White Paper on Liquid Hydrocarbon Drop Out in Natural Gas Infrastructure

NGC+ Liquid Hydrocarbon Drop Out Task Group
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Gas Quality White Paper
Control of Hydrocarbon Liquid Drop Out

Section 1 - Introduction

1.0   Objective

1.1   The objective of this report is to provide background on the issue of **gas quality, specifically hydrocarbon liquid drop out,** and recommend how it can be managed in a way that balances the concerns of all stakeholders in the value chain[1]. These concerns are summarized below:

1.1.1   Producers want the ability to supply natural gas to meet increasing demand. They seek to maximize their natural gas revenue stream by electing the level to process their gas based on market conditions while satisfying pipeline tariff, safety and environmental requirements.

1.1.2   Gas Processors want to know the long term specification requirements for the quality of gas to be delivered into transmission pipelines in order to set operating conditions, evaluate potential investments in reconfiguring their plants to optimize the production of thermal content and meet the pipeline quality specifications and, in many instances, renegotiate the contracts that they have with the gas producers.

1.1.3   Pipelines want to provide transportation flexibility to meet demand but are concerned about operational safety and reliability, system integrity and environmental issues. They are also concerned about whether components of gas they accept for delivery may make the gas in their pipeline unacceptable to distribution systems and end users.

1.1.4   Local distribution companies want to meet customer demand but are concerned about operational safety and reliability, system integrity, and environmental issues as well as the impacts on end use equipment. They have little or no existing capacity to remove or extract hydrocarbons from their systems.

1.1.5   Direct connect customers (e.g., power plants and industrial users directly connected to transmission pipeline) want uniformity of gas quality because of safety and environmental concerns, and potential negative impacts on equipment, end products, and operational reliability. They have little or no existing capacity to remove or extract hydrocarbons from their systems.

1.1.6   End Users (e.g., customers receiving gas from the LDC) expect uniformity of gas quality for appliances, industrial applications, including use as a feedstock or building block in chemical manufacturing.

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[1] There is a separate effort directed at higher heating values, including the role of liquefied natural gas. This effort is referred to as "interchangeability" and is being managed by the Natural Gas Council Interchangeability Task Group.

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1.2   Overview of the Report

This report will examine the occurrence of hydrocarbon liquids in natural gas, the role of gas processing, and historical measures used to control hydrocarbon liquid drop out. There are seven sections, including this Introduction. They are:

1.2.1   Section 2 - Liquid Hydrocarbons in Natural Gas
This section describes the sources of natural gas and shows that all gas as produced is not the same. It describes the role of treatment and processing to provide a more uniform, fungible commodity. It also describes the challenges to controlling hydrocarbon liquid drop out when faced with the influences of pressure reductions and ambient temperature.

1.2.2   Section 3 – Hydrocarbon Liquid Drop Out Control Measures
This section describes measures used historically to control hydrocarbon liquid drop out, including heating value (Btu/volume), and composite concentrations of heavier weight hydrocarbons (such as the mole fraction of heavier weight hydrocarbons measured as the "pentane plus" fraction, referred to as C5+ or the "hexane plus", referred to as C6+)[2]. This section also provides a description of blending, a tool to provide shippers and pipeline operators some flexibility in controlling hydrocarbon liquid drop out.

1.2.3   Section 4 - Overview of Hydrocarbon Dew Point (HDP)
This section defines hydrocarbon dew point and describes how it can be used as a means to understand the behavior of hydrocarbons in a natural gas stream. The section provides a basic description of the thermodynamic principles governing the behavior of compounds found within natural gas. It describes the behavior of hydrocarbons as gas is processed, and as pressure and temperature change downstream in the value chain.

1.2.4   Section 5 - Historical Levels of Hydrocarbons and Hydrocarbon Dew Point
This section provides a summary of historical data on natural gas streams from a variety of sources, including detailed analyses of hydrocarbon constituents in gas as produced and processed. The section also provides historical levels of hydrocarbon dew points.

1.2.5   Section 6 - Determination of Hydrocarbon Dew Point – Measurement and Estimation
This section provides an overview of the direct determination of hydrocarbon dew point. A chilled mirror is used to measure hydrocarbon dew point directly. Alternatively, a combination of sampling, analysis and calculation using a simplified equation of state from chemical thermodynamics is used to estimate the hydrocarbon dew point. The section provides an overview of the value of each in predicting hydrocarbon liquid drop out.

1.2.6   Section 7 – Recommendations
This section provides a set of recommendations developed by the Natural Gas Council HDP Task Group to manage hydrocarbon liquid drop out.

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[2] The abbreviation C6 for example refers to hexanes. The addition of "+" is a term of art used in analytical chemistry that refers to a grouping of compounds (or fraction). For example, C6+ represents C6, as well as C7, C8 and higher molecular weight hydrocarbons. C9+ refers to C9 plus C10, C11 and so forth.

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1.2.7   Appendices
   A. Parameters to be Considered in Establishing a Cricondentherm Hydrocarbon Dew
       Point and C6+ Gallons Per Million (GPM) Cubic Feet Based Limits
   B. Process for Establishing a Cricondentherm Hydrocarbon Dew Point (CHDP) Limit

1.3   Background and Summary of the Issues

1.3.1   Historically, the commercial value of the liquefiable hydrocarbons extracted
from North American natural gas, referred to as natural gas liquids (NGLs), has been greater
than the commercial value of the thermal content that would be added if the NGLs
remained part of gas stream. The infrastructure to extract these NGLs, referred to as the
processing industry, has been built up over time. Some facilities were built to remove NGLs
for operational concerns, but the economic uplift derived from extracting NGLs has resulted
in an entire industry dedicated to production and sales of NGL products.

1.3.2   At times, the value of natural gas has increased dramatically as compared to
the value of the NGLs. Rising natural gas prices relative to NGL prices decrease the
economic incentive to extract NGLs. In this environment suppliers and processors may elect
to reduce extraction levels or bypass processing.

1.3.3.   This economic environment creates two issues for transmission, distribution
and utilization of domestic natural gas. First the decreased level of processing causes the
presence of larger amounts of liquefiable hydrocarbons in the gas stream resulting in a
greater potential for hydrocarbon liquids to drop out of the gas phase while in transit to end
use equipment. This increases the potential for problems in pipeline and LDC operations
with compression, measurement, pressure regulation, over-pressure protection devices and
potential interference with odorization. Second, problems can also occur in end-use
applications such as flame extinguishing or over-firing in home appliances or physical
damage to gas turbines used to generate electricity.

1.3.4. In addition, those LDC's that operate LNG peak shaving liquefaction plants
are concerned about the impact of increasing hydrocarbon dew point has on the overall
thermodynamic process. Feedstock received with an excessive hydrocarbon dew point can
result in adverse plant operations including heat exchanger fouling and excess liquids
collection at points in the process beyond what the plant is designed to handle. This subject
is also directly linked to the Interchangeability issue and will be discussed in detail in the
Interchangeability White Paper.

1.4   Natural Gas – From Wellhead to Burner Tip

1.4.1   This report begins with a brief description of how natural gas makes its way
from the wellhead to the burner tip. Natural gas is produced from one of three sources:
associated gas, recovered in conjunction with oil production, non-associated gas (gas from a
field not producing oil), and as a gaseous stream from coal seams (normally referred to as
coal bed methane). All natural gas is not of the same quality when produced. Each of the
sources exhibits distinct characteristics and even gas produced from a particular source may
vary with the most abundant component being methane. Produced gas will also contain
varying quantities of non-methane hydrocarbons and other constituents that contribute little

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or no heating value. Depending upon the concentrations present, the gas may require treatment to reduce constituents such as water, carbon dioxide, nitrogen, oxygen, total sulfur and hydrogen sulfide. Natural gas that is rich in non-methane hydrocarbon constituents may also be further processed to extract natural gas liquids.

1.4.2   The next step in the path to the burner tip is the custody transfer to a Shipper who contracts for the transportation of the gas through open access pipelines (Transporters) that transport gas to a delivery point at which it is delivered to a distribution company or directly to an end user. Tariffs filed with FERC define the contract and commercial conditions for transporting gas from a specified receipt point to a specified delivery point. Transactions involving transportation of natural gas on pipelines are measured in units of energy called "dekatherms" (MMBtus[3]). Meters measure gas volumes and the heating value is determined by compositional analysis using results of gas chromatography. In general, gas volumes are measured continuously using one of several types of meters. Larger volume onshore receipt points generally use online continuous gas chromatographs (typically daily volumes of about 5 to 50 MMSCF[4], or higher). Manual spot or composite samples are more typical at smaller volume receipt points as well as most offshore transmission receipt points.

1.4.3   Pipeline operators (transporters) have found the need to establish tariff specifications at receipt points for certain constituents affecting gas quality, including water, carbon dioxide, oxygen, total sulfur, hydrogen sulfide, among others, to ensure safe and reliable operations. These constituents, in sufficient quantities, can create a corrosive environment adversely affecting safety and operations in the pipeline system and eventually can create combustion problems in downstream end use equipment. The tariff limits are typically expressed as maximum limits. Gas nominated for transportation must be provided within these limits. Depending upon regulatory issues, operating conditions, and other criteria, pipeline operators may waive tariff limits for a particular shipper on a short-term basis. Natural gas as it is transported in the manner described above is viewed as being fungible; that is, gas transported by one shipper may be interchanged with gas from another shipper without impacting the pipeline's ability to transport gas of acceptable quality to its downstream customers.

1.4.4   As stated earlier, when the commercial value of natural gas liquids is at a discount relative to their value as a thermal contribution in the natural gas, producers may elect to reduce extraction or bypass gas processing if not otherwise obligated. Most pipelines have been designed throughout the years with a variety of means to capture small incidental volumes of liquids so as to protect downstream facilities. Some pipeline companies have installed various two-phase (i.e., gas and liquid) lines to accommodate the presumption of liquid formation. Generally, these facilities are located upstream of compressor stations and measurement stations. Some pipelines have configured their producing area pipelines to handle both liquids and gas. These special lines are located in proximity to and upstream of liquids handling infrastructure such as a condensate removal facilities or a processing plant. With the exception of the specially designed two-phase systems, most pipeline systems anticipated liquid free operation and in many instances found no need to install liquid

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[3] million Btus
[4] million standard cubic feet

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handling equipment. The chemistry and thermodynamics of processed natural gas support operations in this manner. This is because processed gas is sufficiently lean (low liquefiable content) as to be able to provide absorptive capacity in the event that small volumes of liquefiable hydrocarbons are introduced into the pipeline system. However, if the gas temperature becomes sufficiently low at any point in the pipeline system or in the end user system, hydrocarbons can condense to form liquids from the natural gas mixture. Similarly, water vapor in the natural gas stream can condense to free water if the temperature of the gas gets low enough.

1.4.5 The water dew point is the temperature at which water vapor will condense to liquid water. The water content in a pipeline is already covered by tariff provisions and is mentioned here for illustrative purposes. Similarly, the hydrocarbon dew point (HDP) is the temperature at which hydrocarbons will begin to condense (refer to Section 4 – Overview of HDP); hence the expression "hydrocarbon liquid drop out".

1.4.6   The simplest means of controlling small incidental liquid accumulation is through installation of drips; a vessel attached to the pipeline that removes liquids through physical impingement or gravity collection in the pipeline system. The captured liquids accumulate and are periodically pumped or siphoned off and then either recovered as a fuel co-product (if regulation allows) or disposed of as a RCRA[5], TSCA[6] or State-listed hazardous waste. Disposal of these liquids as a hazardous waste may cause a dramatic increase in pipeline operating costs. The trend in recent years has been to remove drips from pipeline systems as they may be subject to corrosion. The Office of Pipeline Safety in some cases has required or encouraged operators to remove drips from their systems since the late-1990s.

1.4.7   Some pipeline operators have installed filtration or separation equipment, or both, on the suction side of compressor stations to collect solids (e.g. rust, weld slag and sand) and small volumes of water and compressor oils carried over from upstream stations. In addition, some LDCs and end users have installed similar equipment to collect small quantities of liquids dropping out as a result of temperature reductions associated with pressure reductions at city gate stations.

1.4.8   LDCs take custody of gas at the transmission pipeline delivery point. Direct connect customers take delivery from a delivery point on the mainline or often a lateral connected to the mainline. The gas must be measured at the point of the custody transfer from the pipeline to the LDC or customer. A metering station will occasionally include knockout vessels to remove any fugitive solids or liquids that may be in the gas prior to flowing through the measurement device. The pressure is normally reduced to the operating pressure of the LDC pipeline system either upstream or downstream of the meter. As the gas pressure is reduced, the temperature also will be reduced (the Joule Thomson effect). If the gas is not processed to specified levels, it is possible that a pressure reduction is enough to chill the gas to below the corresponding hydrocarbon dew point, thereby causing liquids to fall out. Likewise, if the processing is not done to specified levels, existing preheaters, separators or knockout vessels may be overwhelmed as to their capability to handle more than small quantities of hydrocarbon drop out. Heaters will be discussed further in sections

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[5] - Resource Conservation and Conservation Recovery Act
[6] - Toxic Substances Control Act

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2.4.6, 2.4.8 and 3.3. At high velocities, liquids become entrained, forming a mist. The mist may coalesce on the walls of the downstream pipeline and begin to collect in low spots of the pipeline system. Eventually, liquids can be swept along by the gas flow until reaching an exit point on the system -- a customer meter and burner. Liquids reaching a burner are a serious safety concern. They can degrade performance, spew out through the burner ports and either cause a large uncontrolled flame or extinguish the flame altogether and form a puddle in the hot appliance, with the potential to explosively reignite.

1.4.9   Hydrocarbon liquids in sensing lines to the equipment used for controlling pressure can cause erratic pressure variations in the delivered pipeline pressure. Such variations can impact nearby regulating stations upsetting large portions of a gas distribution system. This results in potential adverse impacts on system reliability or safety including overpressure protection devices.

1.4.10   Additional reliability and safety concerns for LDCs and end users due to natural gas liquids include the impact to polyethylene (PE) plastic piping, plastic piping components and current handling / pipe joining methodologies. According to APGA, approximately fifty percent of the typical LDCs distribution system is now comprised of plastic pipe and approximately ninety percent of new pipe installed is now plastic (2003 OPS Annual Report). Hydrocarbon gas constituents that are normally present within historical acceptable levels will have a minimal effect on the long-term strength of the plastic. However, it has been shown that aliphatic gaseous fuels of higher molecular weights ("heavy hydrocarbons") tend to be absorbed to a small extent by PE. This absorption somewhat reduces the long-term strength of PE pipe materials. Further, if the (NGLs) are routinely present, these liquids can cause a greater reduction in long-term strength up to 40% [7]. In addition, it has been reported that during the heat fusion joining of PE piping that has been in service conveying fuel gases that consist of, or include heavier hydrocarbons, the PE surfaces being heated in preparation for fusion on occasion will exhibit a "bubbly" appearance. The bubbling is a result of the rapid expansion (by heat) and passage of absorbed heavier hydrocarbon gases through the molten material, which could compromise the fusion joint if not properly recognized [8][9][10].

1.4.11   Hydrocarbon liquids present in a pipeline may not only cause operational and safety problems but also result in significant measurement error and unaccounted volume/energy losses. If liquids enter the gas sampling points, the sample will not be representative of the flowing gas stream, which results in inaccurate energy data, equipment failure, and costly equipment repair. Some pipeline operators and LDCs have had to install

[7] "Polyethylene Plastic Piping Distribution System Components of Liquefied Petroleum Gases", PPI Technical Report TR-22.
[8] Sudheer M. Pimputkar, Barbara Belew, Michael L. Mamoun, Joseph A. Stets, "Strength of Fusion Joints Made From Polyethylene Pipe Exposed to Heavy Hydrocarbons", Fifteenth International Plastics Pipe Symposium, October 1997.
[9] S.M. Pimputkar, J.A. Stets, and M.L. Mamoun, "Examination of Field Failures", Sixteenth International Plastics Pipe Symposium, November 1999.
[10] Gas Research Institute Topical Report GRI-96/0194, "Service Effects of Hydrocarbons on Fusion and Mechanical Performance of Polyethylene Gas Distribution Piping, May 1997.

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more elaborate filtering systems on the inlet to these instruments at a significant cost that will ultimately be borne by end-use customers. Standards for accurate natural gas measurement are predicated on various principles, including, for example, the absence of liquids. Introduction of hydrocarbon liquids may cause significant degradation of measurement accuracy, thereby leading to incorrect accounting and potentially distorted imbalances between suppliers and those entities receiving the natural gas.

1.4.12   When natural gas is processed to a specified level, the presumption of fungibility is sound and the original design basis of the pipeline infrastructure for managing incidental free liquids is appropriate. However, as processors elect to reduce extraction levels or not to process gas, such as times when natural gas liquids are at a discount to their value in the gas, the increase in liquefiable content may create a dilemma for transporters and end-users. The presumption of fungibility may no longer be appropriate. It is important to recognize that in pipeline systems designed to transport single-phase gas, facilities may not exist to prevent, or accumulate and remove liquids fallout. Any portion of the gas condensed into liquid may not only cause operational or safety problems, but may also result in loss of that portion of the energy quantity (dekatherms) in the process of transportation. The shipper will take receipt of the dekatherms contracted for with the pipeline. Energy lost during transportation because of liquid drop out must be made up by the pipeline in the short term. Where the liquids accumulate in the pipeline or associated equipment, the pipeline operator experiences shortages that must be made up to meet the natural gas demand. This results in increased lost and unaccounted for (UAF) gas. Ultimately, all shippers on the system must contribute their pro-rata share of the UAF.

Section 2 - Liquid Hydrocarbons in Natural Gas

2.1   Sources of Natural Gas Production

2.1.1   Natural gas produced from geological formations comes in a wide array of compositions. The varieties of gas compositions can be broadly categorized into three distinct groups:

- Associated Gas,
- Non-Associated Gas
- Coal Bed Methane.

2.1.2   These produced gases can contain both hydrocarbon based gases (those which contain hydrogen and carbon) and non-hydrocarbon gases. Hydrocarbon gases are Methane (C_1), Ethane (C_2), Propane (C_3), Butanes (C_4), Pentanes (C_5), Hexanes (C_6), Heptanes (C_7), Octanes (C_8), and Nonanes plus (C_9+). The non-hydrocarbon gas portion of the produced gas can contain Nitrogen (N_2), Carbon Dioxide (CO_2), Helium (He), Hydrogen Sulfide (H_2S), water vapor (H_2O), Oxygen (O_2), other sulfur compounds and trace gases. CO_2 and H_2S are commonly referred to as "acid gases" since they form corrosive compounds in the presence of water. N_2, He and CO_2 are referred to as diluents since none of these burn, and thus they have no heating value.

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2.1.3   Associated gas is produced as a by-product of oil production and the oil recovery process. After the production fluids are brought to the surface, they are separated at a tank battery at or near the production lease into a hydrocarbon liquid stream (Crude Oil or Condensate), a produced water stream (brine or salty water) and a gaseous stream. The gaseous stream is traditionally very rich (Rich Gas) in natural gas liquids (NGLs). NGLs are defined as Ethane, Propane, Butanes, and Pentanes and "Heaviers" (higher molecular weight hydrocarbons) (C_5+). The C_5+ product is commonly referred to as Natural Gasoline. Rich gas will have a high heating value and a high HDP. When referring to NGLs in the gas stream, the term GPM (gallons per thousand cubic feet) is used as a measure of hydrocarbon richness. The terms "rich gas" and "lean gas" are commonly used in the gas processing industry. They are not precise indicators but only indicate the relative NGL content.

2.1.4   Non-Associated gas (sometimes called "gas well gas") is produced from geological formations that typically do not contain much, if any, hydrocarbon liquids. This gas generally is lower in NGL content than Associated Gas. Non-Associated Gas can contain all, or none, of the other non-hydrocarbon gases identified above.

2.1.5   Coal Bed Methane is found within geological formations of coal deposits. Because coal is a solid, very high carbon content mineral, there are usually no liquid hydrocarbons contained in the produced gas. The coal bed must first be de-watered to allow the trapped gas to flow through the formation to produce the gas. Consequently, Coal Bed Methane usually has a lower heating value, and elevated levels of CO_2, O_2 and water that must be treated to an acceptable level, given its potential to be corrosive.

2.1.6 Gas quality can have significant effects on the operation of gas storage facilities. Three common types of gas storage facilities are mined salt caverns (either in a salt bed or a salt dome), aquifer, and depleted hydrocarbon reservoirs (geological rock formations). There are two significant ways for high HDP gas to create problems for storage operators. First, if the HDP specification is relaxed at any time, higher HDP gas could be injected into storage. A second cause of high HDP gas being in a storage facility could occur when low HDP pipeline gas is injected into a reservoir whereby it may become enriched if it comes in contact with hydrocarbon liquids existing in the reservoir prior to injection. The absorption is greatest during the first few years after a reservoir has been converted to storage and generally diminishes over time. In either case, when the gas is withdrawn from storage as a higher HDP gas, some of the liquefiable hydrocarbons can drop out through cooling of the withdrawn gas due to pressure reductions or contact with cold winter-time ground temperatures. In the first case, the injection of adequately processed gas would eliminate the problems associated with the withdrawal of high-HDP gas previously injected. In the second case, clean-up or "processing" of the withdrawn storage gas would need to be done at the compressor station used to inject and withdraw the gas from the storage reservoir.

2.1.7   Supply sources connected to interstate/intrastate pipeline systems are usually aggregated to a central delivery point (CDP) in the field through a gathering system. The CDP is the logical point where most gas processing occurs because of the aggregated volumes of gas. CDPs provide producers with economies of scale by centralizing facilities. It is not uncommon for larger CDPs to have connections to multiple interstate pipeline systems. Not all gas enters pipelines through CDPs. Pipelines sometimes have

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interconnects to one or several wells. The economics of conditioning gas from these sources can be problematic depending on the production potential of the well(s).

2.2 Role of Gas Processing

2.2.1 Gas processing is an important step in the journey natural gas makes from the wellhead to the burner tip. The gas processing function is commonly referred to as part of the Midstream Industry, a term used to describe the activities between Upstream – Exploration and Production, and Downstream – Gas Transportation and Marketing. Midstream companies are active in gathering gas from production facilities; aggregate the volumes; and treat and process the gas, before it enters the pipeline transmission system and downstream markets. Offshore, the produced gas enters the pipeline transportation system at the production platform and is transported to an onshore processing plant before being transported further to the downstream markets.

2.2.2 Produced gas can be partially treated at the wellhead to remove solids and liquids through simple, rudimentary physical separation equipment. This treatment is generally done to protect the gathering pipeline facilities used to transport the gas.

2.2.3 Gas processing entails two separate and distinct functions prior to the produced natural gas being deemed marketable. The gas will first be "treated" to remove major "contaminants" such as CO_2, H_2S and water vapor from the hydrocarbon gases if necessary and then, if there are sufficient levels of NGLs, the NGLs will be removed from the hydrocarbon stream.

2.2.4 Gas treating can be done on a stand-alone basis or in an integrated facility in conjunction with recovery of NGLs. Treating and integrated processing plants can be located at the terminus of gathering and aggregating systems. Alternatively, integrated plants can be found on a transmission pipeline near production areas. These plants are referred to as "straddle plants".

2.2.5 If H_2S, CO_2 and O_2 are present in the production gas, the first step is to treat the gas to reduce these gases to acceptable levels. Pipeline tariff specifications establish the acceptable level of contaminants for the pipeline and therefore the processor knows the degree of removal required to make an acceptable natural gas product. Processing plants often reduce the concentration of contaminants below pipeline standards in order to meet NGL product specifications. Water vapor is often reduced to extremely low levels as part of the low temperature extraction process. These gases are removed because they are potentially corrosive to the pipelines delivering the gas to the plant, to the processing equipment inside the plant and downstream transmission and distribution facilities.

2.2.6 Once the gas is cleaned of potentially corrosive gases, it can be processed to remove NGLs or it may be suitable for delivery into pipelines without further processing, as is the case of some non-associated gas and coal bed methane. In most offshore pipelines, natural gas condensate is injected with the gas produced on the offshore platform so that the combined gas and liquids are transported to shore in a single pipeline. This injected condensate, plus additional liquids that drop out as the gas is transported to shore, is

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removed by specially designed condensate removal equipment prior to gas processing or further pipeline transportation.

2.2.7   If the gas contains levels of nitrogen in excess of tariff limits or contains commercial quantities of helium, the next step in gas processing is to reduce the concentrations of these gases. To achieve this, cryogenic plant equipment is required. This is a very costly process, both in operating expense and capital investment. Recovery of helium and rejection (removal) of nitrogen are not commonly used processes and will not be discussed in any greater detail.

2.2.8   There are three common processes to recover NGLs: Refrigeration, Lean Oil Absorption and Cryogenic. Additional processes such as quick-cycle hydrocarbon adsorption are occasionally used and are becoming more common especially in situations with poor processing economics. Today, a processor will select the process to build after evaluating the richness of the gas, the appropriate technology for NGL recovery, market values of the natural gas and NGLs, the costs to get the NGLs to market, capital costs, fuel, and other operating costs. However, as discussed below, the infrastructure that exists today has been constructed over time. Older plants tend to be either refrigeration or lean oil while newer plants tend to be cryogenic. The quick-cycle hydrocarbon adsorption process was commercialized in the late 1940's. It targets recovery of the C5+ to meet hydrocarbon dew point specifications. It can be the simplest technology since no compression is required. A fixed bed of silica gel or other adsorption material is used to remove the liquids from the gas.

2.2.9   Refrigeration plants have the least capital cost but also recover the least NGLs. This process can extract a large percentage of propane and most of the C4+ gases and uses the least amount of fuel, compared to the other processes. The NGLs extracted from this type of plant are lower in vapor pressure and lends itself to trucking if pipelines are not available to move the NGLs to a fractionation plant. In the early days of gas processing, cruder forms of these plants and ambient lean oil plants were referred to as Gasoline Plants.

2.2.10   Lean Oil Absorption plants were the type of processing plant built in the 1960s. These plants were the next evolution from the refrigeration plants and can extract 90%+ of the C3+ in the gas stream and about 30% of the ethane by bubbling the gas through a chilled absorption oil operating at approximately -30°F. The fuel consumption of this type of plant is higher than that of the refrigeration plant. The ethane and propane were recovered to feed the ethylene plants at the infancy of the plastics and petrochemicals industries. Many of these plants are still operating and they straddle the large transcontinental gas pipelines built to transport the rapidly growing gas supplies found in the Gulf of Mexico and the eastern half of Texas during the 60's and 70's to markets in the northern and eastern parts of the U.S.

2.2.11   Cryogenic plants became prevalent in the 1970s as technology enabled higher ethane recoveries and demand for feedstocks increased to feed the growing plastics and petrochemical industries. These first generation cryogenic plants could extract up to 70% of the ethane from the gas, leaving a gas that was 90+% methane with the remainder being ethane and inert gases. To reach these higher extraction levels more expensive metallurgy, compression, and other capital investment are required. Since the early 1990s, modifications to the cryogenic process have allowed ethane recoveries to reach close to a 99% extraction

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level; still, due to the increased pressure reduction involved in the process, there is a higher operating expense due to the added fuel needed to run the compressors.

2.2.12   There are 556 facilities located in the United States that are engaged in processing of natural gas using the technologies described above, comprising approximately 68.4 billion cubic feet (BCF) per day of processing capacity (or approximately 25 trillion cubic feet (TCF) on an annual basis)[11]. There are 263 cryogenic facilities, 72 lean oil, 167 refrigeration and 44 using a quick-cycle or other technology. Approximately 50 percent of the available capacity is operated using cryogenic technology; with 20 percent being lean oil; 20 percent refrigeration and 10 percent other technology. Cryogenic facilities, generally being of a newer vintage are also larger with 100 of the 263 facilities being greater than 100 MMcfd, and 37 having capacities greater than 250 MMcfd. The refrigeration facilities are generally smaller with only 41 of the 167 facilities having capacities of greater than 100 MMcfd; there is one very large facility in Alaska that represents 60 percent of all of refrigeration capacity. The lean oil plants are also generally smaller with only 25 of the 72 being larger than 100 MMcfd.

2.2.12   Gas processing plants at times operated in reduced recovery modes to reduce the NGLs removed from the gas stream. However, the plants were designed to achieve high recoveries of all the NGLs and the "turndown" to lower recoveries has been difficult to attain. Typically, gas plants are not designed to recover only the C5+, or only the butanes, because they are designed to operate in a mode that recovers at least some percentage of all the components. In addition, it is not generally possible to operate the plants to achieve a specific HDP without blending of unprocessed gas.

2.3   Economics of Processing

2.3.1   The basics of NGL processing economics are to evaluate the amount of NGLs available to extract (which is determined by gas composition and the type of plant available to process the gas stream), determine the revenue generated from the sale of those NGLs and deduct the costs of processing. Processing costs include (1) the cost of the gas equivalent used or consumed in the conversion of production gas into NGLs (Shrinkage), (2) the fuel the plant consumes to operate the extraction process, (3) the payment or "processing fee" charged by the plant owner for this service, and (4) the operating costs for the plant. The shrinkage has value as a liquid product, but it also has value as natural gas if it had been left in the gas stream. Shrinkage and plant fuel are calculated both as a volume reduction and as a thermal reduction. Volume reduction occurs because the NGLs removed from the gas stream entering the processing plant and the plant fuel are not in the residue sales stream leaving the plant and therefore the residue gas is less than 100% of the inlet gas stream. Once the gas is processed, there is a gas value and a NGL value to the shrinkage part of the gas. The margin is the difference between the revenues received from selling the residue gas and NGLs, and the cost of the produced gas. If the NGL value less processing costs is greater than the equivalent gas value, then the margin is positive and it makes economic sense to extract the NGLs from the gas. On the other hand, if the NGL value as a liquid is less than the equivalent gas value, then the margin is negative and it does not make

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[11] Oil and Gas Journal, Annual Survey of Gas Processing, 2004.

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economic sense to extract the NGLs from the gas, except when the gas requires processing to meet pipeline specification and user need.

2.3.2   In the early years of the gas industry, producers sold their gas production to gas pipeline companies, normally through long term, fixed price contracts. Gas was generally considered a byproduct of oil exploration and production and a producer would take whatever value they could get for the gas instead of venting or flaring it. The production gas was processed in "Gasoline Plants" which simply compressed the gas, cooled it with either air or water to condense any heavy hydrocarbon gases, (i.e. Natural Gasoline) and then delivered the gas to a pipeline company. This Natural Gasoline was more valuable to the producer since it could be blended into and sold as a more valuable motor gasoline and removal of Natural Gasoline improved the operations of the pipelines. Once the majority of the heavy hydrocarbons were removed from this gas, pipelines took custody of the gas and transported it through their pipelines to markets elsewhere. As pipeline pressures increased, at times, more condensable hydrocarbons were removed at compressor stations and pipeline drips along the route of the pipeline. As pipelines moved gas to regions further from the producing region and the industry became more sophisticated in engineering and materials, gasoline plants began to chill the gas through simple pressure reduction/expansion, by passing the gas through light oil (absorption) or with refrigerants such as ammonia or propane. This evolution continued through the years and was influenced by the price of natural gas, NGLs, crude oil and by many government actions.

2.4   Influence of Ambient Temperatures and Pressure Reductions

2.4.1   Ambient ground and atmospheric temperatures and pressure reductions during transport or at a custody transfer point reduce flowing gas temperatures that in turn can result in hydrocarbon liquid drop out. Ambient temperatures become a concern when they are below *the flowing gas temperature and* the hydrocarbon dew point of a gas stream. Pressure regulation from a high-pressure to a lower pressure results in rapid cooling of the gas stream, a characteristic referred to as the Joule-Thomson effect.

2.4.2   Ambient ground temperature at pipe depth is one of the influential factors in flowing gas temperature. In general, the temperature of the gas exiting a compressor station ranges from 100 to 120 $^{\circ}$F. Once the gas leaves the compressor and travels underground, the temperature of the gas falls rapidly due to the difference between the ambient ground temperature and the flowing gas temperature. The potential for hydrocarbon drop out increases as the ground temperature becomes sufficiently cold as to approach or be below the hydrocarbon dew point. This concern exists in cooler climates where the pipeline may be above the frost line, the depth to which frost penetrates the ground and ground temperatures can reach 32 $^{\circ}$F. Pipelines located above the frost line may have flowing temperatures less than 32 $^{\circ}$F. Transmission pipelines located in the northern part of the country may have been installed at depths below the frost line where the flowing gas temperature is not likely to fall below freezing.

2.4.3   Ambient air temperature is another factor that affects the flowing gas temperature. When the pipeline moves above ground such as at a meter station, compressor station, or aerial crossings, the gas will be heated or cooled based on the ambient air temperature. The concern is whether there will be sufficient heat loss to cause the flowing

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gas temperature to go below the hydrocarbon dew point. Some larger gate stations or gas processing facilities utilize piping insulation where ambient temperature impacts present a specific concern. However, in most cases, piping is not insulated due to pipeline integrity program visual inspection requirements for monitoring atmospheric corrosion. In addition as a practical matter, historically, small diameter piping and appurtenances associated with LDC gas distribution operations are not typically insulated throughout the country due to minimal benefit insulation would provide relatively small surface areas.

     2.4.4   Water crossings also can affect the flowing gas temperature. Pipelines built today are often bored beneath rivers at depths below the mud line. At this depth the temperature of the river has no effect on the flowing gas temperature. However, most pipelines lay on or slightly under the riverbed. Under these conditions the water temperature can affect the flowing gas temperature. As long as the riverbed is not frozen solid, the underwater flowing gas temperature should not fall below 32 °F. It is important to note that many water crossings involve piping offsets that create "low points" which could result in liquid collection from hydrocarbon liquid dropout. If not designed to handle liquid collection, this could result in excessive pressure drops and flow restrictions and ultimately, may result in unscheduled shutdown and supply interruption.

     2.4.5   Pressure reductions such as those that can occur at a meter or regulation station can cause the flowing gas temperature to drop. The rule of thumb is that for every 100 pounds of pressure drop the gas temperature will drop by 7 °F (applicable up to 1000 psig). Thus, if the pipeline is delivering gas at a pressure of 800 psig to an end user who requires a pressure of 200 psig, the gas temperature will drop approximately 42 °F ((800-200)/100 *7) as the pressure is reduced. The example below shows the resultant flowing gas temperature for a delivery to a northern Indiana meter station in January.

Gas temp. based on historic ground temp.		*38 °F*
Temp. drop due to minimal above ground pipe	*less*	*2 °F*
Regulation from 800 psig to 200 psig	*less*	*42 °F*
Resultant gas temp. (without heating)		*-6 °F*

     2.4.6   The resultant low temperature demonstrates how pressure regulation can have significant influence on the flowing gas temperature. In some cases, heaters are used to raise the flowing gas temperature prior to regulation. These heaters are gas fired heat exchangers that heat the gas before it enters the regulator, thereby reducing the potential that hydrocarbon liquids and hydrates will form. The potential increase in temperature of the flowing gas depends on the type of heater employed. In the example above, if the operator or LDC used a heater that only raised the flowing gas temperature by 20 °F, the resultant flowing gas temperature would be 14 °F.

     2.4.7   It is common in the LDC distribution systems to regulate to operating pressures of 60 psig or less. In the example above, this generates an additional temperature drop of almost 10° F causing a resultant temperature of -16° F in the above example. With the added heat of 20° F, the new temperature will be 4° F. Even if a heater is part of the conditioning at a gate station, often further regulation is done immediately downstream into lower pressure systems. Operating at these low temperatures may result in hydrocarbon

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liquefiable condensation, solids blockage or service freeze-up of residential, commercial customers caused not only from potential hydrocarbon fallout but also from water vapor freezing to become a solid methane hydrate.

The formation of hydrates in the small orifices or tubing of regulator control equipment can cause disruption of supply to an LDC and in some cases stop operation of over-pressure protection equipment.

2.4.8   Gas heating prior to pressure reduction has been utilized throughout the industry at strategic locations for decades. The reasoning behind installation of supplemental gas heating equipment varies with industry segments. In general, all segments of the industry recognize that when possible, heating gas prior to a significant pressure reduction provides protection from hydrate formation as well as hydrocarbon liquid dropout. Historically, heaters were strategically installed by some LDC's to help control frost heave of mains and service lines due to subsurface ice formation and freezing of surrounding soils from temperature reductions associated with pressure reduction. In addition, heaters are installed to mitigate external ice ball formation on external piping and equipment surfaces that could interfere with proper operation of control equipment.

2.4.9 While gas heaters do indeed provide immediate protection from the abovementioned problems, gas heating alone should not be considered a system wide hydrocarbon dew point control. Gas heating addresses a specific process condition *at the point of installation* and they may not provide needed protection downstream or upstream. In addition, if heaters need to become more prevalent to reduce hydrocarbon liquid drop out, new as well as retrofit installations will be problematic due to community influences, air permitting, space availability and noise.

Section 3 - Hydrocarbon Liquid Drop Out Control Measures

3.1   Introduction

3.1.1   The 1971 AGA Gas Measurement Committee Report 4a, "Report on Natural Gas Contract Measurement and Quality Clauses" prepared by the Task Group on Gas Contracts established much of the standard Gas Quality language that was originally used in tariffs. It focused on the gas quality requirements that the seller had to meet when delivering gas to the pipeline, including specifications for liquids and solids and limits on non-combustibles or diluents. Many tariffs still contain the phrase originated in this document: "The gas shall be commercially free from dust, gum, gum forming constituents, and liquids at the pressure and temperature at which the gas is delivered." It also made recommendations for levels of water, H2S, total sulfur, CO_2, oxygen and heavy hydrocarbon content (using C5+ GPM (gallons per thousand standard cubic feet) as the reference) because these constituents in concentrations above recognized limits might be detrimental to pipeline integrity. Pipelines addressed the hydrocarbon content of the gas in a variety of ways, but at no time has there ever been a common set of specifications for components such as there has been for CO_2, H_2S and water. Much of this is due to the way the gas industry developed. It is also important to note that while most pipeline tariffs prescribe specifications for CO_2, H_2S and water, the exact specifications vary among pipelines.

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3.1.2   Many, if not all pipelines have minimum specifications for heating value. This resulted from gas historically being produced from fields high in N_2 or CO_2 and long ago from manufactured gas plants N2 and CO_2 are inert and do not have any thermal value, thus they dilute the natural gas and, when in sufficient concentrations, can cause an end user's appliance to experience flame instability. Approximately a third of all interstate pipelines specify a maximum heating value, but there is no differentiation as to whether this is a condition at the receipt point or delivery point. The use of a maximum heating value is an inadequate predictor of hydrocarbon liquid drop out because a gas can have a relatively low heating value and a high C6+ content that can exhibit an elevated HDP and result in hydrocarbon liquid drop out. Conversely, a gas with an elevated ethane level will have a high heating value but a low HDP if the C6+ content is low.

3.1.3   Some pipelines selected another parameter for controlling liquids fallout by establishing a C5+ GPM, C5+ mole percent or C6+ GPM specification. A C6+ GPM specification may in some instances be used as an indicator of the potential for hydrocarbon liquid drop out but as will be discussed in Section 6, there are problems in applying this measure broadly. The C6+ composition varies among gas streams and has the largest effect on the hydrocarbon dew point. It also provides a good indication of liquid volume levels that may condense from the gas if the gas temperature falls below the HDP. By itself, however, the use of a C6+ GPM specification alone does not ensure that the flowing gas will not enter into a two-phase region and cause liquids to drop out. Nonetheless, correlating C6+ GPM levels to HDP can be used as a screening tool or as the basis for establishing a control limit.

3.1.4   More recently, some pipeline operators have elected to establish hydrocarbon dew point limits. As of June 2004, eleven interstate pipeline operators had established currently approved tariff hydrocarbon dew point limits. Three other operators have proposed hydrocarbon dew point limits. An HDP limit can be used to provide a wide range of gas compositions to end-users without compromising the safety, operational reliability, system integrity or environmental compliance within the natural gas infrastructure. The use of an appropriate hydrocarbon dew point specification will provide the information necessary to operate transmission and distribution systems and processing plants.

3.2   Blending

3.2.1   Blending is the mixing of gas streams that yields a volume-weighted average of the concentrations of each constituent. Pipelines and their customers have benefited from blending for years to make the combined quality of its gas stream from the individual gas streams meet gas quality related requirements. Blending has specifically been used in some instances for controlling Btu content and to meet other gas quality requirements. As the industry moves to implement new quality specifications relating to hydrocarbons, blending can play a role in managing hydrocarbon levels and provide the potential to accommodate receipts of gas with varying levels of hydrocarbons. While the overall goal remains to prevent pipeline condensate from forming, blending provides pipelines a mechanism to achieve that goal while still maintaining the flexibility to accept gas streams with varying hydrocarbon levels. In any case, the ability to utilize gas blending to manage HDP is dependent on a number of factors including pipeline configuration, receipt and delivery location, gas supply

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composition, gas markets and shippers' nominations, supply disruption, geographic location, and flowing gas temperature.

3.2.2   Each pipeline operator using blending will define the methodology and process for blending and monitoring the resulting hydrocarbon levels based upon their specific operations. The HDP of the mixed or commingled stream will depend on the volumes and compositions of the two blended streams.   In other words a small amount of low HDP gas will not reduce the HDP temperature of a large volume of high HDP gas significantly. Conversely, even small amounts of high HDP gas when mixed into a low HDP gas can significantly raise the HDP of the mixture. When two different gas streams are mixed/blended, each compositional component of the commingled stream changes the equilibrium of the new mixture creating a completely separate and unique gas quality.

3.2.3   There are two distinctively different types of blending: physical and contractual.

3.2.4   **Physical blending** is when two or more gas streams are mixed together prior to being introduced into or within the pipeline. The combined stream changes in physical composition as discussed above. The blended gas streams may not however, thoroughly mix when combined. It may take some distance and possibly compression or some other mixing event, before they truly become a homogenous blend.

3.2.5   **Contractual blending** is when a producer of rich gas contracts with a lean gas producer or a processing plant upstream of the rich gas producer to reduce its HDP by blending where the resulting HDP is lowered to meet a specific HDP limit through agreement with the pipeline operator. These two volumes may enter at different parts of the pipeline and may not directly blend in the pipe. As such this type of blending does not work on all pipelines. But in theory, the two gas streams do actually blend prior to delivery by the pipeline if they both ultimately flow in the same segment of pipeline. In this type of blending, the overall blended stream of each pipeline segment or area must still meet the pipeline's required limit prior to being delivered. Even though the two combined streams may meet the HDP limit set by the pipeline, the pipeline may not approve this type blending if a section of the pipeline has a HDP limit that cannot be met by one of the contracting parties.

3.3   Heaters

3.3.1   Problematic hydrocarbon condensation often occurs at points of pressure regulation (or immediately downstream). In some cases, water bath heaters can be used to increase the flowing gas temperature prior to pressure regulation. In a water bath heater, as the name indicates, water surrounds and provides heat to a tube bundle (heat exchanger) containing the flowing natural gas, Since these units burn natural gas, they require air permitting. In addition, since pressure regulation often occurs post custody transfer from the pipeline to the LDC or other end user, gas heaters may not be practical or even feasible due to space limitations in urban environments. For example, a measurement and regulating (M&R) station heater could be as large as 8 feet in diameter and 20 feet in length which is simply not practical to install in a small subsurface modern pressure reduction station. As previously mentioned, gas heaters provide specific process temperature control only at the

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point of installation. As a result, some LDC's or end users partially depend on pipeline heat of compression to mitigate temperature decreases associated with local pressure reduction.

3.4   Offshore Gas and Liquids Handling

3.4.1   Handling gas and liquids in the offshore environment is different than onshore gathering because of the way gas and condensate is handled. In the offshore environment most gas pipelines allow for the produced fluids to be separated at the offshore platform, then the condensate is re-injected into the pipeline after the gas is metered so that only one pipeline is necessary to transport both condensate and gas from offshore. Also, since the gas is additionally cooled as it flows in the underwater pipeline systems, additional liquids, commonly called retrograde condensate, are generated by the time the gas arrives at the onshore separation and processing facilities. These liquids must be removed before the gas can be processed for NGL recovery or further transported to market.

3.4.2   When natural gas and condensate are present together in a pipeline, or pressure vessel, they are likely to be intimately mixed and reach a point of "equilibrium" or a saturation point. The gas stream is at its hydrocarbon dew point at the temperature and pressure of the pipeline or pressure vessel. So any time this offshore sourced gas is delivered to a pipeline without processing or without further hydrocarbon content reduction utilizing a JT plant[12], it is most likely to be at its dew point and any cooling of the gas from the ground or water temperature or a pressure reduction (like a pressure regulator) can condense liquids.

Section 4 - Overview of Hydrocarbon Dew Point

4.1   Introduction

4.1.1   The hydrocarbon dew point (HDP) defines whether the natural gas stream in a pipeline at a given pressure and temperature consists of a single gas phase or two phases, gas and liquid. The HDP is defined as the series of matching pressure and temperature points at which hydrocarbons condense into liquid from a natural gas mixture. The hydrocarbon dew point **pressure** is the pressure at which hydrocarbons will begin to condense from a gas mixture at a given temperature. The hydrocarbon dew point **temperature** is the temperature at which hydrocarbons will begin to condense from a gas mixture at a given pressure, and it is usually more important for pipeline operations where the pressure is determined independently.
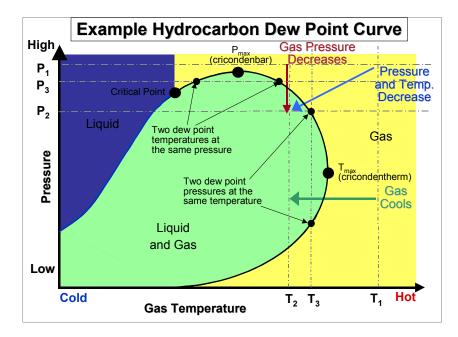
4.1.2   When condensate forms from a gas mixture, the distribution of hydrocarbons changes so that the liquid phase becomes enriched in the heavier components while the gas phase becomes depleted of these heavier components. As the gas is cooled

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[12] A J-T valve has the least capital cost but also recovers the least amount of NGLs. This simple process is used mainly to control HDP temperatures and primarily recovers the C_5 components only. The J-T process (or Joule-Thomson) involves cooling a gas stream by reducing its pressure (adiabatic expansion) through a control valve. Produced liquids are recovered in a cold separator and the gas stream off the top of the separator is used to cool the inlet stream to the J-T valve. This process may require considerable compression to achieve the desired pressure drop across the J-T valve thus resulting in high operating costs.

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below its original dew point temperature, the entire dew point curve shifts cooler for the remaining gas phase that is now depleted in heavier components. The chilled gas temperature becomes the new HDP of the gas stream.

4.2   Hydrocarbon Dew Point Curve

4.2.1   The HDP for natural gas with a given composition is typically displayed on a phase diagram, an example of which is shown in Figure 4-1. The HDP curve is plotted as a function of gas pressure (P) and temperature (T).   The left-hand side of the curve (in blue) 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 (in black) is the dew point line and divides the two-phase gas-liquid region 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. Note that two dew point temperatures are possible at a given pressure (P_3) and two dew point pressures are possible at a given temperature (T_3). This phase envelope phenomenon provides for behavior known as retrograde condensation. The retrograde phenomenon occurs when liquids form at a given temperature when the pressure is lowered (see red arrow). The word "retrograde" means moving backward and this phenomenon was given the name because it is contradictory to the phase behavior of pure components, which condense with increasing pressure and or decreasing temperature. The maximum pressure at which phase change occurs (P_{max}) is called the **cricondenbar**, and the maximum temperature (T_{max}) at which phase change l occurs form is called the **cricondentherm**.

Figure 4-1 - Hydrocarbon Dew Point Curve for a Typical Natural Gas Mixture

4.2.2   The HDP is a function of the composition of the gas mixture and is strongly influenced by the concentration of the heavier hydrocarbons, especially C_6+. The presence

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of heavier hydrocarbons will increase the HDP and failure to include them in a HDP calculation will under–predict the HDP. For most pipeline conditions, the HDP temperature at a given pressure increases as the concentration of heavier hydrocarbons increases. Thus, the **potential** to form liquids at certain pipeline conditions exists for gases rich in C6+. Processing of the gas stream primarily removes or extracts heavy hydrocarbons and thus reduces the HDP of a given mixture. The level of hydrocarbon removal directly impacts the HDP.   Figure 4-2 shows examples of the HDP curve for unprocessed and processed gas mixtures. The unprocessed HDP curve is in red and has a higher cricondentherm temperature while the processed HDP curve is in blue. The difference between the two curves shows the impact of processing on the HDP.
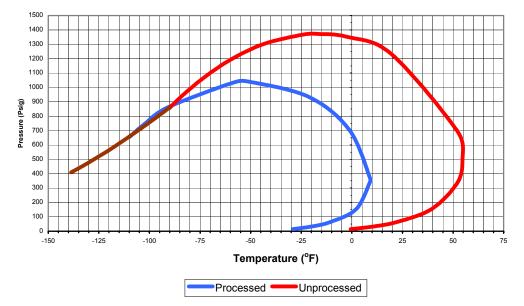

Figure 4-2 – Contrast of Unprocessed and Processed Natural Gas

The significance of the HDP curve for gas transmission and distribution operations lies in the potential transition from the single-phase gas region to the two-phase gas-liquid region. For example, the arrows in Figure 4-1 (Figure numbers to be corrected) show changes in pipeline pressure and temperature in which the end-point lies inside the gas-liquid phase. In this situation, condensate formation inside the pipeline will occur. It is important to recognize, however, that the volume of condensate **cannot** be determined simply by plotting points on the HDP curve. The volume of condensate can be determined by analyzing the gas phase compositions upstream and downstream of a potential condensation location (e.g., regulator, pipeline) and determining the GPM (gallons of liquids per thousand standard cubic feet of gas) for the liquefiable components in each stream.

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Section 5 - Historical Levels of Hydrocarbons and Hydrocarbon Dew Point

5.1 Introduction

5.1.1   There is not an abundance of historical data on hydrocarbon levels or hydrocarbon dew points. However, approximate ranges for cricondentherm hydrocarbon dew points can be estimated by referencing the types of processing in the gas industry from the 1940s to the present.

5.1.2   Prior to the advent of gas processing, hydrocarbon dew points in pipelines and market areas would approach ambient temperatures (between 30 and 60 ^{o}F). Pipelines collected liquids and developed their own methods to force these liquids from the gas prior to its delivery to customers. As uses for gas progressed beyond crude lighting and cooking appliances, processing developed and operators were able to discontinue these practices and remove equipment. The first gas processing plants were really compression plants similar to air conditioning units and operated prior to the advent of refrigeration plants. They compressed the casing head gas and cooled the gas using air or water heat exchangers to condense the heavy NGLs. This resulted in recovery of approximately 25% of the C6+ and reduced the cricondentherm to about 10 °F at the plant outlet.

5.1.3   Propane as a refrigerant became available post 1940 when demand for butane for use in motor gasoline increased and rural heating was converted from butane to propane in the 1940s-1950s. Refrigeration dropped the cricondentherm from 80 °F at the plant outlet to slightly above 0 °F and recovered 50% or so of the propane and 80% or more of the C4^{+}. The gas processed was very rich casing head gas on the order of 1200 -1400 Btu per cubic foot (HHV), (4 - 7 GPM), as this gas was a by-product of oil production.

5.1.4   As ethane became a valued commodity in the early 1960s, the new onshore oil and gas fields discovered in that era had lean oil plants built in the same geographic regions. These lean oil plants had somewhat higher NGL recovery than the older refrigeration plants (70+% of the propane, 90% of the C4+). Cricondentherm of the processed gas was -30 °F or lower.

5.1.5   When oil prices increased dramatically after the 1972 oil embargo, there was a strong economic incentive to recover all the NGLs. During this same time period, cryogenic processing technology developed where it became more economical. Cryogenic plants generally recover from 60% to 99% of the ethane (depending on the technology employed) and essentially all of the C3+ producing cricondentherm temperatures of -100 °F or lower. Due to the increase in value of the NGLs, cryogenic technology was retrofitted at many of the larger, older onshore refrigeration plant sites in the late 1970s to replace the lower recovery refrigeration plants. The lean oil plants built in the 1960s continued to operate until field declines in the 1980s and 1990s, coupled with increased operating expenses, justified the shutdown of some of these older plants. The remaining production formerly processed in these plants was consolidated with other production field gases and processed in the newer plants linked together to create regional processing centers (i.e. the Duke Energy Field Services Oklahoma Super System, the Williams Energy Opal, Wyoming complex, etc.) that afforded the operator a way of efficiently utilizing available capacity.

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Other major changes occurring in the 1990s were the spin-off of producer plant assets to new business entities such as Dynegy, Enterprise Products, GPM, Tejas, UPR Fuels and many of the "Field Service Companies" such as Duke, El Paso and Williams). Since 2000 few new processing facilities have been built. As new facilities are built, the industry has utilized additional plant designs that minimize recoveries of the lighter hydrocarbon constituents (C1 through C3) to maximize sales of gaseous hydrocarbons. These plants are designed more to control HDP than to recover large quantities of NGL's. The three processing technologies that have become popular for this application are refrigeration, short cycle adsorption (molecular sieve) and JT (Joules-Thomson) skids. This trend will probably continue into the future as more pipelines institute HDP specifications and if processing margins are negative due to natural gas prices remaining high relative to NGL's.

5.1.6   On the Outer Continental Shelf (OCS), the Minerals Management Service (MMS) promulgated regulations in the late 1950s that eliminated routine flaring of gas production, primarily associated gas. This created a huge pipeline construction boom to recover the formerly flared gas along with construction of the large lean oil straddle plants on these new pipelines from OCS. As production grew in the 1970s on the OCS, the oil embargo and consequent increased prices provided the incentive for increased NGL recovery. Straddle plants for new pipelines built in the mid 1970s (e.g., Blue Water, UTOS and Sea Robin) employed cryogenic technology while the older plants on the other pipelines were not retrofitted.

5.1.7   In summary, available gas processing technology would have the following approximate cricondentherm HDP at the plant outlet:

Technology Vintage	Processing Technology	Achievable Cricondentherm HDP °F
1940-60	Refrigeration	$\cong 0$
1960-75	Lean oil	$\cong -10$
1975 on	Cryogenic	$\cong -100$

Actual cricondentherm HDP in any pipeline at any point in time is determined by the mix of processed and unprocessed gas and the degree of processing of the processed gas.

Section 6 – Determination of Hydrocarbon Dew Point – Measurement and Estimation

6.1   Introduction

6.1.1   This section provides an overview of the determination of hydrocarbon dew point. It can be done in two ways, measurement or estimation. A method referred to as the "chilled mirror" is used to conduct direct determination of the hydrocarbon dew point. Alternatively, **indirect determination** relies on a combination of sampling, analysis and calculations using a simplified equation of state from chemical thermodynamics. This section provides an overview of the merits of each in managing hydrocarbon drop out. Determining the exact temperature that a vapor component in the gas stream condenses

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does not in and of itself define the basis for controlling hydrocarbon liquid drop out. Knowing the temperature when appreciable amounts of liquids will condense is a useful operational tool. For a procedure establishing a cricondentherm hydrocarbon dew point (CHDP) limit, see Appendix B.

6.2   Direct Determination

6.2.1   The most commonly used direct method of hydrocarbon dew point determination is with a chilled mirror, also known as a dew point tester. The method was developed by the U.S Department of Interior, Bureau of Mines and has been codified into a standard test method by the American Society of Testing and Materials (ASTM)[13]. For many years this device has been used for moisture measurement. A standard for chilled mirror hydrocarbon dew point measurement has also been developed and will appear in the next revision of the American Petroleum Institute (API), Manual of Petroleum Measurement Standards (MPMS) - Chapter 14.1. The device can be used to determine the hydrocarbon dew point at the operating pressure at a specific field location. Some pipeline operators use this as a means of HDP determination and verification. The major advantage of this device is that it provides direct measurement of HDP at a specific operating pressure. For this reason, a pipeline may elect to use chilled mirror as the primary method to determine HDP at a specific pressure, if mutually agreed upon by relevant parties. However, it may not be applicable for determining the cricondentherm. If the cricondentherm is above the operating pressure, the analyst will not be able to determine the exact value. If the cricondentherm is below the operating line pressure, it may be possible to throttle down and determine the cricondentherm but it may take multiple measurements and a considerable amount of time.

6.2.2   The Bureau of Mines dew point tester consists of a small high-pressure chamber (5000 PSI max) through which the gas sample flows. A polished stainless steel mirror is at one end of the chamber and a viewing window is at the other. The chilled mirror is cooled by a refrigerant system. The operator throttles the gas flow through a valve and cools the polished mirror until the hydrocarbon dew point is observed by the formation of a thin film of droplets. The temperature and pressure are then recorded and plotted on a graph.

6.2.3   Determination of the HDP temperature with this apparatus is a subjective test that requires the analyst to watch for the formation of hydrocarbon liquid droplets as the mirror is gradually cooled at the rate of one degree Fahrenheit per minute. This is a very time intensive and tedious process. Chilled mirror dew point testers can be used to determine both water vapor dew point and hydrocarbon dew point. The two types of dew points can be distinguished from one another by the unique location and size of the liquid droplets that form on the mirror surface. These differ because of differences in surface tension between liquid water and liquid hydrocarbons.[14] It may be difficult to distinguish whether the droplets are due to water or hydrocarbons, particularly when the dew points

[13] ASTM D 1142-95. 1995. "Standard Test Method for Water Vapor Content of Gaseous Fuels by Measurement of Dew-Point Temperature," Am Soc for Testing and Materials, Philadelphia.
[14] API Manual of Petroleum Measurement Standards, Chapter 14.1 HCDP Measurement Standard

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either overlap or are within just a few degrees apart. A typical HDP test may last forty-five minutes to one hour and requires uninterrupted attention to the test apparatus.

6.2.4   Experience indicates that trained and experienced operators can generally reproduce each other's results. Among inexperienced operators the results may vary significantly. Even the most skilled operator may make an error due to the appearance of water droplets, methanol droplets, or glycol droplets on the mirror if these exist in the gas sample stream.

6.2.5   Automatic, continuous online dew point detection units are commercially available. These units are expensive relative to the cost of other online analyzers and unlike gas chromatography; these instruments are currently not part of the existing gas quality analytical infrastructure. The decision to deploy them entails consideration of the economics of purchase, installation and maintenance of the online analyzer versus the use of estimation (described below) in conjunction with periodic manual dew point measurements.

6.2.6 In summary, direct reading instruments are useful tools in determining localized observable hydrocarbon dew points at specific pipeline operating conditions. While this technology has proven useful in the field in diagnosing hydrocarbon dew point conditions, universal application of this technology across the grid may prove challenging due to the variables highlighted above in addition to the wide variety of operating conditions that exist across the nation in all sectors of the industry.

6.3   Indirect HDP Determination

6.3.1   Indirect HDP determination relies on a three-step process, sampling, analysis and calculation. The most common means of sampling and analysis (the first two steps) involves a continuous online system. Permanent sample probes (isokinetic) are installed in the pipeline to obtain a representative sample. The sample probe is connected to a heated sample line that transports the gas to a continuous online gas chromatograph. The most common chromatograph found in field applications uses a combination of columns to analyze for methane through pentane and then treats all compounds with molecular weights greater than pentane as a C6+ fraction, generally using a fixed mole fraction average of C6, C7, and C8[15]. This chromatograph is referred to as a C6+ chromatograph.

6.3.2   Manual sampling with off-site analyses of the samples can be used as an alternative. Samples of the gas are collected in a clean sample cylinder (canister) or on charcoal tubes using standard methods published by GPA[16] and referenced by API[17]. Samples are analyzed using a chromatograph typically using C6+ chromatograph described above. Some labs have chromatographic equipment to analyze to C_9 to C_{11}.

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[15] American Society for Testing and Materials standard, ASTM D 1945, and Gas Processors Association, Standard 2261, Analysis for Natural Gas and Similar Gaseous Mixtures by Gas Chromatography.

[16] Gas Processors Association Standard 2166, "Obtaining Natural Gas Samples for Analysis by Gas Chromatography," 1986.

[17] American Petroleum Institute, Manual of Petroleum Measurement Standards, Chapter 14, Section 1. "Collecting and Handling of Natural Gas Samples for Custody Transfer," June 2001.

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6.3.3   The third step, calculation, is conducted by applying thermodynamic principles and accepted equations of state using the gas analysis from above. An equation of state defines the relationship between state variables (pressure and temperature) and gas properties such as density. Two commonly accepted sets of state equations are Peng-Robinson[18] and Soave-Redlich-Kwong[19]. The effects different calculation methods and of gas composition accuracy on the accuracy of Equations of State are also part of an on-going project with the API Chapter 14.1 Working Group.

6.3.4   The degree to which the three-step process reflects the actual hydrocarbon dew point is dependent upon several factors including the characteristics of the natural gas stream, how well the sample represents the stream composition, the chromatographic equipment, how the heavier hydrocarbons are input into the equations of state, and the equations used. HDP are most sensitive to the mole percentage compositions of the hydrocarbons larger than hexane. In applying a HDP limit using C6+ data, it is prudent to conduct periodic validation based on use of an "extended analysis", through C8 to enable demonstration of the "split". The split is the relative proportion of C6 C7, and C8 in a gas mixture. Some commonly used values for these percentage characterizations are published in a GPA standard[20]. However, recent research[21] has shown that use of the GPA 60/30/10 C6/C7/C8 characterization to compute hydrocarbon dew points will usually underestimate the dew point temperatures and cricondentherm. However, the work conducted by a leading chromatograph manufacturer, Daniels, indicates that a 47:36:17 split is generally applicable, but may vary depending on the source of gas or the degree the gas has been processed[22]. The determination of the appropriate characterization for a given pipeline system may be more accurately derived from the weighted average compositions of the regional supply on that pipeline. An alternative approach is to widen the regional observation, such as including all Gulf Coast production. Such a definition may span several operating pipelines in the region. However, the ability of the average characterization to reflect the true composition of a particular gas within a region depends on the variance of the individual components of all gases throughout that region.

The presence and amount of C9+ components is important in determining the HDP as well. Amounts as low as 0.001 mole percent C9 can have significant impact on the calculated HDP. While characterization data available for C9+ show that these components are generally not present or when present are found in relatively small amounts, it is prudent in applying the indirect method to characterize the C9+ fraction as part of the periodic validation process.

6.3.5   Determining the exact temperature that a vapor component in the gas stream condenses is not of as much value as knowing the temperature that an appreciable amount of liquid condenses. For example, if a sample has 0.001 mole percent of C10, the HDP may

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[18] Peng, D. Y. and Robinson, D. B., Industrial and Engineering Chemistry Fundamentals, 15: 59, 1976.

[19] Soave, G., Chemical Engineering Science, 27: 1197, 1972.

[20] Gas Processors Association Standard 2261, "Analysis for Natural Gas and Similar Gaseous Mixtures by Gas Chromatography," 2000.
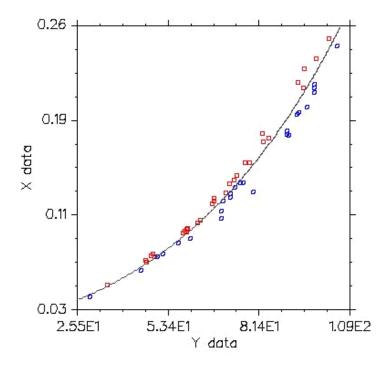
[21] D. L. George et al., "Metering Research Facility Program: Natural Gas Sample Collection and Handling – Phase IV," Gas Research Institute Report No. GRI-03/0049, September 2004.

[22] Standard configurations programmed into Daniels chromatographs, Daniels, a division of Emerson Electric.

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be relatively high, but the volume of liquids contributed will not affect pipeline operations. That is why using a volume number (i.e., GPM) can be of value to pipelines. Developing the relationship between the GPM and HDP is useful, since engineers design and operate facilities around temperature and pressure (7 degrees F per 100 psi reduction). Developing a correlation between C6+ GPM and a CHDP can be done from a gas analysis. This requires a number of different gas streams to be analyzed, using a gas chromatograph. The HDP is calculated for each stream as described in section 6.3.3 above and the C6 + GPM value is calculated by summing the GPM value for each C6+ component using the split methodology described in Section 6.3.4 above. The GPM value for any component in a gas stream is calculated by taking a component's mole fraction (mole percent/100) and dividing it by the volume constant (ft^3) ideal gas/gallon liquid from GPA Standard 2145 and then multiplying by 1000 (scf/Mscf). Once the HDP and C6+ GPM values are determined for the each receipt point in the group, these values can be input to a curve-fitting program. The program can yield a simple equation for determining HDP from the C6+ GPM level. One major advantage to using C6+ GPM is that it correlates to the volume of gas flowing into the pipeline and allows pipelines to determine how much flow reduction from each point will be necessary to maintain an acceptable HDP. To demonstrate the simplicity and accuracy of this approach, a set of 55 unprocessed receipt point samples were selected. The HDP and C6+ GPM values were determined utilizing a Peng-Robinson equation of state program and a 40/40/20 C6+ split. The simplified curve fit equation developed was in the form of

$$y = ax^b + c$$

Where y = HDP, x = C6+GPM
(Constants a, b, and c are 392, 0.159, -210 respectively)

The results are shown in the graph below

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As is seen in the graph, this is a simple method for approximating CHDP and is a way to relate gas quality to gas volumes on a pipeline.

Another practical alternative that draws upon the strengths of direct and indirect HDP determination measurement is being evaluated to predict HDP with lesser uncertainty than the either of the direct or indirect methods[23]. Further studies are required to validate this alternative method.

6.3.6   When indirect determination methods are included in a quality specification, there must be a declaration of the equations of state that will apply, the assumptions for Hexane plus composition, and the hydrocarbon dew point temperature. In addition, assumptions made about the relative proportions of hydrocarbons greater than pentane (i.e.- the hexanes plus fraction) must be validated on a periodic basis.

Section 7 - Recommendations

7.1   Technical

7.1.1   Control of hydrocarbon liquid drop out requires use of a control parameter to ensure operational safety and reliability, system integrity, environmental compliance and to minimize impacts on end use equipment.

7.1.2   Of the various methods which could be used as a control parameter, the NGC Task Group found that the C_5^+ approach and the Heating Value approach were not effective means of predicting and controlling hydrocarbon liquid dropout and should not be used as control parameters.

7.1.3   The NGC Task Group found that cricondentherm HDP and C6+ GPM specification were both valid for use as control parameters to control hydrocarbon liquid dropout and recommends using an equation of state with data derived from gas chromatography for calculation of cricondentherm HDP or C_6+ GPM specification.

7.1.4. The NGC Task Group, however, found that using the cricondentherm HDP as the control parameter offered the greatest operational flexibility for all stakeholders.

7.1.5. If the C_6^+ GPM approach is used as the control parameter, then it must be understood that this approach will not give the end-user all of the information needed to design, install and operate their equipment outside of the two-phase region of the gas stream.

7.1.6   The Task Group recommends that, when using a cricondentherm HDP or C_6+ GPM, a plan must be established by the pipeline operator for periodic validation of the assumptions used including proportions of C6 ,C7 & C8, and where applicable, higher molecular weight hydrocarbons.

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[23] Starling, Kenneth A., Peng-Robinson Equation of State Natural Gas Dew Points, AGA Technical Conference

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7.1.7 The Task Group recognizes that determining HDP using the Bureau of Mines method is not practical for automated applications, and is the subject to the practical limitations described in section 6.3.

7.1.8   The NGC Task Group recognizes that in certain instances, parties may be able, to the extent operationally feasible, to change control parameter limits based on ambient conditions, storage operations, meter station and system pressure drops, and the tolerance for heavy hydrocarbon levels within a specific market area, among others.

7.1.9   The NGC Task Group recommends that additional research be conducted in the following areas:

      1) Build the database to support use of C6+ split assumptions for heavier hydrocarbons, develop better correlation between direct and indirect HDP determination and to improve the accuracy of commonly used equations of state.

      2) Develop a cost effective hydrocarbon-specific direct-reading dew point analyzer because a conventional chilled mirror direct measurement instrument in general can be subject to operator variability and interferences including but not limited to water vapor.

7.1.10 The HDP limits do not presume that gas is Interchangeable. HDP is only one facet of describing gas quality.

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APPENDIX A
PARAMETERS TO BE CONSIDERED IN ESTABLISHING CHDP OR C6+ GPM BASED LIMITS

The Work Group defined a set of parameters that may be useful, in establishing the CHDP or C6+ GPM required to avoid hydrocarbon liquid dropout. The parameters are:

- Minimum Flowing gas temperature

- Minimum Ambient air temperature

- Minimum Ambient ground temperature

- Operating pressure requirements

- Pressure reduction

- CHDP levels of gas supplied including those of downstream pipelines

- Experience with monitoring HDP levels and associated problems caused by hydrocarbon liquid drop at various levels

- Presence of heating systems

- Presence separation equipment

- Prevailing and expected flow patterns

- Impact of storage

- End user applications

- LNG liquefaction peak shaving feedstock requirements

The Work Group recognized that implementation of CHDP may require incremental changes to establish a more flexible CHDP as additional data and experience are gained by pipelines and LDCs.

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APPENDIX B
PROCESS FOR ESTABLISHING A
CRICONDENTHERM HYDROCARBON DEW POINT (CHDP) LIMIT

1. Define an area for which the limit is to be applied (e.g.- market area, energy zone).

2. Review historical data of the area for composition, flowing gas temperature and pressure of delivered gas.

3. Select a candidate CHDP limit based on historical gas quality data
   - Use the full compositional analysis at least through C6.
   - In order of preference, use:

     i. A C6+ split to be established and periodically validated through extended analyses in a C9+ chromatograph, as specified in the White Paper on Control of Hydrocarbon Liquid Drop Out, or
     ii. A 47:36:17% C6/C7/C8 assumed split specified in the White Paper on Control of Hydrocarbon Liquid Drop Out or other published split applicable to a specific region.

4. Develop a phase diagram that represents the gas at the selected CHDP.

5. Apply a line of constant slope that is tangent at a single point to the phase diagram. The slope of the line is the Joule Thomson constant, i.e. – approximately seven (7) degrees of temperature drop per 100 pounds per square inch of pressure drop. This is referred to as the J-T line.

6. Identify the lowest temperature and coinciding highest pressure of flowing gas at each place of pressure reduction and plot the corresponding point on the phase diagram.

   - Consider the effects of existing equipment, such as gas heaters, multi-stage pressure reduction equipment, etc.

7. Applications where the temperature/pressure points fall to the right of the J-T line should not experience liquid drop out.

8. Applications where the temperature/pressure points fall to the left of the J-T line may experience liquid dropout. To prevent hydrocarbon liquid drop out for such applications, either reapply steps 3 through 6 by selecting a lower candidate CHDP or consider alternatives including installation of gas heating or use multi-stage pressure reduction.

9. A review of the established CHDP should be made from time to time as more experience is gained.

   The use of the phase diagram and the J-T line as the bound for liquid drop out provide a reasonable basis to establish a CHDP limit.