on time-related criteria. Figure 14-4 shows the same setpoint response plot, but with the time of certain events labeled.

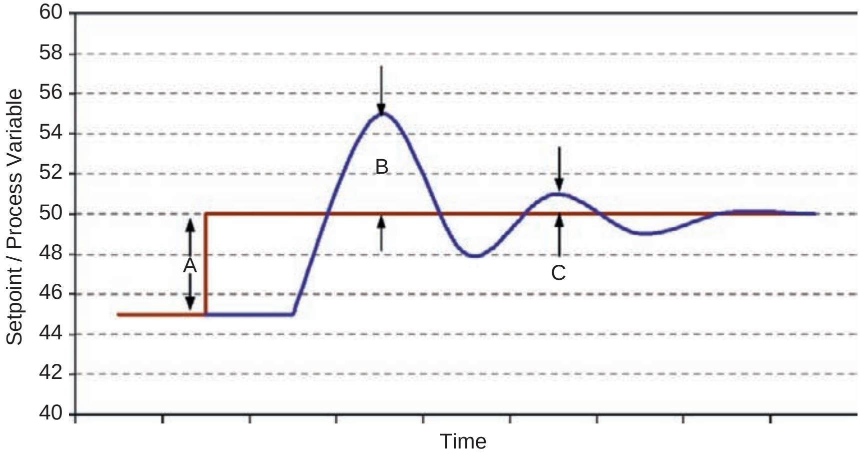


Figure 14-3 Peak-related performance criteria (A = size of the setpoint step, B = height of the first peak, C = height of the second peak).

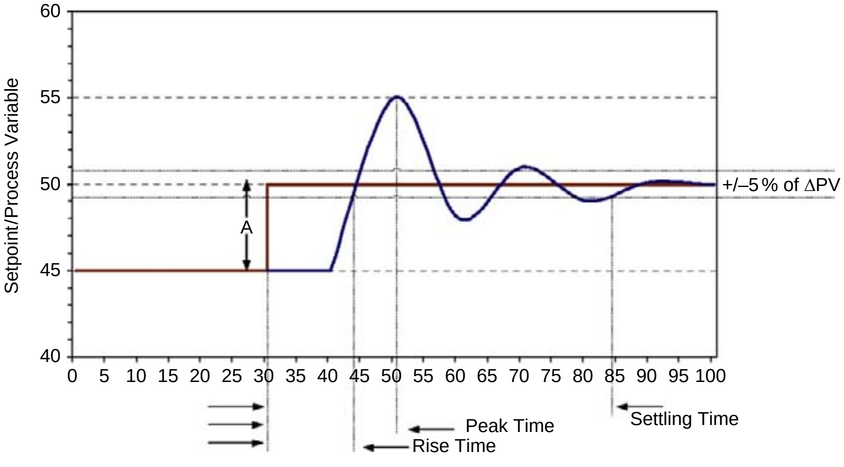


Figure 14-4 Time-related performance criteria.

The clock for time-related events begins when the setpoint is stepped, and as shown in the plot, includes

- Rise time = time until the PV first crosses the setpoint
- Peak time = time to the first peak
- Settling time = time to when a PV first enters and then remains within a band whose width is computed as a percentage of the total change in PV (or ΔPV)

The 5% band used to determine settling time in the preceding plot was chosen arbitrarily. Other percentages are equally valid depending on the situation.

The setpoint was stepped at time $t = 30$ minutes. The time-related criteria are then computed by reading off the time axis as rise time = (44 – 30) = 14 minutes, peak time = (51 – 30) = 21 minutes, and settling time = (85 – 30) = 55 minutes for a ±5% of ΔPV band.

14.4.3.3 When There Is No Overshoot

We should recognize that the peak and time criteria are not independent for the following:

- A process with a large decay ratio will likely have a long settling time.
- A process with a long rise time will likely have a long peak time.

In situations in which we seek moderate tuning with no overshoot in our response plots, there is no peak overshoot ratio, decay ratio, or peak time to compute. Even rise time, with its asymptotic approach to the new steady state, is a measure of questionable value.

In such cases, settling time, or the time to enter and remain within a band of width we choose, remains a useful measure.

The plot illustrated in Figure 14-5 shows the identical process as that in the previous plots. The only difference is that in this case, the controller is tuned for a moderate response.

We compute for this plot:

$$\text{Settling Time} = (68 - 30) = 38 \text{ min for a } \pm 5\% \text{ of } \Delta PV \text{ band}$$

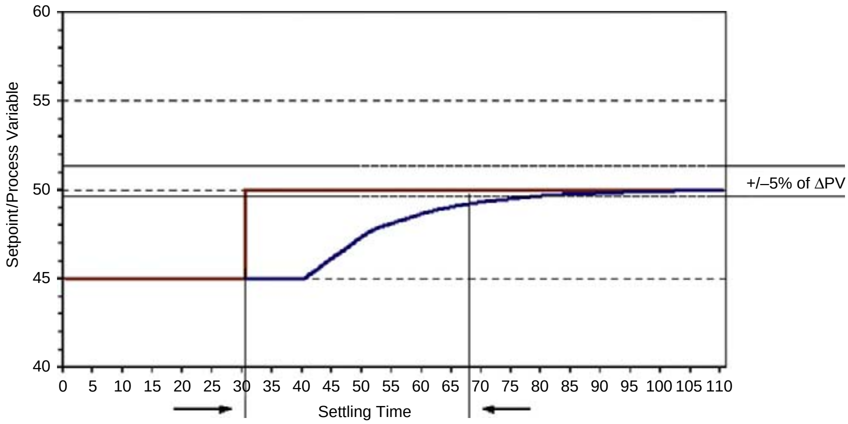


Figure 14-5 Error performance criteria.

14.4.4 Error Performance Criteria

Simple performance criteria, such as decay ratio, use only a few points in the response and are simple to use. However, more complicated error performance criteria describe the entire response of the process. Several criteria or objectives have been proposed. Among the most popular are minimum integral of square error (ISE), minimum integral of absolute error (IAE), and minimum integral of time and absolute error (ITAE).

The integrated square error (ISE) criterion uses the square of the error, thereby penalizing larger errors more than smaller errors. This gives a more conservative response, i.e., faster return to the setpoint. In mathematical terms, with *e* representing the error as a function of time, we can write

$$\mathrm{ISE} = \int_0^\infty e\,(t)^2 dt \tag{14-12}$$

Integrated absolute error (IAE) essentially takes the absolute value of the error:

$$\mathrm{IAE} = \int_0^\infty |e\,(t)| dt \tag{14-13}$$

The integrated time absolute error (ITAE) criterion is the integral of the absolute value of the error multiplied by time. ITAE results in errors existing over time penalized even though they may be small, which results in a more heavily damped response. The mathematical expression for this criterion follows:

$$\mathrm{ITAE} = \int_0^\infty |e\,(t)| \cdot t\, dt \tag{14-14}$$

14.4.5 Tuning Methods

There are a number of methods for tuning single-loop controllers. A few of the methods described are based on simple experiments or simple models and do not require any frequency domain analysis (although such analysis may enhance understanding of the resulting closed-loop behavior).

Many techniques are used for tuning control loops, including experience and a sense for the adequacy of control, heuristics, complex mathematics, and self-tuning systems. Control loop tuning may be accomplished in closed loop where the controller is set on automatic or in open loop where the controller is set on manual.

14.4.5.1 Process Reaction Curve Methods

In process control, the term "reaction curve" is sometimes used to characterize a step response curve. In the process reaction curve methods, a trend is generated in response to a change such as the plot shown in Figure 14-6. This trend reveals the gain, integral time, and derivative time of the process. These methods are performed in an open loop, so no control action occurs and the process response can be isolated.

To generate a process reaction curve, the process reaches steady state or as close to steady state as possible. Then, in an open loop, so there is no control action, one introduces a small disturbance and records the reaction of the process variable.

Methods of process analysis with forcing functions other than a step input are possible and include pulses, ramps, and sinusoids. However, step function analysis is the most common, as it is the easiest to implement.

In this method, the necessary data are generated by introducing a disturbance into the system and analyzing the resulting process reaction curve. When the system reaches steady state, one introduces a disturbance, X_o. The disturbance introduced to the system is a change in either the setpoint or process variable.

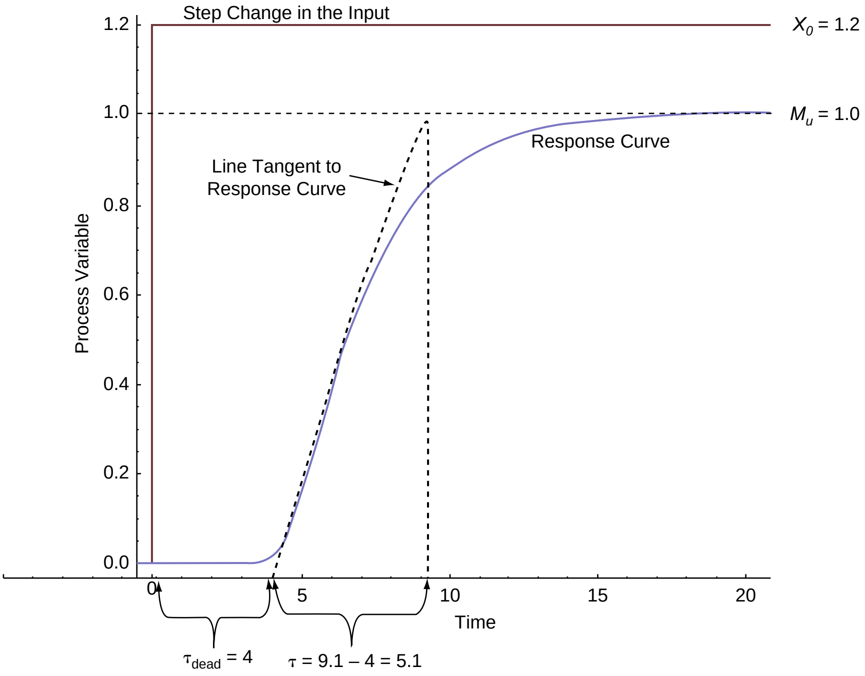


Figure 14-6 Open-loop response to a step change.

The process reaction curve method usually produces a response to a step function change for which several parameters may be measured, revealing transportation lag or dead time, τ_d; the time for the response to change, τ; and the ultimate value that the response reaches at steady state, M_u.

14.4.5.1.1 Ziegler–Nichols Open-Loop Procedure

Ziegler and Nichols (1942) developed controller tuning equations based on field measurements of the ultimate gain and ultimate period. The Ziegler–Nichols tuning method was one of the first formal methods that found wide adoption. This method is also known as the "reaction curve" method. If one wants to use the Ziegler–Nichols method, the process must be stable. With the controller in manual, one changes the output by a small amount outside the noise band. Then one can observe the effect and verify the validity of the dynamic process data collected.

This procedure consists of the two following steps:

1. Determination of the dynamic characteristics, or personality, of the control loop.
2. Estimation of the controller tuning parameters that produce a desired response for the dynamic characteristic determined in the first step; in other words, matching the personality of the controller to that of the other elements in the loop.

In this method, the dynamic characteristics of the process are represented by the ultimate gain of a proportional controller and the ultimate period of oscillation of the loop.

For a manual tuning test, the derivative time is set to zero, and the integral time is set at least 10 times larger than normal so that most of the controller response is from the proportional mode. One increases the controller gain to create equal, sustained oscillations. The controller gain at this point is the ultimate gain (K_u), and the oscillation period is the ultimate period (T_u). In practice, the gain is increased only until decaying oscillations first appear to reduce the disruption to the process. Auto tuners and adaptive controllers can eliminate the need for manual controller tuning. The relay or on/off method is used by auto tuners to automatically compute the ultimate period and gain by switching the controller output when it crosses and departs from a noise band centered on the setpoint (Blevins et al., 2003; McMillan, 2005).

For the Ziegler–Nichols ultimate oscillation method, the controller gain is a fraction of the ultimate gain, and the integral time is a fraction of the ultimate period, as follows for a PI controller.

Defining the open-loop gain as K_o produces

Table 14-1 Tuning Parameters for the Open-loop Ziegler–Nichols Method

	K_c	T_i	T_d
P	K_o		
PI	$0.9K_o$	$3.3\tau_d$	
PID	$1.2K_o$	$2\tau_d$	$0.5\tau_d$

$$K_o = (X_o/M_u) * \tau/\tau_d \qquad (14\text{–}15)$$

The tuning parameters derived by Ziegler and Nichols (1942) are given in Table 14-1.

The advantages of the Ziegler–Nichols open-loop tuning method are

1. It is quick and easy.
2. The process reaction curve method is the least disruptive to implement.

The disadvantages of the Ziegler–Nichols open-loop tuning method are

1. It depends on a purely proportional measurement to estimate I and D parameters.
2. Approximations for the K_c, T_i, and T_d values might not be entirely accurate for different systems.
3. It is not applicable for I, D, and PD controllers.

14.4.5.1.2 Cohen–Coon Tuning Method

Cohen and Coon (1952) modified the Ziegler–Nichols (1942) open-loop tuning rules in 1952. The modifications are insignificant when the dead time is small relative to the time constant but can be important for large dead time. The Cohen and Coon (1952) tuning parameters are as follows:

For P-only control,

$$K_c = (1 + \mu/3)/(\tau_d * RR) \qquad (14\text{–}16)$$

For PI control,

$$K_c = (1 + \mu/11)/(\tau_d * RR) \qquad (14\text{–}17)$$

$$T_i = 3.33 * \tau_d * ((1 + \mu/11)/(1 + 11\mu/5)) \qquad (14\text{–}18)$$

For PID control,

$$K_c = 1.35 * (1 + \mu/3)/(\tau_d * RR) \qquad (14\text{–}19)$$

$$T_i = 2.5 * \tau_d * ((1 + \mu/5)/(1 + 3\mu/5)) \qquad (14\text{–}20)$$

$$T_d = .37 * \tau_d/(1 + \mu/5) \qquad (14\text{–}21)$$

In the preceding equations,

$$\mu = \tau_d/\tau \tag{14-22}$$

$$RR = X_o/M_u * \tau \tag{14-23}$$

As with the Ziegler–Nichols open-loop method recommendations, one can adjust the Cohen–Coon values in closed-loop mode to achieve the quarter decay ratio.

The advantages of the Cohen–Coon method over the Ziegler–Nichols open-loop method are

1. It is useful for systems with a large dead time.
2. It provides faster closed-loop response time.

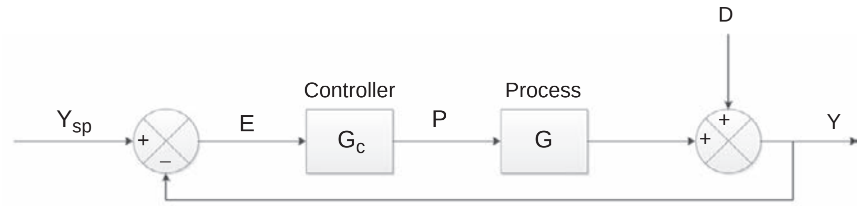
The disadvantages and limitations of the Cohen–Coon method are

1. It can produce an unstable closed-loop system, since this is an open-loop method.
2. It is useful only for first-order models, including large process delays.
3. Approximations for the K_c, τ_i, and τ_d values might not be entirely accurate for different systems.

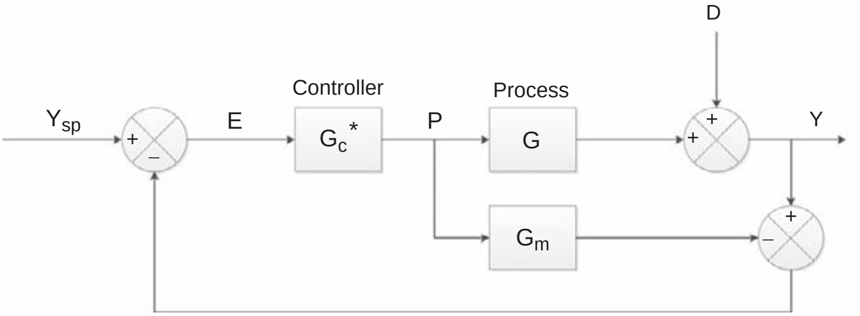
14.4.5.1.3 Internal Model Control Tuning Rules

Internal model control (IMC) tuning rules have proven to be robust and yield acceptable performance when used in the control of common processes. In general, analytical IMC tuning rules are derived for PI and PID compensators by matching an approximate process model to a low-dimensional reference model. The IMC controller structure depends on two factors: the complexity of the model and the performance requirements stated by the designer.

The IMC philosophy relies on the international model principle, which states that a feedback regulator under external disturbances may regain regulation and stability provided a suitably reduplicated model of the disturbance signal is adapted in the feedback path. In other words, if one develops the control scheme based on an exact model of the process, perfect control is theoretically possible. Internal model control uses open-loop step-response Laplace transfer functions with process gain and time constant to predict a measurement change due to a change in setpoint. Use of the IMC philosophy can also generate settings for conventional PI or PID controllers. The algorithm includes a model to bias the setpoint to remove the steady-state error, which becomes a feedback adjustment method. One can filter the biased setpoint to obtain a reference trajectory. A single tuning factor filters the time constant. High tuning factors lead to gentle control action, whereas low tuning factors lead to aggressive control action.



Classical Feedback Control



Internal Model Control

Figure 14-7 Comparison of classical feedback control to internal model control.

Figure 14-7 shows a comparison of the classical feedback control algorithm to the IMC algorithm where Y_{sp} is setpoint, Y is process output, E is setpoint error (the difference between process variable and setpoint), G_c is controller transfer function, P is manipulated input, G is process transfer function, D is system disturbance, G_m is plant model, and G_c^* is internal model controller.

Since the general IMC method is unnecessarily complicated for processes that are well approximated by first-order dead time or integrator dead-time models, simplified IMC rules were developed by Fruehauf et al. for PID controller tuning (Chien and Fruehauf, 1990; Fruehauf et al., 1994).

The IMC-PID tuning rules often apply in industry. However, the widely published IMC tuning rules, while providing adequate suppression of output disturbances, do a poor job suppressing load disturbances when the process dynamics are significantly slower than the desired closed-loop dynamics. Morari and Zafiriou (1989) proposed to address this problem by including an additional integrator in the output disturbance while performing the IMC design procedure. This method was found to provide adequate load disturbance suppression for many processes and has been applied to model predictive control (MPC). However, the resulting controllers do not have PID structure.

IMC models are inherently stationary and linear. Like PID controllers, IMC must be tuned for changes in process gain or time constant. The model should be reparameterized when the process dynamics change substantially.

14.4.5.2 Constant Cycling Methods

14.4.5.2.1 Ziegler–Nichols Closed-Loop Method

The Ziegler–Nichols closed-loop tuning method is probably the most well-known tuning method. It requires a simple closed-loop experiment, using proportional control only. One can increase the proportional gain until a sustained oscillation of the output occurs (neither grows nor decays significantly with time). Then one records the proportional gain that gives the sustained oscillation and the oscillation period (time). The proposed tuning parameters are given in Table 14-2. The ultimate gain, K_u, is the gain at which the oscillations continue with a constant amplitude. The period of these oscillations is the ultimate period, P_u, as shown in Figure 14-8.

In most cases, increasing the proportional gain will provide a sufficient disturbance to initiate the oscillation. Measurement noise may also initiate oscillation. Only if the output is very close to the setpoint will it be necessary

Table 14-2 Tuning Parameters for the Ziegler–Nichols Closed-Loop Method

	Gain	Reset	Derivative
P	$0.5\ K_u$	—	—
PI	$0.45\ K_u$	$1.2/P_u$	—
PID	$0.6\ K_u$	$2/P_u$	$P_u/8$

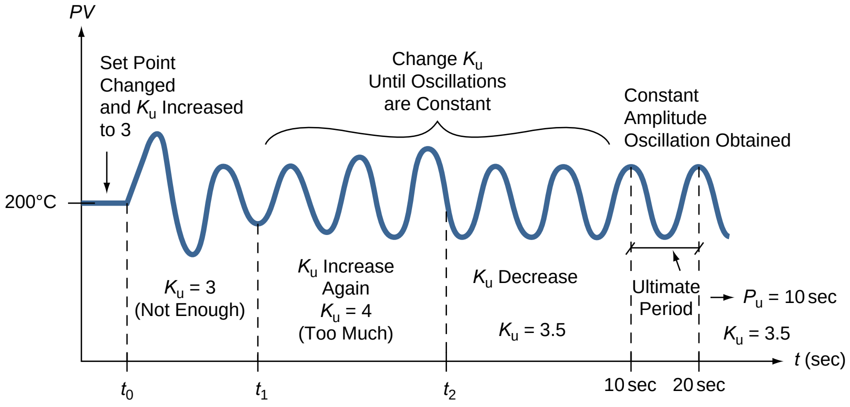


Figure 14-8 Example of determination of the ultimate period.

to introduce a setpoint change after increasing the gain, in order to initiate an oscillation. Note that for controllers giving positive output signals, i.e., controllers giving output signals scaled in the range 0−1 or 0%−100%, a constant bias must be included in the controller in addition to the proportional term, thus allowing a negative proportional term to have an effect. Otherwise, the negative part of the oscillation in the plant input will be cut off and affect the oscillation of the output. In this case, both the input and output of the plant may still oscillate but would show a more complex behavior than the single-frequency sinusoids that the experiment should produce.

The steps required for the Ziegler-Nichols closed-loop methods are

1. Place the controller into automatic with low gain and no reset or derivative.
2. Gradually increase the gain by making small changes in the setpoint until oscillations start.
3. Adjust the gain to force the oscillations to continue with constant amplitude.
4. Note the gain and period.
5. The ultimate gain, G_u, is the gain at which the oscillations continue with a constant amplitude. The period of these oscillations is the ultimate period, P_u.

Advantages of the Ziegler–Nichols closed-loop tuning method are as follows:

1. It is an easy experiment that requires a change only in the gain of a controller.
2. The tuning considers the dynamics of the whole system, which gives a more accurate picture of how the system behaves.

Disadvantages of this method are as follows:

1. The experiment can be time consuming.
2. Unstable regions may be encountered while testing the controller, which could cause the system to become out of control.

14.4.5.3 Autotune Variation Technique

The objective of autotuning methods is to obtain a PID controller capable of satisfying typical requirements such as rapid following, zero steady-state error, and overshoot suppression by means of a practical and robust method. The autotune method has several advantages over open-loop pulse testing methods such that (1) no prior knowledge of the system time constant is needed; and (2) the method is closed-loop tested, so the process will not drift away from the setpoint.

Relay feedback autotuning and frequency domain autotuning by magnitude and phase calculation are some of the most common methods to adjust the parameters of a PID controller. Relay feedback autotuning uses the ultimate frequency. This frequency is a reference, and the design operating frequency will be a fraction of the reference.

Åström and Hägglund (1995) proposed a method for determining the ultimate frequency and ultimate gain commonly called autotune variation (ATV). This method consists of an approximate method called the harmonic balance method based on relay feedback autotuning. Figure 14-9 shows a process transfer function with a feedback ideal relay. In parallel with a feedback relay, there is a PID controller, the achieved virtual controller.

An approximate condition for oscillation can be determined by assuming that there is a limit cycle with period T_u and frequency $w_u = 2\pi/T_u$ such that the relay output is a periodic symmetric square wave. If the relay amplitude is d, a simple Fourier series expansion of the relay output shows that the first harmonic component has the amplitude $4d/\pi$.

The ATV method determines the ultimate gain and period in a manner similar to that of the ultimate method, but ATV tests can be implemented without unduly upsetting the process. Controller settings are calculated, and the controller is then tuned online to meet the selected tuning criterion. The ultimate controller gain, K_u, is calculated by

$$K_u = 4h/\pi a \tag{14-24}$$

where h is the relay height; and a is the amplitude of the process variable. The ultimate period can be directly calculated from the process variable, also.

With K_u and P_u calculated, settings for a PI controller can be calculated using the Ziegler–Nichols method.

An adaptive controller has been achieved through use of the ATV autotuning technique or relay feedback autotuning, which supplies the frequency of the limit cycle and its associated gain, known as ultimate frequency and ultimate gain (Åström and Wittenmark, 1995).

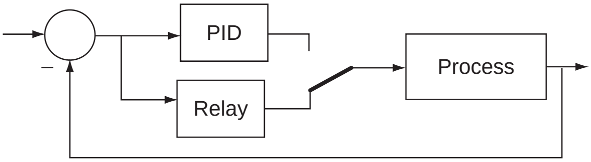


Figure 14-9 Block diagram of a relay autotuner.

14.4.5.4 Lambda Tuning

Another class of tuning method that became popular with the increased use of computing power is lambda tuning. Lambda tuning refers to all tuning methods where the control loop speed of response is a selectable tuning parameter. The closed-loop time constant is "lambda." Lambda tuning, which originated with Dahlin (1968), is based on the same internal model control (IMC) theory as the model predictive control, is model-based, and uses a model inverse and pole-zero cancellation to achieve the desired closed-loop performance.

The lambda tuning equations were developed for simplicity and practicality. For self-regulating processes, the equations are as follows:

$$\lambda = \lambda_f * \tau_1 \tag{14-25}$$

$$T_i = \tau_1 \tag{14-26}$$

$$K_c = T_i/(K_o * (\lambda + \tau_d)) \tag{14-27}$$

$$T_d = \tau_2 \tag{14-28}$$

where K_c is controller gain; K_o is open-loop gain; λ is lambda (closed-loop time constant); λ_f is lambda factor; τ_d is total loop dead time; τ_1 is largest open-loop time constant; τ_2 is second largest open-loop time constant; T_i is integral time setting; and T_d is derivative time setting.

The lambda factor is the ratio of closed-loop time constant to the open-loop time where the open-loop time constant is the largest time constant (τ_1). For maximum load-rejection capability, a lambda equal to the total loop dead time ($\lambda = \tau_d$) can be used, and the loop will still be stable.

The goal of lambda tuning is to match the setpoint response to the first-order time constant or lambda. The response is initially delayed by the process dead time. Since lambda tuning is a model-based method, the tuning parameters are derived from a model of the process. Given a model, the tuning method for an ideal-type PID controller is simply a proper conversion of the units. Parallel- and series-type controllers require different tuning. For PI controllers, series tuning and ideal tuning are the same.

The design concept behind lambda is to cancel the process with the controller and then use a first-order filter to obtain the desired response.

Once the field devices have been checked and corrected as required, an open-loop response test with the controller in manual operation is performed to understand the dynamics of the process. Testing is performed over a range of normal operating conditions. The collected data should be fitted to a simple dynamic model (common models include first order plus dead time and integrator plus dead time).

A unification of lambda, internal model control, and Ziegler-Nichols reaction curve and ultimate oscillation tuning methods has been achieved. The controller tuning equations from diverse methods reduce to a common form, where the maximum controller gain is proportional to the time constant to dead-time ratio (τ_1/τ_d) and is inversely proportional to the open-loop gain (K_o). This common form is easy to remember and provides insight as to the relative effects of process dynamics on tuning and hence on loop performance (Boudreau and McMillan, 2006).

14.4.6 PID Tuning Software

Most modern industrial facilities no longer tune loops using the manual calculation methods discussed previously. Instead, PID tuning and loop optimization software is used to ensure consistent results. These software packages will gather the data, develop process models, and suggest optimal tuning. Some software packages can even develop tuning by gathering data from reference changes.

Mathematical PID loop tuning induces an impulse in the system and then uses the controlled system's frequency response to design the PID loop values. In loops with response times of several minutes, mathematical loop tuning is recommended because trial and error can take too much time to find a stable set of loop values. Optimal values are harder to find. Some digital loop controllers offer a self-tuning feature in which very small setpoint changes are sent to the process, allowing the controller itself to calculate optimal tuning values.

When dynamic process data are generated, it is important that the change in the controller output signal causes a response in the measured process variable that clearly dominates the measurement noise. One way to quantify the amount of noise in the measured process variable is with a noise band. A noise band is based on the standard deviation of the random error in the measurement signal when the controller output is constant and the process is at steady state. Plus or minus three standard deviations of the measurement noise around the steady state of the measured process variable (99.7% of the measurements are contained within the noise band) is conservative when used for controller tuning.

When dynamic process data are generated, the change in controller output should cause the measured process variable to move at least 10 times the size of the noise band. In other words, the signal-to-noise ratio should be greater than 10. For instance, if the noise band is 1°F, then the controller output should be moved enough during a test to cause the measured exit temperature to move at least 10°F.

With today's data historian capabilities, much high-frequency information can be captured for calculating tuning factors. Trending capabilities are also quite helpful for these analyses.

14.4.7 Power Spectrum Analysis

Power spectrum analysis is a technique commonly used by PID tuning software and applies a fast Fourier transform (FFT) to the variation of a particular signal to compute its frequency spectrum. The result is presented as a plot of signal power against frequency and is referred to as its power spectrum. The power spectrum of a signal indicates the relative magnitudes of the frequency components that combine to make up the signal.

The data used to determine the power spectrum must reflect sufficient excitation in the signal. Ideally, the signal should be subject to some form of random excitation in order to generate appropriate data (for example, by the application of a pseudo-random binary sequence to the plant at some appropriate point). The power spectrum is useful for determining the degree of noise that is associated with the signal and also deciding on appropriate sampling rates (following the guidelines of Shannon's sampling theorem).

The power spectrum analysis graphically indicates the frequency content of the PV signal. It does this by performing a fast Fourier transform on the PV signal and plotting the magnitude of each frequency analyzed. Figure 14-10 shows an example of a Power Spectrum plot with frequency. Plots of power over period are more applicable to the process control environment where we would rather think in terms of seconds and minutes instead of hertz, as shown in Figure 14-11.

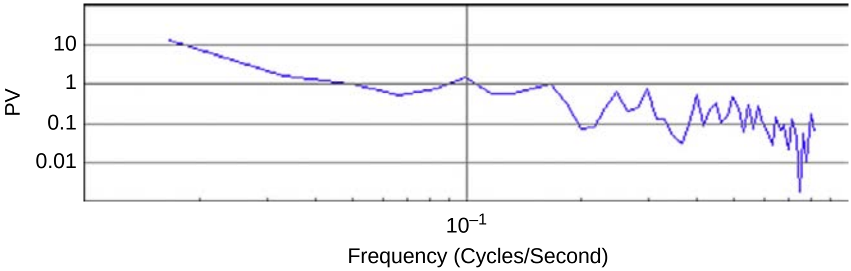


Figure 14-10 Power spectrum (PV) versus frequency.

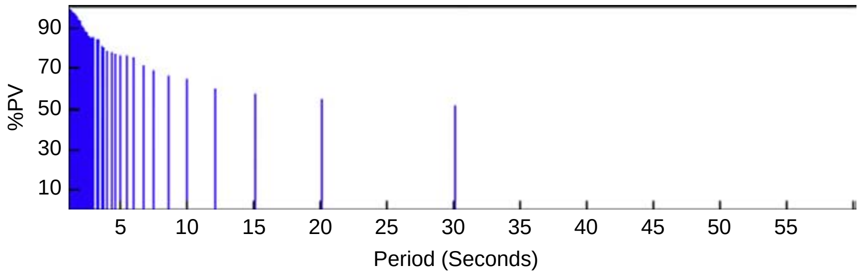


Figure 14-11 Power spectrum (PV) versus period.

The vertical axis of a power spectrum plot is the power detected at each frequency. The span is from 0 Hz to a threshold frequency (f_{thres}) which corresponds to half the logging frequency that is associated with the signal:

$$f_{\text{thres}} = 1/2T \tag{14-29}$$

where T is the log interval in seconds.

According to Shannon's theorem, any information in the signal that relates to a frequency greater than (f_{thres}) cannot be extracted at the particular log interval of sampling. This criterion is, in fact, idealistic, and in practical situations, the threshold frequency is halved in order to guarantee the representative recovery of information:

$$f < f_{\text{limit}} = 1/4T \tag{14-30}$$

A cycle might be completely hidden by the random noise content of the PV signal.

14.4.8 Choosing a Tuning Method

The most popular tuning alternatives employed in today's operations are shown in Table 14-3.

The choice of method will depend largely on whether or not the loop can be tuned offline and the response time of the system. If tuning the system offline, the best method often involves subjecting the system to a step change in input, measuring the output as a function of time, and using this response to determine the control parameters.

If the system must remain online, one tuning method is to first set the I and D values to zero. The P should be increased until the output of the loop oscillates; then the P should be left set to approximately half of that value for a "quarter amplitude decay" type response. Then I should be increased until any error is corrected in sufficient time for the process. However, too much I

Table 14-3 Most Popular Tuning Alternatives

Method	Advantages	Disadvantages
Ziegler–Nichols	Proven online method.	Process upset, some trial and error, very aggressive tuning
Tune by feel	No math required. Online method.	Erratic, not repeatable
Software tools	Consistent tuning. Online or offline method. May include valve and sensor analysis. Allows simulation before downloading.	Some cost and training involved
Cohen–Coon	Good process models.	Some math. Offline method. Only good for first-order processes.

will cause instability. Finally, D should be increased, if required, until the loop is acceptably quick to reach its reference after a load disturbance. However, too much D will cause excessive response and overshoot. A fast PID loop tuning usually overshoots slightly to reach the setpoint more quickly; however, some systems cannot accept overshoot, in which case an "overdamped" tune is required, which will require a P setting significantly less than half that of the P setting causing oscillation.

14.4.8.1 Tuning Flow Loops

Flow loops are too fast to use the standard methods of analysis and tuning. Proper tuning of flow loops depends on whether the control is analog or digital. Some flow loops using analog controllers are tuned with high gain, but this will not work with digital control. With an analog controller, the flow loop has a predominant lag (L) of a few seconds and no subordinate lag. The scan rate of a digital controller can be considered dead time. Although this dead time is small, it is large enough when compared to L to force a low gain. Table 14-4 shows the typical PID settings depending on the type of loop.

14.4.9 Advanced Control Methods

The following sections discuss some advanced control methods, including feedforward control, cascade control, override and selectors, interaction and decoupling, nonlinear control, adaptive control, model predictive control, model-based control, optimizing control, and self-tuning controllers.

Table 14-4 Typical PID Settings for Different Loops

Loop Type	P_b %	Integral min/rep	Integral rep/min	Derivative min	Valve Type
Flow	50–500	0.005–0.05	20–200	none	Linear or modified percentage
Liquid pressure	50–500	0.005–0.05	20–200	none	Linear or modified percentage
Gas pressure	1–50	0.1–50	0.02–10	0.02–0.1	Linear
Liquid level	1–50	1–100	0.1–1	0.01–0.05	Linear or modified percentage
Temperature	2–100	0.2–50	0.02–5	0.1–20	Equal percentage
Chromatograph	100–2,000	10–120	0.008–0.1	0.1–20	Linear

14.4.9.1 Feedforward Control

Feedforward control differs from feedback control in that the load or primary disturbance is measured and the manipulated variable is adjusted so that deviations in the controlled variable from the setpoint are minimized. The controller can then reject disturbances before they affect the controlled variable. For accurate feedforward control, steady-state or dynamic analysis should be the basis for models that relate the effect of the manipulated and disturbance variable on the controlled variable. Since the model is an approximation and not all disturbances are measured, feedforward control should always be used in conjunction with feedback control. This combination will allow compensation for measured and unmeasured disturbances as well as model mismatch.

14.4.9.2 Cascade Control

Processes that respond to disturbances with long time delays or lags are difficult to control with a single feedback controller. One relatively simple way to improve the dynamic response is to use a secondary measurement and controller. The secondary measurement and controller should recognize the effect of a disturbance before the primary controlled variable is affected. Cascade control, where the primary controller (also called master controller) output becomes the setpoint for the secondary controller (also called slave controller), is a readily configured strategy in most computer control systems.

Cascade control is commonly used on distillation towers for composition control. The primary measurement of a key component typically has a time delay, and feedback control is difficult. Temperature of a tray that is sensitive to compositional changes can be used as a secondary measurement. The tray temperature usually responds faster than the analyzer.

In the case of distillation composition control, the response can be further improved by configuring the heat medium flow to a reboiler or reflux rate as a secondary control for the appropriate tray temperature control. As a rule of thumb, the primary loop should be five times faster than the secondary loop.

14.4.9.3 Override and Selectors

One approach to addressing more controlled variables than manipulated variables is to use a selector. The most common types of selective controls are high and low selectors. This allows control action to be based on a selection criterion for multiple measurements.

Override is another type of selective control. In this case, control is dictated by a process variable that reaches a predetermined high or low limit.

14.4.9.4 Interaction and Decoupling

An interactive system exists when a manipulated variable for one controlled variable affects the controlled variable of another loop. Additional feedback loops occur between unpaired manipulated and controlled variables in these cases and destabilize the system.

Relative gain analysis reveals interactions and the degree of interaction. Pressure and level are functions of flow differences, and temperature and composition are functions of flow ratios. These differences and ratios can often be used to decouple and minimize interactions. This technique is rational decoupling. Decoupling is similar to feedforward control except the load variable replaces a manipulated variable.

When ratios and differences are not effective in eliminating interactions, then the controllers can be detuned to restore stability. This detuning will lead to loss in control performance.

Another technique includes linear decouplers calculated from process gains. The issues with this approach are constraints and initialization. When a manipulated variable is constrained, then two controllers compete for an unconstrained variable. One controller will wind down, whereas the other controller winds up. Decoupling using measured values of the manipulated variables overcomes the initialization and constraint issues. Even with these enhancements, decouplers are difficult to match and remain matched with the process, leading again to stability issues.

Partial decoupling can be used to stabilize interactive loops. Best practice is to protect the least likely controlled variable to change setpoint, the slowest, or the most important controlled variable.

Other techniques for decoupling are using an adaptive multivariable controller on feedforward loops and dynamic matrix methods. These methods are described in subsequent sections.

Blending and distillation systems often present interactive systems that require decoupling for effective control.

14.4.9.5 Nonlinear Control

Conventional process control systems utilize linear dynamic models. For highly nonlinear systems, control techniques directly based on nonlinear models provide significantly improved performance.

Most real processes display some nonlinear behavior. The process gain and dead time can change with load, time with equipment degradation, and dead time with transportation lag. In many cases, linear controllers provide adequate control performance. As the degree of nonlinearity increases, then improved control performance may be necessary and desired.

There are two classes of nonlinear control: discontinuous and continuous. The discontinuous methods include on-off and three state devices. These discontinuous methods are adequate only when accurate regulation is not essential. Continuous nonlinear control methods include fuzzy logic, output filtering, characterization, and dead-band or gap action.

14.4.9.6 Adaptive Control

Programmed and self-adaptive controls comprise the two classes of adaptive controls. Programmed adaptive applies when a measured variable yields a predictable response from the control loop. Programmed adjustments are continuous or discontinuous, including gap action and switching controller gains. Adaptive methods include self-tuning, model reference, and pattern recognition.

A classic example of variable, predictable gain on a control loop occurs in plug flow situations commonly found in heat exchangers. At low flow, the dead time, steady-state gain, and dominant time constant are proportionately higher than high flow. For instance, at 50% flow, these parameters are all twice as high as at 100% flow.

The equation for a flow-adapted PID controller is

$$m = K_c * (w * e + w^2/T_i * \int e\,dt + T_d * de/dt) + b \qquad (14\text{-}31)$$

where K_c, T_i, and T_d are the proportional, integral, and derivative settings at full-scale flow; and w is the fraction of full-scale flow.

Gap action, mentioned in the nonlinear control section, can also be adaptive. A dead band is placed around the error used in the calculation of the proportional, integral, and derivative action. No action is taken within the deadband. Level control of surge tanks, pH control, and other systems where stability may be sacrificed with too much control action are typical examples.

Gain scheduling is another form of nonlinear control or programmed adaptation. Variable breakpoint control is another description of this type of control. This method is often similar to gap action. Table 14-5 shows a comparison between gap action and gain scheduling.

Another common method of controller gain adjustments is an error-squared formulation. This produces a nonlinear controller with similar characteristics to the typical implementation of gain scheduling. Little action is taken at small deviations from setpoint compared to the action taken at larger deviations. This method should be used with caution, and limits placed as instability may occur at large deviations from setpoint.

When control loop response is impacted by unknown or immeasurable disturbances, then programmed adaptation is not possible. The industry practices several forms of self-adaptive methods, including self-tuning regulator, model reference adaptive, and pattern recognition.

All self-tuning systems have common elements of an identifier, controller synthesizer, and controller implementation. Types of self-tuning regulators include dead beat, stability theory, fuzzy logic, pole-placement, generalized predictive, and minimum variance algorithms.

The system identifier estimates the parameters of the process. It determines the response of the controlled variable due to a change in the manipulated

Table 14-5 Typical PID Settings for Different Type Loops

Situation	Gap Action	Gain Scheduling
Close to setpoint	No action within a predetermined dead band.	Typically very little action with low controller gain.
Further from setpoint	One set of predetermined PID tuning parameters used outside the dead band.	One or more additional sets of predetermined PID tuning parameters used with higher gain the further the departure from setpoint.

variable. These changes may be deliberately introduced or may be transients that normally occur.

A synthesizer calculates the controller parameters based on a control objective. Recursive estimators determine the optimal proportional, integral, and derivative gains. A function block updates the controller parameters calculated at a predetermined frequency or subject to heuristics.

Model reference adaptive methods are classified as optimal or response specification type algorithms. Minimum variance types employ a least squares method comparing the process variable to its setpoint. The minimum variance adaptive controllers often become unstable when the time delay varies or is unknown. Generalized predictive control uses predictive horizon techniques to overcome the time delay issues experienced with minimum variance methods.

Pole placement and pole assignment routines are common response specification type adaptive controllers. These algorithms use a desired closed-loop frequency to determine preferred controller parameters. This technique requires an excitation of the process that is normally performed in open loop.

Pattern recognition does not use a model to self-adapt a controller. When a change in setpoint occurs, the dead time, sensitivity, and steady-state gain are determined from a positive and negative load response. After identification of the measurement noise, the oscillation, damping ratio, and overshoot are evaluated for desired parameters based on integrated absolute error. Neural network algorithms can also be used for pattern recognition. The radial basis function form of neural networks is ideal due to its ability to analyze time series data.

Optimization can be attempted with single loop controllers using several techniques. However, multivariable algorithms are much more effective for optimization solutions.

Evolutionary optimization and gradient search methods are two approaches with single loop controllers. With evolutionary optimization, online experimentation is conducted to determine the optimum. An independent variable is changed and the response of the optimization objective is measured to determine whether or not there is an improvement. If the response has a negative impact on the objective, then the independent variable will be moved in the opposite direction. If the response has a positive impact on the objective, then the independent variable will continue movement in the same direction. If there is no response, then it is assumed that the optimum has been found.

An example is minimization of a cold separator for a cryogenic natural gas liquids recovery operation. An inlet gas split ratio can be increased or decreased, and the response of the cold separator temperature monitored. If an increase in the split ratio causes an increase in the cold separator temperature, then the split ratio will be decreased. This decrease will continue until there is no change in the cold separator temperature or until the temperature experiences an increase.

Gradient search methods including steepest ascent algorithms are sometimes implemented for optimizing control. Complications arise when a constraint is encountered. Typically, penalty terms for active constraints are imposed on the objective function to force control back within a feasible region.

14.4.9.7 Model Predictive Control

MPC has become the most popular advanced control technique for difficult control problems. The main idea of MPC is to choose the control action by repeatedly solving an optimal control problem online. This technique aims at minimizing a performance criterion over a future horizon, possibly subject to constraints on the manipulated inputs and outputs, where the future behavior is computed according to a model of the plant. Model predictive controllers rely on dynamic models of the process, most often linear empirical models obtained by system identification. The models are used to predict the behavior of dependent variables (i.e., outputs) of a dynamic system with respect to changes in the process-independent variables (i.e., inputs).

If a reasonably accurate dynamic model of the process is available or can be derived, then the model updated with current measurements can be used to predict the future process behavior. The values of the manipulated inputs are calculated so that they minimize the difference between the predicted response of the controlled outputs and the desired response.

Despite the fact that most real processes are approximately linear within only a limited operating window, linear MPC approaches are used in the majority of applications with the feedback mechanism of the MPC compensating for prediction errors due to structural mismatch between the model and the plant. In model predictive controllers that consist only of linear models, the superposition principle of linear algebra enables the effect of changes in multiple independent variables to be added together to predict the response of the dependent variables. This simplifies the control problem to a series of direct matrix algebra calculations that are fast and robust.

Model predictive control offers several important advantages:

1. The manipulated inputs are adjusted based on how they affect the outputs.
2. Inequality constraints on process variables are easily incorporated into the control calculations.
3. The control calculations can be coordinated with the calculation of optimum setpoints.
4. Accurate model predictions provide an indication of equipment problems.

Model predictive control is widely considered the control method of choice for challenging multivariable control problems and normally provides very quick payout.

Identification of the dynamic process model is key to the successful implementation of MPC techniques. In most MPC applications, special plant tests are required to develop empirical dynamic models from input-output data. Although these plant tests can be disruptive, the benefits from implementing MPC generally justify this short-term disruption to normal plant production.

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CHAPTER 15

Process Modeling in the Natural Gas Processing Industry

15.1 INTRODUCTION

Modeling is used in the natural gas industry for simulations of process and equipment design, analysis of system behavior, operator training, leak detection, and design of controllers. The level of model fidelity and rigor depends on the objective. This chapter discusses various modeling techniques, including empirical, rule-based, and first principles, which are commonly used for describing natural gas processes. The presented modeling techniques cover applications for data analysis, expert systems, control systems, and dynamic simulations.

15.2 MODELING AND IDENTIFICATION TECHNIQUES

Traditionally, modeling is a combination of first principles and robust mathematics that yields a usable model. First principles models are more common for steady state but can also be used for dynamic models. The requirement for a detailed description of the first principles is sometimes a limiting factor in practice when complex and poorly understood systems are considered. Difficulties encountered in conventional first principles modeling can arise from poor understanding of the underlying phenomena, inaccurate values of various process parameters, or the complexity of the resulting model. A complete understanding of the underlying mechanisms is virtually impossible for a majority of real systems. However, gathering an acceptable degree of knowledge needed for physical modeling may be a very difficult, time consuming, and expensive or even an impossible task. Even if the structure of the model is determined, a major problem of obtaining accurate values for the parameters remains. It is the task of system identification to estimate the parameters from data measured on the system.

Identification methods are currently developed to a mature level for linear systems only. Empirical models are developed when first principles are difficult to apply. Multivariate regression, principles component analysis, and artificial neural networks are most often used for steady-state empirical models. Some common empirical modeling approaches that can identify dynamics are frequency response, time series, state space, and auto regressive, where the selection of modeling technique depends on purpose.

Most real processes are nonlinear and can be approximated by linear models only locally (Ljung, 1987). Identification of dynamic systems from input/output measurements is an important nonlinear modeling application. There has been much focus on the development of methods for the identification of nonlinear systems from measured data. Artificial neural networks and fuzzy models belong to the most popular model structures used for identifying nonlinear systems.

Regression, artificial neural networks, and elements of fuzzy logic have a "black-box" structure using general function approximators. The modeling problem then reduces to postulating an appropriate structure of the approximator in order to correctly capture the dynamics and nonlinearity of the system. In black-box modeling, the structure of the model is hardly related to the structure of the real system. The identification problem consists of estimating the parameters of the model. If representative process data are available, black-box models usually can be developed quite easily, without requiring process-specific knowledge. A severe drawback of this approach is that the structure and parameters of these models usually do not have any physical significance. Such models cannot be used for analyzing the system's behavior other than by numerical simulation and cannot be scaled up or down when moving from one process scale to another.

There is a range of modeling techniques that attempt to combine the advantages of the first principles and black-box approaches such that the known parts of the system are modeled using physical knowledge, and the unknown or less certain parts are approximated in a black-box manner, using process data and black-box modeling structures with suitable approximation properties. These methods are often denoted as hybrid, semimechanistic modeling.

A common drawback of most standard modeling approaches is that they cannot make effective use of extra information, such as the knowledge and experience of engineers and operators, which is often imprecise and qualitative in its nature. The fact that humans are often able to manage complex tasks under significant uncertainty has stimulated the search for alternative modeling and control paradigms. So-called intelligent modeling and control methodologies, which employ techniques motivated by biological systems and human intelligence to develop models and controllers for dynamic systems, have been introduced. These techniques explore alternative

Table 15-1 Different Modeling Approaches

Modeling Approach	Source of Information	Method of Acquisition	Example	Deficiency
Mechanistic, first principles	Formal knowledge and data	Mathematical (e.g., Lagrange equations)	Differential equations	Cannot use "soft" knowledge
Black-box	Data	Optimization (learning)	Regression, neural network	Cannot use knowledge
Fuzzy	Various knowledge and data	Knowledge-based + learning	Rule-based model	Lack of dimensionality

representation schemes using, for instance, natural language, rules, semantic networks, or qualitative models and possess formal methods to incorporate extra relevant information. Fuzzy modeling and control are typical examples of techniques that make use of human knowledge and deductive processes. Artificial neural networks, on the other hand, realize learning and adaptation capabilities by imitating the functioning of biological neural systems on a simplified level. Some of the different modeling approaches are summarized in Table 15-1.

15.3 ARTIFICIAL INTELLIGENCE AND KNOWLEDGE-BASED SYSTEMS

Knowledge-based systems, also known as expert systems, are a facet of artificial intelligence (AI). These systems emulate the decision-making processes of humans and are one of the most commercially successful AI technologies.

A knowledge-based system may employ any number of approaches to knowledge representation and manipulation from the AI world, including

- **Rule-based systems** capture knowledge in the form of structured if-then statements.
- **Model-based reasoning** uses software models to capture knowledge or to emulate real processes.
- **Neural nets** are a network of nodes and connections used to capture knowledge; they can "learn" by using examples.
- **Fuzzy logic** is used to represent and manipulate knowledge that is incomplete or imprecise.
- **Decision trees** capture decision-making knowledge that can be expressed as sets of order decisions.

15.3.1 Rule-Based Systems

A rule-based system uses heuristics as the knowledge representation for knowledge coded into the system. Rules typically take the form of if–then statements. This is a popular and intuitive knowledge representation. Constraint knowledge, which identifies a set of conditions or a limit, is easily represented using rules. Another form of knowledge, pattern matching, is also a good candidate to be implemented using rules.

15.3.2 Model-Based Reasoning

Model-based reasoning was initially developed to support industrial processes such as gas processes. This technology uses a mathematical model that mimics the real process. Possible control actions can be applied to the model, and the resulting effects can be observed. The model is used to predict the outcomes of various control actions, thus providing a basis for selecting the best control action. One of the key challenges with this technique is ensuring the model has the proper fidelity and captures the important characteristics of the process being modeled.

15.3.3 Artificial Neural Networks

Artificial neural networks were developed from experiments to model the behavior of brain tissue using software. These experiments were some of the earliest forms of artificial intelligence software. Neural networks are good at associative problems. Given partial information, an associative problem is to find items that are associated with the given information. A key advantage of neural networks is that they can be trained with data. Many examples of the desired information can be presented to a neural network to encode knowledge into the neural network. Each example causes the neural network to alter its structure and store the new information. After the neural network is trained, the knowledge is stored in the neural network as a pattern of weights distributed across all the connections between individual neurons. These connected neurons make up the neural network. This easy training is offset by the difficulty in identifying the knowledge stored in the neural network. There is no descriptive form of the knowledge captured in a neural network. The knowledge is only a distribution of connection weights.

A neural network is an interconnected group of nodes, akin to the vast network of neurons in the human brain. Artificial neural networks (ANNs) are nonlinear statistical data modeling tools that can be used to model complex relationships between inputs and outputs or to find patterns in data.

In most cases, an ANN is an adaptive system that changes its structure based on external or internal information that flows through the network during the learning phase.

The main utility of artificial neural network models is that they can be used to infer a function from observations. This is particularly useful in applications in which the complexity of the data or task makes the design of such a function by hand impractical. The tasks to which artificial neural networks are applied tend to fall within the following broad categories:

- Function approximation, or regression analysis, including time series prediction and modeling
- Classification, including pattern and sequence recognition
- Sequential decision making
- Data processing, including filtering, clustering, blind source separation, and compression

Application areas include system identification and process control, data mining (or knowledge discovery in databases), and visualization.

ANNs can be used as an arbitrary function approximation mechanism that "learns" from observed data. The choice of model and the learning algorithm are the two most important aspects of implementing neural networks. The choice of the model depends on the data representation and the application. Overly complex models tend to lead to problems with learning. There are numerous trade-offs between learning algorithms. Almost any algorithm will work well with the correct parameters for training on a particular fixed data set. However selecting and tuning an algorithm for training on unseen data require a significant amount of experimentation. If the model, cost function, and learning algorithm are selected appropriately, the resulting ANN can be extremely robust. With the correct implementation, ANNs can be used naturally in online learning and large data set applications. Their simple implementation and the existence of mostly local dependencies exhibited in the structure allow for fast, parallel implementations in hardware.

15.3.3.1 Background

An artificial neural network contains simple processing elements (neurons) that can exhibit complex global behavior determined by the connections between the processing elements and element parameters. The technique is adapted from the central nervous system and the neurons with their axons, dendrites, and synapses, which constitute one of its most significant information processing elements. In a neural network model, simple nodes (called "neurons," "neurodes," "PEs" ["processing elements"], or "units") are connected together to form a network of nodes—hence, the term "neural network." While a neural network is not required to be adaptive, algorithms designed to alter the strength (weights) of the connections in the network to produce a desired signal flow are the most prevalent and practical.

Artificial neural networks are also similar to the biological neural networks in that functions are performed collectively and in parallel by the units. Currently, the term "artificial neural network" (ANN) tends to refer mostly to neural network models employed in statistics and artificial intelligence.

For software implementations of artificial neural networks, a practical approach based on statistics and signal processing is used. In some of these systems, neural networks or parts of neural networks are used as components in larger systems that combine both adaptive and nonadaptive elements. The algorithms are distributed with parallel local processing and allow identification of nonlinear relationships that are difficult with many other statistical approaches.

15.3.3.2 The Network

The word "network" in the term "artificial neural network" arises because the function $f(x)$ is defined as a composition of other functions $g_i(x)$, which can further be defined as a composition of other functions. This can be conveniently represented as a network structure, with arrows depicting the dependencies between variables. A widely used type of composition is the nonlinear weighted sum:

$$f(x) = K\sum w_i g_i(x) \tag{15-1}$$

where K is some predefined function such as the hyperbolic tangent; w_i are the weight elements; and g_i is a vector.

Figure 15-1 depicts a decomposition of f, with dependencies between variables indicated by arrows. These dependencies can be interpreted in two ways. A functional interpretation describes the input x as transformed into a three-dimensional vector h, which is then transformed into a two-dimensional vector g, which is finally transformed into f. This interpretation is most commonly encountered in the context of optimization. The components of individual layers are independent of each other. The components of g are independent of each other given their input h. This naturally enables a degree of parallelism in the implementation.

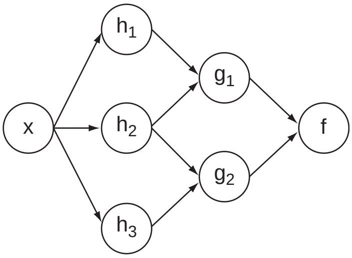

Figure 15-1 Artificial neural network dependency.

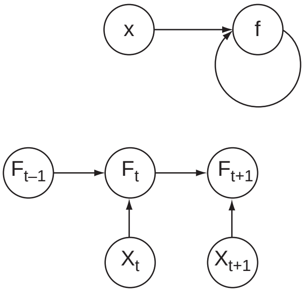

Figure 15-2 Recurrent artificial neural network dependency.

Networks such as that shown in Figure 15-1 are commonly called feed-forward because their graph is unidirectional. Networks with cycles are commonly called recurrent. Such networks are commonly depicted in the manner shown at the top of Figure 15-2, where *f* is shown as being dependent on itself. However, there is an implied dependence that the value of *f* at some point in time *t* depends on the values of *f* at one or more other points in time. The model at the bottom of Figure 15-2 illustrates the case in which the value of *f* at time *t* depends only on its last value. Models such as these, which have no dependencies in the future, are called causal models.

15.3.3.2.1 Learning

The learning capabilities of artificial neural networks allow the identification of nonlinear relationships, which cannot be described conveniently by first principles models. Learning involves using a set of observations to solve a specific task and a class of functions, *F*, in an optimal sense. This requires a cost function and solving for minimum cost to find the optimum. The solution is dependent on some data. The cost must necessarily be a function of the observations; otherwise, the model would not relate to the data. The cost function is frequently defined as a statistic for approximation. A commonly used cost function is the mean squared error, which tries to minimize the average error between the network's output, $f(x)$, and the target value y over all the x–y data pairs. When online learning is used, the cost function is partially minimized as each new set of data is encountered.

There are three major types of learning tasks: supervised learning, unsupervised learning, and reinforcement learning. Any given type of network architecture can be employed in any of these types of learning.

In supervised learning, the aim is to find a function f in the allowed class of functions that matches a given set of data pairs. Minimizing the cost function using gradient descent for the class of neural networks called multilayer perceptrons yields the back propagation algorithm for training neural networks. Other forms of supervised learning are pattern recognition and regression or function approximation.

In unsupervised learning the cost function to be minimized can be any function of the data x and the network's output, f. The cost function is dependent on the model parameters and the observed variables.

In reinforcement learning, data are usually generated by interactions with the environment. At each point in time t, an action y_t is performed, and the environment generates an observation x_t and an instantaneous cost c_t, according to the system dynamics. The objective is to discover actions that minimize some measure of the expected cumulative cost. The environment's dynamics and the cumulative cost for each action are usually unknown but can be estimated. ANNs are frequently used in reinforcement learning as part of an overall algorithm. Control problems are addressed with reinforcement learning.

15.3.3.3 Learning Algorithms

Training a neural network model essentially means selecting one model from the set of allowed models that minimizes the cost criterion. There are numerous algorithms available for training neural network models; most of them can be viewed as a straightforward application of optimization theory and statistical estimation. Most of the algorithms used in training artificial neural networks employ some form of gradient descent. This is performed by taking the derivative of the cost function with respect to the network parameters and then changing those parameters in a gradient-related direction. Evolutionary methods, expectation-maximization, and nonparametric methods are among other commonly used methods for training neural networks.

15.3.3.4 Types of Neural Networks

15.3.3.4.1 Feedforward Neural Network

The feedforward neural network was the first and simplest type of artificial neural network devised. In this network, the information moves in only one direction—forward—from the input nodes, through the hidden nodes (if any), and to the output nodes. There are no cycles or loops in the network.

15.3.3.4.2 Radial Basis Function Network

Radial basis function (RBF) neural networks were developed to identify time series (or dynamic) relationships. This type of neural network has many advantages for control applications. Radial basis functions are powerful techniques for interpolation in multidimensional space. An RBF is a function that has a distance criterion with respect to a center. Radial basis functions have been applied in the area of neural networks where they may be used as a replacement for the sigmoidal hidden layer transfer characteristic in multilayer perceptrons. RBF networks have two layers of processing: in the first, input is mapped onto each RBF in the "hidden" layer. The function is usually a Gaussian. In regression problems, the output layer is then a linear combination of hidden layer values representing mean predicted output. The interpretation of this output layer value is the same as a regression model in statistics. In classification problems, the output layer is typically a sigmoid function of a linear combination of hidden layer values, representing a posterior probability. Performance in both cases is often improved by shrinkage techniques, known as ridge regression in classical statistics and known to correspond to a prior belief in small parameter values leading to smooth output functions in a Bayesian framework. RBF networks also have the advantage of not suffering from local minima in the same way as multilayer perceptrons. The reason is that the only parameters that are adjusted in the learning process are the linear mapping from hidden layer to output layer. Linearity ensures that the error surface is quadratic and therefore has a single easily found minimum. In regression problems, this can be found in one matrix operation. In classification problems, the fixed nonlinearity introduced by the sigmoid output function is most efficiently dealt with using iteratively reweighted least squares.

RBF networks have the disadvantage of requiring good coverage of the input space by radial basis functions. RBF centers are determined with reference to the distribution of the input data but without reference to the prediction task. As a result, representational resources may be wasted on areas of the input space that are irrelevant to the learning task. A common solution is to associate each data point with its own center, although this can make the linear system to be solved in the final layer rather large, and requires shrinkage techniques to avoid overfitting. Associating each input datum with an RBF leads naturally to kernel methods such as support vector machines (SVMs) and Gaussian processes. All three approaches use a nonlinear kernel function to project the input data into a space where the learning problem can be solved using a linear model. Like Gaussian processes and unlike SVMs, RBF networks are typically trained in a maximum likelihood framework by maximizing the probability (minimizing the error) of the data under the model. A support vector machine is a concept in computer science for a set of related supervised learning methods that analyze data and recognize patterns, used for classification and regression analysis. The standard SVM takes a set of input data and predicts its membership class. Given a set of training examples, each marked as belonging to one of two categories, an SVM training algorithm builds a model that assigns new examples into one category or the other. An SVM model is a representation of the examples as points in space, mapped so that the examples of the separate categories are divided by a clear gap that is as wide as possible. New examples are then mapped into that same space and predicted to belong to a category based on which side of the gap they fall on.

RBF networks are outperformed in most classification applications by SVMs. In regression applications, they can be competitive when the dimensionality of the input space is relatively small.

15.3.3.4.3 Recurrent Network

Contrary to feedforward networks, recurrent neural networks (RNs) are models with bidirectional data flow. While a feedforward network propagates data linearly from input to output, RNs also propagate data from later processing stages to earlier stages.

A simple recurrent network (SRN) is a variation on the multilayer perceptron, sometimes called an "Elman network" due to its invention by Jeff Elman. A three-layer network is used, with the addition of a set of "context units," in the input layer. There are connections from the middle, hidden layer to these context units fixed with a weight of one. At each time step, the input is propagated in a standard feedforward fashion, and then a back-propagation learning rule is usually applied. The fixed back connections result in the context units always maintaining a copy of the previous values of the hidden units since they propagate over the connections before the learning rule is applied. Thus, the network can perform such tasks as sequence prediction that are beyond the power of a standard multilayer perceptron.

In a fully recurrent network, every neuron receives inputs from every other neuron in the network. These networks are not arranged in layers. Usually, only a subset of the neurons receives external inputs in addition to the inputs from all the other neurons, and another disjunct subset of neurons reports its output externally as well as sending the output to all the neurons. These distinctive inputs and outputs perform the function of the input and output layers of a feedforward or simple recurrent network and also join all the other neurons in the recurrent processing.

15.3.3.4.4 Committee of Machines

A committee of machines (CoM) is a collection of different neural networks that together "vote" on a given example. This generally gives a much better result compared to other neural network models. In fact, in many cases, starting with the same architecture and training but using different initial random weights gives vastly different networks. A CoM tends to stabilize the result.

15.3.3.5 Dynamic Neural Networks

Dynamic neural networks not only deal with nonlinear multivariate behavior but also include (learning of) time-dependent behavior such as various transient phenomena and delay effects. Regular feedforward perception networks can be generalized with differential equations, using variable time-step algorithms for learning in the time domain and including algorithms for learning in the frequency domain.

15.3.4 Fuzzy Logic

Precise numerical computation with conventional mathematical models makes sense only when the parameters and input data are accurately known. As this is often not the case, a modeling framework is needed that can adequately process not only the given data but also the associated uncertainty. The stochastic approach is a traditional way of dealing with uncertainty. However, it has been recognized that not all types of uncertainties can be dealt with within the stochastic framework. Various alternative approaches have been proposed, fuzzy logic and set theory being among them (Smets et al., 1988).

Fuzzy modeling is a framework in which different modeling and identification methods are combined, providing, on the one hand, a transparent interface with the designer or the operator and, on the other hand, a flexible tool for nonlinear system modeling and control, comparable with other nonlinear black-box techniques. The rule-based character of fuzzy models allows for a model interpretation in a way that is similar to the one humans use. Conventional methods for statistical validation based on numerical data can be complemented by the human expertise that often involves heuristic knowledge and intuition.

From the input-output view, fuzzy systems are flexible mathematical functions that can approximate other functions or just data with a desired accuracy. This property is called general function approximation (Kosko, 1994; Wang, 1994; Zeng and Singh, 1995).

Compared to artificial neural networks, fuzzy systems provide a more transparent representation of the system under study, which is mainly due to the possible linguistic interpretation in the form of rules. The logical structure of the rules facilitates the understanding and analysis of the model in a semiqualitative manner, close to the way humans reason about the real world.

Fuzzy logic has its roots in set theory. It was developed to handle situations in which membership in sets is not clearly defined. Fuzzy models can be used for various aims: analysis, design, control, monitoring, supervision, etc. Approaches have been presented to switch from one model representation to another one, which is more apt for a certain interpretation, allowing a multifaceted use of a model based on one set of data. Rather than as a fully automated identification technique, fuzzy modeling should be seen as an interactive method, facilitating the active participation of the user in a computer-assisted modeling session. This technique is very useful for handling imprecise information. Fuzzy logic can often be combined with other knowledge representations. For example, rules can use fuzzy logic expressions to allow them to more effectively handle imprecise information.

15.3.5 Decision Trees

Decision trees predate computer-based artificial intelligence. This technique has been used for many years to lay out the conditions and steps required for decisions. Decision trees are useful for capturing structured decision-making processes. This technique is useful for troubleshooting and configuration applications. The knowledge for these applications is often structured into a set of steps and decision points.

One problem with this technique is lack of flexibility. Decision trees must be defined ahead of time, thus limiting their flexibility. It is possible to combine decision trees with other AI techniques to lessen this problem. Despite this limitation, decision trees can be very effective representations for specific types of knowledge.

15.4 IMPLEMENTING A KNOWLEDGE-BASED SYSTEM

Knowledge-based systems have their roots in artificial intelligence. Many of the technologies used today to implement knowledge-based systems were originally conceived in the field of artificial intelligence. Software techniques such as rule-based systems or artificial neural networks were first developed as experiments in artificial intelligence. Today these technologies are the building blocks from which knowledge-based systems can be constructed.

15.4.1 Fuzzy Systems

Fuzzy modeling and identification methodologies have been successfully used in a number of applications. Fuzzy models can be used in the design of automatic controllers (Terano et al., 1994), combustion control (Sugeno and Kang, 1986), and pressure control (Babuska et al.,1996). Fuzzy models can also serve as decision support systems to assist operators (Hartog et al., 1997) or can be used to clone the operators based on traces of their behavior (Sugeno and Yasukawa, 1993).

A static or dynamic system that makes use of fuzzy sets or fuzzy logic and of the corresponding mathematical framework is called a fuzzy system. A system can be defined as a collection of if-then rules with fuzzy predicates or as a fuzzy relation. The system can be defined by an algebraic or differential equation in which the parameters are fuzzy numbers instead of real numbers. Fuzzy numbers express the uncertainty in the parameter values.

The input, output, and state variables of a system may be fuzzy sets. Fuzzy inputs can be readings from unreliable sensors ("noisy" data) or quantities related to human perception. Fuzzy systems can process such information, which is not the case with conventional systems.

Fuzzy logic allows for set membership values to range (inclusively) between 0 and 1. Specifically, it allows partial membership in a set. It is related to fuzzy sets and possibility theory and was introduced by Zadeh (1965) at the University of California, Berkeley.

Rule-based fuzzy systems with if-then rules are the most common. Fuzzy truth represents membership in vaguely defined sets. In rule-based fuzzy systems, the relationships between variables are represented by means of fuzzy if-then rules of the following general form:

"If antecedent proposition then consequent proposition."

Rules are usually expressed in the following form:

"IF variable IS set THEN action."

For example, an extremely simple fuzzy logic control for an air cooler that uses a fan might be formulated as follows:

1. IF temperature IS very cold THEN stop fan.
2. IF temperature IS cold THEN slow down fan.
3. IF temperature IS normal THEN maintain fan speed.
4. IF temperature IS hot THEN speed up fan.

These rules contain no "ELSE." All of the rules are evaluated, because the temperature might be "cold" and "normal" at the same time to differing degrees.

The AND, OR, and NOT operators of Boolean logic exist in fuzzy logic, usually defined as the minimum, maximum, and complement; when they are defined this way, they are called the Zadeh operators because they were first defined as such in Zadeh's original papers. So, for the fuzzy variables x and y:

$$\begin{aligned} \text{NOT } x &= (1 - \text{truth}(x)) \\ x \text{ AND } y &= \text{minimum}(\text{truth}(x), \text{truth}(y)) \\ x \text{ OR } y &= \text{maximum}(\text{truth}(x), \text{truth}(y)) \end{aligned}$$

There are also other operators, more linguistic in nature, called hedges, that can be applied. These are generally adverbs such as "small," "very," or "slightly," which modify the meaning of a set using a mathematical formula.

Once fuzzy relations are defined, it is possible to develop fuzzy relational databases.

Some fuzzy querying languages have been defined that include fuzzy aspects in structured query language (SQL) statements, such as fuzzy conditions, fuzzy comparators, fuzzy constants, fuzzy constraints, fuzzy thresholds, and linguistic labels.

The linguistic fuzzy model (Zadeh, 1973; Mamdani, 1977) was introduced as a way to capture available semiqualitative knowledge in the form of if-then rules:

$$R_i\text{: If } x \text{ is } A_i \text{ then } y \text{ is } B_i; \quad i = 1, 2, \ldots., K \tag{15-2}$$

where R is the rule; x is the input (antecedent) linguistic variable; and A_i are the antecedent linguistic terms (constants). Similarly, y is the output (consequent) linguistic variable, and B_i are the consequent linguistic terms. The values of x and y and the linguistic terms A_i and B_i are fuzzy sets defined in the domains of their respective base variables.

The meaning of the linguistic terms is defined by their membership functions, depicted in Figure 15-3.

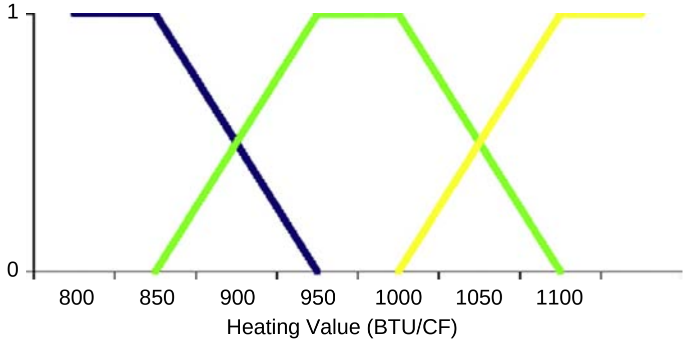

Figure 15-3 Membership functions.

The numerical values along the base variables are selected somewhat arbitrarily. In order to be able to use the linguistic model, we need an algorithm that allows us to compute the output value, given some input value. This algorithm is called the *fuzzy inference* algorithm (or mechanism). For the linguistic model, the inference mechanism can be derived by using fuzzy relational calculus, as shown in the following section.

15.4.2 Multivariable Systems

In the multiple input/multiple output (MIMO) case, all fuzzy sets in the model are defined on vector domains by multivariate membership functions. It is more convenient to write the antecedent and consequent propositions as logical combinations of fuzzy propositions with univariate membership functions. Fuzzy logic operators, such as the conjunction, disjunction, and negation (complement), can be used to combine the propositions. Furthermore, a MIMO model can be written as a set of multiple input/single output (MISO) models.

Most common is the conjunctive form of the antecedent, which is given by

$$\begin{gathered} R_i\text{: If } x_1 \text{ is } A_{i1} \text{ and } x_2 \text{ is } A_{i2} \text{ and } \ldots \text{ and } x_p \text{ is } A_{ip} \\ \text{then } y \text{ is } B_i;\, i = 1, 2, \ldots.\, K \end{gathered} \tag{15-3}$$

The preceding model is a special case of Equation 15-1, as the fuzzy set A_i is obtained as the Cartesian product of fuzzy sets A_{ij}: $A_i = A_{i1} \times A_{i2} \times \cdots A_{ip}$. Hence, the degree of fulfillment (β) is given by

$$\beta_i = \mu_{Ai1}(x_1) \bigwedge \mu_{Ai2}(x_2) \bigwedge \ldots \bigwedge \mu_{Aip}(x_p);\, 1 \leq i \geq K \tag{15-4}$$

where μ = membership degree; and $\wedge$ is a fuzzy conjunction.

Other conjunction operators, such as the product, can be used. A set of rules in the conjunctive antecedent form divides the input domain into a lattice of fuzzy hyperboxes, a Cartesian product-space intersection of the corresponding univariate fuzzy sets, parallel with the axes.

By combining conjunctions, disjunctions, and negations, we can obtain various partitions of the antecedent space; the boundaries are, however, restricted to the rectangular grid defined by the fuzzy sets of the individual variables.

The antecedent form with multivariate membership functions is the most general one, as there is no restriction on the shape of the fuzzy regions. The boundaries between these regions can be arbitrarily curved and opaque to the axes. Also the number of fuzzy sets needed to cover the antecedent space may be much smaller than in the previous cases. Hence, for complex multi-variable systems, this partition may provide the most effective representation.

Another method of reducing the complexity of multivariable fuzzy systems is the decomposition into subsystems with fewer inputs per rule base. The subsystems can be interconnected in a flat or hierarchical (multilayer) structure such as that shown in Figure 15-4. In such a case, an output of one rule base becomes an input to another rule base. This cascade connection will lead to the reduction of the total number of rules. As an example, suppose five linguistic terms for each input. Using the conjunctive form, each of the two subrule bases will have $5^2 = 25$ rules. This is a significant saving compared to a single rule base with three inputs, which would have $5^3 = 125$ rules.

15.4.3 Defuzzification

In many applications, a crisp output y is desired. To obtain a crisp value, we must “defuzzify” the output fuzzy set. With the Mamdani inference scheme, the center of gravity (COG) defuzzification method is used. This method computes the y coordinate of the center of gravity of the area under the fuzzy set B':

$$y' = \mathrm{COG}(B') = \frac{\sum_{j=1}^{F} \mu B'(y_j) y_j}{\sum_{j=1}^{F} \mu B'(y_j)} \tag{15-5}$$

where F is the number of elements y_j in Y. Continuous domain Y must be discretized to be able to compute the center of gravity.

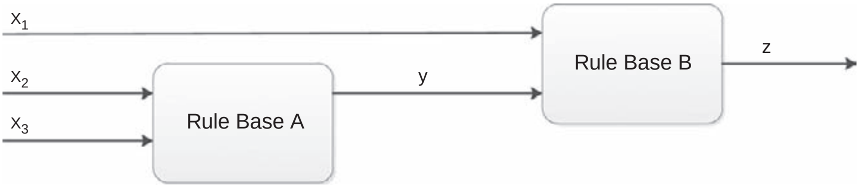

Figure 15-4 Cascade connection of two rule bases.

Consider the output fuzzy set $B = [0{:}2; 0{:}2; 0{:}3; 0{:}9; 1]$, where the output domain is $Y = [0; 25; 50; 75; 100]$. The defuzzified output obtained by applying Equation 15–5 is

$$y = (0.2 \times 0 + 0.2 \times 25 + 0.3 \times 50 + 0.9 \times 75 + 1 \times 100)/(0.2 + 0.2 + 0.3 + 0.9 + 1) = 72.12$$

15.4.3.1 Singleton Model

A special case of the linguistic fuzzy model is obtained when the consequent fuzzy sets β_i are singleton fuzzy sets. These sets can be represented simply as real numbers b_i, yielding the following rules:

$$R_i : \text{If } \bar{x} \text{ is } A_i \text{ then } y = b_i, \mathrm{i} = 1, 2, \ldots., K \tag{15-6}$$

This model is called the singleton model. A simplified inference/defuzzification method is typically used with this model:

$$y = \frac{\sum_{i=1}^{K} \beta_i b_i}{\sum_{i=1}^{K} \beta_i} \tag{15-7}$$

15.4.3.2 Takagi–Sugeno Model

The linguistic model describes a given system by means of linguistic if–then rules with fuzzy proposition in the antecedent as well as in the consequent. The Takagi–Sugeno (TS) fuzzy model (Takagi and Sugeno, 1985) uses crisp functions in the consequents. Hence, it can be seen as a combination of linguistic and mathematical regression modeling in the sense that the antecedents describe fuzzy regions in the input space in which consequent functions are valid. The Takagi–Sugeno model has often been employed in the modeling and identification of nonlinear technical processes from data. The TS rules have the following form:

$$R_i\text{: If } \bar{x} \text{ is } A_i \text{ then } y = f_i(x), \mathrm{i} = 1,\ 2, \ldots., K \tag{15-8}$$

Contrary to the linguistic model, the input x is a crisp variable. The functions f_i are typically of the same structure, except the parameters in each rule are different. A simple and practically useful parameterization is the linear form, yielding these rules:

$$R_{\mathrm{i}}\text{: If } \bar{x} \text{ is } A_{\mathrm{i}} \text{ then } y_{\mathrm{i}} = a_{\mathrm{i}}^{\mathrm{T}}\, x + b_{\mathrm{i}}, \mathrm{i} = 1, 2, \ldots., K \tag{15-9}$$

where a_{i} is a parameter vector; and b_{i} is a scalar offset. This model is called an affine TS model.

Note that if $a_{\mathrm{i}} = 0$ for each i, the singleton model (Equation 15-6) is obtained.

The inference formula of the TS model is an extension of the singleton model inference (Equation 15-7):

$$y = \frac{\sum_{i=1}^{K} \beta_{\mathrm{i}} y_{\mathrm{i}}}{\sum_{i=1}^{K} \beta_{\mathrm{i}}} = \frac{\sum_{i=1}^{K} \beta_{\mathrm{i}} a_{\mathrm{i}}^{\mathrm{T}} X + b_{\mathrm{i}}}{\sum_{i=1}^{K} \beta_{\mathrm{i}}} \tag{15-10}$$

The affine TS model can be regarded as a quasilinear system. The normalized degree of fulfillment is denoted by

$$\gamma_{\mathrm{i}}(x) = \frac{\beta_{\mathrm{i}}(x)}{\sum_{j=1}^{K} \beta_{\mathrm{j}}(x)} \tag{15-11}$$

15.5 MODELING DYNAMIC SYSTEMS

Time-invariant dynamic systems are in general modeled with static functions by using the concept of the system's state. Given the state of a system and given its input, we can determine what the next state will be. In the discrete-time setting, we can write

$$x(k+1) = f(x(k), u(k)); \tag{15-12}$$

where $x(k)$ and $u(k)$ are the state and the input at time k, respectively; and f is a static function, called the state-transition function. Different fuzzy model types can be used to approximate the state-transition function. As the state of a process is often not measured, input-output modeling is usually applied.

The most common is the NARX (Nonlinear AutoRegessive with eXogenous input) model:

$$y(k+1)=f(y(k),y(k-1),\ldots,y(k-n_{\mathrm{y}}+1),u(k),\\ u(k-1),\ldots,u(k-n_{\mathrm{u}}+1)) \tag{15-13}$$

where $y(k), \ldots, y(k-n_{\mathrm{y}}+1)$ and $u(k), \ldots, u(k-n_{\mathrm{u}}+1)$ denote the past model outputs and inputs, respectively; and n_{y}, n_{u} are integers related to the model order.

For example, a linguistic fuzzy model of a dynamic system may consist of rules of the following form:

$$\begin{aligned} R_{\mathrm{i}}:\ &\mathbf{If}\ y(k)\ \text{is}\ A_{\mathrm{i1}}\ \mathbf{and}\ y(k-1)\ \text{is}\ A_{\mathrm{i2}}\ \mathbf{and},\ldots y(k-n+1)\ \text{is}\ A_{\mathrm{in}}\\ &\mathbf{and}\ u(k)\ \text{is}\ B_{\mathrm{i1}}\ \mathbf{and}\ u(k-1)\ \text{is}\ B_{\mathrm{i2}}\ \mathbf{and},\ldots,u(k-m+1)\ \text{is}\ B_{\mathrm{im}}\\ &\mathbf{then}\ y(k+1)\ \text{is}\ C_{\mathrm{i}} \end{aligned} \tag{15-14}$$

In this sense, we can say that the dynamic behavior is taken care of by external dynamic filters added to the fuzzy system in Figure 15-5. In Equation 15-14, the input dynamic filter is a simple generator of the lagged inputs and outputs, and no output filter is used.

Since the fuzzy models can approximate any smooth function to any degree of accuracy (Wang, 1992), models formulated per Equation 15-14 can approximate any observable and controllable modes of a large class of discrete-time nonlinear systems (Leonaritis and Billings, 1985).

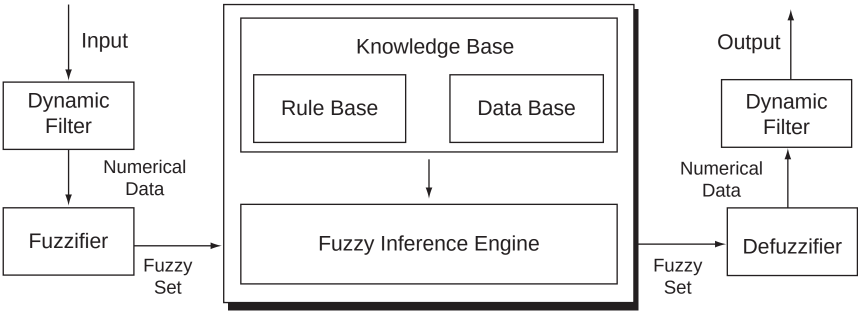

Figure 15-5 Generic fuzzy system with fuzzification, defuzzification, and external dynamic filters.

15.5.1 Knowledge-Based Design

To design a (linguistic) fuzzy model based on available expert knowledge, one can use the following steps:

1. Select the input and output variables, the structure of the rules, and the inference and defuzzification methods.
2. Decide on the number of linguistic terms for each variable and define the corresponding membership functions.
3. Formulate the available knowledge in terms of fuzzy if-then rules.
4. Validate the model (for instance, by using data). If the model does not meet the expected performance, iterate on the preceding design steps.

It should be noted that the success of this method heavily depends on the problem at hand and the extent and quality of the available knowledge. For some problems, the knowledge-based design may lead to useful models quickly; for others, it may be a very time-consuming and inefficient procedure (especially manual fine-tuning of the model parameters). Therefore, it is useful to combine the knowledge-based design with a data-driven tuning of the model parameters.

The following sections review several methods for the adjustment of fuzzy model parameters by means of data.

15.5.2 Data-Driven Development of Fuzzy Models

In this section, we assume that a set of N input-output data pairs $\{(x_i, y_i) \mid i = 1,2, \ldots, N\}$ is available. Input vectors are $x_i \in \mathbf{R}p$ or a set of real numbers $\mathbf{R}$, and y_i are output scalars. Denote $X \in \mathbf{R}Nxp$ a matrix having the vectors x_k^T in its rows, and $y \in \mathbf{R}N$ a vector containing the outputs y_k:

$$X = [x_1, \ldots, x_N]^T, y = [y_1, \ldots, y_N]^T \qquad (15\text{-}15)$$

15.5.2.1 Least-Squares Estimation of Consequents

The defuzzification formulas of the singleton and TS models, Equations 15-7 and 15-10, respectively, are linear in the consequent parameters a_i and b_i. Hence, these parameters can be estimated from the available data by least-squares techniques. We can denote $\Gamma_i \in \mathbf{R}NxN$ the diagonal matrix having the normalized membership degree $\gamma_i(x_k)$ of Equation 15-11 as its k^{th} diagonal element. By appending a unitary column to X, we create the extended matrix $X_e = [X,1]$. Further, we denote X' as the matrix in $\mathbf{R}NxKN$ composed of the products of matrices Γ_i and X_e so

$$X' = [\Gamma_1 X_e, \Gamma_2 X_e, \ldots, \Gamma_K X_e] \qquad (15\text{-}16)$$

The consequent parameters a_i and b_i are lumped into a single parameter vector $\theta \in \mathbf{R}K(p+1)$:

$$\theta = \left[a_1^T, b_1, a_2^T, b_2, \ldots a_K^T, b_K\right]^T \tag{15-17}$$

Given the data X, y, Equation 15-10 can now be written in a matrix form, $y = X'\theta + \varepsilon$. The parameter θ can be solved by

$$\theta = [(X')^T X'\}^{-1}(X')^T y \tag{15-18}$$

This is an optimal least-squares solution that gives the minimal prediction error, and as such is suitable for prediction models. At the same time, however, it may bias the estimates of the consequent parameters as parameters of local models. If an accurate estimate of local model parameters is desired, a weighted least-squares approach applied per rule may be used:

$$\left[a_i^T, \mathrm{b_i}\right]^T = \left[X_e^T\, \Gamma_i X_e\right]^{-1} X_e^T \Gamma_i \mathrm{y} \tag{15-19}$$

In this case, the consequent parameters of individual rules are estimated independently of each other and therefore are not "biased" by the interactions of the rules. When we omit a_i for all $1 \le \mathrm{i} \le K$, Equations 15-18 and 15-19 directly apply to the singleton model (Equation 15-6).

15.6 ARTIFICIAL NEURAL NETWORK APPLICATION

Artificial neural network models have a property called "capacity," which roughly corresponds to their ability to model any given function. It is related to the amount of information that can be stored in the network and to the notion of complexity.

Convergence depends on a number of factors:

1. There may exist many local minima depending on the cost function and the model.
2. The optimization method used might not guarantee convergence when far away from a local minimum.
3. For a very large amount of data or large number of parameters, some methods become impractical.

In general, it has been found that theoretical guarantees regarding convergence are not always a very reliable guide to practical application.

15.6.1 Generalization and Statistics

In applications in which the goal is to create a system that generalizes well in unseen examples, there is sometimes a problem of overtraining. This arises in

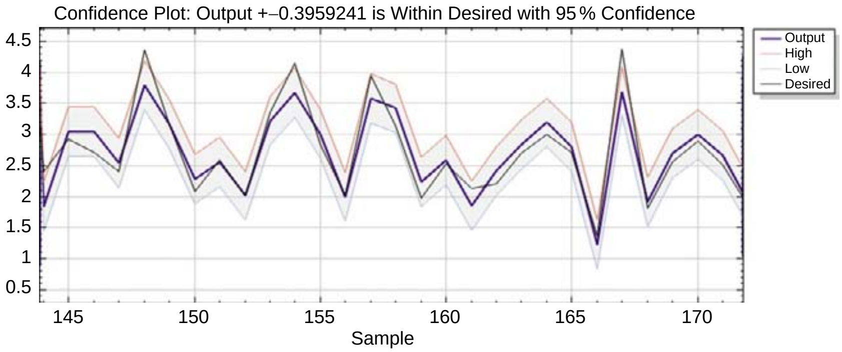

Figure 15-6 Confidence analysis of a neural network.

overcomplex or overspecified systems when the capacity of the network significantly exceeds the needed free parameters.

There are two approaches for avoiding this problem:

1. Use cross-validation and similar techniques to check for the presence of overtraining and optimally select hyper-parameters such as to minimize the generalization error.
2. Use some form of regularization. This is a concept that emerges naturally in a probabilistic (Bayesian) framework, where the regularization can be performed by putting a larger prior probability over simpler models; but also in statistical learning theory, where the goal is to minimize over two quantities: the "empirical risk" and the "structural risk," which roughly correspond to the error over the training set and the predicted error in unseen data due to overfitting.

Supervised neural networks that use a mean squared error (MSE) cost function can use formal statistical methods to determine the confidence of the trained model. The MSE on a validation set can be used as an estimate for variance. This value can then be used to calculate the confidence interval of the output of the network, assuming a normal distribution. A confidence analysis such as shown in Figure 15-6 is statistically valid as long as the output probability distribution stays the same and the network is not modified.

15.7 TIME SERIES MODELS

A time series is a sequence of data points measured at successive times and spaces. Time series analysis comprises methods that attempt to understand the dynamic properties of a system and to make forecasts or predictions.

Time series prediction is the use of a model to predict future events based on known past events: to predict future data points before they are measured.

Frequency response is the measure of any system's response at the output to a signal of varying frequency and constant amplitude at its input. The frequency response is typically characterized by the magnitude of the system's response and the phase versus frequency. The frequency response of a system can be measured by applying an impulse to the system and measuring its response.

Dynamic models for time series data can have many forms and represent different stochastic processes. When the mean of a process is modeled, two broad classes of practical importance are the moving average models and autoregressive models. These two classes depend linearly on previous data points. State-space models can be nonlinear.

Empirical dynamic models are primarily used for model predictive control systems. The main forms of empirical dynamic models for this purpose are finite impulse response, autoregressive moving average, and state-space. Model selection is often based on the underlying assumption on the data-generating process.

15.7.1 Finite Impulse Response Models

Finite impulse response models are based on finite impulse response (FIR) filters, which are a type of a signal processing filter whose impulse response is of finite duration because it settles to zero in finite time. The impulse response of an Nth-order discrete-time FIR filter lasts for $N+1$ samples and then dies to zero.

The output y of a linear time invariant system is determined by convolving its input signal x with its impulse response b.

For a discrete-time FIR filter, the output is a weighted sum of the current and a finite number of previous values of the input. The operation is described by the following equation, which defines the output sequence $y[n]$ in terms of its input sequence $x[n]$:

$$y[n] = b_0 x[n] + b_1 x[n-1] + \ldots + b_{\mathrm{N}} x[n-N] \tag{15-20}$$

or

$$y[n] = \sum_{i=0}^{N} b_{\mathrm{i}} X[n-\mathrm{i}] \tag{15-21}$$

where $x[n]$ is the input signal; $y[n]$ is the output signal; b_{i} are the filter coefficients that make up the impulse response; and N is the filter order.

A FIR filter has a number of useful properties, which are valuable for modeling. FIR filters

- Require no feedback
- Are inherently stable
- Can easily be designed to be linear

The impulse response $h[n]$ can be calculated if we set $x[n] = \delta[n]$ in Equations 15-20 and 15-21, where $\delta[n]$ is the Kronecker delta impulse. The impulse response for a FIR filter then becomes the set of coefficients b_n, as follows:

$$h[n] = \sum_{i=0}^{N} b_{\mathrm{i}}\delta[n-i] = b_{\mathrm{n}} \tag{15-22}$$

for $n = 0$ to N.

The Z-transform of the impulse response yields the transfer function of the FIR filter:

$$H(z) = Z\{h[n]\} = \sum_{n=0}^{N} b_{\mathrm{n}} z^{-\mathrm{n}} \tag{15-23}$$

15.7.2 Autoregressive Moving Average Models

ARMA models are another technique often applied to time series data. An autoregressive model is essentially an infinite impulse response filter with some additional interpretation placed on it.

Given a time series of data X_{t}, the ARMA model is a tool for understanding and predicting future values in this series. The model consists of two parts: an autoregressive (AR) part and a moving average (MA) part. The model is usually then referred to as the ARMA(p,q) model, where p is the order of the autoregressive part and q is the order of the moving average part.

The notation AR(p) refers to the autoregressive model of order p. The AR(p) model is written

$$X_{\mathrm{t}} = c + \sum_{i=1}^{p} \varphi_{\mathrm{i}} X_{\mathrm{t}-1} + \varepsilon_{\mathrm{t}} \tag{15-24}$$

where φ are the parameters of the model; c is a constant; and ε_{t} is an error term.

The notation MA(*q*) refers to the moving average model of order *q*:

$$X_t = \mu + \varepsilon_t + \sum_{i=1}^{q} \theta_i \varepsilon_{t-i} \tag{15-25}$$

where $\theta_1, \ldots, \theta_q$ are the parameters of the model; μ is the expectation of X_t; and ε_t, $\varepsilon_{t-1}, \ldots$ are the error terms. The moving average model is essentially a finite impulse response filter with some additional interpretation placed on it.

The notation ARMA(*p*,*q*) refers to the model with *p* autoregressive terms and *q* moving average terms. This model contains the AR(*p*) and MA(*q*) models:

$$X_t = c + \sum_{i=1}^{p} \varphi_i X_{t-1} + \varepsilon_t + \sum_{i=1}^{q} \theta_i \varepsilon_{t-i} \tag{15-26}$$

The error terms ε_t are generally assumed to be independent identically distributed random variables sampled from a normal distribution with zero mean: $\varepsilon_t \sim N(0,\sigma^2)$, where σ^2 is the variance.

The model parameters may be calculated using least-squares regression. It is generally considered good practice to find the smallest values of *p* and *q* that provide an acceptable fit to the data.

ARMA is appropriate when a system is a function of a series of unmeasured disturbances (identified by the MA part) as well as its own behavior.

15.7.3 Autoregressive Moving Average with Exogenous Inputs Model

The notation ARMAX(*p*,*q*,*b*) refers to the model with *p* autoregressive terms, *q* moving average terms, and *b* exogenous inputs terms. This model contains the AR(*p*) and MA(*q*) models and a linear combination of the last *b* terms of a known and external time series d_t. It is given by

$$X_t = c + \sum_{i=1}^{p} \varphi_i X_{t-1} + \varepsilon_t + \sum_{i=1}^{q} \theta_i \varepsilon_{t-i} + \sum_{i=1}^{q} \eta_i d_{t-i} \tag{15-27}$$

where $\eta_1, \ldots, \eta_b$ are the parameters of the exogenous input d_t.

15.8 STATE-SPACE MODELS

A state-space model is a mathematical representation of a physical system as a set of input, output, and state variables related by first-order differential equations. "State space" refers to the space whose axes are the state variables, and the state of the system can be represented as a vector within that space.

The input, output, and state variables are expressed as vectors, and the differential and algebraic equations are written in matrix form when the dynamic system is linear and time invariant. The state-space representation is a time-domain approach providing a convenient and compact way to model and analyze systems with multiple inputs and outputs. With p inputs and q outputs, we would otherwise have to write down Laplace transforms to encode all the information about a system. Unlike the frequency domain approaches, the use of the state-space representation is not limited to systems with linear components and zero initial conditions.

15.8.1 Linear Systems

The most general state-space representation of a linear system with p inputs, q outputs, and n state variables is written in the following form:

$$\frac{dx}{dt} = A(t)x(t) + B(t)u(t) \tag{15-28}$$

$$y(t) = C(t)x(t) + D(t)u(t) \tag{15-29}$$

where x is the state vector; y is the output vector; u is the control vector; A is the state matrix; B is the input matrix; C is the output matrix; and D is the feedforward matrix, as shown in Figure 15-7.

All matrices are allowed to be time variant. The time variable t can be continuous or discrete. The block with 1/s in Figure 15-7 is an integrator. Depending on the assumptions taken, the state-space model representation can assume the forms described next.

15.8.1.1 Feedback

A common method for feedback is to multiply the output by a matrix K and set this as the input to the system: $u(t) = Ky(t)$. The corresponding block

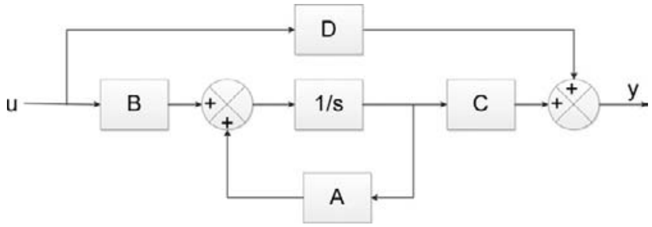

Figure 15-7 Block diagram representation of state-space equations.

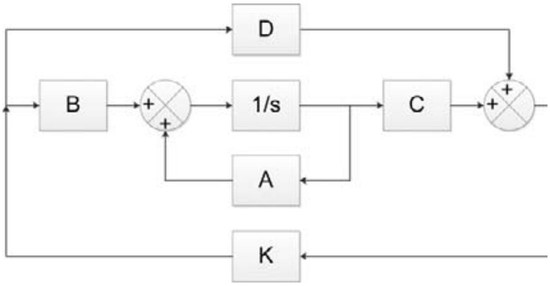

Figure 15-8 Typical state-space model with feedback.

diagram is shown in Figure 15-8. Since the values of K are unrestricted, the values can easily be negated for negative feedback.

Equations 15-28 and 15-29 become

$$\frac{dx}{dt} = Ax(t) + BKy(t) \tag{15-30}$$

$$y(t) = Cx(t) + DKy(t) \tag{15-31}$$

Solving the output equation for $y(t)$ and substituting in the state equation results in

$$\frac{dx}{dt} = (A + BK(I - DK)^{-1}C)x(t) \tag{15-32}$$

$$y(t) = (I - DK)^{-1}Cx(t) \tag{15-33}$$

The advantage of this is that the eigenvalues of A can be controlled by setting K appropriately through eigen-decomposition of $A + BK(I - DK)^{-1}C$. This assumes that the closed-loop system is controllable or that the unstable eigenvalues of A can be made stable through appropriate choice of K.

One fairly common simplification to this system is removing D and setting C to identity, which reduces the equations to

$$\frac{dx}{dt} = (A + BK)x(t) \tag{15-34}$$

$$y(t) = x(t) \tag{15-35}$$

This reduces the necessary eigen-decomposition to just $A + BK$.

15.8.1.2 Feedback with Setpoint (Reference) Input

In addition to feedback, an input, $r(t)$, can be added, as shown in Figure 15-9, such that

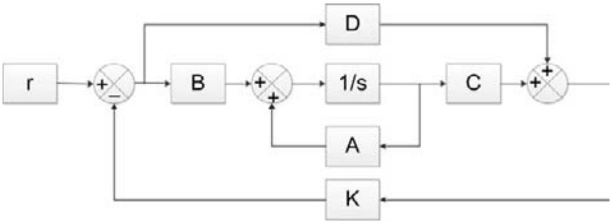

Figure 15-9 Output feedback with setpoint.

$$u(t) = -Ky(t) + r(t) \tag{15-36}$$

Equations 15-28 and 15-29 become

$$\frac{dx}{dt} = Ax(t) - BKy(t) + Br(t) \tag{15-37}$$

$$y(t) = Cx(t) - DKy(t) + Dr(t) \tag{15-38}$$

Solving the output equation for $y(t)$ and substituting in the state equation results in

$$\frac{dx}{dt} = (A - BK(I + DK)^{-1}C)x(t) + B(I - K(I + DK^{-1}D)r(t) \tag{15-39}$$

$$y(t) = (I + DK)^{-1}Cx(t) + (I + DK)^{-1}Dr(t) \tag{15-40}$$

One fairly common simplification to this system is removing the feedforward D, which reduces the equations to

$$\frac{dx}{dt} = (A - BKC)x(t) + Br(t) \tag{15-41}$$

$$y(t) = Cx(t) \tag{15-42}$$

15.8.2 Nonlinear Systems

The more general form of a state-space model can be written as two functions:

$$\frac{dx}{dt} = f(t, x(t), u(t)) \tag{15-43}$$

$$y(t) = h(t, x(t), u(t)) \tag{15-44}$$

Equation 15-43 is the state equation, and Equation 15-44 is the output equation. If the function is a linear combination of states and inputs, then the equations can be written in matrix notation as shown here. The $u(t)$ argument to the functions can be dropped if the system has no inputs.

15.9 PROCESS SIMULATION

Process simulation is used for the design, development, analysis, and optimization of chemical processes and is first principles–based representation of chemical, physical, and other technical processes as well as unit operations in software. Prerequisites are a thorough knowledge of chemical and physical properties of pure components and mixtures, of reactions, and of mathematical models that describe the process.

Process simulation describes processes through flow diagrams where unit operations are positioned and connected by inlet and outlet streams. The simulation mathematics solves the mass and energy balance. Process simulation always uses models that introduce approximations and assumptions but allow the prediction of behavior over a wide range of temperatures and pressures that might not be covered by real data through interpolation and extrapolation within certain limits.

Simulation development focuses on development of models for a better representation of real processes. Model development includes improved predictions of physical properties and thermodynamics as well as improved mathematical simulation techniques for faster and more accurate results.

Two modeling approaches are commonly used:

- Rather simple equations and correlations where parameters are fitted to experimental data
- Predictive methods where properties are estimated

The equations and correlations are preferred because they normally describe the properties adequately. To obtain reliable parameters, one needs to have experimental data, which are usually obtained from data banks or, if no data are publicly available, from laboratory or field measurements.

First principles modeling of chemical properties began in the nineteenth century with development of cubic equation of states and the Antoine equation. The use of process simulation has accelerated with the evolution of computer science, computer hardware, and programming languages. Simple implementations of partial aspects of chemical processes began in the 1960s when, for the first time, suitable hardware and programming software were available. Whereas early simulations required much effort to set up, run, and debug, today's simulations are quite interactive with almost instantaneous results to promote debugging and running of a multitude of cases.

Initially, process simulation was used to simulate steady-state processes. Steady-state models perform a mass and energy balance of a process in an equilibrium state, but any changes over time had to be ignored.

Dynamic simulation is an extension of steady-state process simulation whereby time dependence is built into the models via derivative terms for accumulation of mass and energy. The advent of dynamic simulation means that the time-dependent description, prediction, and control of real processes in real time becomes possible. This includes the description of starting up and shutting down a plant, changes of conditions during a reaction, holdups, thermal changes, and more.

Dynamic simulation requires increased calculation time and is mathematically more complex than a steady-state simulation. It can be seen as a multiply repeated steady-state simulation based on a fixed time step with constantly changing parameters.

Dynamic simulation can be used in both an online and offline fashion. The online case being model predictive control, where the real-time simulation results are used to predict the changes that would occur for a control input change, and the control parameters are optimized based on the results. Offline process simulation can be used in the design, troubleshooting, and optimization of process plants as well as for case studies to assess the impacts of process modifications.

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CHAPTER 16

Gas Processing Plant Controls and Automation

16.1 INTRODUCTION

Automation has become an increasingly important aspect of gas processing. The amount of specific types of automations is a major decision now in the construction of new plants. With many existing plants having debottlenecked and improved process efficiency through mechanical means, automation is now a focus for further improvements. Automation provides the means for fully utilizing the mechanical capabilities of the equipment at all times and running the process at its most efficient points in a stable and reliable fashion. A good automation platform can be leveraged to provide the right information at the right time to the right personnel to make the right decisions in a timely manner. Historical data can be collected in virtually any time frame and analyzed statistically. With these historical data, upset situations can be reconstructed, and production reports can be automatically generated—just to name a couple of the numerous uses for these data. Many processors have upgraded their plants to higher-level control systems such as distributed control systems (DCS). Some of these processors ask, “Where are the benefits?” The correct questions to ask are, “How do I best leverage automation equipment to maximize benefits?” and “What are the best methods for measuring these benefits?”

This chapter discusses the elements of automating today’s gas processing plants, including considerations for instrumentation, controls, data collection, operator information, optimization, and management information. The advantages and disadvantages of various approaches are analyzed in this chapter. Also, strategies for identifying and quantifying the benefits of automation are discussed.

16.2 EARLY METHODS OF GAS PLANT AUTOMATION

The earliest gas processing plants were typically controlled manually by opening, pinching, and closing valves to meet their operating requirements. Pneumatic control systems were quickly adopted. These pneumatic control systems allowed the use of proportional, integral, and derivative (PID) controllers to send analog outputs to control valves to change their opening. As long as a sensor is available as a process variable for the controller, then a setpoint given by the operator can be targeted automatically. Discrete control could also be accomplished with pneumatic control systems by employing devices to fail with or without an air signal. In most cases, compressed air is used as the pneumatic conveyor, but natural gas is used in some remote operations and hydraulic oils are also employed. A reliable source of clean, oil-free air is quite important in the operation of pneumatic air systems.

As electronic controls were developed, these systems became the standard, although some elements of the pneumatic control systems are still in use. Most control valves today are still pneumatically operated for new and existing plants even though electronic control valves have been on the market for decades. Electronic controllers were accepted due to their lower price and greater reliability. These controllers included fewer moving parts to maintain. For new installations, wire was now run instead of pneumatic tubing with electronic to pneumatic (I/P for current to pressure) transducers added at the control valves. Process sensors such as temperature, pressure, level, and flow indicators were converted to electronic types as available and justified. Electronics also brought widespread use of safety systems, such as vibration sensor, burner management, and emergency shutdown systems. Other special controllers such as dedicated surge controllers for rotating equipment and triple modular redundant shutdown devices were developed.

16.3 MICROPROCESSOR-BASED AUTOMATION

16.3.1 Programmable Logic Controllers

Next came development of programmable logic controllers (PLCs). This type of controller was first intended for discrete factory floor applications such as those found in the auto manufacturing industry. These controllers employ programming of a microprocessor to mimic electronic relays. As PLCs gained wider acceptance, the functionality expanded to include PID control and other nondiscrete capabilities. PLCs are still quite effective for batch-type operations in the gas processing industry, such as solid bed gas dehydration and start-up as well as shutdown sequencing for rotating equipment. This automation platform is often used for smaller new facilities and smaller-scope retrofits of existing facilities. A human machine interface (HMI) is highly recommended to monitor the activities of the PLC.

16.3.2 Distributed Control Systems

The definition of a distributed control system (DCS) varies somewhat, but a rather simple definition is a control system method that is spread, or distributed, among several different unit processes on a common computer platform. These systems are typically hard-wired and exist within finite boundaries, although wireless means are becoming accepted, as data security issues are resolved. DCS offers the advantage of centralized control, while retaining the capability of local control. True distributed control systems use localized control, which is, in turn, controlled by the operator located at a central location. A DCS consists of the following:

- Remote control panel or device
- Communications medium
- Central control panel or facility
- Control, interface, and database software

A DCS may be as simple as one PLC remotely connected to a computer located in a field office. Larger systems may be PLC based, but will most likely consist of specially designed cabinets containing all of the equipment necessary to provide input-output (I/O) and communication. One point to consider during the design and specification of a control system is the level of autonomy each node will have in the event of a network or system failure. A true distributed system will allow most remote nodes to operate independently of the central control facility should the facility go offline or lose communication capability. Each remote node should be able to store the minimum process data required to operate in the event of such a failure. In this manner, costly and potentially disastrous process upsets can be avoided. If the system is performing both monitoring and control of a process or facility, it is referred to as a Supervisory Control and Data Acquisition (SCADA) system. Most systems will transfer data and commands using communication protocols such as Ethernet, or some other open standard, depending on the DCS vendor (Capano, 2002).

16.3.2.1 Remote Control Panel

The remote panel for a DCS or SCADA is typically referred to as a remote transmission unit (RTU). A typical RTU contains terminal blocks, input/output modules (both analog and digital), a computer or proprietary processor, and a communications interface. An RTU, depending on where it is located in the world, can perform both monitoring and control of a given process or processes.

16.3.2.2 Communications Medium

The communications medium is a cable or wireless link that serves to connect the RTU to the central control facility. There are several methods of doing this; typically, a cable, either a coaxial or twisted pair, is connected between the central control computer and the remote unit, or RTU. It is considered prudent to run two cables, on different routes, between the two in order to increase the reliability of the system. A network operates by taking data from the sending station, or node, and packaging and routing the information to the proper receiving station. The possibility of electrical noise, physical abuse, and software bugs should be considered when implementing a DCS.

16.3.2.3 Central Control

The control room is the center of activity and provides the means for effectively monitoring and controlling the process or facility. The control room contains the HMI, a computer that runs specialized software designed for that purpose. There may be multiple consoles, with varying degrees of access to data. In most cases, each operator or manager is given specific rights to allow more or less access and control of the system. The plant superintendent, for instance, may have complete control over his facility, while a technician may have access only to specific data on a particular process. This is done to avoid accidents and process upsets. This scheme also affords a degree of security, ensuring that only properly trained and authorized personnel can operate the various parts of the facility. The HMI presents the operator with a graphical version of the remote process. Depending on the skill of the operator and the level of sophistication of the interface, the process may be represented by anything from simple static graphics and displays to animation and voice alerts. Most packages afford the operator wide latitude on the design of the interface. The common thread to each system is the input-output (I/O) database.

The database contains all of the I/O defined for that DCS. This does not mean that all process data will be monitored and controlled; it means that only the data defined by the designers to be monitored and controlled will be available to the DCS. This database is a product of detailed evaluation of the process by the designer, who typically has the responsibility, with operator input, to design the most effective control schemes for a particular facility. The database is the reference the control software uses to correctly address each remote I/O point. Each database entry corresponds to an entity on the system, whether it is a physical point or an internal, or "soft," point such as an alarm, timer, or screen entity. A disadvantage of the early generations of PLC and DCS was the proprietary communications protocols employed. Some protocols such as MODBUS® were adopted as a de facto standard but had limitations as communications complexity increased.

16.3.3 Standards and Protocols

More recently, control systems have standardized on protocols made popular through the wide use of personal computers. Ethernet and Object Linking and Embedding (OLE) for Process Control (OPC) are two examples of highly accepted protocols. Ethernet is used mainly for device-to-device communications, whereas OPC is used primarily for application-to-application communications. Various versions of field buses have been developed to further standardize device-to-device communications. Early DCS was also based on legacy computing platforms such as VMS and the many different UNIX versions. Although Solaris UNIX is still used widely, Microsoft's Windows platform is gaining popularity (Poe and Harris, 2005).

16.4 CONTROL OF EQUIPMENT AND PROCESS SYSTEMS

16.4.1 Gas Gathering

Gas gathering systems are typically controlled through pressure control. Since most gathering systems employ primarily reciprocating compressors (employing a piston or screw), this discussion focuses on control of this type of compression. Rod loadings, maximum discharge temperature, and liquid entry, as well as minimum and maximum speeds, must be considered when operating these machines. It is quite important to prevent liquids from entering the compressor. An adequately sized upstream scrubber with mesh pad should be installed. A level controller to automatically dump liquids is required. The basic purpose of the gas gathering compressors is to keep the wellhead pressures down and keep the pressure to the downstream facilities—whether gas conditioning systems, liquids recovery facilities, or transmission pipelines—up to a minimum. The speed of the reciprocating machine is the main manipulation to control the suction pressure while the discharge pressure is dictated by flow rate and downstream resistance. Speed range is rather limited for these compressors but can be manipulated from about 90% to 100% of maximum speed. Other forms of capacity control include pockets and valve unloaders. These may be manual or automated. Rod loading, which is a function of the pressure differential across the compressor (discharge pressure less suction pressure), must not be exceeded, otherwise damage will result. Since there is no direct way to limit discharge pressure, a recycle line is installed to allow gas to recycle to the suction and raise the suction pressure. This line is often controlled by a minimum flow controller but is more effective from an efficiency perspective when controlled off a differential pressure or even a rod load calculation where the control platform can be configured accordingly. The recycle line should be routed after the discharge cooler and discharge scrubber to increase its effectiveness. The temperature rise across the compressor is dictated by the ratio of discharge pressure to suction pressure. High temperatures can warp piping and destroy packings. Discharge temperature should be monitored and the speed reduced or recycle rate increased to keep the temperature below the maximum allowable. The primary medium for provision of cooling the gas discharged from gas gathering systems is air. The following typically control these gas discharge coolers:

- Louvers
- On-off fans
- Multispeed fans
- Variable-pitch fans
- Variable-speed fans or gas recirculation

In most cases, lower temperatures of gas exiting the cooler are preferred for downstream processes. However, temperatures below the hydrate or freezing point are not desired. In these cases, a temperature controller manipulating any of the preceding coolers can maintain the desired temperature. Logic may be employed to determine when fans are turned on, off, or selected to run a discrete speed.

16.4.2 Gas Treating

The primary method of gas treating is with chemical absorbents. This process is analogous to gas dehydration with absorbents. Here are some differences:

1. The objective is acid gas removal, and, therefore, the contactor outlet gas should be analyzed for H_2S and/or CO_2.
2. Reflux temperature is set lower to maximize the retention of water.
3. In some cases, selective treating of H_2S versus CO_2 is desired, and the contactor outlet analyzer can drive a stripper overhead temperature setpoint, which, in turn, drives the ratio of heat medium used for the reboiler to absorbent flow.

Sometimes physical absorption utilizes a series of flash tanks, which yield better absorbent regeneration at lower temperatures. Vacuum is often pulled on the final stage of flash. In this case, the vacuum driver may need to be on speed control or the eductor on flow control to prevent implosion of the vessel. Hydraulic turbines that use the energy of depressuring to drive recirculation pumps are quite effective. A "helper" pump is always required to make up the horsepower deficiency.

16.4.3 Sulfur Recovery

The most common sulfur recovery process is the Claus process, in which one-third of the H_2S must be converted to SO_2 for proper stoichiometry. The theoretical sulfur recovery efficiency drops sharply when the stoichiometry differs from the desired 2 moles of H_2S to 1 mole of SO_2. A proper amount of oxygen, typically in the form of air, must be introduced to the reaction furnace. A tail gas analyzer is installed after the final reaction stage and before incineration. Two parallel valves, a main valve and a trim valve, are usually available on the discharge of the combustion air blower. Ideally, the main combustion air is manipulated on a feed-forward basis as acid gas feed rate and H_2S percentage vary. Feedback from the tail gas analyzer will control the trim valve. Since the time delay between the air introduction and tail gas analysis may be significant, a model predictive scheme may improve the ability to maintain proper stoichiometry.

Reaction furnace temperature must be maintained between a minimum and a maximum. Infrared measurements are quite effective in this service. Air preheat can be increased or decreased to raise or lower the reaction furnace temperature. Otherwise, air input requires adjustment. Conversion in the reactors is a trade-off between equilibrium, favored at lower temperatures, and kinetics, favored at higher temperatures. Equilibrium usually dictates when catalyst is fresh, whereas kinetics typically dictates when catalyst is near the end of life. Another factor when the temperatures are low is that the converter outlet temperature must be maintained above the sulfur dew point. Converter outlet temperature is controlled primarily by the reheat, which directly controls the converter inlet temperature. Reheat can be classified as direct or indirect. With direct reheat, hot gas bypasses the waste heat boiler at the outlet of the reaction furnace and is injected at the inlet of each converter. Lowering the converted inlet temperature when kinetics allow causes less gas to bypass and gives more shots at additional stages of conversion. Indirect reheat is with a heat medium such as steam or hot oil. Control of the reheat is straightforward with the indirect methods. Condensers are air cooled or water cooled and operated at their lowest temperature to achieve minimum dew point.

16.4.4 Gas Dehydration

The following sections cover two types of gas dehydration typically employed in gas processing operations. The first is absorption, typically with a glycol, and the second is fixed-bed absorption, typically with molecular sieve.

16.4.4.1 Absorption

Dehydration by absorption has several aspects that require control. They include the following:

- Lean absorbent flow rate and temperature
- Contactor pressure
- Flash tank pressure (where applicable)
- Stripper pressure
- Stripper reboiling
- Stripper reflux

Since the objective of gas dehydration is removal of water from the gas stream, the outlet gas stream should be monitored continuously with a moisture indicator. This indication should be monitored to adjust the flow of lean absorbent and the heat input to the stripper reboiler. The flow of absorbent should be based on a ratio of the gas flow that is corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is lower than required, since sensible heat will be saved in the stripper in addition to latent heat. The flow of heating medium to the stripper should be on ratio control and corrected based on the moisture indicator reading. This adjustment should take precedence when the moisture is higher than required, as the ability to dry is driven largely by the water content of the lean absorbent. If the lean absorbent is increased without increasing the ratio of heat to flow, then the lean absorbent's moisture content may not be reduced. Bottom temperature is not a good control basis as it will indicate only the boiling temperature of water at the pressure encountered at the bottom of the stripper. A better indication is a pressure-compensated top tray (or above the packing height) temperature. This has direct correlation with the water content of the lean absorbent. A low absorbent temperature improves its ability to hold water. However, too low a temperature may lead to condensation of hydrocarbons into the absorbent, causing foaming. An ideal strategy is to control the temperature of the absorbent about 5–10°F above the temperature of the inlet gas.

Feed-forward strategies that take into account the water content, flow rate, and temperature of the inlet gas to the absorber can also be employed. Model predictive strategies can account for the relative effects of lean absorbent temperature, flow, and stripper heat. A high pressure on the contactor is desirable to increase the contactor capacity and enhance the absorption of water. A back pressure controller should be employed to maintain a high pressure without causing the relief valve to function. The pressure should be maintained as low as possible when a flash tank exists between the contactor and stripper. A minimum pressure is required to "push" liquids into the contactor while decreasing the demand for stripper reboiler duty. One strategy will raise or lower the flash tank pressure when the liquid level exceeds a desired dead band around the setpoint. If the liquid rises above the dead band, the pressure is increased. Once the liquid falls within the dead band, the pressure can be slowly decreased. This strategy works best with nonlinear level control, which is always recommended for flash or feed tanks.

The stripper pressure should also be maintained as low as possible to lower the boiling point of the water stripped and allow the flash tank, if installed, to run at a lower pressure. Differential pressure measurements should be installed to indicate the onset of column flooding. When the differential pressure approaches setpoint, the stripper pressure should be raised to alleviate flooding. The reflux condenser should be set at a temperature to attain maximum recovery of entrained glycol without condensing excessive water unless required by environmental considerations.

16.4.4.2 Adsorbents

Since drying with fixed-bed adsorbents is a multibed process, the main control is cycle and bed switching. The dryer modes are drying, regenerating, cooling, and standby. Typically, each cycle is set for a fixed time, and the beds are cycled through the use of switching valves per this timer.

16.4.5 Liquids Recovery

16.4.5.1 Condensate Stabilization

Condensate is stabilized by stripping light hydrocarbon components in a fractionation tower. Nonlinear level control is recommended for the upstream flash or feed tank to provide a steady feed rate. Due to the nature of the condensate, online analysis is very difficult. Typically, a bottom temperature, preferably pressure compensated, is used to control the input of heat to the reboiler. A laboratory analysis is required to verify the adequacy of the bottom temperature setpoint. An inferential property predictor can be added to drive the temperature setpoint in between laboratory updates. The reflux temperature, when employed, should be controlled by a sensitive tray above the feed tray. Tower pressure should be driven as low as possible to enhance separation subject to constraints on an overhead compressor.

16.4.5.2 Refrigeration

Refrigeration is used to achieve the bulk condensation of natural gas liquids. Propane refrigerant is the primary medium used in gas processing. The main control aspects are compression, compression driver, refrigerant condenser, economizers, and chillers. Both centrifugal and reciprocating compressors are commonly employed in this service with turbine, electric motor, or gas engine drivers.

Lower temperatures are achieved at lower compressor suction pressures subject to surge conditions on centrifugal compressors and rod loading of reciprocating compressors. The suction pressure directly affects the pressure on the chillers.

Level control of the chillers is cascaded to the flow of refrigerant to the chiller. This control is critical to assure that the chiller tubes are covered without carryover of liquids. A scrubber or economizer before each compression stage is necessary to dump liquids whenever encountered. Economizers should be employed on multistage systems. The economizer pressure should be set to accommodate the compressor load and minimize kickback of high-stage vapor to lower stages. Refer to the sections in this chapter for compressor and driver control considerations.

16.4.5.3 Cryogenic Recovery (Turboexpander Processes)

Expansion with turboexpanders is now the main process employed for recovering natural gas liquids. Turboexpanders can be controlled for various objectives; inlet pressure, demethanizer pressure, and residue pressure are the most common. Guide vanes are manipulated to control the speed of the expander. A Joule–Thomson (JT) valve is always included to allow rapid unloading of the expander. One split-range controller typically operates the guide vanes and JT valve so that the JT valve will open when the manipulation of the guide vanes has been exhausted. The compressor driven by the turboexpander in either inlet compression or residue compression mode requires a recycle valve to maintain a minimum flow for surge protection. Depending on the exact cryogenic recovery processing scheme, additional controls may be required for heat exchanger circuit flow splits, chillers, separator levels, and pressure profiles. Heat exchanger flow splits are typically configured as flow ratios. This ratio may be overridden to prevent "cold spins" or prevent temperatures below the critical temperature in the cold separator upstream of the turboexpander.

16.4.5.4 Demethanizer

The demethanizer is integral to the turboexpansion process. The various feeds to the column are created at multiple points in the process. Side reboil heat sources and sometimes the bottom reboiler heat are integral to the heat exchanger circuit. Seldom are the side reboiler temperatures controlled. Manipulating the heat to the bottom reboiler controls a bottom temperature, preferably pressure compensated. An online chromatograph monitors the methane and/or carbon dioxide content.

Ideally, this output would reset the bottom temperature. Demethanizers are good candidates for model predictive control due to the disturbances caused by the side reboilers, inlet flow rates, and inlet compositions. Minimizing the pressure of the demethanizer based on turboexpander constraints and residue compression constraints is a major opportunity for increasing liquids recoveries.

16.4.6 NGL Fractionation

NGL fractionation consists of deethanization, depropanization, debutanization, and butane splitting (or deisobutanization). The control schemes for each are analogous. The major control points for these fractionators are reboiling heat, reflux, and pressure. Again, nonlinear level control is recommended for feed tanks and bottom surge levels. Reboiling heat is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature below the feed tray. Preferably, the temperature is pressure compensated.

Reflux flow is manipulated to control the bottom composition. The composition is cascaded to a sensitive temperature above the feed tray. Preferably, the temperature is pressure compensated. Minimum reflux schemes are employed to assure that reboiling load is not increased due to excessive reflux and, conversely, excessive reboiling, leading to greater reflux rates for a given separation. Internal reflux calculations and multivariable control schemes can achieve minimum energy consumption for a given separation.

Pressure should be minimized on these towers subject to constraints such as flooding, condenser temperature, and bottom hydraulics. Flooding is indicated by delta pressure measurements across the tower. Reflux is more difficult at lower temperatures as the available duty of the condenser may be limited. There must be sufficient head on the bottom of the tower to allow liquids to feed downstream towers or satisfy minimum head requirements for pumps. Again, multivariable control schemes handle the pressure minimization issue elegantly.

16.4.7 Centrifugal Compressors

Centrifugal compressors (utilizing an impeller to increase the kinetic energy of a vapor) are gaining wider acceptance in a variety of gas processing services, including feed, residue, and refrigeration compression. These compressors are typically driven by gas turbines or electric motors but can sometimes be steam turbines. There are several control considerations for this unit operation. Upstream liquids removal, surge prevention, suction pressure, discharge pressure, and driver speed control are the primary issues. Upstream liquid removal is accomplished with vessels upstream of the compressor that remove any entrained liquids and automatically dump liquids based on a liquid level. A simple on/off level control scheme is adequate for this purpose.

Surge is caused by excessive head requirements for a given suction pressure. The horsepower delivered by the compressor driver can be reduced to prevent surge, or gas can be recirculated from a higher stage of compression. Slowing the compressor driver is usually the most energy-efficient means. However, some drivers are limited in their speed range. There are many schemes that use antisurge or kickback valves to quickly increase the suction pressure by recirculating gas. The most sophisticated will take gas density and head curve characteristics into account for a wide variety of operating conditions. The simplest schemes assure a minimum flow or minimum suction pressure for the compressor. These simpler schemes yield horsepower inefficiencies due to a more conservative approach. Kickback schemes that use cooler gases and minimize the number of stages that gas is recycled are also the most energy efficient.

Suction and discharge pressures can be controlled by adjusting the driver speed, recycling gas, or throttling valves. Driver speed adjustments are the most energy efficient. Gas turbines will typically have a wide speed range. Steam turbines have a moderate speed range. Electric motors may be constant speed; however, variable speed and variable frequency drives are becoming more popular.

16.4.8 Centrifugal Pumps

Centrifugal pumps are analogous to centrifugal compressors but are seldom driven by a gas turbine. The same control considerations exist except, of course, the requirement to remove liquids upstream is replaced by a need to remove entrained vapors upstream. The net positive suction head dictates the surge point. Therefore, the level and density of the liquid at the suction of the pump are important.

16.4.9 Reciprocating Pumps

Reciprocating pumps are typically driven by electric motors, but some applications employ gas engines or steam turbines. These pumps are rather forgiving in their operation and require minimal control. The most common control required is capacity control, since these are positive displacement machines. Variable speed and liquid recycle based on minimum flow or upstream-level considerations are the main forms of capacity control.

16.4.10 Utilities

The most common utilities found in gas processing plants for process purposes are refrigeration systems, heating (hot oil or steam) systems, and cooling water systems. Refrigeration systems were covered previously. Hot oil systems employ heaters, mixing tanks, and headers. It is not uncommon for a plant to employ at least two levels of hot oil temperatures. The various temperatures are distributed through separate header systems. Some processes, such as amines and glycols, will degrade when high skin temperatures are encountered. A lower temperature heat medium minimizes reboiler skin temperatures. A common hot oil temperature scheme is to accumulate all heat medium returning from the process in a surge tank. A portion of the hot oil is routed through the heater to the temperature target for the high-temperature header. Enough of the liquid in the surge tank bypasses the heater and is mixed with the right amount of heated oil to achieve the low-temperature header target. Steam systems also typically employ several levels of temperature (or pressure). Boilers produce sufficient steam to satisfy a high-pressure header. Steam users, heat exchangers, and steam turbines discharge the exhausted steam into a lower pressure header to be reused or into the condensing system for collection and reuse. Steam turbines that exhaust steam into a lower pressure header for reuse at a lower temperature are called topping turbines. Otherwise, the steam turbines are total condensing turbines.

In order to satisfy the balance for the entire steam system for all the temperature (or pressure) levels required, let-down valves are employed to route higher pressure steam into lower pressure headers based on pressure control. The exhausted system is collected in the lowest pressure header, typically at atmospheric pressure; deaerated; replenished with makeup boiler feed water to replace losses; and then boiled again. Process temperature control is usually achieved by simply regulating the flow of heat medium to the heat exchanger.

16.5 AUTOMATION APPLICATIONS

Along with a central control room using electronic means to transmit data came the advent of applications that reside on the automation platform. These applications are focused on collecting information and using it to operate more cost effectively (Poe and Harris, 2005).

16.5.1 Data Historians

The ability to collect and store a large amount of data on a disk is a key advantage of microprocessor-based automation platforms. Even so, several vendors have specialized in developing historians to more efficiently and effectively store and analyze the data. Data compression techniques are used to store a maximum amount of information in the minimum space while maintaining resolution of the data. These historian packages come with tools to assist in mining the data, graphing, tabulating, and analyzing statistically. The historical database can be manipulated to automatically generate reports as well.

16.5.2 Asset and Performance Management

Asset and performance management software has been developed to tap into the wealth of information that is now available with microprocessor-based automation systems. These solutions include

- Computerized maintenance management
- Work order generation
- Predictive maintenance
- Control loop performance and tuning
- Online equipment health monitoring
- Process performance monitoring

Asset management focuses on maintaining plant equipment. Inventory management, work order generation, predictive maintenance programs, and turnaround planning can be accomplished with these tools. Many of these packages include hooks into enterprise planning systems. Control loop performance monitoring and tuning packages are available. These applications can determine whether a control loop is experiencing problems with a valve, positioner, or controller tuning, for example. Other applications trend the vibration, temperatures, and other key parameters for rotating equipment to determine when a failure is expected to occur.

Process models can be run online to determine how well a plant is performing compared to expected performance. Heat exchanger fouling and expander and compressor efficiencies, as well as tower efficiencies, can be calculated and monitored. These packages include data reconciliation features to overcome the problem of how to adjust process models for inconsistent, missing, or bad data.

The opportunities for application of the process performance audit initiative for increased plant profitability in the gas processing and NGL fractionation industries can be defined by analysis of the individual facilities in order to pinpoint the control loops that are the economic drivers of each facility. The following is a list of applications that require minimum control variability and tight adherence to hard spec limits in order to maximize economic performance (Kean, 2000):

Distillation Towers

Feed and reflux flow control loops
Reflux temperature control loop
Reboiler temperature control loop
Reboiler level control loop
Pressure control loop

Gas Compressors

Flow and pressure control loops
Surge control loops
Station recycle control loops
Gas temperature control loops

Acid Gas Treating Systems

Stripper reboiler temperature control loop
Stripper overhead temperature control loop
Stripper reflux flow control loop
Contactor and flash drum level control loops
Hot oil heater fuel and air flow control loops

Steam Boiler Systems

Steam drum level/feedwater flow control loops
Steam pressure and fuel flow control loops
Feedwater heater train control loops
Combustion air/O_2 control loops

Plant Utility Systems

Cooling water flow control loops
Fired heater fuel and air flow control loops
Refrigeration chiller level control loops

Control loop optimization through the employment of a formal process performance audit by skilled process consultants and control engineers can be an effective route to increased plant profitability. This economic improvement initiative could make the difference between a plant being economically viable or one that is considered for temporary shutdown or asset disposition (Kean, 2000).

As discussed previously, the basic control loop greatly affects plant performance due to the following facts:

- Process optimization requires optimization of the entire process, both hardware and software (Rinehart, 1997).
- The final control loop plays a significant role in process optimization (Rinehart, 1997).
- Control loop optimization reduces process variability and also increases process reliability.
- Optimization of the control loop is an essential step for successful application of advanced control.
- Large economic returns result from proper sizing, selection, and maintenance of the process control equipment.
- Continuous online monitoring of both loop equipment and loop performance is a key element for achieving the lowest cost of production, while minimizing the life-cycle cost of the processing facility.

16.5.3 Statistical Process Control

Data from the automation system can be interfaced to statistical process control packages. This software is used to generate run charts, process capability analyses, process characterization, experimental design, and cause-and-effect diagrams. This type of information is quite valuable to determine the causes of plant instability and off-specification products. It is also an excellent tool when baselining the plant performance and determining the benefits of improved control. Statistical process control concepts form a foundation for many of the Six Sigma and other quality initiatives when applied to continuous processes.

16.5.4 Advanced Regulatory Control

Advanced regulatory control was made much easier with the advent of microprocessor-based controllers. This control methodology basically turns single input, single output control into multiple input, single output control through the use of cascading controllers, selectors, feed forwards, ratios, etc. Shinskey (1996) discussed the variety of control strategies that can be employed in this way. Although not impossible with single loop pneumatic and electronic controllers, the software configuration approach with microprocessor-based controllers superseded the tubing runs, wiring, and other devices necessary to accomplish these strategies with earlier controller forms.

16.5.5 Multivariable Predictive Control

A more elegant and robust form of control is multivariable predictive control. This form of control has been used in the petroleum refining industry since the 1970s and provides true multiple input, multiple output control. Multivariable predictive process control provides a structured approach to managing process constraints, such as limits on valves and rates of change of temperature and pressures. A model for long-range prediction is used to ensure that the constraints on these variables are not violated. This enables the maintenance of an operating envelope within which the process is constrained. Recently introduced technology that enhances capability in this area includes constrained quadratic programming. In order to determine the optimal setpoints and constraint values for the controller, one can perform an outer optimization. This optimization can be described as a linear programming technique that is combined with a steady-state model and a cost function, determining the optimum operating point to be derived from a strategy based on minimum energy usage, maximum throughput, or a balance between these or other objectives.

Figure 16-1 shows a multivariable control strategy for a typical cryogenic demethanizer. A multivariable controller takes advantage of the interactive nature of the process. Key controlled variables are modeled dynamically as a function of key manipulated and disturbance variables. Flows depend on

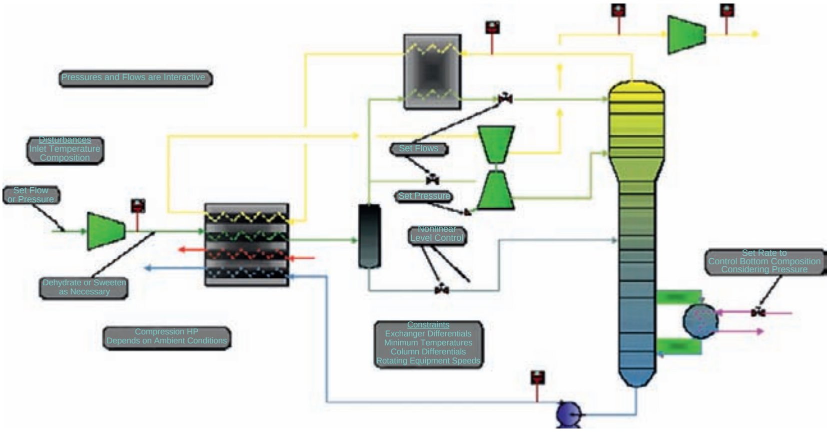

Figure 16-1 Multivariable control on a typical cryogenic demethanizer.

the pressure profile, the compression horsepower available, and the efficiency of the turboexpander. When more horsepower is available, such as at night and in cooler weather, then the flow can be increased or the pressure on the demethanizer can be lowered to increase NGL recovery for a given flow demand. Control of the NGL quality, typically for methane or carbon dioxide in ethane, becomes more difficult as tower pressure is adjusted. The multivariable controller can determine the correct heat input as tower pressure is adjusted to maintain maximum recovery and product quality. Pressure-compensated temperature is a key element of this strategy.

Other controlled variables can be minimum and maximum flows, pressures, temperatures, levels, speeds, etc. Demethanizer pressure, flow splits, reboiler flow, and plant inlet pressure are some manipulated variables, while disturbances such as residue pipeline pressure, inlet flow, and inlet composition are considered.

In the aforementioned manner, a "team" of key controllers pushes the plant to its optimum operating point at all times. Controllers are effectively decoupled, thereby making adjustments similar to an automobile on cruise control. In many regards, the objectives of statistical process control are achieved.

Figure 16-2 shows the beneficial effects of advanced controls. First, the variability of the process is reduced by the ability to control closer to setpoints with the model predictive capability. Once variability is reduced, then the process can be pushed closer to the operating constraints where maximum profitability is attained. This is an area where operators hesitate to operate due to the possibility of exceeding a constraint. With advanced process control, setpoints can be put closer to these constraints without fear of exceeding the constraints.

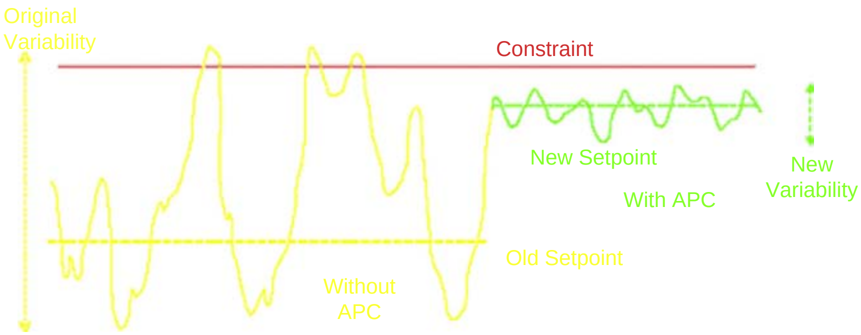

Figure 16-2 Impact of advanced controls.

16.5.6 Optimization

The next level of automation is optimization. For simpler optimization problems, linear programming (LP) techniques can be employed. Nonlinear techniques may be warranted for more complex optimization opportunities where linear methods would miss significant benefits. With the advances in computing power and optimization mathematics, online, rigorous optimization has become a reality. Again, petroleum refining is leading the way, but several gas processing applications have been developed. Optimization systems must rely on a multivariable predictive control system. As the robustness of the multivariable predictive control system improves, so does the effectiveness of the optimization system. With the business focus that optimization brings, multivariable predictive control developments should address the requirements of operations management. Table 16-1 details recent technology developments that have improved operations management and the benefits of multivariable predictive control applications. These advancements have enhanced basic process operations. This enables staff to be redeployed and presents more valuable operation functions that provide increased efficiency, reliability, reduced cost, improved quality, ideal staffing, and responsiveness to changing business requirements.

Networking and web capabilities continue to impact advanced applications. Some companies have central support centers or have contracted application service providers. Automated testing can be easily monitored remotely. Performance monitoring technologies and tools provide clear benchmarks of expected performance. Updates and improvements can be applied remotely. These capabilities reduce project and support costs and further improve the economics of multivariable predictive control initiatives. Mixed

Table 16-1 Recent Developments in Multivariable Predictive Control Technology (Canney, 2004)

Development	Technology
Nonlinear controllers	Neural nets and other empirical modeling techniques
Automated process testing	Multistep, random frequency testing
Performance monitoring	Metrics for model adequacy
Remote implementation and support	Web-enabled applications
Adaptivity	Background testing and model identification
Control system embedded	Powerful processors coupled with programming efficiencies

integer programming and enhanced state-space methods are now employed to improve models and decrease the impacts of unmeasured disturbances. Developments in sensor technology supply new, valuable, low-cost process information. Methods for estimating property and predicting the behavior of multiphase and complex reaction systems are also improving. Multivariate statistical methods continue to progress. Visualization technology may be a key to making the models transparent to the end user.

16.5.7 Leveraging Automation

In order to get the most benefit, the automation platform should be leveraged to its maximum capability. The platform itself is about 80% of the cost for about 20% of the potential benefits. A multitude of additional benefits are there for a fraction of the platform cost by adding applications. Include adequate instrumentation upfront for grassroot plants and during retrofits for existing plants. It is much more cost effective to add instrumentation during construction instead of adding it later. Use control system consultants instead of relying solely on equipment manufacturers. Many equipment manufacturers are penalized for including more instrumentation than their competitors. For control system upgrades, don't just replace in kind: replace and enhance. A DCS gives limited additional value when used simply as a panel board replacement (Poe and Harris, 2005).

16.5.7.1 Automation Upgrade Master Plans

Many gas processors have developed automation upgrade master plans for existing facilities where the relative benefits of each unit and the values of enhancements are estimated. In this way, the area with the most benefits can bear the majority of the initial cost of the platform. Additions to the platform, once the infrastructure is in place, typically cost less per I/O than the initial installation. Master plans are quite effective when the gas processor teams with an automation consultant. The consultant can share his past experience and advise of future developments that should be considered.

16.5.7.2 Determining the Benefits

The benefits of a fully leveraged automation system are numerous and include the following:

- Control room consolidation (reduced manpower)
- Plant reliability (uptime)
- Plant stability (better efficiency due to fewer process upsets)
- Maintenance management (reduced inventories)

- Product quality (fewer off-specification penalties or giveaway of over-processed product)
- Continuous constraint pushing (increased throughput and recoveries as well as energy savings)
- Optimization (energy savings and increased recoveries)

When approached appropriately, these benefits are easily quantified. The key is measuring through baselining and monitoring the results through proper metrics.

16.5.7.2.1 Baselining

A good baseline is a crucial element of determining benefits. Historical data are necessary to derive a good baseline. The historical data should include the primary measurement as well as any factors that should be used to normalize the result. For instance, ethane recovery should be normalized for inlet gas flow and composition as well as ambient temperature. Data during upset periods and outliers should be discarded. True process variability is measured best with frequent data capture on the order of every minute. High-quality baselining requires an excellent understanding of the operation in addition to statistical analysis. In addition to measuring mean performance of a process, baselining will reveal the amount of variability in the process and the source of the variability. Reducing variability is one of the main benefits of automation. With reduced variability, the process has freedom to shift to a more beneficial operating point.

16.5.7.2.2 Statistical Analysis

Statistical analysis is crucial in determining the benefits of automation. A good statistical analysis will give the most accurate assessment of process performance under changing conditions. Mean or average performance is always the final measure between before and after performance. Again, the mean should be evaluated on a normalized basis so that the performance is assessed on a fair or equivalent basis. Total liquids or even total ethane recovered is not a normalized evaluation. Barrels per MMcf of inlet is a better metric, but does not take into account changes in inlet composition. Recovery levels take into account inlet composition, but not process capability under variable inlet rates or ambient conditions that provide more mechanical horsepower availability. Standard deviation gives a measure of the amount of variability for a normal (bell-shaped) response curve. Depending on the process and the sources of variability, anywhere from one-half to two standard deviations of improvement is reasonably expected. One way to check this is to compare best performance or process capability (entitlement) to the mean performance. This difference can be compared to the standard deviation to give an idea of the maximum standard deviation improvement that is possible. Since standard deviation gives an absolute number and is difficult to compare to the mean, the coefficient of variation, which is the standard deviation divided by the mean, gives a relative measure of standard deviation. In other words, the coefficient of variation gives a percentage improvement potential. All responses are not normal or bell-shaped. Kurtosis is a measure of the skewness of the response curve and should be considered when evaluating standard deviation. A common example of an abnormal response is a process that runs close to its physical limits to the upside or downside and far from its physical limits to the opposite side. Elimination of upset conditions can sometimes shift a response curve close to normality. Multivariable control packages also include the capability to further identify the dynamic performance of a process with tools such as autocorrelation, power spectrum, and Fourier transforms. Determining financial benefits in the face of uncertainty is always a challenge. Even though we may have a good gauge on process improvement potential from the standpoint of increased product, the value of these improvements is dependent on economic conditions like pricing, feedstock availability, and product demand. Risk management tools, such as Monte Carlo simulations, can be used to establish the range and certainty within these ranges of financial benefits over the variety of conditions anticipated.

16.5.7.2.3 Performance Improvement Initiatives

Many companies are adopting Six Sigma and other statistically founded performance improvement initiatives. Automation can greatly enhance the effectiveness of Six Sigma and other performance improvement initiatives. Conversely, the rigor of these performance improvement methodologies is useful in documenting the benefits of automation. Consider Six Sigma as an example of how performance improvement methodologies and automation complement each other. The four phases of Six Sigma include measure, analyze, improve, and control (Breyfogle, 1999).

Measurement can sometimes be difficult and painful without automation. With automation and data historians, the measurement and analysis task becomes much easier. After the data are collected, the process capability assessment, run charts, cause-and-effect matrices, etc., can be developed to assess the performance improvement opportunity and where to focus the effort. The many elements of automation can be the enabling technology for the implementation and control phases of Six Sigma. Stabilizing the process is always a key element of performance improvement and is enabled by automation, especially advanced controls. The control phase pretty well speaks for itself.

16.6 CONDENSATE STABILIZER CASE STUDY

The following case study reviews an application of some of the advanced control features and the benefits. This case study is for an actual condensate stabilization process. The main reason for advanced process control on this unit was for quality purposes. Condensate product must meet a Reid vapor pressure (RVP) specification as defined by the customer. The previous method of production relied on laboratory sampling to verify the RVP, which was infrequent. Online analyzers are available but can be expensive. Instead of an online analyzer, the important quality measurement was derived through inferential means. The APC solution provided a stable plant and reduced the RVP variation.

BG Tunisia implemented multivariable predictive control on its gas condensate production system at the Hannibal plant in Sfax, Tunisia. The aims of the project, to maximize condensate yields, improve the stability of the condensate stabilization process, and ensure quality limits for the product, are adhered to at all times. BG Tunisia's Hannibal terminal processes gas at 5.4 MMscmd with the condensate column operating at a typical rate of 550 l/m controlling RVP to a maximum limit of 12 psia. Condensate can be sold as crude oil and is therefore more valuable than the alternative natural gas liquids products (Hotblack, 2004).

Figure 16-3 shows a detailed schematic for the condensate process. The Liquid hydrocarbon condensate is brought into the plant from the slug catcher and preheated in the stabilizer feed/bottoms exchanger E-201. The preheated feed is flashed in the stabilizer feed drum D-201. Liquid from the feed drum is further preheated through the feed/bottoms exchanger and fed to the stabilizer T-201 on local flow control. The feed drum level controller maintains level by manipulating the feed to the drum. The flashed vapor is sent to the high-pressure fuel gas system. A boot is provided on the feed drum to separate any entrained water, which is sent to the warm flare drum under level control. Condensed hydrocarbon liquid from the warm separator in the nitrogen rejection unit (NRU) is fed to the NRU liquids flash drum D-202 under level control—via the warm gas/liquid exchanger in the NRU. The flashed vapor is sent to the high-pressure fuel gas system, and liquid from the drum is fed to the stabilizer.

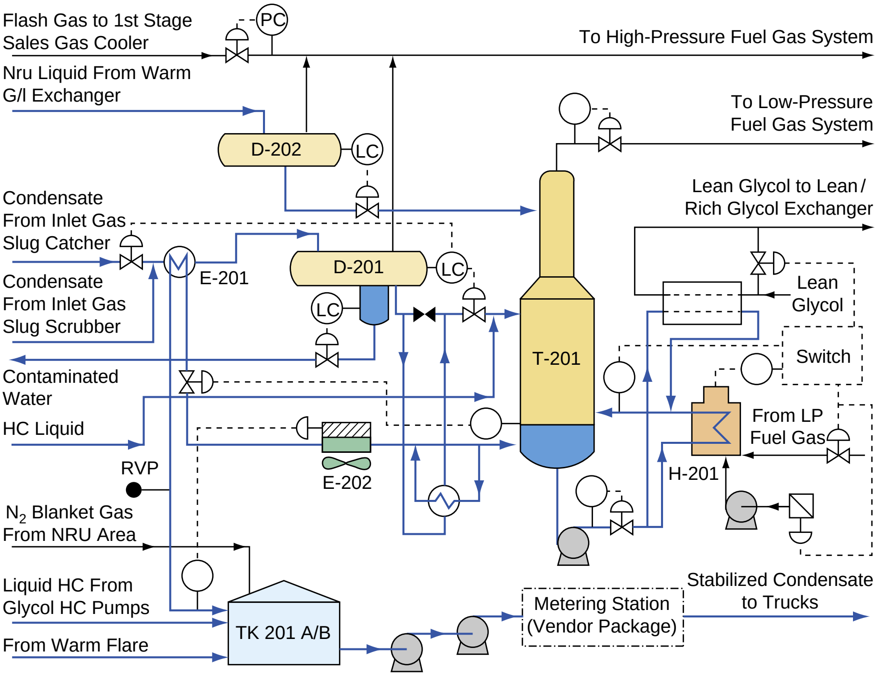

Figure 16-3 Schematic for BG Tunisia's Hannibal terminal condensate process (Hotblack, 2004).

The condensate stabilizer T-201 reduces the vapor pressure of the condensate by removing the lighter components. It is a stripper column with 24 trays. Liquid from the stabilizer feed drum is fed at the mid-point on tray 9, and liquid from the NRU liquids flash drum is fed to the top tray. Overhead vapor from the stabilizer is sent to the low-pressure fuel gas system through a back-pressure control valve that maintains the tower pressure to setpoint under the action of pressure controller.

The bottom part of the tower is divided into two sections by a baffle; the baffle does not extend to the very bottom of the tower, so there is some mixing between the two sections. Liquid from the bottom tray flows into the section that is preferentially pumped through the fired stabilizer reboiler H-201. The two-phase stream from the reboiler is returned to the other compartment, where the liquid is separated as stabilized product and the vapor flows up the tower to provide stripping action. The glycol/condensate exchanger provides additional heat duty if required.

Stabilized condensate leaving the stabilizer is cooled in the condensate cooler E-202 and by exchange with cold inlet feed condensate in the exchanger E-201, and is sent to the condensate storage tanks TK-201 A/B.

The main objectives of the condensate stabilizer connoisseur controller are

- Control RVP to an operator-specified target value
- Enforce any specified unit operating constraints at all times
- Stabilize the unit operation

The controller uses a real-time online estimate (inferential) of RVP. Through more tight control to the target value, excessively high RVP and the associated increased flashing of stabilized condensate in the storage tanks and during tanker transfer can be minimized, thus reducing losses. The product specification for RVP is 10–12 psia. Figure 16-4 shows the before-and-after trends of key process variable. The baseline assessment before APC was a mean RVP of 12.25 psia with one standard deviation of 0.43. After APC was installed, the standard deviation dropped to 0.26, enabling control to 11.5 psia with 95% confidence on the 12-psia limit (Hotblack, 2004).

Constrained optimization drives the process to an optimum without violating process constraints, such as the RVP limit imposed by the condensate purchaser.

A hybrid RVP sensor using both first principle and neural network technology provides a continuous measurement for control. Laboratory samples taken on a periodic basis are integrated into the control scheme, improving accuracy and ensuring product quality is maintained.

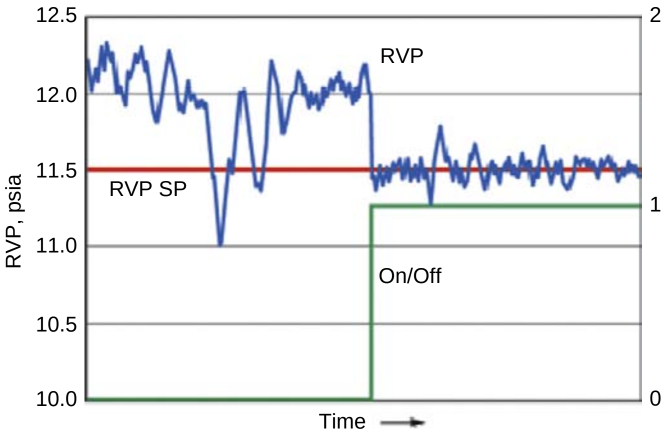

Figure 16-4 Reid vapor pressure (RVP) at BG Tunisia's plant, before-and-after trends of key process variable (Hotblack, 2004).

The inferential estimate is implemented directly within the controller as a hybrid of a first principles model based on the Antoine vapor pressure equation and a radial basis function neural network. Both the first principles model and the neural network require the stabilization column base temperature and top vapor pressure. Compensation for process drift and process measurement error is via feedback from the laboratory analysis. A weekly lab sample is inserted as a minimum for this purpose.

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CHAPTER 17

Dynamic Simulation of Gas Processing Plants

17.1 INTRODUCTION

Modeling has been used for a very long time for the design and for improved operation of gas processing and transmission facilities. The use of steady-state models is universally accepted in all stages of the design and operation of gas processing plants. Dynamic simulation has been used for a long time, but rigorous first principles of dynamic simulation have been confined to use by specialists and control engineers who were using models based on transfer functions that were incapable of representing the nonlinearities in systems and the discontinuities in start-up cases, for example. Only since the late 1990s has dynamic simulation become a more generally accepted tool used by process engineers and control engineers alike. The software available today enables process engineers with some process control knowledge and control engineers with some process knowledge to build dynamic models fairly easily. The constraint to using dynamic simulation is no longer that dynamic simulation is difficult to employ, but rather that the implementation time for a dynamic model is in the order of two to four times as long as the time needed to implement a steady-state model. Frequently, a consultant is employed to develop the model, and one or more engineers of an operating company or engineering company would use the model to run the needed studies.

This chapter discusses the areas of application of dynamic process modeling and modeling considerations, both general and for specific equipment frequently used in gas processing plants. The use of dynamic models in specific gas processing units is analyzed. Some case studies are presented to illustrate the use and impact of dynamic simulation on the design and operation of gas processing and transmission plants.

17.2 AREAS OF APPLICATION OF DYNAMIC SIMULATION

The areas of application have been divided in two large groups. The plant design group highlights applications that are most frequently used by engineering companies, whereas the plant operation group is mostly used by operating companies.

17.2.1 Plant Design

Dynamic models have several applications in plant design. Quite often, it is difficult to quantify the benefits associated with dynamic simulation. It is important to realize that a dynamic model can be reused for various applications in the design of a plant. The dynamic model will evolve as the design and the project evolve. The most detailed and rigorous model will be used close to plant commissioning and beyond (Brown and Hyde, 2001).

17.2.1.1 Controllability and Operability

Decisions taken very early in the design phase of a new plant or a revamp of a plant can have a significant impact on the controllability or operability of that plant. If the design calculations use only steady-state process simulation, the decision to employ a novel design is often a trade-off between the advantage the novel design brings and the potential controllability and operability issues. A dynamic model will shed more light on the problems that can be expected and lets control engineers devise adapted control strategies to mitigate or remove controllability problems. The use of dynamic simulation will increase the adoption of novel designs and ultimately the efficiency of new or revamped plants.

As these issues need to be analyzed in the early development of a process, the models will necessarily be simpler than the models used when the design is completed. At this stage, concepts are tested, and one would not expect quantitative answers from a dynamic model, but rather indications of process stability or control feasibility.

17.2.1.2 Safety Analysis

A lot of effort goes into ensuring a plant will be safe for the operators and the people living in the neighborhood of the plant. Huge liabilities are associated with the safety of a plant. Some dynamic models are standard practice in any design, but surprisingly, in other areas the use of dynamic models is still limited.

For virtually any plant operating at high pressure, depressuring studies will be a standard part of the engineering phase. Depressuring studies are performed to analyze the behavior of pressure and temperature when the pressure of a plant that has been stopped is let down. The depressuring study defines the necessary flaring capacity that is needed, and the results may also impact the choice of materials for vessels. Typically, if the pressure letdown of a vessel results in too low temperatures, carbon steel will have to be replaced with stainless steel to prevent the metal becoming too brittle. The other class of depressuring studies is related to the consequences of a fire in the area where a vessel is located. The main goal here is to determine the minimum vent rate that is needed to keep the vessel pressure under control and to bring down the vessel pressure to prevent a failure of the vessel wall as temperature rises.

All plants have emergency shutdown systems (ESDs). The design of a system to safely stop the operation of part of a plant or a complete plant can be very complex, and it is often difficult to foresee all the consequences of everything that might happen during a shutdown. A dynamic simulation model of the plant is an invaluable tool to properly set up an ESD system. Far too often, a dynamic model is used to analyze the exact causes of an incident that has taken place, but the use of such a model during the design phase might have prevented the occurrence of the incident altogether.

Modeling the behavior of the plant under emergency shutdown conditions requires the model to be quite detailed, and the simulation is usually more challenging than other applications. Under ESD conditions, much of the equipment is shut down, many flows are stopped, and any engineer knows that mathematics and zeros do not go together very well.

17.2.1.3 Start-Up Procedure Definition

The start-up procedure also requires a detailed model. Although generally the dry start condition also involves lots of stopped equipment and zero flows, it tends to be somewhat easier to model than an ESD scenario.

The use of a dynamic model to verify the start-up procedures of a plant can reduce the commissioning time by weeks. This exercise consists of adding the start-up logic to the model and running this start-up logic while observing the behavior of the plant model. When problems occur, the model can be stopped, and the start-up logic can be reviewed and rerun. Not only does the start-up procedure become streamlined, the engineers who have worked on it acquire a detailed understanding of the behavior of the plant, which allows them to make better decisions during plant commissioning and subsequent operation of the plant.

17.2.1.4 Distributed Control System Checkout

Distributed control system (DCS) checkout alone will not warrant the construction of a dynamic model of a plant. But if a model is available, the modifications needed to be able to run a DCS checkout are relatively small.

The purpose of the DCS checkout is to verify that all the cabling connecting the DCS to the plant and the DCS internal TAG allocations are hooked up properly. Obviously, a dynamic model will not be able to help with checking the physical cabling, but the signals from a dynamic model can replace the plant signals. This will help tremendously in verifying the logical connections inside the DCS. If a wrong measurement is routed to a particular controller, this will be seen quite readily, as the dynamic model provides realistic numbers for these. It is much easier to discern an erroneous number among realistic numbers than to match quasirandom numbers.

17.2.1.5 Operator Training

A classical use of dynamic simulation has been for operator training systems. Nowadays this is just one of the applications that can be built on top of a detailed dynamic model. In most new projects, an operator training system is becoming a standard requirement, and it is yet another driver to start using dynamic simulation early in plant design because part of the earlier work can be reused in an operator training system.

Besides the dynamic model, operator training systems are composed of various other parts:

- The operator stations that mimic the real DCS operator stations or that serve as spare operator consoles.
- An instructor station to allow the instructor to monitor the student's progress and to let him introduce a selection of failures or other problems the operator may encounter in the real plant.
- Possibly an automated system to assign a score to the performance of the operator and/or to let the operator run predefined training scenarios.
- Software and hardware for the communication between the various modules.

17.2.1.6 Advanced Process Control

Advanced process control (APC), particularly multivariable predictive control (MPC), normally requires access to plant data and step test results from the operating plant. Hence, MPC is usually implemented only once the plant has been commissioned. With a dynamic model available, this is no longer a limitation, and the necessary information can be obtained from the step tests on that model. This part will be covered in more detail in Section 17.2.2.6.

17.2.2 Plant Operation

A dynamic model in plant operation will usually require justification on a single application. Although future uses of a dynamic model are certainly possible, these will usually not be considered to be a tangible reason or additional justification to create such a model. However, it is quite often easier to quantify the benefits that will be gained from the use of a dynamic model.

17.2.2.1 Troubleshooting

Issues in the control or operability of the plant can be resolved more easily, safely, and with no loss of production using a dynamic model. A dynamic model can be exercised at will where the engineer has very little freedom to test things out on the real plant. Maintaining production on spec will almost always override the need for testing to solve a problem. The solution imagined by the plant engineer is not necessarily the right one, and implementing an untested solution may lead to unsafe operation.

With a dynamic model, the worst that can happen is that the model fails. The engineer can test a large range of operating conditions to ensure that the implemented solution will hold up in abnormal conditions, for example.

17.2.2.2 Plant Performance Enhancement

Most plant engineers will accept as a given that many of the controllers in a plant are operated manually. This is frequently a source of added operating costs for the plant. A typical example would be the reflux rate of a distillation column that is set to a fixed relatively high value to ensure that the product always meets specifications. Most of the time, say 85% of the time, the reflux could be run significantly lower and only during less than 5% of the time the fixed reflux is needed to maintain product quality. It is fairly easy to compute the financial benefit associated with keeping the controller set to automatic operation.

The reason that controllers are put in manual operation is often related to the trust operators have in controllers. There are generally two possible reasons for distrust. Either the operator doesn't understand properly how the controller will cope with upsets, or in the past the controller has shown that it is incapable of dealing with those upsets.

Even without a full operator training system, a dynamic model can be used to show the operators how a series of typical upsets will be handled by the control system. Some of the responses may appear illogical to the operators at first, and with a model, the operators will not feel the pressure to act prematurely in order to ensure product quality. This can help to instill more confidence in the automated control and let it run in automatic mode.

Of course, the distrust of the operators may be well founded, but in this case, the dynamic model can be used as described in Section 17.2.2.1 to improve the controller behavior and subsequently illustrate to the operators that the problem has been solved to restore trust.

17.2.2.3 Incident Analysis

Although an incident analysis is by no means the best use of a dynamic model, it is all too often the first step toward using dynamic simulation. After an incident, there is always the need to know why it happened. If the incident resulted in damage, there will be legal requirements to determine the root cause of the incident. A dynamic process simulation model will often be used in this analysis to determine the sequence of events on the process side that led up to the incident and how adequate the emergency shutdown procedures were to mitigate the consequences of the incident.

17.2.2.4 Operator Decision Support

Operator decision support is an emerging use of dynamic simulation models. In this type of application, the dynamic simulation model is run in real time, and it receives the same input signals as the real plant. It is impossible to cover the entire plant with instrumentation to provide all the information one would like to obtain. The real-time model provides the operators and engineers with simulated measurements throughout the entire plant to better appreciate current operation. Typical parts of a plant that do not have all the instrumentation one would like are long pipelines and high-temperature outlets of reactors or furnaces. A second use of the online model is its predictive capability. Assuming the dynamic simulation model is fast enough, it can be used to predict events occurring minutes or even hours ahead of the actual events. This information can be used to improve the handling of the events and to keep the plant operating within specifications.

17.2.2.5 Operator Training

It is important for the operators to keep their knowledge of the plant operation up to date. Especially with highly automated plants, it is important that operators are confronted with abnormal situations using the simulator. New operators will also benefit from the use of an operator training system.

It is therefore very important to keep the operator training system that was installed as part of the plant commissioning up to date. This means that any changes to the DCS screens and systems must also be made to the operator training system and that any changes made to the plant must be made in the dynamic process model as well.

17.2.2.6 Advanced Process Control

The implementation of an advanced process control (APC) system requires a significant investment, and such a decision is not taken lightly. A dynamic simulation model can assist in determining the relevance of an APC implementation and can help to streamline the implementation itself.

It is fairly straightforward to run the necessary step tests for the implementation of an MPC on a dynamic process simulation. The results can be used to design the MPC and to run the MPC on the dynamic model. A comparison of the plant performance using the existing control system and with the MPC can provide the necessary information to decide whether an MPC implementation is an attractive investment. Running step tests on a model has a number of advantages over step tests on the real plant:

- There is no disturbance of the plant operation.
- Step tests can use a broader range of conditions.
- Step tests do not depend on the availability of plant personnel, plant incidents, and other events that are not compatible with step tests on the plant.
- The dynamic model doesn't suffer from valves that get stuck and other incidents that make life difficult when performing step tests.
- The dynamic model can be run faster than real time, and, hence, step tests that would otherwise take days can be run in an hour or less.

When the final MPC controller has been designed and implemented, it can first be put online using the dynamic simulation model. This setup can be used to discover a significant part of the practical problems that would otherwise surface only during the commissioning of the MPC on the real plant. Although the MPC models should be verified for actual plant operation, this accelerates MPC commissioning and lowers the risk of production loss that may be experienced during commissioning of the MPC.

17.3 MODELING CONSIDERATIONS

17.3.1 Level of Detail in the Model

The level of detail required for a dynamic simulation model is very dependent on the application. Most of the time, a model will contain components that are modeled in great detail (high fidelity), whereas other components capture only the overall dynamic behavior. A typical example is a model of a gas compressor.

An initial application is to use the model to analyze the behavior of the antisurge control logic. In this case, it will be important to properly model all the gas volumes in the main gas lines and also in the antisurge system.

The control logic used in the model will be an exact replication of the commercial system that will be installed. Because compressor surge is a very fast phenomenon, these controllers have sampling times on the order of 50 milliseconds; the model will need to run using a time step that is capable of capturing these phenomena and therefore have a time step that is even smaller than 50 milliseconds. As a consequence, the model may run slower than real time. But because the time span of interest in this case is a few minutes at the most, the slower model performance is not really a problem.

A second application for this dynamic model of a compressor station may be for operator training. It is not relevant to model events that happen so fast that it is impossible for operators to respond during the event. On the one hand, the step size and hence the speed of the model can be increased. On the other hand, it will be important to include manually operated purge valves in the dynamic model to allow the operators to perform all actions they deem necessary. This is a bit of detail that is of no use for the initial application.

The level of detail required should be assessed based on the objectives for the model. This assessment is not global for the complete model, but the assessment should consider the objectives for each section of the plant down to each piece of equipment.

17.3.2 Model Speed

The speed of a model is most frequently expressed as the real-time factor of the model. This is the ratio of the simulated time divided by the real time. The speed requirements vary depending on the application. For an operator training simulator, it is clear that the model needs to be capable of runoff running at least in real time (real-time factor = 1). Quite often, the real-time factor should be higher, up to 10 times real time. This allows the operator to fast forward through periods of stabilization of the process, for example.

For an engineering study, the important factor is the total amount of time it takes to study an event. Ideally, that time would be 10 minutes or less. This means that 3-hour events should have a real-time factor of 18 or higher. If the event to be studied lasts for only one minute, then a real-time factor of 0.1 is acceptable.

The model speed is mainly affected by the following factors:

- Time step of the integrator
- Complexity of the model
- Number of components used to represent the fluids and the complexity of the thermodynamic model

Because the same factors will also affect the accuracy of the model, a balance must be found between speed and accuracy.

17.3.3 Equipment-Specific Considerations

The following sections recommend the information to consider when modeling various pieces of equipment. Depending on the available modeling tool, the recommended level of detail may differ. Recommendations apply only to the main aspects of the model.

17.3.3.1 Valves

The minimum requirement to model a valve properly is to use the correct *Cv* value[1] and the type of valve characteristic. For some studies, it is important to capture the dynamic behavior of the valve. For example, an emergency shutdown valve needs a certain time to close. For safety studies, it is important to consider the time to close. Most plants will have one or more check valves. It is important to include these valves in the model, particularly when the model will be used to run scenarios far away from normal operation.

17.3.3.2 Rotating Equipment

For pumps, compressors, and expanders, it is best to always use the performance curves of the equipment. If this information is not available, it is relatively simple to create a generic performance curve starting from the normal operating point of the equipment. This information is then complemented with either the speed or with the absorbed power. For most motor-driven equipment, a speed specification will be the best option, except when studying start-up and shutdown phenomena. For equipment driven by a gas turbine, a specification of the absorbed power is usually a better choice.

If the study concerns the start-up or shutdown of the equipment, it will be necessary to include details such as the rotor inertia, the friction losses, and the dynamics of the driver (e.g., an electrical motor) in the model.

17.3.3.3 Piping Equipment

The level of detail for modeling the piping depends a lot on the application. For process piping, it is quite often sufficient to model the pressure loss. Most of the time, the volume of the piping is negligible compared to the volume of the process equipment. A notable exception is the modeling of compressor antisurge systems. An accurate representation of the system

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[1] *Cv* is the valve coefficient that is defined by convention in field or imperial units as the number of US gallons that will pass through a control valve in 1 minute, when the pressure differential across the valve is 1 psi.

volume is crucial to obtain correct results. For transport pipelines, the model should usually be more detailed because the expected results may include information such as the time lag of a product in the pipeline, the evolution of the temperature, and the multiphase behavior of the fluid. The required information includes the pipeline elevation profile, the pipeline diameter, pipe schedule, insulation, and environment.

17.3.3.4 Columns

Distillation models should reflect properly the holdup volumes of both the liquid and the vapor phase. A significant difference between tray columns and packed columns is that the much smaller liquid holdup of packed columns will reduce the response time of the column compositional perturbations, for example.

It is customary to use a reduced number of theoretical stages to model distillation columns in steady-state simulations. If the same approach is used in a dynamic simulation, the tray or packing characteristics will be modified to use the correct holdup volumes for liquid and vapor. Another approach is to use tray efficiencies and keep the number of trays used equal to the number of real trays in the column. One must keep in mind that the tray efficiency and the column overall efficiency are not the same.

17.3.3.5 Heat Exchangers

The level of detail for a heat exchanger will strongly depend on the role of the heat exchanger and the phenomena to be studied. For example, if the exchanger serves to cool down a condensate stream before it proceeds to storage and the focus of the study is on the equipment upstream of this exchanger, then it may be enough to use a model that simply assumes the exchanger is always cooling the fluid down to the required temperature. At the opposite side of the spectrum, there would be a plate fin heat exchanger in a natural gas liquefaction plant. In this case, the exchanger is the heart of the plant, and the model needs to accurately represent the construction of the exchanger and take into account elements like the heat capacity of the metal and the dynamics of the metal temperature. The model will need to provide information such as the temperature and pressure profiles inside the exchanger.

17.3.3.6 Control Systems

Contrary to steady-state simulation, the modeling of the control equipment is crucial to the success of a dynamic simulation model. Quite often the control strategy and controller tuning are the final objective of the dynamic simulation, but without proper configuration of the control system, the model will quickly end up in totally abnormal operating conditions.

For regular PID controllers, the main points to consider are correct direction of the action (reverse or direct) taken, realistic tuning constants, and proper span of the instrumentation. Once the simulation model has reached relatively stable conditions, attention can focus on a high-fidelity representation of the control systems.

The high-fidelity representation can come in various forms. For an operator training system, most of the DCS vendors will be able to provide software that emulates the DCS system. The model itself is then used only to represent the noncontrol equipment. The DCS emulation will receive the plant measurements from the model as the real DCS would receive the measurements from the plant, and the DCS will send signals to the valve positioners according to the control algorithms defined in the DCS.

The verification of a surge controller for a compressor is also an area where a high-fidelity representation of the particular controller is crucial. The representation can be built by combining blocks that are part of the dynamic simulator, writing a custom model for the controller, linking the model to an emulation program, or even linking the model to the controller hardware.

17.4 CONTROL OF EQUIPMENT AND PROCESS SYSTEMS

The following sections enumerate some typical applications of dynamic simulation in the various processes that are employed in gas treatment and transportation. The application is usually governed by the particular equipment used in the process.

17.4.1 Gas Gathering and Transportation

The key equipment in gas gathering and transportation are pipelines, valves, and compressors. The range of applications in this area is very wide:

- Risk assessment of condensate accumulation and associated slug sizes
- Line packing capacity studies
- Safety studies on pipeline shutdowns
- Pipeline depressuring studies
- Compressor station antisurge control studies

17.4.2 Gas Treating

The main equipment used for the absorption of CO_2 and H_2S are absorption and regeneration columns using amine solutions to absorb the acid gas components. A dynamic model will prove useful if the quality of the feed gas can fluctuate significantly. In such a case, the product gas quality will be affected directly by the amount of amine solution that is circulated and also by the quality of the lean amine solution that is indirectly influenced by the amount of acid gas in the feed.

17.4.3 Sulfur Recovery

The performance of a sulfur recovery unit is mainly governed by the operation of the various reactors in the process. A key factor in the reactor performance is the correct air-gas ratio of the reactor feed. If the acid gas feed is unstable, dynamic modeling can be used to select the best control strategy to cope with these fluctuations and to improve the controller tuning (Young et al., 2001).

17.4.4 Gas Dehydration

The glycol gas dehydration process is very similar to the acid gas removal process. Fluctuations in water content of the feed gas and gas flow rate affect the product gas quality. The control strategy ensures quality of the product by selecting the appropriate glycol flow rate and by maintaining the quality of the lean glycol.

For dehydration processes using molecular sieve beds, the application of dynamic simulation is similar to an application for plant start-up. The model would include the logic that drives the bed switching and regeneration cycles. Once implemented in the model, the logic can be tested by running it on the model and by tracking critical operating parameters of the mole sieve unit.

17.4.5 Liquids Recovery, Natural Gas Liquefaction

A cold box is a key piece of equipment in these processes. The flows that exchange heat in the cold box exchanger create a multitude of thermal loops in the process, which make it more difficult to control. A detailed dynamic model of the overall process including a detailed model of the cold box will help to understand the severity of the interactions created by these thermal loops. A control strategy designed to cope with these interactions can be tested thoroughly (Valappil et al., 2005).

From the perspective of pressures and flows, the operation of turboexpanders or turboexpanders coupled to compressors is important. The efficiency of a turboexpander drops quickly as one deviates from the design conditions, and the impact of a temporary deviation of the operating conditions on the process dynamics is difficult to understand without help. A dynamic model will aid in understanding the behavior and in selecting the correct control structure and controller tuning to cope with a transient deviation from design conditions.

17.4.6 NGL Fractionation

NGL fractionation is composed of a series of distillation columns. Because the purity specs on the columns are fairly severe on both the top and bottom products, the control of the columns is not straightforward. A dynamic model will provide the capability to select the best control strategy given the particular column operation and specifications and given the expected disturbances in the feeds.

17.5 CASE STUDY I: ANALYSIS OF A FUEL GAS SYSTEM START-UP

This study analyzes the start-up philosophy of the fuel gas system of a Latin American offshore platform. The fuel gas system flow diagram is shown in Figure 17-1. From steady-state analysis, it is thought that the system can be started up using cold pipeline gas without preheating (Wassenhove, 2003).

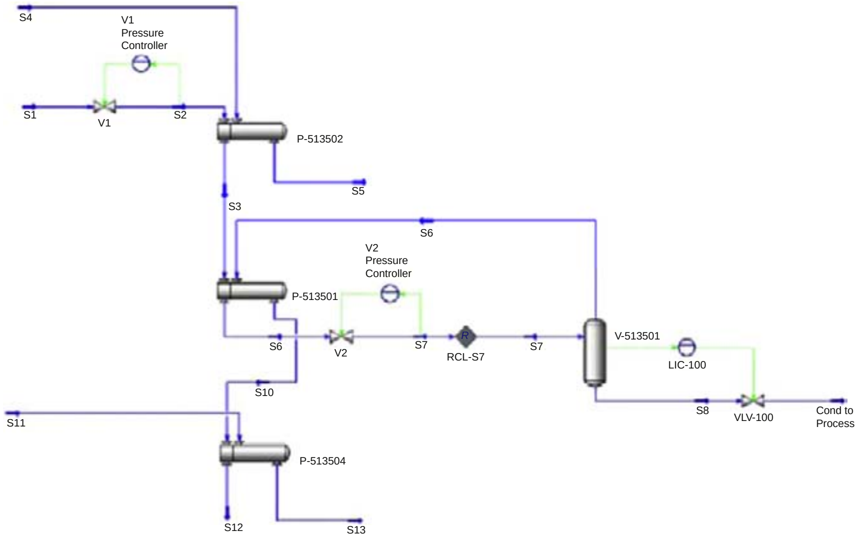


Figure 17-1 Fuel gas system process flow diagram.

17.5.1 Introduction

Under normal operation, the fuel gas burned in the turbogenerators (TGs) comes from the platform main compression system at 180 kg/cm^2 and 38°C. At the system inlet, the pressure is reduced to 100 kg/cm^2, and the temperature goes down to 17°C. The fuel gas is then preheated in P-513502 to 60°C with hot water and another pressure reduction to 45 kg/cm^2 through a Joule–Thomson (JT) valve resulting in a temperature of around 16°C. After this last pressure reduction, the mixture of condensate and gas is sent to a condensate vessel drum (V-513501), the condensate returns to the process, and the gas from the vessel proceeds through a gas-gas heater exchanger (P-513501). This fuel gas from P-513501 at an average temperature of 43°C is reheated with hot water in heat exchanger P-513504 to around 63°C and sent to the turbogenerators.

The objective of the system is to produce a fuel gas stream with a defined rate, a defined pressure, and a temperature at least 20°C above the dew point. This is a minimum value demanded by the turbogenerator vendor. Fuel gas exit temperature must also be maintained above 0°C to meet material temperature limitations.

During the start-up/restart of the platform, there is no gas source on the platform, but cold pipeline gas can be imported at 5°C. Also, there is no hot water available for the preheaters. One solution is to start the turbogenerators with diesel. This solution is undesirable since diesel needs to be imported onto the platform, the turbogenerators need to be adapted to cope with diesel feed, and there are several other problems leading to production inefficiency and unnecessary costs.

The objective of the dynamic simulation is to study the exact behavior of the fuel gas system under the start-up condition of cold pipeline import gas and the circulation of the seawater at 25°C to heat the gas.

17.5.2 Steady-State Analysis

The steady-state results clearly show that thermodynamically and theoretically both normal operating and cold start scenarios result in gas feeds to the turbogenerators that are well above the dew point approach limitation of 20°C. Hence, it seems feasible to provide fuel gas with a temperature at 20°C above dew point. However, this does not account for the transients encountered during start-up. The question can only be answered conclusively by a dynamic analysis.

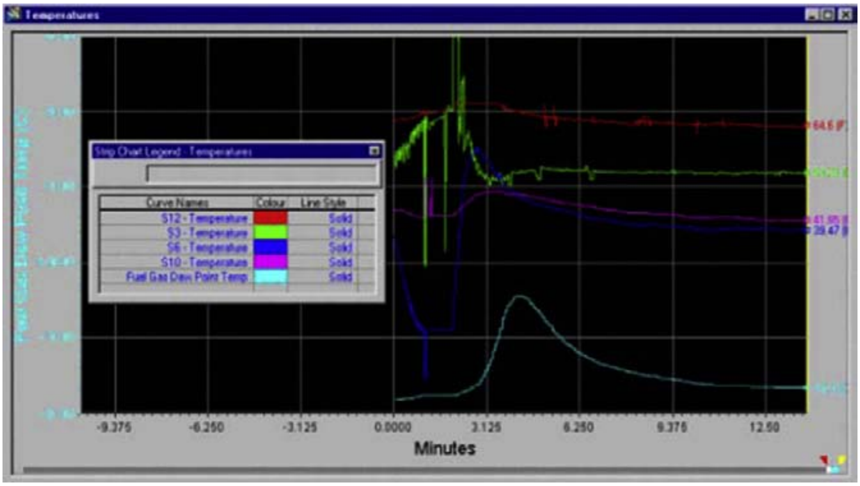
17.5.3 Dynamic Analysis

In the first start-up scenario, water is circulating at 25°C before any gas is fed. Once the water flow is stable, cold pipeline gas is brought on stream. The strip charts from a HYSYS Dynamics model displaying the system temperatures for this case are shown in Figure 17-2.

As the cold import gas hits the warmer exchangers, the gas is heated, and some of the heavier components flash off in the condensate drum and pass back through the gas-gas chiller. The dew point of the fuel gas then rises for about 4 minutes until eventually the chill duty in the recycle and the JT effect in valve V2 stabilize. The temperature of the fuel gas outlet remains relatively constant as the heat exchange in P-513504 is established. Hence, there is a closer approach between the fuel gas temperature and the dew point, reaching 24°C at 4.1 minutes. This is too close to the limit to accept without further study.

The second start-up scenario assumes that the cold pipeline gas flow is established first and that the water system is brought online afterward. The strip charts of the system temperatures in this case are shown in Figure 17-3.

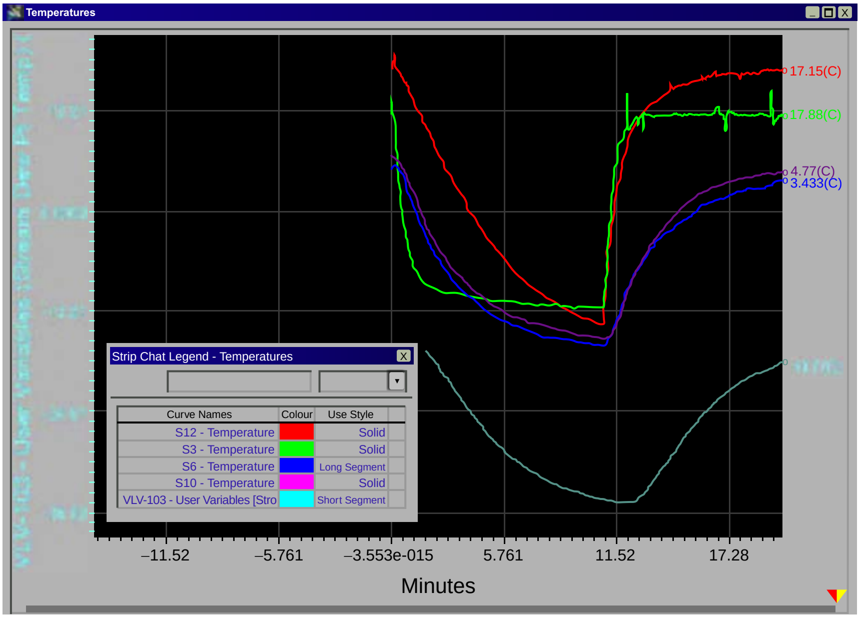
Because of the JT effect and cold duty on the gas-to-gas exchanger, the gas dew point does not initially increase but maintains a difference from stream 12 temperature. In this mode, they would eventually equate. Also, the fuel gas exit temperature is decreasing, and within 4 minutes, it will reach the material temperature limit of 0°C. On start-up of the water system, the



Strip Chart Legend - Temperatures

Curve Names	Colour	Line Style
S12 - Temperature	🟥	Solid
S3 - Temperature	🟩	Solid
S6 - Temperature	🟦	Solid
S10 - Temperature	🟪	Solid
Fuel Gas Dew Point Temp	🟦	Solid

Figure 17-2 Cold water circulating, import pipeline gas system temperatures.



Curve Names	Colour	Use Style
S12 - Temperature	🟥	Solid
S3 - Temperature	🟩	Solid
S6 - Temperature	🟦	Long Segment
S10 - Temperature	🟪	Solid
VLV-103 - User Variables [Stro	🟦	Short Segment

Figure 17-3 Import pipeline gas on, bring up water system temperatures.

temperature recovers after 10 minutes, and the fuel gas always maintains more than a 40°C difference from the dew point.

17.5.4 Conclusion

A dynamic model clearly shows potential problems with all start-up modes of the platform fuel gas system with minimum approaches to dew point or minimum material temperatures. However, the dynamic model also demonstrates a combination of procedures, starting with the water off and then increasing flow quickly, that could maintain all proper flow rates and temperatures for fuel gas start-up without diesel, thus saving millions of dollars.

17.6 CASE STUDY II: ONLINE DYNAMIC MODEL OF A TRUNK PIPELINE

The 235-mile NOGAT trunkline is located in the Dutch part of the North Sea, and it connects eight offshore platforms to onshore gas processing facilities near Den Helder, the Netherlands. Each platform delivers both gas and condensate to the pipeline, so the line operates inside the two-phase region; Two of the platforms sit on oil fields, and the off-gas from the oil stabilization units is compressed and delivered to the pipeline. The current total capacity of the gas transportation system is about 22 million m^3/day, associated with 750 m^3 of condensate/day (La Riviére and Rodriguez, 2005).

The onshore facilities include a 1,000 m^3 slug catcher, condensate stabilization units that remove volatile components from the trunkline produced liquids, and a series of low temperature separation (LTS) units to dry the sales gas prior to delivery to the distribution network. Figure 17-4 shows a representation of the system.

There are two major challenges to operating the system:

1. Controlling the sales gas quality in terms of its Wobbe index.[2] The different platforms produce different gas qualities and quantities that are fed into the line at different locations. Therefore, the quality of the gas that travels through the trunkline varies along its length and also with

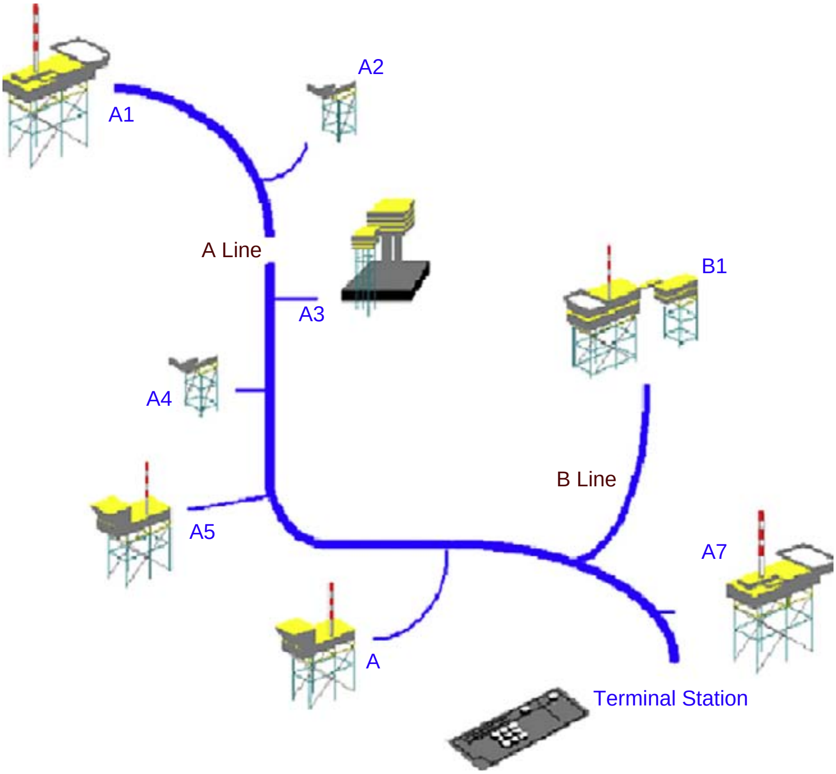


Figure 17-4 NOGAT trunkline system (La Riviére and Rodriguez, 2005).

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[2] The Wobbe index is a measure used to compare the equivalent thermal content characteristics of different gases. It is defined as volumetric high heating value divided by the square root of relative density to air.

time. In spite of this, the sales gas quality must stay within the contractual limits at all times (i.e., Wobbe index values between 49 and 54).

2. Controlling the condensate inventory of the trunkline. The amount of condensate retained inside the line builds up during periods of low gas demand, particularly in trunkline depressions. The available slug catchers and condensate stabilization unit capacities limit the production ramp-up speed and force the scheduling of periodic cleanup cycles to keep the trunkline liquids holdup below certain critical limits.

Neither the Wobbe index nor the condensate holdup profiles along the length of the trunkline are directly measurable through field instrumentation.

A system has been constructed around an executive application that uses dynamic models as the main simulation engine and allows data historians to access the current operating conditions. The executive application handles the necessary data transfer between the historical databases and simulation models. Two dynamic models are integrated in the application. These are the so-called online model to monitor the present-time conditions in the trunkline and the predictive model to perform what-if analyses and determine the consequences of programmed or unexpected variations in operating conditions (typically, gas supply flow rates).

Gas and condensate sources are modeled by means of standard process streams. Gas flow rates are automatically retrieved from the data historians. Condensate flow rates are calculated using average condensate/gas ratio (CGR) values for each platform. Gas and condensate compositions are set as fixed parameters. The multiphase pipelines are modeled using a proprietary model that operates as a unit operation inside the dynamic simulator. The trunkline inside a temperature profile was assumed to be equal to the ambient temperature profile (simulation runs revealed the adequacy of such an assumption). Ambient temperature profiles were obtained from a North Sea temperature map that takes into account depth and date.

The physical properties of the gas condensate mixtures are calculated using proprietary correlations and databases. Although the available chromatographic analysis provided information for up to 49 chemical species, the components slate was simplified by lumping the C_6 and heavier components into a smaller slate to achieve the desired model calculation speed.

Even on today's powerful desktop computers, calculation speed is still a limiting factor for dynamic two-phase flow combined with rigorous physical property calculations. It is therefore necessary to choose a balanced solution between accuracy and speed of calculation by optimizing the number of trunkline cells, time step size, and number of chemical components.

Dedicated simulation runs were performed to select the most suitable values for these parameters. The online trunkline model was given a mathematical grid of 84 cells, and it was able to run as fast as 40 times real time in a single mathematical processor. The predictive model is a copy of the online model, which uses a grid containing fewer cells than the online model (32) and performs its calculations 250 times faster than real time.

While the predictive model, to some extent, sacrifices accuracy, predictions are always found to be influenced more by the estimates and forecasts of future platform flows provided by the user as simulation inputs than by model inaccuracies.

Platform discharge pressures and gas delivery flow are computed by the simulation models and stored by executive application back into the data historian. Space discrete information, such as pressure, temperature, flow rates, velocities, and holdup profiles, are archived in a dedicated database. The historian user interface for the system shows actual conditions versus model-calculated data. Predictive what-if or look-ahead studies are triggered from the system graphical interfaces. The predictive model retains the process and trunkline internal conditions to serve as an initial state for simulations. The user is given the option to edit individual platform flows and onshore landing pressure or flow. The time horizon of the prediction is also an input value.

Simulation results provided by the online and predictive parts of the tool are available to the operators by means of graphical user interfaces (GUIs) that were constructed using historian tools. Predictive results are shown as a series of tables and charts that track the evolution of line conditions (i.e., Wobbe index, compositions, condensate holdup, velocities, flows, and pressure) along the trunkline distance and as a function of time. Figure 17-5 shows a comparison between the observed and the simulated Wobbe index of the gas at the terminal station.

Connecting the process engineering models to online measurement data stored in historians has been successful for the support of operational decisions. The described computer program was installed in the control room of the onshore processing facilities in March 2003, and the operators successfully utilize the system for predicting and managing the gas quality. The calibrated models developed for the online system are now also used by process engineers for offline analysis of the clean-out cycles for managing the large amount of condensate that develops inside the trunkline, for example. Potential slug catcher overflow was prevented by a correct adjustment of platform flows. The model-centric approach of the system and the

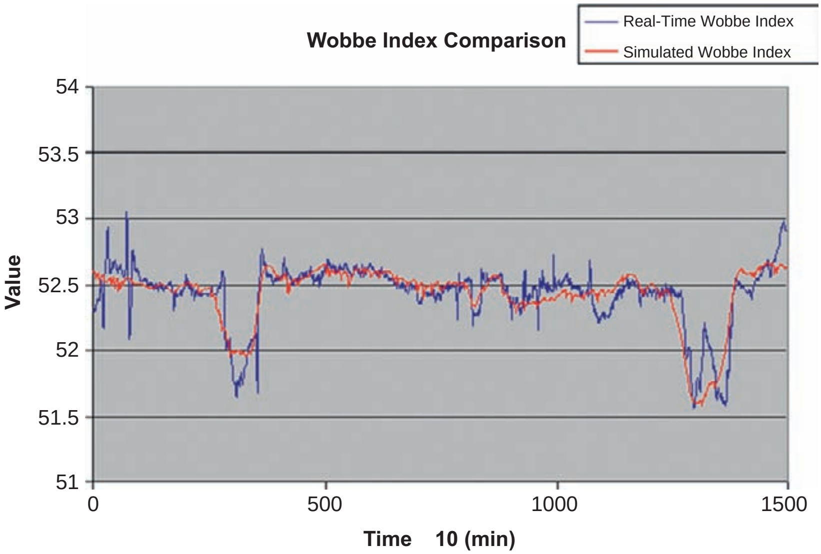

Figure 17-5 Comparison between the real and the simulated Wobbe index (La Riviére and Rodriguez, 2005).

clear split between application components and simulation model also facilitate model maintenance and upgrades. The speed of calculation is more than sufficient to keep the online model synchronized with real-time events. Similarly, the predictive case's time response is satisfactory for the end users. The accuracy of the calculated results for gas quality is more than sufficient to support the necessary operational decisions. Although the accuracy of the results regarding liquid inventories is not clear at present, the availability of these data is already a major improvement with respect to the previous situation when information on this aspect of the process was completely missing.

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Non-Print Items

CHAPTER 18

Real-Time Optimization of Gas Processing Plants

18.1 INTRODUCTION

Gas processing operations constantly experience changing conditions due to varying contracts, feed rates, feed compositions, and pricing. In order to capture the maximum entitlement measured in profits, these operations are prime candidates for real-time optimization (Bullin and Hall, 2000). Real-time optimization enables operating facilities to respond efficiently and effectively to the constantly changing conditions of feed rates and composition, equipment condition, and dynamic processing economics. In fact, world-class gas processing operations have learned how to optimize in real time to return maximum value to their stakeholders. Applications of real-time optimization of gas processing facilities have recently been adopted. Advances in computer power, robust modeling approaches, and the availability of real-time pricing have enabled this technology. An online optimization model also provides a continuously current model for accurate simulations required for offline evaluations and studies. Equipment conditions including fouling factors for heat exchangers and deviation from efficiencies predicted by head curves for rotating equipment are tracked over time. The impact of additional streams under different contractual terms can be evaluated with the most up-to-date process model available.

The objective of this chapter is to introduce the concepts of real-time optimization and describe the considerations for successful application in the gas processing industry.

18.2 REAL-TIME OPTIMIZATION

Real-time optimization (RTO) refers to the online economic optimization of a process plant, or a section of a process plant. An opportunity for implementing RTO exists when the following criteria are met:

- Adjustable optimization variables exist after higher priority safety, product quality, and production rate objectives have been achieved.

- The profit changes significantly as values of the optimization variables are changed.
- Disturbances occur frequently enough for real-time adjustments to be required.
- Determining the proper values for the optimization variables is too complex to be achieved by selecting from several standard operating procedures.

Real-time, online adaptive control of processing systems is possible when the control algorithms include the ability to build multidimensional response surfaces that represent the process being controlled. These response surfaces, or knowledge capes, change constantly as processing conditions, process inputs, and system parameters change, providing a real-time basis for process control and optimization.

Real-time optimization applications have continued to develop in their formative years with 100 or so worldwide large-scale processing applications. RTO systems are frequently layered on top of an advanced process control (APC) system, as shown in Figure 18-1, producing economic benefits using highly detailed thermodynamic, kinetic, and process models, and nonlinear optimization. Whereas an APC system typically pushes material and energy balances to increase feed and preferred products with some elements of linear optimization, RTO systems can trade yield, recovery, and efficiency among disparate pieces of equipment.

Processors frequently use RTO applications for offline studies because they provide a valuable resource for debottlenecking and evaluating changes in feed, catalyst, equipment configuration, operating modes, and chemical costs. Processing RTO has been hampered by the lack of reactor models for major processing units, property estimation techniques for hydrocarbon streams, and the availability of equipment models. Continued technology advancement has removed many of these hurdles, however, and the number of reported RTO successes continues to grow.

Real-time optimization systems perform the following main functions:

- **Steady-state detection:** Monitor the plant's operation and determine if the plant is sufficiently steady for optimization using steady-state models.
- **Reconciliation/parameter estimation:** Collect operating conditions and reconcile the plant-wide model determining the value of the parameters that represent the current state of the plant.
- **Optimization:** Collect the present operating limits (constraints) imposed and solve the optimization problem to find the set of operating conditions that result in the most profitable operation.

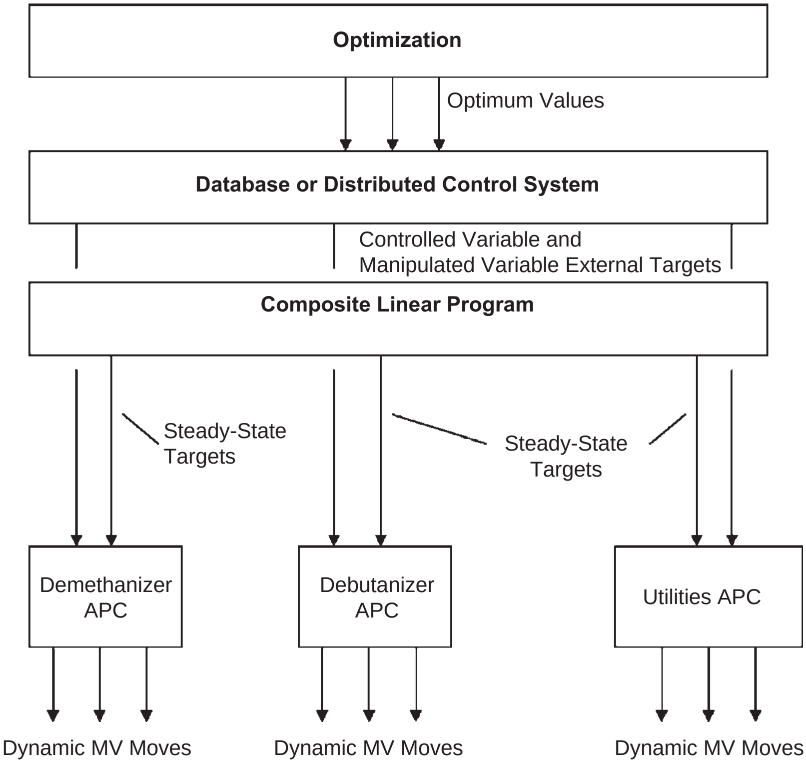


Figure 18-1 Advanced process control with optimization solution architecture.

- **Update setpoints:** Implement the results by downloading the optimized setpoints to the historian for use by the control system.

A good real-time optimization system utilizes the best process engineering technology and operates on a continuous basis. The system constantly solves the appropriate optimization problem for the plant in its present state of performance and as presently constrained.

A typical system consists of an efficient (fast) equation solver/optimizer "engine," coupled with robust, detailed, mechanistic (not correlation-based) equipment models, and a complete graphical interface that contains a real-time scheduling (RTS) system and an external data interface (EDI) to the process computer. Three primary components of a fully integrated graphical interface are shown in Figures 18-2 to 18-4.

The real-time optimization model is composed of separate models for each major piece of equipment. These separate models are integrated and are solved simultaneously. The simultaneous solution (rather than sequential) approach allows for solution of large-scale, highly integrated problems

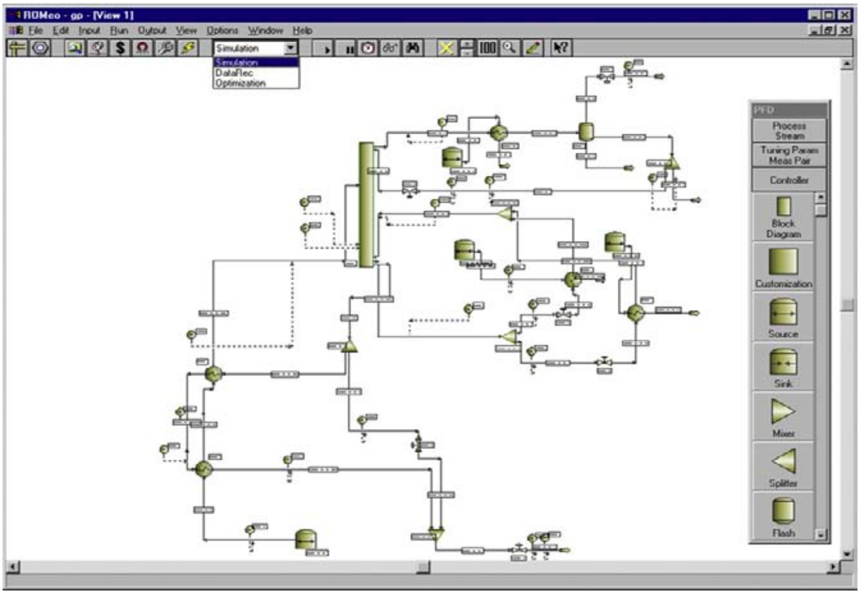
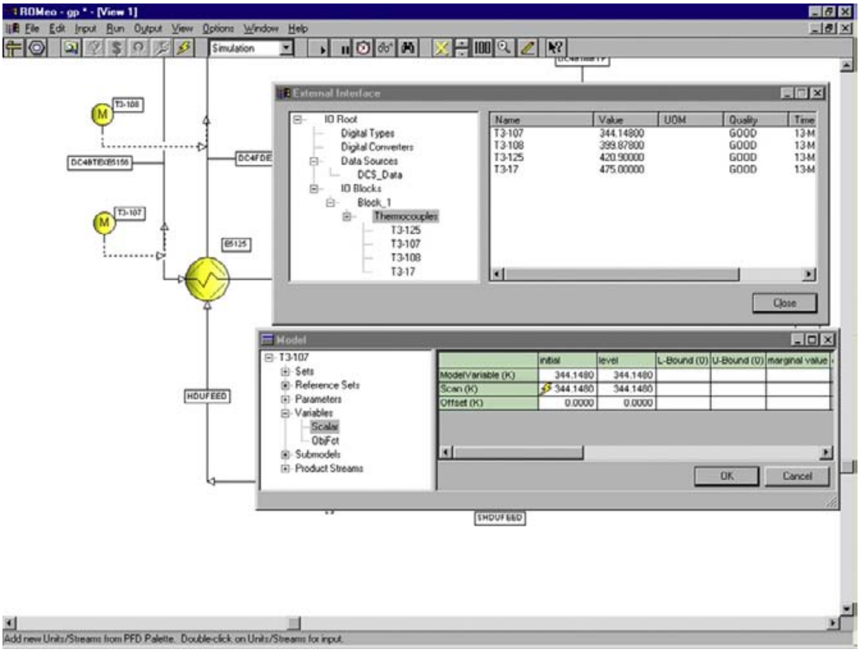


Figure 18-2 Real-time optimization model interface.

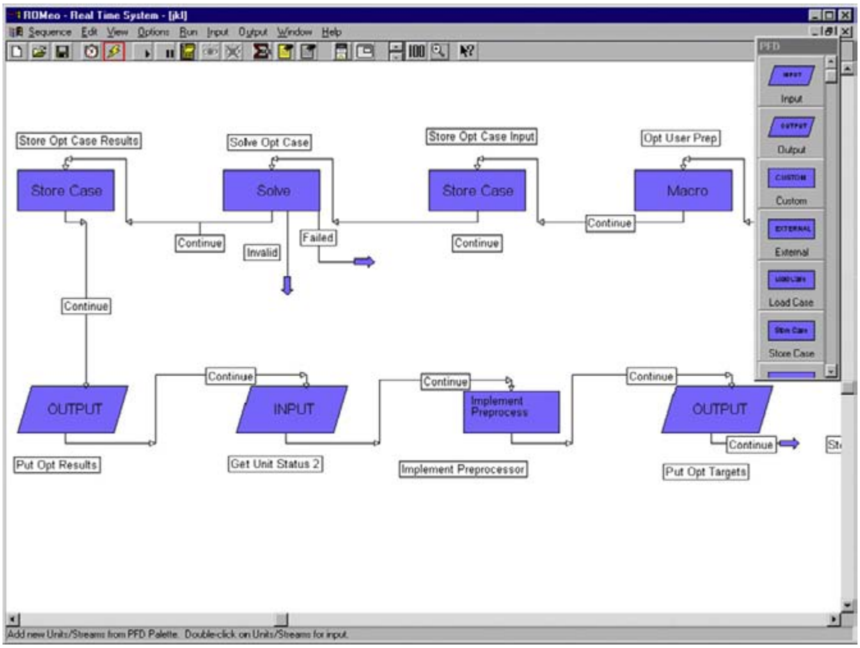


Name	Value	UOM	Quality	Time
T3-107	344.14800		GOOD	13-M
T3-108	399.87800		GOOD	13-M
T3-125	420.90000		GOOD	13-M
T3-17	475.00000		GOOD	13-M

	initial	level	L-Bound (0)	U-Bound (0)	marginal value
ModelVariable (K)	344.1480	344.1480			
Scan (K)	344.1480	344.1480			
Offset (K)	0.0000	0.0000			

Add new Units/Streams from PFD Palette. Double-click on Units/Streams for input.

Figure 18-3 Real-time optimization EDI interface.



Add new Units/Streams from PFD Palette. Double-click on Units/Streams for input.

Figure 18-4 Real-time optimization RTS interface.

that would be difficult or impossible to solve using sequential techniques offered by many flow sheet vendors.

A real-time optimization system will determine the plant optimum operation in terms of a set of optimized setpoints. These will then be implemented via the control system.

18.2.1 Physical Properties

All of the process models for a rigorous real-time optimization system use a mixture of physical properties, such as enthalpy, *K* values, compressibility, vapor pressure, and entropy. Equations of state, such as Soave–Redlich–Kwong (SRK) or Peng–Robinson (PR), are used for fugacities, enthalpy, entropy, and compressibility of the hydrocarbon streams. The enthalpy datum is based on methods such as the enthalpy of formation from the elements at absolute zero temperature. This allows the enthalpy routines to calculate heats of reaction as well as sensible heat changes. Steam and water properties are calculated using routines based on standards such as the National Institute of Standards and Technology.

18.2.2 Optimization Models

Models used in the optimization system must be robust, must be easily specified (in an automated manner) to represent changing plant situations, and must solve efficiently. These models must be able to fit observed operating conditions, have sufficient fidelity to predict the interactions among independent and dependent variables, and represent operating limits or constraints. State-of-the-art models meet these requirements by using residual format ($0 = f(x)$), open equations, fundamental, mechanistic relationships, and by incorporating meaningful parameters within the models. These state-of-the-art systems provide the highly efficient equation solver/optimizer and the interface functionality that automates sensing the plant's operating conditions and situations, and they automate posing and solving the appropriate parameter and optimization cases.

Most plant models are standard models. Sometimes custom models are created specifically for the equipment of a unit. All the models use thermodynamic property routines for enthalpy, vapor-liquid equilibrium, and entropy information.

Rotating equipment models such as compressor, pump, engine, gas turbine, and steam turbine models contain, along with all the thermodynamic relationships, the expected performance relationships for the specific equipment modeled.

18.2.2.1 Optimization Objective Function

The objective function maximized by a high-level optimization system is the net plant profit. This is calculated as product values minus feed costs minus utility costs, i.e., the $P - F - U$ form. When appropriately constrained, this objective function solves either the "maximize profit" or "minimize operating cost" optimization problem. Economic values are required for each of the products, feeds, and utilities. The value of each stream is derived from the composition-weighted sum of its components. Economic values for feeds, products, and utilities in the optimization system are reset on a regular basis for best performance.

18.2.2.2 Custom Models

Often custom models must be incorporated into a standard optimization package to predict proprietary processes and solvents. The custom model may be incorporated with special properties packages or integrated into the system in a semi-open approach where the iteration criterion is handled by the optimization system, but the actual kinetics or thermodynamic equations remain the same as the offline custom model. Proprietary gas sweetening solvent formulations may be the most common example of a custom model in the natural gas industry.

18.2.2.3 Fractionators

Fractionators are modeled using tray-to-tray distillation. Heating and cooling effects as well as product qualities are considered. Column pressure is typically a key optimization parameter. Temperature measurements are used to determine heat transfer coefficients for condensers and reboilers.

18.2.2.4 Absorbers and Strippers

Absorber and stripper units will be modeled using tray-to-tray distillation, if required. Component splitters may be used to simplify the flow sheet where acceptable when these units have little effect on the optimum.

18.2.2.5 Compression Model

The main optimization setpoint variables for the compressors are typically suction or discharge pressures. Important capacity constraints are maximum and minimum speed for the driver; maximum current for electric motors; and maximum power for steam turbines, gas turbines, and engines. Sometimes maximum torque will be considered for engines.

A multistage compressor model consists of models for a series of compressor stages, interstage coolers, and adiabatic flashes. Drivers are included for each compressor machine. For each single compressor stage, the inlet charge gas conditions (pressure, temperature, flow rate, and composition) and the discharge pressure specification are used with the manufacturer's compressor performance curves to predict the outlet temperature and compressor speed. The power required for the compression is calculated from the inlet and outlet conditions.

The first step in the development of the compression model is to fit the manufacturer's performance curves for polytropic head and efficiency to polynomials in suction volume and speed. The equations for each compressor stage (or wheel, if wheel information is available) take the following form:

$$E_\mathrm{p} = A * N^2 + B * N + C * N * V_\mathrm{s} + D * V_\mathrm{s} + E * V_\mathrm{s}^2 + F \tag{18-1}$$

$$W_\mathrm{p} = a * N^2 + b * N + c * N * V_\mathrm{s} + d * V_\mathrm{s} + e * V_\mathrm{s}^2 + f \tag{18-2}$$

where V_s is suction volume flow rate; N is compressor speed; E_p is stage or wheel polytropic efficiency; W_p is stage or wheel polytropic head; and A through F and a through f are correlation constants.

The polytropic head change across the stage or wheel can also be calculated from the integral of V_dP from suction pressure to discharge pressure. This integration can be performed by substituting $V = Z*R*T/P$ and integrating by finite difference approximation:

$$W_p = R * \ln(P_d/P_s) * ((Z_s * T_s) + (Z_d * T_d))/2 \qquad (18\text{-}3)$$

where R is gas constant; P_d is discharge pressure; P_s is suction pressure; Z_s is compressibility at suction conditions; T_s is absolute suction temperature; Z_d is compressibility at discharge conditions; and T_d is absolute discharge temperature.

For simplicity, the preceding equations are formulated in terms of a single integration step between suction and discharge, but in the actual implementation, each stage or wheel is to be divided into at least five sections to describe the true profile. The enthalpy change across the stage or wheel can be calculated from the inlet and outlet conditions, the polytropic head change, and the polytropic efficiency.

This analysis results in three simultaneous equations and three unknowns for each integration step. The unknowns are the discharge temperature, discharge pressure, and enthalpy change for each step except the last. The known discharge pressure for the last step is related to the speed of the machine. In state-of-the-art approaches, all of the integration steps are solved simultaneously.

Measured discharge pressures, speeds, and temperatures are used in the parameter estimation run to update the intercept terms in the polynomials used to represent the manufacturer's polytropic head and efficiency curves. These parameters represent the differences between the actual performance and expected/design performance. As compressors foul, these parameters show increasing deviation from expected performance. It is this difference that has significant meaning since an absolute calculation of efficiency at any moment in time can vary with feed rate and several other factors, which dilute the meaning of the value. By showing a difference from design, we get a true measure of the equipment performance and how it degrades over time.

The discharge flow from each stage is sent through a heat exchanger model coupled with an adiabatic flash. The heat transfer coefficient for each exchanger is based on the measured suction temperature to the next stage, corrected for addition of any recycle streams. Suction and/or discharge flows are measured to fit heat loss terms in the interstage flash drums.

18.2.2.6 Distillation Calculations

Standard tray-to-tray distillation models are used for distillation calculations in an optimization system. The "actual" number of trays is used wherever possible, and performance is adjusted via efficiency. This allows the model to more accurately represent the plant in a way that is understandable to a plant operator.

All distillation models predict column-loading constraints accurately as targets are changed. Condenser and reboiler duties are also calculated for predicting utility requirements and exchanger limitations.

18.2.2.6.1 Tray-to-Tray Distillation Method

An equation-based tray-to-tray distillation method is based on mass, heat, and vapor liquid equilibrium balances on each physical tray. Figure 18-5 depicts a typical distillation column tray.

Component mole balances for each component *i* on tray *j* are

$$F_{i} Z_{ij} + L_{i} + 1X_{(i+1)j} + V_{(i-1)} Y_{(i-1)j} - (L_{i} + LP_{i}) X_{ij} - (V_{i} + VP_{i}) Y_{ij} = 0 \tag{18-4}$$

The overall mole balance is

$$F_{i} + L_{i} + 1 \ + V_{i-1} - (L_{i} + LP_{i}) - (V_{i} - VP_{i}) = 0 \tag{18-5}$$

The vapor-liquid equilibrium definition is written in terms of the liquid mole fractions and *K* values:

$$Y_{ij} = K_{ij}(X_{ij}, T_{jl}, P_{i}) * X_{ij} \tag{18-6}$$

where T_{jl} is the liquid temperature on tray *j*.

The requirement that the mole fractions balance is expressed as

$$\sum [X_{ij}] - \sum [K_{ij}(X_{ij}, T_{jl}, P_{i}) * X_{ij}] = 0 \tag{18-7}$$

The Murphree vapor tray efficiency, which accounts for differences between the equilibrium vapor composition $[K_{ij}(X_{ij}, T_{jl}, P_{i})^{*} X_{ij}]$ and the actual mixed vapor composition $[Y_{ij}]$ leaving the tray, is an important adjustable parameter. It is defined as

$$E_{ij} = (Y_{ij} - Y_{ij-1}) / (K_{ij}(X_{ij}, T_{jl}, P_{i}) * X_{ij} - Y_{ij-1}) \tag{18-8}$$

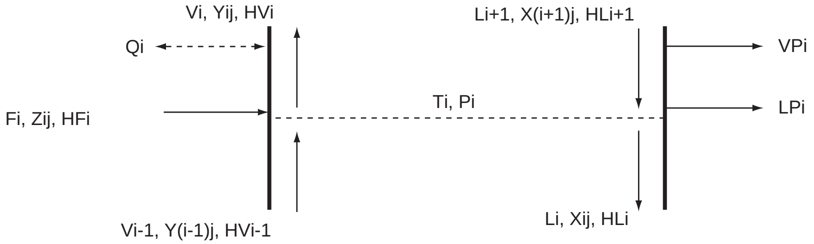


Figure 18-5 Mass transfer operations in a tray distillation column.

The vapor and liquid enthalpies are defined as

$$HV_i - HV(Y_{ij}, T_{jv}, P_j) = 0 \tag{18-9}$$

$$HL_i - HL(X_{ij}, T_{jl}, P_j) = 0 \tag{18-10}$$

where T_{jv} is the vapor temperature on tray j.

The overall tray heat balance is

$$\begin{aligned} F_i HF_i + L_{i+1} HL_{i+1} + V_{i-1} HV_{i-1} - (L_i + LP_i) HL_i \\ -(V_i + VP_i) HV_i + Q_i = 0 \end{aligned} \tag{18-11}$$

The liquid and vapor flow rates on the capacity-limiting trays are used in conjunction with tray-loading calculations to predict column pressure drops. The vapor area factor in the loading correlation is parameterized to enable matching calculated column differential pressure against the measured differential pressure. Upper limits are placed on this calculated differential pressure during the optimization case to ensure avoidance of flooding.

Condenser and reboiler limitations are handled by placing constraints on the minimum approach temperature in these exchangers or on the maximum heat transfer area in the case of variable-level exchangers. Heat transfer coefficients are calculated for the exchangers in the parameter case.

18.2.2.6.2 Demethanizer

Setpoint variables for a cryogenic demethanizer are typically the bottoms methane or carbon dioxide content and the overhead pressure. Constraint variables include overhead condensing capability and minimum column temperatures. The expansion impacts on the demethanizer and the overhead ethane losses are key profit variables.

The demethanizer is modeled with a tray-to-tray distillation model. Rigorous K values for the demethanizer column are recommended due to nonlinear vapor-liquid equilibrium relationships. Overall heat transfer coefficients for the reboilers and feed chillers are calculated from temperature and flow data.

18.2.2.6.3 Deethanizer

The setpoint variables in the deethanizer are the overhead pressure, the overhead propane content, and the bottoms ethane content. Any cooling medium (such as refrigeration, cooling water, or air) and heat medium (such as steam or hot oil) requirements of the deethanizer and the propane loss in the overhead are the major profit considerations.

The deethanizer is modeled with a tray-to-tray distillation model. Parameters include column pressure, column pressure drop, propane content in the deethanizer overhead, reflux flows, deethanizer bottoms ethane content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the exchangers.

18.2.2.6.4 Depropanizer

The setpoint variables for the depropanizer are the overhead butane content and column pressure. Key constraint variables are the propane in the debutanizer overhead, the overhead exchanger capacities, and column pressure drop (flooding). Any cooling and heat medium usage requirements of the depropanizer are the major profit considerations.

The depropanizer is modeled with tray-to-tray distillation models. Parameters considered include column feed temperature, column pressure, column pressure drop, butane content in the tops, reflux flow, bottoms propane content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the reboiler and condenser. A column capacity factor is also parameterized.

18.2.2.6.5 Debutanizer

The setpoint variables for the debutanizer are the overhead pentanes and heavier content as well as column pressure. Key constraint variables are the propane in the debutanizer overhead, the overhead exchanger capacities, and column pressure drop (flooding). Any heating and cooling medium usage requirements of the debutanizer are the major profit considerations.

The debutanizer is modeled with a tray-to-tray distillation model. Parameters considered include column feed temperature, column pressure, column pressure drop, pentanes and heavier content in the overhead, reflux flow, bottoms butanes content, and bottoms draw rate. Overall heat transfer coefficients are calculated for the reboiler and condenser. A column capacity factor is also parameterized.

18.2.2.6.6 Butanes Splitter

The setpoint variables for the butanes splitter are the column pressure, normal butane in overhead, and isobutane in the bottoms. The constraint variables of interest are reboiler and condenser loading and product specifications. The profit variables of interest are the isobutane losses in the bottoms normal butane stream.

The butanes splitter is modeled with a tray-to-tray distillation model. The parameters considered include column pressure, column differential pressure, normal butane in the isobutane product, bottoms isobutane concentration, reflux flow rate, bottoms flow rate, product draws, and bottom reboiler flow rate. Heat transfer coefficients for the exchangers are calculated.

18.2.2.6.7 Refrigeration Models

The main setpoint variable for refrigeration machines is the first-stage suction pressure. Refrigeration system models relate refrigeration heat loads to compressor power. The compressor portions of the refrigeration models use the same basic equations as compressor models discussed previously. Refrigeration systems should use the appropriate composition and will have a component mixture for any makeup gas.

The measured compressor suction flows and the heat exchange duties calculated by the individual unit models are used to determine the total refrigeration loads. The refrigerant vapor flows generated by these loads are calculated based on the enthalpy difference between each refrigerant level. Exchanger models of the refrigerant condensers are used to predict compressor discharge pressures.

18.2.2.6.8 Demethanizer Feed Chilling Models

The demethanizer feed chilling system is modeled as a network of heat exchangers and flash drums. These models are used to predict the flow rates and compositions of the demethanizer feeds. The effects of changing demethanizer system pressures and flow rates are predicted.

The demethanizer feed drum temperatures and feed flow rates are measured to fit the fractions of heat removed from the feed gas by each exchanger section. Flow and temperature measurements on the cold stream side allow the fraction of feed gas heat rejected to each stream to be estimated. The two sides of the feed exchangers are coupled through an overall heat balance. The inlet and outlet temperatures from the refrigerant and other process exchangers are used to fit overall heat transfer factors in the parameter case. A pressure drop model for the feed gas path is also included.

18.2.2.7 Steam and Cooling Water System Models

Heat and material balance models of the steam system are developed. These models include detailed representations of the boilers.

The cooling water system will be modeled with heat exchangers, mixers, and splitters to allow for constraining the cooling water temperature and its effects on the operation of distillation columns and compressors.

18.2.2.8 Turbines

A turbine model, as shown in Figure 18-6, is used for steam turbines (back pressure, condensing, or extraction/condensing) or any expander in which the performance relationship can be expressed using the following equation:

$$\text{Design Power} = A + B \times (\text{Mass Flow}) + C \times (\text{Mass Flow})^2 + D \times (\text{Mass Flow})^3 \tag{18-12}$$

Back pressure and condensing turbine expected performances are usually presented as essentially linear relationships between power and steam flow. Extracting/condensing turbine expected performance relationships are typically presented as power versus throttle steam flow, at various extraction steam flows. This kind of performance "map" can be separated into two relationships of the aforementioned form: one representing the extraction section, and the other the condensing section. An extraction/condensing turbine can be thought of as two turbines in series, with part of the extraction section flow going to the condensing section.

Design power refers to expected power from performance "maps" that are at specific design inlet pressure, inlet temperature, and exhaust pressure. Expected power is the power expected from a turbine operating at other than design conditions. The design power is adjusted by the "power factor," as illustrated here:

$$\text{Expected Power} = \text{Design Power} \times \text{Power Factor} \tag{18-13}$$

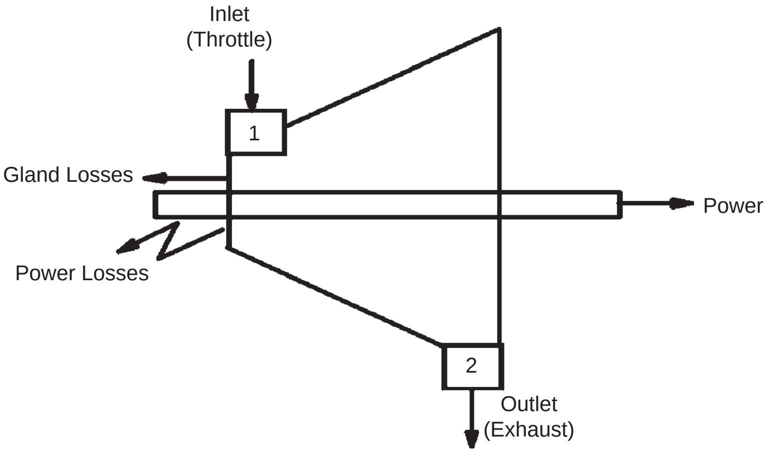


Figure 18-6 Schematic of turbine model.

$$\text{Power Factor} = \frac{\Delta\text{Isentropic Enthalpy at actual conditions}}{\Delta\text{Isentropic Enthalpy at design conditions}} \tag{18-14}$$

$$\text{Brake(shaft)Power} = \text{Expected Power} + \text{Power Bias} \tag{18-15}$$

Design power and expected power are equal if actual expanding fluid (i.e., steam) conditions are the same as design conditions. The power bias can be parameterized using the exhaust temperature for back pressure turbines or the extraction section of an extraction/condensing turbine. In both of these cases, the exhaust steam is superheated (single phase). For condensing turbines or the condensing section of an extraction/condensing turbine, the measured steam flow is calculated because the temperature of the two-phase exhaust cannot be used. The fraction vapor of the two-phase exhaust is determined by energy balance. The total energy demand can be determined from the compressor (or other driven power consumer). The power extracted from the extraction section can be determined from the throttle steam flow and the inlet and outlet (single-phase) conditions. The power extracted from the condensing section is just the difference between the total demand and the extraction section power.

The power loss is taken from the steam and affects its outlet conditions but is not transferred to the shaft or brake power.

To calculate the change of enthalpies needed for the power factor and for the energy balance, we calculate fluid conditions for each turbine section at the inlet (throttle), outlet (exhaust), and at inlet entropy and outlet pressure for both actual and design conditions. The nomenclature associated with the exhaust fluid conditions required for the enthalpy change calculations is

Exh IDM = Exhaust **I**sentropic at **D**esign inlet conditions for vapor/liquid **M**ixture

Exh IDV = Exhaust **I**sentropic at **D**esign inlet conditions for **V**apor

Exh IDL = Exhaust **I**sentropic at **D**esign inlet conditions for **L**iquid

Exh IAM = Exhaust **I**sentropic at **A**ctual inlet conditions for vapor/liquid **M**ixture

Exh IAV = Exhaust **I**sentropic at **A**ctual inlet conditions for **V**apor

Exh IAL = Exhaust **I**sentropic at **A**ctual inlet conditions for **L**iquid

For condensing turbines or the condensing section of extraction/condensing turbines, the exhaust pressure is used to specify the model. The parameter that this measurement updates is the condenser heat transfer coefficient. The exhaust pressure is free to move in the optimization cases, since the actual pressure moves as the condenser calculates the pressure required to condense the steam sent to the condensing turbine or condensing section.

The maximum mass flow at reference (usually design) conditions is used to predict the maximum mass flow at actual conditions as an additional constraint on the turbine performance. Sonic flow relationships are used for this prediction. This is the maximum flow through the inlet nozzles when the inlet steam chest valves are wide open. The nozzle area is not needed if the maximum flow at a set of reference conditions is known. Vendors usually list the maximum flow, not the nozzle area.

The model also includes the gland steam flow, which is used to counterbalance the axial thrust on the turbine shaft and which is lost through labyrinth seals. The gland steam does not contribute to the shaft power.

Typically, turbine performance is not considered to be a function of speed. Consequently, speed is a variable that has no effect on the solution. However, turbine performance is typically a very weak function of speed over a wide range, and its effect on performance is not typically presented on the expected performance "maps." The possibility of adding the weak effect of speed on performance will be considered when the model is being built.

18.2.3 Plant Model Integration

After the plant section models have been developed, they are integrated into the overall plant model. All of the interconnecting streams are specified and checked, and a consistent set of variable specifications is developed. At this stage, the overall validity of the plant model is checked using offline plant data.

Reconciliation/parameter and optimization cases are run, and the results are checked for accuracy and reasonableness. A material balance model is included in the plant model integration work to confirm an absolute model material balance closure. This material balance will include a furnace area balance, a recovery area balance, and an overall balance. In addition, the plant integration allows the objective function to be tested and validated with connections to all feed, product, and utility variables.

It is important to have the engineers that will be responsible for commissioning the optimization system involved in the project during plant model integration. A thorough understanding of the plant model is imperative for a smooth implementation of the online system.

18.2.3.1 Model Fidelity and Measurement Errors

In an optimization system as described here, neither the models nor the measurements need to be absolutely perfect for the system to work well and to deliver significant improvement in profitability. An online optimization system continuously receives feedback from real-time measurements. The model parameters are updated prior to each optimization so that the models fit the plant and the optimum setpoints calculated are valid and can be confidently implemented. Without this constant feedback of plant measurements and regular updating of the plant model, the optimization solution might not be feasible. The fidelity of the models and the accuracy of the measurements are reflected in trends of the parameters. During commissioning of the optimization system, the best available measurements are identified by analyzing many parameter cases, running with real-time data, and prior to closing the loop.

If, for example, significant heat balance discrepancies exist between process side and fuel/flue gas side measurements in a furnace, that discrepancy can be handled by determining the bias required on the fuel gas flow measurement(s) to satisfy the heat balance. That bias could be a parameter that is updated prior to each optimization. The variation of the bias over time (its trend) would be monitored. If this parameter varies significantly, the furnace model and other measurements used by it would be investigated thoroughly. Alternate measurements used to "drive" the parameter case solution would be investigated as well. The outcome of this analysis is that the best available measurements are selected, and the model relationships are thoroughly investigated to ensure that all significant relationships are included. This analysis is a standard and required step in building the optimization system.

Validity checking is an integral part and is built into an online optimization system. It is used to screen out gross errors. If alternate measurements are available, the validity checkers can use these when primary measurements are unavailable or bad. Generic validity checking takes care of common errors, while custom validity checking can respond to site-specific situations. Measurements can be designated as critical, so if they are unavailable, the optimization cycle will be directed to monitor for steady state and will complete its cycle only when the measurement becomes available. Validity checking has several features designed to keep the online service factor of the optimization system high in the face of imperfect measurements. Another feature of the online system is that the measurements used to drive the solution of the parameter case are averages over a specified time window (usually 1 hour) so that measurement noise is suppressed.

The better the measurements and the better the models, the better a real-time optimization can fully exploit the process equipment, and consequently, the more potential profit is realized.

Processing RTO in the future will likely include wider applications driven by demonstrated benefits, reduced implementation costs, and acceptance as a best practice. RTO applications are also becoming tightly intertwined with economic planning systems, where real-time pricing and contractual considerations are available.

Evolving technologies changing the value proposition for refining RTO include

- Detailed kinetic models for all major processing units and configurations, proven by reported applications.
- Optimization technology improvements that incorporate robust solvers, integer variables, and the capability to handle increasing problem sizes. Today's technology can handle applications with several hundred thousand equations; a typical refining application has 100 measurements, more than 100,000 variables, and 25 outputs.
- Greater integration with higher-level systems including shared models and reconciled measurement data.
- Multiunit optimization that leverages shared resources between process units and continues to lead toward rigorous refinerywide optimization.
- Computing technology improvements, which have already shifted RTO from minicomputers to personal computers, and which will allow more solutions per day and more complex formulations. Solver and computing improvements will eventually lead to true dynamic optimization.
- Application and model building tools, operating graphical user interfaces, and sustained performance technologies that will lower cost, improve benefits, and remove other hurdles.

18.3 REAL-TIME OPTIMIZATION PROJECT CONSIDERATIONS

The steps of a real-time project implementation include

- Front-end engineering design
- Flow sheet development
- Model testing and tuning
- Online open loop testing
- Online closed loop testing
- Application sustainment

During front-end design, the project objectives should be clearly defined, including

- Process envelope: The system boundaries need to be determined, including the process equipment included within the boundary. Various equipment lineups that will be considered for optimization are identified. Also, the modeling methodology decisions occur at this stage.

- Economic parameters and objective function: The profit function requires definition, including the specific economic parameters that contribute to the profit function.
- Source of process and economic information: Process and economic information will be available from various sources. Process information is typically acquired from a process historian, whereas economic information will come from various commercial sources. Care should be taken to determine a source that is current and updated frequently.
- Setpoints to be generated from the optimizer: Process setpoints are critical for effective optimization. These setpoints must be selected where process control action will be attained accurately. Advanced process control strategies are usually best for implementing setpoints from a real-time analyzer.
- Metrics required for success: Metrics to track optimizer online time as well as optimizer effectiveness should be determined. An owner of these metrics should be identified to take responsibility for the success and sustainability of the optimizer.

Flow sheet development is obviously a critical step in the implementation of a real-time optimization project. The flow sheet should be developed by subsections of the plant and imported into the main application flow sheet as block diagrams. Redundant streams and equipment used in the initialization phase are removed from the flow sheet, and the sections of the plant are connected together at the block boundaries to fully integrate the subsections into a single flow sheet. Unit and flow sheet customizations are then added to improve the model's representation of the process.

Model testing and tuning include reconciling the imported data in offline mode and testing the real-time sequencing. The standard data reconciliation report is used to identify the measurements with the worst mismatch, and corrective action is taken. The model is then tested by importing multiple sets of data via the electronic data interface before placing the model online. Optimization cases are run and verified. Steady-state detection is fine-tuned. Initial online runs are observed to verify transfer of optimal setpoints to the advanced control and distributed control systems.

Sustainment of the application is critical to the long-term success. An engineer should be assigned responsibility for maintaining the application and working with operations to resolve any concerns immediately. Metrics should be tracked and reported to measure the success. These metrics may include

- The time on optimization as a percentage of time on control, averaged over all optimizable controllers.
- The time the optimizable controllers had optimization setpoints rejected as a percentage of the time on control.

- The time the optimizable controllers ran at expired optimization setpoints as a percentage of the time on control.

Optimization setpoints should expire after a predetermined amount of time, reflecting the nominal period for which a single optimization solution is valid. A standard stream-factor tracking program will calculate and report these statistics.

18.4 EXAMPLE OF REAL-TIME OPTIMIZATION

The Gassled joint venture operated by Gassco and supported by Statoil applied real-time optimization at their Kårstø gas processing plant in Norway (Kovach et al., 2010). The model is called the Plant Production Performance Model (3PM).

Kårstø is the largest natural gas liquids (NGL) plant in Europe. The Gassled owners have first rights to book capacity. Spare capacity is available for any other qualified shipper and subject to published tariffs for transporting and processing gas.

Gassled has a flexible gathering network connecting the respective producers and processing terminals, allowing gas streams from several fields to be routed to different destinations. Mixing of gas streams provides sales gas quality with respect to gross calorific value (GCV), Wobbe index (WI), and CO_2.

Rich gas processing capacity at the Kårstø plant depends on several variables and constraints. One of the most significant variables is the feed gas composition. Simulations demonstrate that the rich gas processing capacity may be significantly lower than the nominal design capacity if the feed has a high NGL content.

New expansion to increase the capacity of the Kårstø facilities has increased the complexity of the Kårstø facilities. The need for an online model to determine plant capacity was recognized to enable the plant to operate with a high degree of capacity utilization, realizing that production regularity is an inherent property of the throughput obligations. Underutilization of capacity sacrifices processing fees. The real-time optimization model allows precise and reliable capacity predictions while reducing the work required to determine the capacity. Operating setpoints required to reach the predicted capacity are generated as well as information for maintenance planning and infrastructure development.

18.4.1 Process Description

Figure 18-7 shows a simplified process diagram of the Kårstø gas processing plant. As can be seen, rich gas enters the plant and is preconditioned by

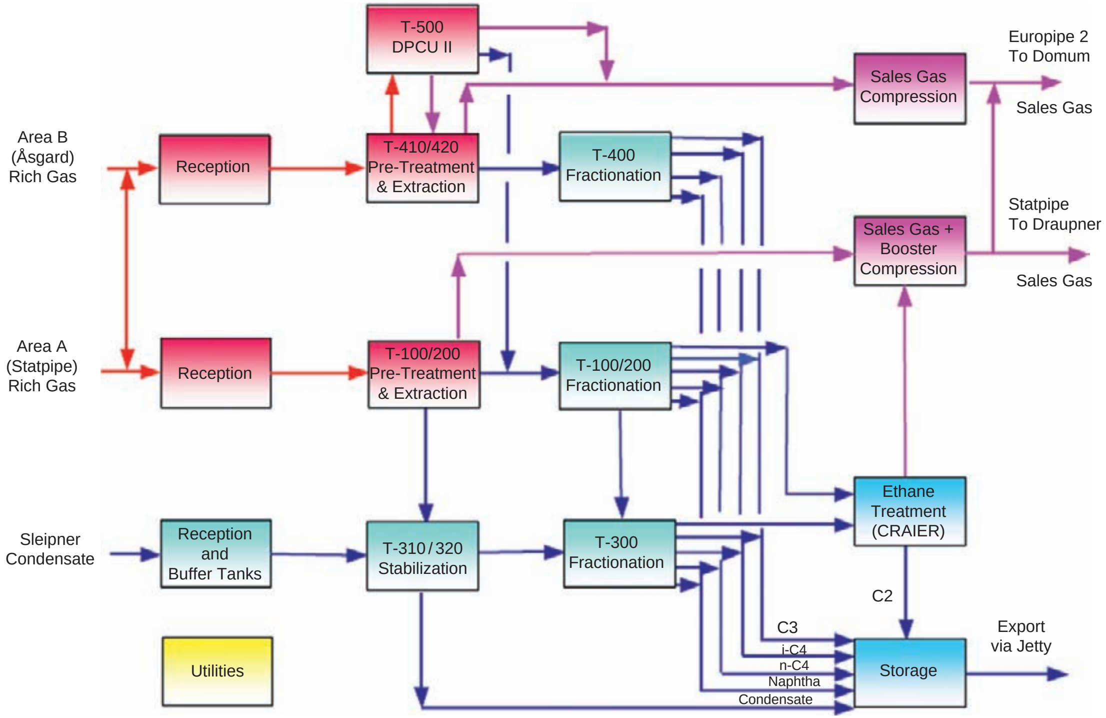


Figure 18-7 Process flow overview of Kårstø gas processing plant.

removal of H_2S and mercury. The gas is then preheated prior to dehydration, and NGL is extracted from the rich gas. The fractionation facilities produce raw ethane, stabilized condensate, propane, n-butane, i-butane, and naphtha. CO_2-rich ethane extracted from each train is routed for purification of ethane.

The processing facilities include 26 distillation columns. Steam is used as the heating medium for the reboilers on the distillation columns and to power turbines. Steam boilers and waste heat recovery systems on gas turbines generate the steam. Three levels of steam are employed. Additional utilities used are sea water for cooling and propane for refrigeration.

Sales gas is exported to two high-pressure subsea pipelines operated at pressures up to 189 barg. The export compression facilities include four compressor manifolds operated at different suction and discharge pressures. The compressors are driven by both gas turbines and electric motors.

In the Kårstø plant, the extraction trains may be bypassed. This allows more gas to be produced and processed. However, the bypass quantities are limited to provide sales gas quality to comply with the GCV, WI, and CO_2 specifications.

18.4.2 Plant Operation

Effective field production and development of new offshore fields require high-capacity utilization of the Kårstø plant. When the throughput margins are squeezed, the plant performance becomes increasingly sensitive to variations in feed gas composition, and the plant constraints become more evident. Operating toward the plant limits challenges plant regularity. A conceptual illustration of the intrinsic relation between these two characteristic parameters is shown in Figure 18-8.

Several parameters impact the rich gas capacity at the Kårstø plant. Some of the most important are rich gas composition, NGL recovery, product purity, and CO_2 recovery. Production rates and the fluid composition determine the rich gas composition to the Kårstø plant. Significant and sometimes rapid changes in the feed composition must be handled continuously. Changes in feed composition may move the capacity constraint from one facility to another. For example, a richer composition of the feed gas may cause the plant to reach a constraint in the NGL fractionation. A leaner gas could hit a constraint in the extraction facilities. Crossover piping allows several combinations of routing of the gas streams.

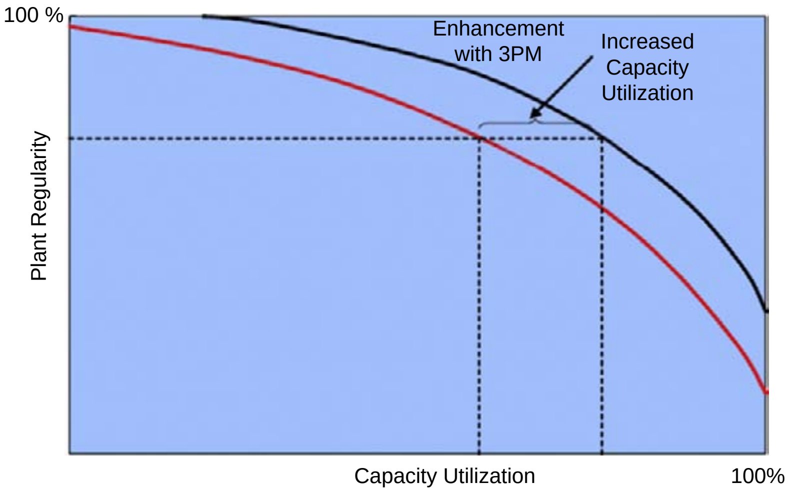


Figure 18-8 Trade-off between plant regularity and capacity utilization (Kovach et al., 2010).

The NGL extraction trains at the Kårstø plant are designed to provide about 90% recovery of NGL. However, recovery of NGL products may be limited for some feed compositions by the fractionation capacity. When the production objective is to maximize rich gas throughput, the NGL recovery can be decreased to process more gas until reaching a constraint in the extraction facilities, compression, preconditioning, or other parts of the plant. Specifications for the liquid products, defined in the transportation and processing agreements, require a minimum purity.

The carbon dioxide concentration of the sales gas is a contractual constraint with a limit of 2.5% mole. The CO_2 removal capacity of the plant is not fixed. Feedstock composition and gas rate are the major factors that affect CO_2 removal capacity. Optimum CO_2 removal can be affected by the routing of gas and liquids through the plant as well. Rich gas can be split between two treating systems or bypassed and delivered to two export pipelines. Also the GCV is controlled between an upper and lower constraint. Rich gas may be routed through one of the bypass connections to increase plant throughput, but this rate is then limited by the upper GCV constraint. To avoid breaching the lower GCV constraint, the plant can reinject extracted ethane in the sales gas. Furthermore, the sales gas streams can be mixed at the compressor manifolds by utilizing the downstream crossover. In addition to increased capacity utilization, the crossover pipelines allow enhanced flexibility of the operations with respect to handling of feedstock variations. It is possible to mix the various feed streams to optimize NGL recovery, CO_2 extraction, and quality of the products.

18.4.3 Production Objectives

The primary production objective is typically plant throughput and to deliver sales gas and products within the specifications. However, the production objectives include maximum daily throughput, high annual capacity utilization, optimum NGL production, and optimum fuel gas consumption.

Most shippers request high production at Kårstø throughout the year, taking into account the seasonal swing in gas demand. This means high demand for processing capacity at Kårstø throughout the year.

The operator is responsible for coordination of the yearly maintenance planning of all the installations connected to the gas transport infrastructure. A primary objective is to obtain a total plan where the availability of gas for deliveries to the market is maximized. Depending on the extent of yearly maintenance at Kårstø as well as at the upstream fields, this could put restrictions on production from certain fields, or allow accelerated production of more NGL rich or CO_2 rich gas from other fields. Maximizing CO_2 production from CO_2 rich fields could imply postponed or reduced investments in future CO_2 removal capacity to meet sales specifications.

In periods when processing demand is below the plant capacity (when Kårstø has no bottleneck effect on the offshore production), the primary operational objective is to maximize NGL recovery in order to provide for increased value creation for the shippers. For the NGL products, this means achieving the minimum product purity.

The Kårstø plant is a large energy consumer, and optimizing energy consumption is an important objective. However, optimizing energy consumption should not compromise the primary production objectives, related to the value creation at Kårstø.

18.4.4 Project Drivers

Value generated by real-time optimization for the Kårstø plant operations comes from the following:

- Increased utilization of the plant capacity, by introducing real-time optimization
- Improved quality of the capacity figures issued for booking

- Reduced time needed to prepare for the booking process
- Improved position in the business development process

On a daily basis, the feed stream compositions will vary. Processing feeds of variable composition require the plant control system to give fast and accurate responses, in order to maintain production targets. Skilled operators have a basic understanding of the plant operational characteristics and learn how to respond to feed disturbances. However, in transition periods, the plant capacity will not be fully utilized. Also, the operators may not push the plant to its full capacity, or they may choose a suboptimal routing of the gas through the processing facilities. On a regular basis, the real-time optimization determines optimized setpoints for the advanced control system, thereby ensuring a rapid and smooth transition period to new optimal plant conditions.

An illustration of the possible benefits with the 3PM employed for RTO to handle feed composition variations is shown in Figure 18-9. In this example, the primary operational objective is to achieve maximum plant throughput. The upper curve illustrates achieved production over a day when the operator employs the 3PM. The lower curve illustrates production with no 3PM implemented. It should be recognized that the upper curve assumes a shorter transition period following changes in the feed conditions. The 3PM model enables an increase of the plant throughput by 1%.

Plant regularity has a very high focus, and a regularity target of at least 98.5% is set for the Kårstø plant. However, when failure occurs, emphasis is put on maintaining the highest possible service degree to limit upstream consequences to oil production and minimize the consequences for the gas

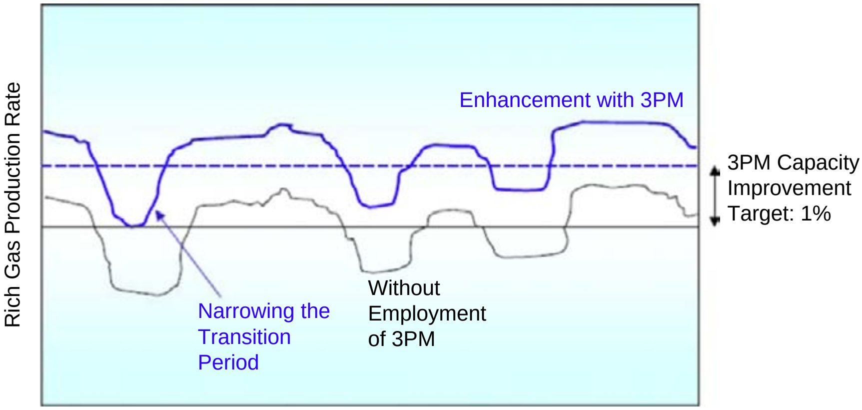


Figure 18-9 Illustration of enhanced processing capabilities.

customers. In situations in which equipment exhibits underperformance such as fouling in heat exchangers or degrading of compressors, this is revealed by the optimization model, and appropriate actions to correct the problem could be executed. If equipment fails to work or is temporarily out of service, the optimization model automatically computes new setpoints of the advanced process control to minimize the consequences.

The operator of the plant is responsible for issuing capacity figures for the booking processes. Often there are requests for processing capacity beyond current capacity at the plant. This puts pressure on the capacity margins of the plant. With the optimization model, which closely mimics the real operation performance, Gassco can provide enhanced confidence to reduce the uncertainty margins.

Extraction of NGL from the rich gas and subsequent fractionation into commercial products adds value to the shippers. Some shippers limit their gas deliveries based on their booked fractionation capacity, whereas shippers with leaner gas are limited by their booked extraction capacity. Any free processing capacity is identified and can be available to the shippers, allowing them to optimize their petroleum portfolio. This allows accelerated production of NGL-rich gas to maximize revenue for the shippers.

Boosting processing capacity within the limits of the facilities is the primary objective. The online model improves planning with a higher accuracy than the current process simulation models. Improved quality and accuracy of the planning tool results in a higher confidence in the calculation of the maximum plant capacity, denoted available technical capacity (ATC). The capacity committable on a long-term basis is denoted maximum available capacity (MAC). This allows for an operational margin for daily operations, as shown in Figure 18-10.

Reduced uncertainty due to the capability of the optimization model may narrow the operational margin and thereby increase the MAC. The biannual capacity estimation process is demanding and time consuming. Performance of existing equipment and systems that limit capacity utilization of the plant requires evaluation. The optimization model allows this estimation process to be run more efficiently.

18.4.5 Features of the Optimization Model

A single model that is applied for various purposes minimizes the inconsistency between various simulation results and the actual operation of the plant. A high degree of confidence in the planning scenarios and simulation results is achieved as the same model is applied for optimization of the plant

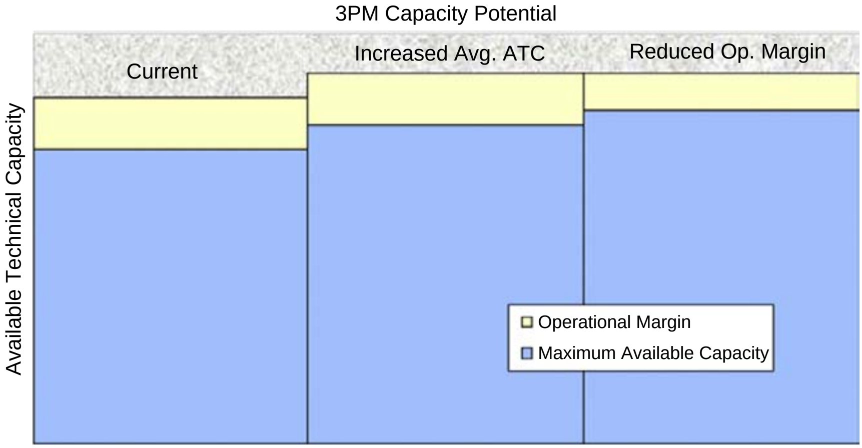


Figure 18-10 Increased available technical capacity and maximum available capacity.

in real time. With the increased complexity of the plant, and thereby the increase in number of opportunities for routing/processing the gas at Kårstø, a model of this type provides a means of improving the plant performance by optimizing the operation by use of a mathematical model to compute the optimum operation point and routing.

18.4.5.1 Implementation and Usage of the Model

The model is built in a commercially available equation-based modeling tool developed specifically for real-time online optimization in the hydrocarbon and chemical industry. This single model is used for all purposes of the 3PM project, i.e., real-time optimization, simulation, and planning. The setup of the system is described briefly here. The online model automatically reads plant measurement tags, and based on these readings, a steady-state test is performed. With the steady plant, a data reconciliation with the actual plant data is performed. Afterward, the reconciled model is subject to optimization with a given objective function.

The online system is located at the Kårstø gas processing facility. The offline system is located at both Kårstø and the Gassco. The model includes the four gas separation trains (serving the Åsgard and Statpipe rich gas pipelines), the Sleipner condensate train, and simplified representations of the utility (steam and refrigeration) systems. This results in the following unit counts:

- Distillation columns: 26
- Single compressors: 18
- Paired expander-compressors: 13

- Process measurements: 1600
- Controllers (regulatory and advanced control, 175 and 110 variables, respectively)
- Associated heat exchangers, drums, motors, furnaces, pumps, pipes, and valves

The output of the optimization is a set of setpoints to the advanced control system. The online model is anticipated to run in intervals of 2–4 hours, depending on solution time, performance, and actual implementation and usage in the daily routines.

The planning interface is a simplified representation of the plant in an external graphical user interface (GUI), where one has the opportunity to generate new feeds by mixing different field flows and compositions, to test out the performance in simulation, as well as optimization mode. The simplified representation of the plant consists in principle of a block diagram, where the blocks represent sections of the plant. The planner then has the option to switch off one or more blocks and check the capacity (performance) of the plant with this very simplified model user interface. The fact that the planner has a simplified overall GUI for the model of the plant, with a number of sections representing logical groups of unit operations in the plant, makes it easy to do scenario analysis for both new fields (new compositions) and for shutdown and maintenance. In each case, it becomes easier for the planning department to give a precise answer to the capacity of any feed with a given layout of the plant.

Simulation is done in the core model GUI, which is also used in the real-time optimization mode. Further, new plant developments may be simulated with the model in order to get a better prediction of throughput capacity than a conventional simulation would give.

The online model is located on a computer in the plant control room. It is running in a fully automated fashion, on a given interval basis. The model immediately sends the solution to a central storage computer. From here, the model can then be accessed by planners and offline simulation engineers, who copy the solved models to their own computers, respectively. This facilitates a common arena for operations as well as the planning and engineering departments.

The online model produces results to various types of end users. In addition to generating the actual setpoints to the advanced control system (and the operators), the model generates a set of reports with different types of information every time it is solved. The type of information to be generated is mass and energy balances, production figures, economical performance, utility consumption, capacity utilization of each unit, shadow prices, and equipment monitoring.

18.4.5.2 Modeling and Optimization Strategy

The plantwide model for the Kårstø plant comprises all processing facilities, as well as the steam boilers and other utilities. The single plant model is based on open equations. The individual unit operations are modeled by use of standard library models. In the cases of rotating equipment and important valves, Statoil and original vendor performance curves are used. The plant instruments that the model reads and applies are carefully chosen, in order to ensure that sufficient and trustworthy signals form the basis of the data reconciliation and, subsequently, the optimization. All alternative operation scenarios in terms of equipment failure, maintenance, etc., are dealt with by lineups defined by a set of macros.

The various crossovers that are normally in use in the plant are treated as continuous variables in order to avoid integer variables in the optimization problem. Only equipment that plays an active role in the plant is modeled, and the set of compounds used are as comprehensive as necessary to achieve representative results. This is in order to maintain a fairly reasonable size (and solution time) of the model.

The model (and the planning facility) comprises a set of objective functions from which one can be applied according to the desired operation mode. The objective functions for the optimizer reflect operation modes that are, or may be, relevant to operation of the Kårstø plant. The objective functions are predefined and can be chosen arbitrarily by the system administrator. Typical examples of objective functions are to maximize sales gas production or to maximize liquid NGL production. For all cases, the general product qualities are modeled as constraints together with the relevant operational limitations.

There are over 100 flags to indicate the routings and equipment status in the plant. While the physical realities prevent this from being a true combinatorial problem, in practice there are easily over a hundred possible plant configurations. This poses challenges in both making sure that the appropriate equipment is on or off in the model, but more importantly, in creating a good starting point for the equation-based solver.

In addition, the CO_2 removal and ethane recovery unit can operate near the CO_2/C_2 azeotrope, and some portions of the unit approach the critical point of the mixture. As with most modern gas plants, it is heavily heat integrated with many heat-pumped columns, cold boxes, and paired expander-compressors.

18.4.5.3 Online Usage

The online system follows the sequence of detecting steady state, checking data consistency, reconciling the data, calculating the optimal setpoints, and sending the optimized setpoints to the controllers.

18.4.5.4 Quantifying Measurement Errors

The 3PM model imports over 1,600 process measurements, which are used in data reconciliation. Of these, several hundred were identified as key process variables and have results from the data reconciliation runs written back to the database. The data exported for these points consist of the sample value used in the data reconciliation, the reconciled value, and the measurement offset.

Some nonmeasured variables such as compressor efficiencies are also saved. The scan and reconciled values are accessible on operator process displays. The same operator displays have links to preconfigured trends of the measurement offsets. This gives operations a quick way of evaluating the accuracy of their process measurements and seeing inferred values (e.g., efficiencies and unmeasured temperatures) in a familiar format.

Data reconciliation detects and runs when the plant is steady and puts the data in a readable format that supports scanning the data over long time periods. Any trended offset that is not distributed around zero and that has a consistent bias may be in error.

A suspected point can be immediately checked. Quantifying measurement errors can also help improve the accuracy of other offline simulation tools. Any process variable that is used as a specification in a process simulator needs to be accurate. Detecting a bias can significantly improve the accuracy of the simulation results.

In addition to highlighting possible measurement errors, an understanding of the behavior of unmeasured variables such as intermediate temperatures or efficiencies can have a positive impact on operations. Data reconciliation results have highlighted differences in efficiencies between parallel compressor trains and furnaces.

18.4.5.5 Offline Usage

The offline system consists of a web-based interface, a SQL data repository for storing cases, and rigorously validated models from the online system. The interface is designed to facilitate queued simulation and optimization runs of the most common plant configurations using various combinations of flows from the 30+ fields that feed the plant. Planning personnel have the option to run the model either through the standard web-based simulation interface or through the more detailed model builder interface. The latter is used to explore special situations that are not supported by the web-based interface. This requires a more detailed knowledge of the model and the simulation software. The model is run in this mode only to conduct special studies with high value.

18.4.5.6 Use for Planners

The user of the planning system can use two strategies to create a starting point. In the first, the user starts with an online model taken from the fully operational plant and then specifies the sections that are to be turned off and the feed rates. In the second, the user begins with an online model that very closely resembles the plant configuration he wishes to model.

For general studies, the first method is simpler and will generate acceptable results. For very specific conditions such as modeling CO_2 constraints, the second method will generate the most accurate results because the model tuning parameters will have been reconciled for the exact conditions of interest.

The second method is feasible because the online system archives valid models to a separate directory. In this way, a library of models for all process operating modes is created automatically.

To locate the time of the desired operating mode, the user can either look at trends of the appropriate equipment flags or use the operator display to identify the time and date that the appropriate model was created. Once the date is determined, the user can then copy the model from the library and move it into the planning system for further study.

18.5 REFERENCES

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CHAPTER 19

Environmental Aspects of the Natural Gas Supply Chain

19.1 INTRODUCTION

Natural gas is the most environmentally friendly fossil fuel. It is cleaner burning than its fuel cousins coal or petroleum because it contains less carbon. Natural gas also has less sulfur and nitrogen compounds, and, when it is burned, it emits fewer ash particulates into the air than coal or petroleum fuels. Natural gas demand in the United States is expected to be 26.55 trillion cubic feet by the year 2035, which indicates an increase of 16% over 2009 demand levels (EIA, 2011). This increase is expected because emissions of greenhouse gases are much lower with the consumption of natural gas relative to other fossil fuel consumption (Speight, 2007). In addition to natural gas being an abundant and secure source of energy worldwide, its use also offers several environmental benefits over other sources of energy, particularly other fossil fuels. When natural gas is burned, it emits lower quantities of greenhouse gases and criteria pollutants per unit of energy produced than do other fossil fuels. The reason is that natural gas is more easily fully combusted, and it contains low carbon content and fewer impurities than any other fossil fuel. For example, U.S. coal contains 1.6% sulfur (a consumption-weighted national average) by weight. The oil burned at electric utility power plants ranges from 0.5% to 1.4% sulfur. Diesel fuel has less than 0.05%. Comparatively, natural gas at the burner tip has less than 0.0005% sulfur compounds. The amount of carbon dioxide produced for an equivalent amount of heat production varies substantially among the fossil fuels, with natural gas producing the least. Nevertheless, gas production and processing operations have the potential for a variety of impacts on the environment. These impacts include atmospheric, aquatic, terrestrial, ecological, and noise impacts. The extent of impact depends on the stage of the process, the size of the plant, the nature of the surrounding environment, and pollution prevention and pollution control techniques.

Emissions from natural gas production accounted for approximately 66% of CH_4 emissions and about 25% of nonenergy CO_2 emissions from the natural gas industry in 2006. Processing facilities accounted for about 6%

of CH_4 emissions and approximately 74% of nonenergy CO_2 emissions from the natural gas industry. CH_4 emissions from the natural gas transmission and storage segment accounted for approximately 17% of emissions, while CO_2 emissions from natural gas transmission and storage accounted for less than 1% of the nonenergy CO_2 emissions from the natural gas industry. Natural gas distribution segment emissions, which account for approximately 10% of CH_4 emissions from natural gas systems and less than 1% of nonenergy CO_2 emissions, result mainly from equipment leaks from gate stations and pipelines (EPA, 2010).

As the global community will rely heavily on gas supplies for the foreseeable future, the challenge is to meet world energy demands while minimizing adverse impacts on the environment. The broad environmental issues faced by the gas production and processing industry are manifested at both local and global levels. They include habitat protection and biodiversity, air emissions, climate change, marine and freshwater discharges, incidents and spills, and soil and groundwater contamination. The gas processing industry has been proactive in the development of management systems, operational practices, and engineering technology targeted at minimizing environmental impact. These practices have been supported by environmental regulatory frameworks, and they reduced the number and impact of environmental incidents worldwide.

The purpose of this chapter is to provide an overview of environmental issues associated with natural gas production, processing, and transmission. It also discusses the environmental management aspects and the best approaches to achieving high environmental performance.

19.2 ATMOSPHERIC IMPACTS

Natural gas imparts the most environmentally friendly fossil fuel presently used as compared to coal or oil. The reason is that natural gas is clean and has a low-carbon quantity. However, the processing of natural gas may instigate emissions of gases or chemicals that have an impact on the environment. Three principal emissions in the gas processing industries as well as their effects on the environment are discussed (IPIECA, 2006).

19.2.1 Air Pollutant Emissions

Natural gas, when compared to other fuels, causes less pollution due to its smaller amount of impurities and less intricate chemical structure. The main component of natural gas is methane. In an elementary scenario, the complete combustion of one molecule of pure methane with two molecules of pure oxygen results in the production of one molecule of carbon dioxide gas, two molecules of water vapor, and energy in the form of heat. However, in practical applications, the combustion of methane takes place in the presence of air and not in pure oxygen, causing the production of a few pollutants. The products of the reaction are composed of carbon particles, carbon dioxide, carbon monoxide, NO_x, water vapor, and heat. The carbon particles, NO_x, and carbon monoxide are the standard indications of various other pollutant emissions. The amount of the products released from the reaction depends on the combustion efficacy (EIA, 1999).

In reality, the composition of natural gas is not solely methane, and air is not merely composed of pure oxygen and nitrogen, which results in the release of supplementary pollutants from the combustion of natural gas.

The U.S. Environmental Protection Agency (EPA) examines the surrounding air throughout the United States and has a dynamic administration system to manage atmospheric emissions from every source, as well as the natural gas industry. Despite the fact that the natural gas industry provides several advantages to the environment when compared to other energy suppliers, predominantly fossil fuels, some air emissions still pass through inspection and generation operations. The emissions and their origins are assessed. In the same way as any production or manufacturing operations, NO_x are developed from the combustion of fossil fuels to generate power to industrial machinery such as compressors and throughout gas flaring processes. Besides the NO_x, volatile organic compounds (VOCs) may be discharged throughout the dehydration process of natural gas. However, this emission of VOCs is less common in natural gas production than oil operations since gas operations are predominantly secured processes from the well to the pipeline with fewer chances for release (see Figure 19-1). Furthermore, emissions of aromatics such as benzene, toluene, and xylene are minor since they are not present in substantial amounts in the gas stream. Overall, the oil and gas industries contribute the least amount of emissions relative to other sources (U.S. DOE, 2009). Nevertheless, emissions that result from flaring, venting, and purging of gases as well as processing and transportation of the natural gas may cause atmospheric pollution and, therefore, affect the public health (E&P Forum/UNEP, 1997; EDF, 2011).

Natural gas contains sulfur in which a portion is removed during the processing of the gas. Yet, the remaining amount results in SO_x, mainly sulfur dioxide, acids that include sulfur and many other sulfuric compounds toxic to the atmosphere. Furthermore, since natural gas is odorless and colorless,

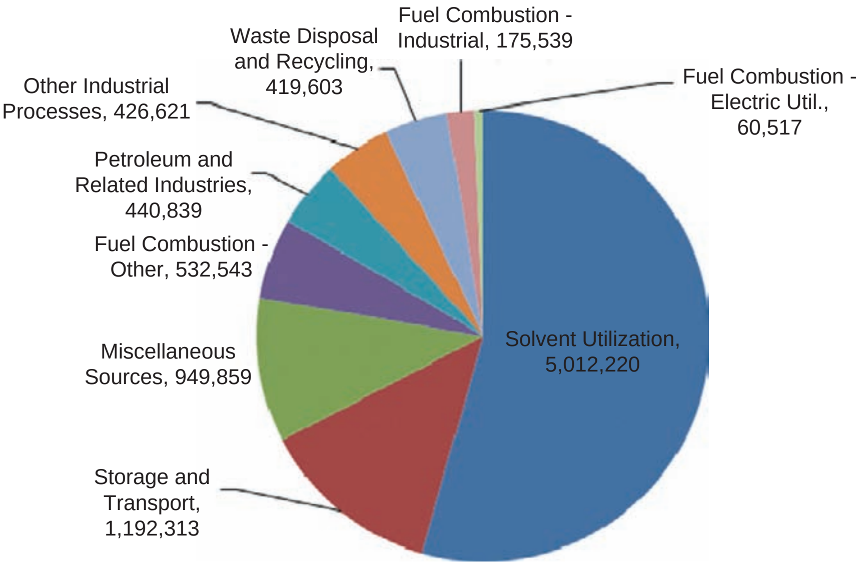

Figure 19-1 VOC emissions from various sources in tons, 1999 (U.S. DOE, 2009).

sulfuric odorants are added to the gas flow in order to detect any leaks (EIA, 1999). Hydrogen sulfide-bearing natural gas shall not be released to the atmosphere. In case of inadvertent release of H_2S, every effort must be made to eliminate the source or ignite it. The effect of H_2S and SO_2 on human health and the environment is shown in Table 19-1 (Younger, 2004).

Table 19-1 Effects of H_2S and SO_2 on Human Health and Environment

	H_2S	SO_2
Noticeable by odor	0.02–0.1 ppm	3.0 ppm
Irritating to the eyes, nose, or throat	20–90 ppm	10.0 ppm
Accepted concentration for industrial working conditions	10 ppm	5.0 ppm
Inhibits the human sense of smell	100 ppm	-
Hazardous to life*	300–1,000 ppm	> 2,000 ppm
Produces instant death	1% (10,000 ppm)	-
Threshold for plant damage	20–40 ppm	0.3 ppm**
Tarnishes paint and silverware	0.1 ppm	-
Accelerates rusting of iron and steel	-	Varies with humidity, but SO_2 is worse than H_2S

*At 300 ppm, the exposure may be over a matter of hours, while at 1,000 ppm there is almost instantaneous unconsciousness.

**The effect of sulfur dioxide varies with the type of vegetation, the time of exposure, and the concentration during the exposure. For sensitive plants, 0.3 ppm for 8 hours is the threshold condition and 1.0 ppm for 1 hour is similar.

19.2.2 Gas Flaring Emissions

Flaring of gas is the chief source of gas emissions in the oil and gas industries (E&P Forum/UNEP, 1997). Flaring of a generated gas is usually carried out when the infrastructure to process the gas is not available or when volumes of produced gas are not enough to be further processed or sold (Shires and Loughran, 2004).

There are several types of flares deployed in gas operations; they vary from miniature open-ended pipes at production wellheads to sizeable horizontal or vertical flares with pilot burners at gas plants. During flaring, the combustion products emitted are primarily CO_2 and N_2O. CH_4 gases may be emitted due to incomplete combustion or from irregular situations in which the flare tip does not give out a flame, resulting from operational difficulties (Shires and Loughran, 2004). In addition, CO emissions can result from flares due to incomplete combustion (U.S. DOE, 2009).

When feasible, produced gases are put to use or managed economically, therefore, reducing the need for flaring significantly (E&P Forum/UNEP, 1997). Flaring is rarely required in natural gas operations except through brief stages of well testing, workovers and completions, and irregular conditions such as transitory pipeline cessation (U.S. DOE, 2009). Flaring may also take place in circumstances as a safety measure during starting points, maintenance, or at processing interruptions (E&P Forum/UNEP, 1997).

19.2.3 Methane Emissions

While natural gas has a comparatively low carbon composition, its major component is methane. Methane is an influential greenhouse gas (GHG) with a radiative potent strength 23 times that of CO_2 on a weight basis throughout a 100-year interval. After CO_2, methane is the lead producer of anthropogenic GHG releases and is responsible for approximately 16% of the total amount on a CO_2 equivalent basis. Consequently, methane emissions are a significant part of every GHG emission mitigation plan (Sakmar, 2010).

In 2003, anthropogenic methane emissions in the United States gave an overall 26.2 million metric tons of methane, which was a 0.1 million metric tons increase from 2002 (see Table 19-2). That said, methane emissions in the United States were 4.6 million tons less than those in 1990, a drop equal to 1.5% of the overall anthropogenic greenhouse gas emissions in the United States (EIA, 2004).

Methane emissions take place in every division of the gas industry, from primary well drilling throughout production and processing and finally to delivery, as shown in Figure 19-2 (INGAA, 2000). Emissions chiefly arise

Table 19-2 Anthropogenic Methane Emissions in the U.S. from 1990–2003 (EIA, 2004)

	Methane	CO_2 Equivalent
Approximated 2003 emissions (million metric tons)	26.2	601.9
Change related to 2002 (million metric tons)	0.1	2.7
Percent change from 2002 (%)	0.5	0.5
Change related to 1990 (million metric tons)	−4.6	−105.9
Percent change from 1990 (%)	−15.0	−15.0

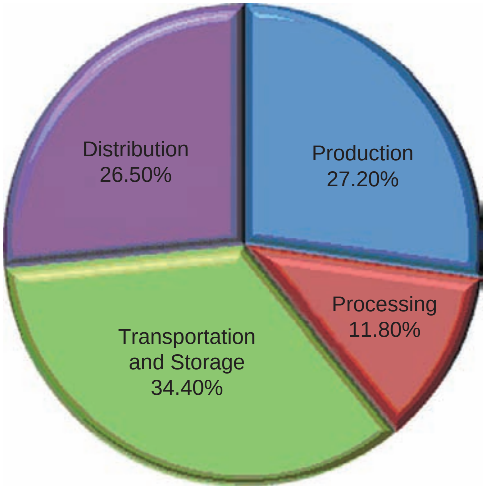

Figure 19-2 Percent of Methane Emissions in Natural Gas Industry Sectors (INGAA, 2000).

from typical procedures, regular maintenance, elusive leaks (fugitive emissions), and process disturbances. While the gas stream passes through the process, emissions may result from deliberate venting and accidental leaks. Venting could develop within construction of equipment or operational procedures—for example, the constant drain of gas from air-filled devices or venting from well terminations throughout production. Besides venting, methane emissions also result from fugitive emissions throughout the entire setup (Sakmar, 2010).

Emissions from each piece of equipment in the natural gas industry can be classified as one of three general emission types: (1) fugitive emissions, (2) vented emissions, and (3) combustion emissions. Fugitive emissions are unintentional leaks emitted from sealed surfaces, such as packings and gaskets, or leaks from underground pipelines resulting from corrosion or faulty connections. Vented emissions are releases to the atmosphere by design or operational practice. Examples of vented emissions include emissions from continuous process vents, such as dehydrator reboiler vents; maintenance practices, such as blowdowns; and small individual sources, such as gas-operated pneumatic device vents. Combustion emissions are exhaust emissions from combustion sources such as compressor engines, burners, and flares (Kirchgessner and Lott, 1997).

Figure 19-3 shows the contribution of different equipment/devices in CH_4 emissions. Pneumatic devices in the natural gas industry are valve actuators and controllers that use natural gas pressure as the force for valve movement. Gas from the valve actuator is vented during every valve stroke, and gas may bleed continuously from the valve controller pilot as well. Pneumatic devices are major sources of CH_4 emissions in both the production and transmission segments of the natural gas fuel cycle. They represent approximately 16% of total emissions. Pneumatic devices are used on heaters, separators, gas dehydrators, and gathering pipelines as liquid-level controllers, pressure regulators, and valve controllers. Some 180,000 to 250,000 pneumatic devices are currently used in the production sector and 90,000 to 130,000 in the transmission sector (INGAA, 2000).

Compressors represent one-third of CH_4 emissions from the natural gas fuel cycle. They are used throughout the natural gas system to move natural gas from production and processing sites to customer distribution systems.

Glycol dehydrator circulation pumps are a major source of methane emissions. These pumps use the high pressure of the rich glycol from the absorber to power pistons that pump the low-pressure, lean glycol from the regenerator. The pumps are an integral part of the glycol dehydrator unit, and their emissions occur through the same point. However, the pumps are the cause for nearly half of the methane emissions from

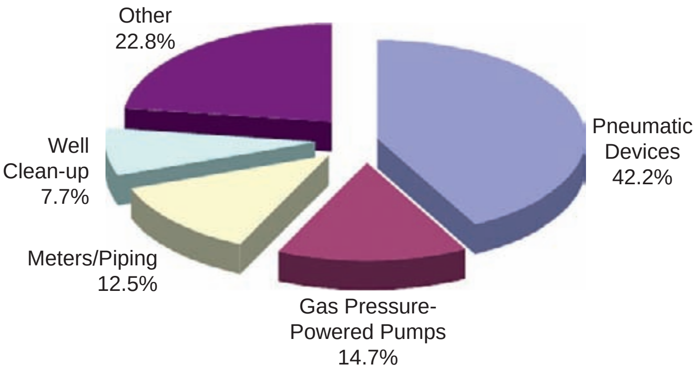

Figure 19-3 Methane emissions sources in gas production, 1998 (INGAA, 2000).

dehydrators, so they are considered separately. Unlike chemical injection pumps, which vent the driving gas directly to the atmosphere, dehydrator pumps pass the driving gas along with the rich (wet) glycol to the reboiler. Therefore, methane emissions from the pump depend on the design of the dehydrator since gas recovery on the dehydrator will also recover gas from the pump. Glycol dehydrator vents are another major source of methane emissions. The majority of the glycol dehydrators are located in production, but dehydrators are also used in gas processing, transmission, and storage. Methane emissions are highest in the production segment; 71% of the total dehydrator vent emissions are attributed to dehydrators in the production segment (Kirchgessner and Lott, 1997). This is due to the high activity and emission factors for this segment. The absence of flash tanks in most production dehydrators leads to an emission rate per volume of gas dehydrated that is higher in production than in the other segments.

Producers of methane have keen commercial inducements to minimize fugitive emissions as much as possible in order to maximize its distribution to the market. Consequently, they depend on best management practices (BMPs), such as state-of-the art equipment, assessment and maintenance programs, and infrared cameras to decrease any possible waste of energy (U.S. DOE, 2009).

19.2.4 Emissions from Boilers and Power Generation Facilities

The three major GHGs pertinent to the natural gas industry are methane, carbon dioxide, and nitrous oxide (N_2O) (EPA, 2010). Power generation is one of the chief sources of CO_2 emissions and is responsible for approximately one-third of CO_2 emissions from fossil fuel consumption, primarily from coal and natural gas (Davison, 2006).

Emissions from sources in the natural gas industry can develop from one of two kinds: combustion-related emissions and equipment leaks or vented emissions (EPA, 2010).

Combustion-related emissions occur from the combustion of natural gas in machinery such as boilers, engines, and other power generation facilities. CO_2 is the major GHG associated with combustion emissions; nevertheless, since combustion equipment is not 100% efficient, methane and additional uncombusted hydrocarbons are also released. N_2O may also be emitted from the nitrogen in natural gas or from nitrogen in ambient air (EPA, 2010). CO_2 and N_2O combustion-related emissions from natural gas are much lower than those of other fossil fuels (EIA, 1999).

GHGs may also result from equipment leaks or vented emissions. As previously discussed, there are both intentional and nonintentional emissions. Intentional emissions involve venting of gases during operational procedures, whereas nonintentional emissions involve leaks of gases from system upsets.

19.2.5 Greenhouse Gases and Climate Change

The earth's exterior temperature is retained at a livable degree by means of particular atmospheric gases, greenhouse gases that aid in trapping the sun's heat near the earth's surface. The majority of GHGs arise naturally; however, intensities of CO_2 and other GHGs have been rising ever since the Industrial Revolution, in which there was an increase in combustion of fossil fuels as well as an increase in agricultural procedures. This led to the apprehension that if this rise was left unrestricted, the final outcome would be that further heat would be trapped, causing a negative change in the earth's climate. As a result, governments globally are developing ways of reducing the anthropogenic emissions of GHGs (EIA, 1999).

The three most potent GHGs are CO_2, CH_4, and N_2O. It is safe to say that water vapor alone is the most intense greenhouse gas in the atmosphere. One of the main GHGs is CO_2, and while it does not trap heat as efficiently as other GHGs, the amount of CO_2 emissions into the atmosphere is large, mainly from the combustion of fossil fuels. Hence, decreasing carbon dioxide emissions can greatly decrease the greenhouse effect and global warming. Natural gas combustion releases about 30% less CO_2 than oil, and approximately 45% less CO_2 than coal. The United States is the primary CO_2 producer globally (Speight, 2007).

After CO_2, methane is the head producer of anthropogenic GHG releases and is responsible for approximately 16% of the total on a CO_2 equivalent basis (Sakmar, 2010). It also has the capability of trapping heat in the atmosphere about 21 times more than CO_2 and is responsible for approximately 10% of the greenhouse effect of U.S. emissions (Speight, 2007).

The large amounts of GHG emissions have contributed to the development of global warming. Global warming is the continuous increase in the average temperature of the earth's surface. Approximately two-thirds of the century's warming has developed over the past 25 years. The effect of global warming can be interpreted in the increase in ocean temperature. The oceans are playing a chief role in heat trapping. Between 1955 and 1995, the Pacific, Atlantic, and Indian Oceans revealed an approximate warming of 0.06°C (0.11°F) between the ocean surface and a depth of 3,000 m (Hill, 2004).

In addition to the increasing ocean temperature, snow and ice are melting due to global warming. Alaska, in the United States, has experienced a temperature increase of 3°C (5.4°F) over the past 30 years, which is four times larger than the average worldwide increase. This increase in temperature has caused the melting of Alaskan glaciers at what a research team from the University of Alaska describes an "incredible rate" (Hill, 2004).

Also, sea levels increased due to the effects of global warming. In the twentieth century, sea levels ascended 10 to 20 cm (4 to 8 inches) worldwide (Hill, 2004).

19.2.6 Acid Rain and Smog Formation

Acid rain and smog might be regarded as the final destination of emissions from natural gas processing and/or consumption (Speight, 2007). Acid rain is developed when nitrogen oxides and sulfur dioxide chemically react with water vapor and oxidants with the existence of sunlight to yield several acidic compounds, such as sulfuric acid and nitric acid. Precipitation in the form of rain, ice, snow, and fog results in nearly half of these atmospheric acids to descend to the ground as acid rain, whereas the other half descends in the form of dry particles and gases (EIA, 1999). Acid rain is an environmental issue that impacts various industrialized regions of the world, causing respiratory problems and poor health in humans as well as harming crops, forests, and wildlife inhabitants. Natural gas is not a major contributor to the development of acid rain since it releases about 80% less NO_x than the use of coal. Therefore, the increased use of natural gas could offer less acid rain-instigating emissions (Speight, 2007).

However, sulfur compounds released from natural gas processing facilities are among the pollutants contributing to acid precipitation, which can negatively affect ecosystems on large regional scales in highly industrialized areas or locally near certain industrial facilities. Effects of elemental sulfur dust on ground cover and shrub vegetation in forests close to sulfur blocks have been well documented (Cárcamo and Parkinson, 2001).

Smog and low air quality are crucial environmental issues, especially for large cities. The major component of smog is ground-level ozone, which is developed by the chemical reaction of CO, NO_x, VOCs, and the presence of sunlight. Smog and ground-level ozone produce a smoggy mist usually found near large cities, mainly in the summer season, and may also promote respiratory problems varying from short-term uneasiness to long-term lung damage. Pollutants responsible for the formation of smog arise from several sources consisting of vehicle emissions, smokestack emissions, paints, and solvents. Since the use of natural gas releases small amounts of NO_x and has nearly no particulate substances, it is not a major contributor to smog formation. Consequently, increased use of natural gas in industries could help in reducing the development of smog at ground level, especially in large cities (Speight, 2007).

19.2.7 Emissions from Gas Transportation and Distribution

The efficient and effective movement of natural gas from producing regions to consumption regions requires an extensive and elaborate transportation system. In many instances, natural gas produced from a particular well will have to travel a great distance to reach its point of use. Emission sources from natural gas transportation include loading and unloading of tank trucks, rail cars, and marine vessels; and transit losses from truck, marine, rail, and pipeline transportation. Pipelines are the most common mode for transporting natural gas from its source or processing plants to end users. The United States is heavily dependent on transmission pipelines to distribute energy because they are the safest mode available for transporting energy fuels. Nearly all natural gas is transported by transmission pipelines, which make up 20% of the 1.8 million total miles of pipelines in the United States (Transportation Research Board, 2004).

Transportation of energy fuels via transmission pipelines is safer than transportation via other modes, but a significant failure can result in loss of life, personal injury, property damage, and environmental damage. Natural gas transmission pipeline incidents have resulted in an annual average of 6 deaths, 10 injuries, and $20 million in property damage. There are many causes and contributors to pipeline failures, including construction errors, material defects, internal and external corrosion, operational errors, malfunctions of control systems or relief equipment, and outside force damage. Excavation and construction-related damage to pipelines remain the leading causes of pipeline failure. Such failures in 2003 were estimated to contribute 22% of hazardous liquids and 24% of natural gas transmission pipeline incidents (Transportation Research Board, 2004). Other emissions sources for this stage of the gas cycle are the internal combustion or turbine engines used to drive the compressors that push the gas through the pipeline. Product losses may occur either as fugitive equipment leaks or as evaporative losses during loading, unloading, and storage operations, and they mainly incorporate CH_4 emissions. Most refined products do not contain significant amounts of CH_4. Emissions of CO_2 and significantly smaller quantities of N_2O occur in transportation and distribution due to combustion of fuels in internal combustion engines, steam boilers on marine vessels, and turbines on gas compressors. It is also possible to have smaller amounts of unburned CH_4 emissions when natural gas is used to fire the internal combustion engines or turbines.

19.3 AQUATIC IMPACTS

The major aqueous waste streams that develop from natural gas operations are produced water, process and drainage water, spills/leaks, and cooling water. Produced water is water trapped in underground formations that is brought to the surface along with oil or gas. It is by far the largest volume byproduct or aqueous waste stream associated with gas production. Its major components may include inorganic salts, heavy metals, solids, production chemicals, hydrocarbons, and naturally available radioactive substances (E&P Forum/UNEP, 1997). These components, both separately or jointly, when present either individually or collectively in high concentrations, can present a threat to aquatic life when they are discharged or to crops when the water is used for irrigation (Veil et al., 2004).

19.3.1 Produced Water

Produced water can have different potential impacts depending on where it is discharged. For example, discharges to small streams are likely to have a larger environmental impact than discharges made to the open ocean by virtue of the dilution that takes place following discharge. Regulatory agencies have recognized the potential impacts that produced water discharges can have on the environment and have prohibited discharges in most onshore or near-shore locations (Veil et al., 2004).

In the United States, produced water comprises approximately 98% of the total volume of exploration and production (E&P) waste generated by the oil and gas industry and is the largest volume waste stream generated by the oil and gas industry. According to the American Petroleum Institute (API), about 18 billion barrels (bbl) of produced water were generated by U.S. onshore operations in 1995 (API, 2000). Additional large volumes of produced water are generated at U.S. offshore wells and at thousands of wells in other countries. In 1999, an average of 210 million bbl of water was produced each day worldwide (Khatib and Verbeek, 2003). This volume represents about 77 billion bbl of produced water for the entire year (U.S. DOE, 2009).

Management of produced water is a key issue because of its sheer volume and its high handling cost. The potential environmental impacts of produced water could be substantial if not properly managed. API's 1995 study indicated that the management and disposal of E&P wastes was following a trend toward less discharge and more reuse, recycling, and reclamation (API, 2000). Nearly all produced water from conventional oil and gas operations onshore is now being injected after the advent of no discharge criteria for produced water in coastal areas. Approximately 71% of all produced water is being injected for enhanced recovery (beneficial use), while 21% is being injected for disposal, resulting in a total of 92% of all produced water generated being returned to the subsurface from whence it came. For the remaining produced water volume, 5% is either treated and discharged or beneficially used for irrigation, livestock/wildlife watering, and other uses. For the last 3% of the produced water, percolation and evaporation ponds are the identified method of disposal (API, 2000).

19.3.2 Cooling Water

Process cooling using seawater or freshwater as a cooling medium is commonplace in gas processing facilities. Once-through cooling is the process in which water is drawn from a body of water, circulated through a condenser, and released back to its source. Approximately 31% of the present U.S. generating operations use open-loop cooling. In fresh water-scarce regions such as the Arabian Peninsula, the majority of onshore gas processing facilities use once-through cooling systems utilizing seawater as the cooling medium.

Cooling water discharge is one of the most profound environmental impacts on water quality and aquatic life. Biocide (chlorine) is added to the seawater to control biofouling of the cooling system. The added chlorine reacts with bromide and other compounds in the water to produce a wide range of chemical oxidants. Some of these oxidants remain in the cooling water after it is discharged, and they pose a major environmental burden. Regrettably, reactions between the residual oxidants and natural organic matter in the water lead to formation of toxic halogenated organic compounds that have detrimental effects on the environment when they are discharged into the receiving water body. Examples of these pollutants are trihalomethanes, haloacetic acids, halophenols, and acetonitriles.

Thermal pollution is another environmental impact resulting from cooling water discharge. Cooling water from industrial effluents is discharged at higher temperatures than that of the receiving waters. The primary effects of thermal pollution are direct thermal shock, changes in dissolved oxygen, and the redistribution of organisms in the local community. Because water can absorb thermal energy with only small changes in temperature, most aquatic organisms have developed enzyme systems that operate in only narrow ranges of temperature. These stenothermic organisms can be killed by sudden temperature changes that are beyond the tolerance limits of their metabolic systems.

19.3.3 Other Wastewater Streams

Other wastewaters routinely generated at gas processing facilities include process wastewater drainage, sewage waters, tank bottom water, fire water, equipment and vehicle wash waters, and general oily water.

Another contamination of water bodies from gas processing facilities may result from the discharge of potentially harmful substances such as hydrocarbons and process chemicals by means of spills or leaks. Similar to oil, gas enters the environment due to both natural and anthropogenic processes. It should be noted that hydrocarbon gases are piped over great distances totaling many thousands of kilometers. These pipelines cross hundreds of water bodies. Possible pipeline damages can lead to hazardous impacts on water ecosystems. The negative fisheries consequences in such cases may go beyond the limits of local scale. Regional problems can emerge if, for example, an accidental gas blowout or leakage blocks the spawning migration of anadromous fish.

Water toxicology of saturated aliphatic hydrocarbons of the methane series has not been developed thus far. However, the toxicity data on different gaseous poisons can help to reveal some general features of interaction between gaseous traces and marine organisms (Patin, 1999). The first important feature is the quick fish response to a toxic gas as compared with fish response to other dissolved or suspended toxicants. Gas rapidly penetrates into the organism (especially through the gills) and disturbs the main functional systems (respiration, nervous system, blood formation, enzyme activity, and others). External evidence of these disturbances includes a number of common symptoms mainly of behavioral nature (e.g., fish excitement, increased activity, scattering in the water).

Another critical environmental factor that directly influences the gas impact on water organisms is the concentration of dissolved oxygen. Numerous studies showed that the oxygen deficit directly controls the rate of fish metabolism and decreases their resistance to many organic and inorganic poisons.

19.4 TERRESTRIAL AND ECOLOGICAL IMPACTS

Gas production and processing sites typically contain intact expanses of natural habitat (such as grassland, forest, or desert), in which the plants are constructed and operated. Through the operation of such sites, releases of gas products and other substances such as brine periodically occur. Often, these releases impact only terrestrial resources, particularly in areas with limited numbers of wetlands or streams. The potential for adverse effects on environmental resources strongly depends on the setting of the area and its existing land and water uses where a gas production and/or processing plant is constructed. For example, industrial areas will have few if any ecological resources or unique habitats that could be affected by construction or operation of gas plants, and no change in existing land use or potential for future land use would be expected. In contrast, recreational areas may be expected to support considerable ecological resources that could be affected by gas processing. Similarly, areas with predominantly recreational water use will likely also contain quality ecological resources and may thus incur greater impacts than surface water areas in industrial or large urban settings.

Ecological resources may be affected during normal activities of all phases of gas production and processing. During normal operations, potential effects could be associated with seismic exploration; the drilling of exploration and production wells; the construction of the drill pads, pipelines, access roads, and utility corridors; and noise during operations. Biota and their habitats may also be affected by the accidental release or spill of oil or gas, as well as by the accidental release of drilling, production, and processing wastes. The magnitude of any adverse effects will depend on a number of factors, including the size and location of the gas facility and on the magnitude and duration of any gas or waste product release that might occur. Both onshore and offshore operations have the potential to affect aquatic and terrestrial biota and habitats (U.S. Army Corps of Engineers, 2005).

Exposure to an accidental natural gas leak may result in asphyxiation as a result of oxygen displacement, and the greatest threats from a natural gas leak are explosion and fire. In contrast, the greatest concern from an accidental release of H_2S is asphyxiation (also as a result of oxygen displacement) and, to a lesser degree, toxic effects from inhalation. Currently, the Occupational Safety and Health Administration (OSHA) has established an acceptable ceiling concentration of 20 parts per million (ppm) for H_2S in the workplace, with a maximum level of 50 ppm allowed for 10 minutes maximum duration if no other measurable exposure occurs. The National Institute for

Occupational Safety and Health (NIOSH) has set a maximum recommended exposure limit ceiling value of 10 ppm for 10 minutes maximum duration (ATSDR, 2006). Concentrations in ambient air from natural sources have been reported to range between 0.11 and 0.33 parts per billion (ppb), with no visible adverse effects on indigenous biota at concentrations of 3.9 ppm, while concentrations at landfills and sewage treatment plants have been reported to reach peaks of 100 ppm (ATSDR, 2006). Exposure to low concentrations of H_2S may cause irritation of the eyes, nose, or throat, as well as difficulty in breathing in asthmatic individuals (ATSDR, 2006). Brief exposures to high concentrations (greater than 500 ppm) can cause respiratory irritation, fluid buildup in the lungs, convulsions, loss of consciousness, and possibly death. Similar effects (to similar exposure levels) have been observed in laboratory animals (ATSDR, 2006) and may be expected for wildlife exposed at similar levels.

While many individuals have been reported to recover from high exposures with no apparent lasting effects, others may experience long-term effects such as headaches, poor attention span, reduced memory, and poor motor function (ATSDR, 2006). Wildlife may be expected to exhibit similar variability in recovery from H_2S exposure (U.S. Army Corps of Engineers, 2005).

Impacts on wetlands from onshore development would be largely associated with the damage or loss of vegetation and habitat from seismic exploration, drilling, and construction activities. Offshore drilling to access gas resources may impact coastal wetlands through the construction of offshore pipeline landfills, onshore pipelines, storage and processing facilities, and access roads and utility corridors (Shires and Loughran, 2004). For both onshore and offshore sites, impacts from normal operation could be associated with maintenance activities or decreases in air or water quality. Wetlands could be affected during any of the phases of developing a gas site by the accidental spill or release of gas during exploratory drilling, well completion, production, and processing, as well as by accidental releases of drilling, production, and processing wastes (U.S. Army Corps of Engineers, 2005).

Interactions between habitats and pollutants can lead to different effects on invertebrates. Forest types may vary in their ability to buffer pollutants; for example, in a study of forest health in western Canada, Maynard and Wilcox (1997) determined that aspen forests were more susceptible to sulfur contamination than pine forests. A regional survey of invertebrates around sour gas plants showed that a number of carabids with reduced populations in polluted pine sites had large populations in an aspen site that was also polluted (Cárcamo and Parkinson, 2001). Tree species are likely to interact with pollutants and differentially affect communities of soil biota. Alterations to the microhabitat and microclimate may result from the loss of vegetational cover at sites under high sulfur pollution (Maynard and Wilcox, 1997). Microclimatic factors such as moisture influence the distribution of invertebrates. Furthermore, lack of plant ground cover reduces the impedance to arthropod mobility and may increase the numbers taken in pitfall traps. This confounding factor needs to be taken into account in studies of forest floor invertebrates based on pitfall trapping. The preceding discussion described potential impacts that may occur to various habitats from gas production and processing.

Plant and animal species may also be indirectly affected by changes in their environment through variation in water, air, and soil/sediment quality and through disturbance by noise, extraneous light, and changes in vegetation cover. Such changes may directly affect the ecology: for example, habitat, food and nutrient supplies, breeding areas, migration routes, vulnerability, or predators or changes in herbivore grazing patterns, which may then have a secondary effect on predators. Soil disturbance, removal of vegetation, erosion, and siltation may have an impact on ecological integrity and may effect nutrient balances and microbial activity in the soil. If not properly controlled, a potential long-term effect is loss of habitat, which affects both fauna and flora, and may include changes in species composition and primary production cycles (UNEP, 1997).

Direct removal of marine habitat would occur due to trenching and blasting associated with the laying of gas pipelines, construction of piled jetty, and the dredging of shipping channels. The risk of marine pest species introductions discharged in ships' ballast water or from hull fouling organisms would increase significantly because of the greater intensity of ship, barge, dredge, and work boat activities associated with the project. Potential disruption to the nesting and breeding patterns for turtles and migration paths for whales and other marine mammals may occur due to noise and disturbance associated with trenching, blasting, and dredging activities during the construction phase.

Light spill during construction activities and from the plant, flare, jetty, and loadout facilities during the operational phase could also pose a significant threat to the nesting and breeding patterns of turtles. The health, abundance, and diversity of benthic marine communities may be impacted by turbidity plumes and smothering of marine organisms with sediment, linked to dredging, dredge spoil disposal and resuspension, reclamation, shipping activities, and runoff and erosion from cleared areas of the island.

Contamination of marine life may occur due to routine discharges, leaching of antifouling paints, accidental spills, or shipping accidents. Thermal loads (cooling water discharge) to marine environment should be discouraged, particularly in summer and early autumn when the corals are near or at the limits of their thermal tolerance and approaching the peak of their reproductive cycle. The presence of a large construction workforce could place increased pressure on the marine and coastal resources of the island, particularly the turtle populations; the intertidal communities; and the territorial demersal fish resources.

Natural gas exhibits negligible solubility in water and thus has little affect on water quality in the event of an underwater leak. For terrestrial releases, exposure to an accidental natural gas leak may result in asphyxiation as a result of oxygen displacement, and the greatest threats from a natural gas leak are explosion and fire. In contrast, the greatest concern from an accidental release of H_2S is asphyxiation (also as a result of oxygen displacement) and, to a lesser degree, toxic effects from inhalation. Currently, the Occupational Safety and Health Administration (OSHA) has established an acceptable ceiling concentration of 20 parts per million (ppm) for H_2S in the workplace, with a maximum level of 50 ppm allowed for 10 minutes maximum duration if no other measurable exposure occurs. The National Institute for Occupational Safety and Health (NIOSH) has set a maximum recommended exposure limit ceiling value of 10 ppm for 10 minutes maximum duration (ATSDR, 2006).

Concentrations in ambient air from natural sources have been reported to range between 0.11 and 0.33 parts per billion (ppb), with no visible adverse effects on indigenous biota at concentrations of 3.9 ppm, while concentrations at landfills and sewage treatment plants have been reported to reach peaks of 100 ppm (ATSDR, 2006).

Other aqueous waste streams such as leakage and discharge of drainage waters may result in pollution of ground and surface waters utilized for household purposes or where fisheries or ecologically important areas are affected.

19.5 NOISE IMPACT

The main sources of noise in natural gas processing facilities include large rotating machines (e.g., compressors, turbines, pumps, electric motors, air coolers, and fired heaters, air coolers at liquefaction facilities, vaporizers used during regasification, and general loading/unloading operations of natural gas carriers/vessels). During emergency depressurization, high noise levels can be generated due to release of high-pressure gases to flare and/or steam release into the atmosphere. Atmospheric conditions that may affect noise levels include humidity, wind direction, and wind speed. Vegetation, such as trees, and walls can reduce noise levels. Installation of acoustic insulating barriers can be implemented where necessary (IFC, 2007).

Noise prevention and mitigation measures should be applied where predicted or measured noise impacts from a project facility or operations exceed the applicable noise level guideline at the most sensitive point of reception. "Guidelines for Community Noise," by the World Health Organization (1999), indicated that noise impacts should not exceed the levels presented in Table 19-3, or result in a maximum increase in background levels of 3 dB at the nearest receptor location off-site (IFC, 2007).

Noise monitoring may be carried out for the purposes of establishing the existing ambient noise levels in the area of the proposed or existing facility, or for verifying operational phase noise levels. Typical monitoring periods should be sufficient for statistical analysis and may last 48 hours with the use of noise monitors that should be capable of logging data continuously over this time period, or hourly or more frequently, as appropriate (or else cover differing time periods within several days, including weekday and weekend workdays). In general, the noise level limit is represented by the background or ambient noise levels that would be present in the absence of the facility or noise source(s) under investigation. The preferred method for controlling noise from stationary sources is to implement noise control measures at the source. Methods for prevention and control of sources of noise emissions depend on the source and proximity of receptors. Noise reduction options that should be considered include (1) selecting equipment with lower sound power levels, (2) installing silencers for fans, (3) installing suitable mufflers on engine exhausts and compressor components, (4) installing acoustic enclosures for equipment casing radiating noise, (5) improving

Table 19-3 Noise Level Guidelines (IFC, 2007)

	One Hour Equivalent Continuous Sound Level (dBA)	
Receptor	Daytime 07:00–22:00	Nighttime 22:00–07:00
Residential, institutional, educational	55	45
Industrial, commercial	70	70

the acoustic performance of constructed buildings, (6) applying sound insulation, (7) installing acoustic barriers without gaps, (8) installing vibration isolation for mechanical equipment, (9) limiting the hours of operation for specific pieces of equipment or operations, especially mobile sources operating through community areas, (10) relocating noise sources to less-sensitive areas to take advantage of distance and shielding, (11) siting permanent facilities away from community areas if possible, (12) taking advantage of the natural topography as a noise buffer during facility design, (13) reducing project traffic routing through community areas wherever possible, (14) planning flight routes, timing and altitude for aircraft (airplane and helicopter) flying over community areas, and (15) developing a mechanism to record and respond to complaints.

19.6 OTHER ENVIRONMENTAL IMPACTS OF NATURAL GAS PROCESSING

19.6.1 Emissions Results from Gas Constituents

Natural gas contains mercury and BTEX, and emissions of these compounds are possible. Although mercury in natural gas is normally present at low levels, some gases contain sufficiently high mercury concentrations to cause health concerns. The environmental impacts of mercury and its compounds are direct harm to individuals from inhalation of mercury vapors or compounds and disposal of contaminated materials and equipment. Ingested mercury, particularly dimethyl mercury, permanently causes brain damage and kidney failure. Mercury accumulates in the tissue of both the fish and shellfish that are consumed by humans. Elemental mercury is removed from natural gas to protect processing equipment and catalysts. However, some of these processes do not remove the mercury compounds. EPA regulations do not require mercury removal in the natural gas industry (EPA, 2001). However, this situation might change in the future, especially for plants that process gases with mercury concentrations higher than normal levels.

Benzene, toluene, ethylbenzene, and xylene (BTEX[1]) cause two possible problems in gas processing. These are environmental impact and potential freezing in cryogenic units. These compounds are volatile and must be controlled to meet EPA clean air regulations. Benzene is also classified as a carcinogen under United States regulations and is considered an air toxin.

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[1] BTEX concentrates in the amines and glycols used to sweeten and dehydrate the natural gas.

19.6.2 Treating Emissions

Gas dehydration with glycol solutions results in BTEX emissions. The dehydration agent, ethylene glycol, is toxic to humans and must be handled properly. Also, most solid scavengers used for gas treating are respiratory and eye irritants. Spent iron sponge is pyrophoric, and great care must be taken in the removal and disposal of reacted iron-sponge material (Veroba and Stewart, 2003). It is important to carefully follow the manufacturer's recommendations for this material in order to prevent a serious incident.

When caustic compounds and solutions are used, the safe handling and disposal of these compounds are major safety and environmental issues involved with treating the hydrocarbon liquids. At present, sulfur specifications are set on the basis of corrosion, and gasoline and diesel will be required to meet increasingly stringent levels of total sulfur in fuels. The primary motivation for sulfur reduction is to reduce NO_X by increasing the performance of a catalyst to destroy NO_X, which is poisoned by sulfur.

19.6.3 Emissions from Natural Gas Compression

Exhaust emissions from compressor power drivers used for natural gas compression are a major environmental concern. The use of "clean-burn" or "lean burn" units in internal-combustion engines that drive reciprocating compressors can meet low emission levels of NO_X and CO. However, catalytic converters, which reduce NO_X to N_2 and oxidize CO to CO_2, are being installed on these units to meet tighter emissions control requirements. CO emission from gas turbines can be minimized by using high air-to-fuel ratios combined with good combustor designs. Unfortunately, high-efficiency systems enhance nitrogen oxides production. NO_X are the most significant exhaust pollutant. Water is often added to reduce NO_X production, but this helps reduce the NO_X formed from N_2 and not that formed from organic nitrogen, which might be in the fuel. Water may actually enhance NO_X production from the organic nitrogen, as well as increase CO production. Modification of the air-fuel ratio and combustor design helps keep the NO_X exit concentration level in the 10–15 ppm range (GPSA, 2004).

19.7 REGULATORY FRAMEWORK

Regulatory control and enforcement is the responsibility of national authorities, and international requirements are implemented by national authorities through primary legislation. Therefore, global and regional treaties and conventions are binding in the first instance on national governments, which are obliged to implement such arrangements through national legislation. Major international conventions include climate change and biodiversity conventions. Other important international instruments include the Montreal Protocol, which is aimed at the phaseout of ozone-depleting substances, and the Basel Convention, which addresses transfrontier movement of hazardous wastes. Also, a number of conventions have been adopted on the protection of migration and endangered species and several conventions concerning the marine environment. At the regional level, there are several agreements or conventions, such as Barcelona, OSPAR, Kuwait, and others, which form the basis of a regional regulatory framework. Oil and gas laws rarely impose detailed requirements for environmental control programs. However, they do provide the framework for subordinate regulations incorporating specific requirements such as preparation of environmental impact assessment, plans for waste disposal and control of emissions and discharges, preparation of emergency plans, control of hazardous substances, and reclamation and rehabilitation of sites at completion of operations and following accidents. The traditional approach of prescriptive legislation is gradually being complemented by performance assessment, goal setting, negotiated agreements, and self-regulation.

Where a country is party to international conventions and environmental treaties, further obligations may arise. In some countries, sophisticated mechanisms exist with single source agencies that act as a focal point for environmental control. The Environmental Protection Agency is the primary source for federal environmental laws and regulations in the United States. In addition, each of the states has its own laws and regulations that are pertinent to the state environment.

The development and production of oil and gas in the United States are regulated under a complex set of federal, state, and local laws that address every aspect of exploration and operation. The U.S. EPA administers most of the federal laws, although development on federally owned land is managed primarily by the Bureau of Land Management (BLM), which is part of the Department of the Interior, and the U.S. Forest Service, which is part of the Department of Agriculture. In addition, each state in which oil and gas are produced has one or more regulatory agencies that permit wells, including their design, location, spacing, operation, and abandonment, as well as environmental activities and discharges, including water management and disposal, waste management and disposal, air emissions, underground injection, wildlife impacts, surface disturbance, and worker health and safety.

Many of the federal laws are implemented by the states under agreements and plans approved by the appropriate federal agencies. The Ground Water Protection Council (Oklahoma) discussed these laws very well in a report that was prepared for the U.S. Department of Energy in 2009 (U.S. DOE, 2009). Those laws and their delegation are discussed in the following sections.

A series of federal laws governs most environmental aspects of gas processing. For example, the Clean Water Act (CWA) regulates surface discharges of water associated with gas drilling and production, as well as storm water runoff from production sites. The Safe Drinking Water Act (SDWA) regulates the underground injection of fluids from gas activities. The Clean Air Act (CAA) limits air emissions from engines, gas processing equipment, and other sources associated with drilling and production. The National Environmental Policy Act (NEPA) requires that exploration and production on federal lands be thoroughly analyzed for environmental impacts.

19.7.1 Clean Water Act

The Clean Water Act (CWA) is the primary federal law in the U.S. governing pollution of surface water. It was established to protect water quality and includes regulation of pollutant limits on the discharge of oil- and gas-related produced water. This is conducted through the National Pollutant Discharge Elimination System (NPDES) permitting process. Although the EPA sets national standards at the federal level, states and tribal governments can acquire primacy for the NPDES program by meeting EPA's primacy requirements.

The CWA establishes the basic structure for regulating discharges of pollutants into the waters of the United States and quality standards for surface waters. The basis of the CWA was enacted in 1948 and was called the Federal Water Pollution Control Act; the act was significantly reorganized and expanded in 1972. "Clean Water Act" became its common name, with additional amendments made in 1977 and later. Under the CWA, the EPA has implemented pollution control programs such as setting wastewater standards for industry. It has also set water quality standards for a variety of contaminants in surface waters.

The CWA made it unlawful to discharge any pollutant from a point source into the navigable waters of the United States, unless done in accordance with a specific approved permit. The NPDES permit program controls discharges from point sources that are discrete conveyances, such as pipes or manmade ditches. Industrial, municipal, and other facilities such as natural gas production sites or commercial facilities that handle the disposal or treatment of gas-produced water must obtain permits if they intend to discharge directly into surface waters (EPA, 2008a).

Most oil and gas production facilities with related discharges are authorized under general permits because there are typically numerous sites with common discharges in a geographic area. A state that meets the federal primacy requirements is allowed to set more stringent state-specific standards for this program. Since individual states can acquire primacy over their respective programs, it is not uncommon to have varying requirements from state to state. This variation can affect how the oil and gas industry manages produced water within a drainage basin located within two or more states. When developing effluent limitations for an NPDES permit, a permit writer must consider limits based on both the technology available to control the pollutants (i.e., technology-based effluent standards) and the regulations that protect the water quality standards of the receiving water (i.e., water quality-based effluent standards). The intent of technology-based effluent limits in NPDES permits is to require treatment of effluent concentrations to less than a maximum allowable standard for point source discharges to the specific surface water body. This is based on available treatment technologies, while allowing the discharger to use any available control technique to meet the limits. For industrial and other nonmunicipal facilities, technology-based effluent limits are derived by (1) using national effluent limitations guidelines and standards established by the EPA, or (2) using best professional judgment (BPJ) on a case-by-case basis in the absence of national guidelines and standards.

Prior to the granting of a permit, the authorizing agency must consider the potential impact of every proposed surface water discharge on the quality of the receiving water, not just individual discharges. If the authorizing agency determines that technology-based effluent limits are not sufficient to ensure that water quality standards will be attained in the receiving water, the CWA and NPDES regulations [40 Code of Federal Regulations (CFR) 122.44(d)] require that more stringent limits be imposed as part of the permit (EPA, 2008b). The EPA establishes effluent limitation guidelines (ELGs) and standards for different nonmunicipal (i.e., industrial) categories. These guidelines are developed based on the degree of pollutant reduction attainable by an industrial category through the application of pollution control technologies.

The CWA requires the EPA to develop specific effluent guidelines that represent the following:

1. Best conventional technology (BCT) for control of conventional pollutants and applicable to existing dischargers.

2. Best practicable technology (BPT) currently available for control of conventional, toxic, and nonconventional pollutants and applicable to existing dischargers.
3. Best available technology (BAT) economically achievable for control of toxic and nonconventional pollutants and applicable to existing dischargers.
4. New source performance standards (NSPS) for conventional pollutants and applicable to new sources.

To date, the EPA has established guidelines and standards for more than 50 different industrial categories (EPA, 2008c). The ELGs for Oil and Gas Extraction, which were published in 1979, can be found at 40 CFR Part 435. The CWA also includes a program to control storm water discharges. The 1987 Water Quality Act (WQA) added Section 402(p) to the CWA requiring the EPA to develop and implement a storm water permitting program. The EPA developed this program in two phases (Phase I: 1990; Phase II: 1999). Those regulations establish NPDES permit requirements for municipal, industrial, and construction site storm water runoff. The WQA also added Section 402(l)(2) to the CWA specifying that the EPA and states shall not require NPDES permits for uncontaminated storm water discharges from oil and gas exploration, production, processing or treatment operations, or transmission facilities. This exemption applies where the runoff is not contaminated by contact with raw materials or wastes. The EPA had previously interpreted the 402(l)(2) exemption as not applying to construction activities of oil and gas development, such as building roads and pads (i.e., an NPDES permit was required) (EPA, 2008d).

The Energy Policy Act of 2005 modified the CWA Section 402(l)(2) exemption by defining the excluded oil and gas sector operations to include all oil and gas field activities and operations, including those necessary to prepare a site for drilling and for the movement and placement of drilling equipment. The EPA promulgated a rule that implemented this exemption. However, on May 23, 2008, the U.S. Court of Appeals for the Ninth Circuit released a decision vacating the permitting exemption for discharges of sediment from oil and gas construction activities that contribute to violations of the CWA. The court based its decision on the fact that the new rule exempted runoff contaminated with sediment, while the CWA does not exempt such runoff. As a result of the court's decision, storm water discharges contaminated with sediment resulting in a water quality violation require permit coverage under the NPDES storm water permitting program. While the EPA storm water permitting rule contains a broad exclusion for oil and gas sector construction activities, it is important to note that individual states may still regulate storm water associated with these activities. The EPA has clarified its position that states and tribes may not regulate such storm water discharges under their CWA authority but are free to regulate under their own independent authorities.

In addition to state and tribal regulation, the industry has a voluntary program of Reasonable and Prudent Practices for Stabilization (RAPPS) of oil and gas construction sites (IPAA, 2004). Producers use RAPPS in order to control erosion and sedimentation associated with storm water runoff from areas disturbed by clearing, grading, and excavating activities related to site preparation.

19.7.2 Safe Drinking Water Act

The Safe Drinking Water Act (SDWA) was originally passed by Congress in 1974 to protect public health by regulating the nation's public drinking water supply. The law was amended in 1986 and 1996 and requires many actions to protect drinking water and its sources, including rivers, lakes, reservoirs, springs, and groundwater wells. SDWA authorizes the U.S. EPA to set national health-based standards for drinking water to protect against both naturally occurring and manmade contaminants that may be found in drinking water. The EPA, states, and municipal water system agencies then work together to make sure that these standards are met (EPA, 2004).

As one aspect of the protection of drinking water supplies, the SDWA establishes a framework for the Underground Injection Control (UIC) program to prevent the injection of liquid wastes into underground sources of drinking water (USDWs). The EPA and states implement the UIC program, which sets standards for safe waste injection practices and bans certain types of injection altogether. The UIC program provides these safeguards so that injection wells do not endanger USDWs. The first federal UIC regulations were issued in 1980. The EPA currently groups underground injection wells into five classes for regulatory control purposes and has a sixth class under consideration. Each class includes wells with similar functions, construction, and operating features so that technical requirements can be applied consistently to the class.

1. Class I wells may inject hazardous and nonhazardous fluids (industrial and municipal wastes) into isolated formations beneath the lowermost USDW. Because they may inject hazardous waste, Class I wells are the most strictly regulated and are further regulated under the Resource Conservation and Recovery Act (RCRA).

2. Class II wells may inject brines and other fluids associated with oil and gas production.
3. Class III wells may inject fluids associated with solution mining of minerals.
4. Class IV wells may inject hazardous or radioactive wastes into or above a USDW and are banned unless specifically authorized under other statutes for groundwater remediation.
5. Class V includes all underground injection not included in Classes I–IV. Generally, most Class V wells inject nonhazardous fluids into or above a USDW and are on-site disposal systems, such as floor and sink drains that discharge to dry wells, septic systems, leach fields, and drainage wells. Injection practices or wells that are not covered by the UIC program include single-family septic systems and cesspools as well as nonresidential septic systems and cesspools serving fewer than 20 persons that inject ONLY sanitary wastewater.
6. Class VI has been proposed specifically for the injection of CO_2 for the purpose of sequestration but has not yet been established.

Most injection wells associated with oil and gas production are Class II wells. These wells may be used to inject water and other fluids (e.g., liquid CO_2) into oil- and gas-bearing zones to enhance recovery, or they may be used to dispose of produced water. The regulation specifically prevents the disposal of waste fluids into USDWs by limiting injection only to formations that are not underground sources of drinking water. The EPA's UIC program is designed to prevent contamination of water supplies by setting minimum requirements for state UIC programs. The basic premise of the UIC program is to prevent contamination of USDWs by keeping injected fluids within the intended injection zone. The injected fluids must not endanger, or have the potential to endanger, a current or future public water supply. The UIC requirements that affect the siting, construction, operation, maintenance, monitoring, testing, and, finally, closure of injection wells have been established to address these concepts. All injection wells require authorization under general rules or specific permits.

19.7.3 Oil Pollution Act of 1990: Spill Prevention Control and Countermeasure

The CWA and the Oil Pollution Act (OPA) include both regulatory and liability provisions that are designed to reduce damage to natural resources from oil spills. The U.S. Congress added Section 311 to the CWA, which in part authorized the president to issue regulations establishing procedures, methods, equipment, and other requirements to prevent discharges of oil from vessels and facilities [Section 311(j)(1)(c)]. In response to the Exxon Valdez oil spill in Alaska, Congress enacted the OPA in 1990 (EPA, 2008e). The OPA amended CWA Section 311 and contains provisions applicable to onshore facilities and operations. Section 311, as amended by the OPA, provides for spill prevention requirements, spill reporting obligations, and spill response planning. It regulates the prevention of and response to accidental releases of oil and hazardous substances into navigable waters, on adjoining shorelines, or affecting natural resources belonging to or managed by the United States. This authority is primarily carried out through the creation and implementation of facility and response plans. These plans are intended to establish measures that will prevent discharge of oil into navigable waters of the United States or adjoining shorelines as opposed to response and cleanup after a spill occurs.

A cornerstone of the strategy to prevent oil spills from reaching the nation's waters is the oil Spill Prevention, Control, and Countermeasure (SPCC) plan. The EPA promulgated regulations to implement this part of the OPA of 1990. These regulations specify that

1. SPCC plans must be prepared, certified (by a professional engineer), and implemented by facilities that store, process, transfer, distribute, use, drill for, produce, or refine oil.
2. Facilities must establish procedures and methods and install proper equipment to prevent an oil release.
3. Facilities must train personnel to properly respond to an oil spill by conducting drills and training sessions.
4. Facilities must also have a plan that outlines steps to contain, clean up, and mitigate any effects that an oil spill may have on waterways (EPA, 2008f).

Before a facility is subject to the SPCC rule, it must meet three criteria:

1. It must be non-transportation-related.
2. It must have an aggregate aboveground storage capacity greater than 1,320 gallons (31.4 bbls) or a completely buried storage capacity greater than 42,000 gallons (1,000 bbls).
3. There must be a reasonable expectation of a discharge into or upon navigable waters of the United States or adjoining shorelines.

19.7.4 Clean Air Act

The Clean Air Act (CAA) is the primary means by which the EPA regulates potential emissions that could affect air quality. The U.S. Congress passed the CAA in 1963 and has amended it on several occasions since, most recently in 1990. The CAA requires the EPA to set national standards to limit levels of certain pollutants. The EPA regulates those pollutants by developing human health-based and/or environmentally and scientifically based criteria for setting permissible levels. Air regulations do not normally include exceptions for a company's size, the age of a field, or the type of operation. As a result of the implementation of the CAA, air quality has improved dramatically across the United States during the last few decades, and existing regulations should continue to reduce air pollution emissions during the next 20 years or longer (EPA, 2008g).

19.7.4.1 Air Quality Regulations

Like any other U.S. industry, natural gas producers must comply with existing and new air regulations including those resulting from the 1990 CAA amendments. These rules pose an ongoing challenge to company resources as producers strive to understand and comply with enforcement, fines, public reaction, and possibly even project cancellations in light of new standards. The EPA has established National Emission Standards for Hazardous Air Pollutants (NESHAPs), which are nationally uniform standards to control specific air emissions. In 2007, the EPA implemented a new standard referred to as the Maximum Achievable Control Technology (MACT) standard for hazardous air pollutants (HAPs) that targeted small area sources such as gas operations located in areas near larger populations. These standards limit HAP emissions (primarily benzene) from process vents on glycol dehydration units, storage vessels with flash emissions, and equipment leaks.

Another example of new or amended federal regulations that will have a direct impact on controlling emissions from gas operations is the Stationary Spark Ignition Internal Combustion Engine new source performance standard and Reciprocating Internal Combustion Engine NESHAP rules, which regulate new and refurbished engines. These rules, passed in 2007, target all internal combustion engines regardless of horsepower rating, location, or fuel (electric engines are not included) and include extensive maintenance, testing, monitoring, recordkeeping, and reporting requirements (EPA, 2007).

19.7.4.2 Air Permits

Air permits are legal documents that facility owners and operators must abide by. The permit specifies what construction is allowed; what emission limits must be met; how the emissions source(s) must be operated; and what conditions specifying monitoring, recordkeeping, and reporting requirements must be maintained to assure ongoing compliance. Shale gas producers may need air quality permits for a number of emissions sources, including gas compressor engines, glycol dehydrators, and flares. A company's permitting responsibility does not end with the issuance of its initial air permit. The company must be constantly vigilant that a new regulation, modification, replacement, or process change does not impact its existing permit and require a permit amendment or a more stringent permit. Although these permits may differ across the country, they all contain specific conditions designed to ensure state and federal standards are met and to prevent any significant degradation in air quality as a result of a proposed activity.

19.7.4.3 Regulation of Impacts to Land

Impacts to land from gas operations include solid waste disposal and surface disturbances that may impact the visual landscape or may affect wildlife habitat. Operations on federal lands are a special case with unique requirements that are discussed in the following sections.

19.7.5 Resource Conservation and Recovery Act

The Resource Conservation and Recovery Act (RCRA) was passed in 1976 to address the growing problems of the increasing volume of municipal and industrial waste. The RCRA established goals for protecting human health and the environment, conserving resources, and reducing the amount of waste. RCRA Subtitle C established a federal program to manage hazardous wastes from cradle to grave to ensure that hazardous waste is handled in a manner that protects human health and the environment. Subtitle D of the RCRA addresses nonhazardous solid wastes, including certain hazardous wastes that are exempted from the Subtitle C regulations (EPA, 2008h).

In 1978, the EPA proposed hazardous waste management standards that included reduced requirements for some industries, including oil and gas, with large volumes of wastes. The EPA determined that these large-volume "special wastes" were lower in toxicity than other wastes being regulated as hazardous waste under the RCRA (EPA, 2002). In 1980, the Solid Waste Disposal Act (SWDA) amended the RCRA to exempt drilling fluids, produced waters, and other wastes associated with exploration, development, and production of crude oil, natural gas, and geothermal energy. The SWDA amendments also required the EPA to provide a report to Congress on these wastes and to make a regulatory determination as to whether regulation of these wastes under RCRA Subtitle C was warranted (EPA, 1993). In 1987, the EPA issued a report to Congress that outlined the results of a study on the management, volume, and toxicity of wastes generated by the oil, natural gas, and geothermal industries. In 1988, the EPA issued a final regulatory determination stating that control of oil and gas exploration and production wastes under RCRA Subtitle C was not warranted. The EPA made this determination because it found that other state and federal programs could protect human health and the environment more effectively. In lieu of regulation under Subtitle C, the EPA implemented a three-pronged strategy to ensure that the environmental and programmatic issues were addressed:

1. Improve other federal programs under existing authorities.
2. Work with states to improve some programs.
3. Work with Congress to develop any additional statutory authorities that may be required (EPA, 1988).

These wastes have remained exempt from Subtitle C regulations, but this does not preclude these wastes from control under state regulations or other federal regulations (EPA, 2002). The exemption applies only to the federal requirements of RCRA Subtitle C. A waste that is exempt from Subtitle C regulation might be subject to more stringent or broader state hazardous and nonhazardous waste regulations and other state and federal program regulations. For example, oil and gas exploration and production wastes may be subject to regulation under RCRA Subtitle D, the Clean Air Act, the Clean Water Act, the Safe Drinking Water Act, and/or the Oil Pollution Act of 1990 (EPA, 2008i, 2008j).

In 1989, the EPA worked with the Interstate Oil and Gas Compact Commission (IOGCC), state regulatory officials, industry representatives, and nationally recognized environmental groups to establish a Council on Regulatory Needs. The purpose of the council was to review existing state oil and gas exploration and production waste management programs and to develop guidelines to describe the elements necessary for an effective state program. This effort was begun by the EPA as part of the second prong of the agency's approach. These groups then worked together with state regulatory agencies to review the state programs, on a voluntary basis, against these guidelines and to make recommendations for improvement. This state review program continues today under the guidance of a nonprofit organization called State Review of Oil & Natural Gas Environmental Regulations (STRONGER). The state programs reviewed to date represent over 90% of the onshore domestic production. Working with the IOGCC, STRONGER (2008) has continued to update the guidelines consistent with developing environmental and oilfield technologies and practices. Under the state review process, state programs have continued to improve, and follow-up reviews have shown significant improvement where states have successfully implemented the recommendations of the review committees.

19.7.6 Endangered Species Act

The Endangered Species Act (ESA) of 1973 protects plants and animals that are listed by the federal government as "endangered" or "threatened" (U.S. DOE, 2009). Sections 7 and 9 are central to regulating oil and gas activities. Section 9 makes it unlawful for anyone to "take" a listed animal, and this includes significantly modifying its habitat. This applies to private parties and private land; a landowner is not allowed to harm an endangered animal or its habitat on his or her property. Section 7 applies not to private parties, but to federal agencies. This section covers not only federal activities but also the issuance of federal permits for private activities, such as Section 404 permits issued by the Corps of Engineers, to people who want to do construction work in waters or wetlands (Albrecht and Christman, 1999). Section 7 imposes an affirmative duty on federal agencies to ensure that their actions (including permitting) are not likely to jeopardize the continued existence of a listed species (plant or animal) or result in the destruction or modification of critical habitat. Both Sections 7 and 9 allow "incidental takes" of threatened or endangered species, but only with a permit. To "take" is to harass, harm, pursue, hunt, shoot, wound, kill, trap, capture, or collect a plant or animal of any threatened or endangered species. Harm includes significant habitat modification when it kills or injures a member of a listed species through impairment of essential behavior (e.g., nesting or reproduction). For any nonfederal industrial activity, the burden is on the owner and/or operator to determine if an incidental take permit is needed. This is typically accomplished by contacting the U.S. Fish and Wildlife Service (FWS) to determine whether any listed species are present or will potentially inhabit the project site (USFWS, 2011). A biological survey may be required to determine whether protected species are present on the site and whether a Section 9 permit may be required (ALL Consulting, 2004; CICA, 2008). The FWS and many state fish and game agencies offer services to help operators determine whether a given project is likely to result in a take and whether a permit is required. FWS can also provide technical assistance to help design a project so as to avoid impacts. For example, the project could be designed to minimize disturbances during nesting or mating seasons.

A Section 9 permit must include a habitat conservation plan (HCP) consisting of an assessment of impacts; measures that will be undertaken to monitor, minimize, and mitigate any impacts; alternative actions considered and an explanation of why they were not taken; and any additional measures that the FWS may require (U.S. DOE, 2009). Mitigation measures, which are actions that reduce or address potential adverse effects of a proposed activity upon species, must be designed to address the specific needs of the species involved and be manageable and enforceable.

19.7.7 Federal Lands

The U.S. Department of Interior's Bureau of Land Management (BLM) is responsible for permitting and managing most onshore oil and gas activities on federal lands. The BLM carries out its responsibility to protect the environment throughout the process of oil and gas resource exploration and development on public lands. Resource protection is considered throughout the land use planning process when resource management plans (RMPs) are prepared and when an Application for Permit to Drill (APD) is processed (ALL Consulting, 2002). The BLM's inspection and enforcement and its monitoring program are designed to ensure that operators comply with relevant laws and regulations as well as specific stipulations set forth during the permitting process.

19.7.8 Comprehensive Environmental Response, Compensation, and Liability Act

The Comprehensive Environmental Response, Compensation, and Liability Act (CERCLA), commonly known as Superfund, was enacted by Congress on December 11, 1980. This law created a tax on the chemical and petroleum industries and provided broad federal authority to respond directly to releases or threatened releases of hazardous substances that may endanger public health or the environment. CERCLA established prohibitions and requirements concerning closed and abandoned hazardous waste sites, provided for liability of persons responsible for releases of hazardous waste at these sites, and established a trust fund to provide for cleanup when no responsible party could be identified. Over 5 years, $1.6 billion was collected and placed in a trust fund for cleaning up abandoned or uncontrolled hazardous waste sites.

CERCLA was amended by the Superfund Amendments and Reauthorization Act (SARA) in 1986. SARA made several changes to the Superfund program that augmented its effectiveness, provided new enforcement authorities, boosted state and citizen involvement, and increased the size of the trust fund. In addition to the provisions for cleaning up hazardous waste sites, CERCLA requires the person in charge of a vessel or facility to immediately notify the National Response Center when there is a release of a hazardous substance in an amount equal to or greater than the reportable quantity (RQ) for that substance [CERCLA Section 103(a)]. The reportable quantity depends on the substance released.

CERCLA Section 101(14) excludes certain substances from the definition of "hazardous substance," thus exempting them from CERCLA regulation. These substances include petroleum (meaning crude oil or any fraction thereof that is not specifically listed as a hazardous substance), natural gas, natural gas liquids, liquefied natural gas (LNG), and synthetic gas usable for fuel. If a release of one of these substances occurs, CERCLA notification is not required. Many states have separate requirements regarding hazardous substances. Reporting of releases of the materials exempted under CERCLA may be required under state law.

19.7.9 Emergency Planning and Community Right-to-Know Act

Congress enacted the Emergency Planning and Community Right-to-Know Act (EPCRA) in 1986 to establish requirements for federal, state, and local governments; tribes; and industry regarding emergency planning and "community right-to-know" reporting on hazardous and toxic chemicals. The community right-to-know provisions of EPCRA are the most relevant part of the law for shale gas producers. They help increase the public's knowledge and access to information on chemicals at individual facilities, along with their uses and potential releases into the environment. Under Sections 311 and 312 of EPCRA, facilities manufacturing, processing, or storing designated hazardous chemicals must make material safety data sheets (MSDSs), describing the properties and health effects of these chemicals, available to state and local officials and local fire departments. Facilities must also provide state and local officials and local fire departments with inventories of all on-site chemicals for which MSDSs exist. Information about chemical inventories at facilities and MSDSs must be available to the public. Facilities that store over 10,000 pounds of hazardous chemicals are subject to this requirement. Any hazardous chemicals above the threshold stored at gas production and processing sites must be reported in this manner.

Section 313 of EPCRA authorizes the EPA's Toxics Release Inventory (TRI), which is a publicly available database that contains information on toxic chemical releases and waste management activities reported annually by certain industries as well as federal facilities. The EPA issues a list of industries that must report releases for the database. To date, the EPA has not included oil and gas extraction as an industry that must report under TRI. This is not an exemption in the law. Rather, it is a decision by the EPA that this industry is not a high priority for reporting under TRI. Part of the rationale for this decision is based on the fact that most of the information required under TRI is already reported by producers to state agencies that make it publicly available. Also, TRI reporting from the hundreds of thousands of oil and gas sites would overwhelm the existing EPA reporting system and make it difficult to extract meaningful data from the massive amount of information submitted (IOGCC, 1996; IPAA, 2000).

19.7.10 Occupational Safety and Health Act

Under the Occupational Safety and Health Act of 1970, employers are responsible for providing a safe and healthy workplace for their employees. The Occupational Safety and Health Administration (OSHA) promotes the safety and health of America's working men and women by setting and enforcing standards; providing training, outreach, and education; establishing partnerships; and encouraging continual process improvement in workplace safety and health (U.S. Department of Labor, 2011). OSHA has developed specific standards to reduce potential safety and health hazards in the oil and gas drilling, servicing, and storage industries. States also have requirements that provide further worker and public safety protections.

19.8 PERFORMANCE INDICATORS AND MONITORING

19.8.1 Emissions and Effluent Guidelines

Tables 19-4 and 19-5 present process emission and effluent guidelines for the gas processing sector (IFC, 2007). Guideline values for process emissions and effluents in this sector are indicative of good international industry practice as reflected in relevant standards of countries with recognized regulatory frameworks. These guidelines are achievable under normal operating conditions in appropriately designed and operated facilities through the application of pollution prevention and control techniques. Combustion source emissions guidelines are associated with steam- and power-generation activities from sources with a capacity equal to or lower than 50 MW. Effluent guidelines are applicable for direct discharges of treated effluents to surface waters for general use. Site-specific discharge levels may be established based on the availability

Table 19-4 Air Emission Levels for Natural Gas Processing Facilities[a] (IFC, 2007)

Pollutant	Units	Guideline Value
NO_x	mg/Nm^3	150[b] 50[c]
SO_2	mg/Nm^3	75
Particulate Matter (PM10)	mg/Nm^3	10
VOC	mg/Nm^3	150
CO	mg/Nm^3	100

[a] Dry gas at 15% oxygen.
[b] 150 mg/Nm^3 NO_x is valid to facilities with a total heat input capacity of up to 300 MWth.
[c] 50 mg/Nm^3 is valid to facilities with a total heat input capacity larger than 300 MWth.

Table 19-5 Effluents Levels for Natural Gas Processing Facilities (IFC, 2007)

Pollutant	Units	Guideline Value
pH	……	6–9
BOD_5	mg/l	50
COD	mg/l	150
TSS	mg/l	50
Oil and grease	mg/l	10
Cadmium	mg/l	0.1
Total residual chlorine	mg/l	0.2
Chromium (total)	mg/l	0.5
Copper	mg/l	0.5
Iron	mg/l	3
Zinc	mg/l	1
Cyanide		
Free	mg/l	0.1
Total		1
Lead	mg/l	0.1
Nickel	mg/l	1.5
Heavy metals (total)	mg/l	5
Phenol	mg/l	0.5
Nitrogen	mg/l	40
Phosphorous	mg/l	3

and conditions in the use of publicly operated sewage collection and treatment systems or, if discharged directly to surface waters, on the receiving water use classification. These levels should be achieved, without dilution, at least 95% of the time that the plant or unit is operating, to be calculated as a proportion of annual operating hours. Deviation from these levels in consideration of specific, local project conditions should be justified in the environmental assessment.

19.8.2 Environmental Monitoring

Environmental monitoring programs for this sector should be implemented to address all activities that have been identified to have potentially significant impacts on the environment, during normal operations and upset conditions. Environmental monitoring activities should be based on direct or indirect indicators of emissions, effluents, and resource use applicable to the particular project. Monitoring frequency should be sufficient to provide representative data for the parameter being monitored. Monitoring should be conducted by trained individuals following monitoring and recordkeeping procedures and using properly calibrated and maintained equipment. Monitoring data should be analyzed and reviewed at regular intervals and compared with the operating standards so that any necessary corrective actions can be taken.

Emissions and air quality monitoring programs provide information that can be used to assess the effectiveness of emissions management strategies. A systematic planning process is recommended to ensure that the data collected are adequate for their intended purposes (and to avoid collecting unnecessary data). This process, sometimes referred to as a data quality objectives process, defines the purpose of collecting the data, the decisions to be made based on the data and the consequences of making an incorrect decision, the time and geographic boundaries, and the quality of data needed to make a correct decision. The air quality monitoring program should consider the following elements:

Monitoring parameters: The monitoring parameters selected should reflect the pollutants of concern associated with project processes. For combustion processes, indicator parameters typically include the quality of inputs, such as the sulfur content of fuel.

Baseline calculations: Before a project is developed, baseline air quality monitoring at and in the vicinity of the site should be undertaken to assess background levels of key pollutants, in order to differentiate between existing ambient conditions and project-related impacts.

Monitoring type and frequency: Data on emissions and ambient air quality generated through the monitoring program should be representative of the emissions discharged by the project over time. Examples of time-dependent variations in the manufacturing process include batch process manufacturing and seasonal process variations. Emissions from highly variable processes may need to be sampled more frequently or through composite methods. Emissions monitoring frequency and duration may also range from continuous for some combustion process operating parameters or inputs (e.g., the quality of fuel) to less frequent, monthly, quarterly, or yearly stack tests.

Monitoring locations: Ambient air quality monitoring may consists of off-site or fence line monitoring either by the project sponsor, the competent government agency, or by collaboration between both. The location of ambient air quality monitoring stations should be established based on the results of scientific methods and mathematical models to estimate potential impact to the receiving airshed from an emissions source, taking into consideration such aspects as the location of potentially affected communities and prevailing wind directions.

Sampling and analysis methods: Monitoring programs should apply national or international methods for sample collection and analysis, such as those published by the International Organization for Standardization, the European Committee for Standardization, or the U.S. Environmental Protection Agency. Sampling should be conducted by, or under, the supervision of trained individuals. Analysis should be conducted by entities permitted or certified for this purpose. Sampling and analysis quality assurance/quality control (QA/QC) plans should be applied and documented to ensure that data quality is adequate for the intended data use (e.g., method detection limits are below levels of concern). Monitoring reports should include QA/QC documentation.

19.9 ENVIRONMENTAL IMPACT ASSESSEMENT OF NATURAL GAS PRODUCTION AND PROCESSING

The environmental impact assessment (EIA) process will identify potential environmental impacts at the early stages before a project is implemented so that negative environmental impacts can be avoided or reduced to an acceptable level. The principal objective of the EIA is to predict the environmental effects, which can be expected should the project proceed. The significance of these environmental effects shall be discussed. The EIA study is also expected to identify methods of optimizing positive environmental effects and minimizing negative environmental effects resulting from the project. The impacts on environmental resources from the proposed project will vary in duration and significance. Three types of impact duration are usually considered as part of the evaluation, including short-term, long-term, and permanent impacts. Short-term impacts are temporary in nature and will occur during the construction phase of the project. Short-term impacts should diminish to cessation as soon as the construction phase of the project is completed. Long-term impacts are those impacts that will be chronic in nature due to ongoing activities of the projet. Permanent impacts are those resulting in a permanent and irreversible change to existing environmental resources as a result of the project. Impacts may be positive or negative, and will be identified as such.

It is the responsibility of the project proponent, through the EIA process, to demonstrate that

1. Best practicable measures have been taken in planning and designing the project to avoid, and where this is not possible, to minimize impacts on the environment.
2. The unavoidable impacts of the project should be found to be environmentally acceptable, taking into account cumulative impacts that have already occurred in the region and principles of sustainability.

To ensure these requirements are met, the proponent should ensure that potential environmental impacts and constraints are identified early in the planning and design of a project, and that adequate time and resources are devoted to carrying out environmental surveys and investigations as part of preparing the EIA document. The proponent is also encouraged to engage in consultation with government authorities and members of the public who may be interested or affected by the project early in the EIA process, to ensure that people are informed about the project and its impacts, and to enable the proponent to respond to issues and make appropriate adjustments to the project during planning and design.

19.9.1 Prediction of Environmental Effects

The main focus of the EIA is to predict environmental effects to the environment, which may result from the proposed undertaking and their significance. Predictions must consider all aspects and phases of the project, and any indirect, cumulative environmental effects and those effects that may result from accidents or malfunctions. These predictions should consider potential effects of the environment on the project such as by extreme weather events, lightning strikes, terrorist attacks, and seismic activity and should include climate change considerations.

EIA predictions are generally based on a combination of objective and subjective evaluation. The use of objective (measurable) analysis is strongly preferred where it is technically feasible and reasonable to do so. However, in recognition of any factor that may limit the ability to predict environmental responses, it is expected that predictions may be based on subjective evaluation using professional judgment and experience. In consideration of this, predictive statements should be accompanied by a discussion of the limitations of the analysis, references to supporting documentation, and the qualifying credentials of those making the predictions. Predictions must be made regarding the nature (negative or positive), magnitude, duration, frequency, geographic extent, and irreversibility of the project's environmental effects. The significance must also be determined. Cumulative environmental effects must be given consideration. Cumulative environmental effects should be considered individually for each valued environmental component selected.

19.9.2 Mitigation, Contingency, and Compensation

The study must describe general and specific measures that are technically and economically feasible to mitigate (eliminate, prevent, avoid, and minimize) the potentially adverse environmental effects of the project. This should include a description of contingency measures (including emergency response plans for construction and operation) that have been designed to address potential accidents and malfunctions that could result in spills or unplanned releases of contaminants or products to the environment. Specific circumstances under which mitigation measures will be implemented must be clearly defined by the proponent. Mitigation options should be considered in a hierarchical manner with a clear priority placed on impact avoidance and pollution prevention opportunities. Opportunities to contribute to a regional approach to management of cumulative effects should be emphasized.

19.9.3 Commitment to Monitoring and Follow-Up

A well-defined program of monitoring and follow-up regarding the potential environmental effects of the project must be outlined in the EIA. The proponent must describe the procedure for monitoring and follow-up programs for the project, including objectives, content, implementation, and reporting of results. Monitoring and follow-up programs should include protocols that would guide interpretation of monitoring results and timely implementation of appropriate corrective actions. The monitoring program must be based on accurate baseline information of the existing physical, biological, and socioeconomic environments. The proponent is expected to collect the necessary information through existing data sources or through primary research such as fieldwork and laboratory testing, as required. Documentation from similar operations elsewhere in the world indicating their ability to achieve standards should be provided. The standards should be included for those other facilities, in addition to the standards to which this project will be constructed, operated, and maintained.

19.9.4 Public Consultation

Public consultation is an essential component of the EIA. Consultation with persons and organizations potentially affected by the project, local governments, and other stakeholders should be done as part of the EIA process. The proponent is expected to hold appropriate public consultation events and to use various media to engage public consultation. The EIA should document the dates and formats for public consultation undertaken, the material presented to the public, the opportunity for receiving public input, a summary review of the concerns expressed by the public, and how these concerns were addressed. It should be clear how the input from consultations was used in the assessment and what changes to the process or project were made as a result of comments provided.

19.9.5 Potential Environmental Impacts

Prediction of potential environmental impacts of the project should be discussed in the EIA, and monitoring and mitigation measures should be described. Potential impacts that should be addressed include but are not limited to

- Effects on air quality
- Effects on the marine environment
- Effects on freshwater fish and fish habitat
- Effects on species of special conservation status
- Effects on terrestrial and wetland environments
- Effects on migratory birds
- Effects on groundwater resources
- Effects of navigation on safety
- Effects on the road transportation network
- Socioeconomic effects
- Effects on public health and safety
- Effects on aboriginal land and resource use
- Effects of natural hazards on the project

19.10 WASTE MINIMIZATION AND POLLUTION PREVENTION IN GAS PROCESSING OPERATIONS

Pollution prevention and energy efficiency, two of the most prominent and cost-effective best management practices, stress economic benefit while protecting the environment and conserving natural resources. Over the last 10 years, a significant shift in focus from pollution control (and waste management) to pollution prevention (and waste minimization) has occurred. The predominant factor causing this shift was the realization that pollution and waste cost money and do not result in any positive revenue stream for the industries that generate them. Today, pollution prevention, energy efficiency, and other environmental best management practices are considered to be quality programs as much as environmental programs. As quality programs, they focus on process (means) as well as product (end), and they require top management commitment. They are instituted company-wide, and they are best utilized through continuous quality improvement principles.

As a proactive approach to environmental protection, these best management practices usually require a level of planning at both the strategic and tactical levels. The major question to answer for pollution prevention and energy efficiency at the strategic level is "Why" and at the tactical level is "How." The requirement of top management commitment is critical to promote and champion the development of the strategy. At this level, issues such as investment in "clean" technologies, return on investment, and appropriate level of risk are typically addressed. Tactical issues typically include training on how to use new technologies and processes, implementation schedule, and monitoring and documenting success.

19.10.1 Source Reduction

Source reduction involves the use of processes, practices, or products to reduce or eliminate the generation of pollutants and wastes. Source reduction includes, but is not limited to, material substitution, process substitution, and process elimination. Examples of some source reduction opportunities are described in the following sections. The EPA and the natural gas industry, through the Natural Gas STAR Program, have identified several best management practices (BMPs) that are cost effective in reducing CH_4 emissions. In addition, a series of "lessons learned" studies of these BMPs provide information on the cost of achieving these emission reductions.

19.10.1.1 Equipment and Materials Selection

Materials and equipment selection is an important factor in minimizing emissions in gas processing facilities.

Numerous options exist to reduce CH_4 emissions from pneumatic devices. The primary methods are to maintain, retrofit, or replace the devices. The lowest cost (i.e., shortest payback) options include the following (INGAA, 2000):

- Plugging holes in pneumatic instruments that exhibit steady-state bleeding. This can be done at very low cost with a rapid payback.
- Replacing orifices or lowering supply pressure in steady-state bleeds from nozzles and flappers in pneumatic instruments. This, too, is an easy and cost-effective option.
- Changing controller gain settings to reduce emissions by about 50% at no cost. However, this can upset loop stability and thereby reduce the benefits.
- Retrofitting single-stage controller-fixed orifices with smaller orifices. This would reduce emissions by about 60% with a 1-month payback.
- Replacing old internal relief valves with zero-bleed, soft-seat relief designs. This would reduce emissions by 100% with a 2-month payback.
- Installing interruptible intermittent-bleed nozzles in place of continuous-bleed nozzles. This could reduce emissions by 75% with a payback of 7 months. Where intermittent-bleed nozzles cannot be used, MIZER nozzle retrofits can be made. Mizer nozzles reduce emissions by 56%–97% with a payback of 1 to 2½ years.

Other, more expensive options also are available. Experience under EPA's Natural Gas STAR Partners Program indicates that replacing high-bleed equipment in level controllers can reduce emissions by approximately 80% with a 16-month payback, and replacing high-bleed equipment in pressure controllers can reduce emissions 80% with a 2-month payback. Retrofitting high-bleed equipment also can reduce emissions by approximately 80% with a payback period of 1–16 months depending on the equipment used. Field experience in EPA's Natural Gas STAR Partners Program indicates that up to 80% of all high-bleed devices can be replaced or retrofitted with low-bleed equipment. However, where once it was thought that mitigating fugitive emissions from pneumatic devices would be easy, experience has shown it to be more difficult. Discussions with representatives from Fisher Controls indicate that many retrofit and replacement options can adversely affect the overall performance of the operating systems, adding costs and reducing performance (INGAA, 2000).

Instrument air and electric valve controllers can also be used in lieu of natural gas-powered pneumatic systems. If this is done, CH_4 emissions would be totally eliminated. Such systems rely on electricity and are significantly more expensive than using low-bleed equipment. However, at facilities that need compressed air or where safety issues are important, these may be viable options. Many options are available to reduce emissions from pneumatic devices by just improving maintenance practices such as reducing system supply pressure and/or removing unneeded positioners from diaphragm actuators. This would eliminate CH_4 emissions at no cost. However, only a fraction of positioners can be replaced without detrimental effects on the system.

Fugitive methane leakage from production and distribution systems (including venting during routine maintenance) accounts for more than one-third of overall methane emissions from the natural gas fuel cycle. The majority of CH_4 emissions arise from leaks in cast iron mains, unprotected steel mains and meters, production well cleanup, and regulators at city gates. The principal options available to reduce CH_4 emissions from these sources are (1) directed inspection and maintenance and (2) reduced pipeline pressures when demand is low.

Good housekeeping and equipment maintenance are two best management practices that are often low-cost/high-benefit approaches to pollution prevention. A common example of good housekeeping practices involves the use of drip pans to catch leaks or drips from equipment. Equipment maintenance is important because routine maintenance will reduce the occurrence of leaks and drips, and will extend the lifetime of the equipment.

Directed inspections can identify the principal sources of leakage and assess how to most cost effectively reduce them. Pinhole leaks, component flaws, loose component joints, and eroded or ineffective valve seals are the major sources of fugitive emissions from gas distribution. Directed inspection and maintenance programs aimed at these sources of leakage can reduce emissions by approximately 40% (INGAA, 2000). Payback periods are expected to be less than a year.

Materials selection is another important element in minimizing emissions. Materials that will result in less toxic wastes can be substituted for materials that are currently being used. For example, the substitution of less toxic drilling fluid additives will result in less toxic drilling wastes, and shifting from solvent-based paints to water-based paints reduces the toxicity of paint wastes.

Process substitution or elimination will also help minimize adverse environmental impacts of gas production and processing. Processes that result in less waste and increased efficiency can be substituted for processes that are currently being used. Also, entire processes can be eliminated if pollution prevention is implemented. Examples of process substitution/elimination are (1) well designs and drilling methods that reduce the volume of cuttings generated, (2) improved transportation methods that reduce the risk of spills and leaks, and (3) improved separation techniques at the well that eliminate the need for several gas processing steps.

19.10.1.2 Energy Efficiency

Renewable energies are not the only relevant factors for ensuring a sustainable energy supply worldwide. About 70% of the total worldwide primary energy used is lost throughout the energy supply chain, starting with the production and transport of energy all the way to its final consumption. There is huge energy-saving potential here. Without clearly improving energy efficiency, the aims of sustainable energy management and a successful climate protection policy cannot be reached.

Energy efficiency is the cleanest, quickest, and most cost-effective way to extend today's energy supply into the future, and it helps improve the industry's environmental performance, including reducing greenhouse gas emissions associated with climate change. The efficient use of energy is not only a core value of the oil and natural gas industry but is also a daily practice. Energy efficiency helps energy companies manage costs, which, in turn, strengthens competitiveness and brings more affordable energy products to the consumer. Through technologies such as combined heat and power (CHP), also known as cogeneration, and the reuse of excess heat from plant processes to produce additional energy, plants can become more efficient, reducing both energy use and emissions. Cogeneration is a critical efficiency option for companies operating in the oil and natural gas sector. The cogeneration process uses steam and electricity produced at facilities and can simultaneously yield efficiency rates far greater than the separate generation of electricity and steam. Between 1990 and 2005, oil and natural gas industry cogeneration capacity grew ninefold. In fact, in 2006 alone, energy savings produced from efficiency tactics were equivalent to the gasoline consumed by more than 530,000 cars or the electricity used by 950,000 homes (http://energytomorrow.org/environment-and-safety/energy-efficiency).

19.10.2 Recycling, Treatment, and Disposal Options

19.10.2.1 Produced Water Management Options

There are many approaches to managing produced water; some of them are discussed here. The primary alternatives being used today are underground injection, discharge, and beneficial reuse, although some other options are used at selected locations. Historically, produced water was managed in ways that were the most convenient or least expensive. Today, many companies recognize that water can be either a cost or a value to their operations. Greater attention to water management allows production of hydrocarbons and the concomitant profits to remain viable.

This chapter discusses water management technologies and strategies in terms of a three-tiered waste management or pollution prevention hierarchy (Veil, 2002). In the first tier (water minimization), processes are modified, technologies are adapted, or products are substituted so that less water is generated. When feasible, water minimization can often save money for operators and results in greater protection of the environment. For the water that is still produced following water minimization, operators next move to the second tier, in which water is reused or recycled. Some water cannot be recycled or reused and must be disposed of by injection or discharge.

In many cases, produced water can be put to other uses. Sometimes the water can be used without treatment, particularly when the produced water is very clean to start with or the end use does not require high water quality. In many other cases, the water must be treated before it can be reused. The cost of treating the water to meet an end use is an important factor in determining the types of reuse options that will be considered. The following sections describe a variety of approaches to recycling and reusing produced water.

19.10.2.2 Underground Injection for Increasing Oil Recovery

The most commonly used approach for managing onshore produced water is reinjection into an underground formation. Although some produced water is injected solely for disposal, most produced water (71%) is injected to maintain reservoir pressure and to hydraulically drive oil toward a producing well. This practice is referred to as water flooding, or if the water is heated to make steam, as steam flooding. When used to improve oil recovery, produced water ceases being a waste and becomes a resource. Without that produced water to use, operators would need to use other surface or groundwater supplies as sources of water for the water or steam flood. Typically, for water flooding, sufficient produced water volumes may not be available for injection. In these instances, other sources of water must be used to supplement the water flooding operation. Historically, freshwater sources have been used for this purpose. However, due to the increasing scarcity of this resource, other brines or water sources are now typically used in lieu of freshwater resources.

19.10.2.3 Injection for Future Use

Some types of produced water are relatively fresh and can be used directly with little or no treatment. This water may be used immediately for beneficial reuse, or it can be injected into an aquifer where it can be recovered for later use. This process is known as aquifer storage and recovery (ASR). The EPA treats ASR wells as Class V injection wells. In a 1999 survey, the EPA identified at least 130 ASR wells in use throughout the country (EPA, 1999), although they were not injecting produced water. Brost (2002) described an operation in the Kern River field of California in which a blend of produced water and treated groundwater was filtered and then sent to the local water district for use in both irrigation and aquifer recharge.

19.10.2.4 Livestock Watering

Livestock can tolerate a range of contaminants in their drinking water. At some concentrations, the animals, although still able to survive, will begin to show some impairment. ALL Consulting (2003) provided a table showing the total dissolved solids (TDS) levels that are appropriate for livestock watering. In general, animals can often tolerate a higher degree of TDS if they are gradually acclimated to the elevated levels. Water with TDS less than 1,000 ppm is considered to be an excellent source water. Water with TDS from 1,000 up to 7,000 ppm can be used for livestock but may cause some diarrhea (ALL Consulting, 2003).

19.10.2.5 Aquaculture and Hydroponic Vegetable Culture

Jackson and Myers (2002) reported on greenhouse experiments to raise vegetables and fish using produced water or potable water as the water source. The system used a combination of hydroponic plant cultivation (no soil) and aquaculture. Tomatoes grown with produced water were smaller than those grown in potable water. The produced water tank grew a larger weight of tilapia fish, although some of the fish died. None of the fish in the potable water tank died. The tests showed that produced water could serve as a water source for vegetables and fish when other potable water sources are not available.

19.10.2.6 Irrigation of Crops

Many parts of the United States and around the world have limited freshwater resources. Crop irrigation is the largest single use of freshwater in the United States, making up 39% of all freshwater withdrawn, or 150 billion gallons per day (USGS, 1998). If produced water has low enough TDS and other characteristics, it can be a valuable resource for crop irrigation.

ALL Consulting (2003) summarized crop irrigation water quality requirements, noting that the three most critical parameters are salinity (affects crops), sodicity (affects soil), and toxicity (affects crops). Salinity is expressed as electrical conductivity in units of mmhos/cm or more currently in micro Siemens per cm (ìS/cm). Crops have varying susceptibility to salinity; as salinity rises above a species-specific salinity threshold, crop yields decrease. Excess sodium can damage soils. Higher SAR[2] values lead to soil dispersion and a loss of soil infiltration capability. When sodic soils are wet, they become sticky, and when dry, they form a crusty layer that is nearly impermeable. Some trace elements in produced water can cause harmful effects to plants when present in sufficient quantities. ALL Consulting (2003) suggested that the most common sources of plant toxicity are chloride, sodium, and boron.

Another source of information on the effects of applying produced water to soils is a manual developed for the American Petroleum Institute on remediation of soils that had experienced produced water spills (API, 1997). The authors of that manual have subsequently taught a series of workshops on the same subject. The manual is a detailed guide with much useful technical information on the impacts of salinity and sodium on soils and vegetation.

Texas A&M University established a program to develop a portable produced water treatment system that can be moved into oil fields to convert produced water to potable or irrigation water. The goal is to produce water suitable for agricultural use (less than 500 mg/L of total dissolved solids and less than 0.05 mg/L of hydrocarbons). Such a system not only augments scarce water supplies in arid regions but also provides an economic payback to operators that could allow the well to produce longer (Burnett et al., 2002; Burnett and Veil, 2004).

19.10.2.7 Industrial Uses of Produced Water

In areas where traditional surface and groundwater resources are scarce, produced water may be substituted in various industrial practices as long

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[2] The Sodium Adsorption Ratio (SAR) defines sodicity in terms of the relative concentration of sodium compared to the sum of calcium and magnesium ions in a sample. The SAR assesses the potential for infiltration problems due to a sodium imbalance in irrigation water (Bauder et al., 2011).

as the quality of the produced water meets the needs of the industrial process with or without treatment. Produced water is already being used for several industrial uses and may be suitable for others. These are discussed next.

Dust control: In most oil fields, the lease roads are unpaved and can create substantial dust. Some oil and gas regulatory agencies allow operators to spray produced water on dirt roads to control the dust. This practice is generally controlled so that produced water is not applied beyond the road boundaries or within buffer zones around stream crossings and near buildings. Surface mining, processing, and hauling are inherently dusty activities. Produced water can be used for dust suppression at those locations, too, if regulators allow the practice (Murphree, 2002).

Oil field use: Peacock (2002) described a program in New Mexico through which produced water is treated to remove hydrogen sulfide and then is used in drilling operations. This beneficial reuse saves more than 4 million bbl per year of local groundwater.

Use for power generation: Produced water can be used to supply water to make steam. About 360,000 bpd of produced water from a Chevron Texaco facility in central California is softened and sent to a co-generation plant as a source of boiler feed water (Brost, 2002). Another potential use of produced water is cooling water. The electric power industry is the second largest user of freshwater in the United States, making up 38% of all freshwater withdrawn, or 150 billion gallons per day (USGS, 1998). Conventional surface and groundwater sources are no longer sufficient to meet increasing power plant needs in many parts of the country. Produced water represents a large-volume source of water that could potentially serve as makeup water for a power plant. Argonne National Laboratory recently completed a study that evaluates the use of another alternative type of water supply for power plant cooling (Veil et al., 2003). Although that study considers underground pools formed in abandoned coal mines, many of the report's discussions concerning water quality, water quantity, and mode of operation are relevant for using produced water as a cooling source.

19.10.3 Water Disposal Options

Ideally, operators will find cost-effective produced water management approaches that employ water minimization, recycling, or reuse. However, much of the world's produced water is not managed in those ways. Instead, it is disposed of through either discharge to surface water bodies or by underground injection. Although there certainly are exceptions, the following premises generally apply:

- Most U.S. onshore oil and natural gas well operators inject their produced water for either enhanced recovery or for disposal. To a large extent, this is necessary because discharge from most onshore wells is prohibited.
- Many U.S. CBM well operators try to discharge produced water to surface water bodies if they can obtain permission to do so.
- Most U.S. offshore operators discharge produced water to the ocean if they are permitted to do so. Offshore produced water is also typically discharged in other parts of the world. Not surprisingly, different countries employ different discharge standards. A small percentage of offshore produced water is used for enhanced recovery operations.

Occasionally disposal of untreated produced water is possible. In most cases, however, it is necessary to treat the produced water first. The type and extent of treatment are determined by a variety of factors:

- Where the water is going (e.g., freshwater surface water body, ocean, ground)
- The applicable regulatory requirements and allowable options
- The cost to transport and treat the water
- Site-specific factors (e.g., climate, availability of infrastructure)
- The potential for long-term liability
- A company's familiarity with or preference for specific options

It is important to ensure that the water being injected is compatible with the formations receiving the water to prevent premature plugging of the formation or other damage to equipment. It may be necessary to treat the water to control excessive solids, dissolved oil, corrosion, chemical reactions, or growth of microbes. Solids are usually treated by gravity settling or filtration. Reynolds and Kiker (2003) reported that a common rule of thumb for solids control is that all particulate matter larger than one-third the average pore-throat size of the receiving formation should be removed.

Residual amounts of oil in the produced water not only represent lost profit for producers but also can contribute to plugging of formations receiving the injected water. Various treatment chemicals are available to break emulsions or make dissolved oil more amenable to oil removal treatment.

Corrosion can be exacerbated by various dissolved gases, primarily oxygen, carbon dioxide, and hydrogen sulfide. Oxygen scavengers and other treatment chemicals are available to minimize levels of undesired dissolved gases. The water chemistry of a produced water sample is not necessarily the same as that of the formation that will receive the injected water. Various substances dissolved in produced water may react with the rock or other fluids in the receiving formation and have undesired consequences. It is important to analyze the constituents of the produced water before beginning a water flood operation to avoid chemical reactions that may form precipitates. If necessary, treatment chemicals can minimize undesired reactions. Bacteria, algae, and fungi can be present in produced water or can be introduced during water handling processes at the surface. These are generally controlled by adding biocides or by filtration.

19.10.4 Process Wastewater Management Options

Process wastewater and other wastewaters, which may be contain dissolved hydrocarbons, oxygenated compounds, and other contaminants, should be treated at the onsite wastewater treatment unit (WWTU). Recommended process wastewater management practices include the following (IFC, 2007):

- Prevention and control of accidental releases of liquids through inspections and maintenance of storage and conveyance systems, including stuffing boxes on pumps and valves and other potential leakage points, as well as the implementation of spill response plans.
- Provision of sufficient process fluids let-down capacity to maximize recovery into the process and to avoid massive process liquids discharge into the oily water drain system.
- Design and construction of wastewater and hazardous materials storage containment basins with impervious surfaces to prevent infiltration of contaminated water into soil and groundwater.

Techniques for treating industrial process wastewater in gas processing facilities include source segregation and pretreatment of concentrated wastewater streams. Typical wastewater treatment steps include grease traps, skimmers, dissolved air floatation, or oil/water separators for separation of oils and floatable solids; filtration for separation of filterable solids; flow and load equalization; sedimentation for suspended solids reduction using clarifiers; biological treatment, typically aerobic treatment, for reduction of soluble organic matter (BOD); chemical or biological nutrient removal for reduction in nitrogen and phosphorus; chlorination of effluent when disinfection is required; and dewatering and disposal of residuals in designated hazardous waste landfills. Additional engineering controls may be required for (1) containment and treatment of volatile organics stripped from various unit operations in the wastewater treatment system; (2) advanced metals removal using membrane filtration or other physical/chemical treatment technologies; (3) removal of recalcitrant organics, cyanide, and nonbiodegradable COD using activated carbon or advanced chemical oxidation; (4) reduction in effluent toxicity using appropriate technology (such as reverse osmosis, ion exchange, activated carbon), and (5) containment and neutralization of nuisance odors.

19.10.5 Management Options of Other Wastewater Streams

19.10.5.1 Storm Water

Storm water may become contaminated as a result of spills of process liquids. Natural gas processing facilities should provide secondary containment where liquids are handled, segregate contaminated and noncontaminated storm water, implement spill control plans, and route storm water from process areas into the wastewater treatment unit.

19.10.5.2 Cooling Water

Cooling water may necessitate high rates of water consumption, as well as the potential release of high-temperature water, residues of biocides, and residues of other cooling system antifouling agents. Recommended cooling water management strategies include (1) adoption of water conservation opportunities for facility cooling systems; (2) use of heat recovery methods or other cooling methods to reduce the temperature of heated water prior to discharge to ensure the discharge water temperature does not result in an increase greater than 3°C of ambient temperature at the edge of a scientifically established mixing zone that takes into account ambient water quality, receiving water use, assimilative capacity, etc.; (3) minimizing use of antifouling and corrosion-inhibiting chemicals through proper selection of depth for placement of water intake and use of screens; selection of the least hazardous alternative with regards to toxicity, biodegradability, bioavailability, and bioaccumulation potential; and dosing according to local regulatory requirements and manufacturer recommendations; and (4) testing for the presence of residual biocides and other pollutants of concern to determine the need for dose adjustments or treatment of cooling water prior to discharge.

19.10.5.3 Hydrostatic Testing Water

Hydrostatic testing of equipment and pipelines involves pressure testing with water to verify their integrity and to detect possible leaks. Chemical additives (typically a corrosion inhibitor, an oxygen scavenger, and a dye) may be added. In managing hydro-test waters, the following pollution prevention and control measures should be implemented: (1) using the same water for multiple tests to conserve water and minimize discharges of potentially contaminated effluent; (2) reducing the use of corrosion inhibiting or other chemicals by minimizing the time that test water remains in the equipment or pipeline; and (3) selecting the least hazardous alternative with regards to toxicity, biodegradability, bioavailability, and bioaccumulation potential; and (4) dosing according to local regulatory requirements and manufacturer recommendations. If discharge of hydro-test waters to the sea or to surface water is the only feasible alternative for disposal, a hydro-test water disposal plan should be prepared considering location and rate of discharge, chemical use (if any), dispersion, environmental risk, and required monitoring. Hydro-test water disposal into shallow coastal waters should be avoided (IFC, 2007).

19.10.6 Hazardous Materials

Natural gas processing facilities use and produce significant amounts of hazardous materials, including raw materials, intermediate/final products, and byproducts. The handling, storage, and transportation of these materials should be managed properly to avoid or minimize the environmental impacts from these hazardous materials. Hazardous waste should be determined according to the characteristics and source of the waste materials and applicable regulatory classification. In gas-to-liquids (GTL) facilities, hazardous wastes may include bio-sludge; spent catalysts; spent oil, solvents, and filters (e.g., activated carbon filters and oily sludge from oil–water separators); used containers and oily rags; mineral spirits; used sweetening; spent amines for CO_2 removal; and laboratory wastes.

Spent catalysts from GTL production are generated from scheduled replacements in natural gas desulphurization reactors, reforming reactors and furnaces, Fischer–Tropsch synthesis reactors, and reactors for mild hydrocracking. Spent catalysts may contain zinc, nickel, iron, cobalt, platinum, palladium, and copper, depending on the particular process. Recommended waste management strategies for spent catalysts include the following: (1) proper on-site management, including submerging pyrophoric spent catalysts in water during temporary storage and transport until they can reach the final point of treatment to avoid uncontrolled exothermic reactions; (2) return to the manufacturer for regeneration; and (3) off-site management by specialized companies that can recover the heavy or precious metals through recovery and recycling processes whenever possible or that can otherwise manage spent catalysts or their nonrecoverable materials. Catalysts that contain platinum or palladium should be sent to a noble metal recovery facility.

Heavy ends from the purification section of the methanol synthesis unit are normally burned in a steam boiler by means of a dedicated burner.

19.10.7 Venting and Flaring

Venting and flaring are important operational and safety measures used in natural gas processing facilities to ensure gas is safely disposed of in the event of an emergency, power or equipment failure, or other plant upset conditions. Unreacted raw materials and byproduct combustible gases are also disposed of through venting and flaring. Excess gas should not be vented but instead sent to an efficient flare gas system for disposal. Recommendations to minimize gas venting and flaring include the following (IFC, 2007):

- Optimize plant controls to increase the reaction conversion rates.
- Recycle unreacted raw materials and byproduct combustible gases in the process or utilize these gases for power generation or heat recovery, if possible.
- Provide backup systems to achieve as high a plant reliability as practical.
- Locate the flaring system at a safe distance from residential areas or other potential receptors and maintain the system to achieve high efficiency.

Emergency venting may be acceptable under specific conditions where flaring of the gas stream is not appropriate. For example, in GTL processing, there could be streams containing high concentrations of carbon dioxide, which, if sent to a flare system, would extinguish the flare's flame; venting of such streams to a safe atmospheric location is an acceptable option. Standard risk assessment methodologies should be utilized to analyze such situations. Justification for not using a gas flaring system should be fully documented before an emergency gas venting facility is considered.

19.10.8 CO_2 Sequestration

The combustion of fossil fuels releases greenhouse gases, principally carbon dioxide (CO_2), methane, and nitrous oxide. Carbon dioxide and methane from the produced gas is released from liquefied natural gas (LNG) operations in significant quantities. Atmospheric concentrations of these gases have increased substantially in the last 200 years. These increases have raised concerns that the related global warming effects will melt polar ice and raise sea levels, change weather patterns and climatic zones, and consequently modify ecosystems. In response, some nations have agreed to limit their greenhouse gas emissions. Australia negotiated to limit its increase in emissions from 1990 to 2010 to 8% of its total 1990 emission figure (Venture, 2003). It is within this context that ChevronTexaco proposes to inject its reservoir CO_2, into the hyper-saline Dupuy Formation aquifer beneath Barrow Island, at depths below 2700 m beneath the surface. The company's commitment to inject CO_2 is qualified by a proviso that it is technically feasible and not cost prohibitive. Should this not be achievable, there are some contingencies discussed in the ESE review for partial offsetting of CO_2 in a future Greenhouse Gas Management Strategy document but no firm commitment to offset the full quantity of reservoir CO_2. Advice from ChevronTexaco indicates that it considers injection beneath Barrow Island to be a unique opportunity. While ChevronTexaco intends to investigate alternative injection locations and other offset mechanisms, none is considered to be viable at this stage. The ESE Review finds that Barrow Island and the Dupuy Formation offer the highest degree of maturity, or confidence, on a combination of economic and geologic factors. Injection would avoid the emission of 4.2 million tonnes of CO_2 annually from two trains of LNG, each rated at 5 mtpa, but it would not offset all the emissions from the project. The quantity proposed for injection amounts to approximately 13% of the life-cycle CO_2 emissions from the production and use of LNG or 57% of the emissions from the production process. The project would still emit 3.2 mtpa, mainly as exhaust gases, from the LNG process plant and power station complex. Injection is thus a partial mitigation of an environmental impact that would otherwise ensue. If gas were to be piped to the mainland at the rate of 300 terajoules per day for domestic consumption, the compressors required would add 0.1 mtpa of CO_2 to the emissions, for a total CO_2 output of 3.3 mtpa from the process plant. Power generated for the project would be by gas turbines using cogeneration technology to maximize efficiency and minimize these emissions, but they would not be captured or returned to the ground. The EPA recognizes that there is a strategic value to a company in establishing on Barrow Island a central repository for subsurface disposal of CO_2, not only for its own gas reserves, but possibly for other developers in the region with unwanted carbon dioxide. This is particularly so if governments should in the future decide to impose a more restrictive greenhouse gas management regime on industry. The EPA notes, however, that there is currently no legal or policy requirement to do so at either the federal or state government level. The EPA believes it is highly desirable for developers to consider innovative means for managing the carbon dioxide generated by industrial developments and encourages the pursuit of sequestration options. In the event that geological sequestration is not technically feasible or is cost prohibitive, then the EPA recommends that other appropriate mechanisms to sequester or otherwise manage an equivalent amount of CO_2 should be required. The EPA notes, however, that there are potential locations for CO_2 injection other than Barrow Island (Venture, 2003). Even if CO_2 could be injected only on Barrow Island and nowhere else, an injection facility could conceivably be established there with the rest of the gas complex being located elsewhere. While such a scenario would require careful assessment, the reduced infrastructure required would reduce the potential threat to Barrow Island from additional development. A key issue for the EPA is the tension between the potential risks to the high conservation values of Barrow Island and the benefits from CO_2 injection. The EPA considers that the CO_2 benefits remain uncertain because there is as yet no unqualified commitment from the developer to inject CO_2. The EPA is also aware that CO_2 injection is not necessarily dependent on Barrow Island. Potential reservoirs exist elsewhere in the general area, possibly at increased cost (Venture, 2003). The developer regards these alternatives as less "mature," meaning there is presently a lower level of certainty that they would be suitable or reliable because they have been subjected to less study. Given the known and unique conservation values of Barrow Island and the uncertain benefits of injection, the EPA considers that potential CO_2 injection benefits are secondary to the risk to the conservation values of Barrow Island in terms of its assessment of relative environmental importance. One issue of particular environmental relevance is the impact of leakage of CO_2 back to the surface. Carbon dioxide is invisible, odorless, and heavier than air; therefore, it tends to collect in depressions. Too much CO_2 is harmful to life forms that use oxygen to exist. Plant roots need oxygen in the soil; CO_2 normally comprises up to 1% of the gas in soils. If the oxygen is replaced by CO_2 seeping upward, the plants will become stressed and may die, as in the case at Mammoth Mountain in California, where it has created dead patches totaling 100 acres in former forest (USGS, 2000). In extreme cases, CO_2 may be lethal to fauna and people. For example, at Lake Nyos in West Africa, at least 1,700 people were killed by asphyxiation in August 1986 as a result of a sudden natural CO_2 release (USGS, 1998).

19.10.9 Life-Cycle Analysis

Pollution prevention often utilizes a principle known as "life-cycle analysis" to address all associated costs and possible solutions associated with a particular process or waste. Life-cycle analysis, sometimes referred to as "cradle-to-grave" analysis, is often used to track a particular material from its inception to its ultimate demise. This tracking usually requires documentation from other companies (both vendors and customers) in the material chain. In material substitution, for example, a possible material alternative that would drastically reduce a particular waste stream may require a process change by the vendor first. Also, a positive pollution prevention approach implemented by a particular company could have negative impacts to its customers or contractors. For these reasons, it is helpful to include vendors, customers, and contractors as part of the pollution prevention team.

19.10.10 Training and Awareness

Training and awareness programs are critical to ensuring that pollution prevention is realized to its fullest potential. The best ideas will come from persons who work with machines, use materials, and generate waste. These people must be aware that often there are alternatives and that they constantly need to be thinking about ways to improve operations, efficiency, etc. It is always more effective to provide pollution prevention training to persons with process knowledge than to provide "pollution prevention experts" with process knowledge to develop a pollution prevention plan.

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CHAPTER 20

Maximizing Profitability of Gas Plant Assets

20.1 INTRODUCTION

Conditions outside the range of data that were used to train a neural network model may be suspect. For example, if a neural network model was trained on data collected while a Joule-Thomson (JT) valve was closed, then the predictions from the model are invalid when the JT valve is open.

Maximizing the return on gas plant assets becomes increasingly difficult because of the rising cost of energy in some cases and the increased demand for operations agility in most cases. On top of these demands, there is the constant need to increase the availability and utilization of these plants. It is worthwhile to consider techniques and the profit improvement of some of the world's best gas plants. This analysis identifies some key manufacturing and business strategies that appear to be necessary and are feasible for many facilities to achieve and sustain satisfactory performance. The essential nature of operating a gas plant is dealing with change. Disturbances such as slugging, trips of gathering compressors, weather changes, and changes in market demands make it challenging to operate reliably, let alone profitably. A stabilization strategy is necessary but insufficient for most gas plants; it is necessary to sustain a degree of flexibility. A useful objective can be tracking desired changes and withstanding undesired changes in a manner that is safe, environmentally sound, and profitable (DeVries et al., 2001).

The main challenges to gas plant profitability are

- Continuous energy and yield inefficiency due to providing operating margin for upsets and plant swings
- Continuous energy and yield inefficiency due to not operating at optimum conditions
- Energy and yield inefficiency due to the consumption during plant swings
- Labor costs to operate and support a gas plant—a poorer performing plant requires more personnel with a higher average salary
- Plant integrity
- Process and equipment reliability

- Poor yield due to low availability
- Maintenance
- Safety

Poorer performance compounds the cost challenges in most gas plants. To ensure these problems are avoided, we need to develop a fundamental understanding of all the technical factors impacting the performance of the plant, how they interact, and how they manifest themselves in business performance. This can only be achieved through the application of an integrated approach utilizing the diverse skills of a multidisciplined team of engineers combined with the application of a robust performance-modeling tool. Technical assessments allow an understanding of the technical risks to which the facility is exposed, ranging from deterioration of equipment due to exposure to corrosive environments to poor process efficiency as a result of poor design. Integration of expertise within a single team, which covers the broad range of technical fields involved in the operation of the gas processing plant, ensures that all the technical risks and interactions are identified.

Other business questions that asset managers are asking themselves include (Howell, 2004):

- Are the assets performing to plan, and how do we know?
- Are we choosing the optimal plans for developing the assets over their lifetime?
- Are we achieving the targeted return on capital employed (ROCE) for the assets?
- Are we meeting all of the ever-growing health, safety, and environment (HSE) guidelines?
- Can we forecast reliably, allocate with confidence, or optimize with knowledge?
- Do we derive enough value from the simulation and engineering model investments?
- How effective is the organization at capital avoidance?
- Are we drowning in data or are we knee-deep in knowledge?

Another area of opportunity is optimization of the process to maximize capacity or yields and minimize energy consumption while maintaining product qualities. Many tools became available in the past decade as powerful, inexpensive computing power became available to run complex software quickly and reliably.

This chapter describes a vision of the integrated gas plant of the future and methods to identify solutions for attaining operational goals to maximize asset values.

20.2 THE PERFORMANCE STRATEGY: INTEGRATED GAS PLANT

Successful gas plants have found that a combination of techniques is necessary:

1. A strategy to influence organizational behavior
2. A strategy to integrate information
3. An operations strategy that uses remote operations and support of unmanned plants
4. Process performance monitoring
5. Asset management
6. Process optimization implementation

Merely adding technology without developing a new organizational behavior and operations strategy has often reduced gas plant performance instead of improving it.

The revolution in digital technologies could well transform the industry. Achieving the vision of the integrated gas plant of the future will require more than new technologies alone. It will require the alignment of strategy structure, culture, systems, business processes and, perhaps most important, the behaviors of people. Visionary companies that truly want to capture the "digital value" will need to create a climate for change and then maintain strong leadership through the change and employ the skills and techniques from vendors to provide the technologies.

Gas plant operators are looking to integrate global operations and the energy supply chain into a cohesive picture. A global enterprise resource planning (ERP) system gives companies the resources they need to better balance supply with demand, reduce inefficiencies and redundancies, and lower the total cost of IT infrastructures. The challenge is to develop overlay solutions with more domain content that can improve the knowledge inside the enterprise systems that producers depend on most for operations, planning, project management, workflow, document management, executive information and decision support, scheduling, database management, data warehousing, and much more.

The industry must capitalize on the opportunities provided by ever more capable and cost-effective digital technology.

20.3 STRATEGIES FOR ORGANIZATIONAL BEHAVIOR AND INFORMATION

Understanding and managing organizational behavior are important elements of effective operations. The effective management and coupling of technology to people are crucial elements in ensuring that the plant's culture will change to support modern manufacturing strategies.

Four typical organizational cultures have been observed (Neumann, 1999). These cultures vary from cultures very resistant to change to cultures that change for the sake of change.

20.4 ORGANIZATIONAL BEHAVIOR MODEL

It is worthwhile to review recent modeling attempts for organizational behavior. Two models, each focusing on different aspects of the organization, deserve exploration. Both attempt to explain the influence of information and management activities on organizational behavior. The first model, as shown in Figure 20-1, attempts to identify and provide a structure for the various major influences on the organization.

This model supports the following observations:

1. Unless constrained by their situation, people will improve their behavior if they have better information on which to act.
2. Behavior is constrained by people's capability to perform. This constraint is tied to the span of their control, their associated physical assets, etc., and what the military calls "readiness."
3. The classical Personal Hierarchy of Needs (Maslow, 1968) that ranges from "survival," "hunger," and "need to belong" applies to organizations as well.

20.4.1 Information Quality

The natural gas processing industry is unusual in its high degree of dependence on information technology in order to meet its business goals. With the enormous quantities of data that it generates and processes, the industry gains an edge by ensuring the quality of the data and using the information

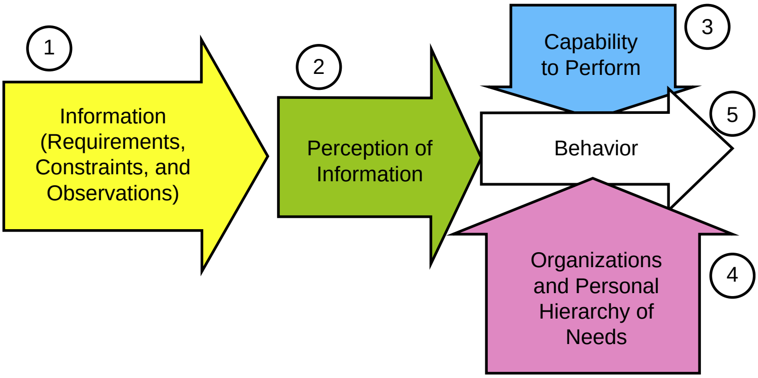

Figure 20-1 Major influences on the organization (DeVries et al., 2001).

intelligently. Alongside this problem, the industry is addressing the changes required due to new and evolving working practices that the popularity of interdisciplinary asset teams has brought about.

This organizational model explains in part why people and organizations behave in a rebellious manner in spite of receiving "better" information. For example, operators and their supervisors are penalized for "tripping" units and major equipment; however, e-business decisions might require them to operate the plant in a manner that would increase the likelihood of trips. Another example is where a plant's or unit's performance is broadcast to a team of marketing and other plants while the plant or unit is chronically "underperforming" based on the other organizations' expectations. This motivates the plant's managers to hide information and resist plans for performance improvement.

This model also points to a key issue in managing technology. Poor-quality information will be discarded, and information that is harder to use will not be used. This points to the need to define the quality of information. Suggested attributes of information quality include the following:

- Available (is the information chain broken?)
- Timely
- In the right context
- Accurate
- To the right place
- Easy to understand

It may be helpful to imagine a fuel gauge in the dashboard of a car. If the fuel gauge is working 75% of the time, the driver will not trust it; the gauge will be ignored. In process plants, information has to be much more reliable than that to be trusted.

Gas plants struggle to convert supply-chain schedules into actual plans. Too often, the marketing department doesn't have appropriate feedback that equipment has failed or that a current swing in recovery mode is taking longer than expected to complete. Conversely, the operations department often lacks the tools to plan which resources and lineups to use, while avoiding poor utilization of equipment. This gap between strategic and tactical scheduling is sometimes referred to as the "operations wall."

Much attention is given to networking information and knowledge software. So much attention is given to making as many components as possible accessible via the Web to support virtual private networks (VPNs) and restructuring knowledge software using application service providers (ASPs). This work is worthwhile, but the underlying information reliability issues are chronic and damage the credibility of major network and software installations.

The old expression "if garbage goes into the computer, then garbage will come out" is especially true for modern information technology strategies that end up with more than seven layers of components that process information. It is extremely important to note the following:

1. Much of the information used to enhance and change plant performance, as well as support supply chain management decisions, is based on sensors that degrade in accuracy and "fail" due to stresses or material that reduces their ability to produce a reliable measurement. A modern hydrocarbon processing plant uses hundreds of these to support advanced control and advanced information software. Simply connecting digital networks to these sensors does not improve information reliability if the sensors don't have enough "intelligence" to detect and possibly compensate for failure or inability to measure.
2. Most components and software packages process online diagnostic information that gives some indication of the reliability of the information. However, most installations don't integrate these diagnostics with the calculations. If the information reliability of each of seven components from the sensor to the top e-business software is 95%, then on average the overall information reliability is no better than 69%. Raising the component reliability to 99% still yields only a maximum overall information reliability of 93%, which is unacceptable for supply-chain management.

Therefore, it is important to use as much "intelligence" as possible at all levels and maintain the integration of online quality statuses to maximize the quality and minimize the time to repair the failures.

20.4.2 Perception of Information

Tables of menus and numbers don't help when a downsized team is asked to open multiple applications and "mine" for information. Most information displays are essentially number tables or bar graphs superimposed on plant flow sheets. Several plant examples exist to show how this interferes with improved performance.

20.4.2.1 Two-Dimensional Curves and Plots

As an example, compressors can be the main constraining factor in major hydrocarbon processing facilities, and it is important to maintain the flow through each stage of compression sufficiently high to avoid surge. Too low a pressure increase is inefficient and limits production throughput. The most efficient is at higher pressure increases but yields operation closer to surge. Operators are normally given a pair of numbers or bar graphs

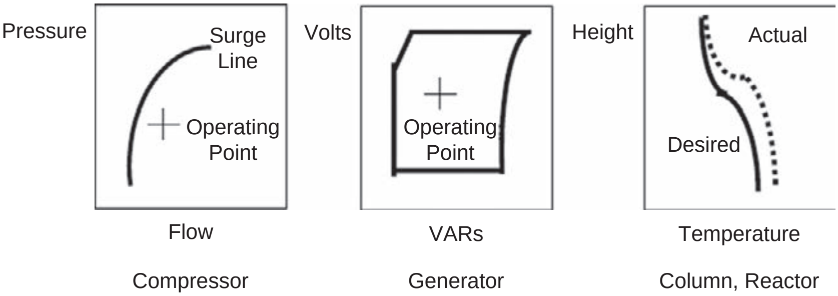

Figure 20-2 Examples of two-dimensional curves and plots (DeVries et al., 2001).

for pressure and flow, and they need to recall from experience where the "operating point" is compared to the "surge line." To make matters worse, the "surge line" changes position as the gas's characteristics change. To solve this, operators can plot the moving surge line and the operating point in real time. This is a suitable solution for distillation and reactor temperature profiles as well (see Figure 20-2).

20.4.2.2 Prediction Trends

The same operators who avoid trips tend to shut off advanced control software when the software is driving the plant toward undesirable states. But often the software is only using a spike to accelerate the transition to best performance, and the top of the spike still keeps the plant in the desirable state. Prediction trends allow operators to see the remainder of the curve, so that if the curve looks safe, they will allow the software to continue with the spike.

20.4.2.3 Dynamic Performance Measures

Operators and their supervisors are given performance information and targets, but often the information isn't "actionable"; i.e., they cannot directly influence the information (see Figure 20-3). Furthermore, as manufacturing

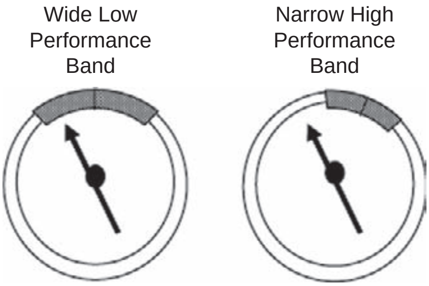

Figure 20-3 Dynamic performance dashboard examples (DeVries et al., 2001).

strategies change, they receive e-mails with a set of numbers. Ideally, the senior operators and their chain of management are seeing the targets and acceptable bands along with each actionable performance measure. In this way, coordination of a change in manufacturing tactic is faster and easier. A common example is changing from maximum throughput (at the expense of efficiency) to maximum efficiency (at the expense of throughput) to take advantage of market opportunities and minimize costs when the demand is reduced.

20.4.2.4 Performance Messages

Operators who monitor long batches, especially batches that will cover a shift change, tend to have lower productivity, especially during times that the main support staff isn't available. Ingredients that exhibit quality problems can be used in applications such as inline blending of fuels and batch blending with online analyzers to support a performance model with a set of messages and procedures that allow operators to modify the batch during the run and avoid a rework.

20.4.3 Capability to Perform

When organizations can more accurately perceive their performance and the distance away from targets, they can develop a culture of learning how to improve. This is different from using monthly accounting reports that show shadow costs and prices. This approach requires a condensed set of reliable information that supports factors that they can change. Examples include reliability, yield, quality, throughput, and often it might be expressed in a different way. For example, reliability may be more "actionable" when expressed as "cycle time," "turnaround time," etc. Cutting the setup time from an overall 5% to 2.5% is actually cutting the time by 50%; this change is far more vivid than showing availability changing from 95% to 97.5%.

Research has been conducted on complex operations where a unit is several steps away from the true customer and attempts to establish internal costs and prices of feedstock, utilities, and internal products have had insufficient credibility or even ability to bring about a change. Some attempts at activity-based costing have tried to increase the percentage of direct costs, with the result that the cost per unit of product is so weighted with costs outside the unit that the unit manager cannot bring about a useful improvement or is motivated to maximize throughput. Maximizing throughput is the only degree of freedom because the "costs" are relatively fixed.

This is incompatible with modern manufacturing strategies that require precise, timely changes in prioritization of throughput, efficiency, and specific formulations to attract and retain key customers and more lucrative long-term contracts.

A key element of an organization's capability to perform is its ability to handle both planned and unplanned disturbances. Modern manufacturing strategies become a nightmare if the organization and facilities can't cope with the accelerated pace of change brought on by the tight coupling with suppliers and customers. The following model, as shown in Figure 20-4, helps to describe different levels of readiness.

Containment is the lowest level of readiness. At this level, unplanned disturbances stop production or cause product rework, but the damage is "contained"; there is no injury, equipment damage, or environmental release. Unplanned disturbances include operator errors, equipment failures, and deviations in feedstock quality, as well as other factors. This level of readiness can be achieved with conventional information strategies:

- Loose or no integration between components from sensors to supply chain software
- Minimal integration of diagnostics of information quality
- Minimal performance information
- Minimal coordination between units or assets to withstand disturbances
- Reactive or scheduled maintenance strategies

Prevention is a level of readiness in which unplanned disturbances rarely affect production availability or product quality. However, this operation cannot consistently support planned disturbances to production rates, feedstock quality, or changes in yield or product mix. It is usually a case of stabilizing

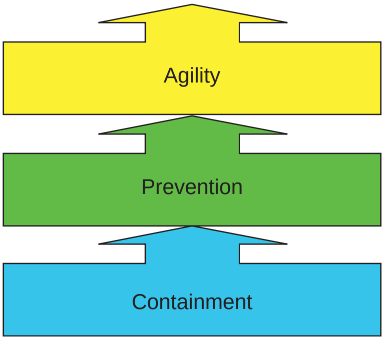

Figure 20-4 Levels of readiness (DeVries et al., 2001).

the plant as much as possible in order to achieve this level of performance. This level of readiness requires more advanced information strategies:

- Moderate integration among components and a reliable connection to supply chain software
- Good management of information quality
- Good coordination among units and assets
- Scheduled maintenance at all levels
- Better performance information

Agility is a level of readiness in which the plant can consistently support wider and faster changes in feedstock quality, production rates, and output mix without reducing production availability or quality. This operation consistently achieves "prevention" readiness so that it can outperform other plants because of its agility. This level of readiness requires the most advanced information strategies:

- The tightest integration of all levels of software and sensors
- Good models that drive performance information and online operations advice
- A reliability-centered maintenance strategy for information technology
- Good coordination with knowledge workers, who will likely come from key suppliers of catalysts, process licensors, information and automation technologies, and workers at other sites of the company

There is a tendency in many operations to try to jump from "containment" to an "agility" level of readiness by adding more information technology, but with the characteristics of "containment" (minimal integration, etc.). This is potentially disastrous. Supporting modern manufacturing strategies can mean that the quality of integration is as important or more important than the quality of the information components themselves.

Due to the complexity of the interactions among technical factors affecting asset performance and the difficulty in converting technical understanding into a business context, the traditional approach has been to try to simplify the problem. Benchmarking, debottlenecking studies, maintenance, and integrity criticality reviews have all been applied to enhance process plant performance.

Each of these approaches tends to focus on a particular element of an asset's operation. As a result, enhancement decisions are made without a full understanding of the impact on the asset's performance and the overall business impact. This can result in opportunities being missed, capital being poorly invested, and delivery of short-term benefits, which are unsustainable. In the worst cases, this can result in a net reduction in plant performance and increased life-cycle costs.

20.4.4 Organizational Hierarchy of Needs

Attempts to evolve plant culture by transplanting methods and equipment have not only failed but actually eroded performance. Key observations include the following:

1. Personnel may not associate performance or knowledge with increased security, wealth, and sense of belonging or esteem. They observe promotion and compensation practices and evaluate the effectiveness of improving their situation against new expectations for better performance.
2. Risk-adverse cultures dread the concept of visual management, benchmarking, or any other strategy that is designed to help a broader teamwork together to continually improve performance. Information shows performance can be used to support rivalries rather than inspire teams to improve.
3. Organizations and personnel worry about surviving. Will the plant be shut down? Will the staffing be reduced? Efforts to change behavior using information have to be coupled with the hierarchy of needs, and then the information can evolve as the organization evolves (and performance improves).

Research on dynamic performance measures has uncovered an effective model of organizational behavior that is especially appropriate for manufacturing. Two of the key issues that have been addressed are overcoming more than 85 years of traditional organization structure that tries to isolate units and departments and the 20 years of cost and performance accounting.

The challenges with cost accounting are assigning costs in the correct manner. A plant manager or operations manager might face an insurmountable fixed cost. This motivates the manager to maximize production, which might be opposite to the current manufacturing requirements. The poor credibility of internal prices for utilities, feedstocks, and products also has motivated middle management to drive performance in the wrong way. These efforts point to the need to redefine key performance indicators (KPIs). The group of research efforts to deal with this is called dynamic performance measures. The U.S. Department of Labor and many large corporations embrace dynamic performance measures.

Vollman et al. (1998) developed a model that associates a manager's strategy and actions to his measures of performance. This combination is called the Vollman triangle, as shown in Figure 20-5.

The manager's subordinates will develop their own triangles to support their supervisor's measures, within the constraints of their capability to perform, etc., as outlined in Figure 20-6.

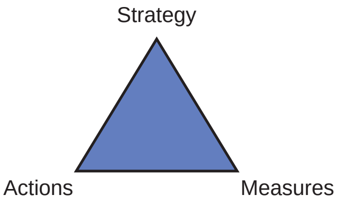

Figure 20-5 Vollman triangle.

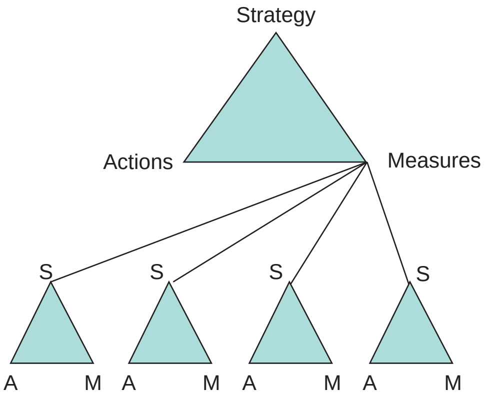

Figure 20-6 Application of the Vollman triangle to subordinates (DeVries et al., 2001).

This behavior occurs at all levels of the organization. The goal is to ensure that a small group of measures (no more than four), which can be consistently affected by that level of operations, is maintained with suitable quality. If this structure is maintained, then it is easier for top management and modern manufacturing strategies to bring about change.

The "actionable" measures tend to conflict with each other. Higher throughput or quality often comes at the expense of efficiency, or each has a different optimum. Each person learns how these interact. Discussions on performance improvement become more effective because personnel can now describe this behavior. The discussion is vastly different from reviewing monthly reports. As far as operations are concerned, something that occurred several days or weeks ago is ancient history.

20.4.5 Behavior

The most important variables for performance are the work culture, job security, and career mobility. In many parts of the world, knowledge, experience, or skill isn't sufficient to achieve promotion. Many cultures do not embrace an openness of sharing performance information, visual management techniques, etc. Therefore, the information strategy needs to be adjusted to reflect the current human resources strategy, and ideally, both will evolve in phases to achieve world-class performance.

Another strong issue is traditions at the plant. It is often much easier to initiate new teamwork and performance initiatives to a new plant with a new organization. Nevertheless, the right information strategy becomes a catalyst of change. Everyone can see the performance against the current targets, and everyone can understand the faster change of priorities, such as changing from maximum production to minimum, where maximum efficiency is desired.

20.5 THE SUCCESSFUL INFORMATION STRATEGY

Technology makes it easier to measure variables such as pressure, speed, weight, and flow. But many operations achieve profitability with very complex equipment, and the key characteristics are properties, not basic measurements. A notable example is found in oil refining. The basic manufacturing strategy is to improve yield by converting less-valuable components from crude oil (molecules that are too short or too long) into valuable components (medium-length molecules). However, the essential measurement to determine the proportion of different molecules is chronically unreliable. New technology now exists to reliably indicate a property called "carbon aromaticity," which helps the operations team ensure that the process will be effective and that the feedstock won't degrade the expensive catalyst materials and erode production.

Reliable property measurement is the key. This also helps to evolve the plant culture from thinking about pressures and flows to thinking about properties. The information strategy becomes the following:

1. Assess the organization's readiness to support the desired manufacturing strategy and use of information to evolve the plant culture.
2. Depending on the organization's hierarchy of needs, develop a phased evolution of the information strategy.
3. Develop appropriate measures that will directly support the actions needed to implement the strategy at all levels.
4. Develop the appropriate information to maximize its quality, reliability, and perception.
5. Manage the quality and reliability of information.

Data integration and visualization will continue to be key areas of focus throughout the next few years in digitizing the asset. Many new applications will bring knowledge-gleaning ability from the data collected.

20.6 THE IMPACT OF LIVING WITH INFORMATION TECHNOLOGY

If an organization invests in "intelligent" components that enable advanced maintenance strategies such as reliability-centered maintenance or performance monitoring centers, then the organization is committing to the advanced strategy. Otherwise, the flood of extra information will be disruptive. Intelligent sensors deliver up to 10 times as much information as conventional ones; a typical downsized organization, with as little as one-fifth the staffing of 10 years ago, would have to deal with 50 times the volume of information—too much. The information is useful only if it is coupled with appropriate software and methods to use the information to optimize the maintenance and maximize overall information and plant availability.

If the organization invests in supply-chain management technology, then the organization is committing to an "agility" level of operations readiness and a team-based performance culture. The organization is also committing to business measurement changes; cost per unit of product gets in the way of profitability measures of reducing procurement costs or increasing return on sales using modern manufacturing strategies.

The organization develops a new language and starts to use new terminology. Furthermore, managers start focusing on managing the higher levels of information. The following model, as shown in Figure 20-7, suggests a new terminology for dealing with information.

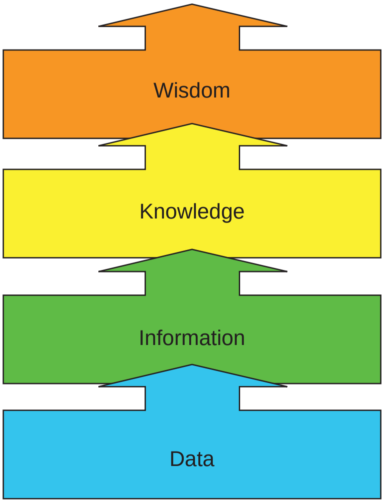

Figure 20-7 Hierarchy of information (DeVries et al., 2001).

This terminology can be defined as follows:

Data: The raw information from sensors and personnel keyboard entry.

Information: The validated data, using diagnostics and techniques to enhance the quality and reliability of information.

Knowledge: A comparison of information to targets and constraints. This answers the set of "how are we doing?" questions.

Wisdom: Guidance for best practices, customer satisfaction, supply chain management, and any other performance guidelines.

Many organizations have managers spending much of their time dealing with the "information" function rather than focusing on "knowledge" and "wisdom." It is extremely important to maximize the quality and type of information to ensure that management can support the "agility" demands of modern manufacturing strategies.

20.7 VISION OF THE MODERN PLANT OPERATION

Four different groups of activities require consistent information for most effective operations:

- Operational decisions on an hourly or more frequent basis by operators, shift supervisors, or engineers.
- Tactical decisions on a 1- to 30-day basis by shift supervisors, engineers, purchasing, trading, and accounting.
- Strategic decisions made on a 1-month to 5-year basis by plant management, purchasing, and accounting.
- Monitoring on at least an hourly, weekly, and monthly basis by all parties involved.

Operational decisions include determination of setpoints for the equipment and switching equipment on or off. These decisions need to consider current pricing for commodities such as fuel and electricity, equipment availability, and environmental constraints. Furthermore, environmental constraints might be accumulative.

Tactical decisions include maintenance scheduling, demand forecasting, production planning, and emissions forecasting and trading. Strategic decisions require evaluation of future investment, budgeting, and long-term contract negotiations with suppliers and customers. Monitoring requirements include tracking plan versus target versus actual performance, such as process energy use, cost accounting based on real costs, emissions accounting, and performance monitoring of utilities equipment.

Given that the quality and perception of information has to be improved in order to be used effectively to change an organization's culture, five key issues highlighted by this model must be addressed:

1. Capability to perform
2. Operations readiness
3. Organizational hierarchy of needs
4. Establishing "actionable" information to act as a catalyst of change
5. Measuring the right things

Technology can be the most effective tool to evolve plant culture. The speed of distributing reliable, timely, and easy-to-use performance and supply information changes the level of discussion and the dynamic behavior of the organization. However, the organization needs to have the right set of performance measures, a persistent effort to enhance operations readiness, an appropriate human resources strategy, and above all, appropriate quality of information. The impact of these on changing a plant culture can be dramatic. As soon as team members realize the ability and need to improve, they learn quickly how to evolve their work methods to achieve it.

Modern manufacturing strategies are viable, and information technology strategies can transform the plant culture to achieve manufacturing success. The evidence of a change in the level of dialog and team dynamics confirms the success. But several key strategies must be consistently maintained in order to sustain this success.

20.8 OPERATIONS STRATEGY

Major installations have been able to consolidate up to 10 control rooms and improve plant output by up to 15%. The goal wasn't to reduce personnel. The goal was to improve flexibility because it is easier for a small team to drive change than a large one. This has been consistently proven in remote jungle sites in Southeast Asia, remote mountain sites in the United States and Canada, and remote sites in northern Africa. There are several techniques necessary to make this feasible:

- Remote, secure, online computer access to all control and monitoring equipment.
- Advanced alarm management that dynamically filters alarms during upsets, thus avoiding "alarm showers" that blur the visibility to key cause-and-effect alarms.
- Stabilization techniques to help units withstand upstream and downstream disturbances.

20.9 MODEL-BASED ASSET MANAGEMENT

Facility simulation has developed quickly over the last several decades. The integrated oil and gas field with production facilities is a complicated system with a high degree of interaction and dynamics that make it impossible for the human mind to control and optimize both technical and business parameters. Integrated discipline workflow and the development of model-based asset management are necessary to effectively deal with these complexities.

Model-based asset management techniques will begin to play a major role in the integrated gas plant of the future in bringing the predictive power of the production engineering toolset to the real-time data platforms. Process and production simulation models will move from the domain of the engineering expert to be used by managers, operators, business development, contracts, and finance. The analysis and prediction of near real-time and future asset performance become a reality in the world of model-based asset management. For this, the complete integrated asset will be modeled dynamically in real time for both slow rigorous and fast proxy loop modes.

Facility simulation has come of age. The ability to predict very complex facility simulations in both steady-state and dynamic mode has been achieved through continued integration of thermodynamic methods, hydraulic simulations, and unit operations. Facility simulation has progressed such that dynamic start-up and shutdown simulations have become routine and trusted among the engineering community. One can begin to think of dynamic simulation as a "virtual plant" for operational, advanced process control, and business analysis by a number of different departments.

Facilities simulation, control, optimization, operator training, collaborative engineering, and planning disciplines have a rather fragmented application of a number of associated technologies. The first step to overcome this fragmentation is the adoption of integrated asset models that use comprehensive existing applications integrated together by a common "glue" layer, allowing a full model of the entire operation.

New concepts are appearing from the software community in the form of workflow solutions that enable assets to be modeled from a suite of selected software adaptors and can bring data from diverse and third-party tools into one common environment, creating the integrated asset model. One can then apply engineering and business applications to identify assetwide improvement opportunities.

A portal environment can be built that will allow multiple asset models to be visualized, compared, and analyzed with distributed data sharing and events management, via proven and accepted enterprise platform message bus technologies. Key visualization and workflow technologies required include look-forward analytics; production scorecards; workflow management; production reporting; capital planning; and scheduling that can be performed on a uniform, companywide, basis and allow for rapid, informed decision making.

20.10 OPTIMIZATION

Maximizing the profit contribution of natural gas processing facilities is challenging given the fluctuating economics, changing ambient conditions, and feed variations that processors must address.

New dynamic markets for gas components lead to a need for stronger analytical capabilities of decision support tools. Also, the supply situation becomes more variable as the gas companies respond to market opportunities. The result is a more rapidly changing environment, and as a consequence, processing plants needs to be reconfigured more often. There is a need for better understanding on how to make plantwide production plans and implement them through process management.

Decision support tools have to combine both optimization and simulation capabilities in order to analyze the consequence of different scenarios. Advanced modeling and optimization are needed in order to address the following new challenges in plant control and optimization systems, which have an impact on the optimal design of gas processing plants:

- Many of the processing plants are short of capacity. It is more important than before to maximize flow through the plants or the profit from final products.
- New opportunities exist because of recent advances in modern control technology; e.g., model predictive control lifts the level of automation and gives better opportunities for process optimization systems at higher levels.
- Tasks and systems in operation support centers are integrated. E-fields give a new perspective on operation of the oil and gas production, and the gas processing plants must prepare for this mode of operation.
- The Man-Technology-Organization (MTO) perspective provides relevant information for the personnel involved.

- The use of models for planning and process control plays a central role.
- There is a need to model the dynamics of the markets and incorporate it in contingent plans. The plans should reflect capacities and production possibilities of the plant. This again depends on the design and operation of the plant control system.
- There is a need to structure the information flow between layers and systems in the process control hierarchy, the real-time optimization level, and the planning and scheduling levels.
- Self-regulating and robust systems optimize the other parameters.
- Limitations are imposed by the process design.

20.10.1 Tools for Optimization

Many of the tools mentioned in Chapter 16 could become integral elements of the operations strategy for maximizing profitability. A steady-state process model is necessary to determine the capability and current performance of the operation. Ideally, a dynamic model would be available to train operators and investigate control options. These models can then be used in a real-time mode to report the capability of the plant under different conditions.

Predictive equipment models can be used to determine when maintenance will yield long-term benefits that more than offset the short-term costs. These models can be an extension of existing steady-state models. These steady-state models can also be key elements of an online optimization strategy.

Real-time control models can be constructed from perturbations of a high-fidelity dynamic model. It is necessary to validate these models constructed from models against actual plant data. Often the control models require detuning to provide adequate and robust control.

20.10.2 Optimization Alternatives

Some optimization alternatives for natural gas processing plants include the following:

- Advanced regulatory control
- Multivariable predictive control
- Neural network controllers
- Offline process simulators
- Online sequential simulation
- Online equation-based optimization
- Linear programs
- Web-based optimization

Advanced regulatory control and multivariable predictive control are discussed in Chapter 16.

Neural network-based controllers are similar to multivariable controls except that they gather plant data from the DCS and use the data to "learn" the process. Neural network controllers are said to handle nonlinearities better than multivariable controllers and are less expensive to commission and maintain.

Neural network-based models are valid only within the range of data in which they were trained. Conditions outside the range of data that were used to train a neural network model may be suspect. For example, if a neural network model was trained on data collected while a Joule-Thomson (JT) valve was closed, then the predictions from the model are invalid when the JT valve is open.

Refining and chemical companies have attempted to use neural networks for control many times since the mid-1990s. The technology hasn't proven to be viable compared to the other approaches, such as advanced regulatory control and multivariable control.

Offline process simulators are used to develop a rigorous steady-state or dynamic model of the process. They are used by process engineering personnel to design and troubleshoot processes. Offline simulations allow for what-if case studies to evaluate process enhancements and expansion opportunities.

Offline simulators are typically not used to support daily operational decisions. They must be updated and calibrated to actual plant conditions for every use. They are not as robust as equation-based optimizers and can have difficulty converging large problems reliably and quickly.

A few of the offline simulation companies offer an inexpensive, sequential-based optimization system. The optimizer is based on a rigorous steady-state model of the process and is typically less expensive than equation-based systems. These systems leverage the work done to develop the offline model for online purposes. The extended convergence times inherent in these systems bring into question the robustness of the technology. The sequential nature of the solving technology also can limit the scope of the systems. These systems require hardware and software to be purchased, installed, commissioned, and maintained on-site and require specialized resources to support them.

Equation-based optimizers use a rigorous steady-state model of the process as the basis for optimization and include an automatic calibration of the model with each optimization run. The equation-based solving technology allows optimizers to execute quickly and robustly, making them viable for larger-scale problems, i.e., multiplant load optimization for plants on a common gathering system.

Equation-based optimizers require a hardware platform, a costly software component, and highly specialized engineering services to install, commission, and maintain the technology. Closed-loop implementation requires a multivariable controller to be installed to effectively achieve the optimal targets.

Online equation-based optimizers, when coupled with a multivariable controller, represent the standard in optimization technology for the refining and petrochemical industries. Most refiners and petrochemical companies are rapidly deploying this technology to improve the profit contribution of their larger-scale processing facilities. Unfortunately for gas processors, this technology is justifiable only for very large gas processing facilities and is not scalable across their asset base.

Linear programs are used for evaluating feed and supply chain options. Linear programs are offline tools that allow for what-if case studies and evaluation of supply chain alternatives. They are relatively inexpensive. Linear programs provide a linear representation of the plant process and do not provide guidance for operators (Roop et al., 2002).

Web-based optimization has been applied to cost-effectively supply equation-based optimization to the natural gas processing industry. However, the time lag in collecting data, calculating the optimum and providing advice to the operators to implement the advice may not be fast enough to keep up with the constantly changing conditions experienced in natural gas processing plants.

Another option is online performance monitoring tools that predict the optimum operating point under all conditions. These tools usually provide dashboards and graphical indicators to show the operator the gap between current plant performance and optimum performance. They may or may not provide advice on what parameters to change to reach optimum performance.

20.11 INDUSTRIAL RELEVANCE

In the upcoming decade, there will be large investments in gas production and in facilities for transportation and processing. Optimal utilization of all these facilities is vital in order to maximize the value from produced natural gas. Advanced process control and operation have raised the level of automation in the process industry significantly in the last decades. For example, in the refinery industry, methods such as model-based predictive control (MPC) and real-time optimization (RTO) have become widely used

(Qin and Badgwell, 2003). The focus on these technologies has given large benefits to the industry in the form of increased throughput and more robust operation. The improvements obtained by the use of better control and decision support tools end up directly on the bottom line for the operating companies. The substance of these tools is, in fact, software realization of process knowledge, control, and optimization methods. MPC and RTO are now more or less off-the-shelf products, although for complex processes, the adoption of this technology requires specialist competence. However, there are significant potentials for further improvements in this area. One challenge for the gas processing plants is being able to quickly adapt the plant operation to dynamic changes in the markets; thus, the plant flexibility and ability to perform rapid production changes become more important. It is also required to know accurately the plant production capability, both on very short-term (today–tomorrow) and on weekly, monthly, and even longer horizons. And this requires use of advanced optimization tools and efficient process calculations. It is important to consider these issues related to dynamic operation also in the design of new processing plants and for modification projects. Investments in this type of project have typically short payback time (Moen, 2004).

The combination of individual units into an integrated plant gives a large-scale control problem that is more than just the sum of the units. Cross-connections, bypasses, and recycling of streams give more flexibility, but at the same time, the operation becomes significantly more complicated, and it is almost impossible to utilize the full potential of a complex plant without computer-based decision support tools. Thus, there is a need for development of new decision support tools that combine optimization technology; realize process calculation models at a suitable level of speed and accuracy; and structure the information flow, both from the process measurements and deduced variables and from the support tool down to the manipulated variables in the control system. Further developing the methodology related to plantwide control in this context is also needed (Skogestad, 2004).

20.12 THE TECHNOLOGY INTEGRATION CHALLENGE

There are three likely scenarios of the deployment of the integrated gas plant:

Scenario 1: Business as Usual. Digital technology, information, data, and models are used in an incremental way to reduce costs, increase recovery, and improve production, but no fundamental changes are made in business models, competitive strategies, or structural relationships.

Scenario 2: Visionary. Those who can best adopt and apply digital technologies and concepts will use them to gain significant competitive advantage. This will involve significant investment in software and IT technologies along with culture and management change. The industry will need to demand leaders in the highly technical processes, and modeling software will contribute new and innovative solutions.

Scenario 3: Symbiotic Relationships. There are those who use the availability of turnkey solutions to optimize production and leverage into larger industry positions. This will involve the use of third-party technology consultants and solution partners in an unprecedented way.

For most companies, an evolution approach that blends all three scenarios may well be the chosen route.

The foundation of the integrated gas plant of the future is engineering simulation, with integrated asset models and portfolio views of the business built on top. At the heart of the digital revolution in the upstream energy industry is a shift from historic, calendar-based, serial processes to real-time, parallel processes for finding, developing, and producing oil and gas assets. Real-time data streams, combined with breakthrough software applications and ever-faster computers, are allowing the creation of dynamic, fast-feedback models. These dynamic models, running in conjunction with remote sensors, intelligent wells, and automated production and facility controls, will allow operators to visualize, like never before, what is happening in the facility and accurately predict what needs to happen next to maximize production and efficiently manage field development.

20.13 SCIENTIFIC APPROACH

Optimal operation of gas processing plants is a challenging multidisciplinary task. Large-scale process optimization is challenging in itself. Thus, when we also consider dynamic conditions in the market and on the supply side, the operation will most certainly run into problems that must be solved. Some will arise from the size of the problem, some from complex process behavior and from requirements to solutions of complex optimization problems.

The starting point in this project is the need for decision support as seen from the personnel in a plant operating company. This defines a set of tasks that requires optimization calculations, process calculations, and measurement data handling. The personnel in question can be plant operators, production planners, sales personnel, maintenance planners, process engineers, management, etc. Experience from other applications such as gas transportation will be utilized (Rømo et al., 2003).

The inclusion of market factors, capacity planning, and scheduling must be focused because this sets new requirements to vertical integration of process control and the optimization layers. We may, in fact, have several such layers, where, for example, the classical real-time optimization is just one element. In planning and capacity assessment, the RTO layer may be accessed by superior layers in order to compute the optimal process targets over a certain horizon.

The requirements to process calculations for each type of task must be classified. This may result in a set of optimization problems with different properties and requirements to solution approaches and to the underlying process calculations and data handling; e.g., one approach from the planning side is to start with empty or extremely simple process models and to refine the models based on the requirements to the planning.

Segmentation into suitable process sections and control hierarchies are central issues. Here, we can apply methods from plantwide control in order to structure the control of the plant units in such a way that the influences from unknown disturbances and model uncertainties are minimized (Skogestad, 2004). A very important output in the first phase is to define high-level targets for the process control. The next important issue is to develop methods to select the variables that should be exchanged between the optimization and the process control layers. This is a control structure design task in which the focus is on selecting the variables that are best suited for setpoint control in order to fulfill the process optimization targets in the presence of unknown disturbances and model parameters and measurement errors. Segmentation of the control into suitable sections and layers is also a part of this task.

Recent advances in process control technology also give a new perspective. For example, with an active MPC controller, information about active and inactive process constraints is high-level information that can be exchanged with the optimization layer instead of representing the constraint equations at the optimization layer.

Efficient use of models is such a wide area that this issue can be subject to extended research in separate programs. For example, in process design, it is industrial practice to use quite detailed process models, including rigorous thermodynamics and representation of detailed phenomena within each process unit. In operator training simulators, detailed dynamic models are used, but these are rarely the same models as used in design, and the built-in process knowledge in the form of model configurations and parameters is usually not interchangeable because of different modeling approaches and different model data representation. The models used in MPCs are normally captured from experiments on the process itself and are not connected to the other two types of models. For RTO, steady-state models are normally used, and in some cases, model tools with rigorous models are used there, too. For capacity assessment, correct representation of potential bottlenecks is important.

20.14 OTHER MISCELLANEOUS INITIATIVES

Maintenance management, field information handling, work process optimization, compensation design, and procurement initiatives are several of the current gas processing management initiatives.

In field information projects, companies considering upgrade should understand that technical support must also be upgraded and that care should be taken to select systems with an eye toward ensuring ongoing availability of support over a reasonable period.

In an effort to determine how they are doing against the competition as well as discovering new areas for potential, some processors have become involved in industry benchmarking activities. Benchmarking tools with a reasonable level of analytical content provide benchmarking against a select peer group as well as individual analysis of various cost components. Most organizations that persevere in the benchmarking process and are diligent to follow up on findings testify that benchmarking is a useful tool.

A company that fully utilizes second-wave technologies to streamline its back office and process support technologies could reap a reduction in selling, general, and administrative (SG&A) costs in the range of 8% to 10%. For a typical, large firm, SG&A costs represent an estimated 10% of the total enterprise costs. Thus, a 10% reduction in SG&A outlays would reduce overall corporate spending by 1%—a major gain, given that these savings would drop to the bottom line.

20.15 CONCLUSION

The goal of an integrated operations environment is to enable direct translation of management strategies into manufacturing performance. The vision is that

- Utilization of raw materials is optimal.
- Overall margin and yield of product(s) are maximized.

- Planning, operational, and monitoring cycles are fully integrated.
- Identification and correction of problems occur rapidly.
- Operational (short- and long-term) factors are fully understood.
- The workforce is well informed and aligned for a common purpose.

The system envisioned is an integrated platform for computing and information processing at the production level. It is built around the premise that information is not to be isolated and that better information, when made widely available, will help people operate the facility closer to the optimum. A key principle is to empower employees to maximize the value of their activities. The production management system provides the tools to help personnel do their job better.

A fully functional system will enable the quality cycle of planning, measuring, analyzing, correcting, and then planning again. To make improvements, the staff must be able to see and measure progress.

The integrated production/management system provides the data and the means to analyze situations, define solutions, and track progress. It is an integrated platform of computers, networks, and applications. It brings together the many individual automated systems that exist today and fills any gaps to bring the overall system to a high level of performance. The production/management system spans the gulf between process control and corporate business systems to support the day-to-day operations.

To achieve these goals, the production management system must do the following:

- Be a single, comprehensive source of real-time and business data addressing all operations and available to all appropriate personnel. This means providing long-term storage of all data (e.g., historical process, laboratory, plan, production, and shipment data), merging of these data, and retrieving data.
- Provide information retrieval tools for the full range of users. This usually means highly graphical tools that provide ease of use and a consistent look and feel to minimize the burden of finding and accessing information.
- Integrate a wide range of computer systems and applications. No one system or set of tools will provide all the functionality needed. Instead, the production management system should allow for the use of the best products from different vendors.
- Provide standard screens and reports that focus attention on problems and opportunities. The system should report by exception, highlighting the unusual, the exceptions, and the opportunities. It should compare actual results with the established plans and economic KPIs.

- Maximize information content while minimizing data volume. This is achieved by the use of performance indices and other numerical, measurable indicators, and the presentation of this information in graphical form whenever possible.
- Present operational data in economic terms whenever possible. Opportunities, problems, and deviations from an operating plan should be prioritized based on their impact on overall profitability and, where possible, include an indication as to the action or followup activity that alleviates the deviation.
- Provide analytical tools that enable users to explore and pursue their own ideas. Much of the value of integrated operations comes not from presenting data about current operations, but from people looking for ways to improve current operations.
- Facilitate plantwide communications and work flow. For instance, plans and economic KPIs set in the planning group should flow automatically to operations to help operate the plant. Beyond the manufacturing issues, implementing a project with the scale and complexity of an integrated production/management system creates several management of technology issues.

These create the need to do the following:

- Balance the selection of individual applications with the need to integrate applications across departmental boundaries.
- Provide a single, accessible look and feel. This is particularly important for users accessing data that originate in systems that belong to other departments (“single pane of glass”).
- Employ the latest proven information system technology as it becomes available and at the pace the operator can assimilate and manage.

Integration is a true example of the total being greater than the sum of the parts. A gas plant can profoundly affect the nature, quality, and profitability of its operations throughout the life of the gas plant with a truly integrated production management system.

Several operations have adopted some or all of the discussed strategies to improve their performance. A couple of these plants were discussed in Kennedy et al. (2002). Other examples include gas processing operations in Tunisia, Norway, Nigeria, and Indonesia.

The future seems to belong to those who will be able to mix vision, intelligence, and understanding of human nature, technology, and the processing business into a formula for success in the new world of natural gas gathering and processing!

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CHAPTER 21

Gas Plant Project Management

21.1 INTRODUCTION

Project management is the application of knowledge, skills, tools, and techniques to project activities in order to meet or exceed stakeholder needs and expectations of a project. The project manager, sometimes referred to as the project coordinator or leader, coordinates project activities on a day-to-day basis. This is an ongoing challenge that requires an understanding of the broader contextual environment of the project and the ability to balance conflicting demands among (1) available resources and expectations (especially with respect to quality, time, and cost); (2) differing stakeholder priorities; (3) identified needs and project scope; and (4) quality and quantity of the project's deliverables. Project management for engineering and construction projects requires the application of principles and techniques of project management from the feasibility study through design and construction to completion. Good project management during the early stages of project development greatly influences the achievement of quality, cost, and schedule.

This chapter covers many aspects of managing capital projects in the gas processing business. For the most part, best practice management for gas plant projects follows generic project management principles applicable to most industrial engineering and construction projects. This chapter reviews many of the standard and accepted practices that lead to successful installations as well as some of the unique considerations for gas plant projects, which arise from relatively complex processes employed in typically remote locations.

21.2 PROJECT MANAGEMENT OVERVIEW

One or more parties can perform the design and/or completion of a project. Regardless of the method that is used to handle a project, the management of a project generally follows these steps:

Step 1: Project definition—Determine the conceptual configurations and components to meet the intended use.

Step 2: Project scope—Identify the tasks that must be performed to fulfill the project definition. This step also clarifies what the project does not include.

Step 3: Project budgeting—Define the permissible budget plus contingencies to match the project definition and scope.

Step 4: Project planning—Determine the strategy and tasks to accomplish the work.

Step 5: Project scheduling—Formalize the product of planning.

Step 6: Project execution and tracking—Complete project tasks and measure work, time, and costs that are expended to ensure that the project is progressing as planned.

Step 7: Project closeout—Final testing, inspection, and payment upon owner satisfaction.

Successful projects require effective management, which means (1) clear objectives, (2) a good project plan, (3) excellent communication, (4) a controlled scope, and (5) stakeholder support. Project management in today's organizations demands multiskilled persons who can handle and manage far more than their predecessors and requires competencies that span all of the critical management fields.

21.3 INDUSTRY PERSPECTIVE

Gas plant project management includes the planning, design, engineering, construction, and commissioning of the plant. Key elements are covered under the general headings of engineering, procurement, and construction (EPC). Today, in many large organizations, there has been a trend away from the owner company performing the whole project management function toward the delegation of EPC in part or whole to engineering consultant organizations. At the same time, smaller companies have almost always subcontracted EPC activities. Most companies, however, specify and procure the major equipment themselves (or at least oversee those activities) to assure technical compatibility and adequate lead times for delivery, and so on. For example, the compressors and drivers will be specified and selected by the owner company as a priority.

A successful project in the gas processing industry is not only one that is profitable, but also one that leads to safe, reliable, predictable, stable, and environmentally friendly operational characteristics.

Gas processing is a service to the oil and gas production business. Oil and gas operations desire to produce into a system that has high availability and can produce salable product in a safe, quality-consistent, and environmentally friendly manner. Gas processors must, of course, provide this service in a profitable manner. Flexibility to operate in various modes to respond to the markets and provide various processing alternatives should provide competitive advantages for a plant.

In many cases, the gas processing facilities are owned by the oil and gas producers first as facilities to enable the production of oil and gas and second as value-added operations. In some cases, particularly the present state of the business in the United States, independent gas processors compete for gas that can be produced into a number of gas gathering systems. Several types of contracts exist, but the prevalent contract type is a "percent of proceeds" arrangement. Processors who can offer the greatest revenue to the oil and gas producers have an advantage in any case. The key is recovery of the greatest percentage of feedstock at the highest availability and at the lowest cost.

Adoption of integrated plant design and engineering is allowing gas processors, licensors, as well as engineering and construction firms to streamline workflows to improve efficiency and execute projects faster (Mullick and Dhole, 2007). Integrated models and toolsets allow rapid learning and ensure reuse of information and knowledge throughout the life cycle of the operation.

Engineering and construction firms are also utilizing their information technology infrastructure to support collaborative engineering environments to manage and execute projects around the clock and across the globe. This global execution capability allows companies to fully utilize available talent in a cost-effective manner and to improve project schedules.

Rapid deployment of new engineering tools across the organization through a "virtualized" environment is another emerging trend.

21.4 THE PROJECT MANAGEMENT PROCESS

At the onset of a project, the owner company will undertake the required economic and business analysis regarding new or expanded facilities in order to receive board approval and budget allocation. Based on this approval and funding, the project definition will be refined. The owner company will initiate the project and set out design objectives, usually embodied in the design basis memorandum (DBM), which will lay down the operating parameters and any key design guidelines and specifications. The owner company will solicit bids from EPC contractors and from these will select the successful bidder. In some cases, owner companies will have a partnership with an EPC contractor for certain types and sizes of projects. Pricing will be prenegotiated according to a range of possible contract models (fixed price, cost plus, risk sharing, etc.) or performed on a reimbursable basis. Both parties, owner and contractor, will set up teams to do the work. The EPC company may be asked to be completely responsible for all aspects of engineering, procurement, and construction or may be required only to do engineering and some procurement with the construction carried out by another company. Estimates and schedules will be set up by the EPC company in consultation with the owner. EPC companies can be large, such as Bechtel, or small, such as Cimarron Engineering or Tartan in Calgary. Each company may have somewhat different expertise and experience, but they tend to operate in very similar ways. Consultants and contract personnel will fill areas where the EPC company lacks resources or expertise. The owner also embeds staff in the contractor's company to ensure oversight of the management process and also may include specialist engineering personnel to monitor progress against plan, including quality assurance as well as training of the owner staff on major projects.

For more complex, large projects that involve elements of innovation or requirements to build a gas facility in a region or country where such projects have not previously been conducted, it is not unusual for the process to include feasibility studies and front-end engineering and design (FEED) studies conducted prior to awarding the EPC contract. The feasibility and FEED studies usually are conducted by competent engineering consulting companies, capable themselves of conducting the EPC work. The deliverable from a FEED study may form the basis of a competitive tender for the EPC contract in which a number of prequalified contractors may compete.

A successful project requires the owner and EPC company to work very closely together. Typically, the owner will bring in operations staff at an early stage in the project to ensure that these staff members contribute to the project specifications and review deliverables as they unfold. In this way, commissioning and operation proceed without serious problems. Companies with experienced staff often prefer to have the operating people closely integrated with the construction work from the outset. Similarly, with the controls people, it is essential that the owner's control philosophy is conveyed to the engineers of the contractor's company and that the machinery suppliers are brought into the loop so that engineering specifications and machine products reflect these perspectives. The owner company will undertake the required economic and business analysis regarding new or expanded facilities to receive board approval and budget allocation. From this, the project will be defined.

21.4.1 Defining Business and Project Objectives

The first step in the project management process is to align the business and project objectives. A project can be installed on time and on budget, but if it does not meet the defined business objects, then the project cannot be deemed a success. Following are some of the questions that the business owners must be asked by the project team:

- How much gas is available for processing (ultimate reserves and daily deliverable quantities)?
- What is the market demand for gas and gas products that can be met by this project?
- What is a realistic gas production schedule?
- What are the production pressure, temperature, and composition of the gas?
- How will the gas pressure, temperature, and composition change over time?
- What products can be sold and at what price?
- What are the product specifications?
- How will the products be delivered to market?
- What are the local environmental policies?
- What are the local safety policies?
- What infrastructure—such as roads, bridges, loading and unloading facilities, and personnel housing—is required?
- What is the skill level of available operations and maintenance personnel?

Since most gas processing plants are services to the oil and gas producers, then collaboration between reservoir and production engineers and gas marketers to obtain answers to these questions is imperative. In cases in which processors compete, then collaboration with those responsible for obtaining the processing and sales contracts (e.g., economists, lawyers, and negotiators) can be critical.

21.4.1.1 The Project Charter

The project charter is a document that demonstrates management support for the project. In particular, it authorizes the project manager to lead the project and allocate resources as required. It simply states the name and purpose of the project, the project manager's name, and a statement of support by management. Senior managers of the responsible organization and the partner organizations sign it. The project charter should be distributed widely—to anyone with an interest in the project. This will help build momentum, encourage questions and concerns early in the project's evolution, reinforce the project manager's authority, and possibly draw other interested and valuable stakeholders into the project.

The project owner may be a joint venture of oil and gas companies, with one designated as project operator. The project charter is then usually signed off by all joint venture partners together with an authority for expenditure (AFE) approving the project budget and/or initial stages of expenditure.

21.4.1.2 Project Team Roles and Responsibilities

Project team size and makeup are dependent on the complexity of the project; however, the basic composition of the project team and its responsibilities is recommended for all projects.

1. **Project Manager.** The project manager is responsible for project development, developing schedule, budget, and deliverables definitions; evaluating alternatives; determining return on investment; adhering to company policies; obtaining funding; acquiring internal and external project resources; selecting contractors; maintaining project schedules and budgets; evaluating quality of project deliverables as they evolve; identifying and mitigating downside risks; identifying and exploiting upside opportunities; reporting to business owners; and creating project closeout reports.
2. **Business Owner Representative.** The business owner representative is responsible for assuring that the project adheres to business objectives because objectives may change or require alteration during the project.
3. **Plant Manager.** A plant manager should be appointed as early as possible to address operability and maintainability issues.
4. **Project Engineer/Construction Engineer/Start-Up Engineer.** A project/construction/start-up engineer can be one role on smaller projects and multiple roles in larger projects. This engineer (or engineers) is responsible for addressing technical specifications for contract bidding purposes, performing technical evaluation of contractor bids, acting as the owner's representative during construction, managing construction inspectors, handling turnover of facility to operations, training operators, determining plant performance, and identifying any project deficiencies.
5. **Purchasing Representative.** The purchasing representative is responsible for commercial evaluation of contractor bids and negotiation of contracts.
6. **Process Engineer.** A process engineer is recommended for evaluation of alternative processing schemes during project development,

assistance with technical specifications, evaluation of contractor bids, and assistance with operator training and start-up issues.

7. **Environmental Engineer.** An environmental engineer is recommended to review and provide advice on environmental issues encountered during the project including addressing technical specifications and obtaining environmental permits.
8. **Safety Engineer.** A safety engineer is recommended to review and advise on safety issues encountered during the project including addressing technical specifications and participating on hazard analysis evaluations.
9. **Production or Reservoir Engineer.** A production or reservoir engineer is recommended to be available to evaluate any oil and gas production issues that may be encountered during the project.
10. **Facilities Planner.** For larger projects, a facilities planner should be available to assist with project economics and to serve as a liaison for economic premises and marketing issues.

21.4.2 Contracting Strategy

There are several alternative contracting stages and strategies. The first stage of contracting may be a front-end engineering design. With this approach, a contract is entered based on the design objectives for an engineering contractor to evaluate process and construction alternatives as well as develop technical specifications for the project. In some cases, the owner's engineers may accomplish the front-end engineering design tasks. The second stage of contracting is for engineering, procurement, and construction services. Either stage can be contracted as a lump sum, fixed-fee price also known as a turnkey project or on a reimbursable basis also known as a time and expense contract. In some cases, more complex contracts, such as risk sharing or gain sharing, will divide risks and rewards more evenly between the contractors and project owners.

21.4.3 Conceptual Estimates and Schedules

Most operating companies have developed estimating tools for budgeting of plants similar to those they are currently operating. Many operating companies have the capabilities and resources to evaluate alternative process and mechanical designs with budgetary or conceptual-level estimates. These estimates typically have an accuracy of +/− 30% to 40%.

Under other circumstances, such as when proprietary processes are in use, unique locations are to be selected, or there is a lack of available resources, an engineering firm may be hired to evaluate alternatives and determine budgetary estimates. After evaluation and selection of a conceptual process and mechanical design, the operating or engineering company will undertake a front-end engineering design. The detailed specifications and request for proposals will be the deliverable from the front-end engineering design.

Conceptual estimates and schedules should take the following into consideration:

- Location
- Operators and operability
- Constructability
- Special materials

The availability of fresh water and electricity are considerations in determining location. Port facilities, roadways, and waterways are other considerations. A qualified and available workforce is always a consideration when determining location. In some locations, qualified operating personnel are difficult to find, so inexperienced and poorly educated operators may be hired. In order to overcome their lack of qualifications, these operators require intensive training. Generally, it is good to include in the project training using high-fidelity simulators, particularly where inexperienced operators are to be hired. In addition, plants with novel processes with which even experienced and highly educated operators are not familiar should include additional training provisions. Such training will impact the project's cost and schedule.

Regardless of schedule, the project team's capability to construct the facility must be addressed. For instance, vessels of large diameter and height will require shop facilities that have appropriate size capacity, as well as trucking and rail facilities that accommodate the finished products. In some cases, the vessel may require field fabrication, or multiple vessels will be needed if shop fabricated. Alternatives for prime mover drivers such as electric motors, steam turbines, gas turbines, and gas engines may be influenced by the availability of infrastructure to support these devices. If electrical service is not provided by a utility, then generation or cogeneration facilities may be required. These must be addressed in the project cost estimates and schedules.

Special materials are often required in gas processing plant construction due to components such as hydrogen sulfide, carbon dioxide, mercury, and water. The availability of these materials and their delivery should be considered. Sometimes cladding or linings may be alternatives to expensive and scarce alloys. In addition, approved welding procedures may not be available, or the workforce may not have the expertise to perform certain procedures. Addressing such obstacles must be part of the project plans.

During the proposal solicitation and award of the construction contract, the cost estimates and project schedule for the prime contractor will be focused on construction activities and therefore are fairly detailed and inclusive. However, the overall project schedule from an operating company point of view must consider nonconstruction activities such as permits, licenses, and other government requirements; staffing, accounting, and other internal issues; as well as contracts with suppliers and customers, to name a few.

As project definition improves, so the uncertainties associated with cost estimates should decrease to a funding level of accuracy of approximately +15% to −10% with a 10% contingency identified. A probabilistic approach to cost estimating identifying percentiles (P90, P50, and P10) is also widely used to illustrate cost uncertainties (McIntire, 2001).

21.4.3.1 HAZOP Analysis

A hazards and operability (HAZOP) analysis or equivalent is good practice even when not a statutory requirement. Such an analysis will most likely recommend the addition or deletion of valves, lines, instrumentation, and equipment needed for safe and reliable operation.

Figure 21-1 illustrates a stages and gates approach to oil and gas facilities project management that emphasizes the importance of the planning stages (feasibility, FEED) leading into EPC contracting, construction, and fabrication activity on to HAZOP and ultimately plant commissioning. To move from one stage to another requires a gate to be passed where decisions and approvals have to be made associated with funding, technical design, and project priority. Such approvals are usually structured in the form of authorities for expenditure (AFEs) to be signed off by the project owners (and often other stakeholders—e.g., government authorities) as positive approval to proceed under stages of a project budget. Although, for simplicity, the diagram suggests a linear process proceeding from one stage to another, in practice there are often loops and feedbacks to the work of earlier stages that require adjustments to design, etc.

21.4.3.2 Scheduling and Cost Estimating Software

Software exists to assist with both conceptual cost and time estimates as well as detailed estimates and complex project networks involving the optimization

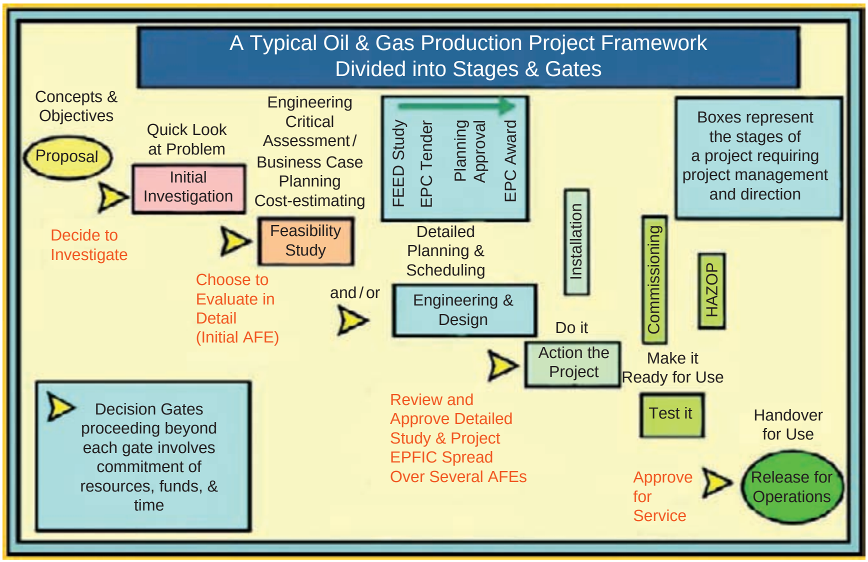

Figure 21-1 Stages and gates approach to project management (Wood, 2001). *(Figure created by David Wood & Associates.)*

of project networks with critical path analysis. Most major engineering, procurement, and construction contractors have their own custom tools. Smaller contractors and operating companies may use products supplied by a vendor specializing in these tools. For larger projects, it is increasingly common for Monte Carlo simulation analysis to be used in conjunction with critical path identification to yield probabilistic estimates of cost and time associated with each project activity and for the project as a whole (Wood, 2001).

21.4.4 Project Execution Planning

A project must be planned and tracked against the plan to assure successful execution. A project plan sets the ground rules and states them in a clear fashion. The project plan helps to control and measure progress and helps to deal with any changes that may occur. Previous experience is the best guide for determining the necessary tasks and the time to complete them. Many engineering and operating companies maintain databases that include previous project plans with actual time to completion and costs. To be able to benefit from such an approach requires good quality record keeping both during a project and following its completion.

Although no two projects are identical no matter how similar they appear, these databases of past experience contain valuable information on which to plan. It is necessary to understand any unique requirements that previous projects met and how the current project compares. Some dissimilarities may include the following:

- Location, which impacts the government regulatory bodies, remoteness, cost of labor, etc.
- Makeup of the project team, including expertise, time available, geographic dispersion of team members, cultural differences, and organizational affiliations/loyalties
- Project scope
- Current economic conditions, which affect inflation and employee availability

The project plan should be relevant, understandable, and complete, and it should reflect the size and complexity of the unique project. The project plan should include the following elements (Hauge and Cramer, 2002):

- A project charter
- A project timeline
- A responsibility matrix
- A project plan budget
- Major milestones with target dates
- A risk management strategy

21.4.5 Preproject Planning Measurement

The project objectives, or the measure for project success or failure, are often defined in terms of cost, schedule, and technical performance. In order to serve as a baseline for project execution, measurements should be in place to identify target completion dates, budgets, and expected technical performance. These measures should be included in a system that allows tracking of actual, target, and projected dates and costs with variances highlighted. The technical expectations should be tracked as well and checked for compliance as the project proceeds.

21.4.6 The Responsibility Matrix

Projects are a collaborative effort between a number of individuals and organizations working together toward a common goal. Managing a diverse team, often spread over several locations (and countries), can present some special challenges. A responsibility matrix is a valuable project management tool to help meet these challenges. The matrix ensures that someone accepts responsibility for each major project activity. It also encourages

Table 21-1 Typical Responsibility Matrix[a]

	Project Team Members				
Task	*Contractor*	*Owner/Operator*	*1*	*2*	*3*
1	A	S	P	–	–
2	–	A	S	P	I
3	–	A	S	P	–
4	A	S	P	–	–
5	A	S	P	I	I
6	A	S	P	–	I
7	–	A	S	–	P

[a] S—Sign-off; A—Accountable; P—Primary Responsible; I—Input.

accountability. The responsibility matrix should correspond with the project timeline. An example is shown in Table 21-1.

The left column lists all the required tasks for your project while the team members (e.g., project manager, project engineer, safety engineer, plant manager, and purchasing agent) are listed across the top. A code entered in each cell represents that team member's involvement in the task in that row. For example, choose codes appropriate to the project; the key is to clearly identify who has a role in every activity, who is accountable, and who must sign off. Make sure the matrix is included in the project plan so that all participants are clearly aware of their responsibilities.

When the construction phase site staffing plan is developed, the contractor's site organization should be analyzed for strengths and weakness that can affect the success of the project. In some cases, it is necessary to supplement the contractor's organization with the project sponsor's own personnel or specialist independent contractors. The need for these adjustments generally is not known until a vulnerability assessment is performed on the contractor. Normally, this is done during the evaluation phase in advance of the final project investment decision and award of the main EPC contracts. Upon completion of such an assessment, project sponsor staffing adjustments should be made to supplement and leverage the contractor's weaknesses and strengths, respectively. This approach helps to reduce project execution risk and improve constructability performance (Wood et al., 2008a).

For project contracts such as reimbursable contracts where the project sponsor is bearing most of the risk exposure, the project sponsor's site teams are commonly developed and aligned to match the structure and function of the contractor's site team. For lump sum and unit price contracts, the project sponsor's and contractor's site teams generally do not match.

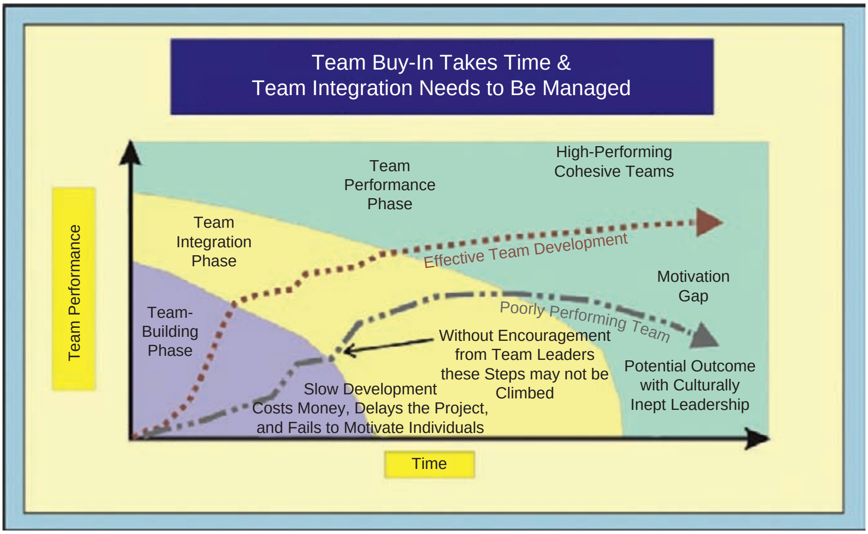

Figure 21-2 Team integration phases (Wood et al., 2011). *(Figure created by David Wood & Associates.)*

A well-disciplined and managed team can make the difference to a project. Figure 21-2 highlights that the time it takes to build a highly performing cohesive project team can be crucial, as project teams do not have time to waste. A focus on early team-building initiatives can pay off as it difficult to assemble a group of skilled staff and expect them to perform as a team without progressing through team-building and integration phases. This becomes even more important in multicultural teams (Safakish and Wood, 2011).

In order to maximize team performance and motivation, team building and development must be given time and priority as well as careful project management attention at an early phase in the execution of any significant project. This is especially the case for mega-scale facilities projects with activities conducted at multiple sites in several countries separated by large time differences and cultural barriers (Safakish and Wood, 2010).

Team members need the following key team attributes to achieve targets by overcoming obstacles:

- They are well trained.
- They are highly motivated.
- They are integrated.
- They are collaborative.
- They are led by a focused project plan.
- They exhibit strong leadership.

- They have shared values obtained through buy-in and alignment.
- They are adequately resourced.

21.5 PROJECT CONTROLS

The two main elements of a project plan are the timeline and cost control. A timeline assures that the project is scheduled properly to meet anticipated and promised dates. For projects that involve several parallel or overlapping activities, the timeline becomes a network. Cost control assures that the project meets its budget.

A checklist approach has been proposed to provide a structured process for incorporating construction and other discipline knowledge that should enhance project safety, quality, schedule, cost, and risk management objectives (Wood et al., 2008a).

This checklist considers 10 key facets of project execution:

1. Schedule
2. Design factors, such as simplicity, standardization, and ease of construction
3. Facilities layouts and arrangements
4. Safety, whether environmental, security, community, or regulatory
5. Specifications
6. Plans and logistics
7. Cost estimates
8. Construction execution issues
9. Modularization
10. Risk mitigation and opportunity exploitation

21.5.1 Project Timeline

By dividing a project into the individual tasks required for completion, the project timeline does the following (Greer, 2002):

- Provides a detailed view of the project's scope.
- Allows monitoring of what has been completed and what remains to be done.
- Allows tracking of labor, time, and costs for each task.
- Allows assigning of responsibility for specific tasks to team members.
- Allows team members to understand how they fit into the "big picture."

This timeline can take on a variety of formats and philosophies. The most prevalent philosophy is to determine the date for project completion and work backward to identify key dates when certain milestones require completion. In practice, the project completion date is first determined through iterations of what is possible going forward with contingencies and identifying an end date. This usually involves establishing a critical path of activities that must be completed on time for a project not to fall behind its scheduled end date. If this end date is not acceptable, then acceleration plans should be explored. Most methods of acceleration require additional expense to accomplish objectives. These methods may include parallel tasking, overtime, and contractor bonuses for meeting an aggressive date, to name a few.

Once the project milestones are set, then subtasks and assignments are identified. On large projects, certain tasks and subtasks will be assigned to an assistant project manager. The overall project manager will become responsible for coordinating the activities of the assistant project managers. The milestones, major tasks, and subtasks are commonly shown in a Gantt chart. This type of presentation presents dependencies in a graphical display such as predecessors (tasks that must be completed prior to commencing the next task) and successors (tasks that cannot be started until a previous task is completed). Milestones are significant events in a project, usually the completion of a major deliverable or tied to vendor or contractor payments. A very good method to analyze subtasks is to consult with an expert on accomplishing the subject task. And the very best way is to receive an estimate with commitment from the person responsible for executing the task. This is where contractors and their bids contribute to the best task analysis possible.

It should be noted that, as the project progresses, there may be tasks that were not foreseeable in the original plan, or tasks are added to enhance the overall project outcome. The effect of the additional tasks, delays, or even acceleration requires consideration of the impact on both the time schedule and resources. When changes to the schedule are warranted and feasible, the project manager should get a written agreement for the revised plan from all the key stakeholders in the project. A regular update of the time schedule is recommended as part of a routine project status report. The time period for such reporting may be weekly or monthly, depending on the project or the stage of the project.

21.5.2 Risk Management

Risk is inherent in all projects. In project management terms, “risk” refers to an uncertain event or condition that has a cause and, if it occurs, has a positive or negative effect on a project’s objectives and a consequence on project cost, schedule, or quality. As discussed in the previous sections, the measure for project success or failure is defined in terms of cost, schedule, and technical performance. Project risk management is intended to increase the likelihood of attaining these project objectives by providing a systematic approach for analyzing, controlling, and documenting identified threats and opportunities both during the planning and execution of a project. The application of project risk management will vary from the operator (owner) or the contractor side. The term "risk management" is used to lump together different activities. These activities may be divided into the following categories (Hauge and Cramer, 2002):

1. Activities related to the day-to-day identification, assessment, and control of uncertainties, i.e., risk management activities related to understanding and controlling the most important risks threatening the achievement of well-defined project objectives. This type of risk management may be based on a qualitative approach in which each risk is assessed separately.
2. Activities related to the periodic assessment of achieving project objectives, i.e., assessing the probability of achieving well-defined project objectives with respect to schedule, budget, or performance. The periodic assessment must use a quantitative approach based on the aggregation of most critical uncertainties.
3. Activities related to the ranking of a set of alternative decision options/ system solutions, i.e., ranking different alternatives with respect to their desirability measured in terms of the corresponding project objectives. Such ranking is typically performed at major decision gates during the conceptual design stage.

A good risk management system identifies, assesses, and controls uncertainties during all phases of the project. This allows the project team to mitigate risks and take advantage of opportunities. The selected system should properly assess all project risks, causes, and consequences including the following issues (Wood et al., 2008b):

- Technical and technological
- Financing and funding
- Joint venture and alliance
- Contractor and vendor performance
- Contract
- Health, safety, and security
- Community and public relations
- Environmental
- Statutory planning and regulation

- Capital and operation costs
- Construction, installation, and assembling schedules
- Completion and commissioning

The preceding risks should be addressed holistically to include all stakeholders and involved parties as well as all phases and all types of project exposures and uncertainties. The system should be developed to properly address all actions necessary to fulfill the requirements of any and all contractual obligations and to incorporate the results from all risk assessments into the requisite project plans.

Risk management is a key to success for project execution but is often constrained by inadequate work processes and tools. An overall understanding of the different risk factors and how these affect the defined project performance goals is critical for successful project management and decision making. Project risk management is a systematic approach for analyzing and managing threats and opportunities associated with a specific project and will increase the likelihood of attaining project objectives. The usage of project risk management will also enhance the understanding of major risk drivers and how these affect the project objectives. Through this insight, the decision makers can develop suitable risk strategies and action plans to manage and mitigate potential project threats and exploit potential project opportunities. Project risk management is based on a number of different analysis techniques. The choice between these techniques is dependent on the quality of the information available and what kinds of decisions project risk management should support. Day-to-day usage of project risk management is typically based on using risk matrices, accounting for both threats and opportunities. With sufficient uncertainty information, the project risk management analysis can be extended to provide more direct decision support through probabilistic cost-benefit analyses.

Gas processing projects are often characterized by large investments, tight time schedules, and the introduction of technology or construction practices under unproven conditions. These challenges can result in a high-risk exposure but also opportunities that bring great rewards.

21.5.2.1 Project Risk Management Methodology

It is important to perform risk management in a structured manner. Indeed, transparent risk management frameworks are becoming statutory requirements for many companies, obliging them to demonstrate how they are managing risks throughout their organizations (enterprise risk management, or ERM) or on an enterprise-wide basis (e.g., the COSO framework in the

United States). It is important to ensure that the project risk management methodology is consistent with ERM frameworks (Wood and Randall, 2005). The project risk management process is often broken down into the following five general steps (Hauge and Cramer, 2002):

1. Initiation and focusing: Initiate the risk management process, including identifying project objectives. The initiation should also assign personnel to the main risk management roles such as risk manager.
2. Uncertainty identification: Identify risks affecting the project objectives. Assign responsibility for assessing and mitigating each risk.
3. Risk analysis: Assess for each risk the probability of its occurring and the corresponding objective consequences given that the risk occurs. Based on the risk assessment, classify each risk in terms of criticality.
4. Action planning: Identify risk-mitigating actions so that the most critical risks are mitigated. Assign responsibility and due dates for each action.
5. Monitoring and control: Review and, if necessary, update risk assessments and corresponding action plans once new and relevant information becomes available.

The "initiation and focusing" step is normally performed once at the start of the project, whereas the other four steps are performed in an iterative manner. The initiation of project risk management in projects has the following set of goals (Hauge and Cramer, 2002):

- Identify, assess, and control risks that threaten the achievement of the defined project objectives, such as schedule, cost targets, and performance of project delivery. These risk management activities should support the day-to-day management of the project as well as contribute to efficient decision making at important decision points.
- Develop and implement a framework, processes, and procedures that ensure the initiation and execution of risk management activities throughout the project.
- Adapt the framework, processes, and procedures so that the interaction with other project processes flows in a seamless and logical manner.

The project risk management process should be assisted by a set of tools that supports these processes.

21.5.2.2 Risk Response Planning

A risk response plan can help maximize the probability and consequences of positive events and minimize the probability and consequences of events adverse to the project objectives. It identifies the risks that might affect the project, determines their effect on the project, and includes responses to each risk. The first step in creating a risk response plan is to identify risks that might affect the project. The project team members should collaborate, referring to the project charter, project timeline, and budget to identify potential risks. Those involved in the project can often identify risks on the basis of experience. Common sources of risk include

- Technical risks, such as unproven technology
- Project management risks, such as a poor allocation of time or resources
- Organizational risks, such as resource conflicts with other activities
- External risks, such as changing priorities in partner or contractor organizations
- Construction risks, such as labor shortages or stoppages and weather

21.5.2.3 Developing Risk Response Strategies

There is no preparation for mitigating all possible risks, but risks with high probability and high impact are likely to merit immediate action. The effectiveness of planning determines whether risk increases or decreases for the project's objectives. Several risk response strategies are available (HRDC, 2003):

- Avoidance—Changing the project plan to eliminate the risk or protect the objectives from its impact. An example of avoidance is using a familiar technology instead of an innovative one.
- Transference—Shifting the management and consequence of the risk to a third party. Risk transfer almost always involves payment of a premium to the party taking on the risk. An example of transference is using a fixed-price contract.
- Mitigation—Reducing the probability and/or consequences of an adverse risk event to an acceptable threshold. Taking early action is more effective than trying to repair the consequences after it has occurred. An example of mitigation is seeking additional project partners to increase the financial resources of the project.
- Acceptance—Deciding not to change the project plan to deal with a risk. Passive acceptance requires no action. Active acceptance may include developing contingency plans for action should the risk occur. An example of active acceptance is creating a list of alternative vendors that can supply materials with little notice.

Since not all risks will be evident at the outset of the project, periodic risk reviews should be scheduled at project team meetings. It is also important not to view risk with a negative mind-set. In addition to the downside consequences associated with many risks lie opportunities, which should be identified and strategies developed to exploit them where possible. Risks that do occur should be documented, along with their response strategies in a risk register that assigns responsibilities for specific risks.

21.5.2.4 Qualitative Project Risk Management

The routine day-to-day identification, assessment, and control of project risks are similar to hazard and operability identification techniques. The identification of risk consists of collecting and examining information on potential events that may influence the achievement of the project objectives. Each such event is categorized as a risk or an opportunity. The identification of these events should involve expertise from all main project competencies to reduce the possibility of important risks being overlooked. These risks will normally be prioritized so that only the most likely and consequential risks will be entered into a formal risk management process. The prioritization of risk should be performed only after thorough assessments and discussions among the project team. New information could mean that risks that have previously been determined as lower likelihood must be inserted into the risk management process.

The assessment of each risk or opportunity is made in terms of scores for probability of occurrence and for consequence, given that it occurs, for each project objective. Based on the probability and consequence scores, the criticality of each risk with respect to achieving the project objectives can be assessed. Typically, a classification consisting of several possible probability scores and several possible consequence scores leads to several different classes of criticality, e.g., "critical," "significant," and "negligible." Figure 21-3 shows an example of how risk can be classified in relation to probability, project cost (economy), project effort or duration (time), and project performance.

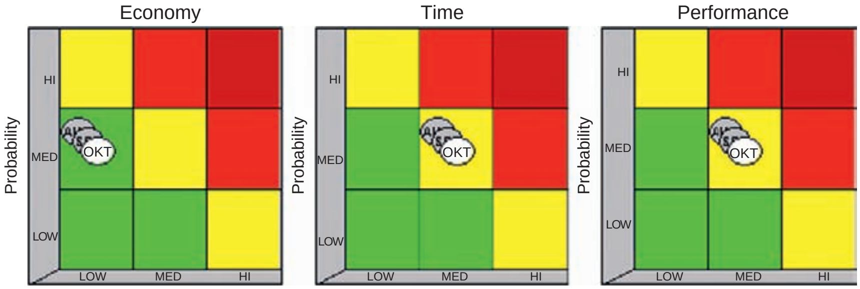

Figure 21-3 Risk matrices (Hauge and Cramer, 2002).

The control of each risk event is normally based on its risk classification. A risk that is classified as "critical" will normally result in actions being identified in order to reduce the risk classification to either "significant" or "negligible." The risk reduction can be caused by either preventive measures (reducing the probability that the event will occur) or corrective measures (lessening the consequences of the event) or both.

21.5.2.5 Quantitative Project Risk Management Assessment

A periodic assessment of the probability of meeting project objectives must be based on quantitative calculation of the aggregated effect of the most important risks on the project objectives. The aggregation must also take into account sequences of scenarios and risk events, as well as the project structures given by the budget, the schedule, and operability. There exist several methodologies with supporting tools that can be used for risk assessment of the total budgets and schedules. The challenge is to apply a methodology and find a tool that supports this methodology so that the integration of risks in the different domains can take place, their mutual independence can be represented, and their aggregate effect on the project objectives can be assessed. Fortunately, the usage of influence diagrams enables such a methodology, and there exist several influence diagram tools. In an influence diagram, each risk is represented as a symbol (or "node") in a graphical diagram. The diagram represents the structural relationship between the different risks and their aggregate effect on the project objectives. In this manner, influence diagrams are well suited to represent risk scenarios. The mathematics used for assessing the aggregate effect is hidden away "behind" the diagram. In this way, the influence diagram also represents a methodology to split a risk management model into two: (1) the structural relationship between the various risks and (2) the mathematics of the risk model, such as probabilistic distribution functions for the risks. Other quantification methods to identify criticality and rank risks include Monte Carlo simulations and presentation through "Tornado diagrams." Tornado diagrams (an example is shown in Figure 21-4) help identify which input parameters, if they were to change, would have most consequential impacts on the analysis. They help to establish materiality of potential outcomes and to illustrate how the project will be impacted by changes, in order or significance, of those selected inputs. This provides valuable insight into which parameters might warrant further investigation to determine how changes would impact the objectives.

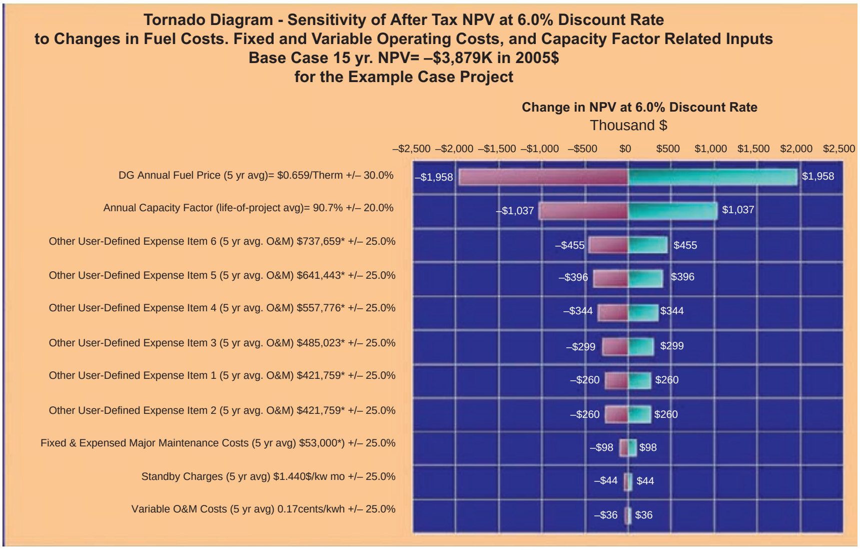

Figure 21-4 Typical Tornado diagram. *(Generated with Palisade @Risk Software.)*

21.5.2.6 Risk Process Modeling

The model of the general risk management process is by no means complete. The two most important omissions are as follows (Hauge and Cramer, 2002):

- No direct representation of the interaction with external organizations and their processes.
- No direct representation of the interaction with other internal processes.

The interaction with other processes should be designed so that the (new) risk management process is integrated as seamlessly as possible into the existing organizations and their already-existing other processes (e.g., enterprise risk management, or ERM, framework). Any required modifications should be as small as possible. Hence, this risk management process is integrated into the organization by having the already-existing weekly management meeting also assessing the weekly risk report. Hence, no new forum for management review of risk is established; only an additional item is added to an existing agenda.

Since different parties in the risk management process have different interests, these interests should correspond to different views in the risk management process. Risk mitigation (and/or exploitation) strategies themselves involve costs to instigate and may lead to secondary risks. It is important to reconcile such costs and secondary risks with the risk management objectives, materiality of the risks being addressed, the project budget, and resources available.

21.5.2.7 *Project Risk Management in Interaction with Other Management Processes*

It is important to note that risk management has major similarities with other common management processes. Examples of such processes are as follows (HRDC, 2003):

- Management of project changes
- Management of public permits
- Management of health, safety, and environmental issues
- Management of decision gates

Similarities exist in the identification of items, the assessment of their criticality, the identification of corresponding mitigating actions, and the follow-up of criticality assessment and corresponding mitigation plans. These similarities should be exploited when setting up the risk management processes and establishing the tools for the management of these processes.

The risk management process will remain constant over the different project phases throughout the life cycle of a construction project. The different risk management techniques for assessing day-to-day risks, calculating the ability to meet defined project objectives, and ranking the different decision alternatives will also remain the same. In the ranking of different investment opportunities for a gas processing project, a number of issues will, in general, need to be looked into:

- Revenue
- Costs (costs of different activities: capital expenditures [CAPEX], operational costs [OPEX])
- Schedule (of project tasks and completion of milestones)
- Taxes and depreciation
- Health, safety, and environment (meeting regulations and company requirements)
- Structural reliability (design that meets requirements)
- On-stream factor (design that meets availability requirements)

What will vary, however, over the life cycle is the quality of the available risk-related information, the kind of competence that is needed to compile and prioritize this information, and the kinds of decisions that are supported by the risk management activities. A believable risk management process must be conducted by personnel with domain knowledge of the project phase in which decisions are to be made. Since the required competence will vary with the project phase, it is unlikely that the same person can fill the risk manager role throughout the project. Whether one or more risk managers or a team is involved in managing risks, it is crucial that analysis, actions, and outcomes are documented in a risk register and widely communicated both within the project team and to project stakeholders.

21.5.2.8 Other Risk Mitigation Concepts

Other methods of risk mitigation include (Hauge and Cramer, 2002):

- Cost overrun protection—Cost overrun insurance can be purchased. This insurance is most often used for infrastructure such as bridges and islands that may be required to access or locate the facility.
- Regulatory risk—A wide-ranging political or regulatory risks form of insurance is available. This is designed to provide an indemnity in the event that any changes occur to the regulatory requirements or political stance perceived at conception of the project during construction and into commercial operations.
- Revenue stream stabilization—In addition to the historic insurance marketplaces, capital markets are available that enable industry to transfer risks to financial vehicles and through structures different from traditional insurance policies. This market convergence has produced a wide range of creatively devised financial products that can be applied to uninsurable risks. Hedging the sale prices for specific volumes of products is now widely used, particularly in the initial years of plant production prior to project payback to help to reduce project finance risks.
- Blended risk solution—This integrated risk solution will result in a more comprehensive package of protection than the traditional set of policies providing limited but specific coverage addressing different areas of risk.

21.6 QUALITY ASSURANCE

An important part of defining the result and performance of the project is the specification of its quality-related features, which the project must then aim to deliver. Quality assurance has been an issue at the forefront of organizational concerns for decades. The development of quality-conscious construction practices has been identified as being of the utmost importance in gaining and retaining a competitive edge. In the context of a project that aims to deliver a complex result, the quality aspects of that result will need to be planned, designed, aspired, and monitored. "Quality assurance" is a term used to incorporate the quality policy, quality management, and quality control functions, which combine to assure that the end result will be consistently achieved to the required condition. Its aim is to attain and assure quality through the adoption of a cost-effective quality control system and through external inspections and audits. Quality planning is an integral part of the planning activity. It manifests itself in the descriptions and in the scheduling of quality-related activities. The results of the quality planning activities are reflected in the resource and technical plans, at each level of the project. Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout their development life cycles. Quality control utilizes measurable quality criteria and is exercised via quality reviews, project reviews, and by the testing of products. Quality assurance requires agreement on the level of quality controls to be adopted, both specifically relating to the project and to the overall organizational policy. It is important that all three interests represented by the project owner are taken into account when deciding the mechanisms to be adopted. The task descriptions should describe the purpose, form, and components of a task. It should also list, or refer to, the quality criteria applicable to that task.

Task descriptions should be created as part of the planning process, to shadow the identification of the tasks that are required by the project. Each task description may either apply to a specific item or to all the tasks of a given type. The component tasks of a complex task may be described in separate descriptions, giving rise to a hierarchy of task descriptions for that task.

Quality criteria should be used to define the characteristics of a task in terms that are quantifiable and therefore allow it to be measured at various points in its development life cycle, if required. The criteria effectively define quality in the context of the product and are used as a benchmark against which to measure the finished task. Quality criteria should be established by considering what the important characteristics of a result or task are in satisfying the need that it addresses, and they should always be stated objectively. Subjective or descriptive criteria such as "quick response" or "maintainable" are unsatisfactory because they do not permit meaningful measurement.

Quality planning should ensure that all quality-related activities are planned and incorporated into the project schedule. The tasks required to ensure the quality of the delivered result are often overlooked, with the result that the project schedule fails to represent quality-related work. This can have serious consequences for either the quality levels achieved, the overall budget, or both.

Quality control is concerned with ensuring that the required qualities are built into all of the tasks throughout the construction cycle. It defines the method of inspection, in-process inspection, and final inspection to determine if the result has met its quality specification. Quality control utilizes measurable quality criteria and is exercised via change control, quality reviews, project reviews, and by the testing of products.

Reviews should be scheduled prior to key decision dates and important milestones such as shipment of rotating equipment and major equipment. For instance, modifications to a gas turbine are more easily made in the shop prior to shipment rather than in the field after transit. Required modifications that could have been found with shop inspections can cause significant delays.

Many of the top engineering consulting companies operate integrated quality, health, safety, and environmental (QHSE) management systems that they apply generically across their operations. This is appropriate because it makes clear to their staff that all four of its components are important to ensure successful outcomes and that the QHSE components all influence each other as well as the budgetary, schedule, and risk issues that drive project decisions. The adage that one should expect what one inspects applies also to gas processing project management. Management and technical peer reviews are good practice. Inspection reports monitor not only quality but also progress and should be used liberally for best results.

21.7 COMMISSIONING AND START-UP

Commissioning and start-up of a new facility or unit are very important phases of any processing project. These activities can be very expensive from a standpoint of project costs as well as deferral of operating revenues, if delays are encountered. For this reason, it is good practice to assign a start-up engineer to plan and coordinate these activities. Operators and maintenance personnel should be hired well before the start-up date and trained on the equipment and process basics. Familiarity with the particular control system can be accomplished through the use of operator training simulators. These simulators can pay off easily by reducing start-up time. More detailed training on particular aspects of the equipment and process can be given to select individuals. Indeed, many now argue that it is important to have operations and maintenance personnel involved in the design and engineering phase of a project as it is much easier to sort out operational control logistics at the planning stage than later on.

Thorough checkout of the equipment should be conducted by the owner and operator of the facility. Punch lists of all deficiencies should be prepared, reviewed with the constructor, and updated on a daily basis. Purging of the equipment is quite important for safe start-up. Process gas or an inert gas, such as nitrogen, can be introduced to remove pockets of oxygen, which could lead to explosions. All high-point vents should be opened and checked with a portable oxygen analyzer. These vents should be closed only when oxygen is no longer detected. Excess water and other liquids should also be removed from the system by opening low-point drains until purge gas escapes through the drains.

Start-up should be a joint effort between the constructor, operator, and process designers. The constructor should be advised of any deficiencies in the equipment and instrumentation as they are identified by operations and have personnel readily available to resolve the issues. It is not uncommon for construction contracts to include a retention portion of the plant cost that the owner will withhold for a period (e.g., 6 months or 1 year) following start-up to motivate the contractor to deal promptly with any teething problems.

21.8 OPERATE AND EVALUATE

The final phase of a processing plant project is continuous operation and evaluation of the project results, including plant capability over a finite period of time. Before the plant is turned over to operation, the performance of the plant should be measured. Sometimes this is required by the contractor or process licensor to meet performance guarantees. The capability of each process and equipment should be calculated and become the baseline for operation. At this point, you may discard the design capability because this was used only as a basis for sizing equipment. Many times, equipment is sized with a contingency, giving better than design capability. Other times, the equipment may not be capable of design. For example, a difference in the actual inlet gas composition from the design basis is a common culprit.

After detailed measurement of the plant's capabilities and deficiencies, these items should be well documented. The capability information is valuable should the plant require expansion in the future, and deficiencies become the basis for possible debottleneck and retrofit projects.

21.9 PROJECT CLOSEOUT

In addition to the evaluation activities mentioned in the previous section, the project activities should be reviewed. This review should comprise the following questions:

- What worked well?
- What didn't work well?
- What were the actual costs?

- What was the actual schedule?
- What assumptions need revision?
- What risks materialized and required attention (as documented in the project risk register)?
- Did risk mitigation strategies employed achieve their objectives?
- What are the project economics as constructed and operated?

As a result of this review or project post mortem, the project manager should write a project closeout report that includes the results of the project closeout review and recommendations for future similar projects.

21.10 CONCLUSION

Planning is critical to project success. Detailed, systematic, team-involved plans are the foundation for such success. When events cause a change to the plan, project managers must make a new one to reflect the changes. Therefore, continuous planning is a requirement of project management. Project managers must focus on three dimensions of project success. Project success means completing all project deliverables on time, within budget, and to a level of quality that is acceptable to sponsors and stakeholders. The project manager must keep the team's attention focused on achieving these broad goals and the stakeholders aligned to the project objectives. It is essential that the project team be composed of all key disciplines that create or use the deliverables. The responsibilities of all team members should be clearly defined. Project managers must feel, and transmit to their team members, a sense of urgency. Because projects are endeavors with limited time, money, and other resources available, they must be kept moving toward completion. Since most team members have lots of other priorities, it's up to the project manager to keep team members' attention on project deliverables and deadlines. Regular status checks, meetings, and reminders are essential. All project deliverables and all project activities must be visualized and communicated in vivid detail. The project manager and project team must create a picture of the finished deliverables in the minds of everyone involved so that all effort is focused in the same direction. Vague descriptions should be avoided, and everyone needs to understand what the final product will be.

Projects require clear approvals and sign-off by sponsors. Clear approval points, accompanied by formal sign-off by sponsors and key stakeholders, should be demarcation points in the evolution of project deliverables. Anyone who has the power to reject or to demand revision of deliverables after they are complete must be required to examine and approve them as they are being built.

Risk management is an essential responsibility of project management. All risks should be identified, and a contingency plan should accompany all critical risks.

A gas processing project is not complete until the plant, unit, or equipment is placed in service. An evaluation of the operability should be a deliverable upon completion of the project.

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APPENDIX 1

Three-Phase Flash Calculation for Hydrocarbon Systems Containing Water

One of the most important engineering problems encountered in modeling chemical and petroleum processes is the multiphase flash problem. The present flash algorithm can be used to calculate phase-equilibria for multicomponent systems with three coexisting phases (water, oil, and gas) at equilibrium with both simplicity and accuracy (Mokhatab, 2003).

This three-phase flash algorithm is based on the thermodynamic condition of equal fugacities for each component in each phase. The resulting phase compositions then provide better values for updating the distribution coefficients, using an equation of state (EOS).

A generalized model of a three-phase equilibrium system is shown in Figure A1-1. The phases are assumed to be in thermodynamic equilibrium with each other, and any component can appear in all three phases.

The overall and component material balance around the flash tank gives

$$F = L_A + L_B + V \tag{A1-1}$$

$$FZ_i = L_A X_{Ai} + L_B X_{Bi} + V_{yi} \tag{A1-2}$$

where F *is* the total moles of feed, lbmole; L_A is the total moles of hydrocarbon-rich liquid, lbmole; L_B is the total moles of water-rich liquid, lbmole; V is the total moles of vapor, lbmole; Z_i is the mole fraction of component *i* in the feed; X_{Ai} is the mole fraction of component *i* in the hydrocarbon-rich liquid phase; X_{Bi} is the mole fraction of component *i* in the water-rich liquid phase; and y_i is the mole fraction of component *i* in the vapor phase.

The relations describing compositions in each phase that must be satisfied are

$$\sum_{i=1}^{n} X_{Ai} = \sum_{i=1}^{n} X_{Bi} = \sum_{i=1}^{n} y_i = 1 \tag{A1-3}$$

where *i* indicates each component; and *n* is number of components.

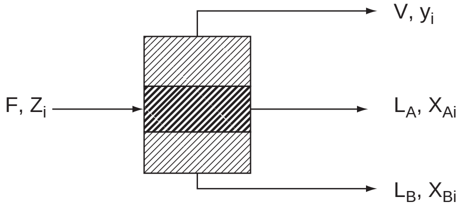

Figure A1-1 Model of a system exhibiting three-phase behavior.

Also, the equilibrium relations between the compositions of each phase are defined by the following expressions (Peng and Robinson, 1976):

$$K_{\mathrm{Ai}} = \frac{y_{\mathrm{i}}}{X_{\mathrm{Ai}}} = \frac{\hat{\varphi}_{\mathrm{i}}^{\mathrm{A}}}{\hat{\varphi}_{\mathrm{i}}^{\mathrm{V}}} \tag{A1-4}$$

$$K_{\mathrm{Bi}} = \frac{y_{\mathrm{i}}}{X_{\mathrm{Bi}}} = \frac{\hat{\varphi}_{\mathrm{i}}^{\mathrm{B}}}{\hat{\varphi}_{\mathrm{i}}^{\mathrm{V}}} \tag{A1-5}$$

where K_{Ai} is the equilibrium ratio of component *i* in the hydrocarbon-rich liquid phase; K_{Bi} is the equilibrium ratio of component i in the water-rich liquid phase; $\hat{\varphi}_{\mathrm{i}}^{\mathrm{A}}$ is the fugacity coefficient of component *i* in the hydrocarbon-rich liquid phase; and $\hat{\varphi}_{\mathrm{i}}^{\mathrm{B}}$ is the fugacity coefficient of component *i* in the water-rich liquid phase.

Combining Equations A1-1 through A1-5 gives the following equations:

$$\sum_{i=1}^{n} x_{\mathrm{Ai}} = \sum_{i=1}^{n} \left[\frac{z_{\mathrm{i}}(1-K_{\mathrm{Ai}})}{\frac{L_{\mathrm{A}}}{F}(1-K_{\mathrm{Ai}}) + \frac{L_{\mathrm{B}}}{F}\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-6}$$

$$\sum_{i=1}^{n} x_{\mathrm{Bi}} = \sum_{i=1}^{n} \left[\frac{z_{\mathrm{i}}K_{\mathrm{Ai}}}{\frac{L_{\mathrm{A}}}{F}(1-K_{\mathrm{Ai}}) + \frac{L_{\mathrm{B}}}{F}\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-7}$$

$$\sum_{i=1}^{n} y_{\mathrm{i}} = \sum_{i=1}^{n} \left[\frac{\mathrm{z}_{\mathrm{i}}}{\frac{L_{\mathrm{A}}}{F}(1-K_{\mathrm{Ai}}) + \left(\frac{L_{\mathrm{B}}}{F}\right)\left(\frac{K_{\mathrm{Ai}}}{K_{\mathrm{Bi}}} - K_{\mathrm{Ai}}\right) - K_{\mathrm{Ai}}} \right] \tag{A1-8}$$

These equations can then be combined to determine the phase and volumetric properties of the three-phase systems.

Peng and Robinson (1980) proposed the following equations for three-phase flash calculations:

$$\sum_{\mathrm{i}=1}^{n} x_{\mathrm{Ai}} - \sum_{i=1}^{n} y_{\mathrm{i}} = 0, \left[\sum_{i=1}^{n} x_{\mathrm{Bi}}\right] - 1 = 0 \tag{A1-9}$$

Provided that the equilibrium ratios and the overall composition are known, the preceding equations can be solved simultaneously by using the modified Rachford and Rice (1952) iterative method. In the course of making phase equilibrium calculations, it is always desirable to provide initial values for the equilibrium ratios so the iterative procedure can proceed as reliably and rapidly as possible. Peng and Robinson adopted Wilson's (1969) equilibrium ratio correlation to provide initial K_{A} -values, as follow:

$$K_{\mathrm{Ai}} = \frac{P_{\mathrm{ci}}}{P} \,\mathrm{EXP}\left[5.3727(1+\omega_{\mathrm{i}})\left(1 - \frac{T_{\mathrm{Ci}}}{T}\right)\right] \tag{A1-10}$$

where P is system pressure, psia; T is system temperature, °F; P_{Ci} is critical pressure of component *i*, psia; T_{Ci} is critical temperature of component *i*, °F; and ω_{i} is acentric factor of component *i*.

For determination of initial K_{B}-values, Peng and Robinson (1980) proposed the following expression:

$$K_{\mathrm{Bi}} = 10^{6}\left[\frac{P_{\mathrm{ci}}.T}{P.T_{\mathrm{ci}}}\right] \tag{A1-11}$$

A logic diagram for three-phase flash calculations is shown in Figure A1-2.

In this method, the fugacity coefficients can be obtained from Peng and Robinson-EOS (1976) for nonpolar compounds. Note that in the association equation of state (AEOS), the compressibility factor of the association component, such as water, is subdivied in two physical and chemical parts as follows (Prausnitz et al., 1986):

$$Z = Z^{\mathrm{ph}} + Z^{\mathrm{ch}} - 1 \tag{A1-12}$$

The physical compressibility factor, Z^{ph}, can be obtained from modified Peng and Robinson-EOS (1980). To evalute Z^{ch} in Equation A1-12, one uses the Shinta and Firoozabadi (1995) association model. Also, the

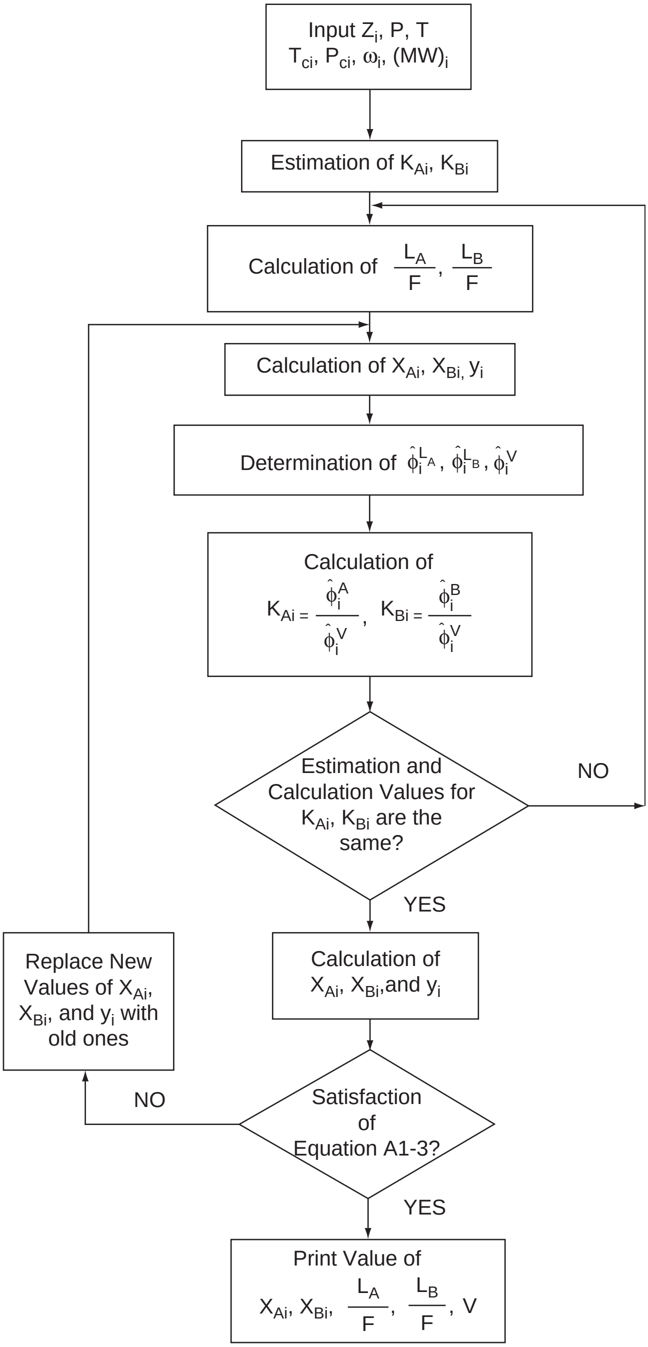

Figure A1-2 Logic diagram for three-phase flash calculations.

fugacity coefficient of an associating component in each phase is the sum of both the chemical and the physical contributions as follows (Anderko, 1991):

$$\ln(\varphi_i Z) = \ln\left(\varphi_i^{ph} Z^{ph}\right) + \ln\left(\varphi_i^{ch} Z^{ch}\right) \tag{A1-13}$$

where chemical fugacity coefficients are obtained by the Shinta and Firoozabadi (1995) association model. Note that in using the AEOS, only binary interaction coefficients between water and hydrocarbon, and nonhydrocarbon components and hydrocarbon cuts are required because calculations are often very sensitive to those. Therefore, in this method, binary interaction coefficients suggested by Nishiumi and Arai (1988), Peng and Robinson (1980), and Shinta and Firoozabadi (1995) are used.

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APPENDIX 2

Conversion Factors

Length

1 m = 39.37 inches = 10^6 μm = 10^{10} A°
1 inch = 2.54 cm
1 ft = 30.48 cm = 0.3048 m
1 mile = 5280 ft = 1760 yds
= 1609.344 m

Mass

1 lbm = 453.6 gr = 0.4536 kg
= 7000 grain
1 kg = 1000 gr = 2.2046 lbm
1 slug = 1 lbf s^2/ft = 32.174 lbm
1 U.S. ton = 2000 lbm
(also called short ton)
1 long ton = 2240 lbm
(also called British ton)
1 tonne = 1000 kg
(also called metric ton)

Force

1 lbf = 4.448 N = 4.448×10^5 dynes
= 32.174 poundals
= 32.174 lbm.ft/s^2
= 1 lbw

Pressure

1 atm = 14.696 psia = 2116 lbf/ft^2
= 29.92 in Hg = 760 mm Hg
= 760 Torr
= 1.013 bar
= 33.9 ft H_2O = 1.013×10^5 Pa
= 101.3 kPa
1 Pa = 1 N/m^2 = 10^{-5} bars

Energy

1 J = 1 W.s = 1 kg m^2/s = 1 N.m^3
= 10^7 dyne.cm = 10^7 erg
1 Btu = 778 ft.lbf = 252 cal
= 1055 J = 10.41 lit.atm
1 HP. hr = 2545 Btu
1 kW. hr = 3412 Btu = 1.341 HP. hr

Power

1 HP = 550 ft.lbf/s = 33000 ft. lbf/min
= 746 W = 0.746 kW

Gas Constant

R = 1.9859 Btu/lbmole °R
= 1.9859cal/grmole. °K
= 0.73024 atm ft^3/lbmole.°R
= 1545.3 ft.lbf/lbmole.°R
= 10.732 psia.ft^3/lbmole. °R
= 0.082057 lit. atm/grmole. °K
= 82.057 atm. cm^3/grmole. °K
= 8314.5 Pa.m^3/kgmole.°K
or J/kg mole.°K
= 8.3145 kJ/kg mole.°K

Dynamic Viscosity

1 cp = 0.01 Poise = 0.01 g/cm s
= 0.01(dyne.s)/cm^2
= 0.001 kg/m.s = 0.001 Pa.s
= 0.001 N.s/m^2
= 2.42 lbm/ft.hr = 0.0752 slug/ft.hr
= 6.72×10^{-4} lbm/ft.s
= 2.09×10^{-5} lbf.s/ft^2
1 Pa.s = 0.0209 lbf.s/ft^2
= 0.672 lbm/ft.s

Volume

1 ft^3 = 7.4805 U.S. gal = 6.23 Imperial gal
= 28.317 lit
1 m^3 = 1000 lit = 264.2 U.S. gal
= 35.31 ft^3
1 bbl = 42 U.S. gal (oil) = 5.615 ft^3
1 lit = 1000 cc

Density

Water = 62.43 lbm/ft^3 = 1000 kg/m^3
= 1 gr/cm^3
= 8.346 lbm/U.S. gal
mercury = 13.6 gr/cm^3

Temperature

°F = 1.8 (°C) + 32
°R = °F + 459.67 = 1.8 (°K)

Kinematic Viscosity

1 St = 1 cm^2/s = 0.0001 m^2/s
1 ft^2/s = 929 St = 0.0929 m^2/s

Force-Mass Conversion Factor

g_c = 1 $kg.m/s^2.N$ = 1 g cm/s^2.dyne
= 32.174 lbm.ft/s^2.lbf = 1 slug.ft/s^2.lbf

Acceleration Due to Gravity

g = 32.2 ft/s^2 = 9.81 m/s^2 = 981 cm/s^2
(varies very slightly with longitude and elevation)

APPENDIX 3

Physical Properties of Fluids

Table A3–1 Properties of Hydrocarbons and Common Gases[*]

Component	Formula	Mol Wt	Boiling Point°F (1 atm)	Vapor Press. @ 100°F, Psia	Critical		Liquid Spec. Grav. 60/60°F	Volume Ratio, Scf gas per gal liquid
					Press. Psia	*Temp. °F*		
Methane	CH_4	16.043	−258.73	(5000)[a]	666.4	−116.67	0.3[a]	59.135[a]
Ethane	C_2H_6	30.070	−127.49	(800)[a]	706.5	89.92	0.35619[c]	37.476[c]
Propane	C_3H_8	44.097	−43.75	188.64	616.0	206.06	0.50699[c]	36.375[c]
iso–Butane	C_4H_{10}	58.123	10.78	72.581	527.9	274.46	0.56287[c]	30.639[c]
n-Butane	C_4H_{10}	58.123	31.08	51.706	550.6	305.62	0.58401[c]	31.790[c]
iso–Pentane	C_5H_{12}	72.150	82.12	20.445	490.4	369.10	0.62470	27.393
n-Pentane	C_5H_{12}	72.150	96.92	15.574	488.6	385.8	0.63112	27.674
n-Hexane	C_6H_{14}	86.177	155.72	4.960	436.9	453.6	0.66383	24.371
n-Heptane	C_7H_{16}	100.204	209.16	1.620	396.8	512.7	0.68820	21.729
n-Octane	C_8H_{18}	114.231	258.21	0.537	360.7	564.2	0.70696	19.580
n-Decane	$C_{10}H_{22}$	142.285	345.48	0.061	305.2	652.0	0.73421	16.326
Nitrogen	N_2	28.013	−320.45	___	493.1	−232.51	0.80940[d]	91.413[c]
Oxygen	O_2	31.999	−297.33	___	731.4	−181.43	1.1421[d]	112.93[c]
Carbon Dioxide	CO_2	44.010	−109.26[b]	___	1071	87.91	0.81802[c]	58.807[c]
Hydrogen Sulfide	H_2S	34.08	−76.50	394.59	1300	212.45	0.80144[c]	74.401[c]
Water	H_2O	18.0115	212.00	0.950	3198	705.16	1.00000[d]	175.62[c]
Air	Mixture	28.9625	−317.8	___	546.9	−221.31	0.87476[d]	95.557[c]

Continued

Table A3–1 Properties of Hydrocarbons and Common Gases*—cont'd

Component	Acentric Factor, ω	Flammability Limits, vol% in Air Mix.		Heating Value @ 60°F 1 atm, Btu/scf [e]		Freezing Point @ 1atm, °F	Heat of Vap'n @ 1 atm, Btu/lb
		Lower	*Upper*	*Net*	*Gross*		
Methane	0.0104	5.0	15.0	909.4	1010.0	−296.44[f]	219.45
Ethane	0.0979	2.9	13.0	1618.7	1769.6	−297.04[f]	211.14
Propane	0.1522	2.0	9.5	2314.9	2516.1	−305.73[f]	183.01
iso–Butane	0.1852	1.8	8.5	3000.4	3251.9	−255.82	157.23
n-Butane	0.1995	1.5	9.0	3010.8	3262.3	−217.05	165.93
iso–Pentane	0.2280	1.3	8.0	3699.0	4000.9	−255.82	147.12
n-Pentane	0.2514	1.4	8.3	3706.9	4008.9	−217.05	153.57
n-Hexane	0.2994	1.1	7.7	4403.8	4755.9	−139.58	143.94
n-Heptane	0.3494	1.0	7.0	5100.8	5502.5	−131.05	136.00
n-Octane	0.3977	0.8	6.5	5796.1	6248.9	−70.18	129.52
n-Decane	0.4898	0.7	5.4	7189.6	7742.9	−21.36	119.65
Nitrogen	0.0372	—	—	—	—	−346.00[f]	85.59
Oxygen	0.0216	—	—	—	—	−361.82[f]	91.59
Carbon Dioxide	0.2667	—	—	—	—	−69.83	246.47
Hydrogen Sulfide	0.0948	4.3	45.5	586.8	637.1	−121.88[f]	235.63
Water	0.3442	—	—	—	—	32.00	970.18
Air	—	—	—	—	—	—	88.20

* *GPSA Engineering Data Book,* 12th Ed., Gas Processors Suppliers Association (GPSA), Tulsa, OK (2004).

[a] Above critical point, extrapolated or estimated.

[b] Sublimation point.

[c] At saturation pressure, 60°F.

[d] At normal boiling point.

[e] Gas at 60°F, liquid at normal boiling point.

[f] At the triple point pressure.

Table A3–2 Physical Data for Common Refrigerants*

Refrigerant	Boiling Point (1 atm.) °F	Critical Temp, °F	Critical Pressure Psia	Freezing Point, °F
R–12, CCl_2F_2	−21.6	233.6	596.9	−252.0
R–11, CCl_3F	74.7	388.4	635.0	−168.0
R–21, $CHCl_2F$	48.0	353.3	750.0	−221.0
R–22, $CHClF_2$	−41.4	204.8	716.0	−256.0
R–113, $CClF_2CCl_2F$	117.6	417.4	495.0	−31.0
R–114, $CClF_2CClF_2$	38.4	294.3	474.0	−137.0
Carbon Dioxide, CO_2	−108.4	37.8	1071	−69.9
Ammonia, NH_3	−28.0	271.2	1651	−108
Methyl Chloride, CH_3Cl	−10.76	289.6	969	−144
Sulfur Dioxide, SO_2	14.0	314.8	1142	−99
Methylene Chloride, CH_2Cl_2	103.7	421	640	−143

*Ludwig, E.E., 1983. Applied Process Design for Chemical and Petrochemical Plants, second edition, Gulf Publishing Company, Houston, TX.

Table A3–3 Approximate Ratio of Specific Heats ("k"-values) for Various Gases[*]

Gas	Symbol	Mol. Wt.	k. @ 14.7 Psia		Density @ 14.7 psi. & 60°F Lbs./cu. ft.
			60°F	*150°F*	
Monatomic	He, Kr, Ne, Hg	—	1.67	—	—
Most Diatomic	O_2, N_2, H_2, etc.	—	1.4	—	—
Acetylene	C_2H_2	26.03	1.3	1.22	0.0688
Air	—	28.97	1.406	1.40	0.0765
Ammonia	NH_3	17.03	1.317	1.29	0.451
Argon	A	—	1.667	—	0.1056
Benzene	C_6H_6	78.0	1.08	1.09	0.2064
Butane	C_4H_{10}	58.1	1.11	1.08	0.1535
Isobutane	C_4H_{10}	58.1	1.11	1.08	0.1578
Butylene	C_4H_8	56.1	1.1	1.09	0.1483
Iso-Butane	C_4H_8	56.1	1.1	1.09	0.1483
Carbon Dioxide	CO_2	44.0	1.3	1.27	0.1164
Carbon Monoxide	CO	28.0	1.4	1.4	0.0741
Carbon Tetrachloride	CCl_4	153.8	1.18	—	0.406
Chlorine	Cl_2	70.9	1.33	—	0.1875
Dichlorodifluoromethane	CCl_2F_2	120.9	1.13	—	—
Dichloromethane	CH_2Cl_2	84.9	1.13	—	0.2215
Ethane	C_2H_6	30.0	1.22	1.17	0.0794
Ethylene	C_2H_4	28.1	1.25	1.21	0.0741
Ethyl Chloride	C_2H_5Cl	64.5	1.13	—	0.1705
Flue Gas	—	—	1.4	—	—
Helium	He	4.0	1.667	—	0.01058
Hexane	C_6H_{14}	88.1	1.08	1.05	0.2276
Heptane	C_7H_{16}	100.2	—	1.04	0.264
Hydrogen	H_2	2.01	1.41	1.40	0.0053
Hydrogen Chloride	HCl	36.5	1.48	—	0.09650
Hydrogen Sulfide	H_2S	34.1	1.30	1.31	0.0901
Methane	CH_4	16.03	1.316	1.28	0.0423
Methyl Chloride	CH_3Cl	50.5	1.20	—	0.1336
Natural Gas (approx.)	—	19.5	1.27	—	0.0514
Nitric Oxide	NO	30.0	1.40	—	0.0793
Nitrogen	N_2	28.0	1.41	1.40	0.0743
Nitrous Oxide	N_2O	44.0	1.311	—	0.1163
Oxygen	O_2	32.0	1.4	1.39	0.0846
Pentane	C_5H_{12}	72.1	1.06	1.06	0.1905
Propane	C_3H_8	44.1	1.15	1.11	0.1164
Propylene	C_3H_6	42.0	1.16	—	0.1112
Sulfur Dioxide	SO_2	64.1	1.256	—	0.1694
Water Vaper (steam)	H_2O	18.0	1.33[a]	1.32	0.04761

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[*] Compiled from: "Plain Talks on Air and Gas Compression, Fourth of Series," Worthington Corp. and "Reciprocating Compressor Calculation Data Cooper-Bessemer Corp. (1956), by permission.

[a] At 212 °F.

Table A3–4 Physical Properties of Selected Amines*

Compound	Monoethanol-amine	Diethanol-amine	Methyldiethanolamine	Diglycol-amine	Diisopropanol-amine
Formula	$HOC_2H_4NH_2$	$(HOC_2H_4)_2NH$	$(HOC_2H_4)_2NCH_3$	$H(OC_2H_4)_2NH_2$	$(HOC_3H_6)_2NH$
Molecular Wt	61.08	105.14	119.16	105.14	133.19
Boiling point @ 760 mmHg, °C	170.5	269	247	221	248.7
Freezing point, °C	10.5	28.0	–23	− 12.5	42
Critical constants					
Pressure, kPa (abs)	5985	3273		3772	3770
Temperature, °C	350	442.1		402.6	399.2
Density @ 20°C, gr/cc.	1.018	1.095		1.058 @ 15.6°C	0.999 @ 30°C
Weight, kg/m^3	1016 @ 15.6°C	1089 @ 15.6°C	1040	1057 @ 15.6°C	
Relative density 20°C/20°C	1.0179	1.0919 (30°/20°C)	1.0418	1.0572	0.989 @ 45°C/20°C
Specific heat @ 15.6°C, kJ/(kg.°C)	2.55 @ 20°C	2.51	2.24	2.39	2.89 @ 30°C
Thermal conductivity @ 20°C, W/(m.°C)	0.256	0.220	0.275	0.209	—
Latent heat of vaporization, kJ/kg	826 @ 760 mmHg	670 @ 73 mmHg	476	510 @ 760 mmHg	430 @ 760 mm Hg
Heat of reaction, kJ/kg of Acid Gas					
H_2S				− 1568	—
CO_2				− 1977	—
Viscosity, mPa.sec	24.1 @ 20°C	350 @ 20°C (at 90% wt. solution)	1.3×10^{-6} m^2/s @ 10°C $0.68\ 10^{-6}$ m^2/s @ 38°C 0.28×10^{-6} m^2/s @ 100°C	40 @ 16°C	870 @ 30°C 198 @ 45°C 86 @ 54°C
Refractive index, N_d 20°C	1.4539	1.4776	1.469	1.4598	1.4542 @ 45°C
Flash point, COC, °C	93	138	129.4	127	124

* *GPSA Engineering Data Book,* 12th Ed., Gas Processors Suppliers Association (GPSA), Tulsa, OK (2004).

Table A3–5 Physical Properties of Selected Glycols and Methanol[*]

Compound	Ethylene Glycol	Diethylene Glycol	Triethylene Glycol	Tetraethylene Glycol	Methanol
Formula	$C_2H_6O_2$	$C_4H_{10}O_3$	$C_6H_{14}O_4$	$C_6H_{18}O_5$	CH_3OH
Molecular Mass	62.1	106.1	150.2	194.2	32.04
Boiling Point[a] @ 760 mmHg, °C	197.3	244.8	285.5	314	64.5
Vapor Pressure @ 77°F (25°C), mmHg	0.12	<0.01	<0.01	<0.01	120
Density @ 77°F (25°C), kg/m^3	1110	1113	1119	1120	790
Freezing Point, °C	−13	−8	−7	−5.5	−97.8
Pour Point, °C	—	−54	−58	−41	
Viscosity in centipoise					
@ 77°F (25°C)	16.5	28.2	37.3	44.6	0.52
@ 140°F (60°C)	4.68	6.99	8.77	10.2	
Specific Heat @ 77°F (25°C), kJ/(kg.K)	2.43	2.30	2.22	2.18	2.52
Flash Point, °C (PMCC)	116	124	177	204	12

Note: These properties are laboratory results on pure compounds or typical of the products, but should not be confused with, or regarded as, specifications.

[*] *GPSA Engineering Data Book,* 12th Ed., Gas Processors Suppliers Association (GPSA), Tulsa, OK (2004).

[a] Glycols decompose at temperatures below their atmospheric boiling point. Approximate decomposition temperatures are

Ethylene Glycol	165°C	Triethylene Glycol	207°C
Diethylene Glycol	164°C	Tetraethylene Glycol	238°C

Table A3–6 Specific Gravities of Aqueous Triethylene Glycol Solutions*

Specific Gravity @ T/60 °F = $A + Bx + Cx^2$, where x = TEG wt%

T°F	A	B	C
−50	1.0502	1.8268E–3	−5.2009E–6
0	1.0319	1.7466E–3	−4.8304E–6
50	1.0121	1.5247E–3	−2.8794E–6
100	0.9920	1.7518E–3	−5.4955E–6
150	0.9804	1.5410E–3	−4.3884E–6
200	0.9627	1.4068E–3	−3.5089E–6
250	0.9413	1.3205E–3	−2.7991E–6
300	0.9177	1.2511E–3	−2.0848E–6

*Union Carbide Chemicals & Plastics Company Inc., "Triethylene Glycol," 39 Old Ridgebury Rd., Danbury, CT 06817 (1989).

Table A3–7 Specific Heat for Aqueous Triethylene Glycol Solutions*

Specific Heat = $A + BT + CT^2$, where T = temperature, °C

TEG wt%	A	B	C
0	1.00540	−2.7286E–4	2.9143E–6
10	0.96705	−2.7144E–5	2.4952E–6
20	0.92490	2.0429E–4	2.4524E–6
30	0.88012	4.3000E–4	1.6952E–6
40	0.83229	6.2286E–4	1.3714E–6
50	0.78229	7.9286E–4	1.0857E–6
60	0.72200	9.4000E–4	8.0000E–7
70	0.66688	1.0871E–3	4.7620E–7
80	0.60393	1.2043E–3	2.8571E–7
90	0.53888	1.2800E–3	1.9048E–7
100	0.48614	1.3929E–3	−5.7140E–8

*Union Carbide Chemicals & Plastics Company Inc., "Triethylene Glycol," 39 Old Ridgebury Rd., Danbury, CT 06817 (1989).

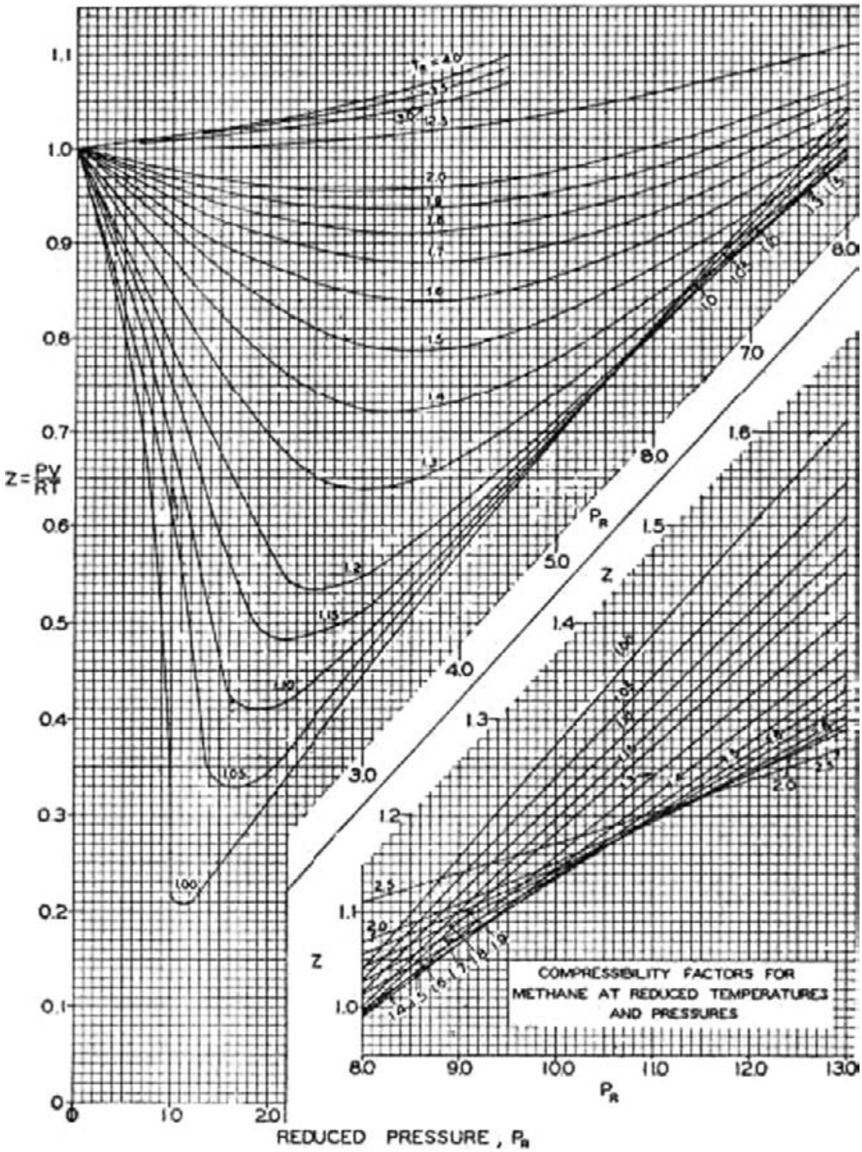

Figure A3–1 Compressibility of natural gases (GPSA, 2004).

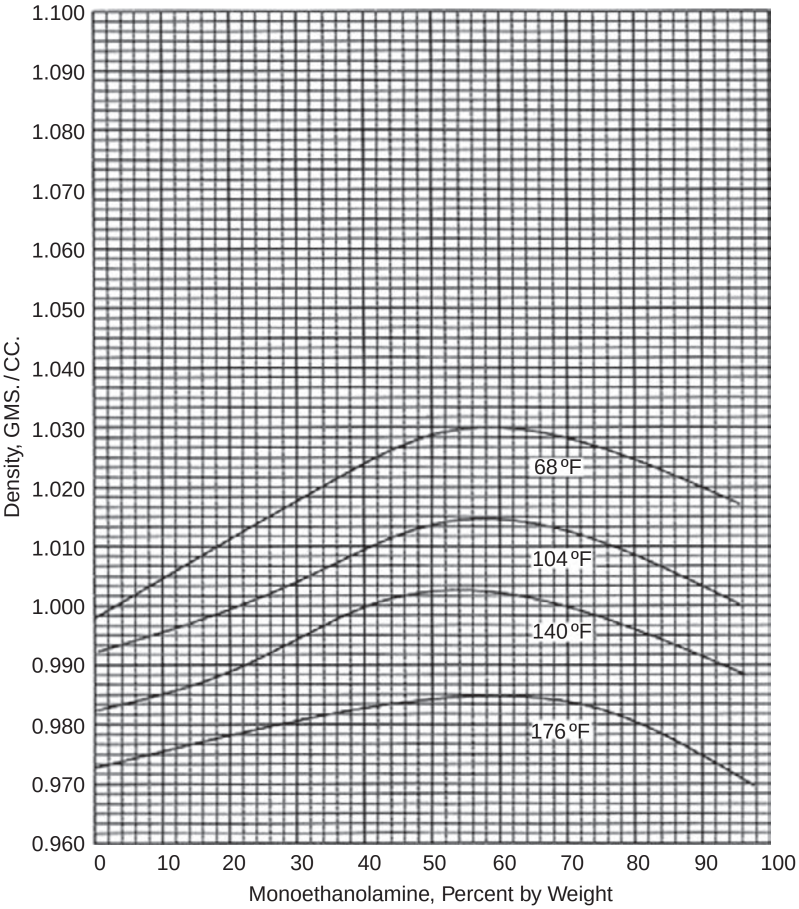

Figure A3–2 Density of MEA solutions (*Gas Conditioning Fact Book.* Dow Chemical Co., Houston, Texas, 1969).

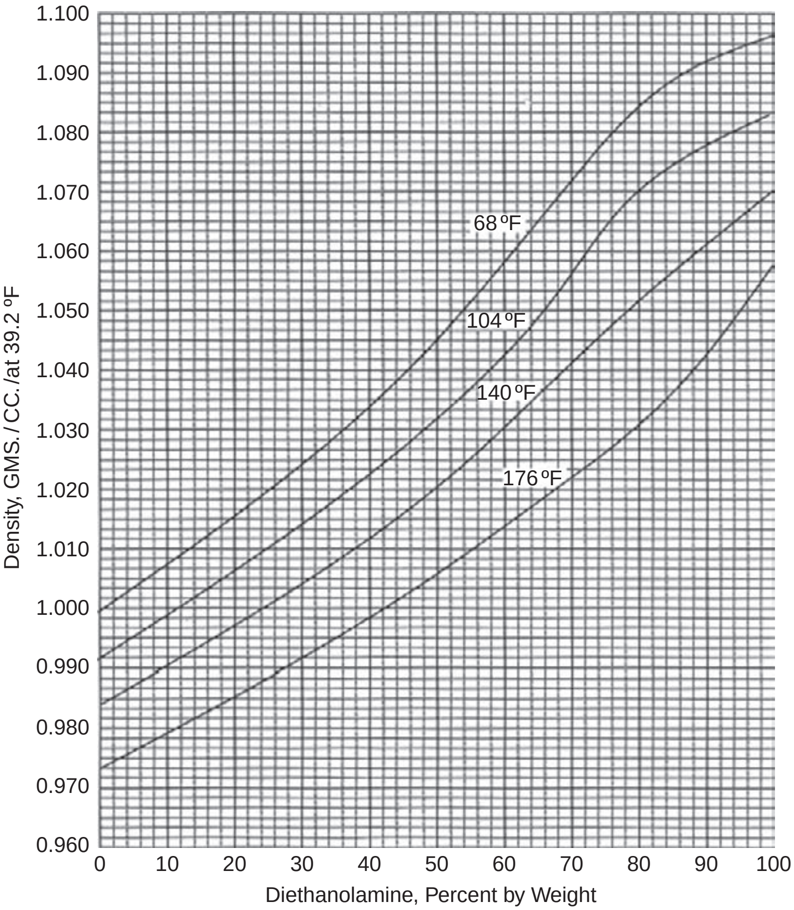

Figure A3–3 Density of DEA solutions (*Gas Conditioning Fact Book*. Dow Chemical Co., Houston, Texas, 1969).

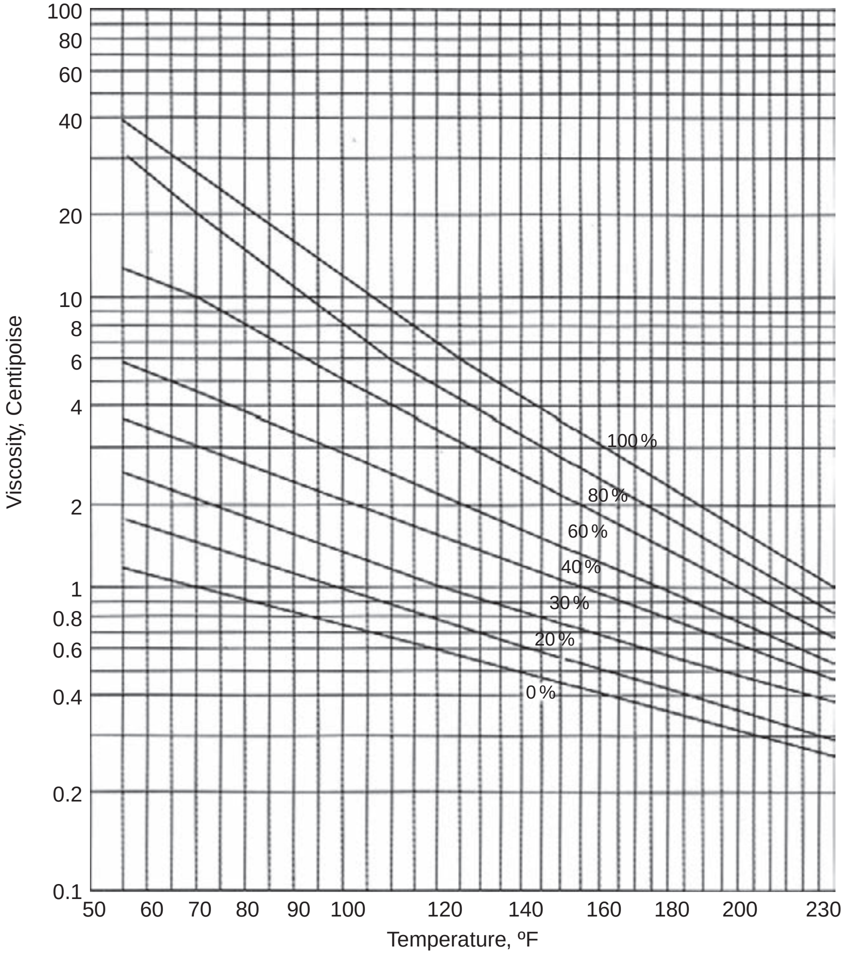

Figure A3–4 Viscosity of aqueous MEA solutions as function of weight percent MEA (*Gas Conditioning Fact Book*. Dow Chemical Co., Houston, Texas, 1969).

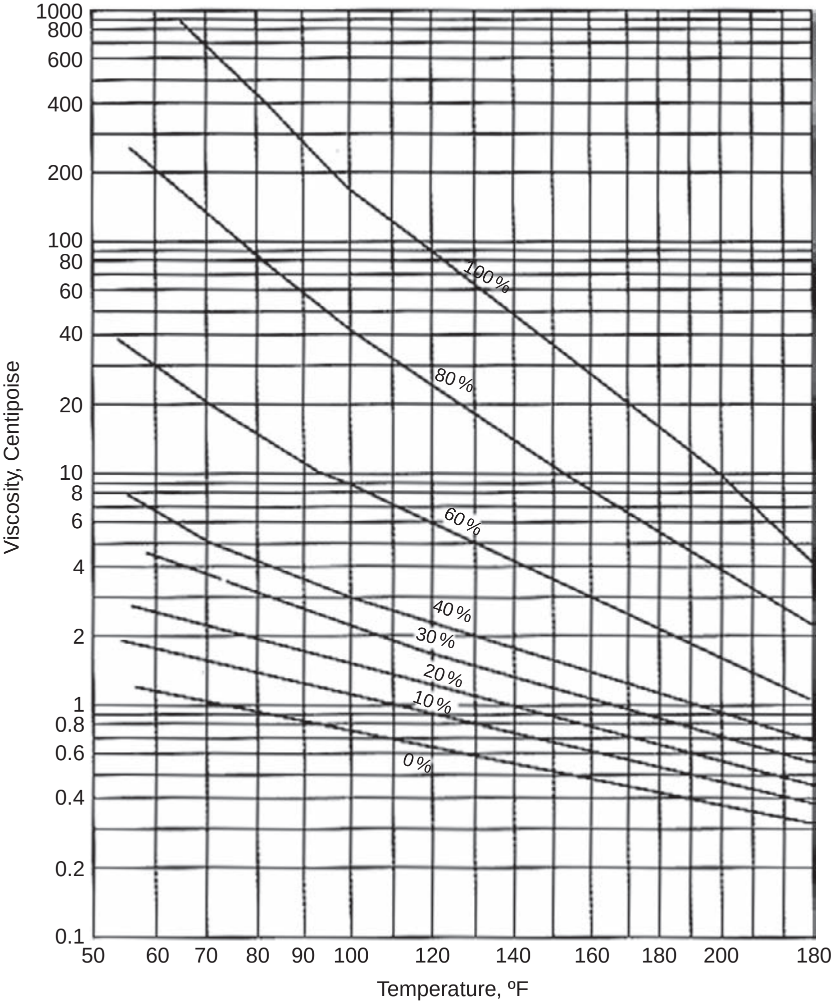

Figure A3–5 Viscosity of aqueous DEA solutions as function of weight percent DEA (*Gas Conditioning Fact Book*. Dow Chemical Co., Houston, Texas, 1969).

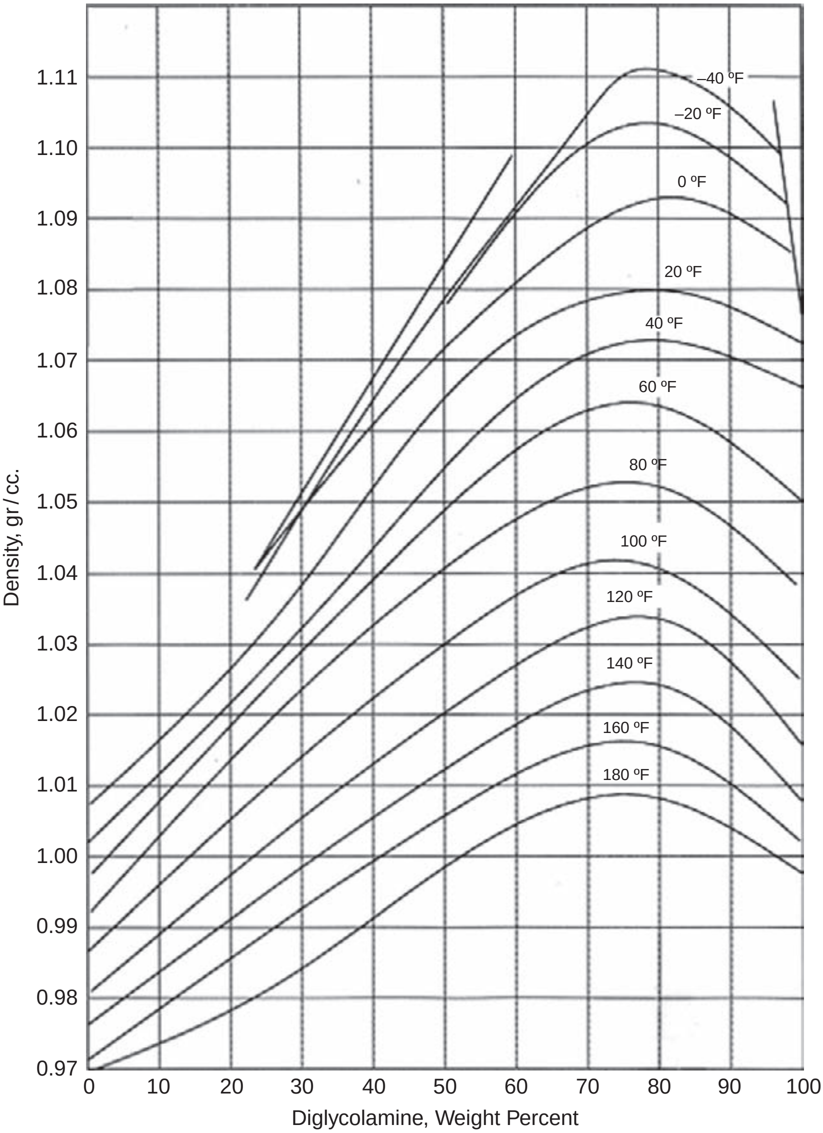

Figure A3–6 Density of aqueous DGA solutions (*Gas Treating Data Book*. Jefferson Chemical Co., Houston, Texas, 1969).

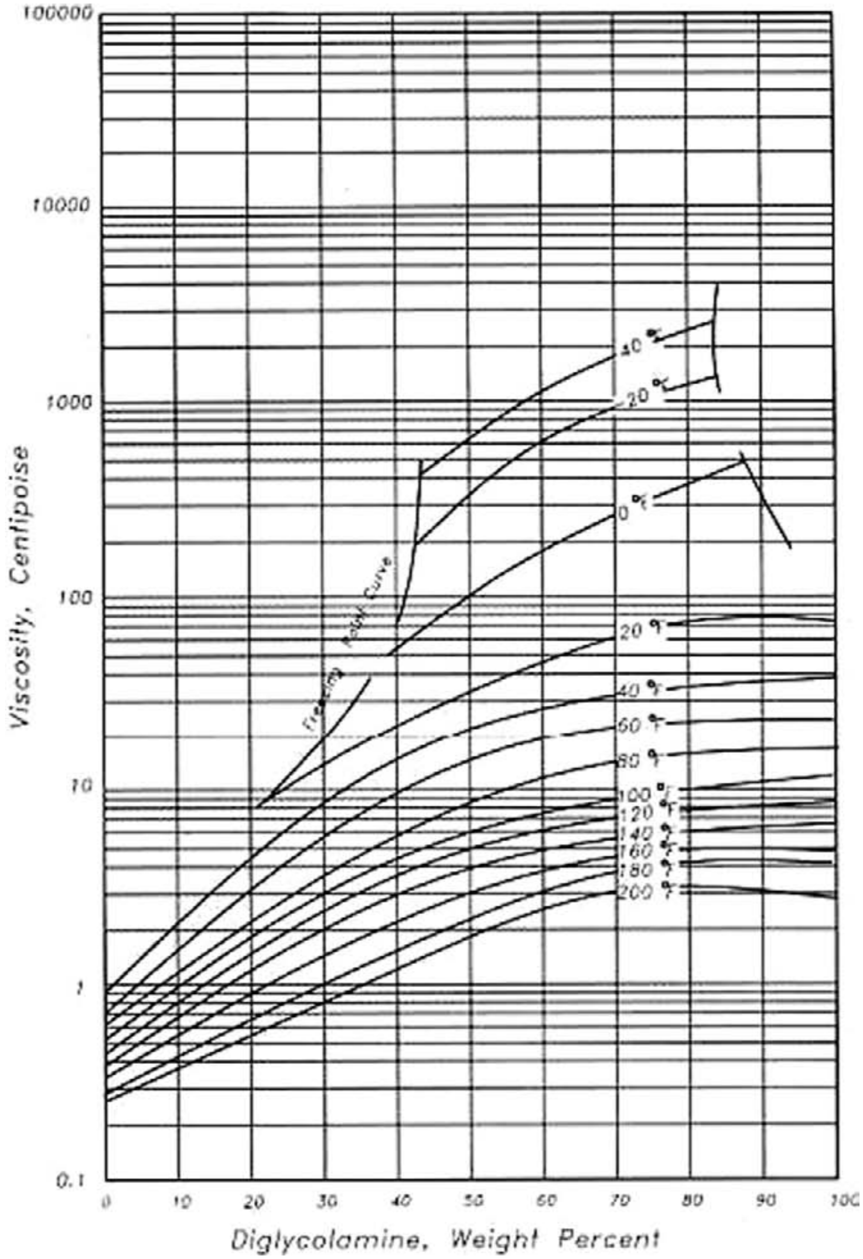

Figure A3–7 Viscosity of aqueous DGA solutions (*Gas Treating Data Book*. Jefferson Chemical Co., Houston, Texas, 1969).

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APPENDIX 4

Glossary

Absorbent: Liquid used to remove specific components from a gas stream.

Absorption: A separation process involving the transfer of a substance from a gaseous phase to a liquid phase through the phase boundary.

Acid gas: The hydrogen sulfide, carbon dioxide, or both contained in or extracted from a natural gas.

Adsorbent: Solid pellets used to remove specific components from natural gas.

Adsorption: The process by which gaseous components are adsorbed on solids because of their molecular attraction to the solid surface. Common examples include: acid gases, water vapor, or heavier hydrocarbon vapors.

Amine: Basic components containing the amino or a substituted amino group. The amines are generally used in water solutions to remove hydrogen sulfide and carbon dioxide from gas and liquid streams.

Antifoam: A substance, usually a silicone or a long-chain alcohol, added to the treating system to reduce the tendency to foam.

Associated gas: Natural gas produced with crude oil and separated in a gas-oil separator.

Blanket gas: Gas introduced into a vessel above a liquid phase usually to protect the liquid from air contamination, for reducing the detonation hazard, or for pressuring the liquid. The source of the gas is external to the vessel.

Blowdown: The act of emptying or depressuring a vessel.

Boiling point: Temperature at which the vapor pressure of a liquid equals the external pressure. When external pressure is 1 atm, temperature is called the normal boiling point.

Breathing: The movement of vapor in or out of a storage tank because of the change of level of the storage liquid, a change in the temperature of the vapor space above the liquid, or of atmospheric pressure.

Bubble point: Temperature and pressure at which the first bubble of vapor forms in a liquid.

Byproduct: A material produced in addition to the main product or result.

Claus process: A process to convert hydrogen sulfide into elemental sulfur by selective oxidation.

Cleaning: Removing liquids and solids from natural gas.

Compressibility factor: Dimensionless ratio of the actual volume occupied by a gas to that predicted using the ideal gas law.

Compression ratio: The ratio of the absolute discharge pressure from a compressor to the absolute suction or intake pressure.

Condensate: Hydrocarbon liquids condensed from natural gas.

Conditioning: Operations used to make produced gas and/or oil suitable for storage and transport.

Copper strip test: A test using a small strip of pure copper to determine qualitatively the corrosivity of a product.

Cricondenbar: The highest pressure at which vapor and liquid phases can coexist in a multicomponent mixture.

Cricondentherm: The highest temperature at which vapor and liquid phases can coexist in a multicomponent mixture.

Critical pressure: The pressure necessary to condense a vapor at its critical temperature.

Critical temperature: The highest temperature at which a fluid can exist as a liquid.

Cryogenic plant: Gas processing plant capable of recovering NGL including ethane using a turboexpander to generate very low operating temperatures.

Custody transfer: Sale (transfer of ownership) of oil or gas.

Cycle gas: Gas that is compressed and returned to the gas reservoir to minimize the decline of reservoir pressure.

Degradation products: Impurities in a treating solution that are formed from both reversible and irreversible side reactions.

Dehydration: The act or process of removing water from gases or liquids.

Desiccant: An adsorbent used to dry a fluid.

Desulfurization: Removal of sulfur and sulfur compounds from gases or liquid hydrocarbons.

Dew point: The temperature and pressure at which a liquid initially condenses from a gas or vapor. Specifically applied to the temperature at which water vapor (water dew point), or hydrocarbons (hydrocarbon dew point) condense.

Doctor test: Qualitative method for detecting hydrogen sulfide and/or mercaptans in petroleum distillates. The test distinguishes between "sour" and "sweet" products.

Dry gas: Gas with water content reduced by dehydration.

Feedstocks: Crude oil, natural gas liquids, natural gas, or other material used as raw materials for the production of gasoline, other refined products, or chemicals.

Field processing: All operations used to make crude oil and/or natural gas suitable for storage and transport.

Flash point: Minimum temperature at which a liquid gives off vapor in sufficient concentration to form an ignitable mixture with air near the surface of the liquid.

Flashing: When a liquid at its bubble point travels across a restricting control valve, its pressure and temperature decrease. Some of the liquid turns into vapor in the process, for which the word "flash" is used.

Gas processing: All the field operations used to separate the constituents of and remove the contaminants in a natural gas.

Gas injection: The injection of natural gas into a reservoir to maintain or increase the reservoir pressure or reduce the rate of decline of the reservoir pressure.

Gas lift: A method of bringing crude oil or water to the surface by injecting gas into the producing well bore.

Gasoline: A mixture of hydrocarbons used in a spark-ignited internal combustion engine.

Glycol: A group of compounds used to dehydrate gaseous or liquid hydrocarbons or to inhibit the formation of hydrates.

Handling: See "Conditioning."

Heavy ends: Relative term used to compare those hydrocarbons that have more carbon atoms than others.

Heptane plus: All hydrocarbon components in a fluid mixture with seven or more carbon atoms in the molecule.

Hexane plus: All hydrocarbon components in a fluid mixture with six or more carbon atoms in the molecule.

Hydrate: A solid (ice-like) material that is the result of the combination of hydrocarbons with water, usually under high pressure and at low temperature.

Hydrocarbons: Organic chemical compounds of hydrogen and carbon atoms that form the basis of all petroleum products, natural gas, and coals. Hydrocarbons may be liquid, gaseous, or solid.

Inert gas: Carbon dioxide, nitrogen, or helium.

Joule-Thomson effect: The change in gas temperature that occurs when the gas is expanded at constant enthalpy from a high to low pressure.

Lean amine: Amine solution stripped of absorbed acid gases.

Lean gas: (1) Residue gas after recovery of NGL. (2) Unprocessed gas containing little or no recoverable NGL.

Lean oil: Absorption oil from which all absorbed NGLs have been stripped.

Light ends: The low-boiling, easily evaporated components of a hydrocarbon mixture.

Liquefied natural gas (LNG): The light hydrocarbon portion of natural gas, predominately methane, that has been liquefied.

Liquefied petroleum gas (LPG): A compressed or liquefied gas produced as a byproduct from refining petroleum. Includes propylene, propane, butylene, butane, isobutane, and other elements.

Mercaptan: Any of a homologous series of compounds of the general formula RSH. R signifies an organic group such as a methyl, ethyl, propyl or other group.

Molecular sieve: Synthetic zeolite (essentially silica-alumina) used in adsorption processes.

Natural gas: Gaseous hydrocarbon mixture found in nature. Principal hydrocarbon components are C1, C2, C3, C4's, and C5+. May also contain N_2, CO_2, H_2O, H_2S.

Natural gas liquid (NGL): Hydrocarbon liquids including ethane, propane, butanes, and natural gasoline.

Natural gasoline: A mixture of pentanes and heavier hydrocarbons extracted from natural gas.

Odorant: Highly odiferous chemical, usually a light mercaptan, added to gas to impart to it a distinctive odor for safety precaution and to detect leaks.

Pentane-plus: All hydrocarbon components in a fluid mixture with five or more carbon atoms in the molecule.

Pigging: Procedure of forcing a solid object through a pipeline for cleaning or other purposes.

Pipeline gas: Gas that meets a transmission company's minimum specifications.

Raw gas: Unprocessed gas or the inlet gas to a plant.

Reid vapor pressure (RVP): Vapor pressure of a liquid determined by laboratory testing in a standard manner in the Reid apparatus at a standard temperature of 37.8°C/100°F.

Residue gas: Gas remaining after the recovery of liquid products.

Rich amine: The amine that leaves the bottom of the absorber. It is the lean amine plus the acid gases removed from the gas by the lean amine.

Rich gas: Raw, unprocessed natural gas that contains large amounts of associated natural gas liquids.

Rich oil: The oil leaving the bottom of an absorber. It is the lean oil plus all of the absorbed constituents.

Sales gas: Same as pipeline gas.

Shrinkage: (1) The reduction in volume, heating value, or both of a gas stream caused by the removal of some of its constituents. (2) Sometimes referred to as the unaccounted loss of products from storage tanks.

Shutdown: Control actions undertaken to stop operation of equipment or a process. Shutdown can be automatically triggered or initiated by voluntary action.

Sour gas: Gas containing undesirable quantities of hydrogen sulfide, mercaptans, and/or carbon dioxide.

Stabilized condensate: Condensate that has been stabilized to a definite vapor pressure in a fractionation system.

Sweet gas: Gas that has no more than the maximum sulfur content defined by the specifications for the sales gas from a plant.

Tail gas: The exit gas from a plant.

True vapor pressure (TVP): The pressure at which the gas and liquid in a closed container are in equilibrium at a given temperature.

Vapor pressure: Pressure exerted by a liquid when in equilibrium with a vapor.

Water dew point: The temperature at which water begins to condense from a gas as it is cooled.

Wet gas: Natural gas that yields hydrocarbon condensate (does not usually refer to water content).

ACRONYMS

acf:	Actual cubic feet
acfm:	Actual cubic feet pet minute
aMDEA:	Activated MDEA
AGA:	American Gas Association
AGEU:	Acid gas enrichment unit
ANSI:	American National Standards Institute
API:	American Petroleum Institute, national trade association of the U.S. petroleum industry, a private standardizing organization
ASME:	American Society of Mechanical Engineers
ASTM:	American Society for Testing and Materials Atmosphere
atm:	Atmosphere
bara:	Unit of pressure, bar absolute
barg:	Unit of pressure, bar gauge
bbl:	Barrel (42 U.S. gallons). The oil industry standard for volumes of oil and its products.
BFW:	Boiler feed water
BHP:	Brake horsepower
BS&W:	Basic sediment and water; water and other contaminants present in crude oil
Btu:	British thermal unit
C1:	Methane
C2:	Ethane
C3:	Propane

C4:	Butane
C5:	Pentane
C5+:	Pentane and heavier
C6:	Hexane
C6+:	Hexane and heavier
C7:	Heptane
C7+:	Heptane and heavier
C8:	Octane
CAPEX:	Capital expenditure
cfm:	Cubic feet per minute
CNG:	Compressed natural gas
CO_2:	Carbon dioxide
COS:	Carbonyl sulfide
CS_2:	Carbon disulfide
DEA:	Diethanolamine
DEG:	Diethylene glycol
DGA:	Diglycolamine
DIPA:	Diisopropanolamine
DMRU:	Dehydration and mercaptans removal unit
DPCU:	Dew-point controlling unit
EG:	Ethylene glycol
ft/sec:	Feet per second
gal:	U.S. gallon
GDU:	Gas dehydration unit
GLR:	Gas liquid ratio, expressed as scf/bbl
GOR:	Gas-oil ratio, combined gas released from stage separation of oil, expressed as scf/Bsto
Gpm/GPM:	(1) gpm (gallons per minute): flowrate in gallons per minute (2) GPM – Preferably gal/Mcf: natural gas liquids content in gallons of liquid products per thousand cubic feet.
GSU:	Gas sweetening unit
GTL:	Gas to liquid
H_2O:	Water
H_2S:	Hydrogen sulfide
HC:	Hydrocarbon
HP:	High pressure
iC4:	Isobutane
iC5:	Isopentane
ID:	Internal diameter
JT:	Joule–Thomson (constant enthalpy) expansion
KW:	Kilowatts
lb:	Pounds
LNG:	Liquefied natural gas
LP:	Low pressure
LPG:	Liquefied petroleum gas
Max:	Maximum
Mcf:	One million cubic feet

Mcfd: Thousand cubic feet per calendar day
MDEA: Methyldiethanolamine
MEA: Monoethanolamine
MEG: Monoethylene glycol
Min: Minimum
MMscfd: Million standard cubic feet per calendar day
MRU: Mercaptans removal and/or recovery unit
MS: Molecular sieve
MSU: Molecular sieve unit
MW: Molecular weight
NACE: National Association of Corrosion Engineers
NRU: Nitrogen rejection (removal) unit
nC4: Normal butane
nC5: Normal pentane
NGL: Natural gas liquids: includes ethane, propane, butanes, pentanes, or mixtures of these
OD: Outside diameter
OPEX: Operating and maintenance expenditure
ppb: parts per billion
ppm: Parts per million
ppmv: Parts per million by volume
ppmw: Parts per million by weight
psi: Pounds per square inch
psia: Pounds per square inch absolute
psig: Pounds per square inch gauge
RP: Recommended practices, e.g., API RP I4E
rpm: Revolutions per minute
RSH: Mercaptan
RVP: Reid vapor pressure
scfm: Standard cubic feet per minute
SG: Specific gravity
SRU: Sulfur recovery unit
TEG: Triethylene glycol
TGTU: Tail gas treating unit
LTGTU: Lurgi tail gas treating unit
VLE: Vapor-liquid equilibrium
VLLE: Vapor-liquid-liquid equilibrium
VOC: Volatile organic compounds
WHB: Waste heat boiler
Δ: Increment or difference
°API: Degrees API gravity
°F: Degrees Fahrenheit

INDEX

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