Section 1 - Introduction

Historically, the commercial value of the liquids (NGLs) extracted from natural gas has been greater than the thermal value of the liquids in the gas. The infrastructure to extract these liquids, referred to as processing has been built up over time. But, recently, the value of natural gas has increased dramatically as compared to the value of the natural gas liquids. Rising natural gas prices have reduced the economic incentive to extract NGLs, such as ethane, propane and butane, from domestically produced natural gas. For example, in 2000, natural gas prices averaged $3.88 per million British Thermal Units (MMBtu) while natural gas liquids prices averaged $5.44 per MMBtu. With pricing at these levels there is clear economic incentive to process gas and recover the NGLs. However, in 2003, natural gas prices averaged $5.38 per MMBtu whereas NGLs were valued at $5.26 per MMBtu[1]. This environment remains with us into 2004. In this economic environment producers may elect to bypass processing.

This economic environment creates two issues for domestic natural gas. The presence of larger amounts of liquefiable hydrocarbons results in higher heating values and a greater volume of liquids. The increased heating value increases the potential for problems in end-use applications such as in home appliances or gas turbines used to generate electricity. The greater volumes of liquids increase the potential for hydrocarbon dropout in the pipeline during transportation. These problems are not likely short-term issues, as natural gas demand in the United States is expected to rise, especially as demand continues to increase for natural gas-fired electric generation.

Objective

The objective of this report is to provide background on the issue of **gas quality** and characterize how it can be managed in a way that balances the concerns of all stakeholders in the value chain[2]. These concerns are summarized below:

Producers want the ability to provide as much natural gas as is possible, especially given increasing demand.
Gas Processors want to know the requirements for the quality of gas to be introduced into transmission pipelines.
Pipelines want to provide flexibility to meet demand but are concerned about operational reliability, system integrity and ultimately safety, given the potential for hydrocarbon drop out.

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[1] Energy Information Agency, U.S. Department of Energy

[2] 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 Force.

Local distribution companies want to meet customer demand but are concerned about operational reliability as well and the impacts on end use equipment.
Direct connect customers are concerned about operational reliability and increased emissions.
End Users are concerned about the uniformity of gas quality for appliances, industrial applications including use as a feedstock or building block in chemical manufacturing.

Background

We begin this report with a brief description of how natural gas makes it way from wellhead to the burner tip. Natural gas is produced from one of three types of formations, associated, recovered in conjunction with oil, non-associated, recovered as a gas, and as a gaseous stream from coal bed seams. Natural gas is not of the same quality when produced. Each of the formation types exhibits distinct characteristics and the gas recovered from within a particular type of formation may vary. The most abundant compound is methane. Produced gas will also contain varying quantities of non-methane hydrocarbons and other constituents that contribute little or no heating value. The gas is generally treated to reduce concentrations of water, carbon dioxide and hydrogen sulfide. Natural gas that is rich in non-methane hydrocarbons may then be further processed to extract natural gas liquids.

Producers work through shippers to contract with common carrier pipelines to transport gas. Tariffs filed with FERC define the conditions under which and charges are applied for transporting gas from a specified receipt point to a specified delivery point. The shipper provides gas to be transported in units called “dekatherms”, which represent the combination of a volume of gas with a specified heating value. Volumes are determined by meters, and the heating value is determined by calorimetry. In general, volumes are measured continuously using one of several types of meters. Larger volume receipt points generally entail the use of on-line continuous calorimetry (typically daily volumes greater than xx MMSCF), whereas long-term composite and manual grab samples are more typical for smaller volume receipt points.

Pipeline operators found the need to establish tariff conditions for parameters effecting quality including water, carbon dioxide, hydrogen sulfide, among others, to ensure safe and reliable operations. These parameters in sufficient quantities can create a corrosive environment in the pipeline system and eventually downstream end use. The tariff limits are typically expressed as maximum limits. Gas contracted for transportation must be provided within these limits. Pipeline operators may apply a waiver 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 Shipper A is generally of the same quality as gas shipped by Shipper B as long as the gas to be transported meets the conditions of the pipeline's tariffs.

As stated earlier, when the value of natural gas liquids is at a discount relative to their thermal value in the natural gas, producers may opt to bypass gas processing. Pipelines have been designed throughout the years with a variety of means to capture small incidental volumes of liquids so as to protect downstream compression. Pipeline systems anticipated generally liquid free operation and in many instances opted to install minimal or no specific equipment to collect liquids that drop out. The chemistry and thermodynamics of processed natural gas support operation in this manner. This is because processed gas is sufficiently dry as to be able to provide absorptive capacity in the event that small volumes of either condensed water or liquid hydrocarbons are introduced into the pipeline system. However, if temperature and pressure become sufficiently low enough at any point in the system, water and hydrocarbons can condense into liquid from a natural gas mixture. The water dew point, is the temperature at which water vapor will condense to liquid water, and similarly, the hydrocarbon dew point is the temperature at which hydrocarbons will begin to condense and hence the term, liquid drop out.

The simplest means of controlling incidental liquid accumulation was through installation of drips; a vessel attached to the pipeline that takes a slip-stream of the flowing gas and removes liquids through physical impingement. The captured liquids accumulate and are periodically pumped or siphoned off and either sold or disposed. There has been a trend in recent years to remove drips from pipeline systems as they may be subject to corrosion. Some Office of Pipeline Safety Regional personnel have encouraged or even required operators to remove drips from their systems.

Some operators did elect to install filtration or separation equipment, or both, on the suction side of compressor stations but to collect small volumes of water and compressor oils carried over from upstream stations.

When natural gas is processed, the presumption of fungibility is sound and the original design basis of the pipeline infrastructure for managing incidental free liquids is appropriate. However, if producers opt not to process gas, such as times when natural gas liquids are at a discount to their value in the gas, a dilemma appears to exist. The presumption of fungibility may no longer be appropriate. In pipeline systems designed to transport single-phase gas, while gas capable of condensing into liquid form may cause operational or safety problems, it is also important to recognize that a portion of the dekatherms received are potentially lost in transport. Where the liquids accumulate in the pipeline or associated equipment, the pipeline operator may experience increased lost and unaccounted for (LAUF). 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. Ultimately, all shippers on the system must contribute their pro-rata share of the LAUF.

Local distribution companies (LDC) 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 before the transfer of custody from the pipeline to the LDC or customer occurs. To measure the gas properly, a metering station will have knockout vessels to remove any fugitive solids or liquids that may be in the gas prior to passing through the measurement device. Once measured, the pressure is normally reduced to the operating pressure of the LDC pipeline system. As the pressure is reduced, energy is released in the form of cooling. It is possible to reduce the pressure enough to cause the gas to chill below the corresponding dew point, thereby causing liquids to fall out. If the low-pressure gas is not properly scrubbed in a separator or a knockout vessel, hydrocarbon liquids in the form of mist can enter the LDC pipeline distribution system. The mist will coalesce on the walls of the pipeline and will begin to collect in the low spot of the pipeline system. If there are no other knockout vessels on the downstream portion of the system, liquids could be swept along by the gas flow until reaching an exit point on the system -- a customer meter and burner. Liquids reaching a burner can degrade performance, spew out through the burner ports and could either cause a large uncontrolled flame or extinguish the flame altogether and form a puddle in the hot appliance.

Pipelines lacking a direct means of controlling hydrocarbons may not only experience operational and safety problems but also experience significant unaccounted for problems. Hence, the need to ensure that the product being transported is in fact truly fungible, that is, that all gas being transported is of the same quality, and the need for a gas quality standard. NOTE: An alternative way of stating this may be, Hence the need to provide operators with a means of controlling levels of condensable hydrocarbons entering the pipeline system, where the operator deems that control to be necessary to ensure operational reliability and pipeline integrity.

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

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 managing hydrocarbon liquid drop out including pressure reductions, ambient temperature and the impact including hydrate formation.

Section 3 – Hydrocarbon Drop Out Control Measures

This section describes measures used historically to control hydrocarbon liquid drop out, including thermal content (Btu/volume), 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+. This section also provides a description of blending, a tool to provide shippers and pipeline operators some flexibility in managing hydrocarbon liquid drop out.

Section 4 - Overview of Hydrocarbon Dew Point

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 produced gas is processed, and as pressure and temperature change along the value chain.

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 detail analyses of hydrocarbon compounds in gas as produced and processed. The section also provides historical levels of hydrocarbon dew point as well from the same sources.

Section 6 - Measurement and Estimation of Hydrocarbon Dew Point

This section provides an overview of the direct measurement of hydrocarbon dew point. A chilled mirror is used to conduct direct measurement of the hydrocarbon dew point. Alternatively, a combination of sampling, analysis and estimation 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 managing hydrocarbon drop out.

Section 7 – Recommendations

This section provides a set of recommendations developed by the Natural Gas Council Task Force on Gas Quality to manage hydrocarbon liquid drop out.

Appendices

A. Glossary
B. Relationship between HCDP and other Historical Measures
C. Estimation of Condensate Volumes
D. Summary of Analytical Methods

Section 2 - Liquid Hydrocarbons in Natural Gas

Sources

Natural gas produced from geological formations within the earth comes in a wide array of compositions. To simplify the varieties of gas compositions, they can be categorized into three distinct groups:

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

These produced gases can contain both hydrocarbon based gases (these burn) 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) and Octanes plus (C_8+). The non-hydrocarbon gas portion of the production gas can contain Nitrogen (N2), Carbon Dioxide (CO_2), Helium (He), Hydrogen Sulfide (H_2S), a corrosive and poisonous gas, water vapor (H_2O) and other trace gases. CO_2 and H_2S are commonly referred to as "acid gases" since they form corrosive liquids in the presence of water. N_2 and CO_2 are also referred to as inert gases contained in the hydrocarbon gas stream since neither burns, thus they have no heating value.

Associated gas is produced as a by-product of oil production and of 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 (C_2), Propane (C_3), Butanes (C_4), 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 thermal content (BTUs – British Thermal Units) per volume of gas and a high hydrocarbon dew point. When referring to NGLs in the gas stream, the term GPM (gallons per thousand cubic feet) is used as a measure of hydrocarbon richness. Note: The terms "rich gas" and "lean gas" are commonly used in the gas processing industry. They are not precise indicators but are used only to describe one gas stream relative to another.

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 and is sometimes referred to as Lean Gas. Non-Associated Gas can contain all the other non-hydrocarbon gases identified above.

Coal Bed or Coal Seam Gas 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. In order to produce the gas, the coal bed must first be de-watered to allow the trapped gas to flow through the formation. Coal Bed gas usually contains elevated levels of CO_2 and hence this gas is generally of lower thermal value and this gas must be treated to remove the CO_2 to acceptable levels given its potential to be corrosive. Need a description of how gas is introduced into the pipeline from process plants, direct connect production and unprocessed gas.

Role of Gas Processing

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 gather the gas from the production facilities, aggregate the volumes treat and process the gas, if required, before it enters the Interstate Pipeline Transportation system and Downstream Markets.

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 pipeline and compression used to transport the gas.

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 “contaminants” such as CO_2, H_2S and water vapor from the hydrocarbon gases if necessary and secondly on the removal of NGLs from the hydrocarbon stream. Not all gas sources contain unacceptable levels of CO_2 or H_2S, therefore, processors do not have to invest in the type of equipment used to remove these contaminants.

Gas treating can be done in a facility 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. Alternatively, integrated plants can be found on a transmission pipeline near production areas. These plants are referred to as “straddle plants”.

If any acid gases such as H2S and CO2 are present in the production gas, the first step is to treat the gas to remove the corrosive gases to acceptable levels. Pipeline tariff specifications establish the acceptable level of contaminants in the pipeline and therefore the processor knows the degree of removal required to make an acceptable natural gas product. Generally speaking, pipelines have historically required merchantable natural gas to contain less than 0.25 grains per hundred cubic feet (4 parts per million) H2S, usually a range of less than 2%-3% CO2 and usually 7 or less pounds H2O per million cubic feet. Processing plants often reduce the quantity of CO_2 and H_2S below these standards in order to meet NGL product specifications. Water vapor is often reduced to less than 1 part per million as part of the low temperature extraction process (discussed later). These gases are removed because they are corrosive to the pipelines delivering the gas to the plant and to the processing equipment inside the plant.

Once the gas is cleaned of corrosive gases, it is processed to remove NGLs or it may be suitable for sale without further processing, as is the case of lean non-associated gas and coal bed gas. 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, must be removed from the gas stream prior to gas processing or further pipeline transportation.

If the gas is rich in NGLs or contains levels of Nitrogen in excess of 2% to 5% or contains commercial quantities of Helium, the next step in gas processing is to remove these gases. To remove the nitrogen or helium, special plant equipment is required to reduce the gas temperature to around –260F. This is a very costly process, both in operating expense and capital investment. Recovery of Helium and rejection of Nitrogen are not commonly used processes and will not be included in this discussion.

To remove NGLs, there are three common processes; Refrigeration, Lean Oil Absorption and Cryogenic. A processor will select the process to build after evaluating the market values of the natural gas and NGLs, the costs to get the NGLs to market, capital costs, fuel, and other operating costs. Refrigeration has the least capital cost but also recovers the least amount of NGLs. This process is able to extract a large percentage of propane and most of the C_4+ 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 that 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.

Lean Oil Absorption plants are another type of processing plant and were the type built in the 1960s. These plants were the next evolution from the refrigeration plants and they can extract 90%+ of the C3s+ in the gas stream and about 30% of the ethane by chilling the gas and an absorption oil to approximately –30F. The fuel consumption of this type of plant is higher than that of the refrigeration plant. The ethane and propane was recovered to begin feeding 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 for the rapidly growing gas supplies found in the Gulf of Mexico and the eastern half of Texas during the 60's and 70's. These pipelines transported the natural gas from the U.S. Gulf Coast to destination in the northern and eastern parts of the U.S. This type of plant was very economical due to the governmental influences affecting the gas industry as discussed in the "Governmental Influence" section following below.

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 were able to extract up to 70% of the ethane from the gas, leaving a sales gas that was 90+% methane and the rest ethane and inert gases. To reach these higher extraction levels more expensive metallurgy, more compression, and more capital investment is required. Also due to the pressure reduction involved in the process, there is a higher operating expense due to the added fuel demand to run the compressors. Since the early 1990s, modifications to the cryogenic process have allowed ethane recoveries to reach close to a 99% extraction level. This increase in recovery comes with a higher fuel consumption and higher operating expense.

Operators of gas processing plants have tried to operate their plants in reduced recovery modes. However, the plants were designed to achieve high recoveries of all the NGLs and the "turndown" to lower recoveries has been difficult to obtain. Typically, gas plants are not designed to recover only the C5+, or only the butanes. The plants 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.

Economics of Processing

Processing costs are typically viewed as part of the cost of production to the producer and revenue to the processing plant owner providing the service. Once a hydrocarbon bearing formation is discovered, a producer must first evaluate the geological structure to determine whether it is large enough to contain a sufficient volume of hydrocarbons (Reserves) to support the investments required to produce the reserves. This evaluation also requires the analysis of the hydrocarbons found and the contaminants within the hydrocarbons. The revenue from the sale of the oil and gas must be expected to exceed the costs of production, including both operating and investment capital expenses. The gas portion of the evaluation takes into consideration the cost of gathering the gas, compressing the gas to sales gas pipeline pressures, the processing and treating of the gas to meet any required sales gas pipeline quality specifications, and the cost of production operations. The removal of CO_2, H_2S, N2 and H_2O involve costs with no by-product revenue. NGLs and Helium are valuable by-products and the extraction of these constituents usually generates revenue that may or may not cover the investment and operating costs of providing the treating and processing services.

The basics of NGL processing economics are to evaluate the amount of NGLs available to extract (which is determined by 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 lost 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 loss and as a thermal loss. Loss 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 that the gas is processed, there is a gas value and a NGL value to the shrinkage part of the gas. The relationship between the gas value of the NGL plus fuel shrinkage and the equivalent gas value of the NGLs as a liquid value is called the margin. If the liquid value after 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 economic sense to extract the NGLs from the gas.

In the early years of the gas industry, producers sold their gas production to a gas pipeline company, normally through long term, fixed low price contracts.

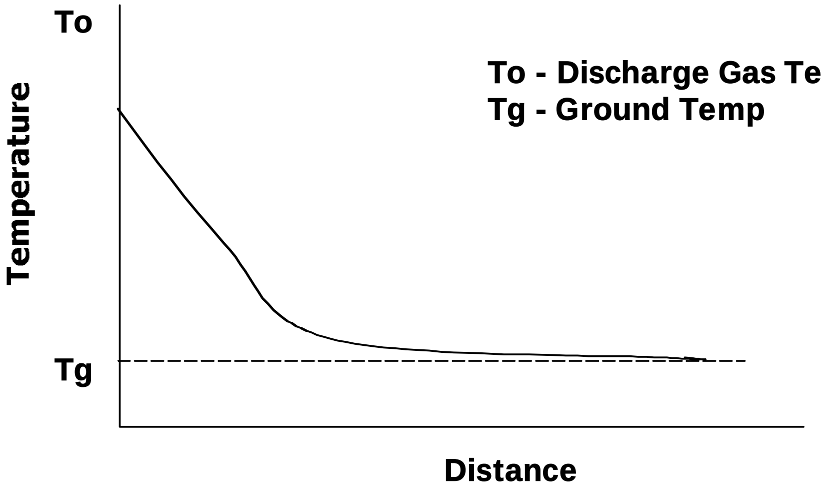
Gas was generally considered a byproduct of oil exploration and production and a producer would take whatever value they could get for the gas instead of venting or flaring it. The production gas was processed in “Gasoline Plants” which simply compressed the gas, cooled it with either air or water to 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 each compressor station 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, the gasoline plants began to chill the gas by the ways of simple pressure reduction/expansion, by passing the gas through a light oil (absorption) or by use of a refrigerant like 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.

Influence of Ambient Temperatures and Pressure Reductions

Ambient temperature and pressure reductions are known to have an impact on flowing gas temperatures that in turn can result hydrocarbon drop out. Ambient temperatures become a concern when they are sufficiently cold as to be below hydrocarbon dew point of a gas stream. It is when the flowing gas temperature comes to equilibrium with the relatively cool ambient temperature that hydrocarbon drop out occurs. Significant pressure reductions during transportation have an impact on the flowing gas temperature. Pressure regulation that for example reduces the gas pressure from a high-pressure source to a lower pressure, creates a rapid cooling of the gas stream; referred to as the Joule-Thompson effect. This section describes the circumstances under which ambient temperature and pressure reductions occur on pipeline systems, examines their impact and describe options that an operator has to manage hydrocarbon drop out in these situations.

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 degrees Fahrenheit. Typically a pipeline's tariff limits the maximum discharge gas temperature to 120 degrees Fahrenheit. Once the gas leaves the compressor and travels underground to pipe depth, the temperature of the gas falls rapidly due to the difference between the ambient ground temperature and the flowing gas temperature. Figure 1, below, shows the typical rate of change of the flowing gas temperature as it approaches ground temperature.

The U.S. Department of Transportation requires that all natural gas pipelines be installed at an appropriate depth below the ground's surface; typically in the range of 30 to 36" depending on the diameter of the pipeline. 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 approach 32 degrees Fahrenheit. Pipelines located in the northern part of the country maybe installed at depths below the frost line, the depth to which frost penetrates the ground. Where the pipeline is below the frost line, the flowing gas temperature should not fall below 32 degrees Fahrenheit.



Ambient air temperature is another factor that affects the flowing gas temperature. Once 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 temperature. The concern is where there will be sufficient heat loss to cause the flowing gas temperature to go below the hydrocarbon dew point. The gain or loss in flowing gas temperature is calculated from heat transfer methods knowing various details such as pipe material, outside temperature, wind velocity, length of exposed pipe, throughput, among other factors. Although most pipeline systems do not have extensive above ground piping, the configuration of the above ground piping can magnify the effect of the ambient temperature on the flowing gas temperature, particularly in the application of air cooling of compressed gas in the winter time.

Water crossings can be subject to fluctuations in the flowing gas temperature. Pipelines built today are typically bored beneath rivers at depths between 25 to 75 feet, certainly below the mud line. At this depth the temperature of the river has no effect on the flowing gas temperature. On the other hand, older pipelines lay on or are 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, just like the underground flowing gas temperature, should not fall below 32 degrees Fahrenheit.

The flowing gas temperature can be determined from a heat transfer calculation. By knowing the ambient temperature, outside diameter of the pipe, overall heat transfer coefficient, specific heat of the gas, flowing gas volume, specific gravity of the gas, and the length of the pipe, the Mean Flowing gas Temperature (MFT) can be computed. Simple charts have been developed to depict the ground temperature at certain depths as a function of the ambient temperature. However, these presume that the flowing gas temperature has reached temperature equilibrium with the soil.

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 degrees Fahrenheit. Thus if the pipeline is delivering gas at a pressure of 800 psia to an end user that requires a pressure of 200 psia, the gas temperature will drop approximately 42 degrees ((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 psia to 200 psia	*less*	*42 °F*
Resultant gas temp (without heating)		*-6 °F*

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 final flowing gas temperature after regulation. Heaters in this application refer to gas fired heat exchangers that heat the gas before entering the regulator, thereby reducing the potential of hydrocarbon liquids from forming. The number of degrees the flowing gas temperature can increase is dependent 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 degrees, the resultant flowing gas temperature would be 14 degrees Fahrenheit.

NOTE: This section provides an interesting historical background but as written does not add value to the effort, except possibly to show that the government found it necessary to intervene, and that is what we are trying to avoid through this cooperative effort. The market conditions of today with respect to natural gas pricing vs. NGL pricing are not a function of past governmental actions.

Governmental Influence

By the early 1930s, the FTC found that over 25% of the interstate pipeline network was owned by 11 utility companies and that these companies could exert market control over gas production, transportation, distribution and electric generation. From this 1935 report Congress passed the Public Utility Holding Company Act, to regulate holding companies from gaining undue influence over the public utility market. This government oversight did not extend to interstate gas sales and hence in 1938, Congress passed the Natural Gas Act, establishing oversight on the prices interstate pipelines could charge their customers. In 1954, the Supreme Court decision in Phillips Petroleum Co. v. Wisconsin initiated wellhead price control of natural gas. Because the gas values were determined through a “cost of service” basis instead of a “market value” basis, prices were fixed and were low relative to crude oil production values. Since C3 and NGLs are generally priced relative to crude oil, this created an environment where the prevailing values for both gas and NGLs rendered the margin equation almost always positive. It was not until after the country experienced massive gas supply shortages in the mid 1970s that Congress enacted the NGPA of 1978. This act had three main goals:

- Creating a single national natural gas market
- Equalizing supply with demand
- Allowing market forces to establish the wellhead price of natural gas

This act established price ceilings and price tiers, established incentives to producers to search for and produce gas from more new natural gas formations. These new pricing tiers for gas produced from different geological formations created processing issues because the margin calculation for some of the new packages of gas were no longer positive. Once producers began to receive higher prices for these new packages of gas, supplies increased. Many times these high priced packages were not processed because the gas shrinkage had a higher value than the NGLs. In 1985, the issuance of FERC Order 436 (Open Access Order) began the unbundling of pipeline marketing companies from pipeline transportation companies, allowing market forces to determine price. This opening of the pipelines was followed in 1989 by the passage of the Natural Gas Wellhead Decontrol Act whereby all gas production was stripped of pricing tiers and spot prices were allowed to float with the market demand. Gas prices entered a period of stability along with abundant supply. This abundant supply of gas led to positive margins for gas processing which motivated the voluntary processing of gas and removal of NGLs.

The liquid fallout problems the pipelines have experienced since 2000 are primarily because negative gas processing margins have caused many gas plants to be shut down. The situation is more pronounced in the Gulf Coast region because many pipelines operate as regulated pipelines and many of the gas plants are straddle plants. In other words, the pipelines have received the gas with a hydrocarbon dew point higher than would normally be acceptable for interstate pipeline transportation. Because the pipelines cannot easily refuse gas that otherwise meets their tariff requirements, the pipelines have issued Operational Flow Orders that impose more stringent gas quality requirements than are contained in the pipelines tariffs. In other parts of the U.S. the pipelines typically receive gas from a plant and gathering system and can expect that the plant has depressed the hydrocarbon dew point to an acceptable level. Gas produced from the Gulf of Mexico is usually in the interstate pipeline grid before it has a chance to be processed.

Since the year 2000, natural gas prices have increased and become more volatile, causing the processing margin to swing negative many times. When a high gas price, low NGL price environment occurs, producers and processors tend to reduce the level of processing in order to reduce the shrinkage due to processing, thus supplying more gas into the marketplace.

Section 3 - Hydrocarbon Drop Out Control Measures

As described in Section 2.0, pipeline operators established specifications for CO_2 and H_2S (acid gases) and water because these compounds were very detrimental to the pipeline integrity. Many pipelines addressed the hydrocarbon content of the gas in a variety of ways, but at no time has there ever been a common interchangeable set of specifications for components such as CO_2, H_2S and water. Much of this is due to the way the gas industry developed. Many, if not all pipelines have minimum specifications for thermal (Btu) content. This resulted from gas historically being produced from fields high in N_2 or CO_2. N_2 and CO_2 are inert and do not have a thermal value, thus they dilute the natural gas and when in high enough concentrations can cause an end user's appliance to experience flame instability. On the other hand, approximately a third of all interstate pipelines specify a maximum thermal content and the majority of these are either 1100 Btus per cubic foot or 1200 Btus per cubic foot. There is not a differentiation as to whether this is a condition at the receipt point or delivery point. In most instances, where gas is processed, the thermal content downstream of the gas plant will be less than 1100 Btus per cubic foot. The use of a maximum thermal content is an inadequate predictor of hydrocarbon drop out because a gas can have a relatively low thermal content and a high hexanes plus content that can result in hydrocarbon liquid drop out.. Furthermore, specifying a maximum thermal content may be detrimental to supply considerations as will be discussed later.

Some pipelines selected another parameter to control liquids fallout by establishing a pentanes plus (C_5+)or hexanes plus (C_6+) specification. A C_6+ specification may in some instances be used as an indicator of potential for hydrocarbon liquid drop out but as will be discussed in Section 6, there are problems in applying this measure broadly. Since the C_6+ composition varies among gas streams and has a large effect on hydrocarbon dew point, analysis and computation are complicated. Additionally, the use of a C_6+ specification alone does not provide the information necessary for end-users to design, install and operate equipment and pressure reduction stations that will keep the flowing gas from entering into a two-phase region and causing liquids drop out.

In more recent times, some pipeline operators have elected to establish hydrocarbon dew point limits. Establishing a hydrocarbon dew point specification, with one or more industry standard test methods, appears to be the most effective means of controlling {is the best control measure for} hydrocarbon drop out. Eight interstate pipeline operators have established hydrocarbon dew point limits. An HCDP limit enables a wide range of gas compositions to be made available to end-users without compromising the safety or the integrity of the natural gas supply infrastructure. The use of a hydrocarbon dew point specification will provide the information necessary for end-users to properly design, install and operate pressure reduction stations.

Blending

Blending (Mixing) is the combination of materials so that separate constituents (parts) cannot be distinguished. Blending is the mixing or commingling of materials to achieve a volume weighted average of the concentrations of each constituent. As it relates to natural gas, pipelines have used blending for years to change the combined quality of its gas stream from the individual gas streams in order to meet gas quality related requirements. Blending has specifically been used for controlling BTU content and to meet other gas quality requirements. As the industry moves to implement new quality specifications relating to hydrocarbons, blending may be one tool that can be used to manage hydrocarbon levels and at the same time provide a potential to accommodate packages 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. However, the ability to utilize gas blending to manage HDP is dependent on a number of factors including pipeline configuration, receipt and delivery points, gas supply quality, geographic location, etc.

Each pipeline operator based upon their unique (or specific) operations will define the exact methodology and process for blending and monitoring the resulting hydrocarbon levels. The following example using blending with HDP is provided to help demonstrate how blending works. When a gas stream with a low HDP is mixed, or commingled with a gas stream with higher HDP, ”rich gas”, the resulting commingled or blended stream will have a HDP somewhere between the two individual HDP levels. In the example shown in Figure 1 below, Gulf Coast unprocessed gas with a HDP temperature of 55°F has been blended with a processed stream with a HDP temperature of 10°F. The resulting blended gas stream has a HDP temperature of approximately 22°F. **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. When two different gas streams are mixed/blended, each compositional component of the commingled stream changes creating a completely separate and unique gas quality.

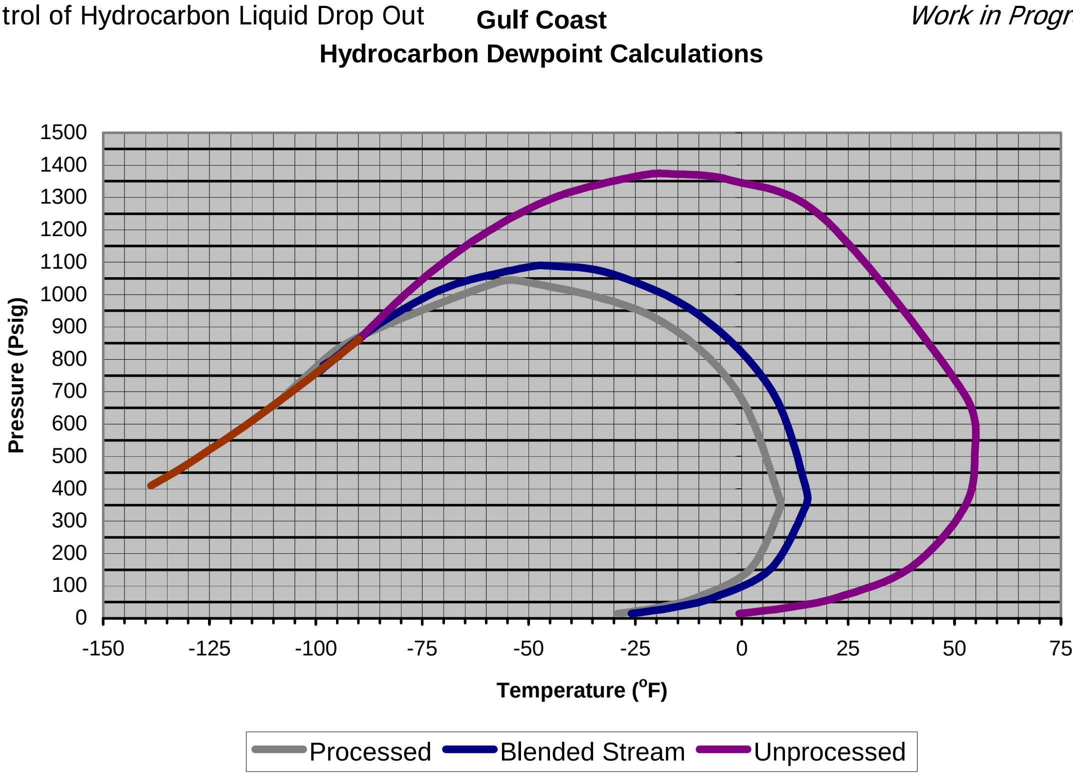


Figure 1.

There are actually two types of blending which are distinctively different when received by the pipeline but may have the same effect on the gas quality. These two types of blending are physical and contractual.

Physical blending is when two gas streams are mixed prior to pipeline injection. In this case, the two gas streams are mixed and actually change in physical composition as illustrated in the example above. The two blended gas streams may not however, thoroughly mix when combined. It may take some distance and possibility after compression or some other mixing event, before they truly become blended. Therefore determination of the combined HDP may be distorted if the HDP was derived by a gas sample taken prior to complete mixing. However, when thorough mixing occurs, the resulting gas stream will have a HDP between the HDP's of the two mixed streams.

Contractual blending is when a producer of high HDP gas contracts with a lower HDP producer to reduce its HDP by blending where the resulting HDP is lowered to meets a specific HDP limit. These two volumes may be on totally 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. Contractual blending is only effective when there is physical blending on the pipeline.

There are many benefits for a pipeline to allow gas to be blended, those benefits, however do not outweigh the enforcement of a HDP limit that reduce the possible formation of pipeline condensate. The major benefit of blending is that it allows producers or production points with high HDP gas that do not have the ability to process the gas to contractually reduce the HDP level through contractual blending without the requirement of adding expensive equipment. Any application or management of HDP specifications must also be implemented in non-discriminatory manner. Therefore, to meet the requirement of non-discrimination and at the same time not to hinder a pipeline's ability to attract gas to the system, a rational HDP management plan must be developed by each pipeline that address the physical limitations and the financial impacts of HDP management. It must also be noted pipelines that have very little control on a day-to-day basis on where the gas flows since customers own the capacity and do not share responsibility for managing the gas quality requirements as provided in the tariff.

Heaters

Most problematic hydrocarbon condensation occurs at points of pressure regulation (or immediately downstream). Water bath heaters, which heat water that surrounds a tube bundle (heat exchanger) containing the flowing natural gas, can be used prior to pressure regulation. Since they burn natural gas, they require air permitting.

Offshore Gas and Liquids Handling

Handling of gas and liquids in the offshore environment is different than onshore gathering because of the way gas and condensate are handled. In the offshore environment most gas pipelines allow for the produced fluids to be separated at the off shore 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 off shore. In addition, 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 on shore separation and processing facilities. These liquids must be removed before the gas can be processed for NGL recovery or further transported to market. Since the condensate and gas are intimately mixed when they reach the shore, the gas that leaves the separator is at its hydrocarbon dew point. Thus, any further cooling of the gas or changes in its pressure may cause liquids to form.

When natural gas and condensate are present together in a pipeline, or pressure vessel, they reach a point of “equilibrium” or a saturation point. The gas stream is at its hydrocarbon dew point at that temperature and pressure. So any time this offshore sourced gas is delivered to a pipeline without processing it is most likely at its dew point and any cooling of the gas from the ground temperature or a pressure reduction (like a pressure regulator) can condense liquids. This common practice in the offshore industry further complicates the application of blending on each pipeline.

The following are ideas relating to a HDP management program. These may not work for all pipelines but are factors that may be considered in the development of such a plan.

System Monitoring of HDP

An integral part of any HDP management program is the ability to monitor the pipeline system at various strategically located points along the system, or “node points” or at individual metered points as necessary. Software is available that works with a pipeline’s measurement system to predict the HDP temperature. This software can calculate the HDP temperature at various locations across the system and at intervals specified by each pipeline. To implement this plan a pipeline must install a quality device at the node or meter point that is capable of determining the gas quality. In selecting a node point several factors should be considered. One factor would be whether the node point is at a location where the stream is completely blended to ensure that the HDP temperature calculations are reflective of the actual blended stream. Another factor is to select a point where the gas quality is readily available from an online chromatograph, such as a measurement point. This requires an online chromatograph and the capability of communicating to a SCADA system. When selecting a node point it is best to identify a series of points so that indications of increasing HDP are noticed before the gas is introduced into the market area. In other words, high HDP must be determined well before the gas gets to the LDC or pressure reduction station.

Section 4 - Overview of Hydrocarbon Dew Point

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. 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 below its dew point temperature then the entire dew point curve shifts cooler for the gas phase that is now depleted in heavier components.

Hydrocarbon Dew Point Curve

The HDP for natural gas with a given composition is typically displayed on a phase diagram, an example of which is shown in Figure 1. The HDP curve is plotted as a function of gas pressure (P) and temperature (T). The left-hand side of the curve is the bubble point line and divides the single-phase liquid region (red-shaded area) from the two-phase gas-liquid region (yellow-shaded area). The right-hand side of the curve is the dew point line and divides the two-phase gas-liquid region and the single-phase gas region (blue-shaded area). 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 and two dew point pressures are possible at a given temperature. This phase envelope phenomenon provides for behavior known as retrograde condensation. The retrograde phenomenon is that liquids form at a given temperature when the pressure is lowered. The word retrograde means generally tracking backward and this phenomenon appears contradictory. The maximum pressure at which liquids can form (P_{max}) is called the **cricondenbar**, and the maximum temperature at which liquids can form (T_{max}) is called the **cricondentherm**.

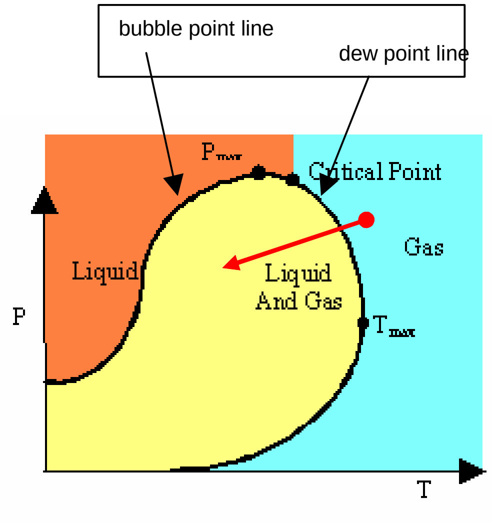


Figure 1. Hydrocarbon Dew Point Curve for a Typical Natural Gas Mixture

QuickTime™ and a
TIFF (LZW) decompressor
are needed to see this picture.

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_5+. The presence 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 C_5+. Processing of the gas stream removes or extracts hydrocarbons and thus reduces the HDP of a given mixture. The level of hydrocarbon removal directly impacts the HDP. Figure 2 shows examples of the HDP curve for unprocessed and processed gas mixtures. The unprocessed HDP curve is in red and has a higher temperature cricondentherm while the processed HDP curve is in blue. The difference between the two curves shows the impact of processing on the HDP.

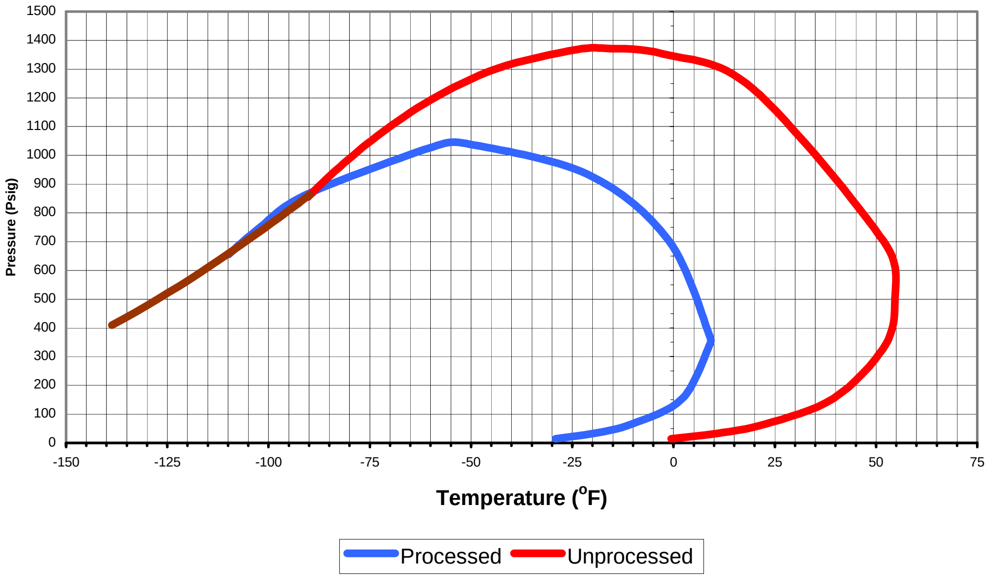


Figure 2 Unprocessed and Processed HDP Curves

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 red arrow in Figure 1 shows a change 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 equivalent liquids per Mcf of gas) for the liquefiable components in each stream.

Pressure Reduction Impact

The reduction in the pressure of a natural gas mixture will cause a Joule-Thompson cooling effect on the mixture. As pressure is decreased the temperature of the gas mixture decreases. The specific temperature reduction coefficient is a function of the gas mixture composition and can be calculated.

The general guideline is that the temperature will drop 7° F for every 100 psi pressure reduction.

HDP increases as the pressure decreases until the cricondentherm point is reached. At pressures below the cricondentherm, the HDP decreases as the pressure declines. For further example, regulator pressure decreases at inlet pressures of approximately 200 psia and less will result in a greater decrease in HDP than a Joule-Thompson expansion temperature decrease, thus resulting in a safer margin.

Section 5 - Historical Levels of Hydrocarbons and Hydrocarbon Dew Point

Industry sources provided a total of 2775 contemporary gas samples from the gas sources shown in the following table to determine the average hydrocarbon composition of the gas type and the cricondentherm of these average compositions. The gas types analyzed were Unprocessed, Processed Plant Outlet, Market, and LNG. Cricondentherm temperatures vary from the low to mid 70 °F for unprocessed gas to below 0 °F for processed gas and market area gas in Massachusetts and New York. C5+/C6+/C7+ mole % are also shown as an indicator of the heavier hydrocarbon liquid content that could potentially condense from the gas.

Average Characteristics for Gas Types

	Extended Unprocessed	C6+ Unprocessed	Processed Plant Outlet	Market MA	Market NY	LNG
Propane mole%	1.61	1.33	1.44	0.24	0.21	0.38
I-butane mole %	0.41	0.33	0.18	0.04	0.03	0.04
N-butane mole%	0.44	0.38	0.18	0.04	0.04	0.03
Neoc5 mole%						
I-pentane mole%	0.14	0.16	0.03	0.01	0.01	0.01
N-pentane mole%	0.14	0.11	0.02	0.01	0.01	0.00
Nitrogen mole%	0.78	0.38	3.29	1.32	1.47	0.01
Methane mole%	90.21	92.93	86.64	95.49	95.27	96.47
Co2 mole%	1.25	0.78		0.74	0.74	0.00

Ethane mole%	4.80	3.28	8.17	2.06	2.20	3.06
Hexane mole%	0.11	.14	0.03	0.03	0.16	0.00
Heptane mole%	0.08	0.10				
Octane mole%	0.01	0.05				
Nonane mole%	0.01	0.03				
Decane mole%	0.01					
C5+ mole%	0.50	0.57	0.06	0.05	0.02	0.01
C6+ mole%	0.23	0.30	0.01	0.03	0.01	0.00
Cricondentherm °F	76.96	72.76	-39.50	-11.37	-42.29	-53.46

Domestic Unprocessed Gas

Two hundred and thirty-two samples of domestic unprocessed gas were analyzed to determine average composition and to determine the distribution of cricondentherm hydrocarbon dewpoint in the samples. 224 samples used a fixed C6/C7/C8/C9 ratio to extend a C6+ analysis in determining cricondentherm hydrocarbon dewpoint (CHDP). The CHDP of the remaining 8 samples was based on full-extended analysis (through C10+). Statistics for the sample cricondentherm, C5+, C6+, and C7+ attributes are shown in these tables.

Standard Analysis – Fixed C6-C9 Ratio

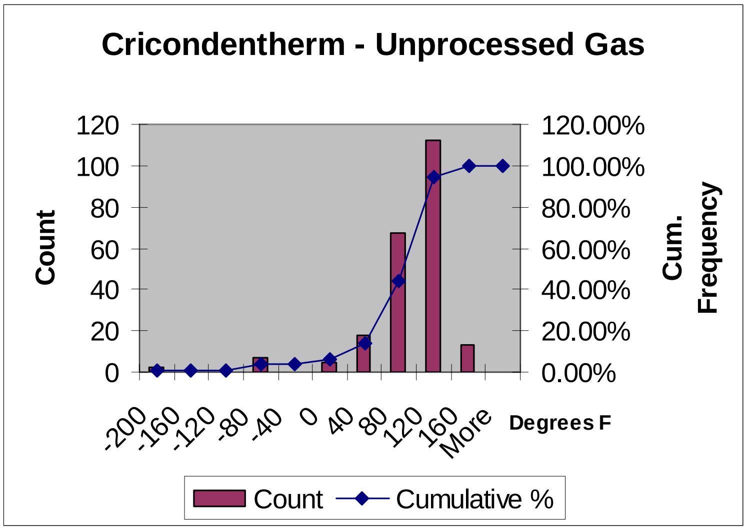
Cricondentherm		*C5+ Mole%*		*C6+ Mole%*		*C7+ Mole%*	
Mean	72.76	Mean	0.57	Mean	0.30	Mean	0.17
Median	83.85	Median	0.52	Median	0.28	Median	0.16
Standard Dev	52.44	Standard Dev	0.41	Standard Dev	0.21	Standard Dev	0.12
Range	373.20	Range	1.89	Range	1.20	Range	0.78
Minimum	-215.90	Minimum	0.00	Minimum	0.00	Minimum	0.00
Maximum	157.30	Maximum	1.89	Maximum	1.20	Maximum	0.78

Count	224.00	Count	224.00	Count	224.00	Count	224.00

Extended Analysis

Cricondentherm		*C5+ Mole%*		*C6+ Mole%*		*C7+ Mole%*	
Mean	76.96	Mean	0.50	Mean	0.23	Mean	0.12
Median	82.70	Median	0.53	Median	0.25	Median	0.10
Standard Dev	27.42	Standard Dev	0.39	Standard Dev	0.15	Standard Dev	0.09
Range	76.10	Range	0.98	Range	0.39	Range	0.24
Minimum	40.90	Minimum	0.07	Minimum	0.05	Minimum	0.03
Maximum	117.00	Maximum	1.04	Maximum	0.44	Maximum	0.26
Count	8.00	Count	8.00	Count	8.00	Count	8.00

The distribution of the cricondentherms for the Standard Analysis data is shown in this chart containing count (frequency) and cumulative frequency information. 75% of the unprocessed gas data points have cricondentherms between 41 and 120 ° F. The range 81 to 120 ° F contains the largest number of samples (112 samples).



Processed Gas, LNGs, and Market Area Gas

Processed gas samples at gas plant outlets contain a small number of samples (9). This data shows that the median cricondentherm temperature is 4.2 °F (half of the samples are above this number and half below this number). Since the mean cricondentherm is -39.5 °F, this means that a few of the samples at the lower end of the range (-116.6 °F) cause the mean to fall well below 0 °F.

For this reason, the median is a more realistic measure of the cricondentherm of the processed plant outlet data set.

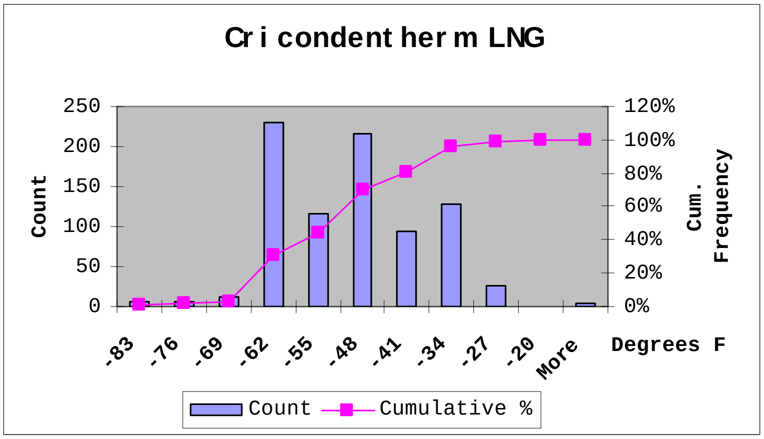
Processed Plant Outlet

Cricondentherm		*C5+ Mole%*		*C6+ Mole%*	
Mean	-39.50	Mean	0.06	Mean	0.01
Median	4.20	Median	0.06	Median	0.00
Standard Dev	67.39	Standard Dev	0.06	Standard Dev	0.02
Range	142.50	Range	0.15	Range	0.04
Minimum	-116.60	Minimum	0.00	Minimum	0.00
Maximum	25.90	Maximum	0.15	Maximum	0.04
Count	9.00	Count	9.00	Count	9.00

The LNG data set consists of 845 samples. The LNG table and chart below show that 95% of the data points have cricondentherms that fall between -69 °F and -34 °F. The maximum cricondentherm temperature in this sample is -16.9 °F. It appears that LNG will not produce liquid hydrocarbon dropout under typical operating conditions in the gas industry because of LNG's low CHDP.

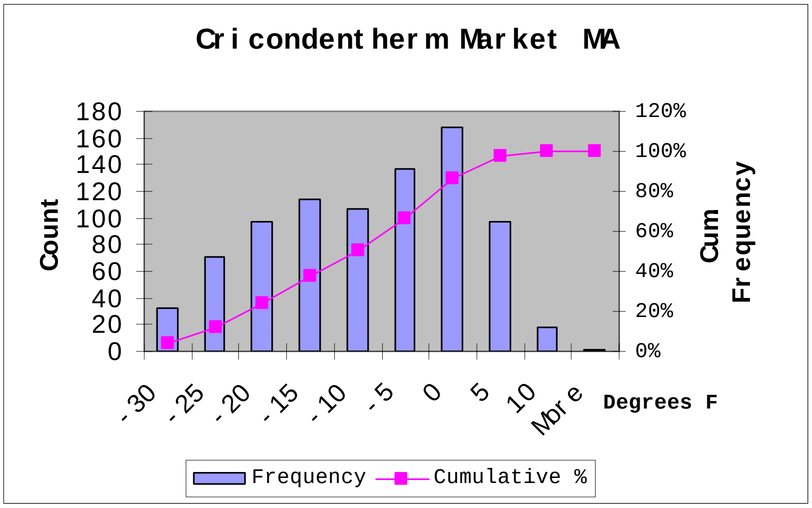
LNG

Cricondentherm		*C5+ Mole%*		*C6+ Mole%*	
Mean	-53.46	Mean	0.01	Mean	0.00
Median	-53.60	Median	0.01	Median	0.00
Standard Dev	12.12	Standard Dev	0.01	Standard Dev	0.00
Range	81.00	Range	0.05	Range	0.02
Minimum	-97.90	Minimum	0.00	Minimum	0.00
Maximum	-16.90	Maximum	0.05	Maximum	0.02
Count	845.00	Count	845.00	Count	845.00



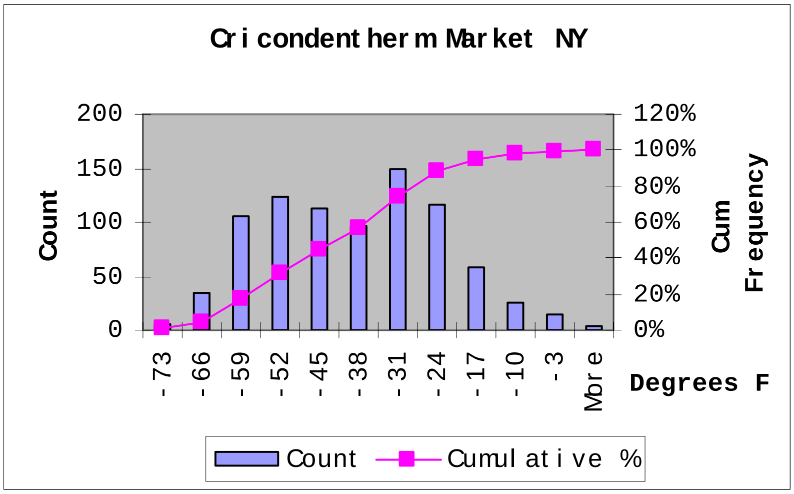
The market area data from Massachusetts (MA) and New York (NY) consists of approximately 1700 samples. The analysis of this data is shown in the tables and charts below. Mean/median cricondentherms are below 0 °F for both areas: -10 °F for New York and -40 °F for Massachusetts. The market gas in the New York area has approximately 30 °F lower mean cricondentherms than the Massachusetts market gas and approximately ½ the C5+/C6+ content. Maximum cricondentherms for both New York and Massachusetts market gas is below 12 °F.

Market MA					
Cricondentherm		*C5+ Mole%*		*C6+ Mole%*	
Mean	-11.37	Mean	0.05	Mean	0.03
Median	-10.00	Median	0.05	Median	0.03
Standard Dev	10.37	Standard Dev	0.02	Standard Dev	0.01
Range	49.70	Range	0.11	Range	0.05
Minimum	-38.00	Minimum	0.01	Minimum	0.01
Maximum	11.70	Maximum	0.12	Maximum	0.06
Count	843.00	Count	843.00	Count	843.00



Market NY

Cricondentherm		*C5+ Mole%*		*C6+ Mole%*	
Mean	-42.29	Mean	0.02	Mean	0.01
Median	-42.30	Median	0.02	Median	0.01
Standard Dev	15.57	Standard Dev	0.02	Standard Dev	0.01
Range	82.30	Range	0.13	Range	0.04
Minimum	-77.00	Minimum	0.00	Minimum	0.00
Maximum	5.30	Maximum	0.13	Maximum	0.04
Count	846.00	Count	846.00	Count	846.00



Historical Use of Hydrocarbon Dewpoint

No data has been developed that tracks historical hydrocarbon dewpoints. However, approximate ranges for cricondentherm hydrocarbon dewpoints can be estimated by referencing the types of processing in the gas industry from the 1940s to the present.

Prior to the advent of gas processing, hydrocarbon dewpoints in pipelines and market areas would approach those for unprocessed gas (cricondentherm ≅ 80 °F or higher). 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 about 10 °F.

Propane as a refrigerant most likely became available post 1940 when demand for butane for motor gasoline increased and rural heating was converted from butane to propane in the 1940s-1950s. Refrigeration dropped the cricondentherm from 80 °F to a little above 0 °F and recovered 50% or so of the propane and 80% or more of the C4+. The gas processed was very rich, on the order of 1200 -1400 Btu, 4 - 7 GPM (gallons per thousand ft^3) casing head gas since this gas was a by-product of oil production.

As ethane became a demand item 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.

When oil prices took off after the 1972 oil embargo, there was great incentive to recover all the NGLs. Cryogenic plant technology became economic. Cryogenic plants can recover 99% of the ethane and essentially all of the C3+ producing cricondentherms 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 (rationalization) of these plants. The remaining production formally processed in the rationalized plants was aggregated with other fields and processed in regional processing centers, i.e. the Phillips-GPM

Oklahoma Super System. The 1990s is also the beginning of the Non-producer plant owner (e.g. GPM, Dynegy, Enterprise, Tejas, Associated (Duke Energy).

On the Outer Continental Shelf (OCS), the Minerals Management Service (MMS) promulgated regulations in the late 1950s that eliminated routine flaring. This created a huge pipeline building boom to recover the formerly flared gas along with the big 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.

In summary, gas in the US pipeline system would have the following approximate cricondentherm temperatures based on the processing technology being used in the indicated time periods if all gas were processed:

Time Period	**Processing Technology**	**Cricondentherm Temperature °F**
1940-60	Refrigeration	≅ 0
1960-75	Lean oil	≅ -30
1975 on	Cryogenic	≅ -100

Actual CHDP 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

This section provides an overview of the determination of hydrocarbon dew point. It can be done in one of two ways, measurement or estimation. A method referred to as the “chilled mirror” is used to conduct direct **measurement** of the hydrocarbon dew point. Alternatively, **estimation** 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 value of each in managing hydrocarbon drop out.

Measurement

The most commonly used method of hydrocarbon dew point detection is with the use of 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)[3]. For many years this device has been used for moisture detection and can be used for hydrocarbon dew point determination also. The device can be used at the pipeline operating pressure in the field location.

The Bureau of Mines dew point tester consists of a polished stainless steel mirror in a pressure-containing chamber having a glass view port. The chilled mirror is cooled by a refrigerant system. It consists of a small high-pressure chamber (5000 PSI max) through which the gas sample flows. A polished mirror is at one end of the chamber and a viewing window at the other. The operator throttles an expandable gas through a valve--cooling the polished mirror until the dew point is observed. The temperature and pressure are then recorded and plotted on a graph.

Determination of the hydrocarbon dew point 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 process is a very time intensive and tedious process. A typical hydrocarbon dew point test may last forty-five minutes to one hour and requires uninterrupted attention to the test apparatus.

Experience indicates that trained and experienced operators can generally reproduce each other’s results within two to five degrees[4]. Among inexperienced operators the results may vary by twenty to fifty degrees. 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.

Automatic, continuous online dew point detection units are commercially available. These units use electronics to control mirror cooling, solid-state optics to detect the formation of dew, a servo control loop to maintain the temperature of the mirror at the dew point, and a PRT to measure the dew point temperature. These units are expensive relative to the cost of other online analyzers. 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.

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[3] 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.

[4] Warner, H. R., Leamer, E. E., Spence, A. P., Bone, R. L., Hubbard, R. A., Bernos, J., and Kriel, W. A., “Hydrocarbon Dewpoint Determination of Lean Natural Gases,” Proceedings, 80th Annual Convention, Gas Processors Association, 2001.

Estimation

Estimation relies on a three-step process, sampling, analysis and calculation. The most common means of sampling and analysis is done with 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 chromatograph. The most common chromatograph uses a combination of columns to analyze for methane through pentane and then treats all compounds with molecular weights greater than hexane as hexane[5]. This chromatograph is referred to as a C_6+ chromatograph.

Manual sampling with off-site analyses can be used as an alternative. Samples of the gas are collected in a canister or on charcoal tubes using standard methods published by API[6]. Samples are analyzed using a chromatograph typically using the method described above.

The third step, calculation, is conducted by applying thermodynamic principals and accepted equations of state using the analysis from above. Two commonly accepted sets of state equations are Peng-Robinson[7] and Souave-Redlich-Kwong[8].

The degree to which the three-step process reflects the actual hydrocarbon dew point is dependent upon the characteristics of the natural gas stream. The parameters of the highest sensitivity are the mole percentage compositions of the hydrocarbons larger than hexane. In particular, having a small percentage of nonane (C_9) or decane (C_{10}) may significantly elevate the calculated hydrocarbon dew point value for the natural gas mixture. The impact of higher molecular weight hydrocarbons is shown in Figure 1. Simply using the C6+ analysis, will yield an estimated value of 15.7 ^oF, whereas inclusion of compounds through decane yield a value of 57.6 ^oF.

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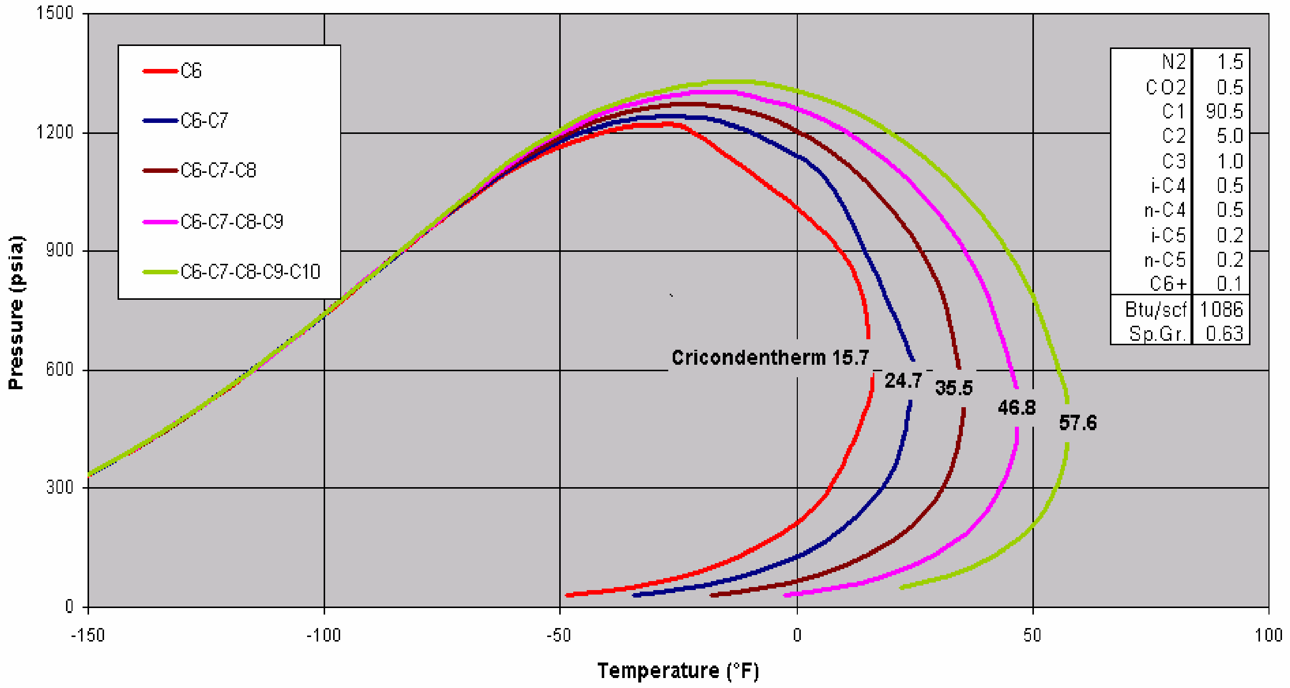
[5] American Society for Testing and Materials standard, ASTM D 1945, and Gas Processors Association, Standard 2261, Analysis for Natural Gas and Similar Gaseous Mixtures by Gas Chromatography.

[6] American Petroleum Institute, Manual of Petroleum Measurement Standards, Chapter 14, Section 1.

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

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

Figure 2. Effect of C_6+ Distribution on Hydrocarbon Dew Point Curves
1086 Btu/scf, 0.1 mol C_6+



Analysis of the discrete factions of higher molecular weight hydrocarbons is accomplished using a method developed by GPA[9]. This provides detail to be able to better estimate the actual hydrocarbon dew point.

Starling has proposed a practical alternativethat draws upon the strength of direct measurement with the versatility of the estimation process[10].
Text that follows probably belongs in recommendations:

When these equations 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. The Peng-Robinson equations are most commonly used. The hexane plus composition is usually characterized as a percentage ratio of hexane, heptane, and octane. Some commonly used values for these percentage characterizations

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[9] GPA Tentative Standard 2286, Tentative Method of Extended Analysis for Natural Gas and Similar Gaseous Mixtures by Temperature Programmed Gas Chromatography, Gas Processors Association.

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

are published in API[11]. The determination of the appropriate characterization for a given pipeline system may be derived from the average compositions of the regional production gathering on that pipeline. An alternative approach is to widen the regional observation, such as, all gulf coast production. Such a definition may span several operating pipelines in the region.

The application of a calculation rule is preferred to the cost prohibitive task of conducting multiple hydrocarbon dew point tests daily or weekly to enforce the pipeline's gas quality specifications. Specification of a precise data point that is independent of the varied operating conditions is most desirable. The cricondentherm is one thermodynamic state condition that meets the operating condition independence requirement. This data point may be calculated from the state equations given the compositional analysis of the natural gas. Using cricondentherm, a conservative rule may be assigned on the basis of market area temperature requirements.

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[11] American Petroleum Institute, Manual of Petroleum Measurement Standards, Chapter 14.