(12) INTERNATIONAL APPLICATION PUBLISHED UNDER THE PATENT COOPERATION TREATY (PCT)

(19) World Intellectual Property Organization
International Bureau

(43) International Publication Date
27 December 2007 (27.12.2007)

PCT

(10) International Publication Number
WO 2007/148122 A2

(51) International Patent Classification: Not classified

(21) International Application Number: PCT/GB2007/002358

(22) International Filing Date: 25 June 2007 (25.06.2007)

(25) Filing Language: English

(26) Publication Language: English

(30) Priority Data:
0612479.6 23 June 2006 (23.06.2006) GB
0612728.6 28 June 2006 (28.06.2006) GB

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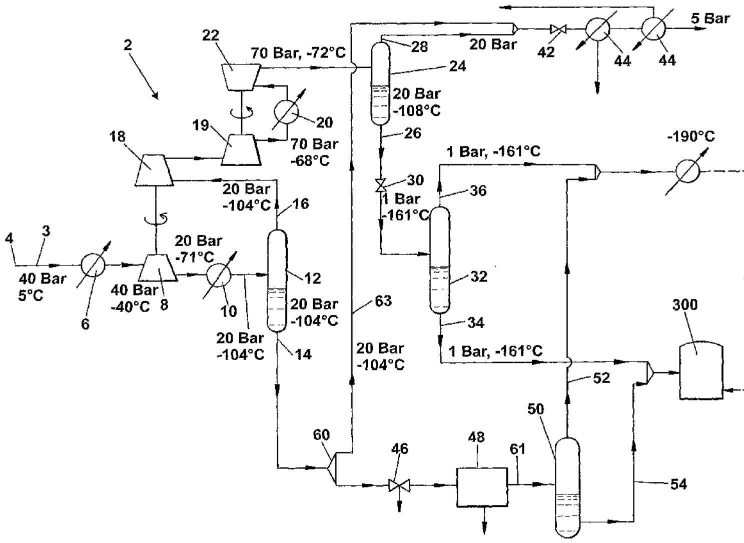
(81) Designated States (*unless otherwise indicated, for every kind of national protection available*): AE, AG, AL, AM, AT, AU, AZ, BA, BB, BG, BH, BR, BW, BY, BZ, CA, CH, CN, CO, CR, CU, CZ, DE, DK, DM, DO, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, GT, HN, HR, HU, ID, IL, IN, IS, JP, KE, KG, KM, KN, KP, KR, KZ, LA, LC, LK, LR, LS, LT, LU, LY, MA, MD, MG, MK, MN, MW, MX, MY, MZ, NA, NG, NI, NO, NZ, OM, PG, PH, PL, PT, RO, RS, RU, SC, SD, SE, SG, SK, SL, SM, SV, SY, TJ, TM, TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, ZA, ZM, ZW.

(84) Designated States (*unless otherwise indicated, for every kind of regional protection available*): ARIPO (BW, GH, GM, KE, LS, MW, MZ, NA, SD, SL, SZ, TZ, UG, ZM, ZW), Eurasian (AM, AZ, BY, KG, KZ, MD, RU, TJ, TM), European (AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR, GB, GR, HU, IE, IS, IT, LT, LU, LV, MC, MT, NL, PL, PT, RO, SE, SI, SK, TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW, ML, MR, NE, SN, TD, TG).

Published:
— *without international search report and to be republished upon receipt of that report*

For two-letter codes and other abbreviations, refer to the "Guidance Notes on Codes and Abbreviations" appearing at the beginning of each regular issue of the PCT Gazette.

(54) Title: LNG PRODUCTION



(57) Abstract: A method for producing Liquefied Natural Gas (LNG) at a Pressure Reduction Station (PRS) comprising the steps of: a) expanding an input stream of natural gas from a gas transmission system; b) separating the input stream of gas into a first liquid stream containing heavy hydrocarbons, and a first gas stream that is rich in methane; c) expanding the first gas stream; d) separating the first gas stream into a spent gas stream and a second liquid steam; e) expanding the second liquid stream; f) separating the second liquid stream into second gas stream and a first LNG product stream.

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LNG PRODUCTION

This invention relates to the production of liquefied natural gas (LNG), and particularly, but not exclusively, to the production of LNG from natural gas distributed at a pressure reduction station (PRS). The invention relates to a method for the production of LNG, to a plant for producing LNG, and to a PRS including a plant for producing LNG.

LNG offers an energy density comparable to petrol and diesel fuels and produces less pollution, but suffers from a relatively high cost of production due to the high electricity consumption necessary in conventional process manufacture.

In known manufacturing processes, the conditions required to condense natural gas depend on its precise composition, the market that it will be sold to, and the process being used. Typically however temperatures of between -100 and -170°C and pressures of between 101 and 6000 kPa are required.

For many applications, it is necessary to produce LNG with high purity. In particular, in order to produce LNG suitable for use as a fuel for vehicles (VLNG) it is necessary to produce LNG having a low content of heavy hydrocarbons ("heavies"). The term "heavies" or "heavy hydrocarbons" and is used herein to describe hydrocarbons having two or more carbon atoms per molecule.

In order to produce LNG having high purity with a low heavies content comprising less than 5% ethane and more preferably less than 3.5% ethane, it is known to be necessary to initially purify a natural gas inlet stream by removing as much ethane as possible.

It is known that natural gas pipeline systems such as National Transmission Systems (NTS) operate at typically three different pressure tiers, a high pressure of approximately 85-70 bar, an intermediate pressure 50-30 bar and a low pressure of approximately 7 bar. In a situation where pipelines carrying natural gas at high pressure feed into an intermediate pressure network, the pressure of the natural gas must be reduced from the high pressure to the intermediate pressure. Similarly, where pipelines carrying natural gas at the intermediate pressure feed into a network of low pressure, the pressure of the natural gas must be reduced from the intermediate pressure to the low pressure. The reduction of pressure is carried out at pressure reduction stations (PRS) which form part of the NTS.

It is known that as the pressure of natural gas reduces, the gas expands. As the natural gas expands, the temperature of the gas decreases. The temperature of the gas may drop to significantly below 0°C. Such a drop in temperature may cause damage to the pipework due to, *inter alia*, ground frost heave at the PRS and induce mechanical stresses leading to possible pipeline fracture.

In order to avoid such damage it is known to preheat the natural gas, for example, passing pipes carrying the natural gas through bath heaters or other heat exchange devices before pressure reduction takes place. Preheating is often applied where the pressure drop across the pressure reduction valve exceeds 14 bar (with a incoming gas temperature of 5 °C).

The overall pressure drop influences the need for preheating. In addition, initial ground temperature, and hence incoming gas temperature is also a factor. In the UK, ground temperature is typically 5°C all year round. However, in tropical climates the ground temperature may be as high as 30°C.

In addition, the type of expansion device used to expand the gas also determines the level of preheating required. Gas expanded by a turbo expander requires more preheating than does gas expanded by a valve, due to the isentropic expansion which occurs when gas is expanded in a turbo expander.

The required pressure drop depends on the upstream gas pressure within the national transmission system, and also the required delivery pressure which is determined by the use to which the gas is to be put. In a typical PRS, a large amount of energy is lost in bringing the natural gas pressure down to a desirable level. This energy may be regarded as a cooling load and could be harnessed in a number of different ways.

According to a first aspect of the present invention there is provided a method for producing liquefied natural gas (LNG) from pressurised natural gas comprising the steps of:

expanding an input stream of the high pressure natural gas;

separating heavy hydrocarbons from the natural gas to produce an LNG product.

The pressurised natural gas may be obtained from any source, but preferably the natural gas is obtained from a pressure reduction station (PRS) forming part of a National Distribution System.

Preferably, the LNG product has a low ethane content.

According to a second aspect of the present invention there is provided a method for producing liquefied natural gas (LNG) from natural gas at a Pressure Reduction Station (PRS) forming part of a gas transmission system, the method comprising the steps of:

expanding an input stream of natural gas from the gas transmission system;

separating heavy hydrocarbons from the natural gas to produce an LNG product that has a low ethane content.

Expansion of the input stream of natural gas results in a reduction in the temperature of the gas as well as a reduction in the pressure of the gas.

The pressure of the input stream of gas from the gas transmission system will vary depending on the parameters under which the PRS is operating. However, typically for example, under medium pressure conditions, the inlet stream of natural gas will be at a pressure of say, 40 bar and a temperature of 5°C. After expansion the gas will be at a lower temperature and pressure. This means that the gas will be closer to its saturation curve and if two phases exist may be separated by, for example a gas/liquid separator, into a methane rich gaseous product stream and a liquid stream containing heavy hydrocarbons.

Under other circumstances, the input stream of gas may be at a pressure of say 85 bar, and it will then typically be necessary to reduce the pressure of the gas to about 40 bar for onward transmission.

Where ground temperature is higher than 5°C, the temperature of the input stream of natural gas will be higher.

The expansion of the gas therefore allows separation of a heavies rich liquid and a methane rich gas stream - which in turn can be liquefied to produce a low heavies content LNG if required. Such LNG is suitable for use as a vehicle fuel and is known as VLNG.

According to a third aspect of the present invention there is provided a method for producing Liquefied Natural Gas (LNG) at a Pressure Reduction Station (PRS) comprising the steps of:

a) expanding an input stream of natural gas from a gas transmission system;

b) separating the input stream of gas into a first liquid stream containing heavy hydrocarbons, and a first gas stream that is rich in methane;

c) expanding the first gas stream;

d) separating the first gas stream into a spent gas stream and a second liquid steam;

e) expanding the second liquid stream;

f) separating the second liquid stream into second gas stream and a first LNG product stream.

The methods according to aspects of the present invention utilise the cooling and power produced when pressure reduction is called for at a pressure reduction station.

The method according to the third aspect of the present invention utilises low temperature separation (LTS) to purify and refine the high pressure (HP) natural gas into an LNG product and a low pressure (LP) gas in the form of a spent gas stream. The spent gas stream may be piped for onward transmission to customers *via* the downward distribution main pipeline, if it is of the correct specification.

The terms “low temperature”, “high pressure” and “low pressure” are used to describe a temperature and pressure, of a gas relative to the temperature and pressure of the input stream of natural gas.

Low temperature separation takes place at a temperature and pressure which is near to the dew point of the particular gas composition.

The ratio of pressure of the input stream of natural gas to the spent gas stream is preferably at least 3:1 but could be for example 7:1 or 8:1. It is likely that a pressure ratio of 2:1 would not be sufficient.

The process according to the invention is a continuous process adapted to produce LNG on a continuous basis. LNG produced by means of the invention may be removed (e.g. by road tankers keeping stored inventories low) and/or stored.

The process according to the invention is adapted to receive the entire output of the PRS station if this is required. Spent gas may be piped for onward transmission via the downward distribution main pipeline thus increasing the efficiency of the invention.

The process of the invention makes use of the cooling which is inherent when natural gas is expanded, and thus carried out at low temperature. Typically the expansion will be carried out within a temperature ratio of -100°C to -160°C.

The gas may be expanded using either a conventional Joule-Thomson (JT) valve or a turbo-expander (TE).

It is known to use JT valves at PRSs.. The use of a turbo-expander provides an opportunity to create shaft power, and also to achieve a greater temperature drop in the gas, due to the fact that JT valves are isenthalpic devices, whereas turbo-expanders are isentropic.

Previously, the greater temperature drop experienced by turbo-expanders to expand natural gas was regarded as a disadvantage, because more preheating is required in order to bring the gas back up to the required temperature.

However, in the field of LNG production, it is an advantage to have gas at a lower temperature. This results in a more efficient process which requires less energy input since the gas temperature and pressure is closer to the saturation curve (or liquefaction point) for the particular gas composition.

Preferably, the input stream of natural gas is at a temperature and pressure that this substantially on the saturation curve of the natural gas following expansion through step a). In addition, the first gas stream is preferably at a temperature and pressure that is substantially on its saturation curve following expansion at step c), and the second liquid stream is preferably at a temperature and pressure that is substantially on its saturation curve following expansion at step e).

It has been found that liquefied natural gas production yields may be maximised if the temperature and pressure conditions to which the natural gas is subjected are kept as close to the saturation curve of the particular gaseous mixture throughout the LNG production process.

A saturation curve for a particular composition shows graphically on a pressure temperature graph, the temperature at a particular pressure at which a liquid will boil (boiling point), and also at which a gaseous composition will liquefy (dew point).

Figures 1 and 2 show schematically the saturation curve for a natural gas composition, and pure methane respectively.

The triple point is the point at which all three phases (gas, liquid and solid) can exist together, and the critical point is the point beyond which pressure can no longer be used to liquefy a gas.

As can be seen from Figure 2, for pure methane, the critical point is about 45 bar and -85°C.

The applicant has realised that if the temperature and pressure of the natural gas are maintained as close to the saturation curve as possible for the particular composition of gas, then under pressure reduction, (in other words a path on the pressure-temperature graph to a lower point) liquid production is maximised.

An equation of state based approach is used to calculate the incipient solid formation point for mixtures containing carbon dioxide. The model can be used for predicting the initial solid formation point in equilibrium with either vapours or liquids. The fugacity of the resultant solid is obtained from the known vapour pressure of solid CO_2. The fugacity of the corresponding phase (in equilibrium with the solid) is calculated from the equation of state.

CO_2 solids prediction is restricted to the Peng Robinson (PR) and Soave Redlich Kwong (SRK) equationsof state.

For natural gas, the CO_2 solidification or freeze out point as a function of CO_2 concentration is set out below.

Percentage of CO_2	**Temperature at which freeze-out occurs**
6%	100°C
5%	103°C
4%	107°C
3%	111°C
1%	131°C
.3%	144°C
200 ppm	161°C

This means that for conventional LNG at -161°C, CO_2 needs to be removed down to 200 ppm or lower to stop freeze-out occurring which will block equipment. If, however, the gas is kept at a higher temperature, a higher percentage of CO_2 may be held in the gaseous composition before freeze out of CO_2 occurs.

Advantageously, the method comprises the further step of maintaining the temperature and pressure of the input stream of nature gas at a temperature that prevents CO_2 freeze out.

CO_2 if present may thus be removed from the first gas stream absorbed in the first high pressure liquid stream and then removed from the first liquid stream if required.

By means of the present invention, therefore, through careful control of the pressure and temperature conditions, it is thus possible to remove components such as carbon dioxide - which could potentially 'freeze-out' as solids and cause blockages in the process equipment. In the present invention this is avoided by virtue of the fact that CO_2 has a higher solubility in a high pressure intermediate LNG produced - typically 5% by volume at 10 to 20 bar. Conventional LNG at close to atmospheric pressure can only hold a few hundred parts per million CO_2 in solution before freeze-out occurs.

Advantageously, the method comprises the further step of cooling the input stream of natural gas before carrying out step a).

The use of pre-cooling before expansion facilitates maintenance of the gas at a temperature and pressure which is at or close to the saturation curve of the particular gas composition. This in turn increases the efficiency of the method.

Preferably, the method comprises the further step of further cooling the input stream between steps a) and b).

The step of cooling the input stream further assists in the maintenance of the gas at or close to the saturation curve of the particular gas composition.

Conveniently, the method comprises the further step of compressing the first gas stream after carrying out the separation of step b) and before the expansion of step c). The compression of the first gas stream at this stage causes an increase in the temperature and pressure of the first gas stream prior to expansion. It is advantageous to compress and then re-expand the first gas stream in this way in order to liquefy the first gas stream to produce high purity LNG.

If the expander is a turbo-expander, the compression may be carried out by a compressor which is partially or completely powered by shaft power from the expander. In other words, energy created through the expansion of the gas may be used to power the compressor, either completely, or partially.

The method according to aspects of the present invention may be operated such that there is a neutral or close to neutral energy balance. This may be achieved by maximising the use of the available pressure energy and cooling/shaft power generated from pressure let down of the input stream of natural gas which exits the PRS. In addition, the method of the invention obviates, or reduces, the need to use additional heat or power to produce LNG.

The method of the present invention splits the input stream of natural gas into two streams before the LNG product is produced. An advantage of having two parallel streams formed prior to LNG production is that shaft power, energy and cooling can be produced from the depressurisation of the non-LNG process stream (i.e., the first liquid stream) which can be used to supply the other stream (i.e., the first gas stream) if appropriate.

Advantageously, the method comprises the further step of cooling the first gas stream before carrying out the expansion of step c). Again, the additional step of cooling facilitates the maintenance of the gas at or close to its saturation curve.

Conveniently, the method comprises the further step of mixing at least a portion of the first liquid stream with the first LNG product stream to produce LNG having a desired composition.

Since the first liquid stream has a high content of heavy hydrocarbons, the mixing of the first LNG product stream with the first liquid stream will alter the composition of the first LNG product stream. The LNG product stream and the first liquid stream may thus be mixed in an appropriate proportions in order to produce an LNG having a desired composition.

Alternatively, the method comprises the further step of removing at least a proportion of the heavy hydrocarbons contains in at least a portion of the first liquid stream; removing CO_2 from the at least a portion of the first liquid stream; and condensing the resulting liquid stream to form a third liquid stream and a third gas stream.

The hydrocarbon content of the third liquid stream may thus be controlled by controlling the amount of heavies removed from at least a portion of the first liquid stream prior to the first liquid stream, or a path thereof being condensed.

Preferably, the method comprises the further step of mixing the third liquid stream with the first LNG product stream to produce LNG having a desired composition.

Preferably, the method comprises the further step of mixing the third gas stream with the second gas stream to produce a low pressure, low ethane gas.

The resultant low pressure, low ethane gas may be liquefied to form a second LNG product stream.

Advantageously, the method comprises the further step of blending the first liquid stream with the spent gas stream to produce a blended gas having a desired composition.

The blended gas may then be piped into the main gas distribution system. In order to be so introduced, it must have the appropriate composition. A suitable blending of the first liquid stream with the spent gas stream will produce a blended gas having appropriate composition.

By means of the present invention therefore the composition of the LNG produced may be varied to produce an LNG product having a desired composition depending on the use to which the LNG is to be put.

In addition, the composition of the spent gas may be varied to produce a gas suitable for piping into the main gas distribution system.

According to a fourth aspect of the present invention there is provided a plant for production of LNG from natural gas at a PRS, the plant comprising:

an expander for expanding the natural gas, and a separator for separating out heavy hydrocarbons in liquid form from the gas.

According to a fifth aspect of the present invention there is provided a pressure reduction station comprising a plant for producing LNG from natural gas, the plant comprising:

an expander for expanding the natural gas, and a separator for separating out heavy hydrocarbons in liquid form.

The plant may comprise a plurality of expanders connected in parallel to one another such that one or more expanders may be operated at the same time depending on the level of gas flow. This may increase the efficiency of the plant.

According to a sixth aspect of the present invention there is provided a plant for producing liquefied natural gas (LNG) comprising:

a) a first expander for expanding the input stream;

b) a first separator for separating the input stream of gas into a first liquid stream containing heavy hydrocarbons, and a first gas stream which is rich in methane;

c) a second expander for expanding the first gas stream;

d) a second separator for separating the first gas stream into a spent gas stream and a second liquid stream;

e) a third expander for expanding the second liquid stream;

f) a third separator for separating the second liquid stream into a second gas stream and a first LNG product stream.

The expander may be any suitable type of expander, but preferably the expander comprises a turbo-expander such as a radial inflow type or similar.

By means of the present invention therefore the composition of both the LNG and the low pressure low ethane gas which constitutes a take off gas can be varied through use of a system of mixers/blenders to produce products having appropriate compositions.

The plant according to aspects of the invention may be connected in parallel with a conventional PRS plant. The plant of the invention may take some, or all, of the gas destined for the PRS station from a national distribution system, depending on prevailing requirements and conditions.

A turbo-expander is particularly suitable for obtaining appropriate temperature drops through expansion of the gas, which temperature drops are advantageous in the production of LNG and for generating shaft power in the process to improve process energy efficiency.

Through use of turbo-expanders, it is possible to expand the gas such that it is at or close to its saturation curve, thus rendering the gas substantially saturated.

The expansion of the natural gas may be harnessed as shaft power as the natural gas is expanded by the turbo-expander.

Advantageously, the plant further comprises a first cooler for cooling the input stream of natural gas before it is expanded by the first expander.

The first cooler may use a cold gas product stream to cool the inlet stream of natural gas.

Advantageously, the plant comprises a second cooler for cooling the input stream of natural gas after it has been expanded by the first expander.

The second cooler may also use heat exchange from a cold gas product stream in order to cool the inlet stream of natural gas.

Advantageously, the plant further comprises a first compressor for compressing the first gas stream after it has been produced by the first separator and before it has been expanded by the second expander.

The first compressor may be powered partially or entirely by shaft power created through the expansion of the inlet stream of natural gas when it is expanded by the first expander.

In some embodiments of the invention, the plant may comprise two compressors, the first compressor and a second compressor working in series with one another.

Each of the first and second compressors is directly coupled to an expander operating in a different part of the plant. Pressure ratio limitations may come into play if only one expander were used. An expander typically achieves a 3:1 pressure ratio.

Advantageously, the second compressor is partially or completely powered by shaft power produced by the expansion of the first gas stream through the second expander.

Preferably the plant comprises a third cooler for cooling the first gas stream prior to expansion by the second expander.

Advantageously, the plant further comprises a first mixer for mixing the first LNG product stream with at least a portion of the first liquid stream to produce LNG having a desired composition.

Advantageously, the plant further comprises a tee positioned downstream of the first separator for splitting the first liquid stream into a first liquid sub-stream and a second liquid sub-stream. In such an embodiment, the first liquid sub-stream is mixed with the first LNG product stream via the first mixer, and the first liquid stream does not mix directly with the first LNG product stream.

Preferably, the plant further comprises a valve for removing at least a proportion of the heavy hydrocarbons contained in the second liquid sub-stream.

Conveniently, the plant further comprises a separation device for removing carbon dioxide from the second liquid sub-stream.

Advantageously, the plant further comprises a fourth separator positioned downstream of the tee for condensing the first liquid sub-stream to form a third liquid stream and a third gas stream.

Conveniently, the plant further comprises a second mixer for enabling mixing of the third gas stream with the second gas stream to produce a low pressure methane rich gas stream.

Advantageously, the plant further comprises a liquefier for liquefying the low pressure methane rich gas stream to form a second LNG product stream.

Conveniently, the plant further comprises a third mixer for allowing mixing of the first liquid sub-stream with the spent gas stream.

The invention will now be further described by way of example only with reference to the accompanying drawings in which:

Figure 1 is a schematic representation of a saturation curve for a typical nature gas composition;

Figure 2 is a schematic diagram of a saturation curve for pure methane gas;

Figure 3 is a schematic representation of a plant for producing liquefied natural gas from natural gas produced at a pressure reduction station (PRS) according to an embodiment of the invention;

Figures 4, 5 to 6 are schematic representations of the further of plants according to embodiments of the invention;

Figure 7 is schematic representation of a plant for producing liquefied natural gas according to a second embodiment of the invention; and

Figure 8 is a schematic representation of an extension to the expansion cycle shown in Figure 7.

Referring to Figure 3, a plant for producing liquefied natural gas from natural gas distributed at a pressure reduction station (PRS) is designated generally by the reference numeral 2.

The plant 2 enables a continuous process to be carried out to produce LNG at desired rates of production.

The flow rate is chosen to obtain a desired output of 679 kilograms per hour of LNG. This output has been specified as being a reasonable daily output for commercial reasons. The input flow is thus calculated on the basis of the efficiency of the process, which in turn depends on the purity requirement of the LNG. In the illustrated embodiments an inlet stream flow of 3461 kilograms per hour is required since the process is approximately 20% efficient. The efficiency of the system is determined by factors such as the overall pressure ratio across the system as a whole which in turns determines how much cooling can be generated to manufacture LNG and increase yields.

The system may be formed integrally with a PRS, or may be formed separately therefrom. If the system is formed separately to the PRS, it must be fluidly connected to the PRS. Preferably however, the plant is embedded in the PRS.

The plant 2 is particularly appropriate for production of LNG suitable for vehicle use. Such LNG must have a low ethane content, typically of less than 3.5%.

Natural gas exiting a PRS enters the plant 2 as an inlet stream of natural gas 3 at inlet 4. The inlet stream of natural gas is then cooled using a first cooler 6. This allows traces of any condensed water to be removed and ensures that the input stream of gas enters the plant as cool as possible. In the illustrated embodiment, the inlet stream of natural gas enters the plant 2 at a temperature of approximately 5°C and a pressure of approximately 40 bar. The cooler 6 then cools the input stream to approximately -40°C.

At this stage traces of water maybe removed since under these conditions of temperature and pressure water will be condensed out of the gas. The gas is however substantially dry from upstream gas processing

The input stream of natural gas is then expanded by expander 8, in this case to a pressure of 20 bar, and a cooled by a second cooler 10 to a temperature of -105°C. The expander will be a Joule-Thomson valve or a turbo-expander. In this example the expander 8 is a turbo-expander, and expansion of the gas creates energy which may be harnessed in the form of shaft power. The temperature and pressure conditions are such that the gas is saturated. It is therefore possible to condense the gas in separator 12, into a first liquid stream 14 containing predominantly heavy hydrocarbons (heavies) and a first gas stream 16 which is rich in methane. In this example, stream 16 has a content of 96.9%.

Any CO_2 contained in the input stream of natural gas will be absorbed into the liquid stream 14 as CO_2 has a greater absorption potential/affinity with the gas liquids produced at elevated temperatures and pressures (at 3 to 6 MOL% CO_2), compared to atmospheric pressure LNG at -160°C where CO_2 removal to 50 to 100 parts per million is required.

The methane rich first gas stream 16 is then recompressed by compressors 18, 19. Compressor 18 may be driven by shaft power from turbo-expander 8. The gas stream 16 is then cooled by a third cooler 20. The cooler 20 may be supplied by a cold gas product stream.

The high methane first gas stream 16 is then expanded using a second turbo-expander 22. Shaft power from this expander may be used to power (or part power) compressor 19. Following cooling and expansion, the methane rich gas stream 16 is at a temperature of -108°C and a pressure of 20 bar. The methane rich gas stream is thus saturated, and it is possible to separate the first gas stream 16 into a second liquid stream 26 and a spent gas stream 28 using separator 24 spent gas which emerges from the separator 24 at a pressure of 20 bar.

The first liquid stream 14 containing heavy hydrocarbons may be mixed with the spent gas 28 in appropriate proportions in order to produce a low pressure gas having an appropriate Wobbe Index to enable the spent gas to be piped into the downstream low pressure gas main.

Before being introduced into the gas main, the blend of spent gas and heavy hydrocarbons is expanded through a Joule-Thomson valve 42 and is then heated (directly or via heat exchange from inlet gas stream) by means of one or more heaters 44 to ensure that the gas mixture is at an appropriate temperature and pressure to be introduced into the gas main.

The second liquid stream 26 is expanded through Joule-Thomson valve 30 such that the liquid is at a pressure of 1 bar and a temperature of approximately -160°C. The liquid is thus saturated and may then be separated by separator 32 into an LNG product stream 34 at 1 bar pressure and approximately -160°C, and a methane rich low pressure (LP) second gas stream 36 at the same pressure and temperature. The LNG product stream 34 may be conveniently stored for example in a tank 300.

The second gas stream 36 is rich in methane and may be liquefied using any convenient process such as a nitrogen loop system or a small mixed refrigerant vapour based liquefaction plant to produce a second high purity LNG product stream 40 which may be stored in tank 300.

First liquid stream 14, may be divided by tee 60 into a first liquid sub-stream 61 and a second liquid sub-stream 63. This means that, as well as being mixed with the spent gas 28, the first liquid stream 14 may also be routed through a valve 46 for separating out some or all of the heavy hydrocarbons contained in the liquid and valve 48 for removing CO_2 as a gas or solid from the liquid stream. Once the CO_2 and any heavies have been removed from the first sub-stream 61 the stream passes through separator 50 where it is separated into a third gas stream 52 which is high in methane content and a third liquid stream 54 containing heavy hydrocarbons. The level of hydrocarbons in the third liquid stream 54 may be adjusted as required using valve 46.

The stream 54 is mixed with the first LNG product 34 to produce LNG having an appropriate level of hydrocarbons in order to ensure the required composition is achieved.

The expanders 8, 22 are, in this example in the form of turbo-expanders which can generally be used for pressure ratios of 3 or 4 to 1. The pressure requirement of the process illustrated in Figure 3 is 2 to 3.5:1.

The expanders should not have any liquid in the inlet stream into each expander, but some condensation within the expander may be tolerable.

The shaft power produced by expanders 8 and 22 corresponds approximately to that required by compressors 18 and 19. However the power may be augmented by power produced from an electrically driven compressor or gas engine driven compressor.

The heat taken out of the spent gas stream 28 may be used to cool the input gas stream 3 at coolers 6, 10, by means of heat exchangers. The heat exchangers may be of any suitable type, such as a shell and tube type.

Natural gas is assumed to have the following molar composition:

methane 93.1024
ethane 3.7041
propane 0.5806
iso-butane 0.3504
N-butane 0.0801
150-pontane 0.2002
nitrogen 1.6618
carbon dioxide 0.3204

Natural gas may however have a different composition under different circumstances.

By means of the present invention the entire flow from a PRS can be handled by the plant 2 if required.

Although the invention has been illustrated in Figure 1 as operating at particular temperatures and pressures, it is to be understood that the invention may be applied to plants and methods operating at different temperatures and pressures.

Figures 4 to 6 are schematic representations of plants according to further embodiments of the invention. Parts of the plants shown in Figures 4 to 6 that correspond to parts shown in the plant illustrated in Figure 3 have been given corresponding reference numerals for ease of reference.

Referring initially to Figure 4, a plant 400 is shown.

The plant 400 differs from the plant illustrated in Figure 3 in that compressors 18, 19 shown in Figure 3 have been replaced by a single compressor 200. Each of the expanders 8, 22 is linked to the single compressor 200. This may be achieved in practice by, for example, the shafts from the expanders 28, 22 entering a gear box adapted to drive a single compressor 200.

Turning now to Figure 5 a plant 500 is shown which has the same components as the plant 400 illustrated in Figure 4. In this example the inlet stream gas 3 enters the system at temperature of 5°C and a pressure of 20 bar. The gas exits as spent gas 28 at a pressure of 3 bar and a temperature of 7°C. The temperatures and pressures throughout the processor differ from those shown in Figures 3 and 4 due to the fact that the inlet stream enters the system at a different pressure (20 bar).

Turning now to Figure 6 another embodiment of the plant according to an aspect of the invention is designated generally by the reference numeral 600. The plant 600 has the same components as plants 400 and 500 illustrated in Figures 4 and 5 respectively. In the embodiment shown in Figure 6, the input stream 3 enters the system at a pressure of 70 bar and a temperature of 5°C, and exits as spent gas at a pressure of 20 bar and a temperature of 5°C. The temperatures and pressures existing throughout the process differs from those shown in Figures 3, 4 and 5 due to the fact that the input gas stream enters the system at a pressure of 70 bar.

Figure 7 depicts a basic LNG scheme, operable with or without chilled fluid production. More specifically, Figure 7 schematic represents a simplified expansion cycle gas/liquids production technique according to an embodiment of the invention using flow pressure energy available at a PRS but which is normally wasted or unutilised.

With an existing PRS facility, the entry and exit gas pressures from the process are often fixed by the requirements for upstream and downstream pressure and gas flow (i.e., the pressure ratio is fixed). Even with an existing or 'brown field' site, a network operator may be able to modify this pressure ratio to some extent, but subject to constraints of pipeline pressure limits, capacity and required gas flows to be delivered to the downstream customers.

For a green field PRS site, there will be greater flexibility to determine pressure ratio(s) beforehand. If low-pressure gas is required, it is feasible to deliver this with a single stage pressure drop which increases the cooling load potential of an individual site.

The various stages in the LNG process of Figure 7 are now discussed.

High-pressure (HP) gas undergoes pressure reduction and partial liquefaction before entering the process at Point 1. If the gas is tapped directly from the national transmission system (NTS), its pressure might typically be 60 to 85 bar (or atmospheres). If the gas is tapped from an intermediate tier pressure system, its pressure might be around 35 to 40 bar (or atmospheres).

In either case, in this embodiment, the gas has firstly to be pre-treated to remove most of the water and carbon dioxide before liquefaction. This is required as these substances can form solids when cooled, potentially resulting in blockage of the equipment.

Following pre-treatment, HP gas enters the recycle heat exchanger (e.g. a shell and tube type, with a series of knock out pots or a carefully designed gas liquids fractionation column) at Point 2.

The heat exchanger pre-cools the incoming HP gas, using a recycle stream of spent cold low-pressure (LP) gas not liquefied following the various expansion stages (about 85-90% of the total flow) at Point 3. This LP gas (on the shell side of the heat exchanger) is fairly close to the LNG temperature of -160°C and pre-cools the incoming HP gas on the tube side of the heat exchanger.

The HP gas flowing counter-current through this heat exchanger progressively cools as it flows through the heat exchanger. The higher hydrocarbon components within the natural gas (these include butane, propane, ethane, etc.) start to liquefy and separate/drop out from the main gas stream as liquids (the gas typically comprising around 93% methane on entry).

A series of knock out pots collect these higher hydrocarbon gas liquids as they drop out of the gas stream after passing through a baffle and demister pad (Points 4a, b, c…). These points are at high pressure (tube side) and have a pressure reduction valve and return leg system to the low pressure side of the gas stream on the outlet shell side of the heat exchanger (Points 5a, b c…..).

The higher hydrocarbons thus collect in these pots, but are also simultaneously 'flashed' back into the low pressure side in a controlled way, so as to maintain a LP gas composition similar to that of the entry HP gas stream.

Some liquids can also be tapped off (for onward sale) to allow for the fact that around 10% to 15% liquid methane (>95mol% pure) is extracted at the end of the liquefaction process. This higher hydrocarbon 'tapping process' ensures the LP gas is not over enriched in higher hydrocarbons, beyond acceptable limits for the LP gas quality.

The various 'knocks out pots' or 'fractionation trays' contain a mixture of higher hydrocarbons similar to liquefied petroleum gas (LPG). The composition will vary from pot to pot (or tray to tray) with the first being butane-rich (lowest dew point temperature) and the latter (expander end) being ethane-rich. Thus each gas-liquid product being tapped for onward sale may serve a different target market niche.

The HP gas entering the recycle heat exchanger will be at around ambient temperature conditions (5°C ground temperature at entry, plus an uplift of around 10-15°C from the adsorption gas treatment process). The HP gas will exit the heat exchanger (prior to the first expander inlet) at around -80°C at Point 6, when a steady state condition is achieved.

At process start-up the gas temperature will be transient at Points 3 and 6 and will become progressively cooler as gas expansion takes place. A steady-state condition is achieved between the incoming ambient temperature HP gas stream and the lower temperature LP recycle gas stream (as cold as -160°C when a steady-state is achieved); the recycle heat exchanger thus achieving a steady-state condition after a period of time.

The pre-chilled/cold HP gas stream at around -80°C enters the first expander at Point 6. Each turbo-expander stage is limited to a maximum expansion of about 2-3:1, owing to choking due to sonic/speed of sound flow and rotor speed limitations. The first stage expansion would take the gas temperature down about 35°C from -80°C at the inlet to approx -115°C at the first expander outlet Point 7.

The outlet at Point 7 from the first expander forms the inlet to the second expander. The gas then undergoes a similar second stage expansion, whereby the exit gas temperature at Point 8 falls to around -145°C.

A Joule-Thomson valve finally reduces the gas temperature from Point 8 to some -160°C at Point 9 where significant (around 10% by mass) liquid (LNG) drop-out takes place into the collection vessel. This vessel also separates out the gas and liquid phases via a baffle plate arrangement and demister pad. The resulting LNG is rich in methane (>95mol%) having separated out the heavies in the earlier recycle heat exchanger/fractionation column. The LNG is then stored and can be removed by road tanker, for onward sale to markets. The cold LP gas (Point 3) is returned to the recycle heat exchanger as shown.

The cold LP gas at Point 3 can then be (a) recompressed and returned to the process (closed loop) or 'dumped' into the low pressure distribution main for onward transmission to LP gas customers. Before this cold LP gas is 'dumped' it can be fed through an indirect heat exchanger (Point 10) to recover thermal or cooling energy. This could be a glycol/water mixture supplying cooling for cold storage applications, district cooling, ice manufacture etc.

Finally, an optional compressor could be installed, driven by power from the turbo-expander shaft(s) to recompress the gas, if the downstream exit gas pressure requirement is above that of the exit heat exchanger gas pressure at Point 11.

Otherwise the shaft power derived from the gas expansion process can be used to drive an alternator for electrical power generation, or a vapour compression chiller to provide additional 'coolth' for ancillary chilling applications, or additional HP gas pre-chilling to improve overall LNG yields. Gas exits the system at Point 12.

Direct outputs from the Figure 7 process, or those generated indirectly include:

- Low temperature gas, a proportion of which (typically some 10-15%) is liquefied, given sufficient pressure ratio at a PRS;
- Mechanical shaft power from the expander(s);
- Chilled fluids generation (e.g. glycol/water) by way of indirect heat exchange by way of coolth recovery on the spent low temperature LP gas;
- Gas liquids production comprising higher hydrocarbons, part of which are flashed back into the LP gas to maintain downstream gas quality/composition (in 'open cycle' mode) and part of which can be tapped off for potential use/sale to markets outside of the process;
- Heat available from the turbo-expander(s) by way of fluidic braking if the available shaft power is not to be utilised for (6) and (7);
- Electrical power generation using alternator from available shaft power;
- Additional cooling/chilling by way of the vapour compression chiller driven from the available expander shaft power.

The degree of overall gas cooling (temperature depression) and ultimately liquefaction is dependant upon the pressure ratio across the process. The higher the pressure cut/ratio, the higher (greater) is the cooling potential.

In addition, pre-cooling by heat exchange of the incoming high-pressure gas using the cold unliquefied LP gas enables:

- Liquefaction to be achieved at a lower overall pressure drop, as the entry temperature into the first stage expander is much lower/colder than would otherwise be the case (i.e., the temperature drop required for liquefaction is less);

- Heat from the incoming HP gas to warm the 'spent' LP gas, which is to be 'dumped' into the downstream distribution main in an 'open cycle' approach.

Re-compression of spent LP gas is also possible using the available expander shaft power in order to:

- Form a completely 'closed cycle' system (i.e. deployed as an end of pipeline liquefier, where no LP gas is dumped into the downstream distribution system);
- Enable LP gas at the end of the process to be reduced in pressure somewhat below that required for the downstream distribution main and then, following liquefaction, re-compressing it to the desired pressure using available expander(s) shaft power,

The expanders may be off-the-shelf turbo-expanders. They would desirably feature a 'radial in-flow turbine', which can accept liquid drop-out without damage to the rotor. Most of the liquefaction would however take place downstream of the final stage Joule-Thomson (J-T) valve expansion of Figure 7.

Gas processing prior to liquefaction is required in this embodiment to ensure H_2O and CO_2 are within specification. Line gas is already treated upstream at the beach terminals, but remaining CO_2 and H_2O can be removed using the following techniques. The recycle heat exchanger will also remove residual water.

Water removal from the gas prior to liquefaction can be undertaken using standard techniques, such as:

Glycol dehydration (TEG and DEG);

Calcium chloride (CaC12) dehydration or lithium and barium chloride;

Dry desiccant dehydrators using molecular sieves, silica gel, activated alumina or activated carbon;

Membrane systems;
Ethylene glycol (MEG) injection;
Methanol (MeOH) injection.

Hydrogen sulphide, carbon dioxide and mercaptans can be removed from natural gas by several processes, including:

Amine treatment, utilising MEA, DEA, DGA, MDEA;
Molecular sieves;
Membranes;
Solvents such as sulfinol and ifpexol.

It is important that the LNG process meets the HCDP (hydrocarbon dew point control) specification of the pipeline operator, to ensure that no free hydrocarbon liquids accumulate in the pipelines.

The Applicant foresees the potential to recover LPG (liquefied petroleum gases), and the constituent products, such as ethane, propane, butane, pentane, etc.

This can be achieved using the plant illustrated in Figure 7, or separately via add-on low temperature refrigeration or modified lean oil absorption plant. Fractionation can be used to separate out the various alkane components. Namely de-methaniser, de-ethaniser, de-propaniser, de-butaniser. For a mixed gas condensate output, only a stabiliser is required.

Figure 3 represents an extension to the expansion cycle LNG process of Figure 7. More specifically, Figure 8 is a pressure cascade LNG scheme, suitable for installation in environments where large gas pressure drops are available, such as:

- Selective pressure reduction stations with high pressure ratios;
- Offshore or 'stranded' gas fields, where well pressures are high, yet it is uneconomic to lay on a conventional pipeline to bring gas to market. It may then be possible using such a system to monetise the asset and bring LNG/gas to market in an economic way.

The pressure cascade system is shown diagrammatically in Figure 8 with 3 stages, respectively high pressure (HP), medium pressure (MP) and low pressure (LP). The first high-pressure stage mirrors that already described for Figure 7. Spent gas leaving this initial process is re-compressed, using the shaft power available from the two expanders E1 and E2.

The re-compressed gas from the first stage then enters the intermediate or medium pressure stage, where it is expanded through expanders E3 and E4. No gas treatment is required, as this is all undertaken at the first stage. The spent gas is then re-compressed (as before) using available shaft power from E3 and E4, before entering the final low-pressure stage.

The re-compressed gas from the intermediate stage then enters the final or low pressure stage, where it is expanded through expanders E5 and E6. No gas treatment is required, as this is all undertaken at the first stage. Spent gas is then re-compressed as before, using available shaft power from E5 and E6, before onward passage into the distribution main, or by total re-compression in the case of a 'closed cycle' system perhaps utilising a gas engine back to the first stage entry point.

Additional re-compression can be provided by a gas engine, its fuel being the LP gas available at the end of the process and/or the LPG gas liquid from stage 1.

Intercoolers or heat exchangers can be installed between stages to increase efficiency and LNG yields. The cold spent gas from the final stage acts as a cooling medium for all stages.

CLAIMS

1. A method for producing Liquefied Natural Gas (LNG) at a Pressure Reduction Station (PRS) comprising the steps of:

a) expanding an input stream of natural gas from a gas transmission system;

b) separating the input stream of gas into a first liquid stream containing heavy hydrocarbons, and a first gas stream that is rich in methane;

c) expanding the first gas stream;

d) separating the first gas stream into a spent gas stream and a second liquid steam;

e) expanding the second liquid stream;

f) separating the second liquid stream into second gas stream and a first LNG product stream.

2. A method according to Claim 1 wherein step a) comprises the step of expanding an input stream of natural gas exiting the PRS to a temperature and pressure that is substantially on the saturation curve of the natural gas.

3. A method according to Claim 1 or Claim 2 wherein step c) comprises the step of expanding the first gas stream to a pressure and temperature that is substantially on the saturation curve of the gas forming the first gas stream.

4. A method according to any one of the preceding claims wherein step e) comprises the step of expanding the second liquid stream to a pressure and temperature that is substantially on the saturation curve of the liquid forming the second liquid stream.

5. A method according to any one of the preceding claims comprising the step of maintaining the temperature and pressure of the input streams to retain CO2 in the liquid 'heavies' phase

6. A method according to any one of the preceding claims comprising the further step of:

g) cooling the input stream of natural gas before step a).

7. A method according to any one of the preceding claims comprising the further step of:

h) further cooling the input stream between steps a) and b).

8. A method according to any one of the preceding claims comprising the further step of:

i) compressing, the first gas stream between steps c) and d).

9. A method according to any one of the preceding claims comprising the further step of cooling the first gas stream before step c).

10. A method according to any one of the preceding claims comprising the further step of mixing at least a portion of the first liquid stream with the first LNG product stream to produce LNG having an appropriate composition.

11. A method according to any one of Claims 1 to 10 comprising the following further steps:

i) removing at least a proportion of the heavy hydrocarbons contained in the at least a portion first liquid stream;

ii) removing CO_2 from the at least portion of the first liquid stream; and

iii) condensing the resultant liquid stream to form a third liquid stream and a third gas stream.

12. A method according to Claim 11 comprising the step of blending the third liquid stream with the first LNG product stream to produce LNG having an appropriate composition.

13. A method according to Claim 11 comprising the further step of mixing the third gas stream with the second gas stream to produce a low pressure low ethane gas.

14. A method according to Claim 13 comprising the further step of liquefying the low pressure gas prior to form a second LNG product stream.

15. A method for producing Liquefied Natural Gas (LNG) from pressurised natural gas comprising the steps of:

expanding an input stream of the high pressure natural gas;

separating heavy hydrocarbons from the natural gas to produce an LNG product.

16. A method for producing Liquefied Natural Gas (LNG) from natural gas at a Pressure Reduction Station (PRS) forming part of a gas transmission system, the method comprising the steps of:

expanding an input stream of natural gas from the gas transmission;

separating heavy hydrocarbon from the natural gas to produce an LNG product that has a low ethane content.

17. A plant for producing Liquefied Natural Gas (LNG) comprising:

a) a first expander for expanding the input stream of natural gas exiting the PRS;

b) a first separator for separating the input stream of gas into a first liquid stream containing heavy hydrocarbons, and a first gas stream which is rich in methane;

c) a second expander for expanding the first gas stream;

d) a second separator for separating the first gas stream into a spent gas stream and a second liquid stream;

e) a third expander for expanding the second liquid stream;

f) a third separator for separating the second liquid stream into a second gas stream and a first LNG product stream.

18. A plant according to Claim 17 wherein each of the first and second expanders comprises a turbo-expander.

19. A plant according to Claim 17 or Claim 18 further comprising a first cooler for cooling the input stream of natural gas before it is expanded by the first expander.

20. A plant according to any one of Claims 17 to 19 further comprising a second cooler for cooling the input stream of natural gas after it has been expanded by the first expander.

21. A plant according to any one of Claims 17 to 20 further comprising a first compressor for compressing the first gas stream after it has emerged from the first separator and before it is expanded by the second expander.

22. A plant according to Claim 21 further comprising a second compressor for compressing the first gas stream after it has been compressed by the first compressor and before it is expanded by the second expander.

23. A plant according to any one of Claims 17 to 22, further comprising a third cooler for cooling the first gas stream prior to the first gas stream being expanded by the second expander.

24. A plant according to any one of Claims 17 to 23 further comprising a first mixer for allowing mixing of at least a portion of the first liquid stream with the first LNG product stream.

25. A plant according to any one of Claims 17 to 24 further comprising a tee positioned downstream of the first separator for splitting the first liquid stream into a first liquid sub-stream and a second liquid sub-stream.

26. A plant according to Claim 25 further comprising a valve or splitter for removing at least a proportion of the heavy hydrocarbons contained in the second liquid sub-stream.

27. A plant according to Claim 25 or Claim 26 further comprising a valve or splitter for removing carbon dioxide from the second liquid sub-stream.

28. A plant according to any one of Claims 25 to 27 further comprising a fourth separator positioned downstream of the tee for condensing the second liquid sub-stream to form a third liquid stream and third gas stream.

29. A plant according to any one of Claims 17 to 28 further comprising a mixer for mixing the third gas stream with the second gas stream to produce a low pressure methane rich gas stream.

30. A plant according to Claim 29 further comprising a liquefier for liquefying the low pressure methane rich gas stream to form a second LNG product stream.

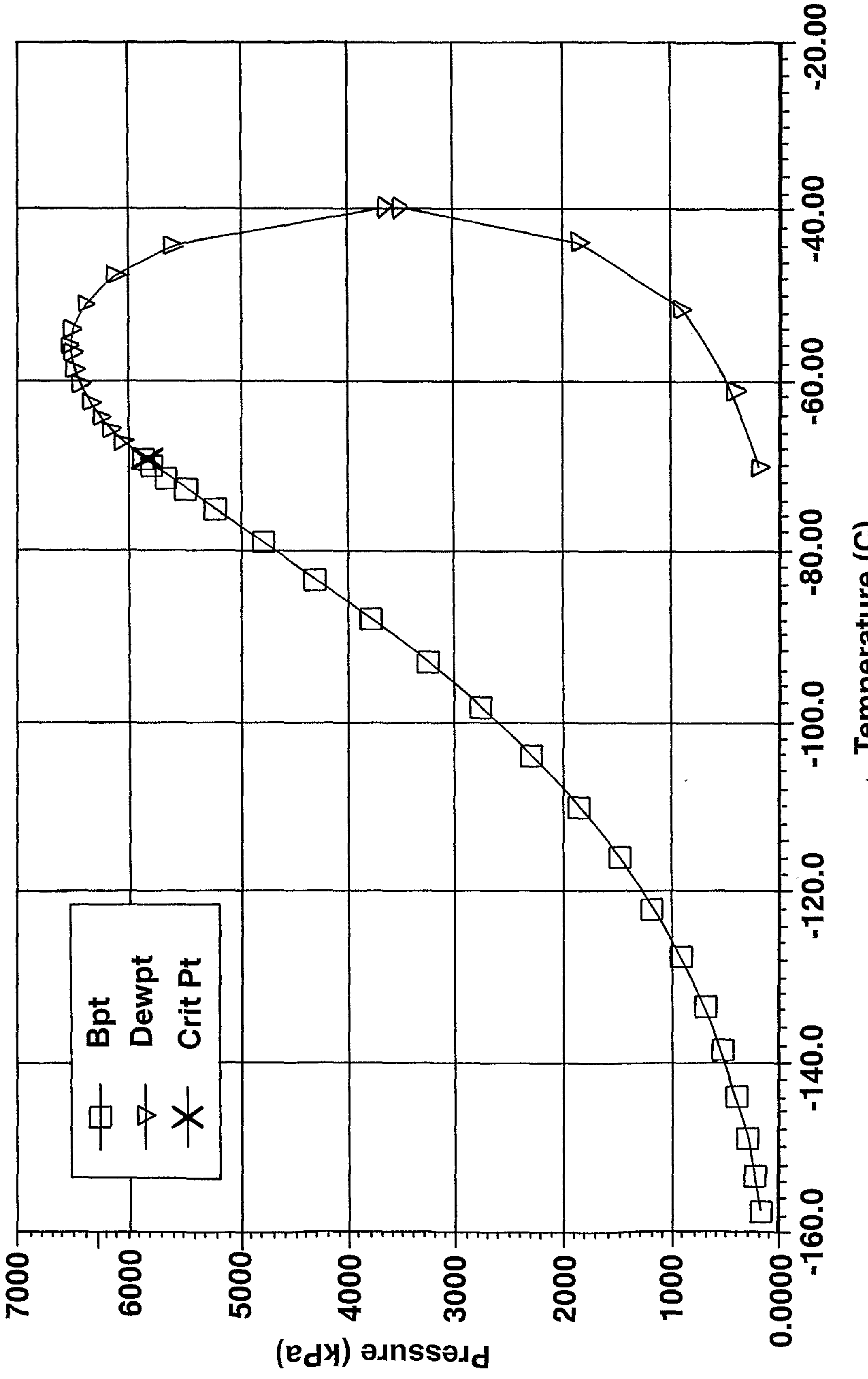
31. A plant according to any one of Claims 17 to 29 further comprising a third mixer for mixing the first liquid sub-stream with the spent gas stream.

32. A plant for production of LNG from natural gas comprising an expander for expanding the natural gas, and a separator for separating out heavy hydrocarbons in liquid form from the gas.

33. A pressure reduction station comprising a plant according to any one of Claims 17 to 33.

34. A method substantially as hereinbefore described with reference to the accompanying drawings.

35. A plant substantially as hereinbefore described with reference to the accompanying drawings.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4000
3000
2000
1000
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-140.0
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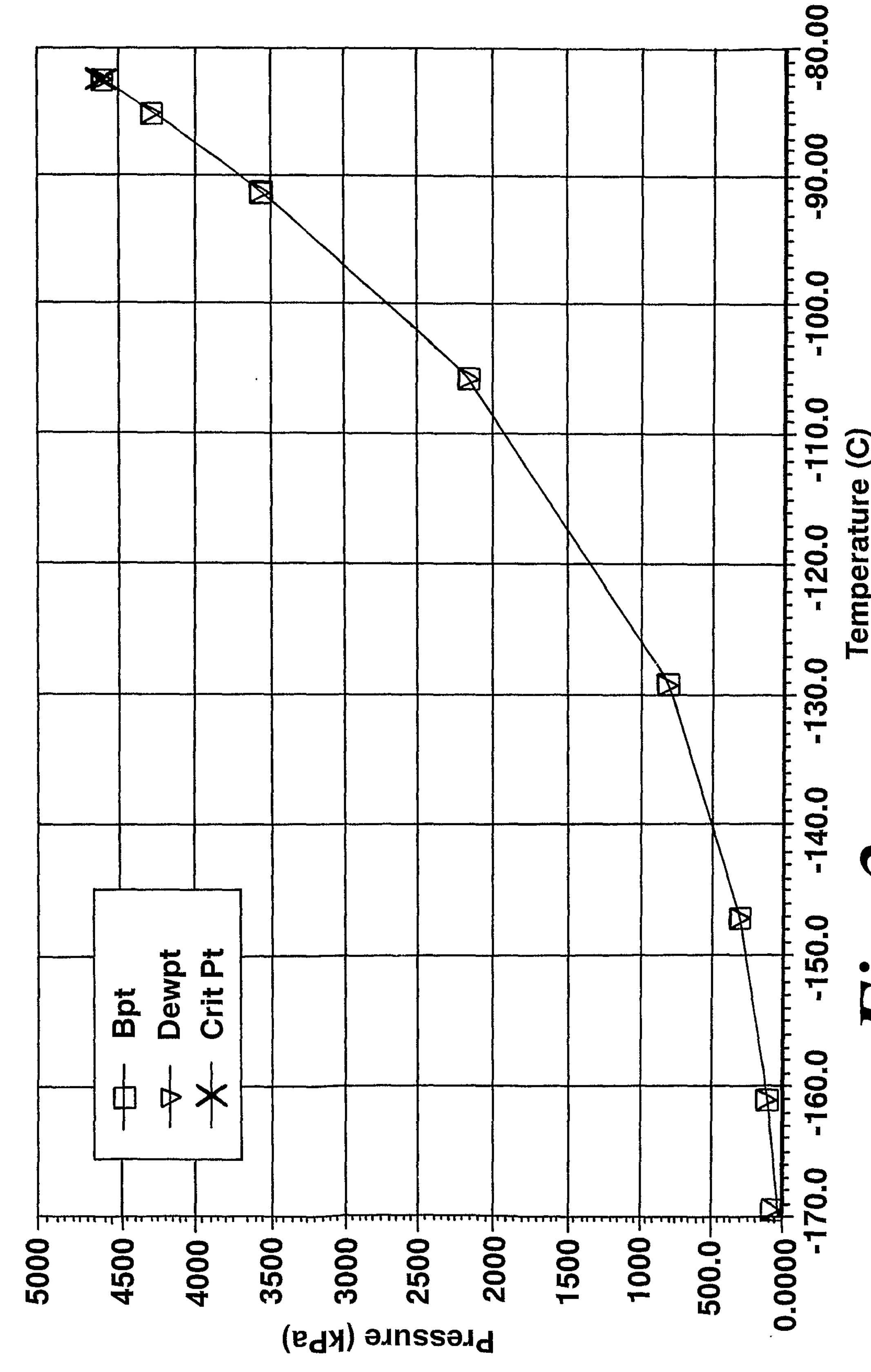
-60.00
-40.00
-20.00
Bpt
Dewpt
Crit Pt

Fig. 1



Pressure (kPa)
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4000
3500
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2000
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1000
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Temperature (C)
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-160.0
-150.0
-140.0
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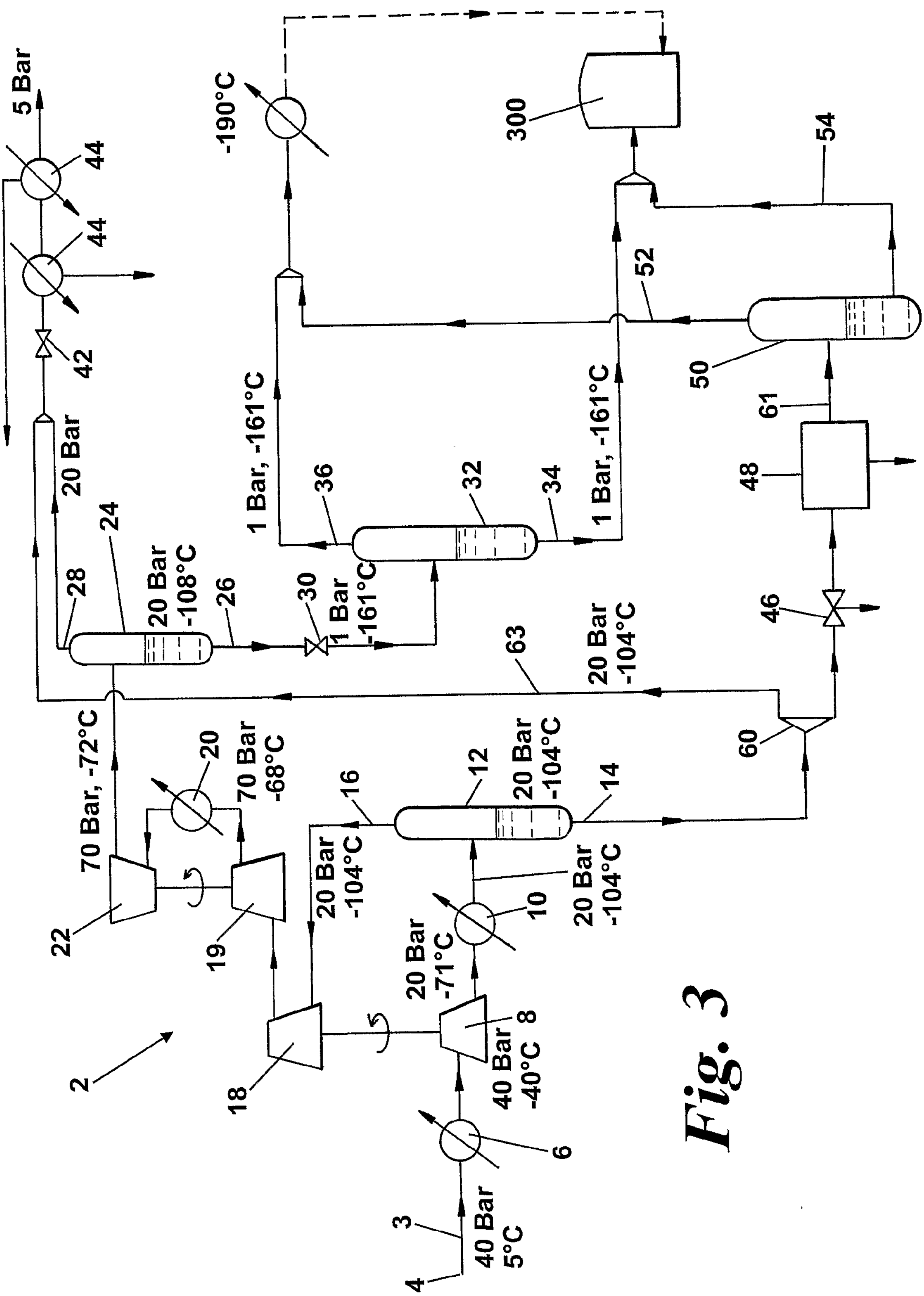
-100.0
-90.00
-80.00
Bpt
Dewpt
Crit Pt

Fig. 2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44
42
20 Bar
-190°C
300
54
52
1 Bar, -161°C
1 Bar, -161°C
36
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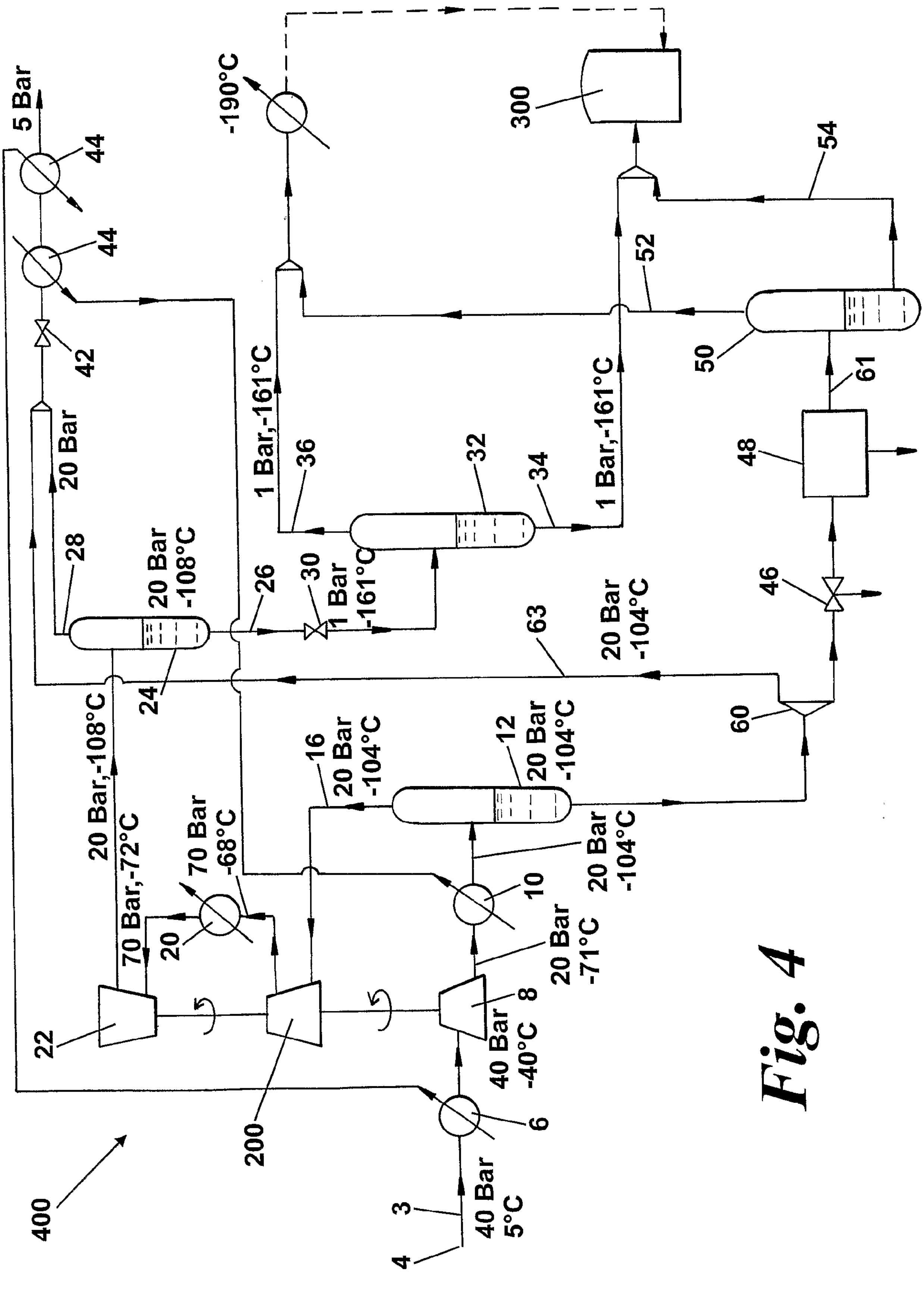
34
50
61
48
24
20 Bar
-108°C
28
26
30
1 Bar
-161°C
46
63
20 Bar
-104°C
60
70 Bar, -72°C
20
70 Bar
-68°C
16
12
20 Bar
-104°C
14
20 Bar
-104°C
20 Bar
-104°C
22
19
10
20 Bar
-71°C
8
40 Bar
-40°C
18
2
6
3
40 Bar
5°C
4

Fig. 3


5 Bar
44
44
42
20 Bar
28
20 Bar
-108°C
24
20 Bar,-108°C
20 Bar,-72°C
70 Bar,-72°C
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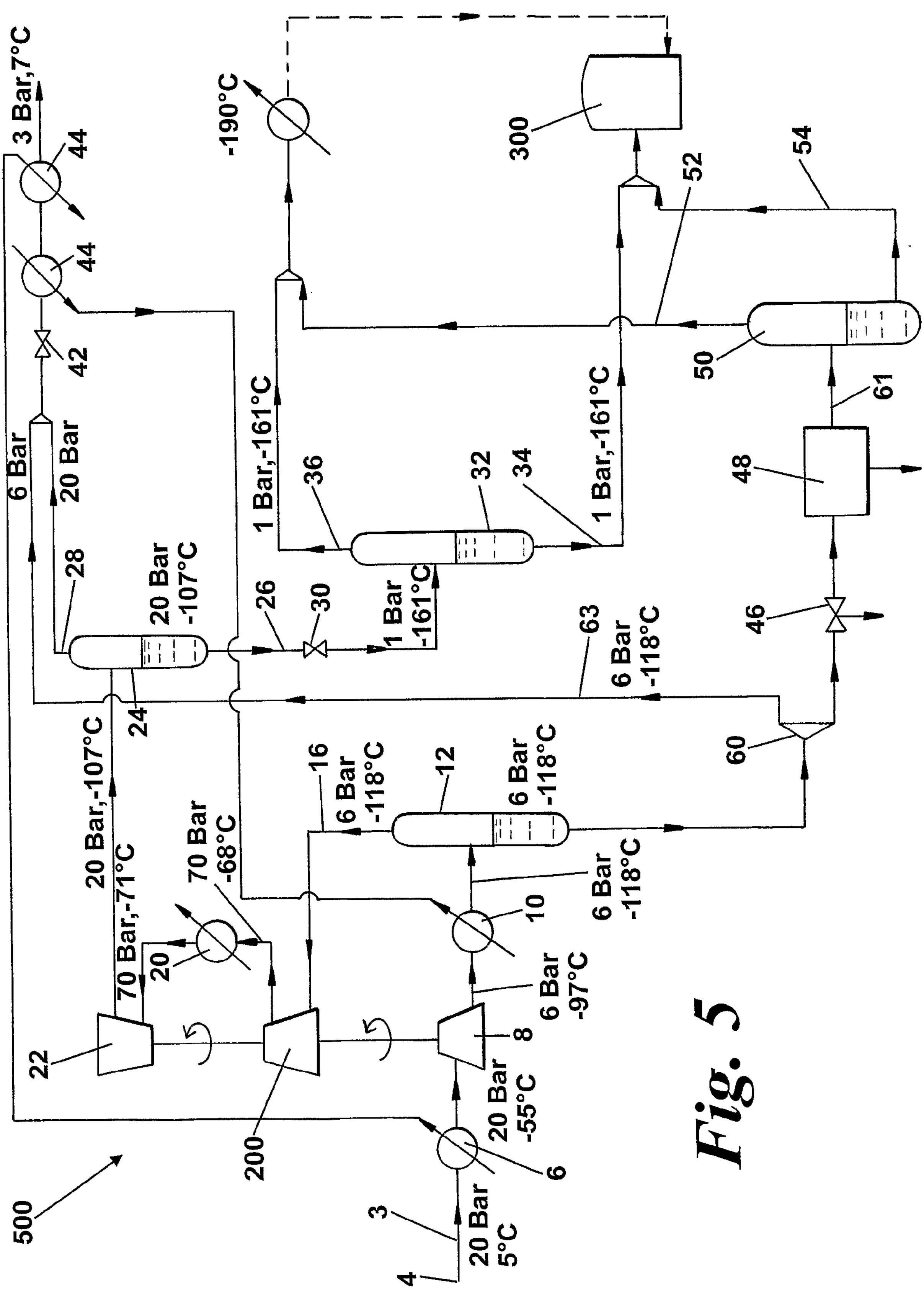
-68°C
20
22
200
400
4
3
40 Bar
5°C
6
40 Bar
-40°C
8
20 Bar
-71°C
10
20 Bar
-104°C
16
20 Bar
-104°C
12
20 Bar
-104°C
26
30
1 Bar
-161°C
32
36
1 Bar,-161°C
34
1 Bar,-161°C
-190°C
300
54
52
50
61
48
46
63
20 Bar
-104°C
60

Fig. 4

5/8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44
44
42
6 Bar
20 Bar
-190°C
300
52
54
1 Bar,-161°C
36
32
34
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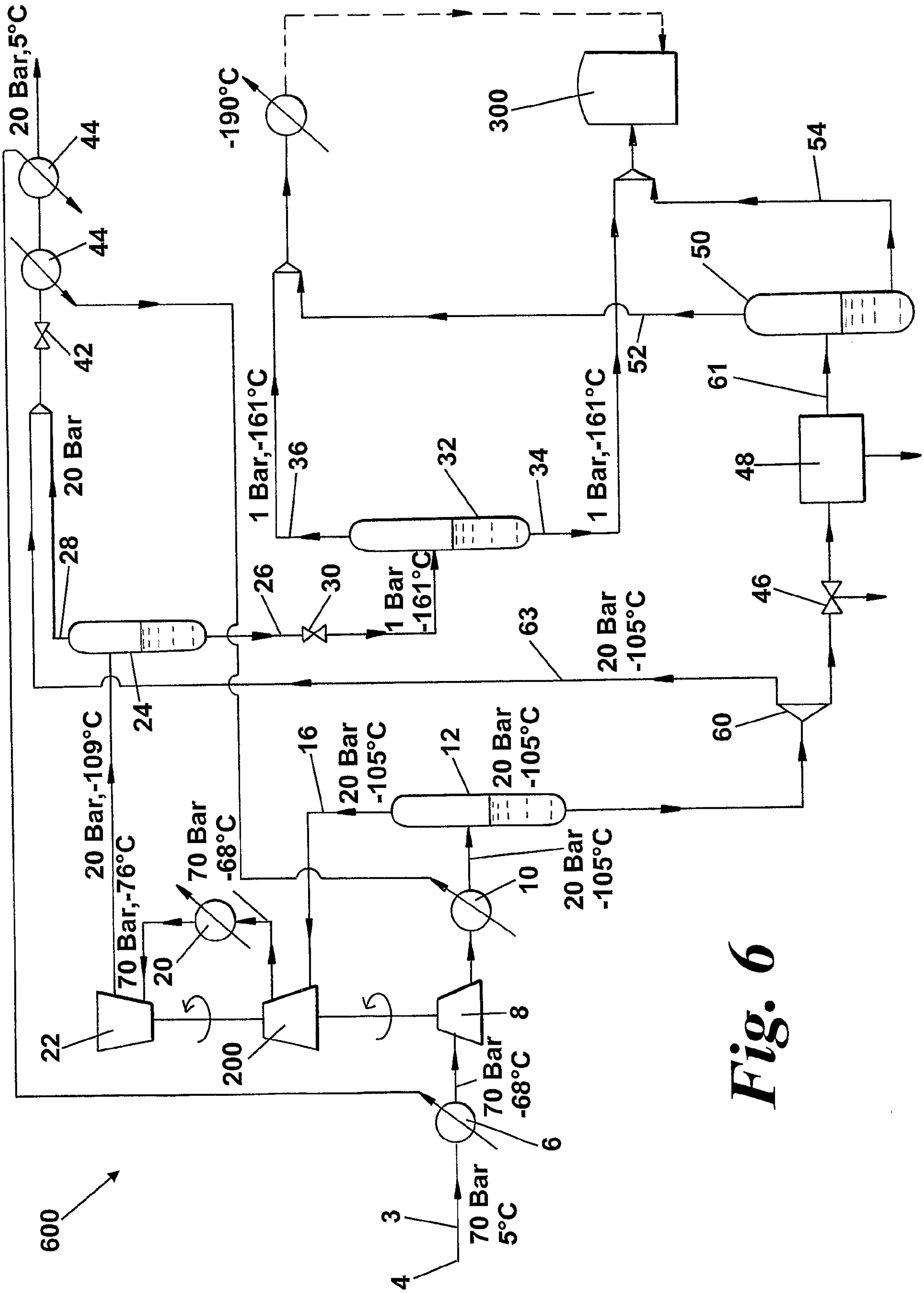
50
61
48
28
20 Bar
-107°C
26
30
1 Bar
-161°C
63
6 Bar
-118°C
46
24
20 Bar,-107°C
16
6 Bar
-118°C
12
6 Bar
-118°C
60
70 Bar
-68°C
20 Bar,-71°C
70 Bar,-71°C
20
10
6 Bar
-118°C
6 Bar
-97°C
22
200
8
20 Bar
-55°C
6
500
3
20 Bar
5°C
4

Fig. 5


600
4
3
70 Bar
5°C
6
70 Bar
-68°C
8
200
22
20
70 Bar
-68°C
70 Bar,-76°C
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24
28
20 Bar
42
44
44
20 Bar,5°C
10
20 Bar
-105°C
12
20 Bar
-105°C
16
20 Bar
-105°C
60
63
20 Bar
-105°C
46
48
61
50
52
54
300
-190°C
26
30
1 Bar
-161°C
32
34
36
1 Bar,-161°C
1 Bar,-161°C

Fig. 6

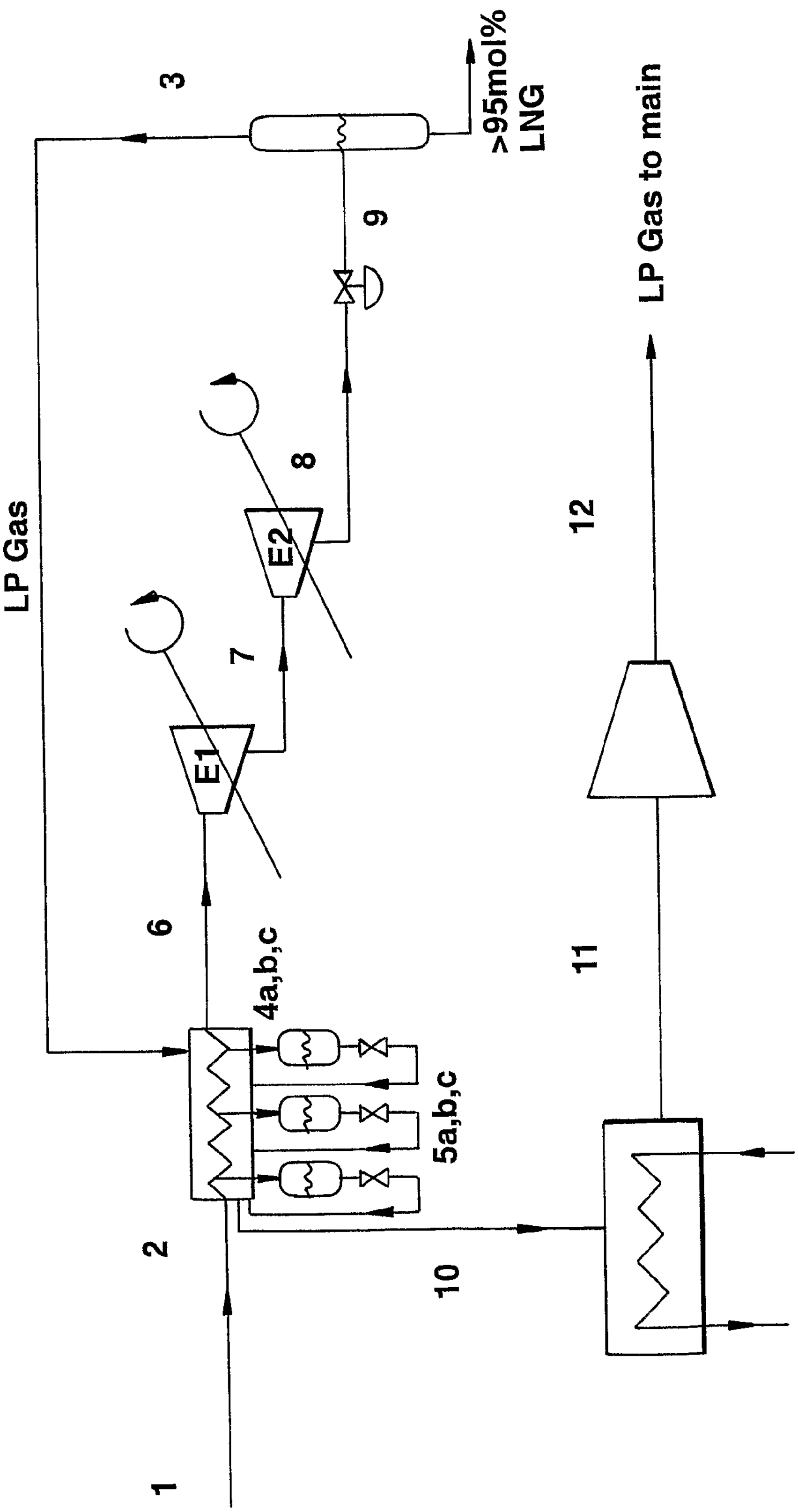

Fig. 7

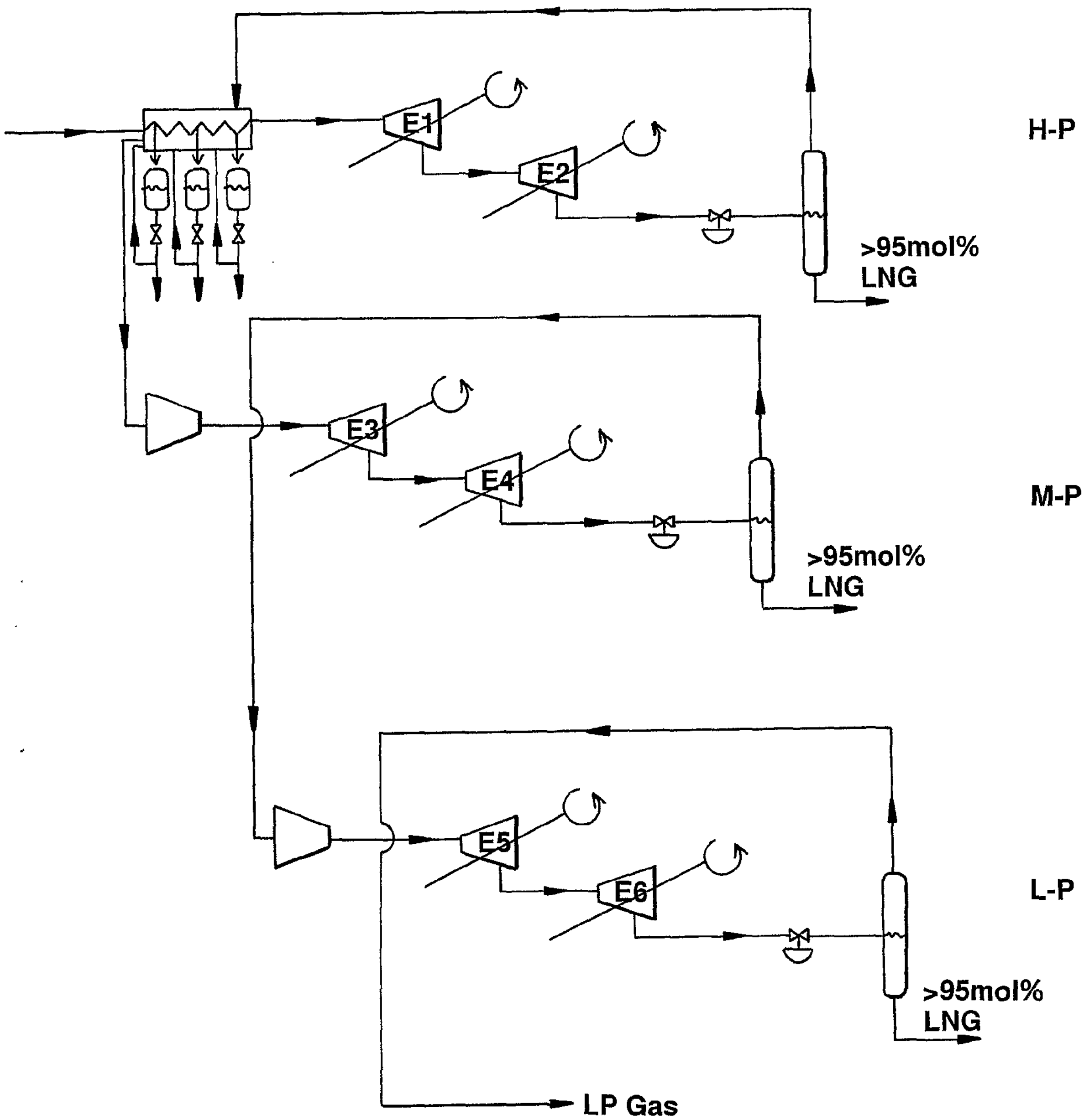

Fig. 8