Related U.S. Application Data

Publication Classification

ABSTRACT

Contemplated plants include a NGL recovery portion and a LNG liquefaction portion wherein the NGL recovery portion provides a low-temperature and high-pressure overhead product directly to the LNG liquefaction portion. Feed gas cooling and condensation are most preferably performed using refrigeration cycles that employ refrigerants other than the demethanizer/absorber overhead product. Thus, cold demethanizer/absorber overhead product is compressed with the turbo-expansion and delivered to a liquefaction portion at significantly lower temperature and higher pressure without net compression energy expenditure.
LNG Liquefaction Composition Curve

Temperature (F)

Enthalpy Change (MMBtu/hr)

Figure 8
CONFIGURATIONS AND METHODS OF INTEGRATED NGL RECOVERY AND LNG LIQUEFACTION

[0001] This application claims priority to our copending U.S. provisional patent application with the Ser. No. 60/697,467, which was filed 7 Jul. 2005.

FIELD OF THE INVENTION

[0002] The field of the invention is natural gas liquids (NGL) recovery and liquefied natural gas (LNG) liquefaction, and particularly integrated plant configurations for such processes.

BACKGROUND OF THE INVENTION

[0003] While the crude oil supply in the world is diminishing, the supply of natural gas is still relatively abundant in many parts of the world. Natural gas is typically recovered from oil and gas production wells located onshore and offshore. Depending on the particular formation and reservoir, natural gas also contains relatively low quantities of non-methane hydrocarbons, including ethane, propane, i-butane, n-butane, pentanes, hexane and heavier components, as well as water, nitrogen, carbon dioxide, hydrogen sulfide, mercaptans, and other gases.

[0004] Natural gas from the wellheads is commonly treated to remove sulfur components, compressed, and transported to consumers in high pressure pipelines. However, in remote locations lacking the necessary pipeline infrastructure, natural gas is commonly transported by liquefying the natural gas and transporting the gas in liquid form (e.g., using LNG cargo carriers). Unfortunately, liquefaction of natural gas is problematic as natural gas also contains aromatics (e.g., benzene) and heavy hydrocarbons, which solidify when chilled to the cryogenic temperatures. Consequently, most aromatic hydrocarbons must be removed to an extremely low level (typically less than 1 ppmv) to avoid freezing and plugging the cryogenic heat exchange equipment. Additionally, at least a portion of the lighter hydrocarbons such as C2, C3 and C4 must be removed when LNG is imported to the North America natural gas market, which typically requires a leaner natural gas. Typical North America pipeline gas contains mostly the clean burning methane gas with higher heating values between 1050 to 1070 Btu/SCF. Recovery of the non-methane components can be economically attractive as these hydrocarbons can be sold at a premium over natural gas. For example, C2 is often used as feedstock for petrochemical manufacture, C3 and C4 are marketed as LPG fuels, and the C5+ hydrocarbons can be further processed to be used for gasoline blending.

[0005] There are numerous configurations and methods known in the art for the recovery of C2 and C3+ NGL from a natural gas feed. However, all past efforts have been focused on the removal of NGL hydrocarbons from natural gas using standalone NGL recovery plants, which operate independently from LNG liquefaction plants. These extraction processes generally produce a low to medium pressure residue gas which then requires compression and further chilling before liquefaction in a liquefaction plant. Typical examples for plants to recover C2 and C3+ components from natural gas include those employing expander processes described in U.S. Pat. No. 4,157,904 to Campbell et al., U.S. Pat. No. 4,251,249 to Gulsby, U.S. Pat. No. 4,617,039 to Buck, U.S. Pat. No. 4,690,702 to Paradowski et al., U.S. Pat. No. 5,275,005 to Campbell et al., U.S. Pat. No. 5,799,507 to Wilkinson et al., or U.S. Pat. No. 5,890,378 to Rambo et al.

[0006] Other known high C2 recovery processes (e.g., U.S. Pat. Nos. 6,116,050), require to let down a portion of the high pressure residue gas to the NGL recovery column as a methane rich reflux using a Joule-Thomson (JT) valve. While these processes improve C2 recovery to at least some degree, the energy spent for residue gas recompression renders the process often uneconomical. To overcome some of these disadvantages, twin-column configurations may be implemented in which a high pressure absorber is fluidly coupled to a lower pressure distillation column to improve NGL recovery efficiency as described in commonly owned U.S. Pat. No. 6,837,070. However, as these NGL processes operate independently from the liquefaction plants, they will generally require additional compression and refrigeration prior to LNG liquefaction of the residue gas.

[0007] In still other known configurations for NGL processing, a scrubbing column is used within the LNG liquefaction plant for the removal of the heavier components (C6+). For example, a side stream from a spiral wound cryogenic exchanger is processed in a scrubber and fractionation unit, as shown in U.S. Pat. No. 6,308,531 to Roberts et al. While such process can advantageously be used to eliminate wax formation by removal of C6+ and heavier components, they are not suitable for the removal of C2+ components, especially at high levels (65% or higher C2 recovery) and consequently fail to produce the lean residue gas that can be liquefied for the North American natural gas market. Still further known NGL recovery processes that are integrated to LNG liquefaction as disclosed by Roberts et al in Pat. No. 6,662,589 teach that a C2 enriched liquid can be used for C3 absorption in a high pressure absorption column. While such processes attempt to operate the absorption column at high feed gas pressure (e.g., 800 psig or higher) to reduce energy consumption, it should be noted that LNG separation suffers significantly due to decreasing relative volatilities of the NGL components, consequently producing NGL with excessive methane content. Moreover, such process schemes typically fail to achieve high C2 and C3 recovery (e.g., greater than 60%).

[0008] Presently known LNG liquefaction processes generally include several steps in which natural gas is cooled and condensed, using either pure component refrigeration cycles or one or more mixed refrigerants cycles. The cascade-refrigeration cycle chills and liquefies the feed gas with several pure component refrigerants having successively lower boiling points, such as propane, ethylene, methane and nitrogen. The mixed refrigerant cycle uses a mixture of refrigerants and can therefore be configured to use a single compressor and heat exchanger, which simplifies the equipment configuration. Alternatively, the feed gas can also be cooled by a propane pre-cooling refrigeration cycle or by expansion of natural gas or nitrogen using either Joule-Thomson expansion valves or an expansion turbine. Unfortunately, most known standalone LNG liquefaction processes using single or multiple refrigeration cycles (either cascade refrigeration or mixed refrigerant cycle) have relatively low efficiencies when C2 or C3 recovery is incorporated upstream of the LNG liquefaction plant.
Thus, while numerous plant configurations and methods for NGL recovery and LNG liquefaction are known in the art, all or almost all of them, suffer from various disadvantages. Thus, there is still a need for improved NGL recovery and LNG liquefaction, and especially plants in which NGL recovery and LNG liquefaction are integrated.

SUMMARY OF THE INVENTION

The present invention is directed to configurations, plants, and methods for natural gas processing and liquefaction in which a cold separator overhead product is directly compressed in a compressor that is driven by a feed gas expander, and wherein the compressed cold separator overhead product is then liquefied in a liquefaction unit. Most advantageously, such plants integrate NGL processing and LNG liquefaction in an efficient, cost effective, and technically simple manner.

In one preferred aspect of the inventive subject matter, a gas processing plant includes a separator that is configured to receive a partially expanded vapor portion of a natural gas feed stream, wherein the separator is further configured to produce a cold overhead product stream. An expander in such plants is operationally coupled to drive a compressor, wherein the expander is configured to produce the partially expanded vapor portion and wherein the compressor is configured to produce a compressed cold overhead product stream from the cold overhead product stream, and wherein the separator and the compressor are fluidly coupled to each other such that the compressed cold overhead product stream has a pressure of at least 700 psig at a temperature of no warmer than -50°F.

Most typically, the separator is configured to receive another expanded vapor portion of the natural gas feed stream at a separate location. In some embodiments, the separator is configured to operate as a demethanizer, and a deethanizer may be coupled to the separator wherein the deethanizer is configured to produce a C3+ product and a C2 product, which may be at least partially combined with the cold overhead product stream (e.g., to adjust heating value). In other embodiments, the separator is configured to operate as a refluxed absorber. In such configurations, the demethanizer is preferably configured to provide a reflux stream to the absorber, wherein the demethanizer is most typically configured to operate at a lower pressure than the absorber. Where desirable, a conduit may be provided that delivers a cooled absorber bottom product to the absorber for C2 recovery and/or a conduit that provides a heated absorber bottom product to the absorber for C2 rejection.

Therefore, in another aspect of the inventive subject matter, a method of producing a liquified natural gas may include a step of producing a cold overhead product stream in a separator and compressing the product stream in a compressor without prior substantial heating of the cold product stream, wherein the compressor is driven by an expander that expands a vapor portion of a natural gas feed, and wherein the expanded vapor portion is fed into the separator. In another step, the cold compressed product stream is then liquefied in a liquefaction unit.

It should be recognized that the separator may be operated as a demethanizer that further receives another portion of a natural gas feed, and that in such methods a deethanizer may be coupled to the separator, wherein the deethanizer receives a bottom product of the separator and optionally provides a C2 product to the cold overhead product stream. In alternative aspects of the inventive subject matter, the separator is operated as a refluxed absorber. Here, it is generally preferred that a demethanizer provides a reflux stream to the absorber and that the absorber provides a bottom product to the demethanizer, wherein the demethanizer is operated at a lower pressure than the absorber. For flexible C2/C3+ recovery, it is contemplated that the absorber bottom product may be heated for C2 rejection prior to entering the deethanizer and/or cooled for C2 recovery prior to entering the demethanizer most typically, a deethanizer is fluidly coupled to the demethanizer in such plants and receives a bottom product of the demethanizer for separation and recovery of C2 and C3. Regardless of the manner of operation of the separator, it is generally preferred that the cold compressed product stream has a pressure of at least 700 psig and a temperature of no warmer than -50°F.

Viewed from a different perspective, a method of producing LNG will include a step of separating in a separator a cold overhead product from a natural gas containing feed gas. In another step, the cold overhead product is compressed using expansion energy from the feed gas to form a cold compressed overhead product at a pressure and temperature suitable for liquefaction in a liquefaction unit, wherein the cold compressed overhead product is formed at neutral or negative net compression energy requirement. Typically, the step of separating is performed in an absorber or a demethanizer, and/or the cold compressed overhead product has a pressure of at least 700 psig and a temperature of no warmer than -50°F.

Various objects, features, aspects and advantages of the present invention will become more apparent from the following detailed description of preferred embodiments of the invention.

BRIEF DESCRIPTION OF THE DRAWINGS

Prior Art FIG. 1 is a schematic of an exemplary known plant configuration for recovery of NGL and LNG liquefaction.

FIG. 2A is a schematic of an exemplary plant configuration using a single column configuration for production of a cold compressed overhead product and separation of C2 and/or C3.

FIG. 2B is a schematic of one exemplary plant configuration using a twin column configuration for production of a cold compressed overhead product and separation of C2 and/or C3.

FIG. 3 is a more detailed schematic of an exemplary plant according to FIG. 2A with a cascade refrigeration cycle and two mixed refrigerant cycles for NGL recovery and LNG liquefaction.

FIG. 4 is a more detailed schematic of an exemplary plant according to FIG. 2B with a cascade refrigeration cycle and two mixed refrigerant cycles for NGL recovery and LNG liquefaction.

FIG. 5 is a more detailed schematic of an exemplary plant according to FIG. 2B with two cascade refrigeration cycles and one mixed refrigerant cycles for NGL recovery and LNG liquefaction.
FIG. 6 is a more detailed schematic of an exemplary plant according to FIG. 2B with two cascade refrigeration cycles, and a mixed refrigerant/cascade cycle for NGL recovery and LNG liquefaction.

FIG. 7 is a more detailed schematic of an exemplary plant according to FIG. 2B with three cascade refrigeration cycles for NGL recovery and LNG liquefaction.

FIG. 8 is a graph depicting a composite heat curve of the LNG liquefaction process.

DETAILED DESCRIPTION

The inventor discovered that natural gas processing and liquefaction can be integrated in various configurations, plants, and methods in which cold compression of lean natural gas in a compressor that is driven by a feed gas expander provides a cold high-pressure natural gas that can be directly liquefied in a liquefaction unit. Therefore, the net compression energy requirement for the lean natural gas is neutral or even negative, while feed gas cooling and condensation are achieved using distinct refrigeration cycles. Among other advantages, it should be appreciated that contemplated configurations and methods allow for an integrated NGL recovery and LNG liquefaction process in which 99% propane and up to 85% ethane can be recovered from a natural feed gas.

Prior Art FIG. 1 shows a standalone C2 NGL recovery process that is coupled with a standalone LNG liquefaction plant. Here, a contaminant-free and dried feed gas stream 1, typically supplied at about 1200 psig is cooled in a feed gas exchanger 51 using refrigeration content of the column overhead vapor, side reboiler stream 22, and external refrigerant 32. Liquid is then removed from the chilled feed gas in separator 52 and sent to the LNG column 58 that acts as a demethanizer. The flashed vapor from separator 52 is split into two portions, with one portion being cooled in exchanger 54 to provide reflux to the column, and the other portion being expanded in turbo-expander 64 to provide a cooled feed stream that is sent to a lower section of the column for rectification. It should be noted that the above standalone gas subcooled process produces a residue gas at about ambient temperature and approximately 450 psig. Such relatively low pressure and high temperature is predominantly due to the use of the residue gas as refrigerant for feed gas cooling and subcooling of a vapor portion of the feed gas and the pressure drops in the heat exchangers. Consequently, substantial recompression in re-compressor 100 and additional cooling (cooler not shown) of the residue gas is typically required prior liquefaction in the plant, which significantly reduces process efficiency and economics.

In contrast, contemplated configurations presented herein preserve substantially all of the refrigeration content in the separator overhead product by directly feeding the residue gas (separator overhead product) into the compressor, without incurring the pressure drops in heat exchangers of prior arts. As the compressor is driven by the vapor expansion of the feed gas and as the residue gas is significantly colder than in heretofore known configurations, substantially higher compressor discharge pressures at notably lower temperatures can be achieved. Moreover, where the separator is operated as an absorber, the compressor discharge pressure can be even higher. It should therefore be appreciated that in most contemplated configurations and methods, the residue gas pressure is higher than 700 psig (typically between 700 and 900 psig) at a temperature of lower than −50°F. (typically −50°F to −80°F) can be achieved.

FIG. 2A depicts one exemplary plant configuration in which the separator is operated as a demethanizer, while FIG. 2B depicts an exemplary plant configuration in which the separator is operated as a refluxed absorber, and in which a demethanizer and deethanizer then operate at a lower pressure to recover C2 and/or C3+ components. As can be seen from both figures, C2 content in the LNG may be adjusted to a predetermined or desired level by either combining separated C2 with the column overhead product as shown in dashed lines in FIG. 2A, or by temperature control of the absorber bottom product that is fed into the demethanizer (as shown in dashed lines for C2 rejection) depicted in FIG. 2B.

It should be recognized that contemplated integrated NGL recovery and liquefaction processes significantly reduce equipment cost and energy consumption of LNG liquefaction, while allowing fractionating the NGL into C2 and C3+ products. Such configurations and processes will produce an LNG predominantly comprising methane that can be used in and/or exported to North America with heating values complying with gas pipeline standards. Moreover, it should be noted that contemplated plants can be operated to produce LNG with variable ethane and propane content for non-US markets.

Using such configurations and methods, high propane recovery (i.e., at least 95%) and high ethane recovery (up to 85%) from a feed gas with relatively high pressure (e.g., between about 800 psig to 1600 psig) can be realized by operating an absorber at a higher pressure than the demethanizer. A compressor is then used to recycle the demethanizer overhead to the absorber, while the absorber bottoms product is expanded to provide cooling to the demethanizer. The overhead vapor from the absorber is compressed using power (preferably exclusively) generated from the feed gas expander. Therefore, it should be appreciated that contemplated configurations and methods significantly reduce the energy consumption of the integrated liquefaction plant. Further configurations related to some aspects of the inventive subject matter are disclosed in our pending U.S. patent application with the Ser. No. 10/478,075, which is incorporated by reference herein.

While not limiting to the inventive subject matter, it is typically preferred that the refrigeration processes for both NGL recovery and residue gas liquefaction can be configured to employ a combination of one or more vaporizing refrigeration cycles to provide chilling at least three temperature ranges: A first temperature range of 10°F to −35°F, a second temperature range of −60°F to −160°F, and a third temperature range of −180°F to −270°F for gas liquefaction.

FIG. 3 depicts a more detailed schematic of an exemplary configuration in which the separator is operated as a demethanizer (see also FIG. 2A). With respect to the feed gas streams it is contemplated that numerous natural and man made feed gas streams are suitable for use in conjunction with the teachings presented herein, and espe-
cially preferred feed gas streams include natural gases, refinery gases, and synthetic gas streams from hydrocarbon materials such as naphtha, coal, oil, lignite, etc. Consequently, the pressure of contemplated feed gas streams may vary considerably. However, it is generally preferred that appropriate feed gas pressures for plant configurations according to the inventive subject matter will generally be in the range between 800 psig and 1600 psig, and that at least a portion of the feed gas is expanded in a turboexpander to provide cooling and/or power for the residue gas recompression. The overall mass balances illustrating gas composition and flow rate for the exemplary feed gas and products are shown in Table 1 below.

<table>
<thead>
<tr>
<th>TABLE 1</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>MOL %</strong></td>
</tr>
<tr>
<td>CO2</td>
</tr>
<tr>
<td>N2</td>
</tr>
<tr>
<td>C1</td>
</tr>
<tr>
<td>C2</td>
</tr>
<tr>
<td>C3</td>
</tr>
<tr>
<td>IC4</td>
</tr>
<tr>
<td>IC5</td>
</tr>
<tr>
<td>Ne5</td>
</tr>
<tr>
<td>C6</td>
</tr>
<tr>
<td>C7</td>
</tr>
<tr>
<td>MMscf/d</td>
</tr>
<tr>
<td>BPD</td>
</tr>
<tr>
<td>Metric Tons/yr</td>
</tr>
</tbody>
</table>

[0034] Here, feed gas stream 1 enters the plant at about 1200 psig and 120° F., and is cooled in exchanger 51 to typically 10° F. to ~30° F., forming stream 2, using multiple cooling streams including liquid stream 5 from separator 52, side reboiler stream 22 from demethanizer 61, flash vapor 70 from LNG storage tank 69, and a propane refrigerant stream 32. The propane refrigerant is typically generated in cascade propane refrigeration system 101, vaporizing at least three different pressure levels to provide chilling for heated stream 33. It should be noted that various exchangers (e.g., plate and fin exchangers or spiral wound exchangers) can be used to achieve a lower temperature approach that provides high thermodynamic efficiency as demonstrated in the integrated composite curves of FIG. 8.

[0035] The chilled feed gas stream 2 is separated in separator 52, forming a gaseous portion 3 and a liquid portion 4. The liquid portion 4 is led down in pressure in JT valve 53 forming stream 5, and optionally heated to stream 6 with the heat content from the feed gas prior to entering the demethanizer 61 (reboiled by reboiler 63). The gaseous portion 3 from separator 52 is split into two portions. One portion (stream 7) is routed to the exchanger 54 to provide reflux to the absorber, and the other portion (stream 8) is expanded in turbo-expander 64 to produce a chilled vapor stream 10, typically at ~80° F. to 100° F. and to generate power to drive the residue gas compressor 65. The chilled vapor 10 is fed to the demethanizer 61, which operates between 400 psig to 650 psig, most typically at 450 psig. It should be appreciated that the flow ratio of stream 8 to stream 3 can be adjusted to tailor to a desired C2 recovery level, and/or to meet desired C2 product rates. Demethanizer 61 is refluxed with top reflux stream 12 (formed from stream 9 via JT valve 55) from exchanger 54. The reflux stream is preferably chilled in exchanger 54 to about ~125° F. to ~155° F. using mixed refrigerant stream 72 and 74 (via stream 73 and JT valve 91) that is generated from stream 72 by compressed mixed refrigerant of refrigeration unit 102. The so heated refrigerant 75 is then returned to refrigeration unit 102.

[0036] The demethanizer 61 produces an overhead vapor stream 28 at about ~120° F. to ~140° F. and a bottoms stream 14 at about 20° F. to 80° F. The overhead vapor is compressed by the residue gas compressor 65 forming a discharge stream 29, typically at about 700 psig to 900 psig and ~50° F. to ~80° F. It should be particularly appreciated that compression of a cryogenic vapor is energy efficient and results in a high compression ratio across the compressor, which significantly reduces the refrigeration consumption for liquefaction (using the third temperature level). Moreover, it should be noted that the compression of the overhead product requires no net energy as the compressor is coupled to the expander 64. Thus, by using a relatively high feed gas pressure (e.g., about 1000 psig) and compression of cold separator overhead product, chilled residue gas can be delivered to the liquefaction unit without net compression energy expenditure at a higher pressure and lower temperature than other known NGL separation processes would allow. The cold compressed residue gas 29 is then further chilled and condensed in exchanger 67 to about ~255° F. to ~265° F. using mixed refrigerant 79 operating at ~250° F. to ~270° F. Refrigerant 79 is produced by the mixed refrigeration system 103, after the compressed stream 76 is chilled in exchangers 54 and 67 (to form stream 78), and JT'd via valve 92. Heated stream 80 is then returned to the refrigeration system 103.

[0037] The liquefied residue gas 81 is further led down in pressure to about 16.0 psig via JT valve 90 to form stream 82, which is stored in LNG storage tank 69. The LNG product is withdrawn as stream 30, optionally combined with JT'd C2 product stream 15 and exported to the ship loading terminal, storage container, or other use. In some cases, and depending on the natural gas composition and temperature from the liquefier exchanger, significant quantities of light gas may evolve, which can be used as refrigeration source in subsequent exchangers to form a fuel gas 71 that is typically compressed to fuel header pressure.

[0038] As pointed out above, a portion of the ethane product stream 15 can be directed from deethanizer 59 to the LNG storage tank, to be blended with the lean LNG to produce a heavier and richer LNG, which may be required to accommodate the various LNG markets. Deethanizer 59 receives the bottom product of the demethanizer and is reboiled by reboiler 34 to produce a C3+ bottom product that is withdrawn as liquid 23 for storage or further processing. Deethanizer overhead condenser 62 provides chilling for the C2 overhead product. One portion of the of the overhead product is provided as deethanizer reflux stream 18 from separator drum 68 to the column via pump 59 while another portion 19 is routed to storage or other use as stream 17.

[0039] Most preferably, the first column (demethanizer) is configured to separately receive a first and a second portion of a feed gas vapor, wherein the first portion of the feed gas vapor is chilled by the first level refrigeration, and the second portion is chilled by the second level refrigeration that provides reflux to the demethanizer. In such configura-
tions, it should be noted that a flow control unit (typically automated and using a controller programmable according to a desired product composition and/or feed gas composition) adjusts at least one of the first and second portions of the feed gas vapor to produce the desirable recovery levels of ethane, from 10% to 85% of the feed gas while maintaining a high C3 (98% or above) recovery.

[0040] It is further contemplated that at least a portion of the demethanizer bottoms product is fed to the deethanizer that fractionates the demethanizer bottom product into an ethane overhead and a C4+ bottoms product. Thus, it should be recognized that contemplated methods and configurations allow production of C2 at variable rates by blending at least a portion of overhead C2 liquid with the LNG. It should further be recognized that blending significantly simplifies NGL recovery plant operation and allows the same process conditions (temperatures and pressures) be maintained, regardless of the net C2 production rates.

[0041] FIG. 4 depicts a more detailed schematic of an exemplary configuration in which the separator is configured as an absorber that operates at a higher pressure than a downstream demethanizer and deethanizer (see also FIG. 2A). With respect to the feed gas streams the same considerations as discussed for configurations according to FIG. 3 above apply. The overall mass balances illustrating gas composition and flow rate for the exemplary feed gas and products are shown in Table 1 above.

[0042] In general, the absorber receives an expanded feed gas and a reflux stream that is produced from the overhead vapor from a demethanizer after the overhead vapor is compressed and cooled by the second level refrigeration. In such configurations, the demethanizer column is fluidly coupled to the absorber and receives a column feed stream and operates at a pressure that is at least 50 to 100 psi lower, more preferably 100 psi to 300 psi lower than the operating pressure of the absorber. Therefore, most typically, the feed gas has a pressure of between 900 psig and 1600 psig, is expanded in a turbo-expander, and is fed to an absorber. The bottom product of the absorber is expanded to a pressure in the range of 50 psi to 350 psi differential pressure (relative to the demethanizer) and thereby chilled by Joule-Thomson effect to −90° F. to −130° F. It is also contemplated that the cooled and expanded absorber bottom stream forms a rectification stream that is fed to the demethanizer for C2 recovery. The demethanizer is reboiled with heat content from the feed gas and an optional external heat source, controlling the methane content in the bottoms product at about 1.5 mol % (or as otherwise needed to meet desired product specifications).

[0043] More particularly, as depicted in FIG. 4, feed gas stream 1 enters the plant at about 1200 psig and 120° F., and is cooled in exchanger 51 to typically 10° F. to −30° F., forming stream 2, using multiple cooling streams including liquid stream 5 from separator 52, side reboiler stream 22 from demethanizer 61, flash vapor 70 from LNG storage tank 69, and a propane refrigerant stream 32 of refrigeration system 101. Propane refrigerant is generated from heated stream 33 with a cascade propane refrigeration system, vaporizing at least three different pressure levels.

[0044] The chilled feed gas stream 2 is separated in separator 52, forming a gaseous portion 3 and a liquid portion 4. The liquid portion 4 is letdown in pressure in JT valve 53 forming stream 5, and optionally heated to stream 6 with the heat content from the feed gas prior to entering the demethanizer 61. The gaseous portion 3 from separator 52 is split into two portions. One portion (stream 7) is routed to the exchanger 54 to provide reflux to the absorber, and the other portion (stream 8) is expanded in turbo-expander 64 to produce a chilled vapor stream 10, typically at −80° F. to −100° F. and to generate power to drive the residue gas compressor 65. The chilled vapor 10 is fed to absorber 58, which operates at a pressure well above 450 psig, typically at between 500 psig to 700 psig, and most typically at 600 psig. The flow ratio of vapor stream 8 to vapor stream 3 can be variably adjusted to achieve a specific C2 recovery level, and/or to meet desired C2 product rates. Table 2 below exemplarily illustrates the effect of flow ratio of vapor stream 8 to vapor stream 3 on C2 and C3 recovery.

**TABLE 2**

<table>
<thead>
<tr>
<th>FLOW RATIO (STREAM 8 TO STREAM 3)</th>
<th>C3 RECOVERY</th>
<th>C2 RECOVERY</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.7</td>
<td>98%</td>
<td>85%</td>
</tr>
<tr>
<td>0.8</td>
<td>98%</td>
<td>62%</td>
</tr>
<tr>
<td>0.9</td>
<td>99%</td>
<td>31%</td>
</tr>
<tr>
<td>1.0</td>
<td>99%</td>
<td>25%</td>
</tr>
</tbody>
</table>

[0045] Absorber 58 is refluxed with two cold streams, wherein the first reflux (top reflux) is supplied by stream 27 (via 56 and 11) from the demethanizer 61, wherein the second reflux is supplied by stream 12 (via 9 and 55) from exchanger 54. The reflux streams are chilled to about −125° F. to −155° F. with mixed refrigerant stream 74 that is generated by the compressed mixed refrigerant from refrigeration unit 102 that is chilled in exchanger 54 and chilled by JT valve 91.

[0046] The absorber produces an overhead vapor stream 28 at about −120° F. to −140° F. and a bottoms stream 14 at about −100° F. to −110° F. The overhead vapor is compressed by the residue gas compressor 65 using the power produced from expander 64 forming a discharge stream 29, typically at about 900 psig and −70° F. to −80° F. It should be especially appreciated that compression of a cryogenic vapor is thermodynamically more efficient resulting in a high compression ratio across the compressor, which reduces the refrigeration consumption for liquefaction. The residual gas is chilled and condensed in exchanger 67 to about −255° F. to −265° F. using mixed refrigerant 79 operating at −180° F. to −270° F. that is produced by the mixed refrigeration system 103, after the compressed stream 76 is chilled in exchangers 54, 67, and JT’d via valve 92. The liquefied residual gas is further letdown in pressure to stream 82 at about 16.0 psig via JT valve 90, and the flashed liquid is stored in LNG storage tank 69. LNG product is withdrawn as stream 30 and withdrawn to storage or transport. In some cases, depending on the natural gas composition and the temperature from the liquefier exchanger, a significant quantity of light gas 70 is evolved which can be recovered as fuel gas after its refrigerant content is recovered. Where desired, a portion of the ethane product stream 15 can be directed from deethanizer 59 to LNG storage or transport. In this way, lean LNG can be converted to a heavier and richer LNG.

[0047] Absorber bottoms product stream 14 is preferably expanded in JT valve (or other expansion device) 60 to a
pressure that is about 50-350 psig less than absorber pressure and enters as cooled stream 20 to the demethanizer at a temperature of between about -90° F. to -130° F. The demethanizer is reboiled using reboiler 63 and produces bottom product 25, which is then fed to deethanizer 59. Demethanizer overhead product 24 is then routed back to the absorber as reflux stream 11. To that end, the overhead product 24 is re-compressed to form stream 26 (to a pressure above absorber pressure) by compressor 66 and chilled in exchanger 54 to form stream 27, which is expanded to reflux stream 11. Alternatively (not shown in FIG. 4, see FIG. 2B), and especially where C2 is rejected, the absorber bottoms product is JT expanded heated against feed gas stream 1. The so heated stream is further heated in the demethanizer overhead condenser and then fed into the demethanizer as feed stream.

[0048] Deethanizer 59 is configured as reboiled column using reboiler 34 to separate C2 from C3+ components, wherein the C3+ components are drawn from the column as stream 23. The C2 overhead product is condensed in overhead condenser 62 and separated in drum 68. One portion 18 of the C2 product is pumped back by pump 59 to the column as reflux while another portion 19 is withdrawn for LNG blending or storage/transport via stream 17. With respect to the remaining components and process conditions, the same considerations apply for like components as described in FIG. 3 above.

[0049] Therefore, and viewed from a different perspective, an absorber in contemplated plants receives a liquid portion of the natural gas feed and a second vapor portion of the natural gas feed, wherein the second portion is reduced in pressure via a turbo expander. Preferred absorbers produce a bottom product that is expanded, chilled, and fed to the demethanizer for absorption of the C2 + components. Preferred demethanizer bottoms products are subsequently fractionated in a deethanizer into a C2 liquid overhead and a C3+ bottoms product. In a still further contemplated aspect, the absorber produces an overhead vapor product that is predominantly methane at cryogenic temperature (~100°F or lower), which is further compressed using power generated by turbo-expansion of the feed gas. Such configurations produce a high pressure cryogenic vapor at ~75°F to ~100°F and 800 psig to 900 psig or higher pressure that is subsequently liquefied forming the LNG using the third temperature level refrigeration.

[0050] While configurations according to FIG. 4 are generally preferred, it should be noted that numerous alternative cooling methods and configurations for the first, second, and/or third cooling stages are also deemed suitable herein. For example, FIG. 5 exemplarily illustrates a plant configuration in which the third temperature range refrigerant at ~180°F to ~270°F is supplied by a cascade methane refrigerant cycle 103, operating with at least three pressure levels. Alternatively, and depending on the residue gas composition and pressure, a pure component refrigerant such as methane may also be appropriate. FIG. 6 illustrates another embodiment in which a propane pre-cooled cascade cycle 104 is added to the discharge of the mixed refrigeration system 102. Such alternative refrigeration system is especially suitable when a very high ethane recovery is required or when the feed gas contains a very high ethane and propane content. FIG. 7 illustrates yet another alternative embodiment in which a cascade propane refrigerant, a cascade ethylene refrigerant, and a methane refrigerant are employed for NGL recovery and LNG liquefaction.

[0051] Thus, the absorber in contemplated plants and methods is configured to separately receive a first and a second portion of a feed gas vapor and a demethanizer overhead, wherein the first portion of the feed gas vapor and the demethanizer column overhead provide reflux to the absorber. In such configurations, a flow control unit adjusts at least one of the first and second portions of the feed gas vapor to produce the desirable recovery levels of ethane, from 10% to 85% of the feed gas, while maintaining a high C3 (98% or above) recovery. It is further contemplated that at least a portion of the demethanizer bottoms product is fed to the deethanizer that fractionate the demethanizer bottoms product into an ethane overhead and a C3+ bottoms product. Thus, preferred configuration can provide variable C2 production rates by blending at least a portion of the excess overhead C2 liquid with the LNG. It should be especially recognized that this blending step simplifies the NGL recovery plant operation and allows the same process conditions (temperatures and pressures) to be maintained, regardless of the net C2 production rates.

[0052] In preferred plants, at least three temperature ranges are provided by one or more vaporizing refrigeration cycles: A first temperature range of ~10°F to ~35°F for refrigeration for the feed gas pre-cooling, a second temperature range of ~60°F to ~160°F for the first column reflux, and a third temperature range of ~180°F to ~270°F for gas liquefaction. It is generally preferred that the refrigerant in contemplated refrigeration circuits comprise one, two, or more hydrocarbon components and may further include nitrogen, halocarbons, and/or other refrigerants. Contemplated refrigeration cycles may also include combinations of refrigeration cycles, and especially combinations of a multi-component mixed refrigerant cycles, a single component cascade cycle, a gas expander cycle, and a propane pre-cooled refrigeration cycle. For example, it is contemplated that the first temperature range refrigeration at ~10°F to ~35°F uses propane pre-cooled refrigeration or cascade refrigeration, and cools at least one portion of the feed gas and the refrigerant of the second temperature level. The second temperature level refrigeration at ~60°F to ~160°F may then use a mixed refrigerant cycle or cascade refrigeration using pure component such as ethylene to chill the absorber reflux, and the third temperature level refrigeration at ~180°F to ~270°F may use a mixed refrigerant cycle or cascade refrigeration using pure component such as methane to liquefy the residual gas. Other preferred refrigeration cycles include letdown devices such as turbo-expanders and Joule-Thomson valves. With respect to the temperature levels, (combination of) refrigeration cycles, and cooling media, it should be noted that they may be adjusted as needed to achieve the lowest energy consumption in the cooling and liquefaction processes.

[0053] With respect to remaining components and process conditions in FIGS. 5-7, the same considerations apply for like components as described in FIG. 3 above. It should be further appreciated that all components of contemplated configurations (e.g., exchangers, pumps, valves, compressors, expanders, reflux absorbers, deethanizers, etc.) are commercially available and suitable for use in conjunction with the teachings presented herein. It is further generally contemplated that configurations according
to the inventive subject matter may find wide applicability in gas plant applications where high propane and ethane recovery are desirable, and feed gas is available at pressure greater than 800 psig. Moreover, such configurations produce a high pressure cryogenic methane rich vapor that will advantageously reduce equipment and operating costs when integrated to a LNG liquefaction plant. Table 3 below illustrates the temperatures and pressures of the residue gas from the NGL recovery plant and the energy savings of contemplated integrated plants versus heretofore known standalone plants on the basis of 70 mol % ethane recovery. The energy savings of contemplated plant configurations are about 10% as compared to known plants, which may be used to produce an equivalent amount of additional LNG.

<table>
<thead>
<tr>
<th>TABLE 3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Standalone NGL Recovery/LNG Plants</td>
</tr>
<tr>
<td>Residue Gas Temperature, °F</td>
</tr>
<tr>
<td>Residue Gas Pressure, psig</td>
</tr>
<tr>
<td>Total Power Consumption of NGL Recovery and Liquefaction, MW</td>
</tr>
</tbody>
</table>

[0054] Thus, specific embodiments and applications for integrated NGL recovery and LNG liquefaction have been disclosed. It should be apparent, however, to those skilled in the art that many modifications besides those already described are possible without departing from the inventive concepts herein. The inventive subject matter, therefore, is not to be restricted except in the spirit of the appended claims. Moreover, in interpreting both the specification and the claims, all terms should be interpreted in the broadest possible manner consistent with the context. In particular, the terms "comprises" and "comprising" should be interpreted as referring to elements, components, or steps in a non-exclusive manner, indicating that the referenced elements, components, or steps may be present, or utilized, or combined with other elements, components, or steps that are not expressly referenced. Furthermore, where a definition or use of a term in a reference, which is incorporated by reference herein is inconsistent or contrary to the definition of that term provided herein, the definition of that term provided herein applies and the definition of that term in the reference does not apply.

What is claimed is:

1. A gas processing plant, comprising:
   a separator that is configured to receive a partially expanded vapor portion of a natural gas feed stream, and that is further configured to produce a cold overhead product stream;
   an expander operationally coupled to drive a compressor, wherein the expander is configured to produce the partially expanded vapor portion and wherein the compressor is configured to produce a compressed cold overhead product stream from the cold overhead product stream; and
   wherein the separator and the compressor are fluidly coupled to each other such that the compressed cold overhead product stream has a pressure of at least 700 psig at a temperature of no warmer than -50 °F.

2. The plant of claim 1 wherein the separator is configured to receive another expanded vapor portion of the natural gas feed stream at a separate location.

3. The plant of claim 1 wherein the separator is configured to operate as a demethanizer.

4. The plant of claim 3 further comprising a deethanizer that is configured to produce a C3+ product and a C2 product.

5. The plant of claim 4 further comprising a conduit that provides at least part of the C2 product to the cold overhead product stream.

6. The plant of claim 1 wherein the separator is configured to operate as a refluxed absorber.

7. The plant of claim 6 wherein the demethanizer is configured to provide a reflux stream to the absorber.

8. The plant of claim 6 wherein the demethanizer is configured to operate at a lower pressure than the absorber.

9. The plant of claim 6 further comprising at least one of a conduit that provides a cooled absorber bottom product the absorber for C2 recovery and a conduit that provides a heated absorber bottom product the absorber for C2 rejection.

10. A method of producing a liquefied natural gas, comprising:

   producing a cold overhead product stream in a separator and compressing the product stream in a compressor without prior substantial heating of the cold product stream;
   wherein the compressor is driven by an expander that expands a vapor portion of a natural gas feed, and wherein the expanded vapor portion is fed into the separator; and
   liquefying the cold compressed product stream in a liquefaction unit.

11. The method of claim 10 wherein the separator is operated as a demethanizer that further receives another portion of a natural gas feed.

12. The method of claim 10 further comprising a deethanizer that receives a bottom product of the separator and that optionally provides a C2 product to the cold overhead product stream.

13. The method of claim 10 wherein the separator is operated as a refluxed absorber.

14. The method of claim 11 wherein a demethanizer provides a reflux stream to the absorber, wherein the absorber provides a bottom product to the demethanizer, and wherein the demethanizer is operated at a lower pressure than the absorber.

15. The method of claim 14 wherein the absorber bottom product is heated for C2 rejection prior to entering the demethanizer or cooled for C2 recovery prior to entering the demethanizer.

16. The method of claim 14 wherein a deethanizer receives a bottom product of the demethanizer for C2 and C3+ recovery.

17. The method of claim 10 wherein the cold compressed product stream has a pressure of at least 700 psig and a temperature of no warmer than -50 °F.

18. A method of producing LNG comprising:

   separating in a separator a cold overhead product from a natural gas containing feed gas;
compressing the cold overhead product using expansion energy from the feed gas to form a cold compressed overhead product at a pressure and temperature suitable for liquefaction in a liquefaction unit;

wherein the cold compressed overhead product is formed at neutral or negative net compression energy requirement.

19. The method of claim 17 wherein the step of separating is performed in an absorber or a demethanizer.

20. The method of claim 17 wherein the cold compressed overhead product has a pressure of at least 700 psig and a temperature of no warmer than -50°F.