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(72) Inventors: ROY E. CAMPBELL
JOHN D. WILKINSON

(54) HYDROCARBON GAS PROCESSING

(71) We, THE ORTLOFF CORPORATION, a corporation of the State of Texas, having its principal place of business at the Wilco Building, Midland, Texas, United States of America, do hereby declare the invention, for which we pray that a patent may be granted to us and the method by which it is to be performed, to be particularly described in and by the following statement:–

This invention relates to the processing of gas streams containing hydrocarbons and other gases of similar volatility to remove desired condensible fractions. In particular, the invention is concerned with processing of gas streams such as natural gas, synthetic gas and refinery gas streams to recover most of the propane and a major portion of the ethane content thereof, together with substantially all of the heavier hydrocarbon content of the gas.

Gas streams containing hydrocarbons and other gases of similar volatility which may be processed according to the present invention include natural gas, synthetic gas streams obtained from other hydrocarbon materials such as coal, crude oil, naphtha, oil shale, tar sands, and lignite. Natural gas typically has a major proportion of methane and ethane (i.e., the combined C_1 and C_2 fractions constitute at least 50% of the gas on a molar bases). There may also be lesser amounts of the relatively heavier hydrocarbons such as propane, butanes, pentanes, and the like as well as H_2, N_2, CO_2 and other gases. A typical analysis of a natural gas stream to be processed in accordance with the invention would be, in approximate mol %, 80% methane, 10% ethane, 5% propane, 0.5% iso-butane, 1.54 normal butane, 0.25% isopentane, 0.25% pentane, 0.5% hexane plus, with the balance made up of nitrogen and carbon dioxide. Sulfur-containing gases are also often found in natural gas.

Recent substantial increases in the market for the ethane and propane components of natural gas has provided demand for processss yielding higher recovery levels of these products, Available processes for separating these materials include those based upon cooling and refrigeration of gas, oil absorption, refrigerated oil absorption, and the more recent cryogenic processes utilizing the principle of gas expansion through a mechanical device to produce power while simultaneously extracting heat from the system. Depending upon the pressure of the gas source, the richness (ethane and heavier hydrocarbons content) of the gas and the desired end products, each of these prior art processes or a combination thereof may be employed.

The cryogenic expansion type recovery process is now generally preferred for ethane recovery because it provides maximum simplicity with ease of start up, operating flexibility, good efficiency, safety, and good reliability. U.S. Patents Nos. 3,360,944, 3,292,380, and 3,292,381 describe relevant processes.

In a typical cryogenic expansion type recovery process a feed gas stream under pressure is cooled by heat exchange with other streams of the process and/or external sources of cooling such as a propane compression-refrigeration system. As the gas is cooled, liquids are condensed and are collected in one or more separators as a high-pressure liquid feed containing most of the desired C_2+ components. The high pressure liquid feed is then

expanded to a lower pressure. The vaporization occurring during expansion of the luquid results in further cooling of the remaining portion of the liquid. The cooled stream comprising a mixture of liquid and vapor is demethanized in a demethanizer column. The demethanizer is a fractionating column in which the expansion-cooled stream is fractionated to separate residual methane, nitrogen and other volatile gases as overhead vapor from the desired products of ethane, propane and heavier components as bottom product.

If the feed stream is not totally condensed, typically it is not, the vapor remaining from this partial condensation is expanded to a lower pressure. Additional liquids are condensed as a result of the further cooling of the stream during expansion. The pressure after the expansion is usually the same pressurs at which the demethanizer is operated. Liquids thus obtained are also supplied as a feed to the demethanizer. Typically, remaining vapor and the demethanizer overhead vapor are combined as the residual methane product gas.

In the ideal operation of such a separation process the vapors leaving the process will contain substantially all of the methane found in the feed gas to the recovery plant, and substantially no hydrocarbons equivalent to ethane or heavier components. The bottoms fraction leaving the demethanizer will contain substantially all of the heavier components and essentially no methane. In practice, however, this ideal situation is not obtained for the reason that the conventional demethanizer is operated largely as a stripping column. The methane produce in the process, therefore, typically comprises vapors leaving the top fractionation stage of the column together with vapors not subjected to any rectification step. Substantial losses of ethane occur because the vapors discharged from the low temperature separation steps contain ethane and heavier components which could be recovered if those vapors could be brought to lower temperatures or if they were brought in contact with a significant quantity of relatively heavy hydrocarbons, for example C_3 and heavier, capable of absorbing the ethane.

As described in our co-pending British Patent application No. 16420/77, serial No. 1 532 336 improved ethane recovery is achieved by pre-cooling the condensed high-pressure liquid prior to expansion. Such pre-cooling will reduce the temperature of the flash-expanded liquid feed supplied to the demethanizer and thus improve ethane recovery. Moreover, as described in application No. 16420/77, serial No. 1 532 336, by pre-cooling the high pressure liquid feed, the temperature of the expanded liquid may be sufficiently reduced that it can be used as top column feed in the demethanizer, while the expanded vapor is supplied to the demethanizer at a feed point intermediate the top feed and column bottom. This variation permits recovery of ethane contained in the expanded vapor which would otherwise be lost.

It will be obvious thah to supply external refrigeration at this stage of the process is difficult because of the extremely low temperatures encountered. In typical demethanizer operations the expanded liquid and vapor feeds are typically at temperatures in the order of −120°F. to −190°F. Accordingly, pre-cooling of the condensed high pressure liquid stream feed can best be achieved by heat exhange of the condensed high pressure liquid stream feed with streams derived within the process as described in Patent application No. 16420/77, (Serial No. 1 532 336).

As already indicated, in modern gas processing plants the vapors remaining from partial condensation of the feed gas are usually expanded to the operating pressure of the demethanizer column in a turbo-expander and, prior to the invention disclosed in Patent application No. 16420/77 (Serial No. 1 532 336) supplied to the demethanizer as the top feed. A turbo-expander is a machine which extracts useful work from the gas during expansion by expanding that gas in a substantially isentropic fashion. Such a work expansion has two advantages. First, it permits cooling the vapor portion to the coldest practicable temperature. Attainment of such cold temperatures is important in the top feed to the demethanizer to provide the most complete recovery of C_2+ components from the incoming feed gas. Second, the useful work recovered by isentropic expansion can be used to supply a portion of the compression requirements ordinarily required in the process.

As explained in the Patent application No. 16420/77, (Serial No. 1 532 336), the liquids recovered from partial condensation of the feed gas may be cooled below their bubble point, and if this is done, upon flash expansion it is possible to achieve flash-expanded temperatures of that sub-cooled liquid even below the temperature achieved by work-expansion of the vapors from partial condensation. Where such low temperatures are achieved in the flash-expanded liquids, it is then usually preferable to supply that flash-expanded liquid as the column feed at a point above the feed point of the work-expanded vapor recovered from partial condensation. The flash-expanded temperature of the sub-cooled liquid may be further reduced by combining the liquid with a process gas stream which reduces the bubble point of the sub-cooled liquid as explained in our co-pending Patent application No. 16399/77 (Serial No. 1 532 335), filed concurrently herewith.

In accordance with the present invention, u method of separating a feed gas containing

hydrocarbons into a volatile residue gas and a relatively less volatile fraction, methane and ethane together comprising a major portion of the feed gas, in which

(a) the feed gas under pressure is cooled to partially condense the gas and form thereby a feed liquid portion and a feed vapor portion;

(b) at least some of the feed liquid portion is expanded to a lower pressure whereby it partially vaporizes to further cool the expanded feed liquid portion;

(c) at least part of the expanded feed liquid portion is supplied at a first feed point to a fractionation column wherein the relativepy less volatile fraction is separated;

(d) the feed vapor portion is expanded to the lower pressure; and

(e) at least part of the expanded feed vapor portion is supplied as a feed to the fractionation column; is characterised in that at least some of the feed vapor portion is reheated prior to expansion thereof, and the reheated portion of the feed vapor portion is expanded in a work expansion machine and then supplied to the fractionation column at a second feed point which is below the first feed point.

This is based on the discovery that when an alternative process stream is available to maintain the cold top column temperature, for example the expanded liquid portion of the partially condensed feed gas, the vapors from work expansion of the partially condensed feed gas can be employed as a feed to the demethanizing column at an intermediate position, and that it is advantageous to turn back some or all of those vapors and warm them prior to work expansion. The mechanical work recovered and refrigeration developed in the expansion machine is greater as a result of expansion beginning at a warmer temperature. Because of the importance of heat economy in natural gas processing, it is generally preferable to turn back the vapors from partial condensation in accordance with the present invention and employ those vapors in a heat exchange relation with all or a portion of the incoming feed stream. This provides the desired warming of the turned-back vapor portion. In this manner vapor turn-back can reduce the need for external refrigeration which might otherwise be required, or alternately could be used to increase recovery of liquid products.

Preferably the feed liquid portion which is expanded is cooled to a temperature below its bubble point before it is expanded. As used above and elsewhere in the specification and claims, the term bubble point refers to a liquid mixture of volatile components and is the temperature at which the first infinitesimal bubble of vqpor will occur.

The present invention will be better understood by reference to the following drawings and examples, in which:–

Figure 1 is a flow diagram of a single-stage cryogenic expander natural gas processing plant of the prior art incorporating a set of conditions for a typically rich natural gas stream;

Figure 2 is a flow diagram of a single-stage cryogenic expander natural gas processing plant of the prior art incorporating a set of conditions for a typically lean natural gas stream;

Figure 3 is a flow diagram of a gas processing plant embodying the inventions forming the subject matter of Patent Applications Nos. 16399/77 and 16420/77 (Serial Nos 1 532 335 and 1 532 336), which is employed as a base case;

Figure 4 is a flow diagram of a gas processing plant in accordance with the present invention;

Figure 5 is a variation of the present invention in which a portion of ths condensed high-pressure liquid feed is sub-cooled and supplied as an intermediate column feed;

Figure 6 is a variation of the present invention in which a portion of the high-pressure vapor is used as vapor turnback and a portion is expanded directly to the demethanizer;

Figures 7A and 7B are graphs showing carbon dioxide as a function of temperature for one embodiment of this invention compared to the prior art.

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 pound moles per hour) have been rounded to the nearest whole number, for convenience. The total stream flow 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.

Referring to Figure 1, for a fuller description of a typical conventional ethane recovery process, plant inlet gas from which carbon dioxide and sulfur compounds have been removed (if the concentration of these compounds in the plant inlet gas would cause the produce stream not to meet specifications, or cause icing in the equipment), and which has been dehydrated enters the process at 120°F. and 910 psia as stream 23. It is divided into two parallel streams and cooled to 45°F. by heat exchange with cool residue gas at 5°F. in exchanger 10; with produce liquids (stream 26) at 82°F. in exchanger 11; and with demethanizer liquid at 53°F. in demethanizer reboiler 12. From these exchangers, the streams recombine and enter the gas chiller, exchanger 13, where the combined stream is cooled to 10°F. with propane refrigerant at °F. The cooled stream is again divided into two

parallel streams and further chilled by heat exchange with cold residue gas (stream 29) at −107°F. in exchange 14, and with demethanizer liquids at −80°F. in demethanizer side reboiler 15. The streams are recombined and enter a high-pressure separator 16 at −45°F. and 900 psia as stream 23a. The condensed liquid (stream 24) is separated and fed to the demethanizer 19 through expansion valve 30. An expansion engine may be used in place of the expansion valve 30 if desired.

The cooled gas from the high pressure separator 16 flows through expander 17 where it is work expanded from 900 psia to 290 psia. The work expansion chills the gas to −125°F. Expander 17 is preferably a turbo-expander, having a compressor 21 mounted on the expander shaft. For convenience, expander 17 is sometimes hereinafter referred to as the expansion means. In certain prior art embodiments, expander 17 is replaced by a conventional expansion valve.

Liquid condensed during expansion is separated in low pressure separator 18. The liquid is fed on level control through line 25 to the demethanizer column 19 at the top and flows from a chimney tray (not shown) as top feed to the column 19.

It should be noted that in certain embodiments low pressure separator 18 may be included as part of demethanizer 19, occupying the top section of the column. In this case, the expander outlet stream enters above a chimney tray at the bottom of the separator section, located at the top of the column. The liquid then flows from the chimney tray as top feed to the demethanizing section of the column.

As liquid fed to demethanizer 19 flows down the column, it is contacted by vapors which strip the methane from the liquid to produce a demethanized liquid product at the bottom. The heat required to generate stripping vapors is provided by heat exchangers 12 and 15.

The vapors stripped from the condensed liquid in demethanizer 19 exit through line 27 to join the cold outlet gas from separator 18 via line 28. The combined vapor stream then flows through line 29 back through heat exchangers 14 and 10. Following these exchangers, the gas flows through compressor 21 driven by expander 17 and directly coupled thereto. Compressor 21 compresses the gas to a discharge pressure of about 305 psia. The gas then enters a compressor 22 and is compressed to a final discharge pressure of 900 psia.

Inlet and liquid component flow rates, outlet liquid recoveries and compression requirements for this prior art process shown in Figure 1 are given in the following table:

TABLE I
(Fig. 1)

Stream Flow Rate Summary – Lb. Moles/Hr.

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL
23	1100	222	163	130	1647
24	795	202	157	129	1300
25	16	10	5	1	32
26	3	162	157	130	453

RECOVERIES

Ethane	72.9%	29,296 GAL/DAY
Propane	96.2%	39,270 GAL/DAY

COMPRESSION HORSEPOWER

Refrigeration	256 BHP
Recompression	892 BHP
Total	1148 BHP

In Figure 2 a typical lean natural gas stream is processed and cooled using a prior art process similar to that shown in Figure 1. The inlet gas stream 33 is cooled to −67°F. and flows to high pressure separator 16 as stream 33a where the liquid contained therein is separated and fed on level control through line 34 and expansion valve 30 to demethanizer

19 in the middle of the column.

Cold gas from separator 16 flows through expander 17 where because of work expansion from 900 psia to 250 psia, the gas is chilled to −153°F. The liquid condensed during expansion is separated in low pressure separator 18 and is fed on level control through line 35 to the demethanizer 19 as top feed to the column.

The data for this case are given in the following table:

TABLE II
(Fig. 2)

Stream Flow Rate Summary – Lb. Moles/Hr.

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL
33	1447	90	36	43	1647
34	280	42	25	39	391
35	133	35	11	4	186
36	2	71	36	43	155

RECOVERIES

Ethane	79.0%	17,355 GAL/DAY
Propane	98.2%	8,935 GAL/DAY

COMPRESSION HORSEPOWER

Refrigeration	0 BHP
Recompression	1180 BHP
Total	1180 BHP

In the prior art cases discussed with respect to Figure 1 and Figure 2 above, recoveries of ethane are 73% for the case of the rich gas feed and 79% for the lean gas feed. It is recognized that some improvement in yield may result by adding one or more cooling steps followed by one or more separation steps, or by altering the temperature of separator 16 or the pressure in separator 18. Recoveries of ethane and propane obtained in this manner, while possibly improved over the cases illustrated by Figure 1 and Figure 2, are significantly less than yields which can be obtained in accordance with the process of the present invention.

As indicated above, the present invention may be used as an improvement in the gas recovery process as set forth in said co-pending application No. 16420/77 (Serial No. 1 532 336) filed concurrently herewith. Figure 3 illustrates a gas recovery facility employing the invention described in these applications and will be employed as a base case for purposes of explaining the present invention. In addition, in the flow plan of Figure 3, the subcooled liquid is combined with a portion of the vapors from partial condensation. Such a further step reduces the bubble point of the subcooled liquid as explained in our co-pending application No. 16399/77 (Serial No. 1 532 335) filed concurrently herewith. With respect to Figure 3, the process flow conditions discussed below and flow rates set forth in Table IV have been calculated on the basis of a lean feed gas composition as set forth in Table II as stream 33.

Referring to Figure 3, plant inlet gas 33 from which carbon dioxide and sulphur compounds have been removed and which has been dehydrated enters the process at 120°F and 910 psia. It is divided into two parallel streams and cooled to −3°F by heat exchange with cool residue gas 43b at −27°F in heat exchanger 10; with liquid product (stream 44) at 46°F in heat exchanger 11; and with demethanizer liquid at 4F in demethanizer reboiler 12. After recombining the combined stream at −3°F is further cooled to −21°F by external refrigeration such as a propane refrigerant at −27°F. The stream is again divided into two parallel streams and is further cooled by heat exchange with cold residue gas stream 43a at −75°F in heat exchanger 14 and with demethanizer liquids at − 139°F in demethanizer side reboiler 15. The streams are combined and supplied as stream 33a to high pressure

separator 16 at −67°F and 900 psia where the condensed liquid is separated. The liquid from separator 16 (stream 34) is at its bubble point and is combined with a portion of the vapor from separator 16 (stream 42), which is at its dew point. The combined stream then passes through heat exchanger 45 in heat exchange relation with overhead vapor stream 43 from the demethanizer. This cools and condenses the combined stream to well below the bubble point of the initial liquid stream 34. The cooled and condensed stream at −146°F is then expanded through an appropriate expansion device such as expansion valve 46 to a pressure of about 290 psia. During expansion, a portion of the feed will vaporize, resulting in cooling of the remaining portion. In the process illustrated in this case, expanded stream 47a leaving expansion valve 46 reaches a temperature of −155°F and is supplied to the demethanizer 19 as the top feed.

The remaining vapor from separator 16 (stream 41) enters a work expansion engine in which mechanical energy is extracted from this portion of the high pressure vapor. As the vapor is expanded from a pressure of about 900 psia to about 290 psia, work expansion cools the expanded vapor 41a to a temperature of approximately −145°F. The expanded and partially condensed vapor 41a is supplied to the demethanizer 19 at an intermediate point.

The temperature and pressure conditions of some of the principal streams are summarized in Table III below as base case A, and a stream flow summary for base case A is set forth in table IV below.

For purposes of further comparison, a base case B has been calculated following the same flow diagram as in Figure 3 but at a somewhat lower column pressure than in base case A. Under the conditions of base case B, more refrigeration could be extracted from residue gas streams 43, 43a and 43b, and the demethanizer reboiler, making it possible to eliminate external refrigeration in heat exchanger 13. This reduced the horsepower required by the process but also reduced the ethane and propane recoveries.

A summary of the process conditions of the principal streams for base case B is set forth below in Table III and a stream flow rate summary for base case B is set forth below in Table V.

TABLE III
(Fig. 3)

STREAM CONDITIONS

Stream	Bas Case A	Base Case B
33	120°F; 910 psia	120°F; 910 psia
33a,34,41,42	–67°F; 900 psia	–67°F; 900 psia
41a	–145°F; 290 psia	–148°F; 275 psia
43	–154°F	–154°F
43a	–75°F	–112°F
43b	–27°F	25°F
43c	98°F	115°F
44	46°F	44°F
47	–146°F; 900 psia	–145°F; 900 psia
47a	–155°F; 290 psia	–155°F; 275 psia

RECOVERIES

	Base Case A	Base Case B
ETHANE	92.56%. 20,323 GAL/DAY	90.52%; 19,876 GAL/DAY
Propane	97.89%; 8,910 GAL/DAY	97.56%; 8.881 GAL/DAY

HORSEPOWER REQUIREMENTS

	Base Case A	Base Case B
Refrigeration	118 BHP	0 BHP
Recompression	1045 BHP	1116 BHP
Total	1163 BHP	116 BHP

TABLE IV
(Fig. 3)

Stream Flow Rate Summary, Base Case A – Lb. Moles/Hr.

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL
33	1447	90	36	43	1647
34	280	42	25	39	391
41	856	36	8	3	921
42	311	12	3	1	335
43	1446	6	0	0	1475
44	1	84	36	43	172

TABLE V
(Fig. 3)

Stream Flow Rate Summary, Base Case B – Lb. Moles/Hr.

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL
33	1447	90	36	43	1647
34	280	42	25	49	391
41	1078	45	10	4	1160
42	89	3	1	0	96
43	1445	8	0	0	1479
44	2	82	36	43	168

The present invention is illustrated by the following examples:

Example 1

Figure 4 sets forth a process diagram for a typical natural gas plant in accordance with the present invention.

The flow plan is similar to the flow plan of Figure 3 except for the provision for vapor turnback. In Figure 4, inlet gas is cooled and partially condensed through heat exchangers 10, 11, 12, 14 and 15 generally as described in connection with Figure 3. It will be noted, however, that it was not found necessary in Figure 4 to make provision for external refrigeration (e.g., heat exchanger 13 of Figure 3). Moreover, in Figure 4, it will also be noted that in the second set of feed gas coolers, the feed is divided into three portions rather than two. A portion of the feed is cooled in heat exchanger 57, as will be further explained below; another portion is cooled in heat exchanger 14 by heat exchange with cool residue gas stream 52a; and the third portion is cooled in heat exchanger 15 by heat exchange with demethanizer liquid in demethanizer side reboiler 15. The cooled and partially condensed feed gas 33a is supplied to separator 16 at −67°F and 900 paia.

Following first the liquid from separator 16, stream 34 at its bubble point is combined with a portion 50 of the vapor from separator 16. The combined stream then passes through heat exchanger 54 in heat exchange relation with the overhead vapor product (stream 52) from demethanizer 19, resulting in cooling and condensation of the combined stream at a temperature well below the bubble point of the stream 34. The cooled stream 55 is then expanded through an appropriate expansion device, such as expansion valve 56, to a pressure of about 290 psia. During expansion, a portion of the feed will vaporize, resulting in cooling of the remaining part. In the process of Figure 4, the expanded stream 55a leaving expansion valve 56 reaches a temperature of −155°F, and is supplied to demethanizer 19 as top feed.

The remaining vapor from separator 16 (stream 51) becomes the turn-back stream. The vapor turn-back 51 flows through heat exchanger 57 in heat exchange relation with part of the plant inlet feed. In the process of Figure 4, the turn-back vapor 51a from exchanger 57 is at about 5°F and flows through expander 17 where because of work expansion from about 895 psia to 290 psia, the gas 51b is chilled to −99°F. The chilled stream 51b from expander 17 flows to demethanizer 19 at an intermediate point.

A summary of the principal streams in this example of the present case is set forth below in Table VI.

As will be seen from Table VI below, in this example of the present invention 92.53% of the ethane and 97.88% of the propane were recovered. 1005 brake horsepower of recompression were required to operate the process. By comparison with base case A above, it will be seen that for substantially the same recovery, the present invention reduces the horsepower requirements for process operation and in the case of this example eliminated the need for external refrigeration.

TABLE VI
(Fig. 4)

Stream summary, Example 1.

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL	CONDITIONS
33a	1447	90	36	43	1647	120°F, 910 psia
33a	1447	90	36	43	1647	–67°F, 900 psia
34	280	42	25	39	391	–67°F
50	311	12	3	1	335	–67°F
51	856	36	8	3	921	–67°F
51a	856	36	8	3	921	5°F., 895 psia
51b	856	36	8	3	921	–99°F., 290 psia
52	1445	6	0	0	1475	–154°F., 290 psia
52a	1445	6	0	0	1475	–75°F
52b	1445	6	0	0	1475	8°F
52c	1445	6	0	0	1475	110°F
53	2	84	36	43	172	45°F
55	591	54	28	40	726	–146°F, 900 psia
55a	591	54	28	40	726	–155°F, 290 psia

RECOVERIES

Ethane	92.53%	20,318 GAL/DAY
Propane	97.88%	8,910 GAL/DAY

HORSEPOWER REQUIREMENTS

Refrigeration	0 BHP
Recompression	1005 BHP
Total	1005 BHP

Example 2

Example 2 (Figure 5) is another illustration of the present invention. In Example 2, a portion of the high-pressure liquid condensate was sub-cooled by residue gas from the demethanizer and flashed directly into the demethanizer at an intermediate feed position in the column.

Referring to Figure 5, the inlet gas is processed and cooled in a manner similar to that of Figure 4 in heat exchangers 10, 11, 12, 14, 15 and 57 to provide a partly condensed feed gas 33a at −67°F at −900 psia. The cooled inlet stream 33a then enters high-pressure separator 16 where the condensed liquid is separated at its bubble point.

The vapor from high-pressure separator 16 is divided into two portions. The first portion 60 is combined with a portion 64 of the liquid 34 stream from exchanger 62 wherein liquid 34 from separator 16 is sub-cooled to below its bubble point. The remaining portion of vapor from separator 16 enters heat exchanger 57 where it is used to cool a portion of plant inlet feed gas. From exchanger 57 the vapor stream 61a enters expander 17 where, because of work expansion from 895 psia to 25 psia, the gas is chilled to −108°F. From expander 17

the stream 61b flows to demethanizer 19 at its lowest feed point.

The cooled liquid 34 from high pressure separator 16 enters heat exchanger 62 where it is sub-cooled to about −150°F by heat exchange with a portion of cold residue gas 70. Following exchanger 62 the sub-cooled liquid is divided into two portions. The first portion 63 flows through expansion valve 65 where it undergoes expansion and flash vaporization and is cooled to −158°F. From expansion valve 65 the stream 63a enters demethanizer 19 at its middle feed point. The remaining liquid portion 64 is combined with a portion 60 of the high pressure separator vapor. The combined stream then flows through heat exchanger 66 where it is cooled to −153°F by heat exchange with a portion of the cold residue gas stream 70. From exchanger 66 the subcooled stream 67 enters expansion valve 68 and undergoes flash vaporization as the pressure is reduced to about 250 psia. From valve 68, the stream 67a now at −163°F flows to demethanizer 19 at its top feed point.

The vapors stripped from the condensed liquid in demethanizer 19 exit as residue gas 70. As already indicated, the residue gas 70 is divided and used as the refrigerant in exchangers 62 and 66. The residue gas from these exchangers is recombined and flows through the balance of the system to exchangers 14 and 10 where it is used to cool and partially condense the feed gas 33.

A summary of the condition of some of the principal streams is set forth in Table VII.

TABLE VII
(Figure 5)

Stream Conditions and Flow Rates

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL	CONDITIONS	
33	1447	90	36	93	1647	120°F.,	910 psia
33a	1447	90	36	43	1647	−67°F.,	900 psia
34	280	42	25	39	391	−67°F.	
60	164	6	2	1	176	−67°F.	
61	1003	42	9	3	1080	−67°F.	
61a	1003	42	9	3	1080	5°F.,	900 psia
61b	1003	42	9	3	1080	−108°F.,	250 psia
63	140	21	12	19	195	−150°F.,	900 psia
63a	140	21	12	19	195	−158°F.,	250 psia
64	140	21	12	19	195	−150°F.,	900 psia
67	304	27	14	20	272	−153°F.,	900 psia
67a	304	27	14	20	272	−163°F.,	250 psia
70	1444	6	0	0	1479	−161°F.,	250 psia
71	3	84	36	43	168	39°F.	

RECOVERIES

Ethane	93.02%	20,426 GAL/DAY
Propane	98.57%	8,972 GAL/DAY

COMPRESSION HORSEPOWER

Refrigeration	0 BHP
Recompression	1111 BHP
Total	1111 BHP

– The fore going invention of turning back some (or all) of the high pressure feed gas vapors separated upon partial condensation is generally applicable in process flow plans where an alternate stream is available to maintain the demethanizer column at the desired overhead operating temperature. Cooling of high-pressure condensate prior to expansion, and supplying the cooled expanded condensate at an upper feed point in the column is a particularly preferred means of maintaining column overhead temperature. As indicated above, in co-pending application No. 16420/77 (Serial No. 1 532 336), a variety of processes are disclosed for coolinv of the high-pressure condensate recovered from the feed gas before expanding that condensate to the demethanizer operating pressure. The advantages of cooling the high pressure condensate before expansion can be enhanced in accordance with our co-pending application No. 16399/77 (Serial No. 1 532 335) by combining that condensate with a portion of the vapor from the high-pressure separator in order to lower the temperature which would be obtained upon expansion of the condensate.

Variations of the invention of this application include the following:

(1) Some or all of the high-pressure condensate may be cooled by auto-refrigeration. In such a procedure, a cooled portion of the high-pressure condensate is divided into two portions. One portion is expanded to the column operating pressure which causes a portion of it to vaporize and to cool the expanded stream. The expanded portion is then directed into heat exchange relation with the high-pressure condensate to obtain the cooled condensate prior to expansion. The second portion of the cooled stream is expanded to a low temperature and supplied to the demethanizer as the column top feed. In this embodiment the cooled high-pressure condensate may be divided into two portions and each portion separately expanded. If more convenient, the entire cooled condensate stream can be expanded to the demethanizer pressure, and the expanded stream resulting then divided into the two portions discussed above.

(2) In another variation, all or a portion of the high-pressure condensate supplied as the top column feed may be heat exchanged prior to expansion with liquid in the demethanizer column in one or more side stream reboilers.

(3) In either variation (1) or (2), the amount of cooling obtained by expansion of the cooled high-pressure liquid can be enhanced by combining the high-pressure condensate with a portion of the high-pressure vapor as explained in our co-pending application No. 16399/77 (Serial No. 1 532 335). This variation is also applicable, as shown in Figures 4 and 5, to cases where the high-pressure condensate is cooled by residue gas. This variation is also applicable, as shown in Figures 4 and 5, to cases where the high-pressure condensate is cooled by residue gas. This variation is particularly valuable in the treatment of lean feed gases, where there is sometimes a limited amount of high-pressure condensate available.

(4) When employing turn-back of the high-pressure vapors, particularly in lean gas cases, very substantial amounts of high-pressure vapor are available and, if heated to too great an extent in the turn-back heat exchanger (e.g., exchanger 57 of Figures 4 and 5), the temperature reached by the turn-back gases after expansion will tend to overheat the demethanizer column and thus raise the column overhead temperature.

A number of expedients are available in such a situation: Only a portion of the high pressure may be turned back through exchanger 57, and the balance of the high-pressure vapors supplied directly to the demethanizer to maintain column overhead temperature. The balance thus supplied directly to the demethanizer may be expanded in a turbo-expander, may be cooled (or partially condensed) by heat exchange against column overhead vapors and expanded into the demethanizer column or it may be used to enrich all or a portion of the high-pressure liquid condensate as explained above. Still another alternate would be to cool the expanded turn-back vapors if a cooling stream is available at an appropriate temperature within the process. In cases where large amounts of vapors are available for use as a turn-back stream, it may be necessary, to avoid overheating the demethanizer, to limit the temperature to which the turned-back vapors are warmed in exchanger 57.

(5) In the illustrations of the present invention set forth in the above examples, the turn-back vapors have been used to cool a portion of the incoming feed gas in the second set of heat exchangers (i.e., in parallel with exchangers 14 and 15 of Figures 4 and 5). It will be appreciated, however, that in a gas treatment process as generally illustrated in Figures 1 through 5 there may be a variety of alternate needs for a cold gas stream, such as is available from the high-pressure separator, where the refrigeratin therein may be used even more effectively than indicated in examples 1 and 2. By way of illustration, the turned-back vapors may be used in lieu of propane refrigeration in a heat exchanger located intermediate between the two sets of feed gas precoolers such as exchanger 13 of Figure 1. Still another variant, the turned-back vapors may be used to cool all or a portion of the incoming feed gas at the initial condition of 120°F such as through exchangers 10, 11, or 12 of Figures 1 and 2. Still another variation, where propane refrigeration is employed, is to use the

turned-back vapors to subcool the condensed propane refrigerant prior to employing the refrigerant in the process operation.

(6) In still another variation of the present invention, the feed gas vapor from separator 16 may be divided into two portions, the first of which is used as the vapor turn-back and the second of which is used to control the column overhead. In this embodiment, the second stream would be heat exchanged against cold residue gas from the demethanizer overhead. This may result in substantial condensation of the cooled feed gas vapor if the vapor is below its critical pressure. If the stream is above the critical pressure, it will remain single phase through the cooling. The second portion would then be expanded and supplied as the top column feed. The vapor turn-back portion would be reheated as previously described, work expanded, and supplied as a lower column feed. In these variations the expanded turn-back vapors may also be heat exchanged with residue gas.

(7) The process flow plans and examples of the present invention have been described for convenience using shell and tube heat exchangers. In cryogenic operations, it is usually preferred to use specially designed heat exchangers such as plate-fin heat exchangers. Such special heat exchangers have improved heat transfer characteristics which may permit closer temperature approaches in the heat exchangers, lower cost, and also permit flow arrangements to accommodate heat exchange of several streams concurrently.

Example 3

Example 2, as illustrated in Figure 6, is an example of the present invention in which a portion of the high-pressure feed gas obtained from partial condensation is employed as turn-back vapor and another portion is expanded directly to column pressure through a work expansion engine.

Referring to Figure 6, a lean feed gas is supplied to the process at a temperature of 120°F and a pressure of 910 psia at stream 33. Lean feed gas 33 is of the same composition referred to above in connection with Figure 2. The feed gas is cooled to a temperature of −67°F and a pressure of 900 psia through heat exchanges 10, 11, 12, 14, 15 and 57, generally as described above in connection with Figures 4 and 5. The partially condensed feed 33a is supplied to separator 16 wherein the liquid and vapor is separated. Liquid portion 34 at its bubble point is drawn off from separator 16, cooled below its bubble point in heat exchanger 75 to a temperature of −150°F (stream 34a), and then passed through expansion valve 76. The expanded stream 34b at −158°F is supplied to demethanizer 19 as a top column feed.

The vapors drawn off from separator 16 are separated into two portions, 77 and 78. Portions 77 is expanded in work expansion engine 79. The expanded stream 77a achieves a temperature of −153°F and is supplied to the demethanizer as an intermediate column feed. Work extracted from stream 77 in expander 79 is in part employed to recompress residue gas by means of associated compressor 80.

Turn-back vapors 78 from separator 16 are directed through heat exchanger 57 to precool a portion of liquid feed 33. The warmed turn-back bapors 78a leave exchangec 57 at a temperature of 50°F. The warmed vapors are then expanded in expansion engine 81 and supplied to the demethanizer column as a second intermediate feed at a feed point below feed 77a at −77°F. Expander 81 is a connected to an associated compressor 82.

Residue gas in the process illustrated in Figure 6 is obtained as a demethanizer overhead 83. Demethanizer overhead is employed to provide a part of the refrigeration required in the process by cooling (i) liquid 34 in exchanger 75, (ii) partly cooled feed gas in exchanger 14, and (iii) hot feed gas in exchanger 10. Thereafter, the residue gas is recompressed to line pressure first in compressor 80 driven by work engine 79; second, in compressor 82 driven by work engine 81; and finally, in supplementary compressor 84.

A summary of the principal stream flow rates and conditions is set forth below in Table VIII. As can be seen, in the process illustrated in Figure 6, an ethane recovery of 89.16% and propane recovery of 97.73% at a total horsepower requirement of 1057 BHP.

TABLE IX
(Fig. 6)

Process Stream Summary

STREAM	METHANE	ETHANE	PROPANE	BUTANES+	TOTAL	CONDITIONS
33	1447	90	36	43	1647	120°F., 910 psia
33a	1447	90	36	43	1647	–67°F., 900 psia
34	280	42	25	39	391	–67°F.
34a	280	42	25	39	391	–150°F., 900 psia
34b	280	42	25	39	391	–158°F., 250 psia
77	584	24	5	2	628	–67°F., 900 psia
77a	584	24	5	2	628	–153°F., 250 psia
78	584	24	5	2	628	– 67°F., 900 psia
78a	584	24	5	2	628	– 50°F., 895 psia
78b	584	24	5	2	628	– 77°F., 250 psia
83	1445	10	1	0	1483	–156°F., 250 psia
83a	1445	10	1	0	1483	–124°F.
83b	1445	10	1	0	1483	70°F.
83c	1445	10	1	0	1483	84°F.
85	2	81	36	43	164	42°F.

RECOVERIES

Ethane	89.16%	19,578 GAL/DAY
Propane	97.73%	8,896 GAL/DAY

COMPRESSION HORSEPOWER

Refrigeration	0 BHP
Recompression	1057 BHP
TOTAL	1057 BHP

As is well known, natural gas streams usually contain carbon dioxide, sometimes in substantial amounts. The presence of carbon dioxide in the demethanizer can lead to icing of the column internals under cryogenic conditions. Even when feed gas contains less than 1% carbon dioxide it fractionates in the demethanizer, and can build up to concentrations of as much as 5% to 10% or greater. At such concentrations, carbon dioxide can freeze out depending on temperature, pressure, whether the carbon dioxide is in the liquid or vapor phase, and the solubility of carbon dioxide in the liquid phase.

In the present invention, it has been found that when the vapor from the high pressure separator is expanded and supplied to the demethanizer below the top column feed position, the problem of carbon dioxide icing can be substantially mitigated. The high-pressure separator gas typically contains a large amount of methane relative to the amount of ethane and carbon dioxide. when supplied as a mid-column feed, therefore, the high-pressure separator gas tends to dilute the carbon dioxide concentration and to prevent it from

increasing to icing levels.

The advantage of the present invention can be readily seen by plotting carbon dioxide concentration and temperature for various trays of the demethanizer when practicing the present invention and when following the prior art. A chart thus constructed for processing the gas as described above in Example 1 (see Figure 4 and Table IV) and containing 0.72% carbon dioxide, can be compared with a similar chart constructed for the process of Figure 2 (prior art) applied to the same gas (see Figures 7-A and 7-B). These charts also include equilibria for vapor-solid and liquid-solid conditions. The equilibrium data given in Figures 7A and 7B are for the methane-carbon dioxide system. These data are generally considered representative for the methane and ethane systems. If the CO_2 concentration at a particular point in the column is at or above the equilibrium level for that temperature, icing can be expected. For practical design purposes, the engineer usually requires a margin of safety, i.e., the actual concentration be less than the "icing" concentration by a suitable safety factor.

As is evident, when following the prior art process of Figure 2 (per Figure 7-A), the vapor conditions at point A touches the line representing solid vapor phase equilibria. By contrast, in Figure 7-B, neither the liquid nor vapor conditions reach or exceed their related equilibria condition. Hence, icing risks are materially reduced.

It should be noted in connection with foregoing that when designing demethanizer columns for use in the present invention the designer will routinely verify that icing in the column will not occur. Even when vapor is fed at a mid column position it is possible that icing may occur if the process is designed for the highest possible ethane recovery. Such designs normally call for the coldest practical temperature at the top of the column. This will result in the carbon dioxide concentrations shifting to the right on the plots of Figures 7-A and 7-B. Depending on the particular application, the result can be an objectionably high concentration of carbon dioxide near the top of the column. For such a circumstance, it may be necessary to accept a somewhat lower ethane recovery to avoid column icing or to pre-treat the feed gas to reduce carbon dioxide levels to the point where they can be tolerated in the demethanizer. In the alternative, it may be possible to avoid icing in such a circumstance by other alterations in the process conditions. For instance, it may be possible to operate the high pressure separator at a different temperature, to change the amount of re-heat, or to increase the quantity of vapor directed through the re-heater. If such alterations can be made within the limitations of the process heat balance, icing may be avoided without significant loss of ethane recovery.

In connection with the foregoing description of our invention, it should be noted that where the feed to the top of the demethanizer is a liquid which is expanded from a high pressure to a lower column operating pressure (as in Figures 4, 5 and 6), liquid may be auto cooled before expansion. Such auto cooling will involve splitting the top liquid feed into two streams either before or after expansion, and directing one of the two streams thus obtained after expansion into heat exchange relation with the top column liquid feed before expansion.

WHAT WE CLAIM IS:–

1. A method of separating a feed gas containing hydrocarbons into a volatile residue gas and a relatively less volatile fraction, methane and ethane together comprising a major portion of the feed gas, in which

(a) the feed gas under pressure is cooled to partially condense the gas and form thereby a feed liquid portion and a feed vapor portion;

(b) at least some of the feed liquid portion is expanded to a lower pressure whereby it partially vaporizes to further cool the expanded feed liquid portion;

(c) at least part of the expanded feed liquid portion is supplied at a first feed point to a fractionation column wherein the relatively less volatile fraction is separated;

(d) the feed vapor portion is expanded to the lower pressure; and,

(e) at least part of the expanded feed vapor portion is supplied as a feed to the fractionation column; characterised in that at least some of the feed vapor portion is reheated prior to expansion thereof, and the reheated portion of the feed vapor portion is expanded in a work expansion machine and then supplied to the fractionation column at a second feed point which is below the first feed point.

2. A method according to claim 1, in which at least some of the feed liquid portion is cooled to a temperature below its bubble point before it is expanded.

3. A method according to claim 2, in which the feed vapor portion is reheated by heat exchange with at least a portion of the feed gas, thereby providing part of the cooling required to partially condense the feed gas.

4. A method according to claim 2, in which the feed liquid portion to be cooled is cooled by heat exchange with residue gas from the fractionation column.

5. A method according to claim 2, in which the feed liquid portion to be cooled is cooled

increasing to icing levels.

The advantage of the present invention can be readily seen by plotting carbon dioxide concentration and temperature for various trays of the demethanizer when practicing the present invention and when following the prior art. A chart thus constructed for processing the gas as described above in Example 1 (see Figure 4 and Table IV) and containing 0.72% carbon dioxide, can be compared with a similar chart constructed for the process of Figure 2 (prior art) applied to the same gas (see Figures 7-A and 7-B). These charts also include equilibria for vapor-solid and liquid-solid conditions. The equilibrium data given in Figures 7A and 7B are for the methane-carbon dioxide system. These data are generally considered representative for the methane and ethane systems. If the CO_2 concentration at a particular point in the column is at or above the equilibrium level for that temperature, icing can be expected. For practical design purposes, the engineer usually requires a margin of safety, i.e., the actual concentration be less than the "icing" concentration by a suitable safety factor.

As is evident, when following the prior art process of Figure 2 (per Figure 7-A), the vapor conditions at point A touches the line representing solid vapor phase equilibria. By contrast, in Figure 7-B, neither the liquid nor vapor conditions reach or exceed their related equilibria condition. Hence, icing risks are materially reduced.

It should be noted in connection with foregoing that when designing demethanizer columns for use in the present invention the designer will routinely verify that icing in the column will not occur. Even when vapor is fed at a mid column position it is possible that icing may occur if the process is designed for the highest possible ethane recovery. Such designs normally call for the coldest practical temperature at the top of the column. This will result in the carbon dioxide concentrations shifting to the right on the plots of Figures 7-A and 7-B. Depending on the particular application, the result can be an objectionably high concentration of carbon dioxide near the top of the column. For such a circumstance, it may be necessary to accept a somewhat lower ethane recovery to avoid column icing or to pre-treat the feed gas to reduce carbon dioxide levels to the point where they can be tolerated in the demethanizer. In the alternative, it may be possible to avoid icing in such a circumstance by other alterations in the process conditions. For instance, it may be possible to operate the high pressure separator at a different temperature, to change the amount of re-heat, or to increase the quantity of vapor directed through the re-heater. If such alterations can be made within the limitations of the process heat balance, icing may be avoided without significant loss of ethane recovery.

In connection with the foregoing description of our invention, it should be noted that where the feed to the top of the demethanizer is a liquid which is expanded from a high pressure to a lower column operating pressure (as in Figures 4, 5 and 6), liquid may be auto cooled before expansion. Such auto cooling will involve splitting the top liquid feed into two streams either before or after expansion, and directing one of the two streams thus obtained after expansion into heat exchange relation with the top column liquid feed before expansion.

WHAT WE CLAIM IS:–

1. A method of separating a feed gas containing hydrocarbons into a volatile residue gas and a relatively less volatile fraction, methane and ethane together comprising a major portion of the feed gas, in which

(a) the feed gas under pressure is cooled to partially condense the gas and form thereby a feed liquid portion and a feed vapor portion;

(b) at least some of the feed liquid portion is expanded to a lower pressure whereby it partially vaporizes to further cool the expanded feed liquid portion;

(c) at least part of the expanded feed liquid portion is supplied at a first feed point to a fractionation column wherein the relatively less volatile fraction is separated;

(d) the feed vapor portion is expanded to the lower pressure; and,

(e) at least part of the expanded feed vapor portion is supplied as a feed to the fractionation column; characterised in that at least some of the feed vapor portion is reheated prior to expansion thereof, and the reheated portion of the feed vapor portion is expanded in a work expansion machine and then supplied to the fractionation column at a second feed point which is below the first feed point.

2. A method according to claim 1, in which at least some of the feed liquid portion is cooled to a temperature below its bubble point before it is expanded.

3. A method according to claim 2, in which the feed vapor portion is reheated by heat exchange with at least a portion of the feed gas, thereby providing part of the cooling required to partially condense the feed gas.

4. A method according to claim 2, in which the feed liquid portion to be cooled is cooled by heat exchange with residue gas from the fractionation column.

5. A method according to claim 2, in which the feed liquid portion to be cooled is cooled

by heat exchange with liquid from the demethanizer column in a side reboiler.

6. A method according to claim 2, in which

(i) the cooled feed liquid portion is divided into a first part and a remaining part;

(ii) The first part is expanded to the lower pressure, whereby a portion thereof vaporizes to further cool the expanded first part;

(iii) the expanded first part is used to cool the portion of the feed liquid portion to be cooled by heat exchange therewith, whereby the cooled feed liquid is obtained; and,

the remaining part is expanded to the lower pressure, and at least some of the expanded remaining part is supplied to the demethanizer at the first feed point.

7. A method according to claim 2, in which

(i) a part of the feed vapor portion is combined with the cooled feed liquid portion prior to expansion;

(ii) the combined stream is expanded to the lower pressure, the temperature of the combined stream prior to expansion being below the bubble point of the feed liquid portion; and,

(iii) at least some of the expanded combined stream is supplied to the fractionation column at the first feed point.

8. A method according to claim 2, in which

(i) the feed vapor portion is divided into at least a first part and a second part, the first part being the portion which is reheated and then expanded to the lower pressure before being supplied to the fractionation column at the second feed point;

(ii) the second part without heating thereof is expanded to ths lower pressure; and

(iii) the expanded second part is supplied to the fractionation column at a feed point above the second feed point.

9. A method according to claim 8, in which the second part is cooled prior to its expansion to the lower pressure.

10. A method according to claim 2 in which the expanded reheated portion of the feed vapor portion is cooled by heat exchange with at least a portion of the feed gas under pressure.

11. A method according to claim 1, in which a first part of the feed vapor portion is expanded to the lower pressure, whereby the first part is cooled, and at least a portion of the expanded first part is supplied to the fractionation column as the column top feed, a second part of the feed vapor portion being the portion which is reheated, expanded, and supplied to the column at the second feed point, the second part being reheated by heat exchange with at least a portion of the feed gas under pressure.

12. A method according to claim 11, in which the first part is cooled prior to expansion thereof by heat exchange with at least a portion of the residue gas from the column.

13. A method according to claim 11, in which the expanded second part is further cooled by heat exchange with at least a portion of the residue gas from the column.

14. An apparatus for separating a feed gas containing hydrocarbons into a volatile residue gas and a relatively less volatile fraction by the method according to claim 1, methane and ethane together comprising a major portion of the feed gas, the apparatus comprising

(a) cooling means for cooling the feed gas under pressure to partially condense the gas sufficiently to form a feed liquid portion and a feed vapor portion;

(b) sub-cooling means connected to the cooling means for receiving at least some of the feed liquid portion and sub-cooling it to a temperature below its bubble point;

(c) a first expansion means connected to the sub-cooling means for receiving the sub-cooled feed liquid portion and expanding it to the lower pressure, whereby some of the liquid portion vaporizes to further cool the expanded sub-cooled feed liquid portion;

(d) a fractionation column connected to receive at least part of the expanded sub-cooled feed liquid portion at a first feed point and arranged to separate the relatively less volatile fraction; and

(e) feed vapor processing means which is connected to the cooling means for receiving the feed vapor portion therefrom and expanding it to the lower pressure, and which is further connected to the fractionation column to supply at least part of the expanded feed gas vapor portion to the column;

characterised in that the feed vapor processing means includes

(i) heat exchange means which is arranged to receive and reheat at least a portion of the feed vapor portion from the cooling means;

(ii) a second expansion means connected to the heat exchange means for receiving the reheated portion of the feed vapor portion and expanding this reheated vapor portion to the lower pressure while extracting work therefrom; and

(iii) means connecting the second expansion means to the fractionation column for supplying the expanded reheated feed vapor portion thereto at a second feed point which is at a

lower position on the column than the first feed point.

15. An apparatus according to claim 14, in which the heat exchange means is connected to reheat the feed vapor portion by heat exchange with a portion of the feed gas under pressure, thereby providing some of the cooling required to partially condense the feed gas.

16. An apparatus according to claim 14, in which the sub-cooling means comprises a heat exchanger connected to effect heat exchange between the feed liquid portion to be sub-cooled and cold residue gas from the column, whereby the feed liquid portion is sub-cooled and the cold residue gas is warmed.

17. An apparatus according to claim 14, in which the sub-cooling means comprises a fractionation column side reboiler for effecting heat exchange between the feed liquid portion to be sub-cooled and liquid from the fractionation column.

18. An apparatus according to claim 14, further comprising

(1) dividing means connected to receive at least some of the sub-cooled feed liquid portion from the sub-cooling means and to divide it into a first part and a remaining part;

(2) means connected to the dividing means for receiving and expanding the first part to the lower pressure, whereby a portion thereof vaporizes to cool the expanded first part;

(3) means for receiving the expanded first part and directing it to the sub-cooling means wherein it passes in heat exchange relation with the feed liquid portion to be sub-cooled; and,

(4) means connected to the dividing means for receiving and supplying the remaining part to the first expansion means.

19. An apparatus according to claim 14, in which the feed vapor processing means further includes

(1) dividing means for dividing the feed vapor portion into a first portion for reheating in the heat exchange means, and a second portion; and,

(2) means for combining the second portion of the feed vapor portion with the feed liquid portion to be sub-cooled prior to expansion thereof, thereby forming a combined stream for expansion by the first expansion means;

the sub-cooling means being arranged to cool at least one of the second portion of the feed vapor portion, the liquid portion to be sub-cooled, and the combined stream, whereby the combined stream is cooled to a temperature below the bubble point of the feed liquid portion.

20. An apparatus according to claim 14, in which the feed vapor processing means further includes

(1) means connected to the cooling means for receiving and dividing the feed vapor portion into at least a first part and a second part;

(2) means connecting the dividing means to the heat exchange means whereby the first part is reheated therein and is subsequently expanded to the lower pressure and supplied to the fractionation column at the second feed point;

(3) third expansion means connected to the dividing means for receiving and expanding the second part of the feed vapor portion to the lower pressure; and,

(4) means connecting the third expansion means to the fractionation column to supply the expanded second part of the feed vapor portion to the fractionation column at a feed point above the second feed point.

21. An apparatus according to claim 14, in which the feed vapour processing means further includes

(1) dividing means connected to the cooling means for receiving and dividing the feed vapor portion into at least a first part and a second part;

(2) means connecting the dividing means to the heat exchange means whereby the first part is reheated therein and subsequently expanded to the lower pressure and supplied to the fractionation column at the second feed point;

(3) second heat exchange means connected to the dividing means for receiving and cooling the second part;

(4) thirf expansion means connected to the second heat exchange means for receiving and expanding the cooled second part to the lower pressure; and,

(5) means connecting the third expansion means to the fractionation column at a feed point above the second feed point.

22. An apparatus according to claim 14, in which the means connecting the second expansion means to the fractionation column includes a heat exchange means connected to effect heat exchange between the expanded reheated feed vapor portion and at least a portion of residue gas from the column, whereby the expanded reheated feed vapor portion is cooled before being supplied to the fractionation column at the second feed point.

23. An apparatus according to claim 14, in which the feed vapor processing means further includes

(1) dividing means connected to the cooling means for receiving and dividing the feed

vapor portion into a first part and a second part;

(2) third expansion means connected to the dividing means for receiving and expanding the first part of the feed vapor portion to the lower pressure; and,

(3) means connecting the third expansion means to the fractionation column so that the expanded first part is supplied as a feed to the fractionation column;

the heat exchange means being connected to the dividing means so as to receive therefrom the second part of the feed vapor portion, and the heat exchange means being further connected to receive a portion of the feed gas under pressure whereby the second part of the feed vapor portion is reheated by heat exchange with the feed gas under pressure prior to expansion of the second part by the second expansion means and its subsequent supply to the fractionation column at the second feed point.

24. An apparatus according to claim 23, in which the sub-cooling means includes means for directing the feed liquid portion to be sub-cooled into heat exchange relation with at least a portion of the residue gas from the fractionation column.

25. An apparatus according to claim 23, comprising a further heat exchange means connected between the third expansion means and the fractionation column to receive the expanded second part, the further heat exchange means being further connected to direct the expanded second part into heat exchange relation with at least a portion of the residue gas from the fractionation column, thereby further cooling the expanded second part prior to supplying it to the fractionation column at the second feed point.

26. A method according to claim 1, substantially as described in any one of examples 1 to 3 herein.

27. An apparatus according to claim 14, substantially as described with reference to any one of Figures 4 to 6 of the accompanying drawings.

GILL, JENNINGS & EVERY
Chartered Patent Agents
53 to 64 Chancery Lane
London WC2A 1HN

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