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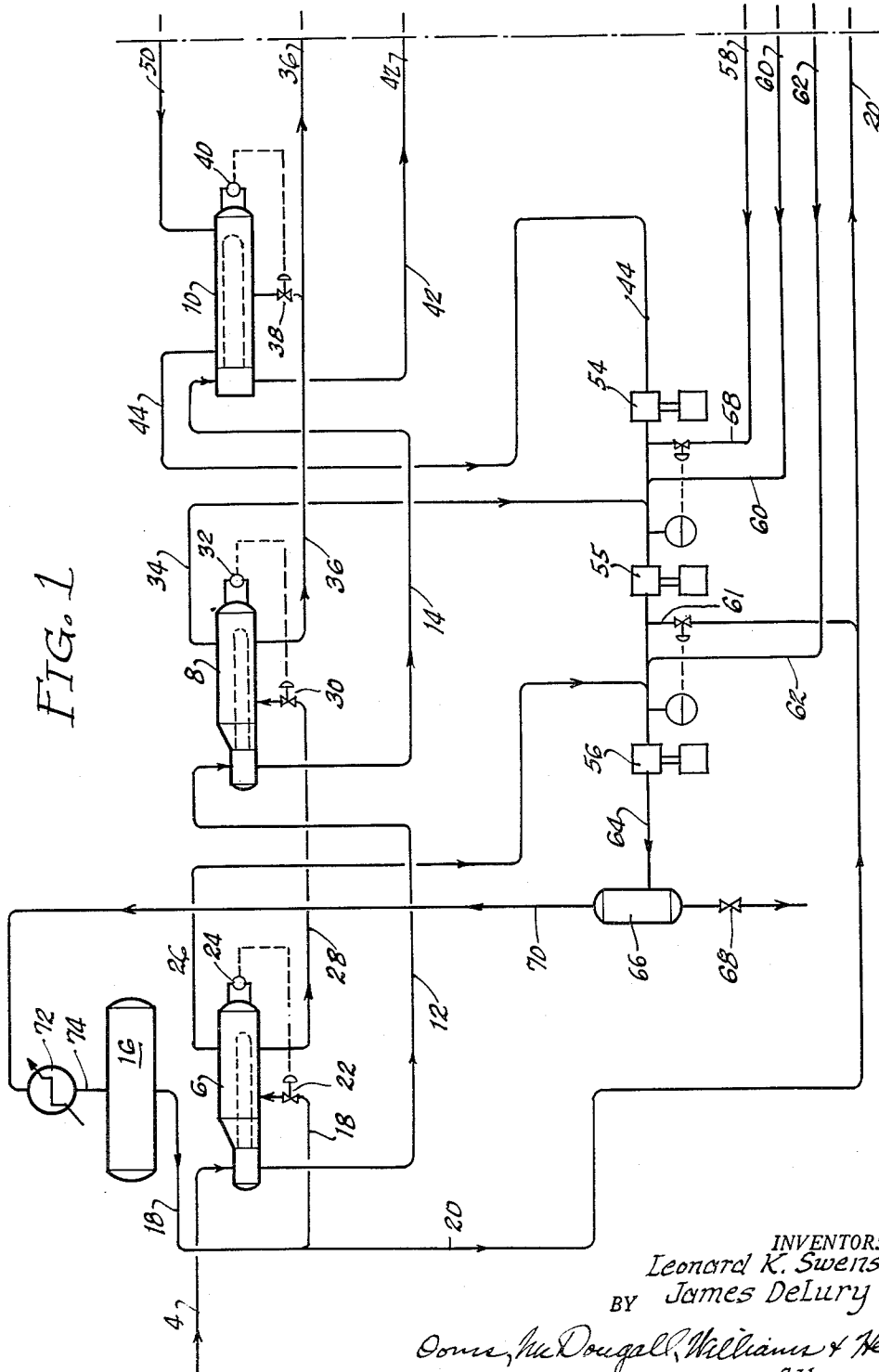
L. K. SWENSON ET AL

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LIQUEFYING NATURAL GAS WITH LOW PRESSURE REFRIGERANTS

Filed July 16, 1958

3 Sheets-Sheet 1



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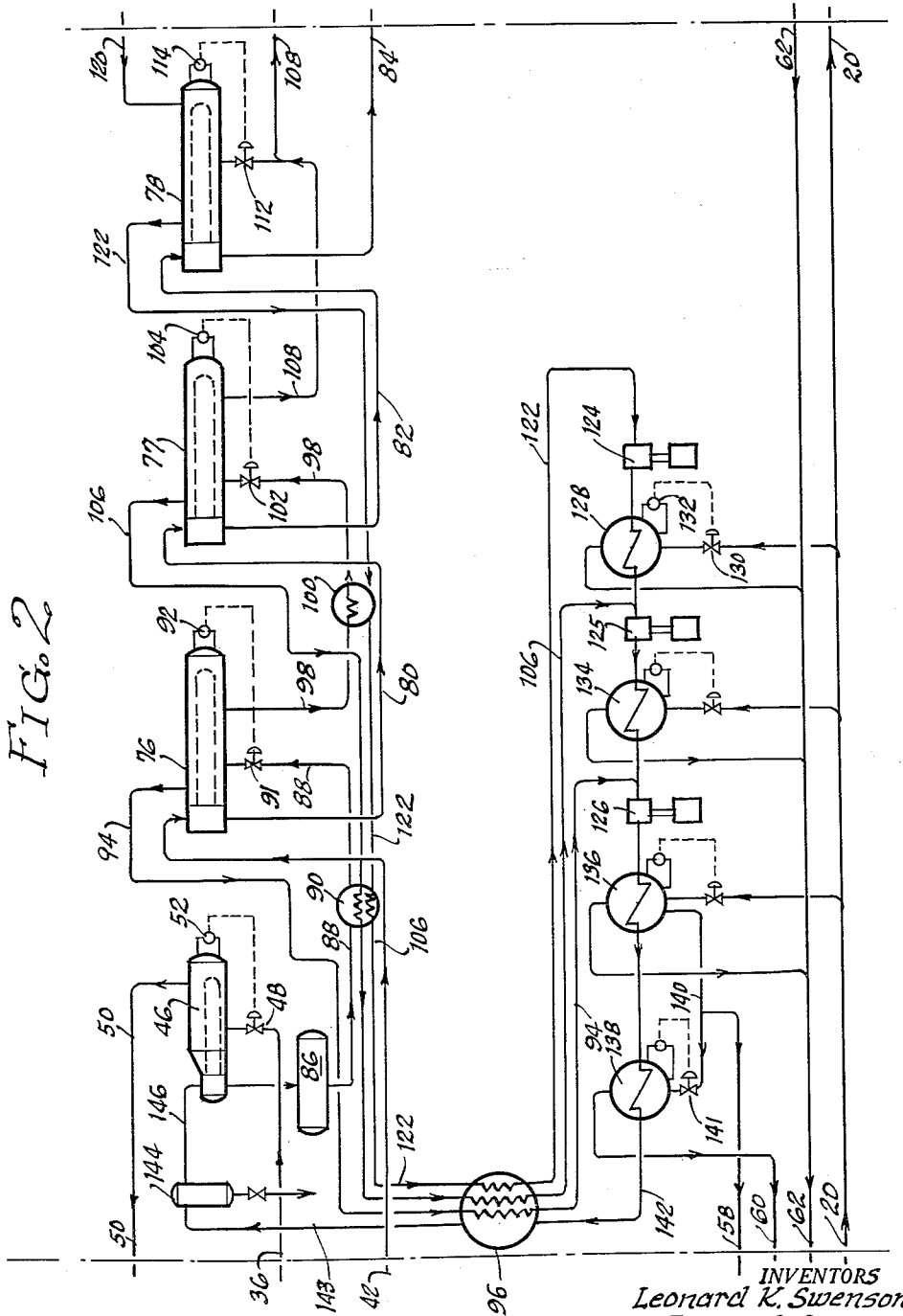
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3 Sheets-Sheet 3

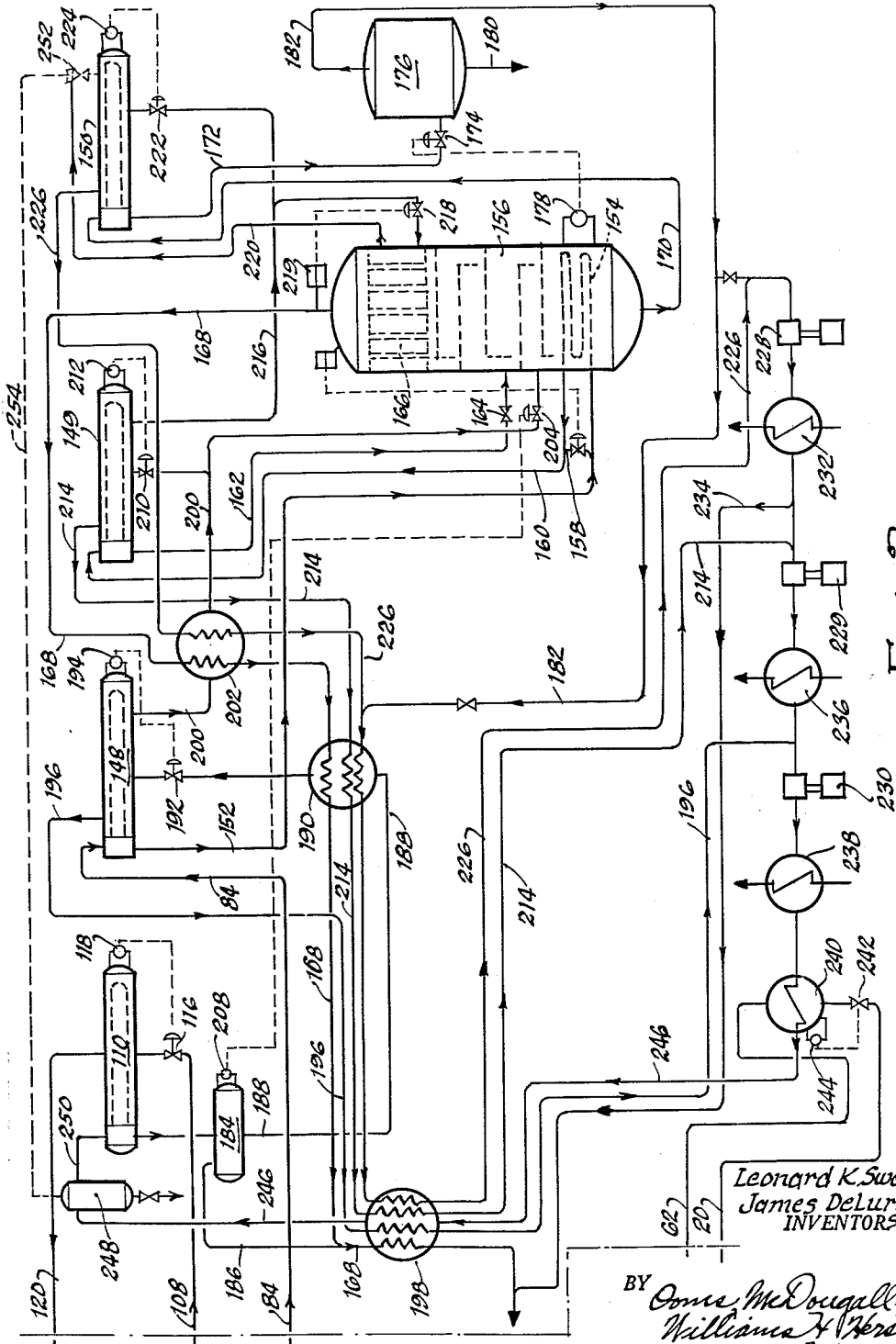


FIG. 3

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1

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LIQUEFYING NATURAL GAS WITH LOW PRESSURE REFRIGERANTS

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16 Claims. (Cl. 62—24)

This invention relates to the liquefaction of a gas and, more particularly, to a method and apparatus for the liquefaction of natural gas which is normally composed mostly of methane but which may contain heavier hydrocarbons such as ethane, propane, butane and the like, small amounts of aromatic hydrocarbons and variable amounts of non-hydrocarbons, such as nitrogen, helium, carbon dioxide, hydrogen sulfide and the like. Illustration of this invention will hereafter be made with reference to the liquefaction of natural gas but it will be understood that the concepts employed are also capable of application to other low boiling liquefiable gases, such as nitrogen, helium, air, oxygen and the like.

There are many purposes for which natural gas is desired to be reduced to a liquefied state. The main reason resides in the resultant reduction, at equivalent pressure, by about $\frac{1}{1000}$ in volume when reduced from the gaseous state to a liquefied state, thereby to enable storage and transportation in containers of more economical and practical design.

For example, when gas is transported by pipeline from the source of supply to a distant market, it is desirable to operate under substantially constant high load factor. Often times the flow capacity will exceed demand, while at other times the demand may exceed the capacity of the line. In order to shave off the peaks where demand would exceed supply, it is desirable to store gas when the supply exceeds demand, whereby peaks in demand can be met by material in storage. For this purpose, it is desirable to provide for storage in a liquefied state and to vaporize liquid in amounts to meet demand.

Liquefaction of natural gas is of even greater importance in making it possible to transport the gas from a source of plentiful supply to a distant market where a deficiency exists, especially when the source of supply cannot be directly joined with the market by a pipeline or the like means for the transportation of the gaseous fuel in a gaseous state. By way of illustration, surplus natural gas is available in the Gulf States of the United States, in Venezuela, and in the Persian Gulf, while deficiencies exist in the northern parts of the United States, the European countries, and Japan, yet these sources of supply cannot be joined by pipeline with some of the markets. Ship transportation in the gaseous state would be uneconomical, unless the gaseous materials were compressed, and then the system would not be commercial because it would be impractical to provide containers of suitable strength and capacity.

It has been determined that natural gas, when shipped from the United States or Venezuela in large volumes in liquefied state, can be made available in Great Britain, for example, at a price which is considerably less than locally manufactured gas. For shipment in large volume, it is desirable to house the liquefied natural gas in suitable insulated containers of large capacity at about atmospheric pressure, or preferably slightly above atmospheric, but not at such high pressures as would unduly limit the economical capacity of the tank. Depending

2

upon the amount of higher boiling heavier hydrocarbons present in the natural gas, the liquefied natural gas will have a boiling point within the range of -240° F. to -258° F. at atmospheric pressure.

5 The present invention contemplates a novel method of converting a gas from a gaseous state to a liquefied state for storage or transportation, wherein the gas is cooled in incremental steps by use of a plurality of refrigerants in such a manner that each refrigerant is used to remove the maximum amount of heat from the gas. Each refrigerant is compressed and expanded in a separate cycle, with each refrigerant being passed in heat exchange relation with the gas being liquefied in a plurality of stages at progressively decreasing pressures of the refrigerant, such that each refrigerant is used at the most efficient temperature levels for removing heat from the gas being liquefied. Each refrigerant is also compressed in a plurality of stages, such that a refrigerant expanded at intermediate pressure stages and passed in heat exchange relation with the gas being liquefied may be by-passed around the lower stages of compression and increase the economy of the refrigeration cycle. This invention further contemplates the distillation of nitrogen from a natural gas being liquefied in such a manner that heat from the natural gas being liquefied and refrigeration from one of the refrigerants used for liquefying the natural gas are used to maintain the desired temperatures in a nitrogen stripping tower for the efficient removal of nitrogen from the natural gas. In a preferred embodiment, the natural gas is used to heat the lower section of a nitrogen stripping tower and is then further cooled by a refrigerant before being expanded into the nitrogen tower for nitrogen removal.

An important object of this invention is to provide an economical and efficient method for converting a gas, and particularly a natural gas, from a gaseous state to a liquefied state for storage and transportation.

Another object of this invention is to provide a method of liquefying a gas by use of a plurality of refrigerants, wherein each refrigerant is circulated through a separate cycle and passed in heat exchange relation with the gas being liquefied at temperature levels which provide the most efficient removal of heat from the gas being liquefied.

Another object of this invention is to provide a method of liquefying natural gas by use of a plurality of separate refrigeration cycles, wherein each refrigeration cycle utilizes a plurality of stages of compression and expansion and wherein the heat of compression in the lower temperature level refrigeration cycles is at least partially removed by a higher temperature level refrigerant which is also used for removing heat from the gas being liquefied.

A further object of this invention is to provide a method of liquefying natural gas, wherein nitrogen is distilled from the natural gas and refrigeration from the removed nitrogen is utilized to subcool a refrigerant used in the liquefaction of the natural gas. It is also an object to utilize heat from the natural gas being liquefied for removal of a nitrogen from the natural gas, and then further cooling the natural gas prior to expansion of the natural gas into the distillation zone where the nitrogen is removed.

A still further object of this invention is to provide a novel method of liquefying natural gas by use of a plurality of refrigerants, wherein one of the refrigerants is methane, and boil-off vapor from the storage vessel receiving the liquefied gas is utilized in the methane refrigeration cycle. Further, it is an object of the invention to return excess methane in the refrigeration cycle to the natural gas stream being liquefied in such a man-

3

ner as to facilitate the removal of nitrogen from the natural gas.

Other objects and advantages of the invention will be evident from the following detailed description, when read in conjunction with the accompanying drawings which illustrate this invention.

In the drawings:

Figure 1 is a partial flow diagram illustrating a high temperature level refrigeration cycle which may be used in a practice of this invention.

Figure 2 is a partial flow diagram of an intermediate temperature level refrigeration cycle which may be used in a practice of this invention, with Fig. 2 being a continuation from the right hand end of Fig. 1.

Figure 3 is a partial flow diagram illustrating a low temperature level refrigeration cycle which may be used in a practice of this invention, with Fig. 3 being a continuation from the right hand end of Fig. 2, such that Figs. 1, 2 and 3 form a complete flow diagram illustrating a practice of this invention.

The process will hereinafter be described in detail with reference to the liquefaction of natural gas at a source of supply using an operative set of temperature and pressure conditions. It should be understood, however, that the conditions set forth are merely illustrative and may easily and properly be varied in consonance with the design and capacity of the apparatus, the character of the gas from the standpoint of composition, temperature and pressure, and the conditions under which the liquefaction is carried out as influenced by the volume of material, types of refrigerants and the like, all within the scope of the invention. In the example, the gas liquefied will be a natural gas from which moisture, acid gases, such as carbon dioxide, hydrogen sulfide and the like, will previously have been removed by pretreatment in the form of desiccators, amine extractors and the like. In this typical example, a cleaned natural gas is used having about 73 mol. percent methane, about 12 mol. percent ethane, about 8 mol. percent propane, about 2 mol. percent nitrogen and minor percentages of heavier hydrocarbons. It will be understood that natural gas capable of being processed in accordance with the teachings of this invention may have up to 20-25 mol. percent heavier hydrocarbons, up to 20 mol. percent nitrogen, and up to 5 mol. percent carbon dioxide or hydrogen sulfide, but usually the amount of methane will be from 70 to more than 90 mol. percent of the natural gas.

Referring to the drawings in detail, and particularly Fig. 1, reference character 4 designates a line leading from a source of supply of natural gas (not shown) for conveying the gas to a liquefaction system practicing the present invention. The natural gas, as previously indicated, will be passed from the producing wells through a clean-up system where moisture and acid gases are removed from the natural gas stream, and then the natural gas will be supplied to the liquefaction system, such that the natural gas will be supplied at a substantial pressure. For purposes of illustration, it will be assumed that the natural gas conveyed through the line 4 will be at a pressure of about 715 p.s.i.a. and have a temperature of about 94° F.

The natural gas stream is passed in series through the tubes of three heat exchangers 6, 8 and 10 by means of lines 12 and 14 for reducing the temperature of the natural gas to about -53° F. The three heat exchangers 6, 8 and 10 are cooled by a relatively high temperature level refrigeration cycle, preferably utilizing propane as the refrigerant. Condensed propane is stored in a surge drum 16 at a pressure of about 183.7 p.s.i.a. and a temperature of about 98° F. This propane is supplied through a line 18 to the shell of the heat exchanger 6 for initially cooling the natural gas feed stream passed through the tubes of the respective heat exchanger. It may also be noted that a portion of the propane refrigerant at about 98° F. is conveyed on through a line 20 for another refrigeration operation, as will be described.

4

The propane being conveyed through the line 18 to the heat exchanger 6 is expanded from about 183.7 to about 87.2 p.s.i.a. into the shell of the heat exchanger by means of a suitable expansion valve 22. The valve 22 is controlled by a liquid level controller 24 mounted on one end of the heat exchanger to control the liquid level of the propane in the heat exchanger 6. The propane in the heat exchanger 6 cools the natural gas feed stream from about 94° F. down to about 52° F., such that a portion of the liquid propane in the shell of the heat exchanger 6 will be boiled to provide both vapor and liquid propane in the heat exchanger 6. The propane vapor is withdrawn from the top of the heat exchanger 6 through a line 26 and returned to an intermediate stage of the compression portion of the refrigeration cycle, as will be hereinafter described. Liquid propane is withdrawn from the heat exchanger 6 through a line 28 and in turn fed to the shell of the medium pressure propane heat exchanger 8.

The propane conveyed through the line 28 will be at a pressure of about 87.2 p.s.i.a. and a temperature of about 47° F. As this propane is being fed into the heat exchanger 8 it is expanded down to a pressure of about 33.5 p.s.i.a. by use of an expansion valve 30, to substantially reduce the temperature of the propane entering in the heat exchanger 8, and remove an additional amount of heat from the natural gas feed stream flowing through the tubes of the heat exchanger 8. The natural gas will be reduced from about 52° F. down to about -2° F. by passage through the tubes of the heat exchanger 8. The expansion valve 30 is controlled by a liquid level controller 32 mounted on one end of the heat exchanger 8. The transfer of heat from the natural gas to the propane in the heat exchanger 8 will, as before, provide a boiling of the propane in the shell, thus the heat exchanger 8 will contain both vaporous and liquid propane. The propane vapor is withdrawn from the top of the shell through a line 34 and returned to another intermediate stage of the compression portion of the refrigeration cycle, as will be described. The liquid propane in the heat exchanger 8 is withdrawn through a line 36 and partially passed through an expansion valve 38 to the shell of the heat exchanger 10.

The liquid propane flowing through the line 36 will be at about 33.5 p.s.i.a. and have a temperature of about -6° F. The expansion valve 38 is controlled by a liquid level controller 40 on one end of the heat exchanger 10 to reduce the pressure of the propane being fed to the heat exchanger 10 down to about 10.8 p.s.i.a., such that the temperature of the propane is reduced to about -56° F. upon entering the heat exchanger 10. Thus, the propane refrigerant in the heat exchanger 10 will further reduce the temperature of the natural gas stream flowing through the tubes of the heat exchanger, such that the natural gas discharged from the exchanger 10 through the line 42 will be at a temperature of about -53° F. As before, the transfer of heat to the propane in the heat exchanger 10 will provide a boiling of the propane, such that propane vapors will collect in the top of the shell of the exchanger 10. These propane vapors are withdrawn through a line 44 and returned to the inlet of the compression portion of the cycle, as will be described.

The major portion of the liquid propane in the line 36 is by-passed around the heat exchanger 10 and fed on to a condenser 46 (Fig. 2) used in condensing the intermediate temperature level refrigerant, as will be more fully hereinafter set forth. As shown in Fig. 2, the propane entering the condenser 46 is expanded by a suitable valve 48 down to a pressure of about 11 p.s.i.a., with an accompanying drop in temperature to about -55° F. which, it will be noted, is slightly higher than the temperature and pressure of the propane fed to the heat exchanger 10. Thus, the propane vapor produced in the condenser 46 may be withdrawn through a line 50 and returned to the shell of the heat exchanger 10 for joiner with the vapor boiled off in the heat exchanger 10

5

through the line 44 to the compression portion of the cycle. The expansion valve 48 is controlled by a suitable liquid level controller 52 on one end of the condenser 46 to maintain the desired liquid level in the condenser 46.

As illustrated in the lower portion of Fig. 1, the compression portion of the propane refrigeration cycle utilizes three stages of compression, referred to as 54, 55 and 56, to progressively compress the propane vapors. The low pressure stage 54 receives propane vapors at about 9 p.s.i.a., and at a temperature of about -20°F. , from the line 44 and increases the pressure of these vapors to about 32.5 p.s.i.a., with a resulting increase in temperature to about 59°F. These partially compressed propane vapors are fed to the intake of the intermediate stage 55, along with propane vapors from the medium pressure and intermediate temperature level exchanger 8 directed through line 34, as well as propane vapors from lines 58 and 60. The propane flowing through lines 58 and 60 is used in cooling the intermediate temperature level refrigerant, as will be described. It may be noted at this point that the pressure of the vapors withdrawn from the medium pressure heat exchanger 8 are at a pressure of about 33.5 p.s.i.a., slightly higher than the discharge pressure of the low pressure compressor 54, such that the propane vapors generated in the heat exchanger 8 are returned to an intermediate stage of the compression portion of the cycle to minimize the horsepower required for operating the compressor 54.

The intermediate stage compressor 55 increases the pressure of the propane vapors to about 86 p.s.i.a., with a resulting temperature rise to about 80°F. These partially compressed vapors are fed to the intake of the high pressure stage 56, along with vapors conveyed through the line 26 from the high pressure exchanger 6, and along with propane conveyed through by-pass line 61 connected to the previously mentioned line 20 and propane vapors conveyed through the line 62 from the intermediate temperature level refrigeration cycle, as will be described. It should be noted here, however, that the vapors produced in the high pressure heat exchanger 6 are recirculated only through the high pressure compressor 56 to minimize the required horsepower for the compressor stages 54 and 55.

The high pressure stage 56 increases the pressure of the propane vapors to about 187 p.s.i.a., with a resulting temperature rise to about 122°F. The compressed vapors are conveyed through a line 64 to a suitable separator 66 wherein any lubricating oil which may have been picked up by the propane vapors in their passage through the stages 54, 55 and 56 is separated and collected in the bottom of the separator 66. It will be noted that at a temperature level of 122°F. , lubricating oil present in the propane vapor stream as a vapor mist and/or fog will be condensed at least in part to permit separation from the propane vapors. The separated oil is selectively discharged from the lower end of the separator 66 through a drain line 68. The cleaned propane vapors are withdrawn from the top of the separator 66 through a line 70 and conveyed to a suitable heat exchanger 72 operating as a condenser for condensing the propane. The condenser 72 may be easily cooled by water at the temperature conditions existing, such that the propane passing through the condenser 72 will be reduced in temperature to about 98°F. and condensed. The condensed propane is fed through a line 74 to the propane surge drum 16 for subsequent passage through the propane refrigeration cycle.

The intermediate temperature level refrigeration cycle illustrated in Fig. 2 may utilize either ethane or ethylene as the refrigerant, although ethane is the preferred refrigerant, principally due to its availability is most natural gas. In other words, the ethane may normally be obtained as a fraction from natural gas, the same as the propane, such that a commercial system practicing the present invention will not require refrigerants supplied

6

from an outside source. This intermediate temperature level refrigeration cycle utilizes three main heat exchangers 76, 77 and 78 through which the natural gas feed stream is passed in series to incrementally reduce the temperature of the natural gas from about -53°F. to about -145°F.

The stream is fed to the tubes of the heat exchanger 76 by the line 42 and is withdrawn from the exchanger 78 through a line 80 at a temperature of about -77°F. The natural gas stream is in turn conveyed by the line 80 to the heat exchanger 77 wherein the temperature of the gas is further reduced to about -107°F. , and, in the example taken for illustration, at this temperature the natural gas will be substantially totally condensed. The feed stream is transferred from the heat exchanger 77 to the tubes of the heat exchanger 78 through a line 82; whereupon the temperature of the natural gas feed stream is reduced on down to about -145°F. and the stream is withdrawn from the tubes of the heat exchanger 78 through a line 84 for further cooling, as will be hereinafter described.

The intermediate temperature level refrigerant, which will be hereinafter described as ethane, is stored in a surge drum 86 in liquid form at about -49°F. and at a pressure of about 96 p.s.i.a. The liquid ethane is withdrawn from the lower end of the surge drum 86 through a line 88 and passed through a suitable heat exchanger 90 for a decrease in temperature to about -62°F. , as will be described. The subcooled ethane is conveyed on through the line 88 and expanded through a suitable expansion valve 91 into the shell of the heat exchanger 76. The expansion valve 91 reduces the pressure of the ethane to about 50.2 p.s.i.a. which reduces the temperature of the ethane entering the heat exchanger 76 to about -80°F. for efficiently removing heat from the natural gas feed stream flowing through the tubes of the exchanger 76. The transfer of heat to the ethane in exchanger 76 provides a boiling of the ethane at the pressure conditions existing to, in turn, provide both vaporous and liquid ethane in the exchanger 76. The expansion valve 91 is operated by a suitable liquid level controller 92 mounted on an end of the exchanger 76 to control the liquid level in the exchanger 76. The ethane vapor is withdrawn from the top of the exchanger 76 through a line 94 and returned through a heat exchanger 96 for refrigerating higher temperature ethane, as will be described. This expanded ethane vapor is conveyed on through the line 94 to an intermediate stage of the compression portion of the ethane refrigeration cycle, as will be more fully hereinafter described.

Liquid ethane is withdrawn from the bottom of the heat exchanger 76 through a line 98 at about -80°F. , sub-cooled in a heat exchanger 100 to about -90.5°F. and then fed to the shell of the medium pressure ethane exchanger 77. An expansion valve 102 is interposed in the line 98 adjacent the exchanger 77 and is controlled by a suitable liquid level controller 104 for reducing the pressure of the ethane fed to the exchanger 77 down to a pressure of about 24.1 p.s.i.a., with a resulting decrease in temperature of the ethane to about -110°F. The ethane refrigerant in the exchanger 77 reduces the temperature of the natural gas feed stream to about -107°F. , as previously indicated.

The transfer of heat from the natural gas feed stream to the ethane in the exchanger 77 provides a boiling of the ethane refrigerant to in turn provide both vaporous and liquid ethane in the shell of the exchanger 77. The ethane vapor is withdrawn from the top of the exchanger 77 through a line 106 at a pressure of about 23.5 p.s.i.a. and a temperature of about -85°F. This ethane vapor is passed through the heat exchanger 90 to provide a cooling of the liquid ethane being fed to the heat exchanger 76, as previously described, and is then fed on through the line 106 to the heat exchanger 96 to provide a refrigerating action, as will be described. The slightly warmed ethane vapor is then conveyed on through the line 106

to an intermediate stage of the compression portion of the ethane refrigeration cycle, as will be described.

Liquid ethane is withdrawn from the shell of the heat exchanger 77 through a line 108 at a temperature of about -110° F. and a pressure of about 23.5 p.s.i.a. for transfer to the shell of the low pressure ethane heat exchanger 78 and a condenser 110 (Fig. 3) for a further cooling of the natural gas feed stream and for condensing the low temperature level refrigerant, respectively, as will be described. The ethane fed to the heat exchanger 78 (Fig. 2) is expanded by a suitable expansion valve 112 down to a pressure of about 7.7 p.s.i.a., with a resulting decrease in the temperature of the ethane to about -148° F. The expansion valve 112 is controlled by a liquid level controller 114 mounted on an end of the exchanger 78 to control the liquid level in the exchanger 78 in the usual manner. The ethane refrigerant in the exchanger 78 reduces the temperature of the natural gas feed stream to about -145° F., as previously indicated, to provide a boiling of the ethane in the exchanger.

The liquid ethane fed to the condenser 110 (Fig. 3) is also reduced in pressure by an expansion valve 116 to a pressure of about 7.9 p.s.i.a., with a resulting decrease in temperature to about -147° F. The expansion valve 116 is controlled by a liquid level controller 118 mounted on the condenser 110 in the usual fashion. Heat is transferred to the ethane in the condenser 110, as will be described, to boil the ethane in the condenser. The resulting ethane vapors generated in the condenser 110 are withdrawn through a line 120 and directed into the shell of the low pressure ethane heat exchanger 78 (Fig. 2). It will be noted that the pressure of the ethane vapors in the condenser 110 is slightly higher than the pressure of the vapors in the heat exchanger 78, such that the ethane vapors will readily flow from the condenser 110 through the line 120 into the exchanger 78. All of the vapors in the exchanger 78 are withdrawn through a line 122 and fed back through the heat exchanger 100 for sub-cooling the liquid ethane being fed to the heat exchanger 77. These ethane vapors are conveyed on through the line 122 for passage through the heat exchangers 90 and 96, and then the ethane vapors are fed to the low pressure side of the compression portion of the ethane refrigeration cycle. The pressure of the ethane vapors in the line 122 will be reduced to about 5 p.s.i.a. by passage through the exchangers 100, 90 and 96.

As illustrated in the lower portion of Fig. 2, the ethane refrigeration cycle utilizes three compressor stages, referred to as stages 124, 125 and 126, to progressively increase the pressure of the ethane vapors and provide an economical refrigeration cycle. The low stage 124 compresses ethane vapors from the line 122 to a pressure of about 22 p.s.i.a., with a resulting temperature rise of the vapors to about 121° F. This heat of compression is partially removed by a heat exchanger 128 preferably cooled by propane from the propane refrigeration cycle. As previously indicated, see Fig. 1, a portion of the propane liquid stored in the propane surge drum 16 is conveyed through the line 20 to the intermediate temperature level refrigeration cycle. A portion of the liquid propane in the line 20 is expanded through an expansion valve 130 into the heat exchanger 128 down to a pressure of about 87.2 p.s.i.a., with a resulting temperature drop to about 47° F. The expanded propane removes heat of compression from the ethane vapors passing through the coil of the heat exchanger 128 to such an extent that the temperature of the ethane vapors being fed to the inlet of the intermediate stage 125 is about 60° F. After passing through the heat exchanger 128, the propane is returned through the line 62 to the inlet of the high pressure propane compressor 56 in the manner previously described.

The ethane vapors entering the intermediate stage 125 are obtained from the low stage 124 and the line 106 leading from the medium pressure main ethane heat exchanger 77, such that the ethane vapors withdrawn from

the heat exchanger 77 are by-passed around the low stage 124 to minimize the required horsepower of the compressor 124. The intermediate stage 125 increases the pressure of the ethane vapors to about 50 p.s.i.a., with a resulting rise in temperature to about 123° F. The heat of compression in the ethane vapors is again removed by propane from the line 20 expanded into a heat exchanger 134 in the same manner as previously described in connection with the heat exchanger 128. Thus, the ethane vapors flowing from the stage 125 to the stage 126 are reduced in temperature to about 60° F.

The high pressure stage 126 also receives ethane vapors through the line 94 leading from the high pressure main ethane heat exchanger 76, such that the vapors withdrawn from the exchanger 76 are by-passed around the stages 124 and 125. The high pressure stage 126 increases the pressure of the ethane vapors to about 107 p.s.i.a., with a resulting temperature increase to about 118° F. Another heat exchanger 136 is interposed in the outlet of the high pressure stage 126 and cooled by propane expanded from the line 20 in the same manner as in the exchangers 134 and 128. Thus, the ethane vapors discharging from the high pressure stage 126 are cooled down to about 60° F. These compressed ethane vapors are in turn preferably passed through another heat exchanger 138 acting as an afterchiller to further reduce the temperature of the vapors to about 7° F. The heat exchanger 138 receives propane through a line 140 leading from the heat exchanger 136, with the propane fed to the exchanger 138 being further expanded by a suitable expansion valve 141 to reduce the temperature of the propane entering the exchanger 138 to about -6° F. It should be noted that a substantially greater amount of propane is fed to the exchanger 136 than is fed to the exchangers 134 and 128, and the propane fed to the exchanger 136 is reduced in pressure only to about 87.2 p.s.i.a. Therefore, this propane may be withdrawn in liquid form from the exchanger 136 and in turn expanded into the exchanger 138 for a further reduction in temperature.

A portion of the propane is passed through the line 58 and expanded down to about 32.5 p.s.i.a. for entry into the intermediate stage propane compressor 55, for the purpose of reducing the suction temperature to the compressor, as previously indicated. In the same manner, liquid propane is flashed through line 61 for the purpose of lowering the temperature of gas to compressor 56. The propane vapors discharging from the top of the exchanger 138 are returned through the line 60 to the intermediate stage propane compressor 55. It will thus be apparent that the propane refrigerant is utilized to remove all of the heat of compression from the ethane refrigerant by use of heat exchangers between the various stages of compression and a pair of heat exchangers after the last stage of compression.

The ethane vapors discharging from the exchanger 138 are conveyed by a line 142 to the heat exchanger 96 for a further reduction in temperature of the vapors to about -40° F. As previously indicated, the exchanger 96 is cooled by ethane vapors withdrawn from the main ethane exchangers 76, 77 and 78. The cold methane vapors flowing from the exchanger 96 are directed on through a line 143 to a suitable separator 144 wherein any lubricating oil which may have been picked up by the vapors in passage through the compressors 124, 125 and 126 is removed. It will be apparent that with the ethane vapors at a temperature of -40° F., any oil which may be entrained in the vapors will be in liquid form and may be easily separated in the separator 144. These condensates are selectively drained from the separator 144 when and as required. The remaining ethane vapors are directed from the separator 144 through a line 146 to the coils of the ethane condenser 46.

As previously described, the condenser 46 is main-

tained at a temperature of about -55° F. by expansion of propane through the valve 48, such that the ethane circulating through the coil of the condenser will be converted to a liquid state. The condensed ethane flows from the condenser 46 into the ethane surge drum 86 for re-use in the ethane refrigeration cycle.

It will thus be apparent that the ethane refrigeration cycle utilizes three progressive expansions of liquid ethane to provide three separate cooling steps for the natural gas feed stream, as well as three separate compression stages corresponding in pressure levels to the expansion steps. Thus, the ethane vapors produced in each of the first two cooling steps may be returned to intermediate stages of the compression portion of the cycle to minimize the required horsepower for the compression of the ethane vapors. It will also be noted that the ethane refrigerant is used through a temperature and pressure range which permits the most efficient heat transfer from the natural gas feed stream to the ethane refrigerant and provide the maximum removal of heat from the natural gas feed stream by the ethane refrigeration cycle. Further, the heat of compression is removed from the ethane refrigerant by the propane refrigerant used in the higher temperature level removal of heat from the natural gas feed stream.

The low temperature level refrigeration cycle preferably utilizes methane as a refrigerant and three main heat exchangers 148, 149 and 150. The natural gas feed stream is fed to the high pressure methane exchanger 148 by the line 84 to provide a further cooling of the natural gas feed stream from about -145° F. to about -182° F. The natural gas feed stream is withdrawn from the exchanger 148 through a line 152 and conveyed to the re-boiler section 154 of a nitrogen stripping tower 156. The natural gas feed stream flowing through the re-boiler 154 provides a warming of the contents of the lower end portion of the tower 156, as will be more fully hereinafter set forth, such that the temperature of the natural gas stream is reduced to about -204° F. upon discharge from the re-boiler 154. It may also be noted that a valved by-pass line 138 is provided between the inlet and the outlet of the re-boiler 154 to by-pass all or a portion of the natural gas feed stream around the re-boiler 154 when the conditions in the tower 156 so require.

The natural gas feed stream discharging from the re-boiler 154 and through the by-pass 158 is conveyed by a line 160 to the coils of the medium pressure methane exchanger 149 for a further cooling of the natural gas feed stream. The natural gas feed stream is cooled to about -216° F. by passage through the exchanger 149 and is withdrawn from the exchanger 149 through line 162. It will thus be noted that instead of a natural gas feed stream being conveyed directly from the high pressure methane exchanger 148 to the low pressure methane exchanger 149, the feed stream is first passed through the re-boiler section of the nitrogen stripping tower to obtain the benefit of the heat content of the natural gas feed stream in the operation of the stripping tower.

The natural gas feed stream is conveyed through the line 162 to a medial portion of the tower 156 for a distillation of nitrogen from the feed stream. The natural gas feed stream entering the tower 156 is expanded through a suitable expansion valve 164 down to an intermediate pressure, such as about 66 p.s.i.a., to enhance the vaporization of the nitrogen component of the feed stream. The tower 156 operates in the usual manner to provide a downward flow of liquid and an upward flow of vapors. A reflux condenser 166 is provided in the upper section of the tower 156 and is maintained at a temperature below the temperature of the expanded natural gas feed stream entering the tower 156, such that natural gas vapors rising through the tower 156 will tend to become condensed and flow downwardly through the tower, and

the vapors collecting in the upper end of the tower will be enriched with nitrogen. The operation of the reflux condenser 166 will be described below. Also, the nitrogen-enriched vapors collecting in the upper end of the tower 156 are withdrawn through a line 168 and utilized to cool the methane refrigerant, as will be described below.

The substantially nitrogen-free liquefied natural gas collecting in the lower end of the tower 156 is withdrawn through a line 170 and conveyed to the coils of the low pressure methane exchanger 150. Methane refrigerant in the exchanger 150 further reduces the temperature of the natural gas feed stream passing through the exchanger from a temperature of about -209° F. to about -246° F. Thus, the low pressure methane exchanger 150 acts as a sub-cooler. The liquefied natural gas feed stream is withdrawn from the exchanger 150 through a line 172 and expanded through an expansion valve 174 down to about atmospheric pressure, or slightly above, such as 17.7 p.s.i.a., into a suitable insulated storage vessel 176. The final expansion of the liquefied natural gas by the valve 174 will, since the liquefied natural gas is sub-cooled, provide a minimum of flashing and minimize the vapors fed to the storage vessel 176. It may also be noted that the expansion valve 174 is controlled by a suitable liquid level controller 178 on the side of the nitrogen stripping tower 156 to control the liquid level in the tower 156.

The storage vessel 176 is suitably insulated, and the liquid product, comprising substantially nitrogen-free liquefied natural gas at a pressure of about 17.7 p.s.i.a. and a temperature of about -246° F., is selectively withdrawn from the bottom of the tank through a line 180, such that the liquid product may be either transported by pipeline, or loaded into suitable shipping containers (not shown) for transportation to distant markets. As it is well known in the art, the insulation of the storage vessel 176 will not be a perfect insulation. Therefore, at least a minor amount of the liquefied natural gas will be boiled off as a vapor. This boil-off vapor is withdrawn from the top of the vessel 176 through a line 182 and fed into the methane refrigeration cycle, as will be hereinafter set forth.

The methane refrigerant used for cooling the main heat exchangers 148, 149 and 150 is stored in a suitable surge drum 184 at a temperature of about -143° F. and a pressure of about 415 p.s.i.a., such that the stored methane refrigerant will be in liquid form. However, at least a minor portion of the methane refrigerant in the surge drum 184 will be boiled off, and this boil-off vapor is withdrawn through a line 186 to join with the nitrogen-enriched vapors withdrawn from the top of the nitrogen stripping tower 156 through the line 168. These vapors are used as a fuel for various units of equipment required in a commercial installation practicing the present invention.

Liquid methane refrigerant is withdrawn from the bottom of the surge drum 184 through a line 188 and fed to the high pressure methane heat exchanger 148. However, the refrigerant being fed to the exchanger 148 is preferably sub-cooled by an exchanger 190 down to a temperature of -154.5° F. prior to being expanded by a valve 192 into the exchanger 148. The heat exchanger 190 is cooled by the nitrogen-enriched vapors withdrawn from the top of the nitrogen stripping tower 156, as well as vaporized methane refrigerant, as will be described. The expansion valve 192 is controlled by suitable liquid level controller 194 mounted on an end of the exchanger 148, such that the liquid methane refrigerant being fed to the exchanger 148 will be expanded down to a pressure of about 170 p.s.i.a., with a resulting decrease in temperature to about -185° F. The transfer of heat from the natural gas feed stream flowing through the coils of the exchanger 148 will provide a boiling of the methane refrigerant in the shell of the exchanger 148. The

resulting methane vapor refrigerant is withdrawn through a line 196 and passed through a heat exchanger 198 to an intermediate stage of the compression portion of the methane refrigeration cycle. The heat exchanger 198 is also cooled by the nitrogen-enriched vapors flowing through the line 168, as will as additional methane refrigerant vapors, as will be described, to cool compressed methane vapors leaving the compressor portion of the refrigeration cycle.

Liquid methane refrigerant is withdrawn from the shell of the exchanger 148 through a line 200 and passed through an exchanger 202 for a reduction in temperature to about -194.7°F . The exchanger 202 is cooled by methane refrigerant vapors, as will be described. The liquid methane refrigerant leaving the exchanger 202 through the line 200 is partially fed to the medium pressure methane exchanger 149 and partially to an intermediate portion of the nitrogen stripping tower 156 at a point below the point at which the natural gas feed stream is expanded into the tower 156. As will be apparent, the methane refrigerant will be substantially nitrogen-free; therefore this refrigerant may be expanded into a lower portion of the nitrogen stripping tower 156 and act as a nitrogen purge to facilitate the removal of nitrogen from the natural gas feed stream in the stripper 156. The expansion valve 204 used for expanding the liquefied methane refrigerant into the tower 156 is controlled by a suitable liquid level controller 208 mounted on the side of the methane refrigerant surge drum 184, such that only excess methane refrigerant is fed to the tower 156 for joinder with the main natural gas feed stream.

The liquid methane refrigerant fed from the line 200 into the shell of the medium pressure exchanger 149 is expanded through a suitable expansion valve 210 to reduce the pressure of the refrigerant entering the exchanger to about 66 p.s.i.a., with a resultant reduction in temperature to about -219°F . The expansion valve 210 is controlled by a liquid level controller 212 mounted on the end of the exchanger 149 in the usual fashion. As before, heat removed from the natural gas feed stream flowing through the coils of the exchanger 149 provides a boiling of the methane refrigerant contained in the shell of the exchanger 149. The resulting methane refrigerant vapors are withdrawn from the exchanger 149 through a line 214 and passed through the heat exchangers 190 and 198 to an intermediate stage of the compressor portion of the refrigeration cycle, as will be more fully hereinafter described.

Liquid methane refrigerant is withdrawn from the bottom of the exchanger 149 through a line 216 and directed partially to the reflux condenser 166 in the upper section of the tower 156, and partially to the low pressure methane heat exchanger 150. That portion of the liquefied methane refrigerant directed to the reflux condenser 166 is expanded by a suitable expansion valve 218 down to a pressure of about 23.5 p.s.i.a., with a resulting decrease in temperature to about -248°F . It will be observed that this temperature level is below the temperature level (-219°F) of the natural gas feed stream expanded into the medial portion of the tower 156, such that any methane vapors rising through the tower will tend to be condensed and flow back downwardly to the lower end of the tower. It may also be noted that the expansion valve 218 is controlled by a temperature controller 219 mounted on the vapor line 168 leading from the tower, such that the temperature maintained in the reflux condenser 166 will be governed by the temperature of the overhead vapors discharging from the tower. Methane refrigerant vapors are withdrawn from the reflux condenser 166 through a line 220 and conveyed to the shell of the low pressure methane exchanger 150.

That portion of the liquefied methane refrigerant flowing through the line 216 which is fed to the low pres-

sure exchanger 150 is expanded through a suitable expansion valve 222 down to a pressure of about 21.5 p.s.i.a., with a resulting decrease in temperature to about -250°F . for cooling the natural gas feed stream in the coils of the exchanger to -246°F . The expansion valve 222 is controlled by a liquid level controller 224 mounted on one end of the exchanger 150.

Methane refrigerant vapor in the shell of the exchanger 150 is withdrawn through a line 226 and passed through the heat exchangers 202, 190 and 198 to the low pressure side of the compressor portion of the methane refrigeration cycle. These methane vapors will be provided both by cooling of the natural gas feed stream in the coils of the exchanger 150, and by the vapor withdrawn from the reflux condenser 166.

The methane refrigerant is compressed by three stage compressors 228, 229 and 230. The low pressure stage 228 receives methane refrigerant vapor from the line 226 which leads from the low pressure methane heat exchanger 150. It will also be noted that the boil-off vapor from the storage vessel 176 is conducted through the line 182 and joined with the vapor in the line 226, such that this boil-off vapor is also fed to the low pressure stage 228. The compressor 228 increases the pressure of the methane refrigerant vapor from about 15 p.s.i.a. to about 64 p.s.i.a., with a resulting temperature rise to about 231°F . The temperature level of the partially compressed methane vapor is sufficiently high that the heat of compression may be removed by a water cooled heat exchanger 232 and the temperature of the vapor reduced to about 105°F . The major portion of the vapor leaving the exchanger 232 is fed to the intake of the intermediate stage 229. However, a portion of the refrigerant may be by-passed through a line 234 and joined with the nitrogen-enriched vapors in the line 168 to provide make-up fuel as necessary.

Methane vapors from the line 214 are joined with the partially compressed vapors discharging from the exchanger 232 at the intake of the intermediate stage 229. It will be noted that the vapors in the line 214 are taken from the medium pressure methane exchanger 149, such that these vapors are by-passed around the low pressure stage 228 to minimize the horsepower requirements for the low stage of the compressor.

The intermediate stage 229 increases the pressure of the methane vapors from about 60 p.s.i.a. to about 170 p.s.i.a., with a resulting temperature rise to about 228°F . This temperature level is again sufficiently high that the heat of compression may be removed by another water-cooled exchanger 236, such that the temperature of the vapors will again be reduced to about 105°F .

The partially compressed methane refrigerant vapors discharging from the exchanger 236, along with vapor from the line 196, are fed to the intake of the high pressure compressor 230. It will be noted that the vapors in the line 196 are taken from the high pressure methane exchanger 148, such that these vapors by-pass the lower pressure stages 228 and 229 to minimize the horsepower requirements of these compressors. The compressor 230 increases the pressure of the methane vapors from about 165 p.s.i.a. to about 430 p.s.i.a., with a resulting temperature increase to about 210°F . This temperature level is again sufficiently high that the heat of compression may be removed by a water-cooled exchanger 238 to reduce the temperature of the vapors to about 105°F .

It is also preferred to interpose an afterchiller 240 in the discharge from the exchanger 238 to further cool the compressed methane vapors down to about 60°F . The chiller 240 is cooled by propane refrigerant fed to the cooler from the line 20 through an expansion valve 242. The expansion valve 242 is operated by a liquid level controller 244 mounted on a side of the cooler to control the operation of the valve 242 and expand the propane refrigerant from about 183.7 p.s.i.a. down to

about 82.2 p.s.i.a., with a temperature drop to about 47° F. Propane refrigerant is withdrawn from the cooler 240 through the line 62 and returned with the propane refrigerant from the various other coolers in the ethane compression cycle to an intermediate point in the compressing portion of the propane refrigeration cycle.

The cooled methane refrigerant vapors discharging from the afterchiller 240 are conveyed through a line 246, through the exchanger 198, and on to a suitable separator 248. The compressed methane refrigerant vapors are cooled to about -120° F. by passage through the exchanger 198, such that the methane will still be in vaporous form, but lubricating oil which may have been picked up as a mist, fog or vapor by the methane in passing through the stages 228, 229 and 230 is condensed and removed in the separator 248. The condensates are drained from the separator 248 when and as required. The remaining methane refrigerant vapors are fed through a line 250 to the coil of the methane condenser 110, wherein a sufficient amount of heat is removed from the methane vapors by the ethane refrigerant in the shell of the condenser 110 to convert the methane refrigerant to a liquefied state. The liquefied methane refrigerant is drained into the methane surge drum 184 for a re-use in the methane refrigeration cycle.

It will thus be observed that in the low temperature methane refrigeration cycle, the methane is expanded through three separate stages for providing three separate cooling steps for the natural gas feed stream. Also, the methane refrigerant is compressed in three separate stages, such that the expanded methane refrigerant may be returned to intermediate portions of the compressor cycle to minimize the horsepower requirements for compressing the refrigerant. The methane refrigerant vapor is passed in heat exchange relation with the liquefied methane refrigerant prior to expansion of the refrigerant to obtain the maximum refrigeration from the methane.

From the foregoing it will be apparent that the present invention provides a novel method of liquefying a natural gas wherein the natural gas is cooled in incremental steps by separate refrigerants, with the refrigerants being utilized through temperature and pressure ranges which provide the most efficient transfer of heat from the natural gas to the respective refrigerants. The heat exchange between the various refrigerants and the natural gas feed stream is accomplished with the refrigerants being in liquid form to provide the most efficient heat transfer operation. It will also be apparent that the present invention provides a novel method of liquefying natural gas wherein nitrogen is distilled from the feed stream by use of heat from the feed stream and by use of refrigeration made available in a refrigerant used for providing a normal cooling of the feed stream. The natural gas feed stream is passed through the re-boiler section of a nitrogen stripping tower and then through a medium pressure methane exchanger to obtain the benefits of heat from the feed stream, and yet provide operation of the various refrigerants at the maximum temperature levels to minimize the horsepower requirements of the refrigeration cycles. Boil-off vapor from the storage vessel is fed to one of the refrigeration cycles, and excessive refrigerant is returned to the natural gas feed stream in such a manner as to facilitate the distillation of nitrogen from the feed stream.

Changes may be made in the combination and arrangement of steps and procedures as heretofore set forth in the specification and shown in the drawings, it being understood that changes may be made in the precise embodiment disclosed without departing from the spirit and scope of the invention as defined in the following claims. For example, and as shown in dashed lines in Fig. 3, a methane jet ejector 252 may be mounted on the low pressure methane exchanger 150 and connected to the line

220 leading from the reflux condenser 166. The ejector 252 is also connected by a line 254 to the top of the separator 248. In this embodiment, the pressure available in the separator 248 may be used to operate the ejector 252 and decrease the pressure of the methane refrigerant vapor in the line 220 and the reflux condenser 166. This will in turn reduce the temperature of the reflux condenser and reduce the natural gas content in the overhead vapor line 168, with a corresponding proportional increase in the nitrogen content of the overhead vapors. Such a system is desirable when the overhead vapors from the tower 156 are vented and not used as a fuel in the system.

While reference is made to three-stage compressors in the foregoing description, it will be understood that compressors having two or more stages may be employed and that each stage may be embodied in separate or multiple-stage compressor units. It will be further understood that changes may be made in the details of construction and operation without departing from the spirit of the invention, especially as defined in the following claims.

We claim:

1. In a method of liquefying a natural gas composed mostly of methane, the steps of:
 - (a) supplying the natural gas in a process stream at an elevated temperature and pressure,
 - (b) condensing the natural gas without substantially reducing the pressure thereof by passing the process stream in heat exchange relation with a series of separate progressively decreasing temperature level refrigerants for incremental reduction in temperature of the stream,
 - (c) expanding the liquefied natural gas into a storage vessel to a pressure suitable for transportation of the product in liquid form,
 - (d) alternately compressing, condensing and expanding each refrigerant in a separate closed cycle wherein at least one of the refrigerants is expanded to sub-atmospheric pressure, and
 - (e) passing the expanded refrigerants in heat exchange relation with the process stream at temperature and pressure levels such that the separate refrigerants are in liquid form when passed in heat exchange relation with the process stream and receive principally latent heat from the process stream to provide the condensation called for in step (b).
2. The method defined in claim 1 characterized further in that the refrigerants are fractions of natural gas.
3. The method defined in claim 1 characterized further in that at least one of the refrigerants is methane, boil-off vapor from the storage vessel is added to the methane refrigeration cycle, and excess liquefied methane in the methane refrigeration cycle is fed into the process stream.
4. The method defined in claim 1 characterized further in that the refrigerants are propane, ethane and methane.
5. The method defined in claim 1 characterized further in that nitrogen is distilled from the process stream after the natural gas is condensed, but before the stream is expanded to a pressure suitable for transportation.
6. The method defined in claim 1 characterized further in that a higher temperature level refrigerant is passed in heat exchange relation with a lower temperature level refrigerant for removing heat of compression from the lower temperature level refrigerant.
7. The method defined in claim 1 characterized further in that each refrigerant is compressed and expanded in a plurality of stages, and that portion of each refrigerant vaporized at an intermediate pressure is returned to an intermediate stage of compression.
8. The method defined in claim 1 characterized fur-

15

ther in that a vaporized portion of each of the lower temperature level refrigerants is passed in heat exchange relation with a condensed portion of the same refrigerant for subcooling the condensed refrigerants.

9. The method defined in claim 1 characterized further in that three separate refrigerants are used at progressively lower temperature levels for cooling the process stream, each refrigerant is compressed and expanded through a plurality of stages, and the higher temperature level refrigerant is passed in heat exchange relation with the intermediate temperature level refrigerant between compression stages and after the final stage of compression of the intermediate temperature level refrigerant to remove heat of compression from the intermediate temperature level refrigerant.

10. The method defined in claim 9 characterized further in that the highest temperature level refrigerant is passed in heat exchange relation with the lowest temperature level refrigerant to remove heat from the lowest temperature level refrigerant prior to condensation thereof.

11. The method defined in claim 1 characterized further in that three separate refrigerants are used at progressively lower temperature levels for cooling the process stream, and each of the higher temperature level refrigerants is passed in heat exchanger relation with the next lower temperature level refrigerant following compression of the respective lower temperature level refrigerant for condensing the respective lower temperature level refrigerant.

12. In a method of liquefying a natural gas containing nitrogen, the steps of:

(a) supplying the natural gas in a process stream at an elevated temperature and pressure,

(b) condensing the natural gas without substantially reducing the pressure thereof by passing the process stream in heat exchange relation with a series of progressively lower temperature level refrigerants for incremental reduction in temperature of the stream, with each series containing more than one refrigeration stage,

(c) expanding the liquefied stream to an intermediate pressure into the medial portion of a nitrogen stripping tower having a reboiler in the lower section thereof maintained at a temperature above the temperature of the expanded stream and a reflux condenser in the upper section thereof maintained at a temperature below the tem-

16

perature of the expanded stream for vaporizing and removing nitrogen from the stream,

(d) withdrawing nitrogen enriched vapors from the top of the nitrogen stripping tower and passing said vapors in heat exchange relation with at least one of the refrigerants,

(e) withdrawing the remaining process stream from the lower end of the nitrogen stripping tower,

(f) subcooling said remaining process stream, and

(g) expanding said remaining process stream into a storage vessel to a pressure suitable for transporting the product in liquid form.

13. The method defined in claim 12 characterized further in that the process stream is passed through said re-boiler prior to expansion thereof into the medial portion of the stripping tower for maintaining the re-boiler at the desired temperature, and the lowest temperature level refrigerant is passed through said reflux condenser for maintaining the reflux condenser at the desired temperature.

14. The method defined in claim 13 characterized further in that the process stream is passed in heat exchange relation with a refrigerant for further cooling thereof after passage through the re-boiler and before expansion into the stripping tower.

15. The method defined in claim 12 characterized further in that the lowest temperature level refrigerant is methane, boil-off vapor from the storage vessel is combined with the methane refrigerant, and excess methane refrigerant is bled into the process stream.

16. The method defined in claim 15 characterized further in that the excess methane refrigerant is expanded into the stripping tower at a point below the expansion of the process stream into the tower for combining with the process stream and acting as a nitrogen purge in the stripping tower.

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