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(54) **ENHANCED HEAVIES REMOVAL/LPG RECOVERY PROCESS FOR LNG FACILITIES**

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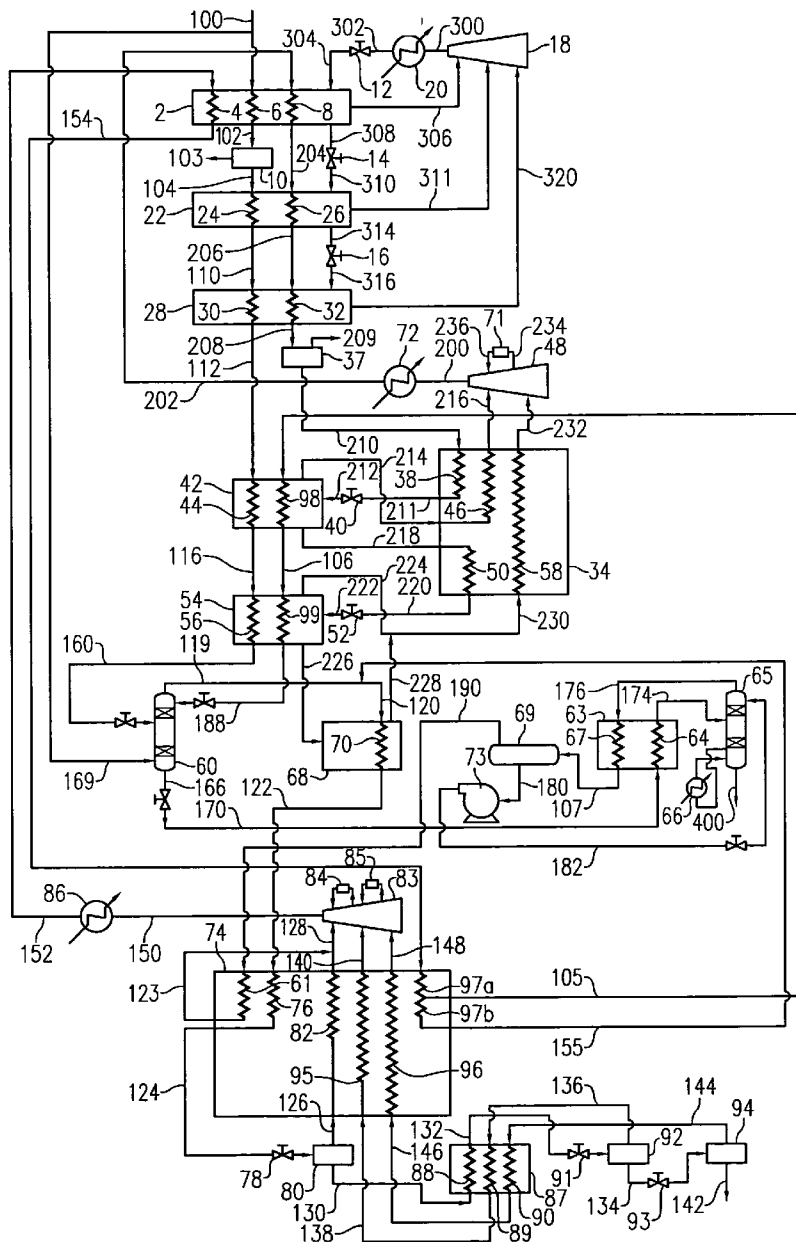
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(57) **ABSTRACT**

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A LNG facility, and operating method therefor, capable of more efficiently and/or effectively removing heavy hydrocarbon components from the processed natural gas stream and recovering the removed heavy components as LPG.

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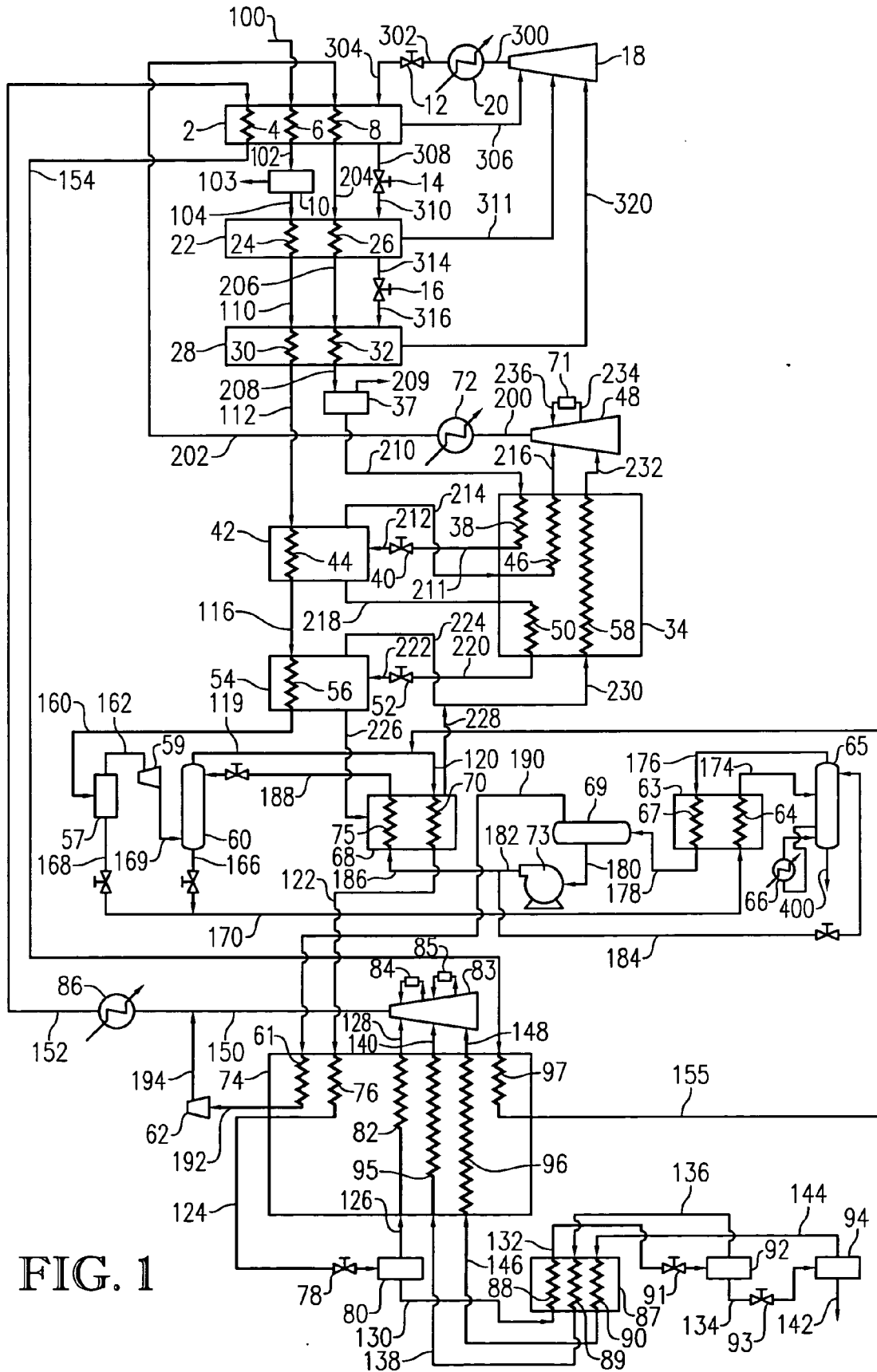


FIG. 1

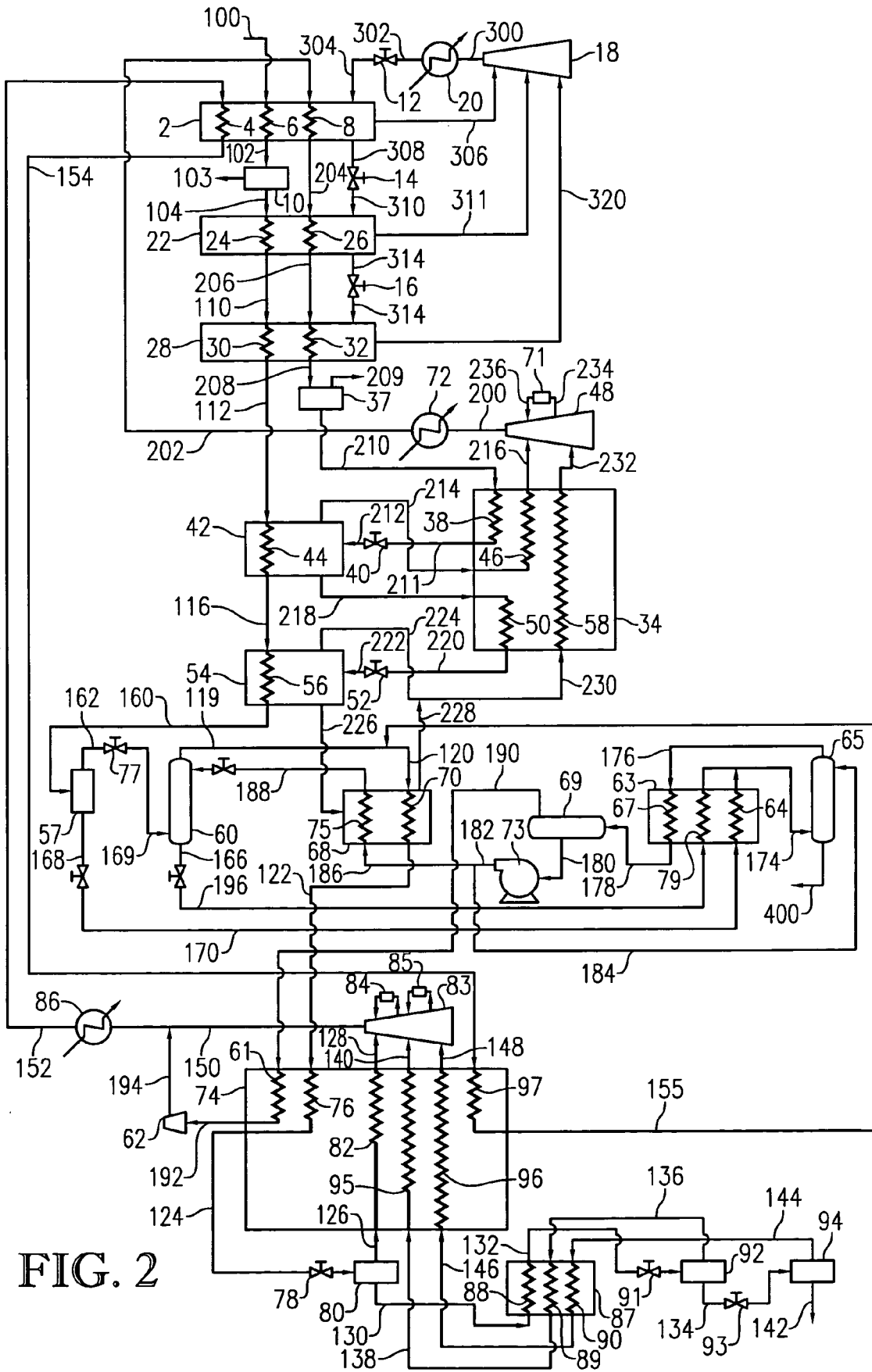


FIG. 2

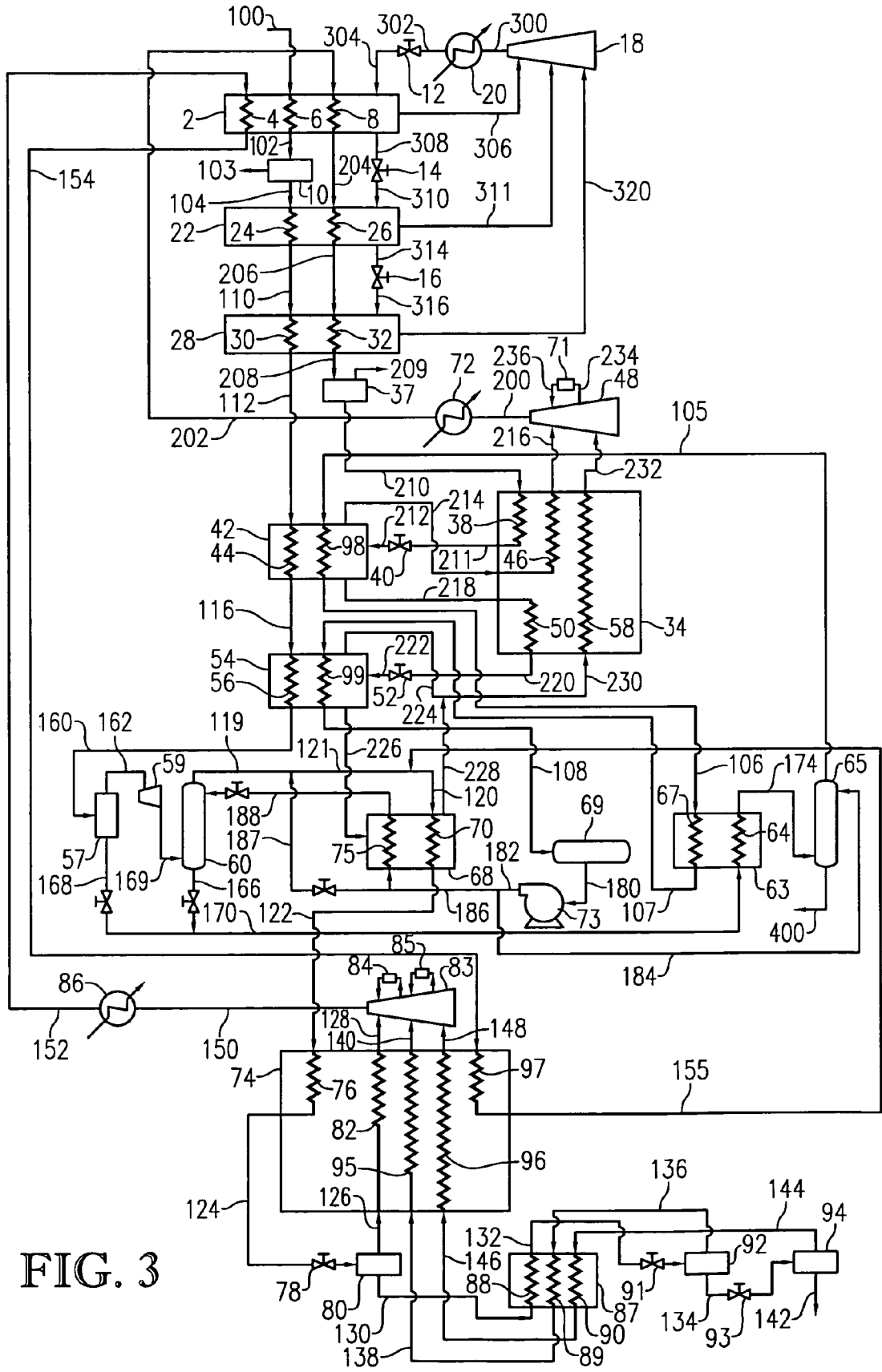


FIG. 3

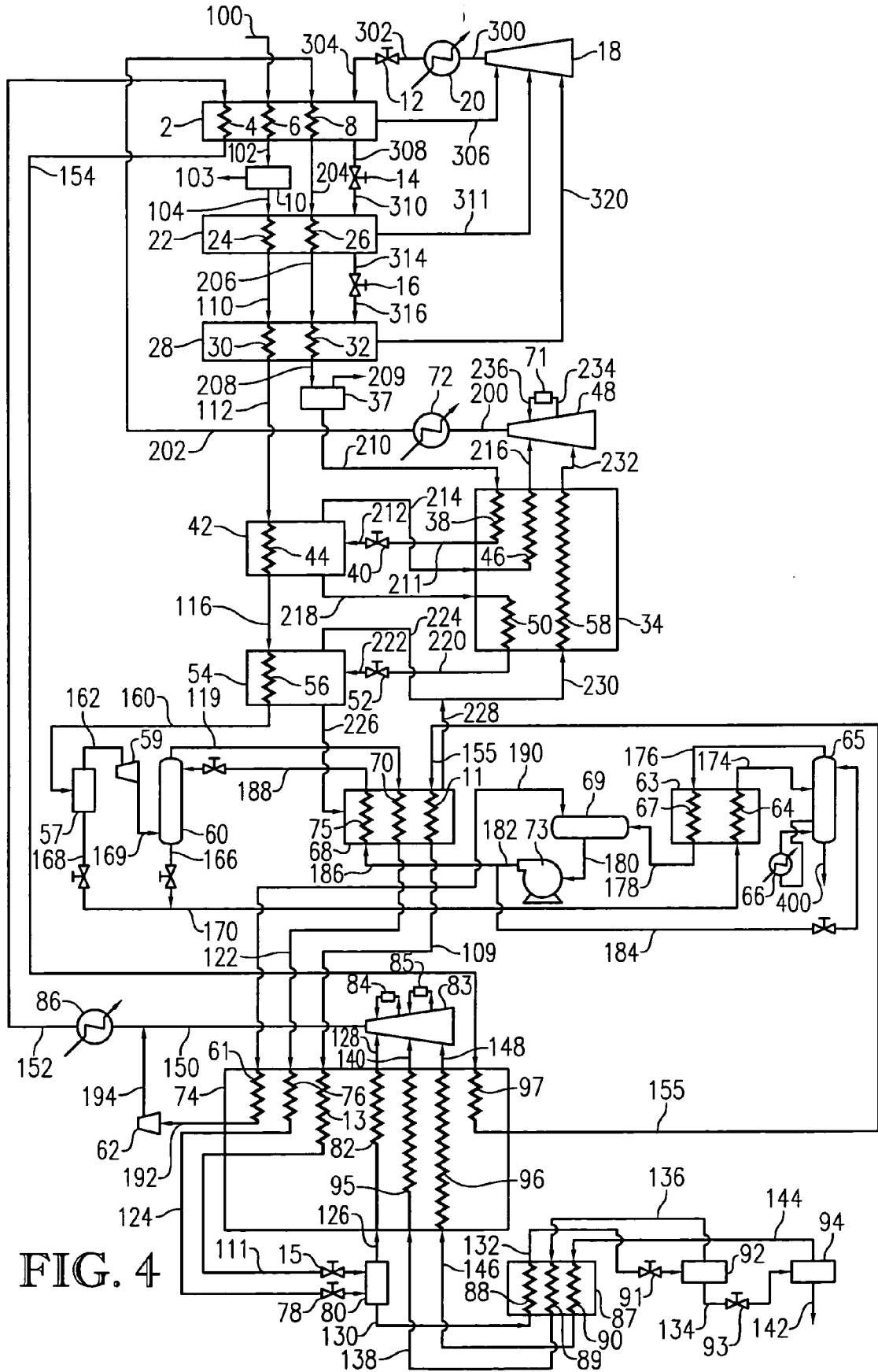


FIG. 4

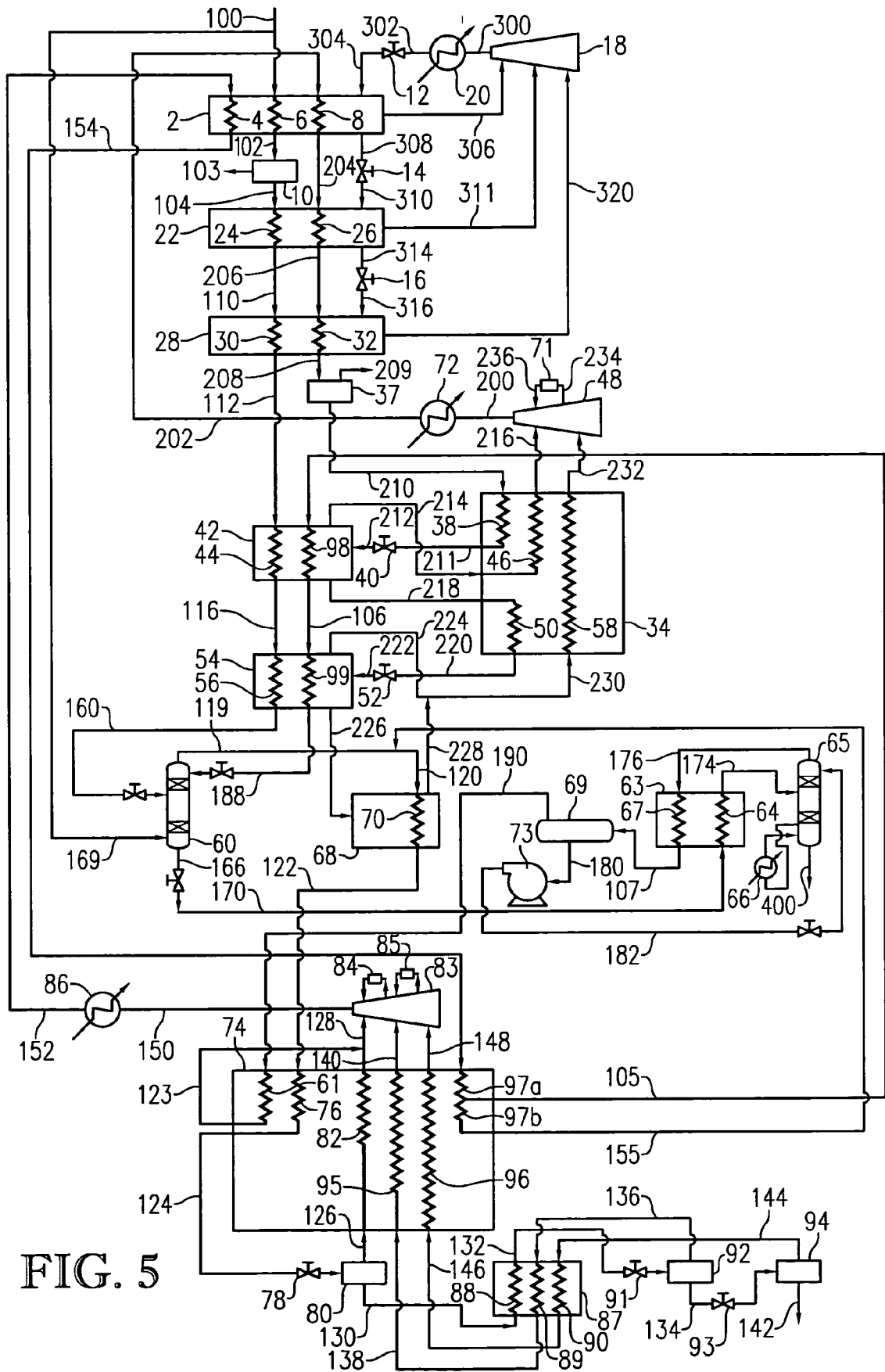


FIG. 5

ENHANCED HEAVIES REMOVAL/LPG RECOVERY PROCESS FOR LNG FACILITIES

BACKGROUND OF THE INVENTION

[0001] 1. Field of the Invention

[0002] This invention relates to a method and apparatus for liquefying natural gas. In another aspect, the invention concerns a liquefied natural gas (LNG) facility that provides more efficient and/or effective removal of heavy hydrocarbon components from the processed natural gas stream. In a further aspect, the invention concerns an LNG facility that provides more efficient and/or effective recovery of liquefied petroleum gas (LPG) from the processed natural gas stream.

[0003] 2. Description of the Prior Art

[0004] The cryogenic liquefaction of natural gas is routinely practiced as a means of converting natural gas into a more convenient form for transportation and storage. Such liquefaction reduces the volume of the natural gas by about 600-fold and results in a product which can be stored and transported at near atmospheric pressure.

[0005] Natural gas is frequently transported by pipeline from the supply source to a distant market. It is desirable to operate the pipeline under a substantially constant and high load factor but often the deliverability or capacity of the pipeline will exceed demand while at other times the demand may exceed the deliverability of the pipeline. In order to shave off the peaks where demand exceeds supply or the valleys when supply exceeds demand, it is desirable to store the excess gas in such a manner that it can be delivered when demand exceeds supply. Such practice allows future demand peaks to be met with material from storage. One practical means for doing this is to convert the gas to a liquefied state for storage and to then vaporize the liquid as demand requires.

[0006] The liquefaction of natural gas is of even greater importance when transporting gas from a supply source which is separated by great distances from the candidate market and a pipeline either is not available or is impractical. This is particularly true where transport must be made by ocean-going vessels. Ship transportation in the gaseous state is generally not practical because appreciable pressurization is required to significantly reduce the specific volume of the gas. Such pressurization requires the use of more expensive storage containers.

[0007] In order to store and transport natural gas in the liquid state, the natural gas is preferably cooled to -240° F. to -260° F. where the liquefied natural gas (LNG) possesses a near-atmospheric vapor pressure. Numerous systems exist in the prior art for the liquefaction of natural gas in which the gas is liquefied by sequentially passing the gas at an elevated pressure through a plurality of cooling stages whereupon the gas is cooled to successively lower temperatures until the liquefaction temperature is reached. Cooling is generally accomplished by indirect heat exchange with one or more refrigerants such as propane, propylene, ethane, ethylene, methane, nitrogen, carbon dioxide, or combinations of the preceding refrigerants (e.g., mixed refrigerant systems). A liquefaction methodology which is particularly applicable to the current invention employs an open methane cycle for the final refrigeration cycle wherein a pressurized LNG-bearing stream is flashed and the flash vapors (i.e., the flash gas

stream(s)) are subsequently employed as cooling agents, recompressed, cooled, combined with the processed natural gas feed stream and liquefied, thereby producing the pressurized LNG-bearing stream.

[0008] During cooling of the processed natural gas stream in the LNG facility, certain heavy hydrocarbon components are typically removed in order to avoid freezing of these heavy components (benzene in particular) in downstream heat exchangers. The removed heavy hydrocarbon components can be further fractionated and used as fuel gas and/or sold as liquefied petroleum gas (LPG). A variety of schemes have been used in the past to remove heavies from the processed natural gas stream and/or recover LPG from the stream. However, many of the conventional systems are difficult and/or expensive to operate. Further, many of the existing heavies removal/LPG recovery systems do not recover enough LPG to be economically feasible.

OBJECTS AND SUMMARY OF THE INVENTION

[0009] It is, therefore, an object of the present invention to provide a system for efficiently and effectively removing heavy hydrocarbon components from a natural gas stream undergoing liquefaction in an LNG facility.

[0010] A further object of the present invention to provide a system for efficiently and effectively recovering LPG from a natural gas stream undergoing liquefaction in an LNG facility.

[0011] It should be understood that the above-listed objects are only exemplary, and not all the objects listed above need be accomplished by the invention described and claimed herein.

[0012] One embodiment of the present invention concerns a process for liquefying a natural gas stream, the process comprising the following steps: (a) using a first distillation column to separate at least a portion of the natural gas stream into a first relatively more volatile fraction and a first relatively less volatile fraction; (b) using a second distillation column to separate at least a portion of the first relatively less volatile fraction into a second relatively more volatile fraction and a second relatively less volatile fraction, the second distillation column operating at a lower pressure than the first distillation column; (c) using a reflux portion of the second relatively more volatile fraction as reflux in the first and/or second distillation columns; and (d) cooling at least a portion of the reflux portion via indirect heat exchange with at least a portion of the first relatively less volatile fraction.

[0013] Another embodiment of the present invention concerns a process for liquefying a natural gas stream, the process comprising the following steps: (a) using a vapor/liquid separator to separate at least a portion of the natural gas stream into a predominately vapor separated portion and a predominately liquid separated portion; (b) using a first distillation column to separate at least a portion of the predominately vapor separated portion into a first relatively more volatile fraction and a first relatively less volatile fraction; (c) introducing a first reflux stream into an upper section of the first distillation column; and (d) cooling at least a portion of the first reflux stream via indirect heat exchange with at least a portion of the predominately liquid separated portion.

[0014] Yet another embodiment of the present invention concerns a process for liquefying a natural gas stream, the process comprising the following steps: (a) using a distillation column to separate at least a portion of the natural gas stream into a relatively more volatile fraction and a relatively less volatile fraction; (b) cooling at least a portion of the relatively more volatile fraction in a first heat exchange pass via indirect heat exchange with a first refrigerant; (c) flashing at least a portion of the cooled relatively more volatile fraction to thereby produce a flash gas; (d) compressing at least a portion of the flash gas; and (e) cooling at least a portion of the compressed flash gas in a second heat exchange pass via indirect heat exchange with the first refrigerant, the first and second heat exchange passes being separate from one another.

[0015] A further embodiment of the present invention concerns an apparatus for liquefying natural gas. The apparatus includes a vapor/liquid separator, a first distillation column, a heat exchanger, and a second distillation column. The vapor/liquid separator has a vapor outlet and a liquid outlet. The first distillation column is fluidly coupled to the vapor outlet and has a first overhead outlet and a first bottom outlet. The heat exchanger has a heating pass and a cooling pass, the heating and cooling passes are configured to facilitate indirect heat exchange between fluids flowing therethrough. The heating pass is fluidly coupled to the liquid outlet. The second distillation column is fluidly coupled to the first bottom outlet and has a second overhead outlet and a second bottom outlet. The cooling pass is fluidly coupled to the second overhead outlet.

BRIEF DESCRIPTION OF THE DRAWING FIGURES

[0016] FIG. 1 is a simplified flow diagram of the Oman LNG facility employing an improved system for removing heavies and recovering LPG from the processed natural gas stream.

[0017] FIG. 2 is a simplified flow diagram of a first alternative embodiment of the LNG facility.

[0018] FIG. 3 is a simplified flow diagram of a second alternative embodiment of the LNG facility.

[0019] FIG. 4 is a simplified flow diagram of a third alternative embodiment of the LNG facility.

[0020] FIG. 5 is a simplified flow diagram of a fourth alternative embodiment of the LNG facility.

DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENT

[0021] A cascaded refrigeration process uses one or more refrigerants to transfer heat energy from a natural gas stream to the refrigerants, and ultimately transferring the heat energy to the environment. In essence, the overall refrigeration system functions as a heat pump by removing heat energy from the natural gas stream as the stream is progressively cooled to lower and lower temperatures. The design of a cascaded refrigeration process involves a balancing of thermodynamic efficiencies and capital costs. In heat transfer processes, thermodynamic irreversibilities are reduced as the temperature gradients between heating and cooling fluids become smaller, but obtaining such small temperature gradients generally requires significant increases in the amount

of heat transfer area, major modifications to various process equipment, and the proper selection of flow rates through such equipment so as to ensure that both flow rates and outlet temperatures are compatible with the required heating/cooling duty.

[0022] As used herein, the term “open-cycle cascaded refrigeration process” refers to a cascaded refrigeration process comprising at least one closed refrigeration cycle and one open refrigeration cycle where the boiling point of the refrigerant employed in the open cycle is less than the boiling point of the refrigerant employed in the closed cycle and a portion of the cooling duty to condense the compressed open-cycle refrigerant is provided by one or more of the closed cycles. In one embodiment of the present invention, a predominately methane stream is employed as the refrigerant in the open cycle. This predominately methane stream is preferably derived from the processed natural gas feed stream and can include compressed open methane cycle gas streams. As used herein, the terms “predominantly,” “primarily,” “principally,” and “in major portion,” when used to describe the presence of a particular component of a fluid stream, shall mean that the fluid stream is comprised of at least 50 mole percent of the stated component. For example, a “predominantly” methane stream, a “primarily” methane stream, a stream “principally” comprised of methane, or a stream comprised “in major portion” of methane each denote a stream comprised of at least 50 mole percent methane.

[0023] One of the most efficient and effective means of liquefying natural gas is via an optimized cascade-type operation in combination with expansion-type cooling. Such a liquefaction process involves the cascade-type cooling of a natural gas stream at an elevated pressure, (e.g., about 650 psia) by sequentially cooling the gas stream via passage through a multistage propane refrigeration cycle, a multistage ethane or ethylene refrigeration cycle, and an open-end methane refrigeration cycle which utilizes a portion of the feed gas as a source of methane and which includes therein a multistage expansion cycle to further cool the same and reduce the pressure to near-atmospheric pressure. In the sequence of cooling cycles, the refrigerant having the highest boiling point is utilized first followed by a refrigerant having an intermediate boiling point and finally by a refrigerant having the lowest boiling point. As used herein, the terms “upstream” and “downstream” shall be used to describe the relative positions of various components of an LNG facility along the main flow path of natural gas through the facility.

[0024] Various pretreatment steps can be provided to remove certain undesirable components, such as acid gases, mercaptans, mercury, and moisture from the natural gas feed stream delivered to the LNG facility. The composition of this gas stream may vary significantly. As used herein, a natural gas stream is any stream principally comprised of methane which originates in major portion from a natural gas feed stream, such feed stream for example containing at least 85 mole percent methane, with the balance being ethane, higher hydrocarbons, nitrogen, carbon dioxide, and a minor amount of other contaminants such as mercury, hydrogen sulfide, and mercaptan. The pretreatment steps may be separate steps located either upstream of the cooling cycles or located downstream of one of the early stages of cooling in the initial cycle. The following is a non-inclusive

listing of some of the available means which are readily known to one skilled in the art. Acid gases and to a lesser extent mercaptans are routinely removed via a chemical reaction process employing an aqueous amine-bearing solution. This treatment step is generally performed upstream of the cooling stages in the initial cycle. A major portion of the water is routinely removed as a liquid via two-phase gas-liquid separation following gas compression and cooling upstream of the initial cooling cycle and also downstream of the first cooling stage in the initial cooling cycle. Mercury is routinely removed via mercury sorbent beds. Residual amounts of water and acid gases are routinely removed via the use of properly selected sorbent beds such as regenerable molecular sieves.

[0025] The pretreated natural gas feed stream is generally delivered to the liquefaction process at an elevated pressure or is compressed to an elevated pressure generally greater than 500 psia, preferably about 500 psia to about 3,000 psia, still more preferably about 500 psia to about 1,000 psia, still yet more preferably about 600 psia to about 800 psia. The feed stream temperature is typically near ambient to slightly above ambient. A representative temperature range being 60 to 150° F. (10-65° C.).

[0026] As previously noted, the natural gas feed stream is cooled in a plurality of multistage refrigeration cycles (preferably three) by indirect heat exchange with a plurality of different refrigerants (preferably three). The overall cooling efficiency for a given cycle improves as the number of stages increases but this increase in efficiency is accompanied by corresponding increases in net capital cost and process complexity. The feed gas is preferably passed through an effective number of refrigeration stages, nominally two, preferably two to four, and more preferably three stages, in a first closed refrigeration cycle utilizing a relatively high boiling refrigerant. Such relatively high boiling point refrigerant is preferably comprised in major portion of propane, propylene, carbon dioxide, or mixtures thereof, more preferably the refrigerant comprises at least about 75 mole percent propane, even more preferably at least 90 mole percent propane, and most preferably the refrigerant consists essentially of propane. Thereafter, the processed feed gas flows through an effective number of stages, nominally two, preferably two to four, and more preferably two or three, in a second closed refrigeration cycle in heat exchange with a refrigerant having a lower boiling point. Such lower boiling point refrigerant is preferably comprised in major portion of ethane, ethylene, or mixtures thereof, more preferably the refrigerant comprises at least about 75 mole percent ethylene, even more preferably at least 90 mole percent ethylene, and most preferably the refrigerant consists essentially of ethylene. Each cooling stage comprises a separate cooling zone. As previously noted, the processed natural gas feed stream is preferably combined with one or more recycle streams (i.e., compressed open methane cycle gas streams) at various locations in the second cycle thereby producing a liquefaction stream. In the last stage of the second cooling cycle, the liquefaction stream is condensed (i.e., liquefied) in major portion, preferably in its entirety, thereby producing a pressurized LNG-bearing stream. Generally, the process pressure at this location is only slightly lower than the pressure of the pretreated feed gas to the first stage of the first cycle.

[0027] The pressurized LNG-bearing stream is then further cooled in a third refrigeration cycle referred to as the open methane cycle via contact in a main methane economizer (heat exchanger) with flash gases (i.e., flash gas streams) generated in this third cycle via sequential expansion of the pressurized LNG-bearing stream to near atmospheric pressure. The flash gases used as a refrigerant in the third refrigeration cycle are preferably comprised in major portion of methane, more preferably the flash gas refrigerant comprises at least 75 mole percent methane, still more preferably at least 90 mole percent methane, and most preferably the refrigerant consists essentially of methane. During expansion of the pressurized LNG-bearing stream to near atmospheric pressure, the pressurized LNG-bearing stream is cooled via at least one, preferably two to four, and more preferably three expansions where each expansion employs an expander as a pressure reduction means. Suitable expanders include, for example, either Joule-Thomson expansion valves or hydraulic expanders. The expansion is followed by a separation of the gas-liquid product with a separator. When a hydraulic expander is employed and properly operated, the greater efficiencies associated with the recovery of power, a greater reduction in stream temperature, and the production of less vapor during the flash expansion step will frequently more than off-set the higher capital and operating costs associated with the expander. In one embodiment, additional cooling of the pressurized LNG-bearing stream prior to flashing is made possible by first flashing a portion of this stream via one or more hydraulic expanders and then via indirect heat exchange means employing said flash gas stream to cool the remaining portion of the pressurized LNG-bearing stream prior to flashing. The warmed flash gas stream is then recycled via return to an appropriate location, based on temperature and pressure considerations, in the open methane cycle and will be recompressed.

[0028] Generally, the natural gas feed stream fed to the LNG facility will contain such quantities of C₃₊ components so as to result in the formation of a C₃₊ rich liquid in one or more of the cooling stages. This liquid can be removed via gas-liquid separation means. Generally, the sequential cooling of the natural gas in each stage is controlled so as to remove as much of the C₃ and higher molecular weight hydrocarbons as possible from the gas to produce a gas stream predominating in methane and a liquid stream containing significant amounts of ethane and heavier components. An effective number of gas/liquid separation means are located at strategic locations downstream of the cooling zones for the removal of liquids streams rich in C₃₊ components. The exact locations and number of gas/liquid separation means will be dependant on a number of operating parameters, such as the C₃₊ composition of the natural gas feed stream, the desired BTU content of the LNG product, the value of the C₃₊ components for other applications, and other factors routinely considered by those skilled in the art of LNG plant and gas plant operation. The C₃₊ hydrocarbon stream or streams may be demethanized via a silPGe stage flash or a fractionation column. In the latter case, the resulting methane-rich stream can be directly returned at pressure to the liquefaction process. In the former case, this methane-rich stream can be repressurized and recycled or can be used as fuel gas. The C₃₊ hydrocarbon stream or streams or the demethanized C₃₊ hydrocarbon stream may be used as fuel or may be further processed, such as by

fractionation in one or more fractionation zones to produce individual streams rich in specific chemical constituents (e.g., C₃, C₄, and C₅₊).

[0029] The liquefaction process described herein may use one of several types of cooling which include but are not limited to (a) indirect heat exchange, (b) vaporization, and (c) expansion or pressure reduction. Indirect heat exchange, as used herein, refers to a process wherein the refrigerant cools the substance to be cooled without actual physical contact between the refrigerating agent and the substance to be cooled. Specific examples of indirect heat exchange means include heat exchange undergone in a shell-and-tube heat exchanger, a core-in-kettle heat exchanger, and a brazed aluminum plate-fin heat exchanger. The physical state of the refrigerant and substance to be cooled can vary depending on the demands of the system and the type of heat exchanger chosen. Thus, a shell-and-tube heat exchanger will typically be utilized, where the refrigerating agent is in a liquid state and the substance to be cooled is in a liquid or gaseous state or when one of the substances undergoes a phase change and process conditions do not favor the use of a core-in-kettle heat exchanger. As an example, aluminum and aluminum alloys are preferred materials of construction for the core but such materials may not be suitable for use at the designated process conditions. A plate-fin heat exchanger will typically be utilized where the refrigerant is in a gaseous state and the substance to be cooled is in a liquid or gaseous state. Finally, the core-in-kettle heat exchanger will typically be utilized where the substance to be cooled is liquid or gas and the refrigerant undergoes a phase change from a liquid state to a gaseous state during the heat exchange.

[0030] Vaporization cooling refers to the cooling of a substance by the evaporation or vaporization of a portion of the substance with the system maintained at a constant pressure. Thus, during the vaporization, the portion of the substance which evaporates absorbs heat from the portion of the substance which remains in a liquid state and hence, cools the liquid portion. Finally, expansion or pressure reduction cooling refers to cooling which occurs when the pressure of a gas, liquid or a two-phase system is decreased by passing through a pressure reduction means. In one embodiment, this expansion means is a Joule-Thomson expansion valve. In another embodiment, the expansion means is either a hydraulic or gas expander. Because expanders recover work energy from the expansion process, lower process stream temperatures are possible upon expansion.

[0031] The flow schematics set forth in FIGS. 1-4 represent preferred embodiments of an inventive LNG facility providing enhanced heavies removal and LPG recovery. Those skilled in the art will recognize that FIGS. 1-4 are schematics only and, therefore, many items of equipment that would be needed in a commercial plant for successful operation have been omitted for the sake of clarity. Such items might include, for example, compressor controls, flow and level measurements and corresponding controllers, temperature and pressure controls, pumps, motors, filters, additional heat exchangers, and valves, etc. These items would be provided in accordance with standard engineering practice.

[0032] To facilitate an understanding of FIGS. 1-4, the following numbering nomenclature was employed. Items

numbered 1 through 99 are process vessels and equipment which are associated with the liquefaction process. Items numbered 100 through 199 correspond to flow lines or conduits which contain predominantly methane streams. Items numbered 200 through 299 correspond to flow lines or conduits which contain predominantly ethylene streams. Items numbered 300 through 399 correspond to flow lines or conduits which contain predominantly propane streams.

[0033] Referring to FIG. 1, gaseous propane is compressed in a multistage (preferably three-stage) compressor 18 driven by a gas turbine driver (not illustrated). The three stages of compression preferably exist in a siLPGe unit although each stage of compression may be a separate unit and the units mechanically coupled to be driven by a siLPGe driver or combination of drivers. Upon compression, the compressed propane is passed through conduit 300 to a cooler 20 where it is cooled and liquefied. A representative pressure and temperature of the liquefied propane refrigerant prior to flashing is about 100° F. (38° C.) and about 190 psia. The stream from cooler 20 is passed through conduit 302 to a pressure reduction means, illustrated as expansion valve 12, wherein the pressure of the liquefied propane is reduced, thereby evaporating or flashing a portion thereof. The resulting two-phase product then flows through conduit 304 into a high-stage propane chiller 2 wherein gaseous methane refrigerant introduced via conduit 152, natural gas feed introduced via conduit 100, and gaseous ethylene refrigerant introduced via conduit 202 are respectively cooled via indirect heat exchange means 4, 6, and 8, thereby producing cooled gas streams respectively produced via conduits 154, 102, and 204.

[0034] The vaporized propane gas from chiller 2 is returned to the high-stage inlet port of compressor 18 through conduit 306. The remaining liquid propane is passed through conduit 308, the pressure further reduced by passage through a pressure reduction means, illustrated as expansion valve 14, whereupon an additional portion of the liquefied propane is flashed. The resulting two-phase stream is then fed to an intermediate-stage propane chiller 22 through conduit 310, thereby providing a coolant for chiller 22. The cooled feed gas stream from chiller 2 flows via conduit 102 to separation equipment 10 wherein gas and liquid phases are separated. The liquid phase, which can be rich in C₃₊ components, is removed via conduit 103. The gaseous phase is removed via conduit 104 and fed to propane chiller 22. Ethylene refrigerant from chiller 2 is introduced to chiller 22 via conduit 204.

[0035] In chiller 22, the feed gas stream, also referred to herein as a methane-rich stream, and the ethylene refrigerant streams are respectively cooled via indirect heat transfer means 24 and 26, thereby producing cooled methane-rich and ethylene refrigerant streams via conduits 110 and 206. The vaporized portion of the propane refrigerant in chiller 22 is separated and passed through conduit 311 to the intermediate-stage inlet of compressor 18. Liquid propane refrigerant from chiller 22 is removed via conduit 314, flashed across a pressure reduction means, illustrated as expansion valve 16, and then fed to a low-stage propane chiller/condenser 28 via conduit 316.

[0036] As illustrated in FIG. 1, the methane-rich stream flows from intermediate-stage propane chiller 22 to the low-stage propane chiller 28 via conduit 110. In chiller 28,

the methane-rich stream is cooled via indirect heat exchange means 30. In a like manner, the ethylene refrigerant stream flows from the intermediate-stage propane chiller 22 to low-stage propane chiller 28 via conduit 206. In the latter, the ethylene refrigerant is totally condensed or condensed in nearly its entirety via indirect heat exchange means 32. The vaporized propane is removed from low-stage propane chiller 28 and returned to the low-stage inlet of compressor 18 via conduit 320.

[0037] As illustrated in FIG. 1, the methane-rich stream exiting low-stage propane chiller 28 is introduced into high-stage ethylene chiller 42 via conduit 112. Ethylene refrigerant exits low-stage propane chiller 28 via conduit 208 and is preferably fed to a separation vessel 37 wherein light components are removed via conduit 209 and condensed ethylene is removed via conduit 210. The ethylene refrigerant at this location in the process is generally at a temperature of about -24° F. (-31° C.) and a pressure of about 285 psia. The ethylene refrigerant then flows to an ethylene economizer 34 wherein it is cooled via indirect heat exchange means 38, removed via conduit 211, and passed to a pressure reduction means, illustrated as an expansion valve 40, whereupon the refrigerant is flashed to a preselected temperature and pressure and fed to high-stage ethylene chiller 42 via conduit 212. Vaporized ethylene is removed from chiller 42 via conduit 214 and routed to ethylene economizer 34 wherein the vapor functions as a coolant via indirect heat exchange means 46. The ethylene vapor is then removed from ethylene economizer 34 via conduit 216 and fed to the high-stage inlet of ethylene compressor 48. The ethylene refrigerant which is not vaporized in high-stage ethylene chiller 42 is removed via conduit 218 and returned to ethylene economizer 34 for further cooling via indirect heat exchange means 50, removed from ethylene economizer via conduit 220, and flashed in a pressure reduction means, illustrated as expansion valve 52, whereupon the resulting two-phase product is introduced into a low-stage ethylene chiller 54 via conduit 222.

[0038] The methane-rich stream introduced into high-stage ethylene chiller 42 via conduit 112 is cooled in indirect heat exchange means 44 and subsequently removed from high-stage ethylene chiller 42 via conduit 116. The stream in conduit 116 is then carried to a low-stage ethylene chiller 54 wherein the methane-rich stream is cooled and partially condensed in indirect heat exchange means 56. The cooled and partially condensed methane-rich stream exiting low-stage ethylene chiller 54 is directed to a phase separator 57 via conduit 160 for separation of the vapor and liquid fractions. Phase separator 57 can be any equipment known in the art that is capable of separating a stream containing vapor and liquid fractions into two distinct vapor and liquid streams. A predominately vapor stream exits an upper vapor outlet of phase separator 57. This predominately vapor stream is carried to expander 59 via conduit 162. In expander 59, the predominately vapor stream is expanded and partially condensed to thereby form an expanded stream that is transported to a first distillation column 60 via conduit 169. In a preferred embodiment of the present invention, expander 59 is a hydraulic expander providing substantially isentropic expansion of the predominately vapor stream introduced therein. The expanded stream in conduit 169 preferably has a vapor fraction greater than about 0.7 on a molar basis. More preferably the vapor fraction of the expanded stream in conduit 169 is in the range of from about

0.8 to about 0.995, still more preferably in the range of from about 0.9 to about 0.99, and most preferably in the range of from 0.95 to 0.985 on a molar basis.

[0039] First distillation column 60 is preferably equipped with internals, such as trays and/or packing, disposed between a lower section of column 60, where the predominately vapor expanded feedstream is introduced, and an upper section of column 60, where a liquid reflux stream is introduced as described in further detail below. First distillation column 60 preferably provides at least five theoretical stages, more preferably at least seven theoretical stages, and most preferably in the range of from 9 to 20 theoretical stages. It is preferred for the expanded stream in conduit 169 to be introduced into first distillation column 60 at a location near the bottom of column 60. In particular, it is preferred for a predominate portion of the expanded stream to be introduced into first distillation column 60 in or below one or more of the bottom three theoretical stages, more preferably in or below one or more of the bottom two theoretical stages, and most preferably in or below the bottom theoretical stage.

[0040] First distillation column 60 is operable to separate the expanded stream introduced via conduit 169 into a first relatively more volatile overhead vapor fraction produced via conduit 119 and a first relatively less volatile bottom liquid fraction produced via conduit 166. Typically, the relatively more volatile overhead vapor fraction contains primarily methane (preferably >than 85 mole % methane), while the relatively less volatile bottom liquid fraction contains primarily C_{2+} hydrocarbons. The methane-rich overhead vapor fraction in conduit 119 is then combined with a condensed flash gas stream in conduit 155, described in further detail below, and the combined stream is carried to an ethylene condenser 68 via conduit 120. The combined stream is cooled in an indirect heat exchange means 70 of ethylene condenser 68 to thereby produce a pressurized LNG-bearing stream which is produced via conduit 122. Further processing of the LNG-bearing stream in conduit 122 is discussed in detail below.

[0041] The first relatively less volatile bottom liquid stream exiting first distillation column 60 via conduit 166 is preferably combined with the predominately liquid stream exiting the bottom outlet of phase separator 57 via conduit 168. The combined liquid stream is then conducted to a reflux condenser 63 via conduit 170. In condenser 63, the combined stream from conduit 170 acts as a coolant as it passes through indirect heat exchange means 64. Thus, as the stream passes through indirect heat exchange means 64 it cools the stream in indirect heat exchange means 67 by indirect heat exchange. The cooling provided by the stream in heat exchange means 64 warms the combined stream and, preferably, causes at least a portion of the combined stream to vaporize. The warmed combined stream is removed from reflux condenser 63 via conduit 174, which carries the warmed combined stream to an inlet of a second distillation column 65.

[0042] Second distillation column 65 is operable to separate the combined stream from conduit 174 into a second relatively more volatile overhead vapor fraction produced via conduit 176 and a second relatively less volatile bottom liquid fraction produced via conduit 400. The less volatile bottom liquid stream in conduit 400 preferably contains

predominately C₂₊ hydrocarbons. The stream in conduit 400 can be subjected to further separation to produce LPG and/or various individual hydrocarbon components.

[0043] Second distillation column 65 includes an upper rectification section disposed above the inlet of the combined stream from conduit 174 and a lower reboiled section disposed below the inlet of the combined stream from conduit 174. Second distillation column 65 preferably includes upper internals, such as packing and/or trays, disposed in the upper section of the column and lower internals, such as packing and/or trays, disposed in the lower section of the column. Second distillation column 65 is preferably equipped with a reboiler 66 which removes a portion of the liquid in the lower section of column 65, vaporizes the removed liquid, and reintroduces the vaporized reboil stream into the lower section of column 65 at a location below the location where the liquid stream was removed. Further, second distillation column 65 includes an upper inlet for receiving a liquid reflux stream introduced via conduit 184, as described in further detail below.

[0044] It is preferred for second distillation column 65 to be operated at a pressure that is less than the operating pressure of first distillation column 60. Preferably, the operating pressure of second distillation column 65 is at least about 25 psi less than the operating pressure of first distillation column 60. Most preferably, the operating pressure of second distillation column 65 is at least 100 psi less than the operating pressure of first distillation column 60.

[0045] The second relatively more volatile overhead vapor fraction produced from second distillation column 65 is transported to reflux condenser 63 via conduit 176. As described above, the second relatively more volatile fraction is cooled in indirect heat exchanges means 67 via indirect heat exchange with the combined stream passing through indirect heat exchange means 64. This cooling in indirect heat exchange means 67 preferably condenses a predominate portion of the second relatively more volatile fraction. Conduit 178 removes the cooled second relatively more volatile fraction from reflux condenser 63 and carries it to a reflux accumulator vessel 69. Reflux accumulator vessel 69 can simply be a substantially empty tank having a sufficient volume to account for fluctuations in the rate of fluids supplied thereto and fluids withdrawn therefrom.

[0046] A liquid reflux portion is preferably withdrawn from the bottom of reflux accumulator vessel 69 via conduit 180. This liquid reflux portion is then pumped, via cryogenic pump 73, to both the first and second distillation columns 60, 65 for use as liquid reflux streams. The liquid reflux portion is discharged from pump 73 via conduit 182 and subsequently split into a first reflux stream, which is carried to final ethylene chiller 68 via conduit 186, and a second reflux stream, which is carried to the upper section of second distillation column 65 via conduit 184. The second reflux stream in conduit 184 is introduced directly into the upper section of second distillation column 65 as a liquid reflux stream. The first reflux stream in conduit 186 undergoes further cooling in indirect heat exchange means 75 of ethylene chiller 68 prior to introduction into an upper section of first distillation column 60, via conduit 188, as a liquid reflux stream.

[0047] A vapor portion is withdrawn from reflux accumulator vessel 69 via conduit 190. This vapor portion is

transported to main methane economizer 74 and cooled in indirect heat exchange means 61 via indirect heat exchange with the flash gas streams described in further detail below. The resulting cooled stream is withdrawn from methane economizer 74 via conduit 192 and thereafter introduced into a compressor 62. The resulting compressed stream is removed from compressor 62 via conduit 194 and subsequently combined with a compressed predominately-methane stream exiting main methane compressor 83 via conduit 150, as described in further detail below. In one embodiment of the present invention, compressor 62 is powered by work generated in hydraulic expander 59. Compressor 62 and hydraulic expander 59 can be directly mechanically coupled to one another. Alternatively, hydraulic expander 59 can drive a generator (not shown) which provides electricity to a motor (not shown) for driving compressor 62.

[0048] The gas exiting high-stage propane chiller 2 via conduit 154 is fed to main methane economizer 74 wherein the stream is cooled via indirect heat exchange means 97. The stream cooled in heat exchange mean 97 is removed from methane economizer 74 via conduit 155 and combined with the relatively more volatile overhead stream from first distillation column 60 flowing in conduit 119. The resulting combined stream is fed to ethylene chiller 68 via conduit 120. In ethylene chiller 68, this methane-rich stream is cooled and condensed via indirect heat exchange means 70 with the liquid effluent from low-stage ethylene chiller 54, which is routed to ethylene chiller 68 via conduit 226. The vaporized ethylene from low-stage ethylene chiller 54, withdrawn via conduit 224, and ethylene chiller 68, withdrawn via conduit 228, are combined and routed, via conduit 230, to ethylene economizer 34 wherein the vapors function as a coolant via indirect heat exchange means 58. The stream is then routed via conduit 232 from ethylene economizer 34 to the low-stage inlet of ethylene compressor 48.

[0049] As illustrated in FIG. 1, the compressor effluent from vapor introduced via the low-stage side of ethylene compressor 48 is removed via conduit 234, cooled via inter-stage cooler 71, and returned to compressor 48 via conduit 236 for injection with the high-stage stream present in conduit 216. Preferably, the two-stages are a siLPGe module although they may each be a separate module and the modules mechanically coupled to a common driver. The compressed ethylene product from compressor 48 is routed to a downstream cooler 72 via conduit 200. The product from cooler 72 flows via conduit 202 and is introduced, as previously discussed, to high-stage propane chiller 2.

[0050] The pressurized LNG-bearing stream, preferably a liquid stream in its entirety, in conduit 122 is preferably at a temperature in the range of from about -200 to about -50° F. (-130° C. to -45° C.), more preferably in the range of from about -175 to about -100° F. (-115° C. to -73° C.), most preferably in the range of from -150 to -125° F. (-100° C. to -85° C.). The pressure of the stream in conduit 122 is preferably in the range of from about 500 to about 700 psia, most preferably in the range of from 550 to 725 psia. The stream in conduit 122 is directed to main methane economizer 74 wherein the stream is further cooled by indirect heat exchange means/heat exchanger pass 76 as hereinafter explained. It is preferred for main methane economizer 74 to include a plurality of heat exchanger passes which provide for the indirect exchange of heat between various predominantly methane streams in the

economizer 74. Preferably, methane economizer 74 comprises one or more plate-fin heat exchangers. The cooled stream from heat exchanger pass 76 exits methane economizer 74 via conduit 124. The pressure of the stream in conduit 124 is then reduced by a pressure reduction means, illustrated as expansion valve 78, which evaporates or flashes a portion of the gas stream thereby generating a two-phase stream. The two-phase stream from expansion valve 78 is then passed to high-stage methane flash drum 80 where it is separated into a flash gas stream discharged through conduit 126 and a liquid phase stream (i.e., pressurized LNG-bearing stream) discharged through conduit 130. The flash gas stream is then transferred to main methane economizer 74 via conduit 126 wherein the stream functions as a coolant in heat exchanger pass 82. The predominantly methane stream is warmed in heat exchanger pass 82, at least in part, by indirect heat exchange with the predominantly methane stream in heat exchanger pass 76. The warmed stream exits heat exchanger pass 82 and methane economizer 74 via conduit 128.

[0051] The liquid-phase stream exiting high-stage flash drum 80 via conduit 130 is passed through a second methane economizer 87 wherein the liquid is further cooled by downstream flash vapors via indirect heat exchange means 88. The cooled liquid exits second methane economizer 87 via conduit 132 and is expanded or flashed via pressure reduction means, illustrated as expansion valve 91, to further reduce the pressure and, at the same time, vaporize a second portion thereof. This two-phase stream is then passed to an intermediate-stage methane flash drum 92 where the stream is separated into a gas phase passing through conduit 136 and a liquid phase passing through conduit 134. The gas phase flows through conduit 136 to second methane economizer 87 wherein the vapor cools the liquid introduced to economizer 87 via conduit 130 via indirect heat exchanger means 89. Conduit 138 serves as a flow conduit between indirect heat exchange means 89 in second methane economizer 87 and heat exchanger pass 95 in main methane economizer 74. The warmed vapor stream from heat exchanger pass 95 exits main methane economizer 74 via conduit 140 and is conducted to the intermediate-stage inlet of methane compressor 83.

[0052] The liquid phase exiting intermediate-stage flash drum 92 via conduit 134 is further reduced in pressure by passage through a pressure reduction means, illustrated as an expansion valve 93. Again, a third portion of the liquefied gas is evaporated or flashed. The two-phase stream from expansion valve 93 is passed to a final or low-stage flash drum 94. In flash drum 94, a vapor phase is separated and passed through conduit 144 to second methane economizer 87 wherein the vapor functions as a coolant via indirect heat exchange means 90, exits second methane economizer 87 via conduit 146, which is connected to the first methane economizer 74 wherein the vapor functions as a coolant via heat exchanger pass 96. The warmed vapor stream from heat exchanger pass 96 exits main methane economizer 74 via conduit 148 and is conducted to the low-stage inlet of compressor 83. The liquid stream exiting the bottom of low-stage flash drum 94 via conduit 142 is liquefied natural gas (LNG) at approximately atmospheric pressure. The resulting LNG can then be stored and/or transported, and subsequently vaporized as needed for use as gaseous natural gas.

[0053] As shown in FIG. 1, the high, intermediate, and low stages of compressor 83 are preferably combined as a siLPGe unit. However, each stage may exist as a separate unit where the units are mechanically coupled together to be driven by a siLPGe driver. The compressed gas from the low-stage section passes through an inter-stage cooler 85 and is combined with the intermediate pressure gas in conduit 140 prior to the second-stage of compression. The compressed gas from the intermediate stage of compressor 83 is passed through an inter-stage cooler 84 and is combined with the high pressure gas provided via conduit 128 prior to the third-stage of compression. The compressed gas (i.e., compressed open methane cycle gas stream) is discharged from high stage methane compressor through conduit 150, is cooled in cooler 86, and is routed to the high pressure propane chiller 2 via conduit 152 as previously discussed.

[0054] FIG. 2 illustrates a first alternative embodiment of the present invention. The LNG facility shown in FIG. 2 is very similar to the facility illustrated in FIG. 1. Therefore, common components of FIG. 1 and FIG. 2 are identified with the same reference numerals, and the written description of these components provided above with reference to FIG. 1 also applies to FIG. 2. There are two main differences between the embodiment illustrated in FIG. 2 and the embodiment illustrated in FIG. 1. The first difference is the substitution of an expansion valve 77 (FIG. 2) for hydraulic expander 59 (FIG. 1) between the upper outlet of phase separator 57 and the inlet of first distillation column 60. The second difference is that in the embodiment of FIG. 2, reflux condenser 63 is modified to include an additional indirect heat exchange means 79. This additional heat exchange pass 79 allows the separated liquid stream exiting the bottom of phase separator 57 via conduit 170 and the first relatively less volatile bottom liquid fraction exiting first distillation column 60 via conduit 196 to function as coolants in separate heat exchange means 64 and 79. After use as separate coolants in reflux condenser 63, the warmed streams from indirect heat exchange means 64 and 79 can be combined and routed to the inlet of second distillation column 65 via conduit 174. The use of two separate heat exchange means 64 and 79 can improve the efficiency of reflux condenser 63.

[0055] FIG. 3 illustrates a second alternative embodiment of the present invention. The LNG facility shown in FIG. 3 is very similar to the facilities illustrated in FIGS. 1 and 2. Therefore, common components of FIGS. 1-3 are identified with the same reference numerals, and the written description of these components provided above with reference to FIGS. 1 and 2 also applies to FIG. 3. The main difference between the embodiment illustrated in FIG. 3 and the embodiments illustrated in FIGS. 1 and 2 deals with the treatment of the second relatively more volatile overhead vapor fraction from second distillation column 65. In the embodiment of FIG. 3, the second relatively more volatile overhead stream is totally condensed prior to introduction into reflux accumulator vessel 69. The additional refrigeration duty required to totally condense the second relatively more volatile overhead stream from second distillation column 65 is provided by additional indirect heat exchange means 98 and 99, which are added to high and low stage ethylene chillers 42 and 54, respectively. Thus, in the embodiment of FIG. 3, conduit 105 carries the second relatively more volatile overhead stream from second distillation column 65 to high-stage ethylene chiller for cooling

in indirect heat exchange means 98. The resulting cooled stream is removed from high-stage ethylene chiller 44 and transported to reflux condenser 63 via conduit 106 for cooling in indirect heat exchange means 67. The cooled stream from indirect heat exchange means 67 of reflux condenser 63 is transported to low-stage ethylene chiller 54 via conduit 107 for further cooling and condensing in indirect heat exchange means 99. The resulting condensed stream is then transported to reflux accumulator vessel 69 via conduit 108 and the entire stream is subsequently employed as reflux streams to first and second distillation columns 60,65.

[0056] FIG. 4 illustrates a third alternative embodiment of the present invention. The LNG facility shown in FIG. 4 is very similar to the facilities illustrated in FIGS. 1-3. Therefore, common components of FIGS. 1-4 are identified with the same reference numerals, and the written description of these components provided above with reference to FIGS. 1-3 also applies to FIG. 4. The main difference between the embodiment illustrated in FIG. 4 and the embodiments illustrated in FIGS. 1-3 concerns the way that the compressed and cooled flash gas stream in conduit 155 is processed. In FIG. 1, the stream in conduit 155 is simply combined with the first overhead fraction from first distillation column 60 in conduit 119 and then the combined stream in conduit 120 is subsequently subjected to further cooling and expansion. In the embodiment of FIG. 4, the stream in conduit 155 is kept separate from the first overhead fraction in conduit 119 during cooling in ethylene chiller 68, cooling in methane economizer 74, and flashing upstream of high-stage flash drum 80. Thus, the embodiment illustrated in FIG. 4 includes an additional heat exchange means 11 in ethylene chiller 68, an additional indirect heat exchange means 13 in methane economizer 74, and an additional expansion means 15 upstream of high-stage flash drum 80. It is preferred for ethylene chiller 68 to be a core-in-kettle heat exchanger, wherein heat exchange means 75, 70, and 11 are separate cores.

[0057] In the embodiment of FIG. 4, the compressed and cooled flash gas stream in conduit 155 is cooled in indirect heat exchange means 11 of ethylene chiller 68. The resulting cooled stream is removed from ethylene chiller 68 in conduit 109 and transported to main methane economizer 74 for further cooling in indirect heat exchange means 13. In indirect heat exchange means 13, the stream is cooled via indirect heat exchange with the flash gas streams in indirect heat exchange means 82,95,96. The resulting cooled stream exits methane economizer and is carried to expansion means 15 via conduit 111. In expansion means 15, the condensed stream is flashed, and the resulting stream is introduced into high-stage flash drum 80 via an inlet that is separate from the inlet for the flashed stream from expansion means 78. Enhanced efficiency, control, and/or operability can be provided by keeping the main predominately-methane stream and the compressed flash gas streams separate during cooling in ethylene chiller 68 and methane economizer 74.

[0058] In one embodiment of the present invention, the LNG production systems illustrated in FIGS. 1-4 are simulated on a computer using conventional process simulation software. Examples of suitable simulation software include HYSYS™ from Hyprotech, Aspen Plus® from Aspen Technology, Inc., and PRO/II® from Simulation Sciences Inc.

[0059] The preferred forms of the invention described above are to be used as illustration only, and should not be used in a limiting sense to interpret the scope of the present invention. Obvious modifications to the exemplary embodiments, set forth above, could be readily made by those skilled in the art without departing from the spirit of the present invention.

[0060] The inventors hereby state their intent to rely on the Doctrine of Equivalents to determine and assess the reasonably fair scope of the present invention as pertains to any apparatus not materially departing from but outside the literal scope of the invention as set forth in the following claims.

What is claimed is:

1. A process for liquefying a natural gas stream, said process comprising:
 - (a) using a first distillation column to separate at least a portion of said natural gas stream into a first relatively more volatile fraction and a first relatively less volatile fraction;
 - (b) using a second distillation column to separate at least a portion of said first relatively less volatile fraction into a second relatively more volatile fraction and a second relatively less volatile fraction, said second distillation column operating at a lower pressure than said first distillation column;
 - (c) using a reflux portion of said second relatively more volatile fraction as reflux in said first and/or second distillation columns; and
 - (d) cooling at least a portion of said reflux portion via indirect heat exchange with at least a portion of said first relatively less volatile fraction.
2. The process of claim 1,

said second distillation column being operated at a pressure at least about 25 psi less than the operating pressure of said first distillation column.
3. The process of claim 1,

said second distillation column being operated at a pressure at least about 100 psi less than the operating pressure of said first distillation column.
4. The process of claim 1,

said at least a portion of said natural gas stream separated in said first distillation column having a vapor fraction greater than about 0.7 on a molar basis when introduced into said first distillation column.
5. The process of claim 1; and
 - (e) prior to step (a), condensing a portion of said natural gas stream;
 - (f) prior to step (a) and subsequent to step (e), separating said natural gas stream into a predominately vapor stream and a predominately liquid stream; and
 - (g) introducing at least a portion of said predominately vapor stream into said first distillation column.
6. The process of claim 5,

said first distillation column having at least 5 theoretical stages,

said at least a portion of said predominately vapor stream introduced into said first distillation column in accordance with step (g) entering said first distillation column in or below one or more of the bottom 3 theoretical stages of said first distillation column.

7. The process of claim 5; and

(h) expanding said predominately vapor stream prior to introduction into said first distillation column,

said expanding of step (h) causing at least a portion of said predominately vapor stream to condense.

8. The process of claim 5; and

(i) combining at least a portion of said first relatively less volatile fraction with at least a portion of said predominately liquid stream to thereby form a combined stream; and

(j) introducing at least a portion of said combined stream into said second distillation column.

9. The process of claim 8,

said combining of step (i) occurring subsequent to using said first relatively less volatile fraction for said cooling of step (d).

10. The process of claim 1; and

(k) separating said reflux portion into a first reflux stream and a second reflux stream;

(l) introducing said first reflux stream into an upper section of said first distillation column; and

(m) introducing said second reflux stream into an upper section of said second distillation column.

11. The process of claim 10,

said separating of step (k) including introducing at least a portion of said second relatively more volatile fraction into a reflux accumulator vessel, removing a liquid stream from said reflux accumulator, and removing a vapor stream from said reflux accumulator,

said liquid stream removed from said reflux accumulator being said reflux portion.

12. The process of claim 11; and

(n) combining at least a portion of said vapor stream removed from said reflux accumulator with at least a portion of said first relatively more volatile stream.

13. The process of claim 12; and

(o) cooling the combined stream resulting from step (n) in a methane refrigeration cycle employing a predominately methane refrigerant.

14. The process of claim 10; and

(p) prior to introduction into said second distillation column, cooling said second reflux stream via indirect heat exchange with a mechanical refrigeration cycle.

15. The process of claim 14,

said mechanical refrigeration cycle employing a predominately ethylene or ethane refrigerant.

16. The process of claim 1; and

(q) cooling at least a portion of said first relatively more volatile fraction in a methane refrigeration cycle employing a predominately methane refrigerant.

17. The process of claim 1,

said process being a cascade-type LNG process.

18. The process of claim 17,

said cascade-type LNG process including an open methane refrigeration cycle.

19. The process of claim 1; and

(r) vaporizing liquefied natural gas produced by the process of claim 1.

20. A liquefied natural gas product produced by the process of claim 1.

21. A computer-implemented simulation process comprising: using a computer to simulate the process of claim 1.

22. A process for liquefying a natural gas stream, said process comprising:

(a) using a vapor/liquid separator to separate at least a portion of said natural gas stream into a predominately vapor separated portion and a predominately liquid separated portion;

(b) using a first distillation column to separate at least a portion of said predominately vapor separated portion into a first relatively more volatile fraction and a first relatively less volatile fraction;

(c) introducing a first reflux stream into an upper section of said first distillation column; and

(d) cooling at least a portion of said first reflux stream via indirect heat exchange with at least a portion of said predominately liquid separated portion.

23. The process of claim 22; and

(e) using at least a portion of said first relatively less volatile fraction as said first reflux stream.

24. The process of claim 22; and

(f) using at least a portion of said first relatively more volatile fraction as said first reflux stream.

25. The process of claim 22; and

(g) cooling at least a portion of said first reflux stream via indirect heat exchange with at least a portion of said first relatively less volatile fraction.

26. The process of claim 25; and

(h) combining at least a portion of said predominately liquid separated portion with at least a portion of said first relatively less volatile fraction.

27. The process of claim 26,

said combining of step (f) occurring subsequent to using said predominately liquid separated stream and said first relatively less volatile fraction for said cooling of steps (d) and (g).

28. The process of claim 22; and

(i) using a second distillation column to separate at least a portion of said first relatively less volatile fraction into a second relatively more volatile fraction and a second relatively less volatile fraction.

29. The process of claim 28; and

(j) using at least a portion of said second relatively more volatile fraction as said first reflux stream.

30. The process of claim 28; and

(k) introducing a second reflux stream into an upper section of said second distillation column.

- 31.** The process of claim 30; and
- (l) using a first portion of said second relatively more volatile fraction as said first reflux stream and a second portion of said second relatively more volatile fraction as said second reflux stream.
- 32.** The process of claim 28,
said second distillation column being operated a pressure that is at least about 25 psi less than the operating pressure of said first distillation column.
- 33.** The process of claim 22; and
- (m) prior to step (a), condensing at least a portion of said natural gas stream;
- (n) prior to step (a), separating said natural gas stream into a predominately vapor stream and a predominately liquid stream; and
- (o) feeding at least a portion of said predominately vapor stream to said first distillation column.
- 34.** The process of claim 33; and
- (p) expanding said predominately vapor stream prior to introduction into said first distillation column,
said expanding of step (p) causing at least a portion of said predominately vapor stream to condense.
- 35.** The process of claim 22,
said process being a cascade-type LNG process.
- 36.** The process of claim 22; and
- (q) vaporizing liquefied natural gas produced by the process of claim 22.
- 37.** A liquefied natural gas product produced by the process of claim 22.
- 38.** A computer-implemented simulation process comprising: using a computer to simulate the process of claim 22.
- 39.** A process for liquefying a natural gas stream, said process comprising:
- (a) using a distillation column to separate at least a portion of said natural gas stream into a relatively more volatile fraction and a relatively less volatile fraction;
- (b) cooling at least a portion of said relatively more volatile fraction in a first heat exchange pass via indirect heat exchange with a first refrigerant;
- (c) flashing at least a portion of the cooled relatively more volatile fraction to thereby produce a flash gas;
- (d) compressing at least a portion of said flash gas; and
- (e) cooling at least a portion of the compressed flash gas in a second heat exchange pass via indirect heat exchange with a second refrigerant, said first and second heat exchange passes being separate from one another.
- 40.** The process of claim 39,
said first and second refrigerants comprising predominately propane, propylene, carbon dioxide, ethane, ethylene, or methane.
- 41.** The process of claim 39,
said first and second refrigerants comprising predominately ethylene or ethane.
- 42.** The process of claim 39,
said first and second refrigerants being the same refrigerant.
- 43.** The process of claim 39,
said cooling of steps (b) and (e) being carried out in a common heat exchanger.
- 44.** The process of claim 43,
said common heat exchanger being a core-in-kettle type heat exchanger,
said first heat exchange pass being defined by a first core of said common heat exchanger,
said second heat exchange pass being defined by a second core of said common heat exchanger.
- 45.** The process of claim 39,
- (f) prior to step (a), cooling at least a portion of said natural gas stream via indirect heat exchanger with a second refrigerant; and
- (g) prior to step (e) cooling at least a portion of said compressed flash gas via indirect heat exchange with said second refrigerant.
- 46.** The process of claim 45,
said second refrigerant comprising predominately propane, propylene, or carbon dioxide.
- 47.** An apparatus for liquefying natural gas, said apparatus comprising:
- a vapor/liquid separator having a vapor outlet and a liquid outlet;
- a first distillation column fluidly coupled to said vapor outlet and having a first overhead outlet and a first bottom outlet;
- a heat exchanger having a heating pass and a cooling pass, said heating and cooling passes being configured to facilitate indirect heat exchange between fluids flowing therethrough, said heating pass being fluidly coupled to said first bottom outlet; and
- a second distillation column fluidly coupled to said heating pass and having a second overhead outlet and a second bottom outlet, said second overhead outlet being fluidly coupled to said cooling pass.
- 48.** The apparatus of claim 47; and
an expander fluidly coupled between said vapor outlet and said first distillation column.
- 49.** The apparatus of claim 47,
said first distillation column including a first reflux inlet, said cooling pass being fluidly coupled between said first reflux inlet and said second overhead outlet.
- 50.** The apparatus of claim 49,
said second distillation column including a second reflux inlet,
said cooling pass being fluidly coupled between said second reflux inlet and said second overhead outlet.

51. The apparatus of claim 50; and
a reflux accumulator fluidly disposed between said cooling pass and said first and second reflux inlets.

52. The apparatus of claim 47,
said second heating pass being fluidly coupled to said first bottom outlet.

53. The apparatus of claim 47,
said heat exchanger having a second heating pass,
said second heating pass being fluidly coupled to said first bottom outlet.

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