

# **White Paper on Liquid Hydrocarbon Dropout in Natural Gas Infrastructure**

**NGC+ Liquid Hydrocarbon Dropout Task Group  
July 21, 2004**

DRAFT

## Section 1 - Introduction

### Objective

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

**Producers** want the ability to supply natural gas to meet increasing demand. They seek to maximize their natural gas revenue stream by electing to process or not process their gas based on market conditions while satisfying pipeline tariff, safety and environmental requirements.

**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 maximize 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.

**Pipelines** want to provide transportation flexibility to meet demand but are concerned about operational safety and reliability, system integrity and environmental issues.

**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.

**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.

**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.

---

<sup>1</sup> 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.

## **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:

### **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.

### **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), composite concentrations of heavier weight hydrocarbons (such as the mole fraction of heavier weight hydrocarbons measured as the “pentane plus” fraction, referred to as  $C_5^+$  or the “hexane plus”, referred to as  $C_6^+$ ). This section also provides a description of blending, a tool to provide shippers and pipeline operators some flexibility in controlling hydrocarbon liquid drop out.

### **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.

### **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.

### **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 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.

## **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.

## **Appendices (Under Development)**

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

## **Background and Summary of The Issues**

Historically, the commercial value of the hydrocarbon liquids 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 liquids remained part of gas stream. The infrastructure to extract these liquids, referred to as processing, 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 additional processing capacity.

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

This economic environment creates two issues for transmission, distribution and utilization of domestic natural gas. The decreased level of processing has caused 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 the end user. This increases the potential for problems in pipeline and LDC operations with compression, measurement and regulation or over-pressure protection devices, end-use applications such as in home appliances extinguishing or physical damage to gas turbines used to generate electricity. If not addressed these may represent continuing problems because natural gas demand in the United States is expected to rise, especially as demand for natural gas-fired electric generation increases.

## **Natural Gas – From Wellhead to Burner Tip**

This report begins with a brief description of how natural gas makes its way from 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 recovered from within 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 or no heating value. Depending upon the concentrations present, the gas may be treated 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.

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<sup>2</sup>). 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 receipt points generally use on-line continuous gas chromatographs (typically daily volumes greater than for example 5 to 50 MMSCF<sup>3</sup>), whereas manual spot or composite samples are more typical at smaller volume receipt points as well as many offshore receipt points.

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 affects 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 contracted 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

---

<sup>2</sup> million Btus

<sup>3</sup> million standard cubic feet

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.

As stated earlier, when the commercial value of natural gas liquids is at a discount relative to their 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. Generally, these special lines are located in proximity to and upstream of liquids handling infrastructure such as a processing plant. Some pipeline companies have installed various two-phase (i.e., gas and liquid) lines to accommodate the presumption of liquid formation. 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 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, 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. 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 term hydrocarbon liquid drop out. The water content in a pipeline is already covered by tariff provisions and is mentioned here for illustrative purposes.

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 or gravity collection at a low point 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<sup>4</sup>, TSCA<sup>5</sup> or State-listed hazardous waste. Pipeline operating costs may increase dramatically since these liquids often must be disposed of as a hazardous waste. 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.

---

<sup>4</sup> - Resource Conservation and Conservation Recovery Act

<sup>5</sup> - Toxic Substances Control Act

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 liquids dropping out as a result of temperature reductions associated with pressure reductions at city gate stations.

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 install knockout vessels to remove any fugitive solids or liquids that may be in the gas prior to passing 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). It is possible to reduce the pressure enough to chill the gas to below the corresponding dew point, thereby causing liquids to fall out. If the high-pressure gas is not pre-heated sufficiently to keep the low-pressure gas above the dew point or 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 system. 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. 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 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.

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 near-by regulating stations upsetting large portions of a gas distribution system. This results in potential adverse impacts on system reliability or safety.

Additional reliability and safety concerns for LDCs and end users involving gas liquids include the impact to plastic piping and plastic piping components and processes. Hydrocarbon liquids can be absorbed into plastic that can impact plastic pipe fusing processes resulting in joint failures if not properly recognized.

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 billing data, equipment failure, and equipment repair costs. Some LDCs have had to install 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.

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 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, a dilemma exists. 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. 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 may result in increased lost and unaccounted for (LAUF). Ultimately, all shippers on the system must contribute their pro-rata share of the LAUF.

## **Section 2 - Liquid Hydrocarbons in Natural Gas**

### **Sources**

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.

These produced gases can contain both hydrocarbon based gases (those that burn) and non-hydrocarbon gases. Hydrocarbon gases are Methane (C<sub>1</sub>),



Ethane (C<sub>2</sub>), Propane (C<sub>3</sub>), Butanes (C<sub>4</sub>), Pentanes (C<sub>5</sub>), Hexanes (C<sub>6</sub>), Heptanes (C<sub>7</sub>), Octanes (C<sub>8</sub>), and Nonanes plus (C<sub>9</sub>+). The non-hydrocarbon gas portion of the produced gas can contain Nitrogen (N<sub>2</sub>), Carbon Dioxide (CO<sub>2</sub>), Helium (He), Hydrogen Sulfide (H<sub>2</sub>S), water vapor (H<sub>2</sub>O), Oxygen (O<sub>2</sub>), other sulfur compounds and trace gases. CO<sub>2</sub> and H<sub>2</sub>S are commonly referred to as "acid gases" since they form corrosive compounds in the presence of water. N<sub>2</sub>, He and CO<sub>2</sub> are also referred to as diluents since none of these burn, and thus they have no heating value.

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<sub>5</sub>+). The C<sub>5</sub>+ 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.

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 of the other non-hydrocarbon gases identified above.

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<sub>2</sub>, O<sub>2</sub> and water that must be treated to an acceptable level, given the potential to be corrosive.

Hydrocarbon liquids can be produced from gas reservoirs that have been converted to gas storage. Relatively lean pipeline gas injected into the reservoir may become enriched if it comes in contact with hydrocarbon liquids in the reservoir. When the gas is withdrawn from storage as a richer gas, the absorbed hydrocarbons can drop out through cooling of the withdrawn gas due to pressure reductions or contact with cold winter-time ground temperatures. The absorption is greatest during the first few years after a reservoir has been

converted to storage and generally diminishes over time. Clean-up or “processing” of the withdrawn storage gas is usually done at the compressor station used to inject and withdraw the gas from the storage reservoir.

Supply sources connected to interstate 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. 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 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).

### **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 gas from production facilities; aggregate the volumes; and treat and process the gas, 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 gathering pipeline facilities 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 major “contaminants” such as CO<sub>2</sub>, H<sub>2</sub>S and water vapor from the hydrocarbon gases if necessary and then, the NGLs will be removed from the hydrocarbon stream.

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”.

If H<sub>2</sub>S and CO<sub>2</sub> 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 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 facilities.

Once the gas is cleaned of potentially corrosive gases, it is 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, must be removed from the gas stream prior to gas processing or further pipeline transportation.

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 of nitrogen are not commonly used processes and will not be discussed in any greater detail.

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 present 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.

Refrigeration has the least capital cost but also recovers the least NGLs. This process can extract a large percentage of propane and most of the C<sub>4</sub>+ 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.

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 C<sub>3</sub>+ 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 destination in the northern and eastern parts of the U.S.

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 is required. Since the early 1990s, modifications to the cryogenic process have allowed ethane recoveries to reach close to a 99% extraction 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.

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.

## **Economics of Processing**

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 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 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 shrinkage plus fuel use 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 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, the gasoline plants began to chill the gas by the ways of simple pressure reduction/expansion, by passing the gas through 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.

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 gathering, compression, treating and processing. These costs include (1) the cost of the gas lost in the conversion of production gas into residue gas and NGLs. (This process is also known as Shrinkage), (2) the fuel the plant consumes to operate the gathering, compression, treating and extraction processes, (3) the payment or fee collected by the gas processor for these service, and (4) the operating costs for the pipeline and compression that collects the gas from the wellhead and brings it to the gas processing plant; as well as the operating costs for the gas plant. The shrinkage and plant fuel can be 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 heating value of the residue gas is less than 100% of the heating value of the inlet gas stream. Margin is the difference between the revenues received from selling the residue gas and NGLs and the cost of the produced gas. If the margin is positive, it makes economic sense to extract the NGLs from the gas. If the margin is negative, it does not make economic sense to extract the NGLs from the gas.

### **Influence of Ambient Temperatures and Pressure Reductions**

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 referred to as the Joule-Thomson effect.

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 °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 °F. Pipelines located above the frost line may have flowing temperatures less than 32 °F. 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 fall below freezing.

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 gas temperature to go below the hydrocarbon dew point.

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. On the other hand, most 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 should not fall below 32 °F.

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 calculations 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 °F (applicable up to 1000 psig). Thus if the pipeline is delivering gas at a pressure of 800 psig to an end user that 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.

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

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 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.

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 providing a temperature in the example of -16° F with the added heat of 20° F provides a new temperature of 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. This may result in

liquid condensation and blockage or service freeze-up of residential, commercial customers. Of greater significance is the formation of hydrates in the small orifices or tubing of regulator control equipment causing disruption of supply to an LDC.

Figure to be inserted

Because LDC operators have generally required that they receive “dry” gas at the city gate, the reason that heaters have been employed has been to avoid frost heaving or frost balls on the equipment. If heaters need to become more prevalent to reduce hydrocarbon drop out, it may be problematic due to community influence (NIMBY), air permitting, space availability. Also in this case, the LDC bears the burden of making the case for recovering the costs it has incurred to preserve gas quality while receiving no added value for themselves or their ratepayers.

### **Section 3 - Hydrocarbon Liquid Drop Out Control Measures**

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, H<sub>2</sub>S, total sulfur, CO<sub>2</sub>, oxygen and heavy hydrocarbon content (using C<sub>5</sub><sup>+</sup> GPM as the reference) because these constituents in concentrations above recognized limits may 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 CO<sub>2</sub>, H<sub>2</sub>S and water. Much of this is due to the way the gas industry developed.

Many, if not all pipelines have minimum specifications for heating value. This resulted from gas historically being produced from fields high in N<sub>2</sub> or CO<sub>2</sub>. N<sub>2</sub> and CO<sub>2</sub> are inert and do not have a 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. There is no 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 1,100 Btus per cubic foot. The use of a maximum heating value is an inadequate predictor of hydrocarbon drop out because a gas can have a relatively low heating value and a high C<sub>6</sub>+ 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.

Some pipelines selected another parameter for controlling liquids fallout by establishing a C<sub>5</sub>+ or C<sub>6</sub>+ specification. A C<sub>6</sub>+ 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 C<sub>6</sub>+ 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 C<sub>6</sub>+ 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. Nonetheless, correlating C<sub>6</sub>+ levels to HDP can be used as a screening tool or as the basis for establishing a control limit.

More recently, 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 accurate method for determining the potential for hydrocarbon liquid drop out. As of June 2004, eight interstate pipeline operators have established hydrocarbon dew point limits. An HDP limit enables a wide range of gas compositions to be made available to end-users without compromising the safety, operational reliability, system integrity or environmental compliance within the natural gas infrastructure. The use of a hydrocarbon dew point specification will provide the information necessary to design, install and operate pressure reduction stations.

## **Blending**

Blending is the mixing of gas streams that yields a volume-weighted average of the concentrations of each constituent. Pipelines 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 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

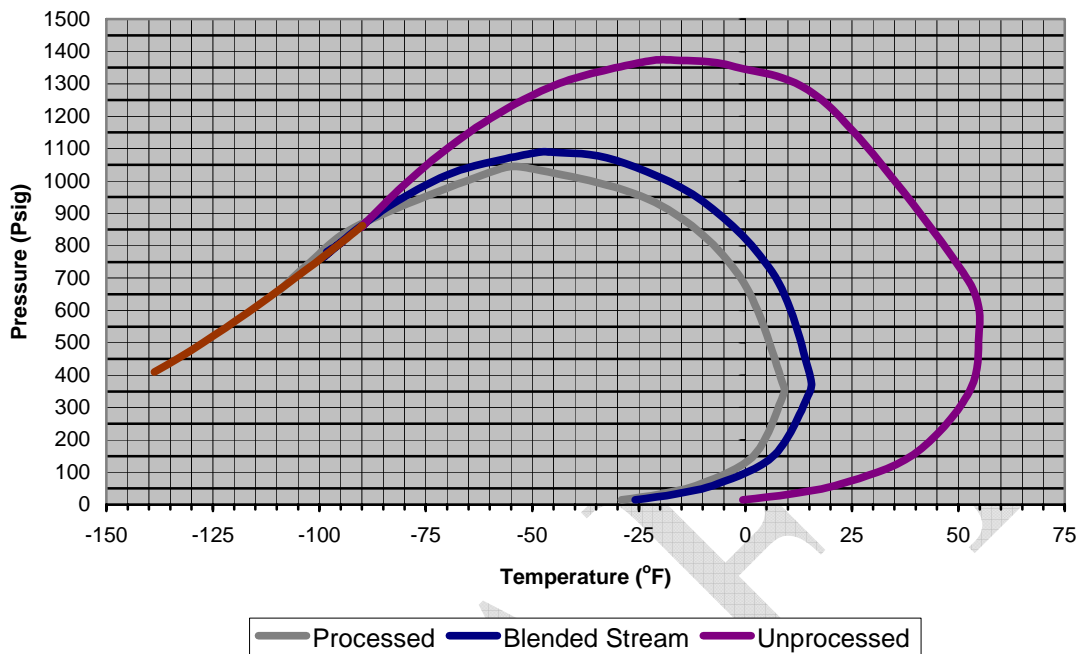
condensate from forming, blending provides pipelines a mechanism to achieve that goal while still maintaining the flexibility to accept gas streams with varying hydrocarbon levels. However, the ability to utilize gas blending to manage HDP is dependent on a number of factors including pipeline configuration, receipt and delivery location, gas supply composition, gas markets and shippers' nominations, geographic location, flowing gas temperature.

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 following example of how a pipeline can benefit from blending is provided to help demonstrate how the effect of blending. When a gas stream with a low HDP is mixed, or commingled with a gas stream with higher HDP, the resulting commingled or blended stream will have a HDP somewhere between the two individual HDP levels. In the example shown in Figure 1<sup>6</sup> below, a volume of Gulf Coast unprocessed gas with a HDP temperature of 55°F has been blended with a processed gas stream with a HDP temperature of 10°F (need to add blending basis). The resulting blended gas stream has a HDP temperature of approximately 15°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 the equilibrium of the new mixture creating a completely separate and unique gas quality.

---

<sup>6</sup> Cricondentherm

Figure 1 - Gulf Coast  
Hydrocarbon Dewpoint Calculations



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

**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 illustrated in the example 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 homogenous blend.

**Contractual blending** is when a producer of rich gas contracts with a lean 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 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 within the pipeline.

## **Heaters**

Most problematic hydrocarbon condensation occurs at points of pressure regulation (or immediately downstream). Water bath heaters, in which heated water 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. In addition, since pressure regulation typically occurs post custody transfer from the pipeline to the LDC or other end user, gas heaters may not be available due to space limitations in urban environments. As a result, some LDC's or end users partially depend on pipeline heat of compression to mitigate temperature associated with local pressure reduction. A problem for LDCs may be that a system may not have the physical space to install a heater. For a large system, an M&R station heater could be as large as 8 feet in diameter and 20 feet in length.

## **Offshore Gas and Liquids Handling**

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.

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<sup>7</sup>, it is most

---

<sup>7</sup> 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<sub>5</sub> components only. The J-T process (or Joule-Thompson) 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.

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. This common practice in the offshore industry further complicates the application of blending on each pipeline.

## 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 2. 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 from the two-phase gas-liquid region. The right-hand side of the curve is the dew point line and divides the two-phase gas-liquid region 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 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 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 (e.g.- water, as shown in Figure 2). 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**.

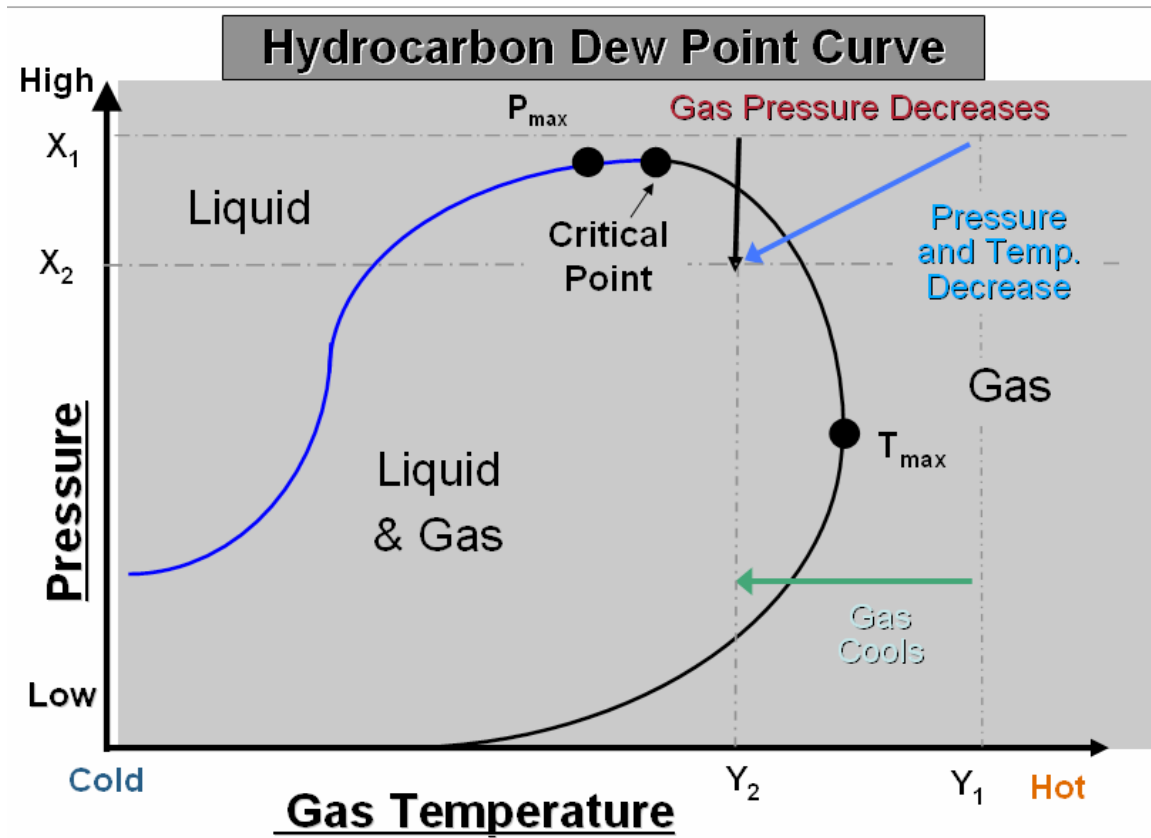


Figure 2 - Hydrocarbon Dew Point Curve for a Typical Natural Gas Mixture

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 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  $C_6+$ . 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 3 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.

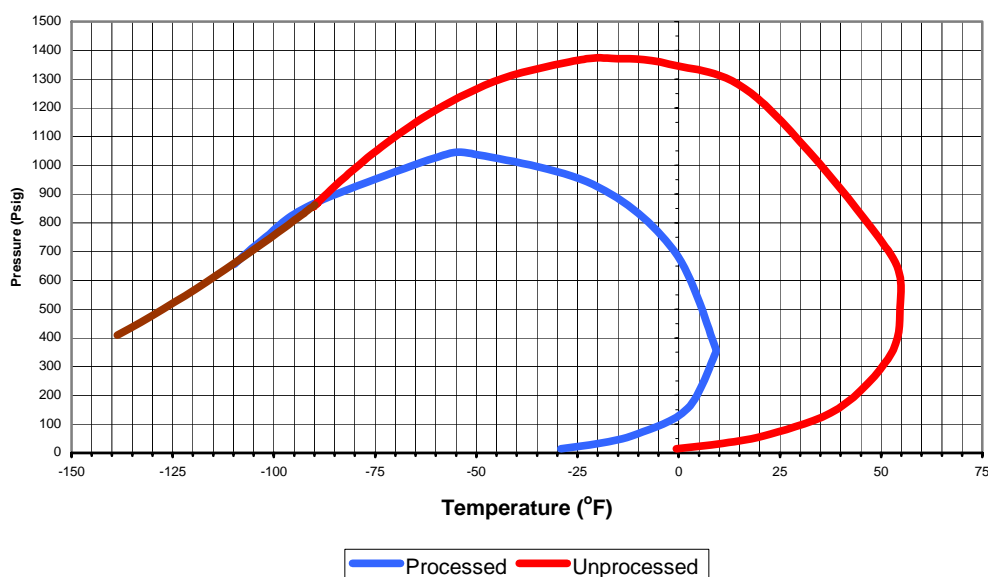


Figure 3 – Contrast of Unprocessed and Processed Natural Gas

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 1 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 million standard cubic feet of gas) for the liquefiable components in each stream.

## Section 5 - Historical Levels of Hydrocarbons and Hydrocarbon Dew Point

No data has been developed that tracks historical 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.

Prior to the advent of gas processing, hydrocarbon dew points in pipelines and market areas would approach ambient temperatures (between 30 and 60 °F). Pipelines collected liquids and developed their own methods to force these liquids from the gas prior to its delivery to customers. As processing developed

with increasing natural gas production, 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  $C_6+$  and reduced the cricondenthem about 10 °F at the plant outlet.

Propane as a refrigerant 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 cricondenthem from 80 °F at the plant outlet to level slightly above 0 °F and recovered 50% or so of the propane and 80% or more of the  $C_4+$ . The gas processed was very rich, on the order of 1200 -1400 Btu, 4 - 7 GPM 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  $C_4+$ ). Cricondenthem of the processed gas was -30 °F or lower.

When oil prices increased dramatically 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  $C_3+$  producing cricondenthem 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 plants. The remaining production formerly processed in these plants was aggregated with other fields and processed in regional processing centers, i.e. the Phillips-GPM Oklahoma Super System. Non-producer plant owners (e.g. GPM, Dynegy, Enterprise, Tejas, Associated (Duke Energy)) emerged at the beginning of the 1990s.

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.

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

Technology Vintage	Processing Technology	Achievable Cricondenthem HDP °F
1940-60	Refrigeration	≅ 0
1960-75	Lean oil	≅ -10
1975 on	Cryogenic	≅ -100

Actual cricondenthem 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

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 merits of each in managing hydrocarbon drop out.

### Measurement

The most commonly used method of hydrocarbon dew point detection 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)<sup>8</sup>. For many years this device has been used for moisture detection and can be used for hydrocarbon dew point determination also. 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

<sup>8</sup> 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.

14.1. 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 the 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 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. A conventional chilled mirror direct measurement instrument in general can be subject to operator variability and interferences including but not limited to water vapor. A typical HDP 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  $\pm 0.5^\circ\text{F}$ <sup>9</sup>. 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.

Automatic, continuous online dew point detection units are commercially available. 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.

## **Estimation**

Estimation relies on a three-step process, sampling, analysis and calculation. The most common means of sampling and analysis 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

---

<sup>9</sup> Warner, H. R., Leamer, E. E., Spence, A. P., Bone, R. L., Hubbard, R. A., Bernos, J., and Kriel, W. A., "Hydrocarbon Dew Point Determination of Lean Natural Gases," Proceedings, 80th Annual Convention, Gas Processors Association, 2001.

uses a combination of columns to analyze for methane through pentane and then treats all compounds with molecular weights greater than hexane as a fixed mole fraction average of C<sub>6</sub>, C<sub>7</sub>, and C<sub>8</sub><sup>10</sup>. This chromatograph is referred to as a C<sub>6</sub><sup>+</sup> chromatograph.

Manual sampling with off-site analyses of the samples can be used as an alternative. Samples of the gas are collected in a sample cylinder (canister) or on charcoal tubes using standard methods published by GPA<sup>11</sup> and referenced by API<sup>12</sup>. Samples are analyzed using a chromatograph typically using C<sub>6</sub><sup>+</sup> chromatograph described above.

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 (e.g.- pressure, temperature, etc.). Two commonly accepted sets of state equations are Peng-Robinson<sup>13</sup> and Soave-Redlich-Kwong<sup>14</sup>. The effects of Equations of State and of gas composition are also part of an on-going project with the API under section 14.1.

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 representative the sample, 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. For example, having a small percentage of Nonane (C<sub>9</sub>) or Decane (C<sub>10</sub>) may significantly elevate the calculated hydrocarbon dew point value for the natural gas mixture. In applying a HDP limit it is prudent to conduct periodic validation based on use of an "extended analysis", through C<sub>9</sub><sup>+</sup> to enable demonstration of the "split". The split is the relative proportions of C<sub>6</sub>, C<sub>7</sub>, and C<sub>8</sub> in a gas mixture. Some commonly used values for these percentage characterizations are published in a GPA standard<sup>15</sup>.

---

<sup>10</sup> 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.

<sup>11</sup> Gas Processors Association Standard 2166, "Obtaining Natural Gas Samples for Analysis by Gas Chromatography," 1986.

<sup>12</sup> American Petroleum Institute, Manual of Petroleum Measurement Standards, Chapter 14, Section 1. "Collecting and Handling of Natural Gas Samples for Custody Transfer," June 2001.

<sup>13</sup> Peng, D. Y. and Robinson, D. B., Industrial and Engineering Chemistry Fundamentals, 15: 59, 1976.

<sup>14</sup> Soave, G., Chemical Engineering Science, 27: 1197, 1972.

<sup>15</sup> Gas Processors Association Standard 2261, "Analysis for Natural Gas and Similar Gaseous Mixtures by Gas Chromatography," 2000.

However, recent research<sup>16</sup> has shown that use of the GPA 60/30/10 C<sub>6</sub>/C<sub>7</sub>/C<sub>8</sub> characterization to compute hydrocarbon dew points will usually underestimate the dew point temperatures and cricondentherm. Work conducted by the PRCI (an industry research organization) indicates that a 47:37:16 split is generally applicable<sup>17</sup>. The determination of the appropriate characterization for a given pipeline system may be more accurately derived from the weighted average compositions of the regional production gathering 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.

Dr. Ken Starling, an expert in this area, has proposed a practical alternative that draws upon the strength of direct measurement with the versatility of the estimation process<sup>18</sup>.

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 hexane-plus composition is usually characterized as a percentage ratio of Hexane, Heptane, and Octane for calculating heating value and specific gravity. Some commonly used values for these percentage characterizations are published in a GPA standard<sup>19</sup>. However, recent research<sup>20</sup> has shown that use of the GPA 60/30/10 C<sub>6</sub>/C<sub>7</sub>/C<sub>8</sub> characterization to compute hydrocarbon dew points will usually underestimate the dew point temperatures and cricondentherm. The determination of the appropriate characterization for a given pipeline system may be more accurately derived from the weighted average compositions of the regional production gathering 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.

---

<sup>16</sup> 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, July 2004.

<sup>17</sup> PRCI

<sup>18</sup> Starling, Kenneth A., Peng-Robinson Equation of State Natural Gas Dew Points, AGA Technical Conference

<sup>19</sup> Gas Processors Association Standard 2261, "Analysis for Natural Gas and Similar Gaseous Mixtures by Gas Chromatography," 2000.

<sup>20</sup> 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, July 2004.

## Section 7 - Recommendations

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

The NGC Task Group found that HDP has the broadest applicability and is the preferred control parameter.

An HDP may be applied to control hydrocarbon liquid drop out.

The NGC Task Group found that the cricondenthem HDP, i.e., the temperature above which hydrocarbon liquids will not drop out regardless of pressure, provides the greatest operational flexibility for all stakeholders.

The cricondenthem HDP may be applied to control hydrocarbon liquid drop out.

The cricondenthem HDP is the preferred control parameter.

An HDP temperature at a specified pressure or pressure range may also be applied to control hydrocarbon liquid drop out.

The NGC GQ Task Group also found that  $C_6^+$  can be a valid means of controlling hydrocarbon liquid drop out as it tends to correlate with HDP temperature.

A  $C_6^+$  limit may be applied to control hydrocarbon liquid drop out.

The NGC GQ Task Group also found that other approaches such as  $C_5^+$  or heating value are not effective means of predicting and controlling hydrocarbon liquid drop out.

The NGC Task Group recommends that the control parameters be determined by one of two means:

- 1) direct measurement - using the Bureau of Mines Chilled Mirror method for HDP temperature at operating pressure
- 2) indirect measurement - calculation using an equation of state with data derived using gas chromatography for HDP or  $C_6^+$

The NGC Task Group found that in applying either an HDP or  $C_6^+$  control parameter, a plan must be established for periodic validation of the assumptions used in calculations as detailed below.

- 1) It is recommended that in estimating HDP, when using  $C_6^+$  data, that the periodic validation entails use of an "extended analysis" to enable demonstration of the "split". The split refers to the relative proportions of  $C_6$ ,  $C_7$ , and  $C_8$  in a gas mixture.
- 2) It is recommended that in applying a  $C_6^+$  limit that a pipeline operator establish a plan for periodic validation. {Periodically validating that analysis of hydrocarbons with gas chromatography through  $C_5$  and quantification of  $C_6$  and heavier hydrocarbons as  $C_6^+$ , correlates with HDP}.

The NGC Task Group recognizes that in certain instances, parties may, to the extent operationally feasible, 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.

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

- 1) Build the database to support use of  $C_6^+$  split assumptions for heavier hydrocarbons 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.