HYDROCARBON GAS PROCESSING

of which the following is a

SPECIFICATION

BACKGROUND OF THE INVENTION

[0001] This invention relates to a process for the separation of a gas containing hydrocarbons. The applicants claim the benefits under Title 35, United States Code, Section 119(e) of prior U.S. Provisional Application Numbers 60/848,299 which was filed on September 28, 2006 and 60/897,683 which was filed on January 25, 2007.

[0002] Ethylene, ethane, propylene, propane, and/or heavier hydrocarbons can be recovered from a variety of gases, such as natural gas, refinery gas, and synthetic gas streams obtained from other hydrocarbon materials such as coal, crude oil, naphtha, oil shale, tar sands, and lignite. Natural gas usually has a major proportion of methane and

ethane, i.e., methane and ethane together comprise at least 50 mole percent of the gas. The gas also contains relatively lesser amounts of heavier hydrocarbons such as propane, butanes, pentanes, and the like, as well as hydrogen, nitrogen, carbon dioxide, and other gases.

[0003] The present invention is generally concerned with the recovery of ethylene, ethane, propylene, propane, and heavier hydrocarbons from such gas streams. A typical analysis of a gas stream to be processed in accordance with this invention would be, in approximate mole percent, 90.5% methane, 4.1% ethane and other C_2 components, 1.3% propane and other C_3 components, 0.4% iso-butane, 0.3% normal butane, 0.5% pentanes plus, and 2.6% carbon dioxide, with the balance made up of nitrogen. Sulfur containing gases are also sometimes present.

[0004] The historically cyclic fluctuations in the prices of both natural gas and its natural gas liquid (NGL) constituents have at times reduced the incremental value of ethane, ethylene, propane, propylene, and heavier components as liquid products. This has resulted in a demand for processes that can provide more efficient recoveries of these products, for processes that can provide efficient recoveries with lower capital investment and lower operating costs, and for processes that can be easily adapted or adjusted to vary the recovery of a specific component over a broad range. Available processes for separating these materials include those based upon cooling and refrigeration of gas, oil absorption, and refrigerated oil absorption. Additionally, cryogenic processes have become popular because of the availability of economical equipment that produces power while simultaneously expanding and extracting heat from the gas being processed.

Depending upon the pressure of the gas source, the richness (ethane, ethylene, and heavier hydrocarbons content) of the gas, and the desired end products, each of these processes or a combination thereof may be employed.

[0005] The cryogenic expansion process is now generally preferred for natural gas liquids recovery because it provides maximum simplicity with ease of startup, operating flexibility, good efficiency, safety, and good reliability. U.S. Patent Nos. 3,292,380; 4,061,481; 4,140,504; 4,157,904; 4,171,964; 4,185,978; 4,251,249; 4,278,457; 4,519,824; 4,617,039; 4,687,499; 4,689,063; 4,690,702; 4,854,955; 4,869,740; 4,889,545; 5,275,005; 5,555,748; 5,568,737; 5,771,712; 5,799,507; 5,881,569; 5,890,378; 5,983,664; 6,182,469; 6,712,880; 6,915,662; 7,191,617; 7,219,513; reissue U.S. Patent No. 33,408; and co-pending application no. 11/430,412 describe relevant processes (although the description of the present invention in some cases is based on different processing conditions than those described in the cited patents and applications).

[0006] In a typical cryogenic expansion recovery process, a feed gas stream under pressure is cooled by heat exchange with other streams of the process and/or external sources of refrigeration such as a propane compression-refrigeration system. As the gas is cooled, liquids may be condensed and collected in one or more separators as high-pressure liquids containing some of the desired C_2+ or C_3+ components. Depending on the richness of the gas and the amount of liquids formed, the high-pressure liquids may be expanded to a lower pressure and fractionated. The vaporization occurring during expansion of the liquids results in further cooling of the stream. Under some conditions, pre-cooling the high pressure liquids prior to the expansion may be desirable

in order to further lower the temperature resulting from the expansion. The expanded stream, comprising a mixture of liquid and vapor, is fractionated in a distillation (demethanizer or deethanizer) column. In the column, the expansion cooled stream(s) is (are) distilled to separate residual methane, nitrogen, and other volatile gases as overhead vapor from the desired C_2 components, C_3 components, and heavier hydrocarbon components as bottom liquid product, or to separate residual methane, C_2 components, nitrogen, and other volatile gases as overhead vapor from the desired C_3 components and heavier hydrocarbon components as bottom liquid product.

[0007] If the feed gas is not totally condensed (typically it is not), a portion of the vapor remaining from the partial condensation can be passed through a work expansion machine or engine, or an expansion valve, to a lower pressure at which additional liquids are condensed as a result of further cooling of the stream. The pressure after expansion is essentially the same as the pressure at which the distillation column is operated. The combined vapor-liquid phases resulting from the expansion are supplied as feed to the column.

[0008] The remaining portion of the vapor is cooled to substantial condensation by heat exchange with other process streams, e.g., the cold fractionation tower overhead. Some or all of the high-pressure liquid may be combined with this vapor portion prior to cooling. The resulting cooled stream is then expanded through an appropriate expansion device, such as an expansion valve, to the pressure at which the demethanizer is operated. During expansion, a portion of the liquid will vaporize, resulting in cooling of the total stream. The flash expanded stream is then supplied as top feed to the demethanizer.

Typically, the vapor portion of the expanded stream and the demethanizer overhead vapor combine in an upper separator section in the fractionation tower as residual methane product gas. Alternatively, the cooled and expanded stream may be supplied to a separator to provide vapor and liquid streams. The vapor is combined with the tower overhead and the liquid is supplied to the column as a top column feed.

[0009] In the ideal operation of such a separation process, the residue gas leaving the process will contain substantially all of the methane in the feed gas with essentially none of the heavier hydrocarbon components and the bottoms fraction leaving the demethanizer will contain substantially all of the heavier hydrocarbon components with essentially no methane or more volatile components. In practice, however, this ideal situation is not obtained for two main reasons. The first reason is that the conventional demethanizer is operated largely as a stripping column. The methane product of the process, therefore, typically comprises vapors leaving the top fractionation stage of the column, together with vapors not subjected to any rectification step. Considerable losses of C_3 and C_4+ components occur because the top liquid feed contains substantial quantities of these components and heavier hydrocarbon components, resulting in corresponding equilibrium quantities of C_3 components, C_4 components, and heavier hydrocarbon components in the vapors leaving the top fractionation stage of the demethanizer. The loss of these desirable components could be significantly reduced if the rising vapors could be brought into contact with a significant quantity of liquid (reflux) capable of absorbing the C_3 components, C_4 components, and heavier hydrocarbon components from the vapors.

[0010] The second reason that this ideal situation cannot be obtained is that carbon dioxide contained in the feed gas fractionates in the demethanizer and can build up to concentrations of as much as 5% to 10% or more in the tower even when the feed gas contains less than 1% carbon dioxide. At such high concentrations, formation of solid carbon dioxide can occur depending on temperatures, pressures, and the liquid solubility. It is well known that natural gas streams usually contain carbon dioxide, sometimes in substantial amounts. If the carbon dioxide concentration in the feed gas is high enough, it becomes impossible to process the feed gas as desired due to blockage of the process equipment with solid carbon dioxide (unless carbon dioxide removal equipment is added, which would increase capital cost substantially). The present invention provides a means for generating a liquid reflux stream that will improve the recovery efficiency for the desired products while simultaneously substantially mitigating the problem of carbon dioxide icing.

[0011] In recent years, the preferred processes for hydrocarbon separation use an upper absorber section to provide additional rectification of the rising vapors. The source of the reflux stream for the upper rectification section is typically a recycled stream of residue gas supplied under pressure. The recycled residue gas stream is usually cooled to substantial condensation by heat exchange with other process streams, e.g., the cold fractionation tower overhead. The resulting substantially condensed stream is then expanded through an appropriate expansion device, such as an expansion valve, to the pressure at which the demethanizer is operated. During expansion, a portion of the liquid will usually vaporize, resulting in cooling of the total stream. The flash expanded stream

is then supplied as top feed to the demethanizer. Typically, the vapor portion of the expanded stream and the demethanizer overhead vapor combine in an upper separator section in the fractionation tower as residual methane product gas. Alternatively, the cooled and expanded stream may be supplied to a separator to provide vapor and liquid streams, so that thereafter the vapor is combined with the tower overhead and the liquid is supplied to the column as a top column feed. Typical process schemes of this type are disclosed in U.S. Patent Nos. 4,889,545; 5,568,737; and 5,881,569, and in Mowrey, E. Ross, "Efficient, High Recovery of Liquids from Natural Gas Utilizing a High Pressure Absorber", Proceedings of the Eighty-First Annual Convention of the Gas Processors Association, Dallas, Texas, March 11-13, 2002. Unfortunately, these processes require the use of a large amount of compression power to provide the motive force for recycling the reflux stream to the demethanizer, adding to both the capital cost and the operating cost of facilities using these processes.

[0012] The present invention also employs an upper rectification section (or a separate rectification column in some embodiments). However, the reflux stream for this rectification section is provided by using a side draw of the vapors rising in a lower portion of the tower. By modestly elevating its pressure, a significant quantity of liquid can be condensed in this side draw stream, often using only the refrigeration available in the cold vapor leaving the upper rectification section. This condensed liquid, which is predominantly liquid methane, can then be used to absorb C_2 components, C_3 components, C_4 components, and heavier hydrocarbon components from the vapors

rising through the upper rectification section and thereby capture these valuable components in the bottom liquid product from the demethanizer.

[0013] Heretofore, such a side draw feature has been employed in C_2+ recovery systems, as illustrated in the assignee's U.S. Patent No. 7,191,617. Surprisingly, applicants have found that elevating the pressure of the side draw feature of the assignee's U.S. Patent No. 7,191,617 invention improves C_3+ recoveries without sacrificing C_2 component recovery levels and improves the system efficiency, while simultaneously substantially mitigating the problem of carbon dioxide icing.

[0014] In accordance with the present invention, it has been found that C_3 and C_4+ recoveries in excess of 99 percent can be obtained with no loss in C_2+ component recovery. The present invention provides the further advantage of being able to maintain in excess of 99 percent recovery of the C_3 and C_4+ components as the recovery of C_2 components is adjusted from high to low values. In addition, the present invention makes possible essentially 100 percent separation of methane and lighter components from the C_2 components and heavier components while maintaining the same recovery levels as the prior art and improving the safety factor with respect to the danger of carbon dioxide icing. The present invention, although applicable at lower pressures and warmer temperatures, is particularly advantageous when processing feed gases in the range of 400 to 1500 psia [2,758 to 10,342 kPa(a)] or higher under conditions requiring NGL recovery column overhead temperatures of -50°F [-46°C] or colder.

[0015] For a better understanding of the present invention, reference is made to the following examples and drawings. Referring to the drawings:

[0016] FIG. 1 is a flow diagram of a prior art natural gas processing plant in accordance with United States Patent No. 7,191,617;

[0017] FIG. 2 is a flow diagram of a natural gas processing plant in accordance with the present invention;

[0018] FIG. 3 is a concentration-temperature diagram for carbon dioxide showing the effect of the present invention;

[0019] FIG. 4 is a flow diagram illustrating an alternative means of application of the present invention to a natural gas stream;

[0020] FIG. 5 is a concentration-temperature diagram for carbon dioxide showing the effect of the present invention with respect to the process of FIG. 4;

[0021] FIGS. 6 through 9 are flow diagrams illustrating alternative means of application of the present invention to a natural gas stream; and

[0022] FIG. 10 is a partial flow diagram illustrating alternative means of accomplishing the splitting of the vapor feed in accordance with the present invention.

[0023] In the following explanation of the above figures, tables are provided summarizing flow rates calculated for representative process conditions. In the tables appearing herein, the values for flow rates (in moles per hour) have been rounded to the nearest whole number for convenience. The total stream rates shown in the tables include all non-hydrocarbon components and hence are generally larger than the sum of the stream flow rates for the hydrocarbon components. Temperatures indicated are approximate values rounded to the nearest degree. It should also be noted that the process design calculations performed for the purpose of comparing the processes

depicted in the figures are based on the assumption of no heat leak from (or to) the surroundings to (or from) the process. The quality of commercially available insulating materials makes this a very reasonable assumption and one that is typically made by those skilled in the art.

[0024] For convenience, process parameters are reported in both the traditional British units and in the units of the Système International d'Unités (SI). The molar flow rates given in the tables may be interpreted as either pound moles per hour or kilogram moles per hour. The energy consumptions reported as horsepower (HP) and/or thousand British Thermal Units per hour (MBTU/Hr) correspond to the stated molar flow rates in pound moles per hour. The energy consumptions reported as kilowatts (kW) correspond to the stated molar flow rates in kilogram moles per hour.

DESCRIPTION OF THE PRIOR ART

[0025] FIG. 1 is a process flow diagram showing the design of a processing plant to recover C_2+ components from natural gas using prior art according to assignee's U.S. Pat. No. 7,191,617. In this simulation of the process, inlet gas enters the plant at 120°F [49°C] and 1040 psia [7,171 kPa(a)] as stream **31**. If the inlet gas contains a concentration of sulfur compounds which would prevent the product streams from meeting specifications, the sulfur compounds are removed by appropriate pretreatment of the feed gas (not illustrated). In addition, the feed stream is usually dehydrated to prevent hydrate (ice) formation under cryogenic conditions. Solid desiccant has typically been used for this purpose.

[0026] The feed stream **31** is cooled in heat exchanger **10** by heat exchange with cool residue gas at -28°F [-33°C] (stream **48a**), demethanizer reboiler liquids at 35°F [2°C] (stream **41**), demethanizer lower side reboiler liquids at -10°F [-23°C] (stream **40**), and demethanizer upper side reboiler liquids at -79°F [-62°C] (stream **39**). The cooled stream **31a** enters separator **11** at -15°F [–26°C] and 1030 psia [7,102 kPa(a)] where the vapor (stream **32**) is separated from the condensed liquid (stream **33**). The separator liquid (stream **33**) is expanded to the operating pressure (approximately 432 psia [2,976 kPa(a)]) of fractionation tower **19** by expansion valve **12**, cooling stream **33a** to -39°F [-39°C] before it is supplied to fractionation tower **19** at a lower mid-column feed point.

[0027] The vapor (stream **32**) from separator **11** is divided into two streams, **35** and **36**. Stream **35**, containing about 36% of the total vapor, passes through heat exchanger **15** in heat exchange relation with the cold residue gas at -127°F [-88°C] (stream **48**) where it is cooled to substantial condensation. The resulting substantially condensed stream **35a** at -123°F [-86°C] is then flash expanded through expansion valve **16** to the operating pressure of fractionation tower **19**. During expansion a portion of the stream is vaporized, resulting in cooling of the total stream to -134°F [-92°C]. The expanded stream **35b** is supplied to fractionation tower **19** at an upper mid-column feed point.

[0028] The remaining 64% of the vapor from separator **11** (stream **36**) enters a work expansion machine **17** in which mechanical energy is extracted from this portion of the high pressure feed. The machine **17** expands the vapor substantially isentropically to

the tower operating pressure, with the work expansion cooling the expanded stream **36a** to a temperature of approximately -90°F [-68°C]. The typical commercially available expanders are capable of recovering on the order of 80-88% of the work theoretically available in an ideal isentropic expansion. The work recovered is often used to drive a centrifugal compressor (such as item **18**) that can be used to re-compress the residue gas (stream **48b**), for example. The partially condensed expanded stream **36a** is thereafter supplied as feed to fractionation tower **19** a second lower mid-column feed point.

[0029] The demethanizer in tower **19** is a conventional distillation column containing a plurality of vertically spaced trays, one or more packed beds, or some combination of trays and packing. The demethanizer tower consists of two sections: an upper absorbing (rectification) section **19a** that contains the trays and/or packing to provide the necessary contact between the vapor portion of the expanded streams **35b** and **36a** rising upward and cold liquid falling downward to condense and absorb the C_2 components, C_3 components, and heavier components; and a lower stripping (demethanizing) section **19b** that contains the trays and/or packing to provide the necessary contact between the liquids falling downward and the vapors rising upward. The stripping section **19b** also includes reboilers (such as trim reboiler **20** and the reboiler and side reboilers described previously) which heat and vaporize a portion of the liquids flowing down the column to provide the stripping vapors which flow up the column to strip the liquid product, stream **42**, of methane and lighter components. Stream **36a** enters demethanizer **19** at an intermediate feed position located in the lower region of absorbing section **19a** of demethanizer **19**. The liquid portion of the expanded stream

commingles with liquids falling downward from the absorbing section **19a** and the combined liquid continues downward into the stripping section **19b** of demethanizer **19**. The vapor portion of the expanded stream rises upward through absorbing section **19a** and is contacted with cold liquid falling downward to condense and absorb the C_2 components, C_3 components, and heavier components.

[0030] A portion of the distillation vapor (stream **43**) is withdrawn from the upper region of stripping section **19b**. This stream is then cooled from -112°F [-80°C] to -130°F [-90°C] and partially condensed (stream **43a**) in heat exchanger **22** by heat exchange with the cold demethanizer overhead stream **38** exiting the top of demethanizer **19** at -134°F [-92°C]. The cold demethanizer overhead stream is warmed slightly to -126°F [-88°C] (stream **38a**) as it cools and condenses at least a portion of stream **43**.

[0031] The operating pressure in reflux separator **23** (428 psia [2,951 kPa(a)]) is maintained slightly below the operating pressure of demethanizer **19**. This provides the driving force which causes distillation vapor stream **43** to flow through heat exchanger **22** and thence into the reflux separator **23** wherein the condensed liquid (stream **45**) is separated from the uncondensed vapor (stream **44**). Stream **44** then combines with the warmed demethanizer overhead stream **38a** from heat exchanger **22** to form cold residue gas stream **48** at -127°F [-88°C].

[0032] The liquid stream **45** from reflux separator **23** is pumped by pump **24** to a pressure slightly above the operating pressure of demethanizer **19**, and stream **45a** is then supplied as cold top column feed (reflux) to demethanizer **19**. This cold liquid reflux

absorbs and condenses the propane and heavier components rising in the upper rectification region of absorbing section **19a** of demethanizer **19**.

[0033] In stripping section **19b** of demethanizer **19**, the feed streams are stripped of their methane and lighter components. The resulting liquid product (stream **42**) exits the bottom of tower **19** at 52°F [11°C], based on a typical specification of a methane to ethane ratio of 0.025:1 on a molar basis in the bottom product. The distillation vapor stream forming the tower overhead (stream **38**) is warmed in heat exchanger **22** as it provides cooling to distillation stream **43** as described previously, then combines with stream **44** to form the cold residue gas stream **48**. The residue gas passes countercurrently to the incoming feed gas in heat exchanger **15** where it is heated to -28°F [-33°C] (stream **48a**), and in heat exchanger **10** where it is heated to 107°F [42°C] (stream **48b**) as it provides cooling as previously described. The residue gas is then re-compressed in two stages, compressor **18** driven by expansion machine **17** and compressor **27** driven by a supplemental power source. After stream **48d** is cooled to 120°F [49°C] in discharge cooler **28**, the residue gas product (stream **48e**) flows to the sales gas pipeline at 1040 psia [7,171 kPa(a)].

[0034] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 1 is set forth in the following table:

Table I

(FIG. 1)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	C. Dioxide	Total
31	25,382	1,161	362	332	743	28,055
32	25,241	1,131	336	220	733	27,736
33	141	30	26	112	10	319
35	9,087	407	121	79	264	9,985
36	16,154	724	215	141	469	17,751
43	3,598	96	5	1	113	3,816
44	2,963	33	0	0	59	3,058
45	635	63	5	1	54	758
38	22,395	164	5	0	262	22,897
48	25,358	197	5	0	321	25,955
42	24	964	357	332	422	2,100

Recoveries*

Ethane	83.05%
Propane	98.50%
Butanes+	99.94%

Power

Residue Gas Compression	12,464	HP	[20,490 kW]

* (Based on un-rounded flow rates)

DESCRIPTION OF THE INVENTION

Example 1

[0035] FIG. 2 illustrates a flow diagram of a process in accordance with the present invention. The feed gas composition and conditions considered in the process presented in FIG. 2 are the same as those in FIG. 1. Accordingly, the FIG. 2 process can be compared with that of the FIG. 1 process to illustrate the advantages of the present invention.

[0036] In the simulation of the FIG. 2 process, inlet gas enters the plant as stream **31** and is cooled in heat exchanger **10** by heat exchange with cool residue gas at -66°F [-54°C] (stream **38b**), demethanizer reboiler liquids at 48°F [9°C] (stream **41**), demethanizer lower side reboiler liquids at 5°F [-15°C] (stream **40**), and demethanizer upper side reboiler liquids at -70°F [-57°C] (stream **39**). The cooled stream **31a** enters separator **11** at -38°F [–39°C] and 1030 psia [7,102 kPa(a)] where the vapor (stream **32**) is separated from the condensed liquid (stream **33**). The separator liquid (stream **33**) may in some cases be divided into two streams, stream **47** and stream **37**. In this example of the present invention, all of the separator liquid in stream **33** is directed to stream **37** and is expanded to the operating pressure (approximately 470 psia [3,238 kPa(a)]) of fractionation tower **19** by expansion valve **12**, cooling stream **37a** to -68°F [-56°C] before it is supplied to fractionation tower **19** at a lower mid-column feed point. In other embodiments of the present invention, all of the separator liquid in stream **33** may be directed to stream **47**, or a portion of stream **33** may be directed to stream **37** with the remaining portion directed to stream **47**.

[0037] The vapor (stream **32**) from separator **11** is divided into two streams, **34** and **36**. Stream **34**, containing about 22% of the total vapor, may in some embodiments be combined with a portion (stream **47**) of separator liquid stream **33** to form combined stream **35**. Stream **34** or **35**, as the case may be, passes through heat exchanger **15** in heat exchange relation with the cold residue gas at -105°F [-76°C] (stream **38a**) where it is cooled to substantial condensation. The resulting substantially condensed stream **35a** at -101°F [-74°C] is then flash expanded through expansion valve **16** to the operating pressure of fractionation tower **19**. During expansion a portion of the stream is vaporized, resulting in cooling of the total stream. In the process illustrated in FIG. 2, the expanded stream **35b** leaving expansion valve **16** reaches a temperature of -128°F [-89°C] and is supplied to fractionation tower **19** at an upper mid-column feed point.

[0038] The remaining 78% of the vapor from separator **11** (stream **36**) enters a work expansion machine **17** in which mechanical energy is extracted from this portion of the high pressure feed. The machine **17** expands the vapor substantially isentropically to the tower operating pressure, with the work expansion cooling the expanded stream **36a** to a temperature of approximately -102°F [-74°C]. The partially condensed expanded stream **36a** is thereafter supplied as feed to fractionation tower **19** a second lower mid-column feed point.

[0039] The demethanizer in tower **19** is a conventional distillation column containing a plurality of vertically spaced trays, one or more packed beds, or some combination of trays and packing. The demethanizer tower consists of two sections: an upper absorbing (rectification) section **19a** that contains the trays and/or packing to

provide the necessary contact between the vapor portion of the expanded streams **35b** and **36a** rising upward and cold liquid falling downward to condense and absorb the C_2 components, C_3 components, and heavier components; and a lower stripping (demethanizing) section **19b** that contains the trays and/or packing to provide the necessary contact between the liquids falling downward and the vapors rising upward. The stripping section **19b** also includes reboilers (such as trim reboiler **20** and the reboiler and side reboilers described previously) which heat and vaporize a portion of the liquids flowing down the column to provide the stripping vapors which flow up the column to strip the liquid product, stream **42**, of methane and lighter components. Stream **36a** enters demethanizer **19** at an intermediate feed position located in the lower region of absorbing section **19a** of demethanizer **19**. The liquid portion of the expanded stream commingles with liquids falling downward from the absorbing section **19a** and the combined liquid continues downward into the stripping section **19b** of demethanizer **19**. The vapor portion of the expanded stream rises upward through absorbing section **19a** and is contacted with cold liquid falling downward to condense and absorb the C_2 components, C_3 components, and heavier components.

[0040] A portion of the distillation vapor (stream **43**) is withdrawn from the upper region of stripping section **19b** at -108°F [-78°C] below expanded stream **36a** and is compressed to approximately 609 psia [4,199 kPa(a)] by vapor compressor **21**. The compressed stream **43a** is then cooled from -78°F [-61°C] to -125°F [-87°C] and substantially condensed (stream **43b**) in heat exchanger **22** by heat exchange with the cold demethanizer overhead stream **38** exiting the top of demethanizer **19** at -129°F

[-89°C]. The cold demethanizer overhead stream is warmed to -105°F [-76°C] (stream **38a**) as it cools and condenses stream **43a**.

[0041] Since substantially condensed stream **43b** is at a pressure greater than the operating pressure of demethanizer **19**, it is flash expanded through expansion valve **25** to the operating pressure of fractionation tower **19**. During expansion a small portion of the stream is vaporized, resulting in cooling of the total stream to -132°F [-91°C]. The expanded stream **43c** is then supplied as cold top column feed (reflux) to demethanizer **19**. The vapor portion (if any) of stream **43c** combines with the distillation vapor rising from the upper fractionation stage to form residue gas stream **38**, while the cold liquid reflux portion absorbs and condenses the C_2 components, C_3 components, and heavier components rising in the upper rectification region of absorbing section **19a** of demethanizer **19**.

[0042] In stripping section **19b** of demethanizer **19**, the feed streams are stripped of their methane and lighter components. The resulting liquid product (stream **42**) exits the bottom of tower **19** at 66°F [19°C]. The distillation vapor stream forming cold residue gas stream **38** is warmed in heat exchanger **22** as it provides cooling to compressed distillation stream **43a** as described previously. The residue gas (stream **38a**) passes countercurrently to the incoming feed gas in heat exchanger **15** where it is heated to -66°F [-54°C] (stream **38b**), and in heat exchanger **10** where it is heated to 110°F [43°C] (stream **38c**) as it provides cooling as previously described. The residue gas is then re-compressed in two stages, compressor **18** driven by expansion machine **17** and compressor **27** driven by a supplemental power source. After stream **38e** is cooled to

120°F [49°C] in discharge cooler **28**, the residue gas product (stream **38f**) flows to the sales gas pipeline at 1040 psia [7,171 kPa(a)].

[0043] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 2 is set forth in the following table:

Table II

(FIG. 2)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	C. Dioxide	Total
31	25,382	1,161	362	332	743	28,055
32	25,050	1,096	310	180	720	27,431
33	332	65	52	152	23	624
34/35	5,473	239	68	39	157	5,994
36	19,577	857	242	141	563	21,437
43	3,936	114	7	1	109	4,171
38	25,358	197	2	0	403	26,034
42	24	964	360	332	340	2,021

Recoveries*					
Ethane	83.06%				
Propane	99.33%				
Butanes+	99.97%				
Power					
Residue Gas Compression	11,111	HP	[	18,266	kW]
Vapor Compression	278	HP	[	457	kW]
Total Compression	11,389	HP	[	18,723	kW]

* (Based on un-rounded flow rates)

[0044] A comparison of Tables I and II shows that, compared to the prior art, the present invention maintains essentially the same ethane recovery (83.05% versus 83.06%), but improves both the propane recovery (99.33% versus 98.50%) and butanes+ recovery (99.97% versus 99.94%). Comparison of Tables I and II further shows that these increased yields were achieved using less horsepower than the prior art (11,389 HP versus 12,464 HP, or more than 8% less).

[0045] There are three primary factors that account for the improved efficiency of the present invention. First, the boost in pressure provided by vapor compressor **21** allows the column overhead (stream **38**) to condense all of distillation vapor stream **43**, unlike the prior art process which can condense only a fraction of the stream. As a result, the top reflux stream (stream **43c**) for the present invention is more than 5 times greater than that of the prior art (stream **45a**), providing much more efficient rectification in the upper region of absorbing section **19a**. Second, with the increase in the quantity of the

top reflux stream possible with the present invention, the quantity of secondary reflux stream **35b** can be correspondingly less without reducing the product yields. This in turn results in more flow (stream **36**) to expansion machine **17** and the resultant increase in the energy recovered to power compressor **18**, thereby reducing the power requirements of compressor **27**. Third, the more efficient rectification provided by stream **43c** in the upper region of absorbing section **19a** allows operating demethanizer **19** at a higher pressure without reducing the product yields, further reducing the power requirements of compressor **27**.

[0046] A further advantage of the present invention is a reduced likelihood of carbon dioxide icing. FIG. 3 is a graph of the relation between carbon dioxide concentration and temperature. Line **71** represents the equilibrium conditions for solid and liquid carbon dioxide in methane. (The liquid-solid equilibrium line in this graph is based on the data given in FIG. 16-33 on page 16-24 of the Engineering Data Book, Twelfth Edition, published in 2004 by the Gas Processors Suppliers Association, which is often used as a reference when checking for potential icing conditions.) A liquid temperature on or to the right of line **71**, or a carbon dioxide concentration on or above this line, signifies an icing condition. Because of the variations which normally occur in gas processing facilities (e.g., feed gas composition, conditions, and flow rate), it is usually desired to design a demethanizer with a considerable safety factor between the expected operating conditions and the icing conditions. (Experience has shown that the conditions of the liquids on the fractionation stages of a demethanizer, rather than the conditions of the vapors, typically govern the allowable operating conditions in most

demethanizers. For this reason, the corresponding vapor-solid equilibrium line is not shown in FIG. 3.)

[0047] Also plotted in FIG. 3 is a line representing the conditions for the liquids on the fractionation stages of demethanizer **19** in the prior art FIG. 1 process (line **72**). As can be seen, a portion of this operating line lies above the liquid-solid equilibrium line, indicating that the prior art FIG. 1 process cannot be operated at these conditions without encountering carbon dioxide icing problems. As a result, it is not possible to use the FIG. 1 process under these conditions, so the prior art FIG. 1 process cannot actually achieve the recovery efficiencies stated in Table I in practice without removal of at least some of the carbon dioxide from the feed gas. This would, of course, substantially increase capital cost.

[0048] Line **73** in FIG. 3 represents the conditions for the liquids on the fractionation stages of demethanizer **19** in the present invention as depicted in FIG. 2. In contrast to the prior art FIG. 1 process, there is a minimum safety factor of 1.2 between the carbon dioxide concentration in the column liquids for the anticipated operating conditions of the FIG. 2 process versus the concentrations at the liquid-solid equilibrium line. That is, it would require a 20 percent increase in the carbon dioxide content of the liquids to cause icing. Thus, the present invention could tolerate a 20% higher concentration of carbon dioxide in its feed gas than the prior art FIG. 1 process could tolerate without risk of crossing the liquid-solid equilibrium line. Further, whereas the prior art FIG. 1 process cannot be operated to achieve the recovery levels given in Table I

because of icing, the present invention could in fact be operated at even higher recovery levels than those given in Table II without risk of icing.

[0049] The shift in the operating conditions of the FIG. 2 demethanizer as indicated by line **73** in FIG. 3 can be understood by comparing the distinguishing features of the present invention to the prior art process of FIG. 1. While the shape of the operating line for the prior art FIG. 1 process (line **72**) is similar to the shape of the operating line for the present invention (line **73**), there is a key difference. The operating temperatures of the critical upper fractionation stages in the demethanizer in the FIG. 2 process are warmer than those of the corresponding fractionation stages in the demethanizer in the prior art FIG. 1 process, effectively shifting the operating line of the FIG. 2 process away from the liquid-solid equilibrium line. The warmer temperatures of the fractionation stages in the FIG. 2 demethanizer are mainly the result of operating the tower at higher pressure than the prior art FIG. 1 process. However, the higher tower pressure does not cause a loss in C_2+ component recovery levels because the distillation vapor stream **43** in the FIG. 2 process is in essence an open direct-contact compression-refrigeration cycle for the demethanizer using a portion of the inter-column vapor as the working fluid, supplying refrigeration to the process needed to overcome the loss in recovery that normally accompanies an increase in demethanizer operating pressure.

[0050] Another advantage of the present invention is a reduction in the amount of carbon dioxide leaving demethanizer **19** in liquid product stream **42**. Comparing stream **42** in Table I for the prior art FIG. 1 process to stream **42** in Table II for the FIG. 2

embodiment of the present invention reveals that there is nearly a 20% reduction in the quantity of carbon dioxide captured in stream **42** with the present invention. This generally reduces the product treating requirements by a corresponding amount, reducing both the capital cost and the operating cost of the treating system.

[0051] One of the inherent features in the operation of a demethanizer column to recover C_2 components is that the column must fractionate between the methane that is to leave the tower in its overhead product (vapor stream **38**) and the C_2 components that are to leave the tower in its bottom product (liquid stream **42**). However, the relative volatility of carbon dioxide lies between that of methane and C_2 components, causing the carbon dioxide to appear in both terminal streams. Further, carbon dioxide and ethane form an azeotrope, resulting in a tendency for carbon dioxide to accumulate in the intermediate fractionation stages of the column and thereby cause large concentrations of carbon dioxide to develop in the tower liquids.

[0052] The reflux streams for absorbing section **19a** in demethanizer **19** of the prior art FIG. 1 process are streams **45a** and **35b**, while those for the present invention shown in the FIG. 2 process are streams **43c** and **35b**. Comparing these streams in Table I and Table II, note that the total amounts of C_2 components and carbon dioxide in the reflux streams in the prior art FIG. 1 process are 470 and 318 Lb. Moles/Hr [470 and 318 kg moles/Hr], respectively, versus 353 and 266 Lb. Moles/Hr [353 and 266 kg moles/Hr], respectively, for the reflux streams in the FIG. 2 process of the present invention. Thus, significantly less of the azeotrope forming components enter absorbing section **19a** in the cold liquid reflux streams, entering instead into the warmer, lower

region of absorbing section **19a** with stream **36a** so that there is less accumulation of carbon dioxide in the fractionation stages of absorbing section **19a**. This allows more of the carbon dioxide to escape in overhead stream **38** instead of being captured in liquid product stream **42**.

Example 2

[0053] An alternative embodiment of the present invention is shown in FIG. 4. The feed gas composition and conditions considered in the process presented in FIG. 4 are the same as those in FIGS. 1 and 2. Accordingly, FIG. 4 can be compared with the prior art FIG. 1 process to illustrate the advantages of the present invention, and can likewise be compared to the embodiment displayed in FIG. 2.

[0054] In the simulation of the FIG. 4 process, inlet gas enters the plant as stream **31** and is cooled in heat exchanger **10** by heat exchange with cool residue gas at -66°F [-55°C] (stream **38b**), demethanizer reboiler liquids at 51°F [11°C] (stream **41**), demethanizer lower side reboiler liquids at 10°F [-12°C] (stream **40**), and demethanizer upper side reboiler liquids at -65°F [-54°C] (stream **39**). The cooled stream **31a** enters separator **11** at -38°F [–39°C] and 1030 psia [7,102 kPa(a)] where the vapor (stream **32**) is separated from the condensed liquid (stream **33**). The separator liquid (stream **33**) may in some cases be divided into two streams, stream **47** and stream **37**. In this example of the present invention, all of the separator liquid in stream **33** is directed to stream **37** and is expanded to the operating pressure (approximately 480 psia [3,309 kPa(a)]) of fractionation tower **19** by expansion valve **12**, cooling stream **37a** to -67°F [-55°C] before it is supplied to fractionation tower **19** at a lower mid-column feed point. In other

embodiments of the present invention, all of the separator liquid in stream **33** may be directed to stream **47**, or a portion of stream **33** may be directed to stream **37** with the remaining portion directed to stream **47**.

[0055] The vapor (stream **32**) from separator **11** is divided into two streams, **34** and **36**. Stream **34**, containing about 23% of the total vapor, may in some embodiments be combined with a portion (stream **47**) of separator liquid stream **33** to form combined stream **35**. Stream **34** or **35**, as the case may be, passes through heat exchanger **15** in heat exchange relation with the cold residue gas at -106°F [-77°C] (stream **38a**) where it is cooled to substantial condensation. The resulting substantially condensed stream **35a** at -102°F [-74°C] is then flash expanded through expansion valve **16** to the operating pressure of fractionation tower **19**. During expansion a portion of the stream is vaporized, resulting in cooling of the total stream. In the process illustrated in FIG. 4, the expanded stream **35b** leaving expansion valve **16** reaches a temperature of -127°F [-88°C] and is supplied to fractionation tower **19** at an upper mid-column feed point.

[0056] The remaining 77% of the vapor from separator **11** (stream **36**) enters a work expansion machine **17** in which mechanical energy is extracted from this portion of the high pressure feed. The machine **17** expands the vapor substantially isentropically to the tower operating pressure, with the work expansion cooling the expanded stream **36a** to a temperature of approximately -101°F [-74°C]. The partially condensed expanded stream **36a** is thereafter supplied as feed to fractionation tower **19** a second lower mid-column feed point.

[0057] A portion of the distillation vapor (stream **43**) is withdrawn from the lower region of absorbing section **19a** of demethanizer **19** at -113°F [-81°C] above expanded stream **36a** and is compressed to approximately 619 psia [4,266 kPa(a)] by vapor compressor **21**. The compressed stream **43a** is then cooled from -84°F [-65°C] to -124°F [-87°C] and substantially condensed (stream **43b**) in heat exchanger **22** by heat exchange with the cold demethanizer overhead stream **38** exiting the top of demethanizer **19** at -128°F [-89°C]. The cold demethanizer overhead stream is warmed to -106°F [-77°C] (stream **38a**) as it cools and condenses stream **43a**.

[0058] Since substantially condensed stream **43b** is at a pressure greater than the operating pressure of demethanizer **19**, it is flash expanded through expansion valve **25** to the operating pressure of fractionation tower **19**. During expansion a small portion of the stream is vaporized, resulting in cooling of the total stream to -131°F [-91°C]. The expanded stream **43c** is then supplied as cold top column feed (reflux) to demethanizer **19**. The vapor portion (if any) of stream **43c** combines with the distillation vapor rising from the upper fractionation stage to form residue gas stream **38**, while the cold liquid reflux portion absorbs and condenses the C_2 components, C_3 components, and heavier components rising in the upper rectification region of absorbing section **19a** of demethanizer **19**.

[0059] In stripping section **19b** of demethanizer **19**, the feed streams are stripped of their methane and lighter components. The resulting liquid product (stream **42**) exits the bottom of tower **19** at 70°F [21°C]. The distillation vapor stream forming cold residue gas stream **38** is warmed in heat exchanger **22** as it provides cooling to

compressed distillation stream **43a** as described previously. The residue gas (stream **38a**) passes countercurrently to the incoming feed gas in heat exchanger **15** where it is heated to -66°F [-55°C] (stream **38b**), and in heat exchanger **10** where it is heated to 110°F [43°C] (stream **38c**) as it provides cooling as previously described. The residue gas is then re-compressed in two stages, compressor **18** driven by expansion machine **17** and compressor **27** driven by a supplemental power source. After stream **38e** is cooled to 120°F [49°C] in discharge cooler **28**, the residue gas product (stream **38f**) flows to the sales gas pipeline at 1040 psia [7,171 kPa(a)].

[0060] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 4 is set forth in the following table:

Table III

(FIG. 4)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	C. Dioxide	Total
31	25,382	1,161	362	332	743	28,055
32	25,050	1,096	310	180	720	27,431
33	332	65	52	152	23	624
34/35	5,636	247	70	40	162	6,172
36	19,414	849	240	140	558	21,259
43	3,962	100	3	0	125	4,200
38	25,358	197	2	0	425	26,055
42	24	964	360	332	318	2,000

Recoveries*					
Ethane	83.06%				
Propane	99.50%				
Butanes+	99.98%				
Power					
Residue Gas Compression	10,784	HP	[	17,728	kW]
Vapor Compression	260	HP	[	428	kW]
Total Compression	11,044	HP	[	18,156	kW]

* (Based on un-rounded flow rates)

[0061] A comparison of Tables II and III shows that, compared to the FIG. 2 embodiment of the present invention, the FIG. 4 embodiment maintains the same ethane recovery while improving the propane recovery (99.50% versus 99.33%) and butanes+ recovery (99.98% versus 99.97%) slightly. However, comparison of Tables II and III further shows that these yields were achieved using about 3% less horsepower than that required by the FIG. 2 embodiment of the present invention. The drop in the power requirements for the FIG. 4 embodiment is mainly due to the lower content of C_2+ components in top reflux stream **43c**, which provides more efficient rectification in the upper region of absorbing section **19a** so that demethanizer **19** can be operated at a slightly higher operating pressure (thereby reducing compression requirements) without reducing product yields. Comparing distillation vapor stream **43** in Table III for the FIG. 4 embodiment of the present invention to stream **43** in Table II for the FIG. 2 embodiment of the present invention, the concentrations of C_2 components and

particularly the C_3+ components in stream **43** of the FIG. 4 embodiment are significantly lower, so that higher product yields are achieved using less power than the FIG. 2 embodiment. The lower concentrations of C_2 components and C_3+ components in stream **43** of the FIG. 4 embodiment are the result of withdrawing the distillation vapor from the lower region of absorbing section **19a** rather than from the upper region of stripping section **19b** as in the FIG. 2 embodiment. The distillation vapor at the higher column location has been subjected to more rectification than the distillation vapor lower in the column, and so is closer to being the pure methane stream that would be the ideal reflux stream for the top of the column. In the prior art process of FIG. 1, the column overhead (stream **38**) could not condense a pure methane stream, but with the elevation in pressure provided by vapor compressor **21** of the present invention, column overhead stream **38** is cold enough to totally condense the distillation vapor stream **43** even though it is almost pure methane.

[0062] When the present invention is employed as in Example 2, the advantage with respect to avoiding carbon dioxide icing conditions is maintained compared to the FIG. 2 embodiment. FIG. 5 is another graph of the relation between carbon dioxide concentration and temperature, with line **71** as before representing the equilibrium conditions for solid and liquid carbon dioxide in methane and line **72** representing the conditions for the liquids on the fractionation stages of demethanizer **19** in the prior art process of FIG. 1. Line **74** in FIG. 5 represents the conditions for the liquids on the fractionation stages of demethanizer **19** in the present invention as depicted in FIG. 4, and shows a safety factor of 1.2 between the anticipated operating conditions and the icing

conditions for the FIG. 4 process. Thus, this embodiment of the present invention could also tolerate an increase of 20 percent in the concentration of carbon dioxide without risk of icing. In practice, this improvement in the icing safety factor could be used to advantage by operating the demethanizer at lower pressure (i.e., with colder temperatures on the fractionation stages) to raise the C_2+ component recovery levels without encountering icing problems. The shape of line **74** in FIG. 5 for the FIG. 4 embodiment is very similar to that of line **73** in FIG. 3 for the FIG. 2 embodiment. The primary difference is the significantly lower carbon dioxide concentrations of the liquids on the fractionation stages in the lower section of the FIG. 4 demethanizer due to withdrawing the distillation vapor stream at a higher location on the column in this embodiment. As can be seen by comparing stream **42** in Tables II and III, even less of the carbon dioxide in the feed gas is captured with the bottom liquid product in the FIG. 4 embodiment of the present invention, which generally means still less product treating will be required compared to the FIG. 2 embodiment of the present invention.

Other Embodiments

[0063] In accordance with this invention, it is generally advantageous to design the absorbing (rectification) section of the demethanizer to contain multiple theoretical separation stages. However, the benefits of the present invention can be achieved with as few as one theoretical stage, and it is believed that even the equivalent of a fractional theoretical stage may allow achieving these benefits. For instance, all or a part of the expanded substantially condensed distillation stream **43c** from expansion valve **25**, all or a part of the expanded substantially condensed stream **35b** from expansion valve **16**, and

all or a part of the expanded stream **36a** from work expansion machine **17** can be combined (such as in the piping joining the expansion valve to the demethanizer) and if thoroughly intermingled, the vapors and liquids will mix together and separate in accordance with the relative volatilities of the various components of the total combined streams. Such commingling of the three streams shall be considered for the purposes of this invention as constituting an absorbing section.

[0064] In some cases it may be advantageous to split the substantially condensed distillation stream **43b** into at least two streams as shown in FIGS. 6 through 9. This allows a portion (stream **51**) to be supplied above the location where vapor distillation stream **43** is withdrawn (and perhaps also above the feed location of expanded stream **36a**), either lower in the absorbing section of fractionation tower **19** (FIGS. 6 and 7) or lower on absorber column **19** (FIGS. 8 and 9), to increase the liquid flow in that part of the distillation system and improve the rectification of stream **43**. In such cases, expansion valve **26** is used to expand stream **51** to the column operating pressure (forming stream **51a**), while expansion valve **25** is used to expand the remaining portion (stream **50**) to the column operating pressure so that the resulting stream **50a** can then be supplied to the top of the absorbing section in demethanizer **19** (FIGS. 6 and 7) or to the top of absorber column **19** (FIGS. 8 and 9).

[0065] FIGS. 8 and 9 depict a fractionation tower constructed in two vessels, absorber (rectifier) column **19** (a contacting and separating device) and stripper column **29** (a distillation column). In FIG. 8, the overhead vapor (stream **46**) from stripper column **29** is split into two portions. One portion (stream **43**) is routed to compressor **21**

and thence to heat exchanger **22** to generate reflux for absorber column **19** as described earlier. The remaining portion (stream **49**) flows to the lower section of absorber column **19** to be contacted by expanded substantially condensed stream **35b** and the expanded substantially condensed distillation stream (either stream **50a**, or streams **50a** and **51a**). Pump **30** is used to route the liquids (stream **52**) from the bottom of absorber column **19** to the top of stripper column **29** so that the two towers effectively function as one distillation system. In FIG. 9, all of the overhead vapor (stream **46**) flows to the lower section of absorber column **19**, and distillation vapor stream **43** is withdrawn from a location higher in absorber column **19**, above the feed location of expanded stream **36a**. The decision whether to construct the fractionation tower as a single vessel (such as demethanizer **19** in FIGS. 2, 4, 6, and 7) or multiple vessels will depend on a number of factors such as plant size, the distance to fabrication facilities, etc.

[0066] Feed gas conditions, plant size, available equipment, or other factors may indicate that elimination of work expansion machine **17**, or replacement with an alternate expansion device (such as an expansion valve), is feasible. Although individual stream expansion is depicted in particular expansion devices, alternative expansion means may be employed where appropriate. For example, conditions may warrant work expansion of the substantially condensed portion of the feed stream (stream **35a**) and/or the substantially condensed distillation stream (stream **43b**).

[0067] As described in the earlier examples, distillation stream **43** is substantially condensed and the resulting condensate used to absorb valuable C_2 components, C_3 components, and heavier components from the vapors rising through the upper region of

absorbing section **19a** of demethanizer **19** (FIGS. 2, 4, 6, and 7) or absorber column **19** (FIGS. 8 and 9). However, the present invention is not limited to this embodiment. It may be advantageous, for instance, to treat only a portion of these vapors in this manner, or to use only a portion of the condensate as an absorbent, in cases where other design considerations indicate portions of the vapors or the condensate should bypass absorbing section **19a** of demethanizer **19** (FIGS. 2, 4, 6, and 7) or absorber column **19** (FIGS. 8 and 9). Some circumstances may favor partial condensation, rather than total condensation, of distillation stream **43a** in heat exchanger **22**. Other circumstances may favor that distillation stream **43** be a total vapor side draw from fractionation column **19** rather than a partial vapor side draw. It should also be noted that, depending on the composition of the feed gas stream, it may be advantageous to use external refrigeration to provide some portion of the cooling of distillation stream **43a** in heat exchanger **22**.

[0068] Under some circumstances, it may be advantageous to heat distillation stream **43** before it is compressed, as this may reduce the capital cost of compressor **21**. One means to accomplish this is to use compressed distillation stream **43a** (which is warmer due to the heat of compression) to supply this heating using a cross exchanger. In such cases, it may be possible to supplement the cooling of compressed distillation stream **43a** by the use of aerial cooling or other means, thereby reducing the cooling that must be supplied in heat exchanger **22** by overhead stream **38**. The potential reduction in the capital cost of compressor **21** must be weighed against the capital cost of the additional heating and cooling means for each application to determine whether this embodiment is advantageous.

[0069] In accordance with this invention, the splitting of the vapor feed may be accomplished in several ways. In some embodiments, vapor splitting may be effected in a separator. In the processes of FIGS. 2, 4, and 6 through 9, the splitting of the vapor occurs following cooling, and perhaps after separation of any liquids which may have been formed. The high pressure gas may be split, however, prior to any cooling of the inlet gas as shown in FIG. 10. Streams **35b**, **36a**, and **37a** in FIG. 10 may all be fed to a distillation column (such as demethanizer **19** in FIGS. 2, 4, 6, and 7), or streams **35b** and **36a** may be fed to a contacting and separating device and stream **37a** may be fed to a distillation column (such as absorber column **19** and stripper column **29**, respectively, in FIGS. 8 and 9). The cooling of stream **53** in heat exchanger **10** in FIG. 10 may be accomplished or supplemented by additional process streams (such as streams **39**, **40**, and **41** in FIGS. 2, 4, and 6 through 9) and/or external refrigeration.

[0070] When the inlet gas is leaner, separator **11** in FIGS. 2, 4, and 6 through 10 may not be needed. Depending on the quantity of heavier hydrocarbons in the feed gas and the feed gas pressure, the cooled feed stream **31a** leaving heat exchanger **10** in FIGS. 2, 4, and 6 through 9 or the cooled stream **53a** leaving heat exchanger **10** in FIG. 10 may not contain any liquid (because it is above its dewpoint, or because it is above its cricondenbar), so that separator **11** shown in FIGS. 2, 4, and 6 through 10 is not required.

[0071] The high pressure liquid (stream **33**) in FIGS. 2, 4, and 6 through 9 need not be expanded and fed to a mid-column feed point on the distillation column. Instead, all or a portion of it (dashed stream **47**) may be combined with the portion of the

separator vapor (stream **34**) to form combined stream **35** that flows to heat exchanger **15**. Any remaining portion of the liquid (dashed stream **37**) may be expanded through an appropriate expansion device, such as expansion valve **12**, to form stream **37a** which is then fed to a mid-column feed point on distillation column **19** (FIGS. 2, 4, 6, and 7) or stripper column **29** (FIGS. 8 and 9). Stream **33** in FIGS. 2, 4, and 6 through 9 and/or stream **37** in FIGS. 2, 4, and 6 through 10 may also be used for inlet gas cooling or other heat exchange service before or after the expansion step prior to flowing to the demethanizer.

[0072] In accordance with this invention, the use of external refrigeration to supplement the cooling available to the inlet gas and/or the distillation stream from other process streams may be employed, particularly in the case of a rich inlet gas. The use and distribution of separator liquids and demethanizer side draw liquids for process heat exchange, and the particular arrangement of heat exchangers for inlet gas cooling must be evaluated for each particular application, as well as the choice of process streams for specific heat exchange services.

[0073] It will also be recognized that the relative amount of feed found in each branch of the split vapor feed will depend on several factors, including gas pressure, feed gas composition, the amount of heat which can economically be extracted from the feed, and the quantity of horsepower available. More feed to the top of the column may increase recovery while decreasing power recovered from the expander thereby increasing the recompression horsepower requirements. Increasing feed lower in the column reduces the horsepower consumption but may also reduce product recovery. The

relative locations of the mid-column feeds may vary depending on inlet composition or other factors such as desired recovery levels and amount of liquid formed during inlet gas cooling. Moreover, two or more of the feed streams, or portions thereof, may be combined depending on the relative temperatures and quantities of individual streams, and the combined stream then fed to a mid-column feed position.

[0074] The present invention provides improved recovery of C_3 components and heavier hydrocarbon components per amount of utility consumption required to operate the process. An improvement in utility consumption required for operating the demethanizer process may appear in the form of reduced power requirements for compression or re-compression, reduced power requirements for external refrigeration, reduced energy requirements for tower reboilers, or a combination thereof.

[0075] While there have been described what are believed to be preferred embodiments of the invention, those skilled in the art will recognize that other and further modifications may be made thereto, e.g. to adapt the invention to various conditions, types of feed, or other requirements without departing from the spirit of the present invention as defined by the following claims.

WE CLAIM:

1. In a process for the separation of a gas stream containing methane, C_2 components, C_3 components, and heavier hydrocarbon components into a volatile residue gas fraction and a relatively less volatile fraction containing a major portion of said C_2 components, C_3 components, and heavier hydrocarbon components or said C_3 components and heavier hydrocarbon components, in which process

(a) said gas stream is cooled under pressure to provide a cooled stream;

(b) said cooled stream is expanded to a lower pressure whereby it is further cooled; and

(c) said further cooled stream is directed into a distillation column and fractionated at said lower pressure whereby the components of said relatively less volatile fraction are recovered;

the improvement wherein following cooling, said cooled stream is divided into first and second streams; and

(1) said first stream is cooled to condense substantially all of it and is thereafter expanded to said lower pressure whereby it is further cooled;

(2) said expanded cooled first stream is thereafter supplied at a first mid-column feed position to said distillation column;

(3) said second stream is expanded to said lower pressure and is supplied to said distillation column at a second mid-column feed position;

(4) a vapor distillation stream is withdrawn from a region of said distillation column below said expanded second stream and is compressed to higher pressure;

(5) said compressed vapor distillation stream is cooled sufficiently to condense at least a part of it, thereby forming a condensed stream;

(6) at least a portion of said condensed stream is expanded to said lower pressure and is thereafter supplied to said distillation column at a top feed position;

(7) an overhead vapor stream is withdrawn from an upper region of said distillation column and is directed into heat exchange relation with said compressed vapor distillation stream and heated, thereby to supply at least a portion of the cooling of step (5), and thereafter discharging at least a portion of said heated overhead vapor stream as said volatile residue gas fraction; and

(8) the quantities and temperatures of said feed streams to said distillation column are effective to maintain the overhead temperature of said distillation column at a temperature whereby the major portions of the components in said relatively less volatile fraction are recovered.

2. The process according to claim 1, wherein prior to cooling,

said gas is divided into said first and second streams;

said second stream is cooled and thereafter expanded to said lower pressure and supplied to said distillation column at said second mid-column feed position; and

said vapor distillation stream is withdrawn from the region of said distillation column below said expanded cooled second stream and is compressed to higher pressure.

3. The process according to claim 1, wherein prior to said division,

said cooled gas stream is partially condensed;

said partially condensed gas stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is thereafter divided into said first and second streams; and

at least a portion of said at least one liquid stream is expanded to said lower pressure and is supplied to said distillation column at a third mid-column feed position.

4. The process according to claim 3, wherein

said first stream is combined with at least a portion of said at least one liquid stream to form a combined stream, and said combined stream is cooled to condense substantially all of it and is thereafter expanded to said lower pressure whereby it is further cooled;

said expanded cooled combined stream is thereafter supplied at said first mid-column feed position to said distillation column; and

any remaining portion of said at least one liquid stream is expanded to said lower pressure and is supplied to said distillation column at said third mid-column feed position.

5. The process according to claim 2, wherein

said second stream is partially condensed;

said partially condensed second stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is expanded to said lower pressure and supplied to said distillation column at said second mid-column feed position; and

at least a portion of said at least one liquid stream is expanded to said lower pressure and is supplied to said distillation column at a third mid-column feed position.

6. In a process for the separation of a gas stream containing methane, C_2 components, C_3 components, and heavier hydrocarbon components into a volatile residue gas fraction and a relatively less volatile fraction containing a major portion of said C_2 components, C_3 components, and heavier hydrocarbon components or said C_3 components and heavier hydrocarbon components, in which process

(a) said gas stream is cooled under pressure to provide a cooled stream;

(b) said cooled stream is expanded to a lower pressure whereby it is further cooled; and

(c) said further cooled stream is directed into a distillation column and fractionated at said lower pressure whereby the components of said relatively less volatile fraction are recovered;

the improvement wherein following cooling, said cooled stream is divided into first and second streams; and

(1) said first stream is cooled to condense substantially all of it and is thereafter expanded to said lower pressure whereby it is further cooled;

(2) said expanded cooled first stream is thereafter supplied at a mid-column feed position to a contacting and separating device that produces an overhead vapor stream and a bottom liquid stream, whereupon said bottom liquid stream is supplied to said distillation column;

(3) said second stream is expanded to said lower pressure and is supplied to said contacting and separating device at a first lower feed position;

(4) a vapor distillation stream is withdrawn from an upper region of said distillation column to form at least a first distillation stream;

(5) said first distillation stream is compressed to higher pressure;

(6) said compressed first distillation stream is cooled sufficiently to condense at least a part of it, thereby forming a condensed stream;

(7) at least a portion of said condensed stream is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at a top feed position;

(8) any remaining portion of said vapor distillation stream is directed to said contacting and separating device at a second lower feed position;

(9) said overhead vapor stream is directed into heat exchange relation with said compressed first distillation stream and heated, thereby to supply at least a portion of the cooling of step (6), and thereafter discharging at least a portion of said heated overhead vapor stream as said volatile residue gas fraction; and

(10) the quantities and temperatures of said feed streams to said contacting and separating device are effective to maintain the overhead temperature of said

contacting and separating device at a temperature whereby the major portions of the components in said relatively less volatile fraction are recovered.

7. The process according to claim 6, wherein prior to cooling,

said gas is divided into said first and second streams; and

said second stream is cooled and thereafter expanded to said lower pressure and is supplied to said contacting and separating device at said first lower feed position.

8. The process according to claim 6, wherein said cooled gas stream is partially condensed;

said partially condensed gas stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is thereafter divided into said first and second streams; and

said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at a mid-column feed position.

9. The process according to claim 8, wherein

said first stream is combined with at least a portion of said at least one liquid stream to form a combined stream, and said combined stream is cooled to condense substantially all of it and is thereafter expanded to said lower pressure whereby it is further cooled;

said expanded cooled combined stream is thereafter supplied at said mid-column feed position to said contacting and separating device; and

any remaining portion of said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at said mid-column feed position.

10. The process according to claim 7, wherein

said second stream is cooled under pressure sufficiently to partially condense it;

said partially condensed second stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is expanded to said lower pressure and is supplied to said contacting and separating device at said first lower feed position; and

said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at a mid-column feed position.

11. The process according to claim 6, wherein

said further cooled stream is directed into said distillation column and fractionated at said lower pressure whereby the components of said relatively less volatile fraction are recovered and a first overhead vapor stream is produced;

said expanded cooled first stream is thereafter supplied at said mid-column feed position to said contacting and separating device that produces a second overhead vapor stream and said bottom liquid stream, whereupon said bottom liquid stream is supplied to said distillation column;

a vapor distillation stream is withdrawn from a region of said contacting and separating device above said expanded second stream and is compressed to higher pressure;

said compressed vapor distillation stream is cooled sufficiently to condense at least a part of it, thereby forming said condensed stream;

said first overhead vapor stream is directed to said contacting and separating device at said second lower feed position; and

said second overhead vapor stream is directed into heat exchange relation with said compressed vapor distillation stream and heated, thereby to supply at least a portion of the cooling of step (6), and thereafter discharging at least a portion of said heated second overhead vapor stream as said volatile residue gas fraction.

12. The process according to claim 11, wherein prior to cooling,

said gas is divided into said first and second streams;

said second stream is cooled and thereafter expanded to said lower pressure and is supplied to said contacting and separating device at said first lower feed position; and

said vapor distillation stream is withdrawn from a region of said contacting and separating device above said expanded cooled second stream and is compressed to higher pressure.

13. The process according to claim 11, wherein prior to said division,

said cooled gas stream is partially condensed;

said partially condensed gas stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is thereafter divided into said first and second streams; and

said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at a mid-column feed position.

14. The process according to claim 13, wherein

said first stream is combined with at least a portion of said at least one liquid stream to form a combined stream, and said combined stream is cooled to condense substantially all of it and is thereafter expanded to said lower pressure whereby it is further cooled;

said expanded cooled combined stream is thereafter supplied at said mid-column feed position to said contacting and separating device; and

any remaining portion of said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at said mid-column feed position.

15. The process according to claim 12, wherein

said second stream is partially condensed;

said partially condensed second stream is separated thereby to provide a vapor stream and at least one liquid stream;

said vapor stream is expanded to said lower pressure and is supplied to said contacting and separating device at said first lower feed position; and

said at least one liquid stream is expanded to said lower pressure and supplied to said distillation column at a mid-column feed position.

16. The improvement according to claim 1, 3, or 4 wherein said vapor distillation stream is withdrawn from a region of said distillation column above said expanded second stream and is thereafter compressed to higher pressure.

17. The improvement according to claim 2 wherein said vapor distillation stream is withdrawn from a region of said distillation column above said expanded cooled second stream and is thereafter compressed to higher pressure.

18. The improvement according to claim 5 wherein said vapor distillation stream is withdrawn from a region of said distillation column above said expanded vapor stream and is thereafter compressed to higher pressure.

19. The improvement according to claim 1, 3, or 4 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said distillation column at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said distillation column at a mid-column feed position above that of said expanded second stream.

20. The improvement according to claim 2 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said distillation column at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said distillation column at a mid-column feed position above that of said expanded cooled second stream.

21. The improvement according to claim 5 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said distillation column at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said distillation column at a mid-column feed position above that of said expanded vapor stream.

22. The improvement according to claim 6, 8, or 9 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at a mid-column feed position above that of said expanded second stream.

23. The improvement according to claim 7 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at a mid-column feed position above that of said expanded cooled second stream.

24. The improvement according to claim 10 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at a mid-column feed position above that of said expanded vapor stream.

25. The improvement according to claim 11, 12, 13, 14, or 15 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said contacting and separating device at a mid-column feed position above the region wherein said vapor distillation stream is withdrawn.

26. The improvement according to claim 16 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said distillation column at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said distillation column at a mid-column feed position above the region wherein said vapor distillation stream is withdrawn.

27. The improvement according to claim 17 or 18 wherein

(1) said condensed stream is divided into at least a first portion and a second portion;

(2) said first portion is expanded to said lower pressure and is thereafter supplied to said distillation column at said top feed position; and

(3) said second portion is expanded to said lower pressure and is thereafter supplied to said distillation column at a mid-column feed position above the region wherein said vapor distillation stream is withdrawn.