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(71) Applicant *(for all designated States except US)*: **SHELL INTERNATIONALE RESEARCH MAATSCHAPPIJ B.V.** [NL/NL]; Carel van Bylandtlaan 30, 2596 HR The Hague (NL).

(71) Applicant *(for US only)*: **SHELL OIL COMPANY** [US/US]; 150 N. Dairy Ashford, Houston, Texas 77079 (US).

(72) Inventors: **LIU, Mi Chan**; Lange Kleiweg 40, 2288 GS Rijswijk (NL). **SLEESWIJK VISSER, Fenna Iona**; Lange Kleiweg 40, 2288 GK Rijswijk (NL). **GARG, Prerak**; Plot No 7, Bengaluru Hardware Park, Devanahalli Industrial Park. Mahadeva Kodigahalli, Bangalore North 562149 (IN). **BOSMA, Paul**; Lange Kleiweg 40, 2288 GS Rijswijk (NL).

(74) Agent: **SHELL LEGAL SERVICES IP**; P.O. Box 384, 2501 CJ The Hague (NL).

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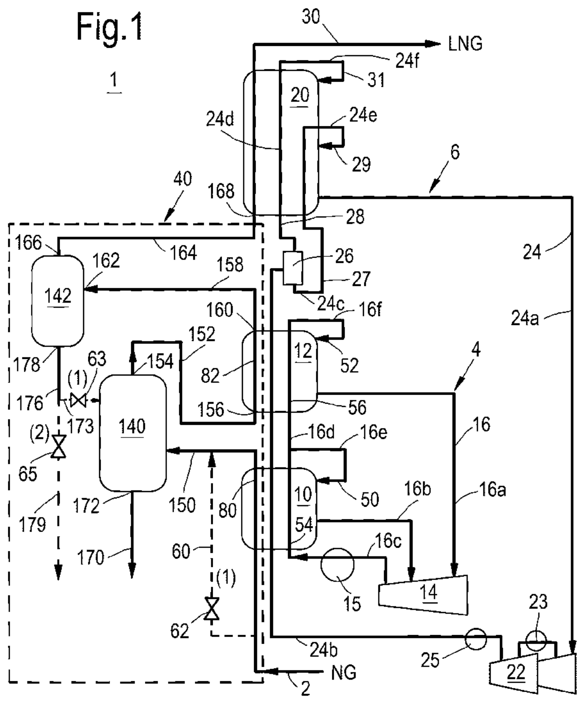
(54) Title: METHOD AND SYSTEM FOR NATURAL GAS LIQUEFACTION WITH IMPROVED REMOVAL OF HEAVY HYDROCARBONS

Fig.1



(57) Abstract: A method for production of liquefied natural gas includes the steps of: providing a natural gas stream; cooling the natural gas stream in indirect heat exchange with respect to a first refrigerant to produce a cooled gas stream having a first temperature; guiding the cooled gas stream to a first gas-liquid separator; in the first separator, separating the cooled gas stream in a first liquid stream and a first gaseous overhead stream; cooling the first gaseous overhead stream in indirect heat exchange with respect to the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature; guiding the further cooled gas stream to a second gas-liquid separator; in the second separator, separating the further cooled gas stream in a second liquid stream and a second gaseous overhead stream; and cooling the second gaseous overhead stream in indirect heat exchange with respect to a second refrigerant to produce liquefied natural gas.

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METHOD AND SYSTEM FOR NATURAL GAS LIQUEFACTION WITH IMPROVED REMOVAL OF HEAVY HYDROCARBONS

FIELD OF THE INVENTION

The present invention is directed to a method and system for the production of liquefied natural gas, wherein heavy hydrocarbons are removed from the gas.

BACKGROUND TO THE INVENTION

Natural gas is a general term that may refer to mixtures of light hydrocarbons and optionally other gases (nitrogen, carbon dioxide, helium) derived from natural gas wells. The main component of natural gas is methane. In addition to methane, natural gas may comprise higher hydrocarbons, such as ethane, propane and butane. In addition to these so-called liquid petroleum gases (LPG), (relatively small) amounts of heavier hydrocarbons may be comprised in the natural gas. Heavier herein typically relates to hydrocarbons including five of more carbon atoms, also referred to as C5+. When produced together with oil, the natural gas may be referred to as associated gas. Other compounds that may be present as contaminants in natural gas in varying amounts include carbon dioxide, hydrogen sulfide, and aromatic compounds.

Natural gas can be liquefied for purposes of storage and transportation, as the gas occupies a smaller volume as a liquid than in the gaseous state. Liquefaction takes place in an LNG (liquified natural gas) plant, in which a natural gas feed stream is typically first treated (including for instance the removal of contaminants) and subsequently liquefied. The section for liquefaction typically includes one or more heat exchangers to cool the (natural) gas by heat exchange with one or more refrigerants. Of these heat exchangers, the last heat exchanger for cooling the natural gas to the liquid state is typically referred to as the main cryogenic heat exchanger (MCHE).

The removal of heavy hydrocarbons (HHCs) including aromatic hydrocarbons, for example Benzene, Toluene, Ethylbenzene, Xylene (BTEX) components, from natural gas prior to liquefaction is required to avoid the formation of solids at cryogenic temperatures. If not removed properly, solids may form in, for instance, an upper part of the Main Cryogenic Heat Exchanger (MCHE), in an end flash system and in the storage tank for LNG. All hydrocarbons having five carbon atoms or more (also referred to as C5+), as well as water and carbon dioxide, are in their pure state solid at temperatures at or below the boiling point of methane. However, due to their low concentration these components will stay dissolved in the Liquefied Natural Gas (LNG) unless their solubility at a given temperature is exceeded. Therefore, it becomes necessary to meet a certain HHC specification in the LNG process. The HHC specification may vary depending on the temperature at a respective position or location in the process. Herein, the HHC specification typically becomes more stringent when progressing from the warm end of the process towards the cold end where the natural gas has been liquefied.

For the purpose of this disclosure, the term heavy hydrocarbons may relate to components having at least six carbon atoms (C6+). Pentane and other components having five carbon atoms (C5) may be considered a cross-over between LPG and HHC (C5 may be included in both LPG and HHC). For the avoidance of doubt, an HHC removal method should be able to remove not only C5, but respective C6+ components to below a preset threshold level. Said threshold level may differ per respective component.

US6370910 discloses a liquefaction system and process referred to as Double Mixed Refrigerant (DMR) process. Herein, both a pre-cooling cycle and a main cooling cycle use a mixed or multicomponent refrigerant. The pre-cool cycle uses a pre-cool multicomponent refrigerant and the main cooling loop uses another multicomponent refrigerant to liquefy a gaseous feed stream. The process comprises supplying a natural gas stream to a scrub column, and removing in the scrub column heavier hydrocarbons from the natural gas stream to obtain a gaseous overhead stream withdrawn from the top of the scrub column. The process comprises partly condensing the gaseous overhead stream and removing from it a condensate stream, which is returned to the upper part of the scrub column as reflux. The gaseous overhead stream withdrawn from the top of the scrub column is partly condensed to a temperature of about -50° C. Said 'condensate stream' is the liquid that is resultant from gas condensation, which typically contains a relatively high amount of LPGs. The composition of the condensate is dependent on the feed gas and phase envelope.

As exemplified by the disclosure referenced above, it is known that the heavy hydrocarbon components in natural gas feed streams can be removed using a scrub column containing a number of separation stages. The feed stream can be introduced into the scrub column as a gaseous stream or as a two phase, gas-liquid stream. In the scrub column, the feed stream is brought into countercurrent contact with a falling liquid reflux stream.

A scrub column can be effective in HHC removal, but stable operation requires sufficient liquid (i.e. reflux) to vapor flow ratio in order to avoid column dryout. The reflux for the column is typically provided by condensing a portion of the gas stream from the top of the column. If the natural gas feed is in particular too lean, containing more methane and considerably less C2 to C5s and aromatic hydrocarbons, it becomes very energy inefficient to maintain the downward liquid traffic in the scrub column as colder temperatures and therefore increasing cooling duty are required with limited natural gas liquids (NGL) components present.

WO2016150827 discloses a process, wherein a purified natural gas stream is cooled and de-pressurized. Flash gas is obtained using a single stage flash vessel, while obviating the use of a scrubber or demethaniser. The flash gas comprising less than 3 ppmv BTEX can be subjected to liquefaction to obtain LNG.

GB1572899 discloses a method of liquefying natural gas by cooling it under pressure in stages. The natural gas, which is at a relatively high pressure of, for example, 50 bar and at a temperature of, for example, 20 degrees centigrade, is passed through a coil of a heat-exchanger. The cooled natural gas leaving the coil is passed to a phase separator, wherein condensed heavier hydrocarbon components are separated from the gas. The condensed heavier hydrocarbons are removed from the phase separator through a discharge line. Together with the condensate some lighter hydrocarbons, such as methane, ethane and propane are removed from the separator through the discharge line. Cooled natural gas, which contains a small quantity of liquid, is passed through a line to a coil of a heat-exchanger. In the coil the temperature of the natural gas is reduced to a lower value, so that more liquid is formed. From the coil the natural gas is passed through a line to a phase separator. In the phase separator condensed hydrocarbons are separated from the natural gas. The condensed hydrocarbons, which contain, for example, mainly methane, ethane, propane and butane as well as some pentane, are removed from the separator. The condensed hydrocarbons are passed to a demethanizer via an expansion device, or are alternatively pumped back and recombined with the natural gas.

As exemplified by the disclosures above, for a feed gas that is lean in NGL components, it can be considered to obviate the scrub column line-up and use a simple flash or phase separator. The latter involves passing the gaseous feed stream through one or more heat exchangers to achieve a colder stream and then de-pressurizing the stream resulting in a cold two-phase stream. The two-phase stream is then separated into a vapor phase and a liquid phase in the flash separator.

In the case of natural gas processing, a flash separator is typically less effective in removing HHC and therefore must be operated at lower temperature and pressure to achieve the required level of condensation and HHC liquid drop-out. As a result of lower operating pressure, the optimum natural gas liquefaction cannot be achieved which leads to lower liquefaction process energy efficiency. Notably, there is a limiting process operating condition of the flash separator where the vapor and liquid density difference is too small and effective phase separation cannot be achieved.

An additional disadvantage of conventional flash separation is that a moderate amount of methane, which is a valuable component in LNG, is lost in the liquid product from the separator drum containing most of the NGL components. Additional processing e.g. fractionation would be required for recovery of the methane as LNG product.

The leaner the feed gas, the more unsuitable the above known heavy hydrocarbon removal methods become. Accordingly, there is a need for an improved method and apparatus, particularly when the natural gas is relatively lean while having a relatively high concentration of HHCs.

It is an aim to provide an improved method and system for the production of liquefied natural gas, wherein heavy hydrocarbons are removed.

SUMMARY OF THE INVENTION

In one aspect, the present disclosure is directed to a method for production of liquefied natural gas, the method comprising the steps of:

providing a natural gas stream;

producing a cooled gas stream having a first temperature, comprising cooling at least a part of the natural gas stream in indirect heat exchange with respect to a first refrigerant;

guiding the cooled gas stream to a first gas-liquid separator;

in the first gas-liquid separator, separating the cooled gas stream in a first liquid stream and a first gaseous overhead stream;

cooling the first gaseous overhead stream in indirect heat exchange with respect to at least the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature;

guiding the further cooled gas stream to a second gas-liquid separator;

in the second gas-liquid separator, separating the further cooled gas stream in a second liquid stream and a second gaseous overhead stream; and

cooling the second gaseous overhead stream in indirect heat exchange with respect to a second refrigerant to produce liquefied natural gas.

According to another aspect, the disclosure provides a system for production of liquefied natural gas, the system comprising:

a natural gas conduit to provide a natural gas stream;

a first heat exchanger for cooling at least a part of the natural gas stream in indirect heat exchange with respect to a first refrigerant to produce a cooled gas stream;

a first conduit for guiding the cooled gas stream from the first heat exchanger to a first gas-liquid separator and feeding the cooled gas stream into the first gas-liquid separtor at a first temperature, the first gas-liquid separator being adapted for separating the cooled gas stream in a first liquid stream and a first gaseous overhead stream;

a second conduit for guiding the first gaseous overhead stream to a second heat exchanger or second heat exchange section adapted to cool the gaseous overhead stream in indirect heat exchange with respect to the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature;

a third conduit for guiding the further cooled gas stream to a second gas-liquid separator, the second gas-liquid separator being adapted to separate the further cooled gas stream in a second liquid stream and a second gaseous overhead stream; and

another heat exchanger for cooling the second gaseous overhead stream in indirect heat exchange with respect to a second refrigerant to produce liquefied natural gas.

The system as described above can be adapted to implement the method as outlined above.

BRIEF DESCRIPTION OF THE DRAWINGS

The drawing figures depict one or more implementations in accord with the present teachings, by way of example only, not by way of limitation. In the figures, like reference numerals refer to the same or similar elements.

Figure 1 schematically shows an embodiment of a system according to the present disclosure;

Figure 2 schematically shows another embodiment of a system according to the present disclosure;

Figure 3 schematically shows yet another embodiment of a system according to the present disclosure;

Figure 4 schematically shows an embodiment of a system according to the present disclosure;

Figure 5 schematically shows another embodiment of a system according to the present disclosure;

Figure 6 schematically shows an embodiment of a system according to the present disclosure comprising a wash section; and

Figure 7 schematically shows another embodiment of a system according to the present disclosure comprising a liquid spray section.

DETAILED DESCRIPTION OF THE INVENTION

It is presently proposed to integrate a two-stage heavy hydrocarbon (HHC) removal in a natural gas liquefaction system and/or process. First, a cooled gas stream is produced from a natural gas stream, by cooling at least a part of the natural gas stream in indirect heat exchange with respect to a first refrigerant. The cooled natural gas stream is fed into a first gas-liquid separator at a first temperature, and separated in the first gas-liquid separator. A first gaseous overhead stream from the first gas-liquid separator is subsequently cooled in indirect heat exchange with respect to at least the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature. The further cooled gas stream is then separated in a second gas-liquid separator. A second gaseous overhead stream expelled from the second gas-liquid separator is subsequently cooled even further, in indirect heat exchange with respect to a second refrigerant, to produce liquefied natural gas.

Heavy hydrocarbons can be extracted from the process and/or system as a first liquid stream from the first gas-liquid separator. Other heavy hydrocarbons (on average lighter than those in the first liquid stream) can be extracted the second gas-liquid separator as a second liquid stream. Heavy hydrocarbon streams may optionally be passed to a fractionation unit for further separating in HHC fractions.

Any part (up to all) of the second liquid stream may be fed back, optionally by means of a pump, into the first separator, whereby contacting the cooled gas stream with this part of the second liquid stream in the first separator. The first gas-liquid separator may comprise a spray header, which can be used to spray the fed-back part of the second liquid stream into the first separator using the spray header. Alternatively, the first gas-liquid separator may comprise a liquid distributor, which can be used to distribute the fed-back part of the second liquid stream over the internal cross-sectional area of the first separator.

In an embodiment, the step of producing the cooled gs stream comprises an additional step of controlling a flow of a part of the natural gas stream through a bypass line to bypass and feeding the part of the natural gas stream from the bypass line into the at least part of the natural gas stream after the indirect heat exchange with respect to the first refrigerant, to bypass the step of cooling the at least part of the natural gas stream, and thereby to control the first temperature. Herein, the first temperature may be controlled within a margin of 2°C, prefereably within a margin of ±1°C.

Certain terms used herein are defined as follows:

"NG" refers to natural gas. Natural gas is a naturally occurring hydrocarbon gas mixture primarily comprising methane, but commonly including varying amounts of other higher alkanes, and sometimes a small percentage of carbon dioxide, nitrogen, hydrogen sulfide, or helium;

"LNG" refers to liquefied natural gas, which is typically cooled to at least a temperature whereat the gas can be in the liquid phase at about 1 bar pressure; for liquefied methane this temperature may be about -162 ℃;

"Mixed refrigerant" or "MR" refers to a refrigerant comprised of two or more components. Depending on the stage of the heat exchanger (pre-cooler or main cryogenic heat exchanger), the refrigerant may include components such as methane, ethane, propane, and nitrogen.

"HMR" and "LMR" refer to "heavy mixed refrigerant" and "light mixed refrigerant" respectively, indicating mixed refrigerant separated into light and heavy mixed refrigerant streams, wherein the terms “light” and “heavy” indicate average component weight of each stream relative to each other;

"PMR" may refer to a pre-cool mixed refrigerant, i.e. a mixed refrigerant used in a pre-cool circuit of a liquefaction system;

"Bar" is a metric unit of pressure, defined as equal to 100 kPa. "Bar(a)" and "bara" are sometimes used to indicate absolute pressures and "bar(g)" and "barg" for gauge pressures. Herein, "2 barg" is similar to fuller descriptions such as "gauge pressure of 2 bar" or "2-bar gauge".

"Natural Gas Liquids" or "NGLs" are hydrocarbon components of natural gas that are separated from the gas state in the form of liquids. NGLs may also be referred to as condensate.

"LPG" relates to Liquid Petroleum Gases. LPG is a subset of NGL. LPG are C2 to C4s with high vapor pressure. LPG typically includes ethane, propane, and butane. Trace amounts of C5 can be found in LPG due to the fractionation process.

Another subset of NGL is condensate which has lower vapor pressure than LPG, and is mainly composed of C4+.

"Heavy Hydrocarbons" or "HHC" are hydrocarbon components comprising five carbon atoms or more (C5+ components), including aromatics.

The phrase 'lean gas' typically refers to natural gas including a relatively low amount of NGL components per unit of volume of the natural gas. Low herein may imply 10% or less, or even 5% or less NGL. Methane is the main component of natural gas, usually accounting for 70%–90% of the total volume produced. If gas contains, for instance, more than 90% methane, it may be termed lean gas. Natural gas, including lean natural gas, may be defined as having a high concentration of unwanted heavy hydrocarbons, such as C5+ and aromatics, when comprising - for instance - about 1300 ppmv or more unwanted HHCs per unit of volume.

When the description below mentions respective process streams, typically the respective system comprises corresponding conduits. In use, the process streams are routed through the system via said corresponding flow conduits.

Different liquefaction schemes are known, such as C3MR, SMR (single mixed refrigerant), DMR (double mixed refrigerant), or cascade-based liquefaction processes. Many of these schemes comprise a coil wound heat exchanger, typically the main cryogenic heat exchanger, in which a substantial part of the cooling of the natural gas takes place. Suitable coil wound heat exchangers are commercially available from a variety of vendors, including Air Products and Chemicals Inc. (APCI), Pennsylvania (USA), and Linde AG (Germany).

A mixed refrigerant process may include a precooling circuit as well as a main cooling circuit, both cooling circuits comprising one or more heat exchangers. A similar approach as described in the present disclosure with respect to the main cooling cycle can be applied to the pre-cooling cycle. Examples of mixed refrigerant liquefaction processes and systems include, for instance, a single mixed refrigerant process (see for instance US6658891), a parallel mixed refrigerant process (see for instance US20080156037), or a C3MR process (see for instance US20090301131). In a DMR process, typically, the refrigerant used for the pre-cooling heat exchanger is a first mixed refrigerant and the refrigerant used for the main cryogenic heat exchanger is a second mixed refrigerant. For a detailed description of the DMR system and process, reference is made to, for instance, patent publications US6370910 or US6658891.

Fig. 1 schematically depicts a system 1 and associated method for liquefying a gas 2, such as natural gas. The system of Fig. 1 may be referred to as a version of a DMR liquefaction process. Please note that various changes to the scheme of Figure 1 are conceivable. For instance, the gas 2 in its associated conduit may have been pre-treated. Several types of gas treatment can precede the scheme of Fig. 1. Various other types of equipment can be added downstream of the system to further process liquefied gas provided by the scheme of Fig. 1. Such equipment may include a nitrogen removal section or addition of an end-flash section. Also, there may be differences between respective liquefaction systems, for instance with respect to thermodynamics, pressure ranges, temperature ranges, and flow rates at various locations in the system during operation, equipment sizes, maximum capacity, etc. Despite these potential differences, for simplicity, equipment parts with substantially the same function have been annotated with the same reference number throughout the Figures.

At a generic level, the liquefaction system 1 comprises two consecutive cooling cycles 4 and 6. The pre-cool cycle 4 may comprise at least one, for instance two heat exchangers 10 and 12 and at least one pre-cool refrigerant compressor 14. The cycle 4 may also include at least one cooler 15. Pre-cool refrigerant conduits 16 guide pre-cool refrigerant from the compressor 14 through the two heat exchangers 10 and 12 and back. Refrigerant conduits 16 may comprise conduits 16a to 16f. Together, conduits 16 constitute a loop to circulate a refrigerant. The refrigerant may be referred to as precool refrigerant or first refrigerant.

The first pre-cool heat exchanger 10 may operate at a first pressure and the second pre-cool heat exchanger 12 may operate at a second pressure. Herein, the first pressure exceeds the second pressure. As a result, the first heat exchanger 10 may be referred to as high-pressure (HP) pre-cool heat exchanger. The second heat exchanger 12 may be referred to as low-pressure (LP) pre-cool heat exchanger.

The main cooling cycle 6 comprises at least one main heat exchanger 20 and at least one main refrigerant compressor 22. The cycle 6 may include one of more coolers 23, 25. Main refrigerant conduits 24 extend from the compressor 22 through the two heat exchangers 10 and 12, subsequently through the at least one main heat exchanger 20, and back to the compressor 22. The cycle 6 may comprise a separator 26 to split the mixed refrigerant of the main cycle 6 in a heavy mixed refrigerant 27 and a light mixed refrigerant 28. Refrigerant conduits 24 may comprise conduits 24a to 24f. Together, conduits 24 constitute a loop to circulate a refrigerant. The refrigerant circulated in loop 6 via conduits 24 may be referred to, for instance, as main refrigerant, mixed refrigerant or second refrigerant.

The gaseous feed stream 2 is routed via conduits extending through the two heat pre-cool exchangers 10 and 12, and subsequently through the at least one main heat exchanger 20, to provide at least partially condensed or liquefied gas 30.

Both the pre-cooling cycle 4 and the main cooling cycle 6 may use a mixed or multi-component refrigerant to pre-cool and subsequently condense or liquefy the gaseous feed stream 2. Expanded pre-cool mixed refrigerant 50, 52 provides cooling duty to the pre-cool heat exchangers 10, 12 respectively and to all streams routed through the inside of said heat exchangers. Expanded heavy mixed refrigerant 29 and expanded light mixed refrigerant 31 provide cooling duty to the main cryogenic heat exchanger 20 and to all streams routed through the inside of the MCHE 20.

For detail of the cooling cycles and the working thereof, reference is made to, for instance, US6370910.

The system 1 also includes a HHC removal system 40. In a conventional system, a HHC removal system may comprise a scrub column. For a person skilled in the art of gas liquefaction, a vertical scrub column for removing hydrocarbon liquids from natural gas refers to a relatively large and expensive piece of equipment, having multiple (at least five, but typically about ten) layers of stages with packing or trays as internals. There is an upper limit on the column diameter size due to fabrication feasibility. Because the gas velocity is typically higher in a vertical scrub column than in a simpler separator or flash vessel (for the same diameter of the vessel, and with all other parameters the same), a vertical scrub column tends to be taller than a flash vessel as the tray spacing is increased to allow for sufficient tray vapor capacity. A typical scrub column provided with ten trays typically has a height exceeding 10 m. Height of a typical scrub column may be at least 1.5 to 2 times the height of a flash vessel or relatively simple separator with other parameters the same. See for instance Oil and Gas Separation Design Manual by C. Richard Sivalls, of 10 February 2009. In addition, a scrub column introduces a pressure drop per stage, typically about 0.01 bar per tray in a scrub column. Said pressure drop may be in the order of 0.1 bar.

The embodiment of system 1 shown in Figure 1 is provided with a liquids removal section 40 comprising, for instance, a first separator 140, and a second separator 142. First and second separator 140, 142 may be flash vessels. A conduit 150 connects an outlet of the first pre-cool heat exchanger 10 with an inlet of the first separator 140. A second conduit 152 connects an upper outlet 154 of the first separator 140 with an inlet 156 of the second pre-cool heat exchanger 12. A third conduit 158 connects an outlet 160 of the second pre-cool heat exchanger 12 with an inlet 162 of the second separator 142. A fourth conduit 164 connects an upper outlet 166 of the second separator 142 with an inlet 168 of the main heat exchanger 20.

An optional bypass conduit 60 may be provided, connecting the inlet of the bundle 80 of the first heat exchanger 10 to an outlet of said bundle 80. The conduit 60 may be provided with a valve 62. The valve 62 may be controllable within a range between a closed position and an open position, allowing to adjust a flow of gas through the bypass conduit 60. The latter allows at least a section or part of the gas 2 to bypass the bundle 80 in the heat exchanger 10.

A fifth conduit 170 is connected to a lower outlet 172 of the first separator. Liquids can be removed from the first separator 140 via the conduit 170. A sixth conduit 176 is connected to a lower outlet 178 of the second separator 142. Liquids can be removed from the second separator 142 via the conduit 176. Optionally, the liquids from the second separator 142 can be recycled back to the first separator via line 173, either by gravity or optionally aided with a pump 175 (Fig. 2). Liquids may be removed, bypassing the first separator 140 via line 179. The conduits 173 and/or 179 may be provided with valves 63 and 65 respectively, allowing to control the flow in each conduit.

Figure 2 shows an embodiment, wherein conduit line 173 is provided with a pump 175. Optionally, an expansion device may be provided in conduit 150 and/or in conduit 158. The expansion device may be an expander, or a JT valve 64, 66.

In the embodiment shown in Figure 3, a turbo expander 70, 72 may be arranged in conduit 158 and/or in conduit 164.

In an embodiment (Figure 4), the system 1 may comprise two stages of a cold flash separator 140, 142 with extended cooling. Herein, vapor conduit 152 connects to a bundle or conduit 82 of the precool loop 4. Conduit 88 connects an outlet of bundle 82 to a conduit 104 in the heat exchanger 20. Said conduit 104 may be referred to as warm bundle. An outlet of the warm bundle 104 may be connected to conduit 158 guiding the process stream to the second separator 142. This setup allows deeper cooling, i.e. cooling to lower temperatures, of the natural gas in addition to cooling in the pre-cool cycle 4.

Figure 5 shows system 101, wherein the pre-cool cycle 4 includes a single heat exchanger 10 provided with two bundles 80, 82 for guiding and cooling the natural gas feed stream 2. The first bundle 80 is connected to a second outlet 84 of the heat exchanger 10 which in turn is connected to conduit 150. The conduit 152 is connected to a second inlet 86 of the heat exchanger 10 connected to the second bundle 82.

In use, the single heat exchanger 10 is set to operate at a predetermined internal pressure. Precool refrigerant is passed through conduit 16d. Main refrigerant is passed through conduit 24b. The natural gas 2 is passed through the bundles 80 and

82. Cooled precool refrigerant in conduit 16f is expanded. The expanded precool refrigerant is diffused into the heat exchanger 10 near the top of the heat exchanger 10. The expanded refrigerant cools the conduits 16d, 24b and 80, 82 (as shown in Fig. 5).

Figures 6 and 7 respectively show system 201 and 301 according to the present disclosure, wherein the first separator 140 comprises an integrated wash section 144.

As shown in Figure 6, the wash section may comprise a header 146 for distributing liquid into the separator vessel. The separator 140 may be provided with one contactor 148. The contactor section 148 may comprise, for instance, packing or similar device. The contactor 148 of separator 140 of the present disclosure can comprise for instance a single layer of packing. Thus, the separator 140 is relatively simple and inexpensive. The contactor is arranged to improve contact between vapor and liquid, in this case natural gas and wash liquid. The conduit 176 may be connected to an inlet 147 of the header 146 for the supply of wash liquid. The separator 140 in Fig. 6 may be referred to as a wash vessel.

As shown in Figure 7, the wash section 144 may comprise a spray header 180 for spraying dispersed liquid into the vessel of the first separator 140. The conduit 176 may be connected to an inlet of the spray header 180 for even distribution and desired droplet size for optimum contact of wash liquid.

The separator 140 may be provided with a distributor 182, connected to the gas conduit 150. The distributor 182 may be arranged for receiving pre-cooled natural gas and distributing the gas evenly over the internal area of the separator 140. The separator 140 in Fig. 4 may be referred to as a spray vessel. Liquid distributors are available on the market, and are for instance marketed by HMDS PROCESS in France and many other companies.

In embodiments wherein liquid from the second separator 142 is fed back to the first separator 140 (via conduits 176 and 173), the operating temperature of the first separator will change with varying quantity of liquids routed via conduit 173. This will cause a wider variation in temperatures of the process stream in line 152. Said temperature variation translates to a temperature difference (also referred to as a delta temperature) at the warm end approach to, i.e. the inlet of, the second bundle 82 or the LP Precooler 12 with respect to the temperature at the other tube-side inlet streams at the warm end, for example the PMR (at the inlet of tube bundle 56) and

MR (at the inlet of the respective tube bundle in heat exchanger 12). Said inlet for the natural gas may be inlet 86 of the cold bundle 82 in heat exchanger 10 (single stage DMR, see Fig. 5) or inlet 156 of LP Precooler 12 (two stage DMR, see Fig. 1). Such delta temperature is preferably limited as the precooling step may become less efficient with increasing temperature difference between respective process streams .

Therefore, a temperature bypass line 60, including a controllable valve 62, may be included, allowing for a portion of the natural gas 2 to bypass the first bundle 80 or the first heat exchanger 10 and feed this portion of the natural gas stream into cooled part of the natural gas stream which is discarged from bundle 80. Hereby a direct heat exchange is accomplished between the bypass portion of the natural gas and the cooled part of the natural gas stream as discharged from bundle 80. Controlling the amount of said bypass portion of the gas 2 can control the temperature of the feed to the first separation system 140, whilst the mid-bundle temperature is allowed to match the temperature of the natural gas at the inlet of the second bundle 82, therefore reducing or eliminating the said delta temperature.

The feedback of liquid from the second separator may cause a variation in temperature in the first vessel. The temperature variation or delta may exceed for instance ±5 to ±10°C. Controlling the bypass conduit 60 allows mixing of cooled gas from bundle 80 with uncooled gas from conduit 60. In a practical embodiment, the bypass via conduit 60 enables to control the feed temperature to the first separator. Controlling the temperature in the first separator allows to control the mid-bundle temperature (the temperature at the inlet of bundle 82). The latter allows to limit the temperature difference between the tube-side inlet streams (i.e. difference between the temperature of streams in conduits 152, 56 and 24b respectively at the level of inlet 156) at the warm end of the second bundle 82 or LP Precooler 12. The warm end of the bundle 82 herein relates to the end at inlet 156. The temperature difference may be controlled within a margin of, for instance, ±2°C, or even ±1°C. As a result, the process can operate at or near an optimum setting, with optimal yield or efficiency. The temperature will vary depend on amount of liquids which also depends on the feed gas composition and PMR cut point. For the cases considered for two-stage HHC, this was around 5 to 10 °C. However, it is noted for the scrub column example, the temperature difference between inlet streams to the heat exchanger 12 (or alternatively to the second bundle in heat exchanger 10) was higher, in the order of 20°C.

The high-pressure (HP) precooler 10 (Fig. 1) may operate approximately between 5 and 11 bar. The low-pressure (LP) precooler 12 (Fig. 1) may operate approximately between 1.5 to 5 bar. The pressures mentioned herein relate to the pressure in the shell side of the heat exchanger.

A pressure drop across first bundle 80 may be about 3 bar. A pressure drop across second bundle 82 may also be about 3 bar. Therefore, the second separator 142 may operate at an internal pressure in the order of 3 bar below an internal pressure in the first separator 140. A pressure drop in the vessel 140, 142 may be about 10 – 15 kPa (0.1 to 0.15 bar). The pressure drop in vessels 140, 142 may be due to entrance, equipment internals, and exit losses only. A flash may be created in separator 140 when conduit 150 provides a two-phase stream. A two-phase stream in conduit 150 may result from reducing the temperature and/or pressure in the heat exchanger 10. The temperature in heat exchanger 10 can be reduced, for instance, by increasing the flow of expanded refrigerant 50, 98. In addition, the temperature in heat exchanger 10 can be reduced by reducing the pressure inside the heat exchanger vessel.

Conduit 150 may optionally be provided with a Joule-Thompson valve or JT valve 64 to introduce a pressure reduction of the process stream passing through said conduit 150. Said pressure reduction lowers the temperature thereby increases the flash condensation of the stream entering the first separator 140. If using pressure reduction to create a two-phase stream in conduit 150, then a pressure drop across the optional JT valve 64 is preferably minimized as it may negatively impact liquefaction efficiency. The pressure drop across the JT valve 64 can be around 3 to 5 bar. The pressure drop may depend on the feed gas composition and extent of HHC removal required.

Conduit 158 may optionally be provided with a Joule-Thompson valve or JT valve 66 to introduce a pressure reduction of the process stream passing through said conduit 158. Said pressure reduction lowers the temperature thereby increases the flash condensation of the stream entering the second separator 142. If using pressure reduction to create a two-phase stream in conduit 158, then a pressure drop across the optional JT valve 66 is preferably minimized to limit negative impact on liquefaction efficiency. The pressure drop across the JT valve 66 can be around 3 to 5 bar. The pressure drop may depend on the feed gas composition and extent of HHC removal required.

The heavy hydrocarbon removal method according to the present disclosure comprises the steps of cooling the natural gas feed stream 2 by indirect heat exchange with a first refrigerant. The first refrigerant may include the Precool Mixed Refrigerant (PMR) of the PMR loop 4. The gas in conduit 150 may have been cooled to a temperature in the range of -10 to -40°C. In a practical embodiment, the temperature of the process stream in line 150 is about -35°C. This first cooling step can be done in the bundle 80 of the heat exchanger 10 (Figures 1 to 7). Subsequently, the cooled natural gas (in conduit 150) is introduced into the first gas-liquid separation system 140.

Optionally, the first gas-liquid separator 140 has the option to comprise of a wash section 144 where a countercurrent flow of hydrocarbon liquid (originating from conduit 176) absorbs, at least, gaseous heavy hydrocarbons (HHC).

A next step comprises further cooling of the natural gas vapor stream (in conduit 152) resulting in a gas-liquid stream (in conduit 158). Further cooling may be by cooling in the cold bundle 82. The bundle 82 may be included in the second heat exchanger 12 (see Figures 1-4 and 6-7) or as a second bundle following a bundle break in a single heat exchanger 10 (Fig. 5). Cooling medium may be the first refrigerant (e.g. the PMR). The gas is further cooled in bundle 82 to partially condense the gas. The gas in conduit 158 may be cooled to a temperature in the range of -40 to -60°C.

Further cooling may optionally comprise additional cooling in the first bundle 104 of the main heat exchanger 20 using the second refrigerant (e.g. the MR) as the cooling medium (see. Fig. 4). Additional cooling allows to cool the gas in the line 158 to a lower temperature, for instance in the range of -60 to -100°C.

The gas-liquid stream in conduit 158 is subsequently separated in the second separator 142 thereby producing a lean natural gas vapor stream (in conduit 164) and a heavy hydrocarbon enriched liquid stream (in conduit 176). Lean natural gas vapor herein means predominantly methane depleted in hydrocarbons heavier than butane. Herein, depleted means depleted to a predetermined maximum threshold or below said threshold for each respective component of pentane and heavier. Said threshold may be about 0.1 ppmv or less, or even less than 0.01 ppmv, for individual HHC components. A maximum threshold for all heavy hydrocarbon components combined may be in the order of 500 ppmv.

In an embodiment, the liquid stream in conduit 176 can be returned to the first separator 140. The liquid returned to the first separator 140 can be used as wash liquid. The liquid can also recycle to the first separator for enrichment of the natural gas only (for altering the feed gas composition and phase envelope).

The natural gas stream lean in heavy hydrocarbons (in conduit 164) is supplied to the main heat exchanger 20 to undergo liquefaction by indirect heat exchange with a refrigerant. The refrigerant can be referred to as Mixed Refrigerant or multicomponent refrigerant (MR). The multicomponent refrigerant is cycled in the main liquefaction loop 6.

The advantage of incorporating a hydrocarbon wash section 144 in the first separator 140 thereby increasing the functionality to include gas and liquid mass transfer provides more effective HHC removal. The second separator 142 enables production of a hydrocarbon liquid in conduit 176 suitable for gas absorption. This enhanced capability for HHC removal comes at a relatively low cost and has the added benefit of increased flexibility and robustness to handling variations in feed gas composition. A relatively high operating pressure can be sustained as there is reduced or no requirement for pressure let-down.

In a practical embodiment, relatively high pressure herein is a typical liquefaction feed pressure (of feed gas 2) in the range of, for instance, 50 to 75 bar, and higher. As mentioned above, pressure drop across heat exchanger bundles 80 and 82 may be in the order of 3 bar respectively. Operating pressures in the first separator 140 and the second separator 142 can therefore be substantially about 3 bar below the feed pressure and 3 bar below the pressure in the first separator respectively. Optional JT valves 64, 66 may introduce an additional limited pressure reduction.

Conventionally, pressure let-down or pressure reduction to a lower pressure (than the feed pressure according to the method of the present disclosure) is used to enable partial condensation for flash separation. Pressure let-down may typically involve expansion such as passing the feed through a Joule-Thompson valve or other expansion device. The higher feed pressure according to the method of the present disclosure allows for optimum liquefaction feed conditions and process energy efficiency due to the lower latent heat of condensation of the natural gas.

PMR is used to precool the natural gas and the second refrigerant loop, however there is a limit to how low the temperature can be reached with the PMR. The lowest achievable temperature depends on the pre-cool refrigerant composition, but is typically in the order of -45 to -60°C. To establish lower temperatures, in an alternative embodiment of the above, the gaseous overhead in conduit 152 from the first separator 140 can undergo cooling using PMR and subsequent extended cooling by heat exchange with MR in the Main Cryogenic Heat Exchanger (MCHE), see Figure 4. When cooling of the PMR is limited, this variation allows lower temperatures to be established thereby increasing the gas condensation and therefore HHC liquid drop-out for removal.

The second separator 142 will be operating at a pressure at least 3 bar less than the operating pressure in the first separator 140. Light ends that can be recovered from the NGL stream 170 from the HHC separation with an optional downstream fractionation unit may be compressed, cooled (to a similar temperature to the PMR cutpoint) and combined with the main NG line prior to liquefaction in the MCHE. The PMR cutpoint herein is the temperature at the outlet 160 of the precool loop. In some conventional system, instead the lighter components of the NGLs are fed as a separate tube in the MCHE 20. The system and method of the present disclosure enable a simpler design of the MCHE 20, allow to reduce the size and thus capital costs of the heat exchanger 20, or for the same size allow for increased throughput of the process stream for higher LNG capacity.

The PMR loop 4 and tubes in the precool loop 4 are also different. Conventionally, the PMR can be compressed and partially condensed, vapor and liquid is separated and fed into the precooler in separate tubes. The present invention allows to compress and completely condense the refrigerant in loop 4, which may be a precool multicomponent refrigerant (PMR), and to feed the precool refrigerant into the precool heat exchanger 10, 12 thereby reducing the number of tubes. This simplifies the design of the heat exchangers 10, 12, reduces heat exchanger size or allows for increased throughput for higher LNG production.

Another embodiment of the invention allows deeper cooling by pressure letdown of the precooled high-pressure feed gas upstream of the first and/or second separators by flashing over a Joule-Thomson (JT) valve 64, 66 or expansion in an expander. The latter results in a stronger temperature reduction as enthalpy is removed from the system.

The expander 70 can be coupled directly to a product gas (re)compressor 72, sometimes referred to as a turbo-expander or compander, or can also be used to drive a power generator. Thus, the natural gas lean in heavy hydrocarbons in line 164 can optionally be compressed, using compressor 72, prior to liquefaction in the main cryogenic heat exchanger to achieve greater liquefaction efficiency. See Figure 3.

A ratio of condensed liquids from the second separator to liquids from the first separator, i.e. a ratio of liquid streams in conduits 179 with respect to liquids in conduit 170 in case the reflux in conduit 178 is closed, may depend on the feed gas composition, and on optimization of the overall process. For instance, the process can have more or less cooling in the precool heat exchanger 10 which alters the ratio. The system and method of the disclosure have been simulated for a range of operational parameters. In a base case, said ratio may be about 7:1. In case the feed gas 2 comprises relatively large amounts of C5+, the ratio may be in the order of 17:1. The system and method of the invention provide the advantages as mentioned above over this entire range.

(a) In some conventional methods, a scrub column is used. The present invention uses a flash separator (140, 142), optionally with an integrated wash section (144). A flash separator is cheaper than a scrub column (lower capital expenditure) and is simpler to operate.

(b) In another conventional method, the purified gas stream is cooled and depressurized. The flash vessel of the prior art lacks a wash section. The present invention has two stages of flash, with an intermediate cooling step. The present invention does not require a pressure let-down step as it does not solely rely on partial condensation for HHC removal. The system of the present disclosure benefits from the wash section increasing the functionality of the flash separator with a gas and liquid mass transfer zone.

(c) Some prior art documents lack details on how cooling of the pre-treated gas stream is achieved. The method of the present disclosure uses PMR, preferably via an integrated heat exchanger to achieve the cold temperatures required. Integrated heat exchanger herein means that the respective heat exchanger also provides cooling duties for other streams than just the natural gas feed stream, for instance to the multicomponent refrigerant of the main cycle 6.

(d) In the system of GB1572899, the liquid streams from each of the separators are routed to independent demethanisers. In the present disclosure, the liquid stream from the second separator 142 may be routed to the first separator 140 where it may include an integrated wash section. The liquids from the first separator 140 (in conduit 170) may be subject to further processing for recovery of light ends which may include a demethaniser or similar fractionation column. The process enables to recover valuable hydrocarbon components which can then be either liquefied (this can be in a separate heat exchanger to the MCHE) and mixed with the LNG stream (condensed product 30 or further downstream but before end flash) or it can be precooled (this can be in a separate exchanger to the precooler) and mixed with the main natural gas stream (somewhere between feed 2 and 164).

A scrub column typically contains multiple layers of packing or trays (the latter conventionally for high pressure columns, operating at an internal pressure exceeding 35 bar). Also, a scrub column often is provided with a stripping section (reboiler). The present disclosure of the wash vessel 140 (Figure 6) is limited to one layer of packing 148 and has no stripping section. The first separator 140 of the disclosure reduces or may obviate the requirement for an expansion device, such as a JT valve, depending on feed gas composition and HHC content.

The need for pressure reduction increases when the feed gas is leaner and contains more unwanted HHCs. An example is provided with a feed gas comprising, for instance, 1300 ppmv or more C5+ in total. Herein, the feed gas may comprise 15 ppm Benzene. No additional pressure drop other than unavoidable equipment loss is required for HHC removal. This obviates a requirement for JT valve 64 (Fig. 2). Another example is feed gas containing more than 1800 ppm C5+ and 30 ppm Benzene or more. The two-stage HHC separation line-up of the present disclosure allows, for instance, for a liquefaction feed pressure at inlet 168 of the main heat exchanger 20 that is at least 3 bar greater in comparison to a single stage flash separator line-up (conventional).

The present invention can optionally be complemented with heavy hydrocarbon removal methods for removing heavy hydrocarbons in a gas pre-treatment section. Gas pre-treatment is not shown, but typically precedes the above described embodiments and provides the treated gas stream 2. For instance, heavy hydrocarbons may be partly removed in an acid gas removal unit (AGRU).

In the method of the present disclosure, the wash stream 178 is generated by further cooling and partial condensation of the gaseous overhead 152. The separation of the resultant vapor and liquid stream produce a hydrocarbon liquid 176 suitable as wash stream. As the method does not rely on downstream units to supply the liquid wash stream 178, the heavy hydrocarbon removal system of the present disclosure is self-sustaining.

The purpose of components of the system and method of the present disclosure, and the qualitative advantages and disadvantages of each component compared to alternatives are detailed in table below:

Component	**Purpose**	**Advantages**
1st Separator	Separation of vapor and liquid; removal of HHC via the liquid stream.	Simpler and cheaper than scrub column.
Wash within 1st separator	Liquid wash is used for absorption of gaseous HHC.	Improved ability to remove HHC from gaseous stream compared to phase separator; less concern on achieving sufficient liquid traffic for operational stability compared to scrub column.
2nd Separator	Separation of vapor and liquid; removal of HHC from gaseous stream; liquid stream can be used for hydrocarbon wash for 1st separator.	Further HHC removal at different pressure and temperature, allows production of liquid enriched in NGL and HHC suitable as a wash; obviates stream from fractionation or condensate stabilizer.
Temperature Bypass (line 60)	Control the temperature of the feed in the first separation system	Enables equalization of the temperature of the natural gas exiting the first bundle of a Precooler or the HP Precooler with the natural gas entering the second bundle of a Precooler or the LP Precooler

PMR Loop	Precools and partially condenses natural gas and MR	Can cool streams to approximately -60°C (colder than propane, which is used in C3MR liquefaction process, and cools to -35°C).
MR Loop	Liquefies and sub-cools natural gas	Deeper cooling.

The performance of the new line-up has been assessed by comparing it against alternative line-ups using a process simulation tool. The base case is the conventional scrub column with a Dual Mixed Refrigerant (DMR) liquefaction process with lean feed gas. Processing of lean gas requires deeper cooling of the vapor to reach the heavy hydrocarbons specification. This is due to the lower concentration of NGL components (C2 to C4) in the feed gas, which require a lower temperature to generate liquids for sufficient reflux flow for a scrub column line-up.

The DMR liquefaction process of the present disclosure has improved flexibility in this respect by improved ability to vary the precooling temperature (the cut-point) available to achieve deeper cooling and balance the power requirements of the two refrigerant systems (refrigerant loops 4 and 6 respectively). The cut-point temperature is the temperature of the process stream when leaving the precool loop 4. For instance, the temperature at outlet 160 of the (LP) precooler 12. This additional cooling can be provided via a single precooler 10 (Fig. 2). Alternatively, tuning the temperature can be provided with more than one precooler, for example the two separate PMR heat exchangers 10 and 12 (Figures 1, 3 and 4).

The following changes led to the evolution of the new line-up:

1. Two stage flash separator.

2. Two stage HHC separator with liquid wash

a. with (a single layer of) packing (Fig. 3) or

b. with liquid spray (Fig. 4).

In the assessment of the HHC Removal System 40, the following parameters were kept constant:

• Feed flowrate (flux of stream 2); and

• PMR cut-point.

For modelling and comparison with a scrub column setup (See US6370910), the scrub column inlet temperature is set at about -10°C, and the overhead temperature (i.e. the temperature of gas exiting a top end of the scrub column; see line 8 in Fig. 1 of US6370910) is about -29°C.

Line-Up	PMR Cooling Duty (MW)	Liquefaction Feed Gas			LPG Flow (t/h)	Remarks
		Pressure (kPaa)	C5+ spec (ppm)	Bz spec (ppm)		
Scrub Column (Base Case)	29.9	5973	106	0.006	3.8	Minimum liquid wetting rate marginally met[1]. Vapor and liquid density difference satisfactory. Column operating 6 bar below critical point.
Single Flash Separator (Case 1)	30.4	5973	380	3	15.9	Less equipment pressure drop however pressure across J-T valve required to achieve colder temperature. Increased NGL production. Vapor and liquid density difference reduced. Operating 2 bar below critical point. SC operating pressure: Base Case is 6288 kPaa, Case 1 is 5973 kPaa
Two Stage Separators (line-up of present disclosure)						
Flash Only (Case 2A)	30.3	5968	331	3	15.7	Increased equipment pressure drop due to second flash separator. Keeping the same mid-bundle temperature of Precooler resulted in no partial condensation in first separator.
Flash Only (Case 2B)	30.2	5968	337	2.8	14.6	Lowered mid-bundle temperature of Precooler to achieve liquid dropout in first separator improves Bz removal. Less NGLs produced compared to single flash.
Flash Only (Case 2C)	30.1	5948	331	2.7	14.6	Lowered temperature and pressure to achieve liquid dropout in first separator improves C5+ and Bz removal. Less NGLs

						produced compared to single flash.
Flash Only (Case 2D)	30.2	5948	328	2.7	15.1	Lowered temperature to achieve liquid dropout in first separator and lowered pressure in second separator further improves C5+ and Bz removal. Slightly less NGLs produced compared to single flash.
Flash with recycle (Case 2E)	29.9	5948	275	1.3	7.0	Lowered temperature to achieve liquid dropout in first separator and lowered pressure in second separator (as per Case 2D). Liquid from second separator recycled back to first separator feed. Significant improvement in HHC removal with less C1 slip into NGL stream.
Wash Section with packing (Case 3A)	29.9	5966	276	0.1	6.1	Minimum liquid wetting rate marginally met(2). Increased equipment pressure drop due to packing material. Significantly less NGLs produced compared to single flash. Operating 2 bar below critical point. Flexible to lowering the mid-bundle temperature of the Precooler for greater Bz removal.
Wash Section with liquid spray (Case 3B)	29.9	5966	282	0.5	7.4	An assumed inefficiency in mass transfer taken into account(2). Liquid to gas ratio is 2 L/m3. Significantly less NGLs produced compared to single flash. Flexible to lowering the mid-bundle temperature of the Precooler for greater Bz removal.
Wash Section with liquid spray (Sensitivity Case 3B)	29.9	5966	300	0.6	7.2	Assumed 5% liquid carryover in spray tower and 2% liquid carryover in second separator due to phase separation inefficiency.

(1) A minimum weir load of 2 and a maximum of 60 m3/(hm) [1].

(2) Reasonable minimum wetting rates of 3m3/m2h for random packing, and even lower wetting rates of 2m3/m2h known in industry.

(3) Spray tower removal efficiency conservatively taken as 40% based on comparison to tray inefficiency typically 30 – 50%. Literature indicates the mass transfer efficiency of spray towers can vary between 70 and 99%, depending on particle distribution [2].

References:

[1]. The Effect of Outlet Weir Height on Sieve Tray Performance, A. Shariat, T.J.Cai (available via, for instance: http://folk.ntnu.no/skoge/prost/proceedings/aiche-2008/data/papers/P121394.pdf).

[2]. See, for instance, https://emis.vito.be/nl/node/19426.

From this assessment, it can be deduced that for a particular lean feed gas composition, the HHC specification can be met using the simple single flash separator line-up therefore obviating the scrub column line-up. But for the same cut-point the pressure needs to be reduced to further remove HHC negatively impacting the liquefaction efficiency. It is also noted that this line-up requires more PMR cooling duty and incurs high loss of methane in the LPG or condensate stream.

A two-stage flash separator line-up gives the system flexibility to remove HHC liquid at different pressure and temperatures. Lowering the mid-bundle precool temperature and/or reducing the pressure increases HHC removal.

Supplementing the two-stage line-up with a liquid recycle or wash in the first separator enhances the HHC separation considerably, in particular for benzene component. The wash section can be with or without a built-in device. The built-in device could comprise of random or structured packing material. The liquid wash flows downwards in a thin film over the packing material whilst the gas flows upwards through the remaining space. The liquid and gas do not disperse into one another.

A wash section without a built-in device could be a spray tower. See for instance Figure 7. The liquid is dispersed in fine droplets, normally via sprayers at the top of the vessel while the gas is continuously fed from underneath, typically via the distributor 182. The spray 180 can be relatively simple to implement and there is no risk of fouling or blockage in the wash section. The mass transfer may be less efficient than packing and with higher risk of liquid entrainment in the gas outlet stream. However, the mass transfer between the continuous gas phase and liquid droplets in a spray tower may be improved using specialist internals. The separation performance of the vessel can be achieved by the addition or other internals such as a demister.

The present disclosure is not limited to the embodiments as described above and the appended claims. Many modifications are conceivable within the scope of the appended claims. For instance, features of respective embodiments may be combined. Operational parameters such as temperatures, pressures and flow rates mentioned with respect to an embodiment may be applied in other embodiments of the method and system of the present disclosure.

CLAIMS

1. A method for production of liquefied natural gas, the method comprising the steps of:

providing a natural gas stream;

producing a cooled gas stream having a first temperature, comprising cooling at least a part of the natural gas stream in indirect heat exchange with respect to a first refrigerant;

guiding the cooled gas stream to a first gas-liquid separator;

in the first gas-liquid separator, separating the cooled gas stream in a first liquid stream and a first gaseous overhead stream;

cooling the first gaseous overhead stream in indirect heat exchange with respect to at least the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature;

guiding the further cooled gas stream to a second gas-liquid separator;

in the second gas-liquid separator, separating the further cooled gas stream in a second liquid stream and a second gaseous overhead stream;

cooling the second gaseous overhead stream in indirect heat exchange with respect to a second refrigerant to produce liquefied natural gas.

2. The method of claim 1, comprising the steps of:

providing at least part of the second liquid stream to the first gas-liquid separator; and

contacting the cooled gas stream with the second liquid stream in the first gas-liquid separator.

3. The method of claim 2, wherein the step of providing the second liquid stream to the first gas-liquid separator includes using a pump to pump the at least part of the second liquid stream to the first gas-liquid separator.

4. The method of claim 2 or 3, wherein the step of providing the second liquid stream to the first gas-liquid separator includes receiving the second liquid stream at a spray

header and spraying the second liquid stream into the first gas-liquid separator using the spray header.

5. The method of claim 2 or 3, wherein the step of providing the second liquid stream to the first gas-liquid separator includes receiving the second liquid stream at a liquid distributor and distributing the second liquid stream over an internal cross-sectional area of the first gas-liquid separator using the liquid distributor.

6. The method of claim 2, wherein the step of producing the cooled gas stream comprises an additional step of controlling a flow of a part of the natural gas stream through a bypass line and feeding the part of the natural gas stream from the bypass line into the at least part of the natural gas stream after the indirect heat exchange with respect to the first refrigerant, to bypass the step of cooling the at least part of the natural gas stream and thereby control the first temperature.

7. The method of claim 6, wherein the first temperature is controlled within a margin of 2°C, prefereably within a range of 1°C.

8. The method of any of the previous claims, wherein the step of cooling at least part of the natural gas stream includes cooling the second refrigerant in indirect heat exchange with respect to the first refrigerant.

9. The method of any of the previous claims, wherein the step of cooling the first gaseous overhead stream includes cooling the second refrigerant in indirect heat exchange with respect to the first refrigerant.

10. System for production of liquefied natural gas, the system comprising:

a natural gas conduit to provide a natural gas stream;

a first heat exchanger for cooling at least a part of the natural gas stream in indirect heat exchange with respect to a first refrigerant to produce a cooled gas stream;

a first conduit for guiding the cooled gas stream from the first heat exchanger to a first gas-liquid separator and feeding the cooled gas stream into the first gas-liquid separator at a first temperature, the first gas-liquid separator being adapted for separating the cooled gas stream in a first liquid stream and a first gaseous overhead stream;

a second conduit for guiding the first gaseous overhead stream to a second heat exchanger or second heat exchange section adapted to cool the gaseous overhead stream in indirect heat exchange with respect to the first refrigerant to produce a further cooled gas stream having a second temperature lower than the first temperature;

a third conduit for guiding the further cooled gas stream to a second gas-liquid separator, the second gas-liquid separator being adapted to separate the further cooled gas stream in a second liquid stream and a second gaseous overhead stream; and

another heat exchanger for cooling the second gaseous overhead stream in indirect heat exchange with respect to a second refrigerant to produce liquefied natural gas.

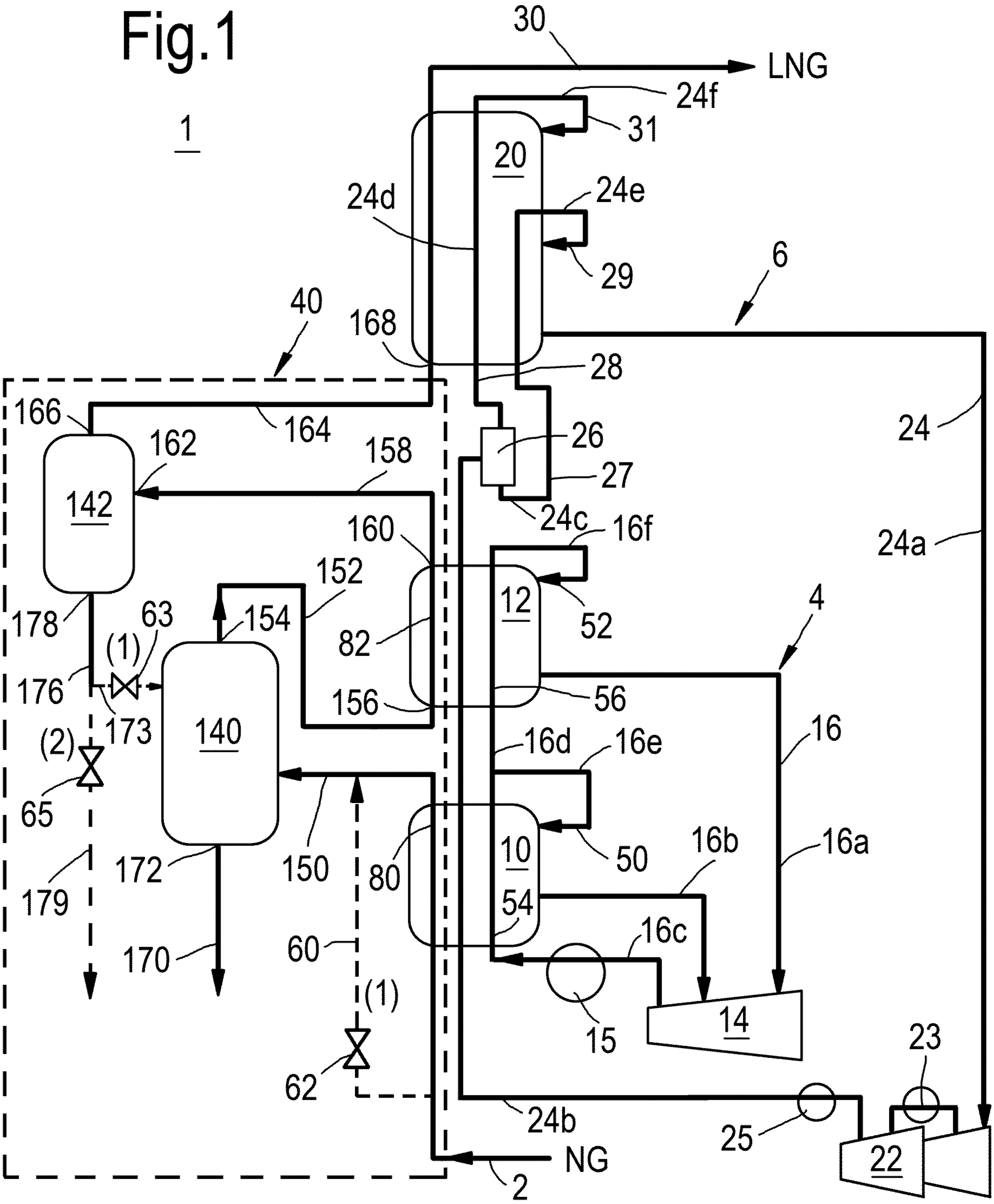
11. The system of claim 10, comprising a feedback conduit for providing at least part of the second liquid stream to the first gas-liquid separator; the first gas-liquid separator being adapted to contact contact the cooled gas stream with the second liquid stream in the first gas-liquid separator.

12. The system of claim 11, the first gas-liquid separator comprising a spray header for spraying the second liquid stream into the first gas-liquid separator or a liquid distributor for distributing the second liquid stream over an internal cross-sectional area of the first gas-liquid separator.

13. The system of claim 11 or 12, comprising a bypass line allowing a part of the natural gas stream to bypass the first heat exchanger and to be fed into the first conduit and thereby control the first temperature.

14. The system of one of claims 10 to 13, the system being adapted to implement the method of any one of claims 1 to 9.

Fig.1


1
30
LNG
24f
31
20
24e
24d
29
6
40
168
28
166
164
162
158
26
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24
142
160
24c
16f
24a
152
12
52
4
178
63
154
82
(1)
140
156
56
16
176
173
16d
16e
(2)
65
150
10
50
16b
16a
172
80
54
179
16c
170
60
(1)
15
14
23
62
24b
25
22
NG
2

Fig.2

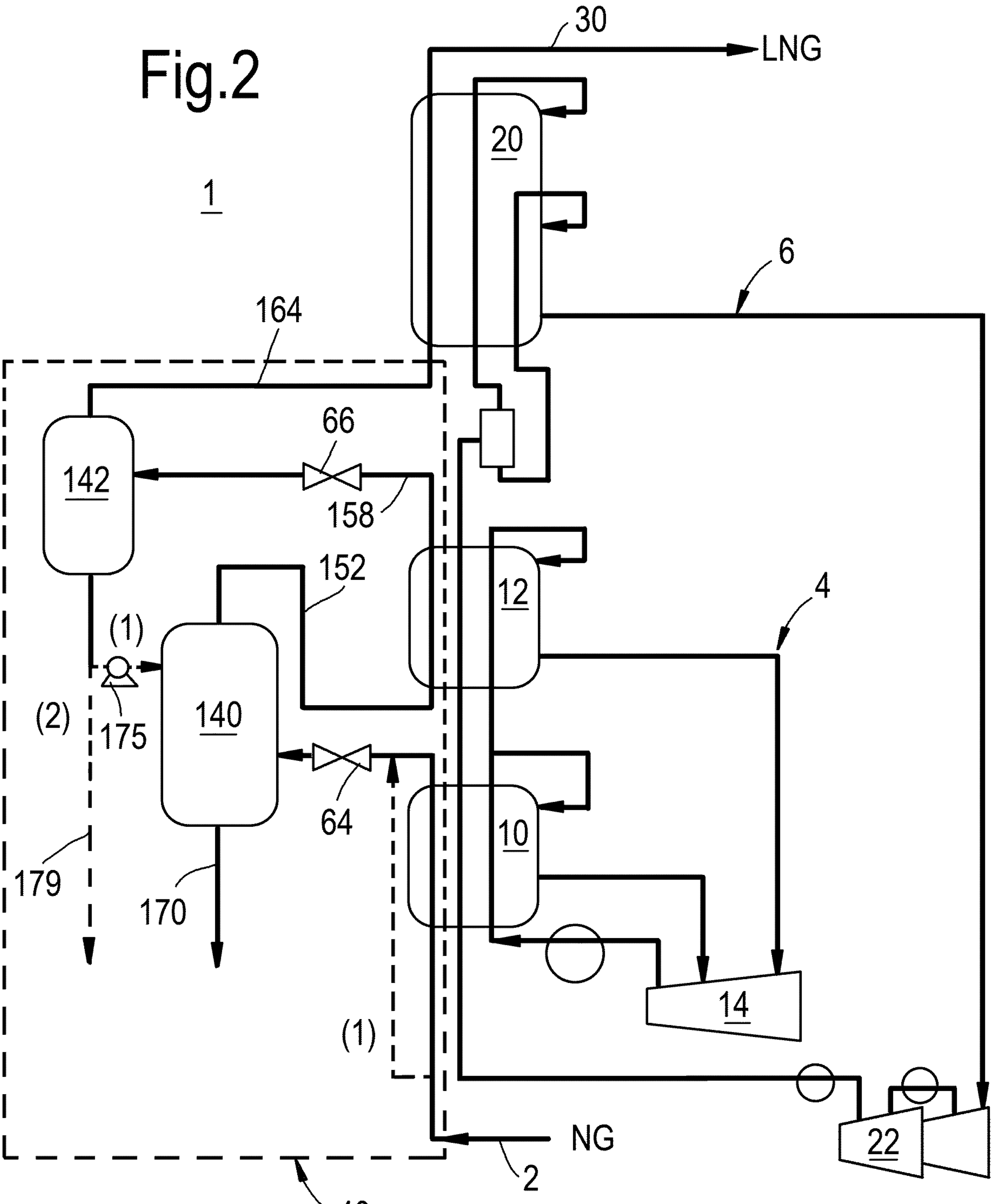

30
LNG
20
1
6
164
66
142
158
152
12
4
(1)
140
(2)
175
64
10
179
170
14
(1)
22
NG
2
40

Fig.3

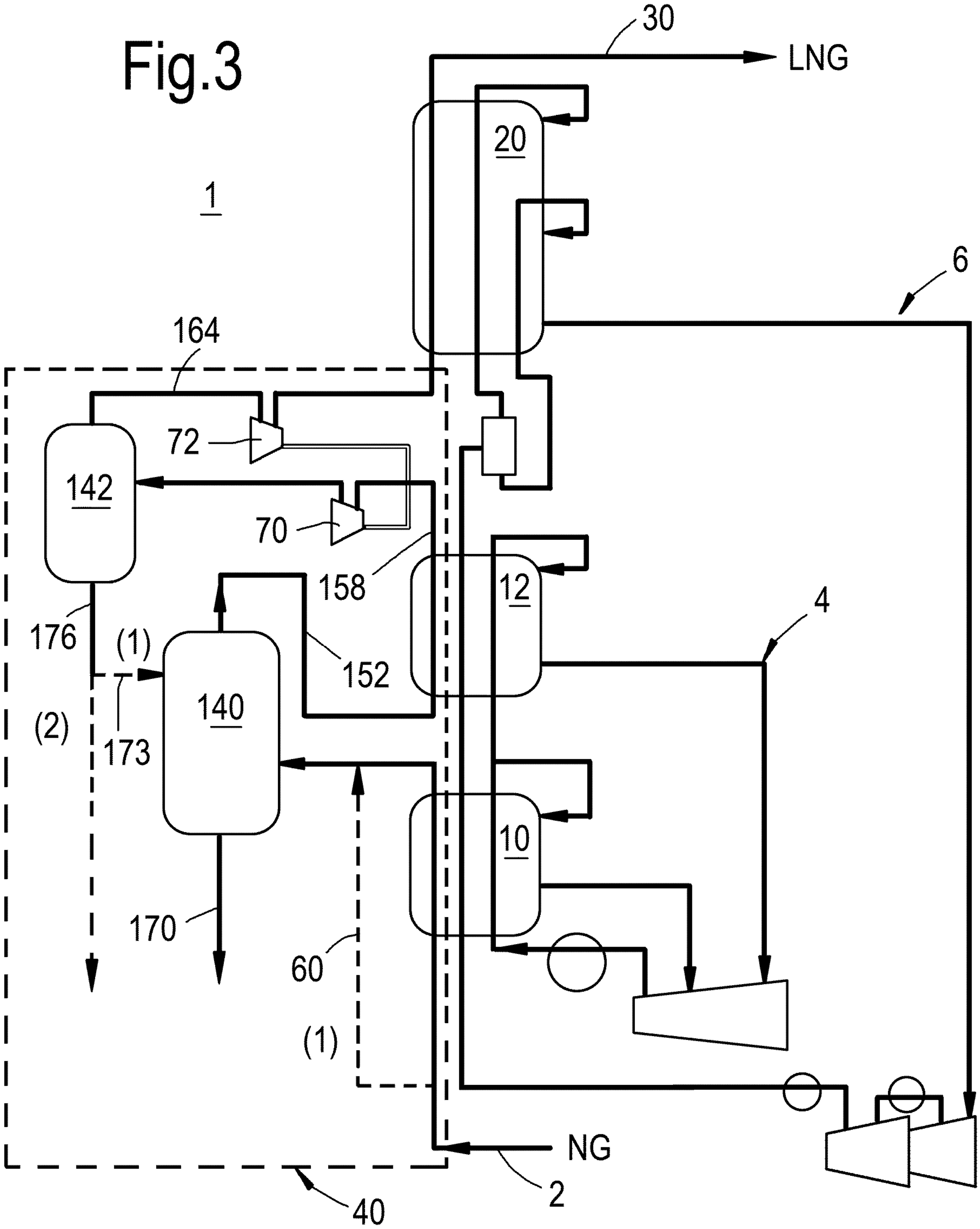

30
LNG
20
1
6
164
72
142
70
158
12
4
176
(1)
152
140
(2)
173
10
170
60
(1)
NG
2
40

Fig.4

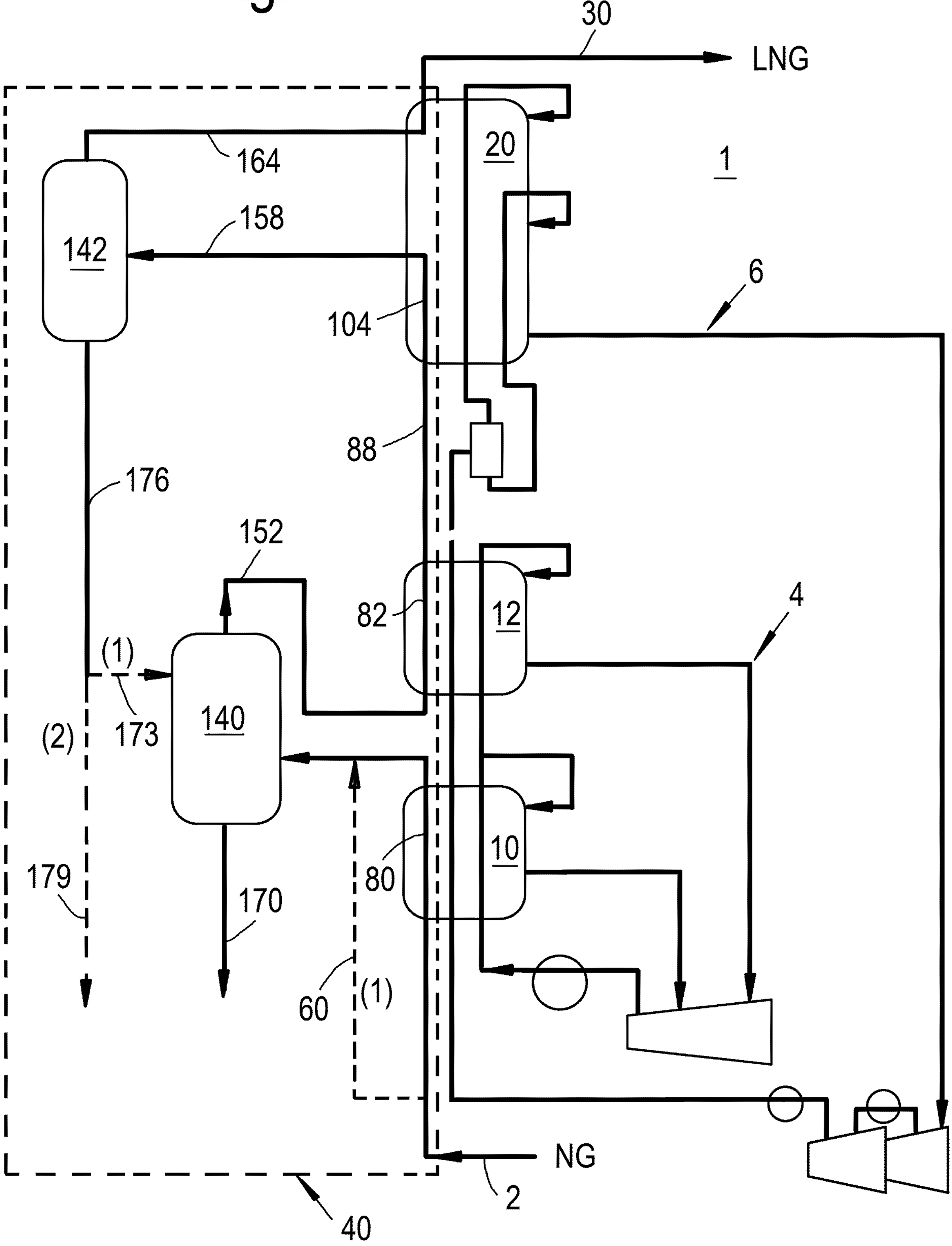
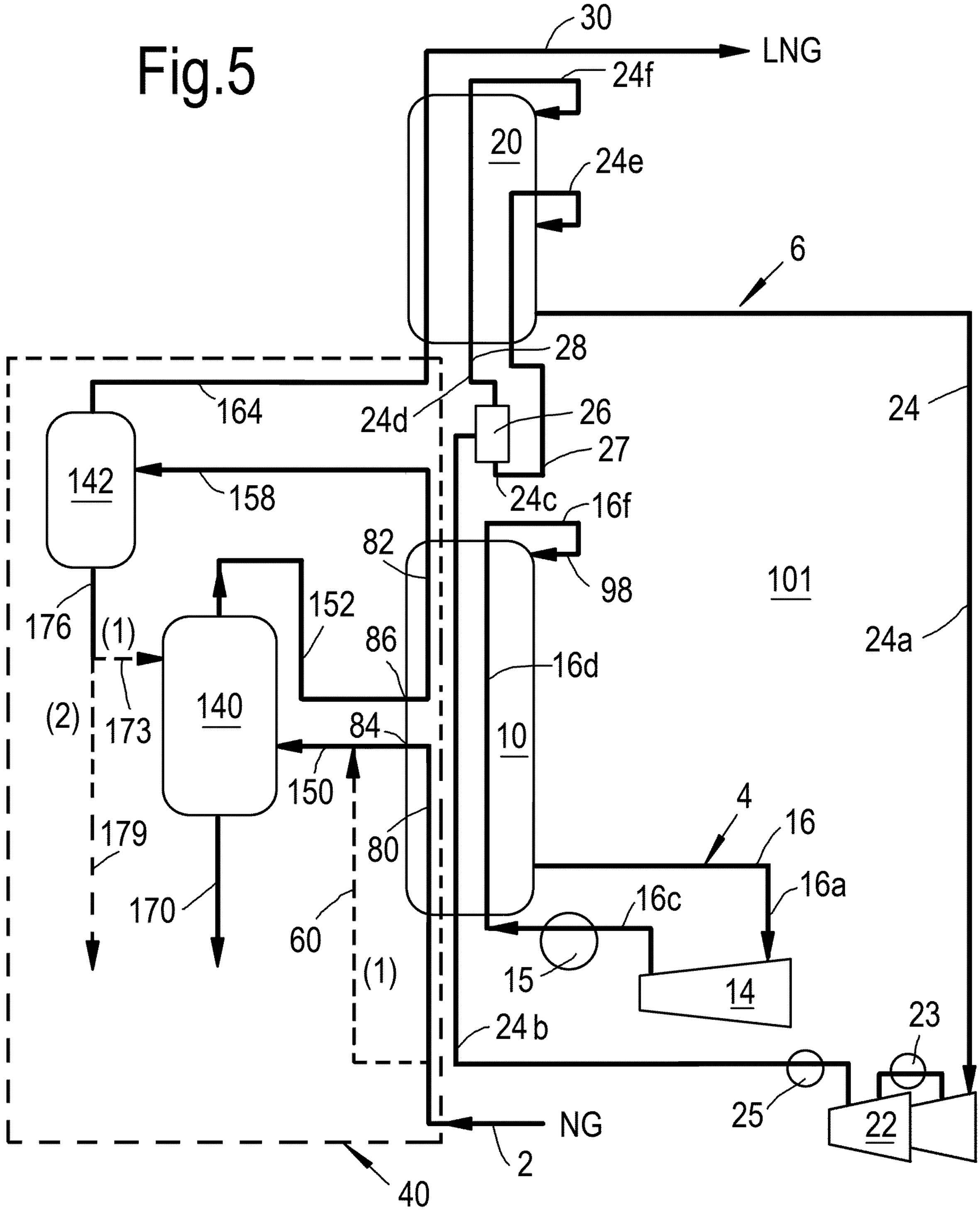

30
LNG
20
1
164
142
158
104
6
88
176
152
82
12
4
(1)
173
140
(2)
80
10
179
170
60
(1)
NG
2
40

Fig.5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24
142
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158
24c
16f
82
98
101
176
(1)
152
140
86
24a
16d
(2)
173
84
10
150
80
4
16
179
16a
170
60
16c
15
(1)
14
24 b
23
25
22
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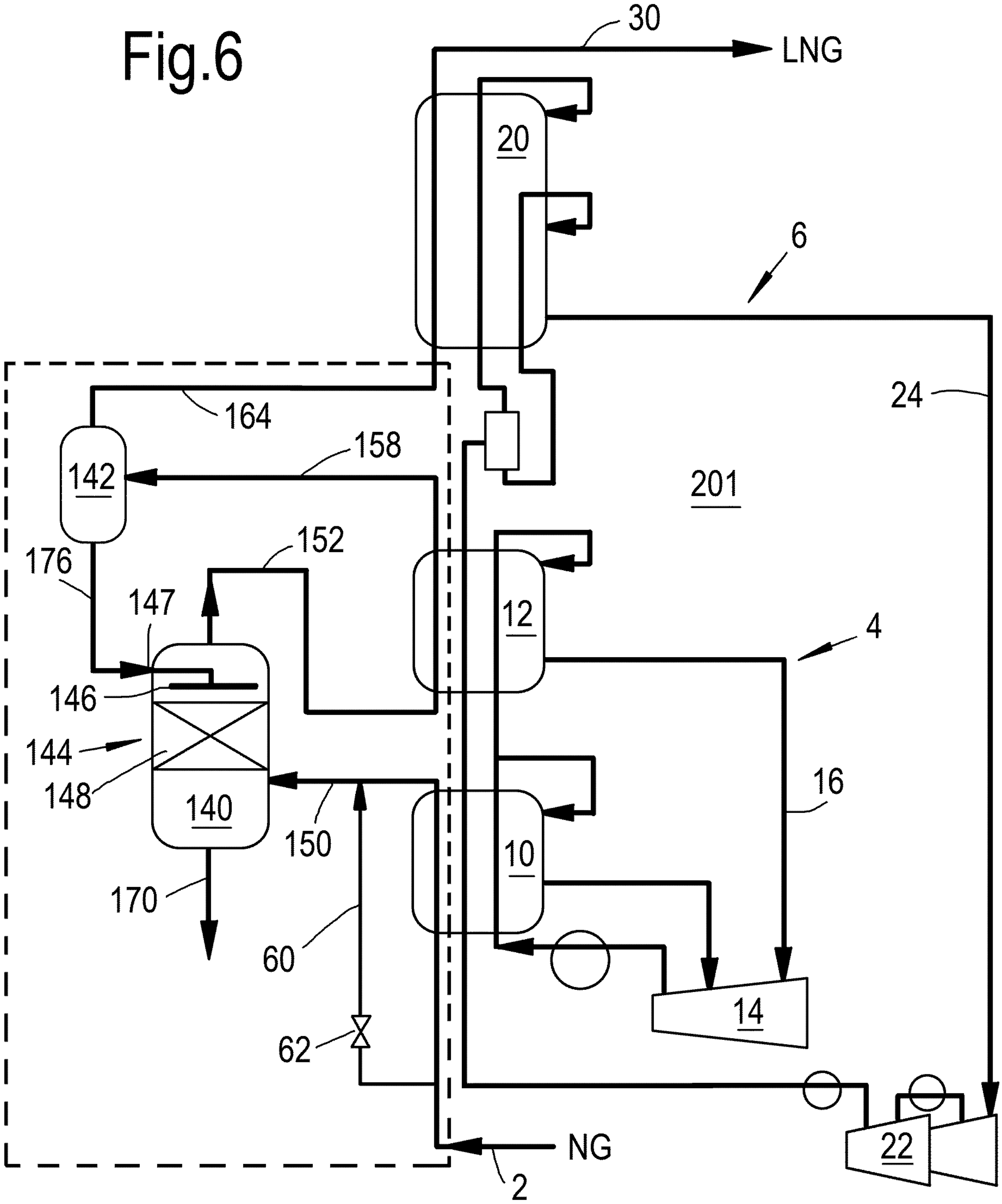
2
40

Fig.6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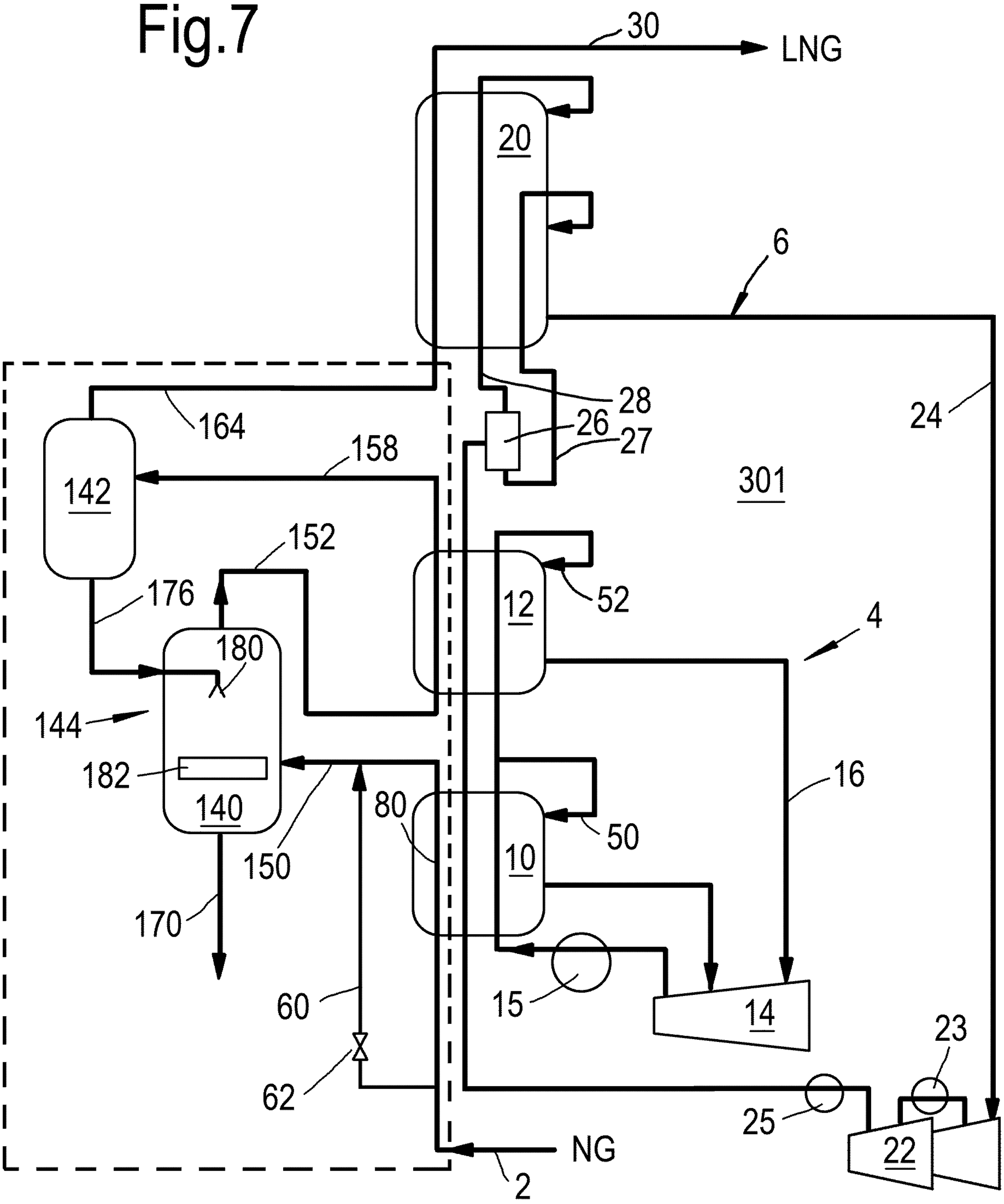
NG
2
22

Fig.7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15
14
62
23
25
22
NG
2

INTERNATIONAL SEARCH REPORT

International application No
PCT/EP2021/069914

A. CLASSIFICATION OF SUBJECT MATTER
INV. F25J1/00 F25J1/02
ADD.

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
F25J

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)

EPO-Internal, WPI Data

C. DOCUMENTS CONSIDERED TO BE RELEVANT

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X	US 2019/271501 A1 (PARAMASIVAM SENTHILKUMAR [IN] ET AL) 5 September 2019 (2019-09-05)	1-3, 8-11,14
Y	paragraphs [0054] - [0056], [0062] - [0069]; figure 1	4-7,12, 13
X	FR 2 920 529 A1 (TOTAL SA [FR]) 6 March 2009 (2009-03-06)	1-3, 8-11,14
Y	figure 1	4-7,12, 13
X	US 6 272 882 B1 (HODGES DEREK WILLIAM [AU] ET AL) 14 August 2001 (2001-08-14)	1-4, 8-12,14
Y	column 5, line 54 - column 6, line 5; figure 1	4,7,12, 13

-/--

[X] Further documents are listed in the continuation of Box C.

[X] See patent family annex.

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"&" document member of the same patent family

Date of the actual completion of the international search	Date of mailing of the international search report
7 October 2021	15/10/2021

Name and mailing address of the ISA/	Authorized officer
European Patent Office, P.B. 5818 Patentlaan 2 NL - 2280 HV Rijswijk Tel. (+31-70) 340-2040, Fax: (+31-70) 340-3016	Göritz, Dirk

2

C(Continuation). DOCUMENTS CONSIDERED TO BE RELEVANT

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