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**Bakker et al.**

(54) **METHOD FOR LIQUEFYING A HYDROCARBON STREAM**

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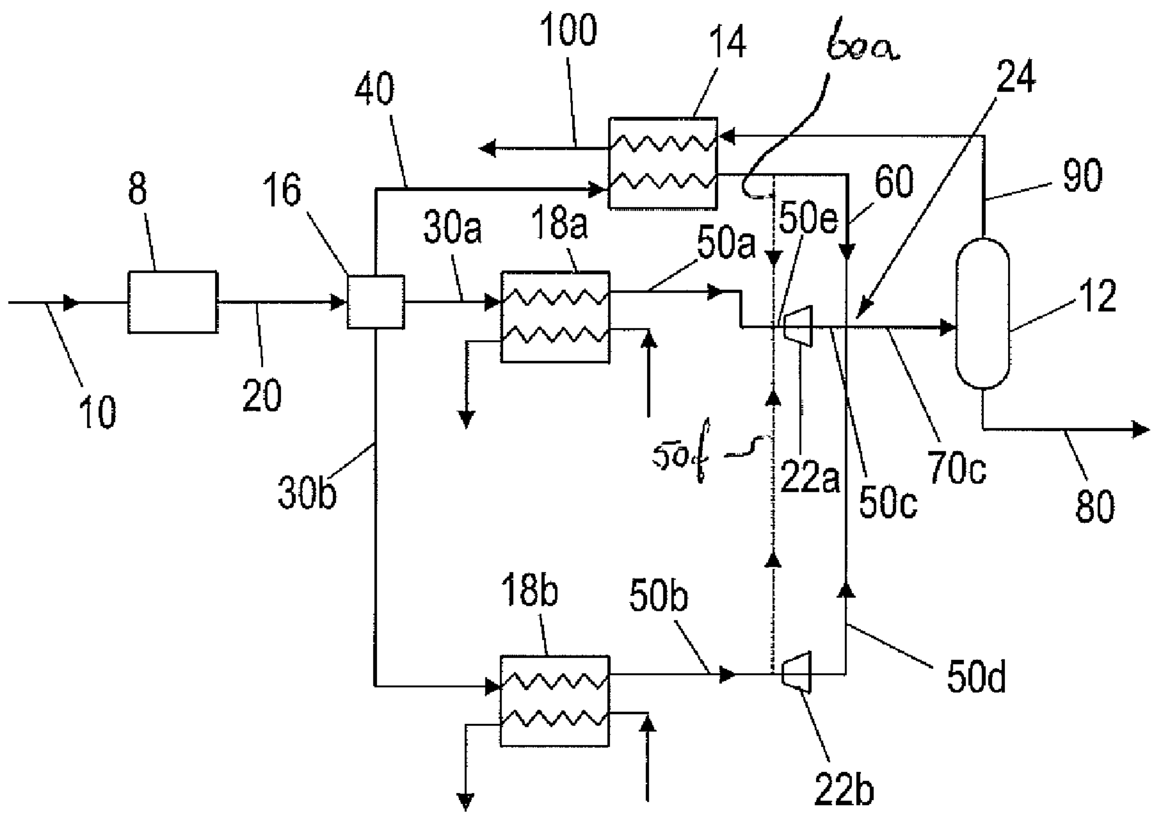
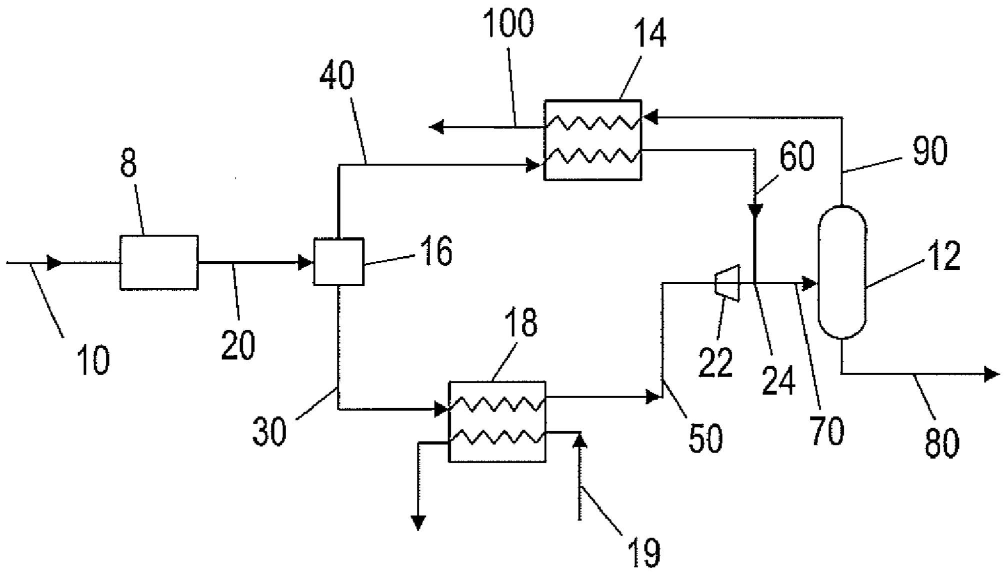
(57) **ABSTRACT**

A hydrocarbon stream is liquefied in a method wherein a feed stream is provided, from which feed stream a first stream and a second stream are produced. The first stream is liquefied in a main liquefaction system to provide a first liquefied stream, while the second stream is cooled separately to provide a cooled second stream. The first liquefied stream is combined with the cooled second stream to produce a combined stream, which is separated in a gas/liquid separator into a liquefied hydrocarbon product stream and a gaseous stream. The gaseous stream is warmed to a temperature of above −40° C. by heat exchanging against any other stream used in the method, wherein the warming of the gaseous stream employs at least heat exchanging against the second stream to provide at least part of the cooling of the second stream.




100
14
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12
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19

*Fig. 1*

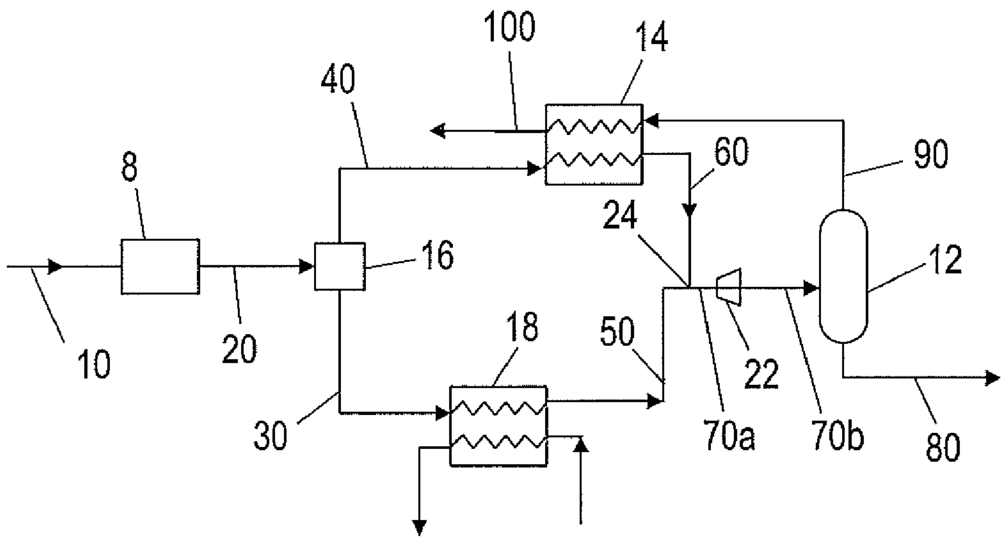
100
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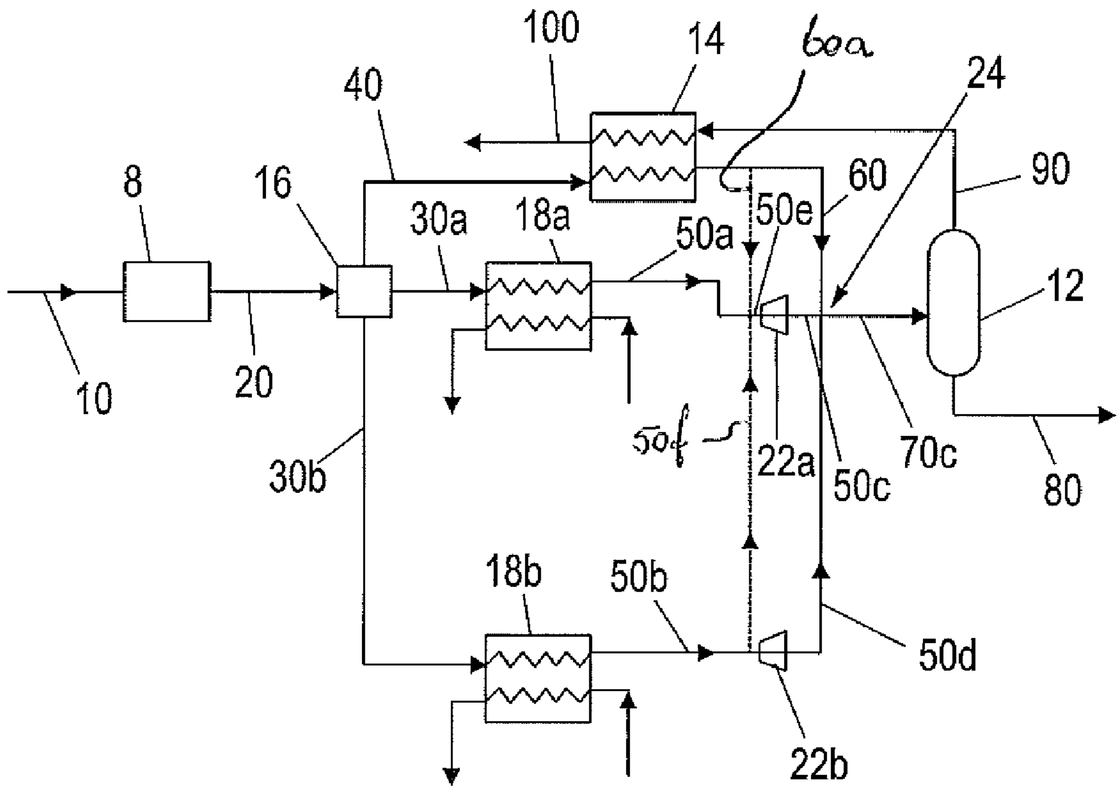
*Fig. 2*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50e
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10
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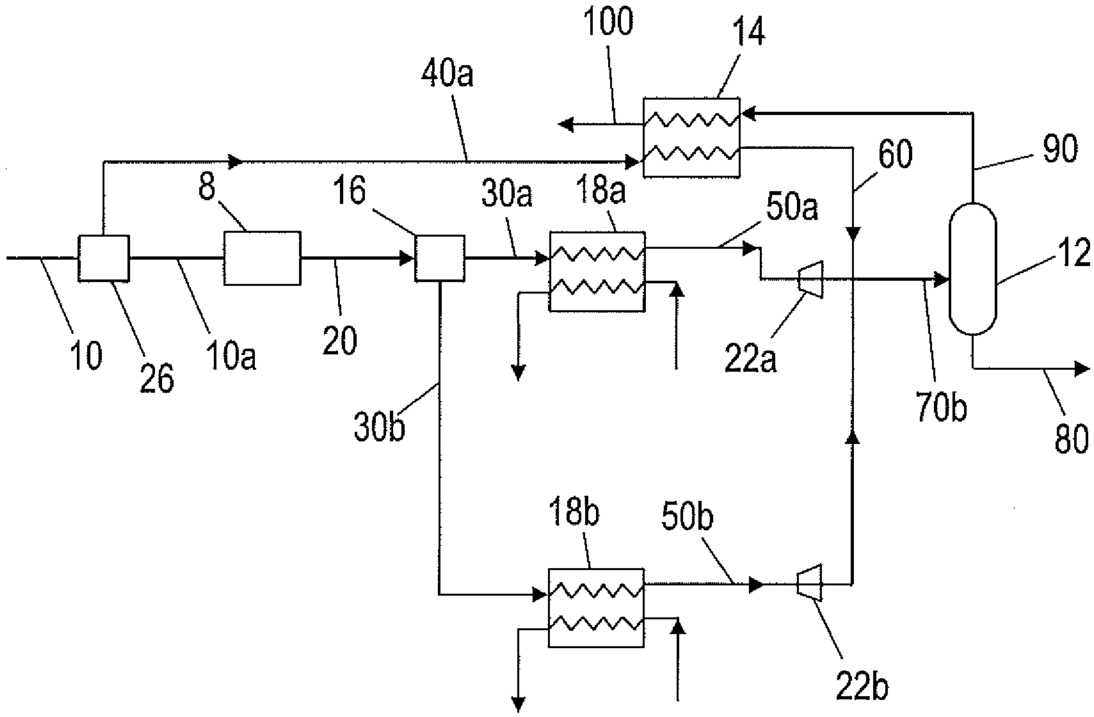
50f
22a
50c
70c
80
30b
18b
50b
50d
22b

*Fig. 3*


100
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40a
90
60
8
16
30a
18a
50a
12
10
26
10a
20
22a
70b
30b
80
18b
50b
22b

*Fig. 4*

# METHOD FOR LIQUEFYING A HYDROCARBON STREAM

[0001] This application is a continuation-in-part of International application PCT/EP2007/053681, filed 16 Apr. 2007.

[0002] The present invention relates to a method for liquefying a hydrocarbon stream.

[0003] Several methods of liquefying a hydrocarbon stream, typically in the form of a natural gas stream, thereby obtaining liquefied natural gas (LNG), are known. It is desirable to liquefy a natural gas stream for a number of reasons. As an example, natural gas can be stored and transported over long distances more readily as a liquid than in gaseous form, because it occupies a smaller volume and does not need to be stored at a high pressure.

[0004] Usually natural gas, comprising predominantly methane, enters an LNG plant at elevated pressures and is pre-treated to produce a purified feed stock suitable for liquefaction at cryogenic temperatures. The purified gas is processed through a plurality of cooling stages using heat exchangers to progressively reduce its temperature until liquefaction is achieved. The liquid natural gas is then further cooled (to reduce flashed vapour through one or more expansion stages) to final atmospheric pressure suitable for storage and transportation. The flashed vapour from each expansion stage can be used as a source of plant fuel gas.

[0005] The costs in creating and running a liquefying natural gas (LNG) plant or system are naturally high, and much is for the cooling configurations. Thus any reduction in the energy requirements of the plant or system has significant cost benefit. Reducing the cost of the cooling configuration is particularly advantageous.

[0006] U.S. Pat. No. 6,658,892 B2 relates to processes and systems for liquefying natural gas wherein a common separator and vapor compressor are used by multiple trains within the system to recover vapor both for cooling and for use as a fuel gas. A problem of U.S. Pat. No. 6,658,892 is having a separate line feeding directly into the common flash tank. It also has two separate liquefying systems, which requires full duplication of the liquefying apparatus.

[0007] The present invention provides a method of liquefying a hydrocarbon stream, the method at least comprising the steps of:

(a) providing a feed stream;

(b) producing, from the feed stream, a first stream and a second stream;

(c) liquefying the first stream in a main liquefaction system to provide a first liquefied stream;

(d) cooling the second stream of step (b) to provide a cooled second stream;

(e) combining the first liquefied stream of step (c) with the cooled second stream of step (d) to produce a combined stream;

(f) separating the combined stream in a gas/liquid separator into a liquefied hydrocarbon product stream and a gaseous stream;

(g) warming the gaseous stream of step (f) to a temperature above −40° C. by heat exchanging against any other stream used in the method, wherein the warming of the gaseous stream comprises at least heat exchanging against the second stream of step (b) to provide at least part of the cooling of step (d).

[0008] Embodiments of the present invention will now be described by way of example only, and with reference to the accompanying non-limiting drawings in which:

[0009] FIG. **1** is a general scheme of part of an LNG plant according to one embodiment;

[0010] FIG. **2** is a general scheme of part of an LNG plant according to a second embodiment;

[0011] FIG. **3** is a general scheme of part of an LNG plant according to a third embodiment; and

[0012] FIG. **4** is a general scheme of part of an alternative LNG plant arrangement.

[0013] For the purpose of this description, a single reference number will be assigned to a line as well as a stream carried in that line. Same reference numbers refer to similar components.

[0014] The method and apparatus described herein are capable of reducing the energy requirement of a liquefying plant or method by bypassing a second stream away from the main liquefaction system, and separately cooling the second stream prior to its re-combination with the other stream(s) and introduction into a gas/liquid separator. The second stream may comprise or consist of a fraction of the feed stream and/or it may be optionally already pre-cooled in a first cooling stage.

[0015] The producing, in a step (b), of the first and second streams may comprise pre-cooling, preferably commonly pre-cooling, the feed stream in a first cooling stage. The first and second streams may be produced from the pre-cooled feed stream.

[0016] The gaseous stream provided from the gas/liquid separator [in a step (f)] may be used elsewhere in the method or plant, even without recycling any part of it to the feed stream. For instance, it may be used to provide at least part of the cold necessary for the cooling of the second stream.

[0017] The cold vested in the gaseous stream of step (f) could be directly used to provide part, substantial or full cooling for any part, stream, unit, stage or process of a liquefying plant or method. This could be carried out possibly as one cooling stream or as multiple cooling streams, either in parallel or in series. This could include at least part of the liquefying of the first stream, or indeed any feed or cooled stream. It could also include cooling a refrigerant. This could be carried out by passing the gaseous stream of step (f) through one or more heat exchangers.

[0018] Thus, the gaseous stream from the gas/liquid separator can advantageously provide direct cooling to a feed or other stream without requiring any intermediate refrigerant processes or streams.

[0019] A further advantage of the present invention is that by warming the gaseous stream to a temperature higher, preferably substantially higher, than −40° C., more cold recovery is possible from the gaseous stream, thereby increasing the efficiency of the cold recovery and therefore further reducing the energy requirements of the overall liquefying plant.

[0020] Hitherto, the cold (energy) of flashed vapour from the expansion or end flash stages has been recovered in one or more heat exchangers by cooling down a fraction of a refrigerant stream, usually a Light Mixed Refrigerant (LMR) stream, in a counter-current heat exchanger. In this way, the flashed vapour is brought from a temperature level of typically about −160° C. to about −40° C., such that the full cold of the flashed vapour is not recovered prior to it being used as fuel gas. The cooled LMR stream is then used in one or more other heat exchangers to cool another stream in the plant or system.

**[0021]** Coming from the end flash of the LNG production process, the gaseous stream (which stream may also be termed a reject gas stream) may generally have a temperature of between −150° C. and −170° C., usually about −160° C. to −162° C. The temperature of the gaseous stream after the heat exchanging against one or more other streams used in the method (e.g. passing through one or more heat exchangers) may be warmed to above −40° C., preferably to above −30° C., preferably immediately following any heat exchange with the second stream.

**[0022]** In one embodiment of the present invention, the method further comprises the step of passing the second stream and the gaseous stream through a heat exchanger to at least partly provide the cooling of the second stream in step (d).

**[0023]** An advantage of this embodiment is that the second stream does not require a separate cooling system or apparatus, reducing the plant installation and energy requirements.

**[0024]** Preferably, the method of the present invention further comprises the step of:

(h) using the warmed gaseous stream, such as the exit stream of the gaseous stream from the heat exchanger, as a fuel gas stream.

**[0025]** An advantage of this embodiment is that the gaseous stream is still a useable product in an overall plant without the need to recycle to the feed stream.

**[0026]** The second stream may be cooled to a temperature sufficient to provide a combined LNG stream, preferably a combined sub-cooled LNG stream, upon combining the cooled feed stream with the first LNG stream. For instance, the second stream may be cooled by the heat exchanging in step (g) to a temperature of about −100° C. or lower, and preferably the same or similar temperature to that of the first LNG stream.

**[0027]** The hydrocarbon stream may be any suitable hydrocarbon-containing stream to be liquefied, but is usually a natural gas stream obtained from natural gas or petroleum reservoirs. As an alternative the natural gas stream may also be obtained from another source, also including a synthetic source such as a Fischer-Tropsch process.

**[0028]** Usually the natural gas stream is comprised substantially of methane. Preferably the feed stream comprises at least 60 mol % methane, more preferably at least 80 mol % methane.

**[0029]** Depending on the source, the natural gas may contain varying amounts of hydrocarbons heavier than methane such as ethane, propane, butanes and pentanes as well as some aromatic hydrocarbons. The natural gas stream may also contain non-hydrocarbons such as $H_2O$, $N_2$, $CO_2$, $H_2S$ and other sulphur compounds, and the like.

**[0030]** If desired, the feed stream may be pre-treated before using it in the present invention. This pre-treatment may comprise removal of any undesired components such as $CO_2$ and $H_2S$, or other steps such as pre-cooling, pre-pressurizing or the like. As these steps are well known to the person skilled in the art, they are not further discussed here.

**[0031]** The first stream and the second stream may be produced from the feed stream upstream of a natural gas liquids extraction system. Producing the first and second streams upstream of any natural gas liquids extraction system, or in absence of any natural gas liquids extraction system, may be accomplished by simply dividing the feed stream. Splitting of the feed stream is considered to be a form of dividing.

**[0032]** Alternatively, the producing of the first stream and the second stream may involve a natural gas liquids extraction system or process. For instance, the natural gas liquids extraction system may comprise a separator column, for example a scrub column, to divide the feed stream into a methane-enriched overhead stream and an extracted natural gas liquids bottom stream. As a first example, the overhead stream from the separator column may be divided into said first and second streams. Another example is to fractionate the extracted natural gas liquids stream and use one or more of the fractions, preferably the overhead stream from a de-methanizer, for the second stream. The separator column overhead stream could then be the first stream.

**[0033]** The division of the—optionally pre-cooled—hydrocarbon stream could be provided by any suitable divider, for example a stream splitter. Preferably the division creates two or more streams having the same composition and phases.

**[0034]** The ratio or ratios between the two or more streams produced in step (b), e.g. by dividing the feed stream, can have any value. Preferably, however, the first stream comprises a majority of the feed stream, preferably at least 60 mass %, more preferably at least 90 mass %, of the feed stream.

**[0035]** Generally, there are two streams created, whereby the smaller stream could be regarded as a 'bypass stream' because it bypasses the main liquefaction system. However, the first stream could itself be divided into a number, e.g. two or three, of part streams to be similarly liquefied and recombined. Each such part stream may be equal the other(s) in terms of mass and/or volume and/or flow rate. The dividing may be after or simultaneously with step (b).

**[0036]** In various embodiments, the first feed stream comprises at least 95 mass %, preferably at least 97 mass %, of the initial feed stream. In the alternative, the second feed stream is between 1-5 mass % of the feed stream containing natural gas, preferably between 2-3 mass % of the feed stream.

**[0037]** The gas/liquid separator may be provided in the form of any suitable vessel for obtaining a product LNG stream and a gaseous stream. Such vessels are known in the art, and may be referred to as a flash vessel.

**[0038]** In one embodiment of the present invention, the combined stream that is sent through the gas/liquid separator is expanded, that is there is a reduction of pressure prior to the gas/liquid separator. Any such expansion generally applies to the first stream which is being liquefied, and its expansion may be prior to or after combining with the cooled second stream of step (d), and/or before recombining any divided or split first stream, or any others streams of step (b).

**[0039]** Thus, the combination of streams may be before any expansion. Where the size and/or capacity of expansion is limited by design or otherwise, especially in relation to the flow of liquefied hydrocarbon to pass therethrough, then expansion of the liquefied hydrocarbon is preferably carried our prior to combination with the cooled second stream.

**[0040]** The sequence of combination and expansion of the various streams prior to passing them through a gas/liquid separator may be influenced by other parameters of the process and/or apparatus or equipment.

**[0041]** The combining of the first liquefied stream and the cooled second stream may be done using a combiner upstream of the gas/liquid separator.

**[0042]** The person skilled in the art will understand that the step of reducing the pressure of any stream may be performed in various ways using any expansion device (e.g. using a flash valve or, more preferably, a common expander or even more preferably a combination of both). Preferably, the reduction in pressure is carried out in one or more two-phase expanders.

**[0043]** Although the method according to the present invention is applicable to various hydrocarbon feed streams, it is particularly suitable for natural gas streams to be liquefied. As the person skilled readily understands how to liquefy a hydrocarbon stream, this is only briefly discussed here.

**[0044]** The liquefaction of the first feed stream is generally carried out between 20-100 bar, typically 40-80 bar. Also preferably, there is no real or significant pressure change (other than any de minimus or normal operational change, for example of 10 bar or less) of the first feed stream between its production and its expansion and/or subsequent recombination with the second feed stream.

**[0045]** The product LNG stream is preferably at a low pressure such as 1-10 bar, more preferably 1-5 bar, even more preferably ambient pressure. The person skilled in the art will readily understand that after liquefaction, the liquefied natural gas may be further processed, if desired. As an example, the obtained LNG may be depressurized by means of a Joule-Thomson valve or by means of a cryogenic turbo-expander. Also, further intermediate processing steps between the gas/liquid separation in the first gas/liquid separator and the liquefaction may be performed.

**[0046]** Typically, the main liquefaction system in the main cooling stage has a separate refrigeration circuit, and generally includes one or more separate refrigerant compressors. A non-limiting example of a typical main refrigerant is a mixture of compounds having different boiling points in order to obtain a well-distributed heat transfer. One mixture is nitrogen, ethane and propane.

**[0047]** FIG. **1** shows a general schematic arrangement of part a plant for producing a liquefied hydrocarbon gas stream from a feed stream (**10**), the plant at least comprising:

**[0048]** a first cooling stage (**8**) to cool the feed stream (**10**) and provide a pre-cooled stream (**20**);

**[0049]** a stream splitter (**16**) to divide the pre-cooled stream (**20**) into at least a first stream (**30**) and a second stream (**40**);

**[0050]** a liquefying stage including at least one heat exchanger (**18**) for liquefying the first stream (**30**) to provide a first liquefied stream (**50**);

**[0051]** a heat exchanger (**14**) to at least partly cool the second stream (**40**) to provide a cooled second stream (**60**);

**[0052]** a gas/liquid separator (**12**) to receive the first liquefied stream (**50**) and the further cooled second stream (**60**) and to provide a product stream (**80**) and a gaseous stream (**90**).

**[0053]** The plant may be part of a liquefied natural gas plant.

**[0054]** The initial feed stream **10** may contain and/or essentially consist of natural gas. In addition to methane, natural gas usually includes some heavier hydrocarbons and impurities, e.g. carbon dioxide, nitrogen, helium, water and non-hydrocarbon acid gases. The feed stream is usually pre-treated to separate out these impurities as far as possible, or at least as far as appropriate to meet LNG quality specifications; to prevent fouling/damage to equipment downstream and to prevent ice formation in equipment downstream of feed stream **10**, and provide a purified feed stock suitable for liquefying at cryogenic temperatures.

**[0055]** In operation, the feed stream **10** passes through the first cooling stage **8** to provide the pre-cooled stream **20**. The first cooling stage **8** may comprise one or more heat exchangers, which heat exchanger(s) are usually supplied with a refrigerant circulating in a first refrigerant circuit (not shown) to provide cooling.

**[0056]** The first cooling stage **8** will generally cool the feed stream **10** to a temperature of between −20° C. to −50° C.

**[0057]** The pre-cooled stream **20** is, in the embodiment as depicted in FIG. **1**, divided by stream splitter **16** into the at least first and second streams **30**, **40**, which each may have wholly or substantially the same composition, such as the same components and/or phase or phases.

**[0058]** The division of the feed stream containing the natural gas can be any ratio or ratios between the two or more streams formed by division. Generally, there is a first stream or major stream which is a major part of the initial feed stream, that is generally at least 60% by mass or volume of the initial feed stream, preferably at least 90 mass %. The second stream is generally smaller, preferably significantly smaller, than the first stream. The second stream could be regarded as a minor stream or 'bypass stream', generally being <40% by mass or volume of the initial feed stream, in some embodiments being less than 10% by mass or volume of the initial feed stream.

**[0059]** In one embodiment of the present invention, the first or major stream(s) comprises at least 95 mass %, preferably at least 97 mass %, of the initial feed stream. In the alternative, the second stream is between 1-5 mass % of the feed stream containing natural gas, preferably between 2-3 mass % of the feed stream.

**[0060]** In FIG. **1**, the majority of the pre-cooled stream **20** provides a first stream **30**, generally being at least 95 mass % of the feed stream **10**, preferably more than 97 mass %. This first stream **30** is liquefied at a pressure of between 20-100 bar, for example about 55 bar in the main liquefaction system that here for the sake of clarity is represented by the heat exchanger **18** and refrigerant line **19**. The liquefaction system is not fully drawn, as liquefaction systems are known in the art, and may include one or more cooling and/or refrigeration processes, generally including at least one heat exchanger **18** having a refrigerant stream **19** also passing therethrough. Such means are well known in the art, and are not described further herein. The liquefaction system provides the first liquefied stream **50**.

**[0061]** Meanwhile, the second stream **40** created by the stream splitter **16** is passed through another heat exchanger **14**. Heat exchangers are well known in the art, and generally involve the passage of at least two streams therethrough, wherein cold energy from one stream is recovered to cool and/or refrigerate at least one other stream running concurrently or countercurrently to the first stream. In FIG. **1**, the heat exchanger **14** cools the second stream **40** to produce a cooled second stream **60**.

**[0062]** The heat exchanger **14** could comprise more than one heat exchanger to cool the second stream **40**. Cooling of the second stream **40** may also be assisted by one or more other heat exchangers or coolers or refrigerants (not shown in FIG. **1**), either related to and/or unrelated to the scheme of the LNG plant shown in FIG. **1**.

**[0063]** Generally, the second stream **40** is cooled by the heat exchange to a temperature of less than −100° C., and preferably the same or similar temperature to that of the first liquefied stream **50**.

[0064] The first liquefied stream **50** may be reduced in pressure by passage through an expander **22**. Expanders are well known in the art and are adapted to reduce the pressure of a fluid stream passing therethrough so as to create a liquid stream and a gaseous or vapour stream therefrom.

[0065] The first liquefied stream **50** and cooled second stream **60** are combined before entering into a gas/liquid separator **12** as described hereinafter. In alternative embodiments, the first liquefied stream **50** and the cooled second stream **60** may be combined at or in the gas/liquid separator. The combining of the streams may not require full integration or mixing for their passage through the gas/liquid separator.

[0066] As shown in FIG. **1**, the cooled second stream **60** is combined with the first liquefied stream **50** by a combiner **24** such as a junction or union to produce a combined stream **70**. The combiner may be any suitable arrangement, generally involving a union or junction or piping or conduits, optionally involving one or more valves.

[0067] In FIG. **1**, the combined stream **70** can pass through a flash valve (not shown) and then on to a gas/liquid separator **12** such as an end flash vessel, wherein the liquid stream is generally recovered as a liquefied hydrocarbon product stream **80**, and a gaseous stream **90**. The product liquefied hydrocarbon stream **80**, typically having a pressure of between 1-10 bar, more typically being at ambient pressure, is then passed by one or more pumps (not shown) to storage and/or transportation facilities.

[0068] The resultant gaseous stream **90** from the end flash vessel **12** can be passed through a conduit or line to the heat exchanger **14**, through which it passes, usually counter-currently to the second stream **40**. The exit stream **100** of the gaseous stream **90** from the heat exchanger **14** can then be used as a fuel gas and/or used in other parts of the LNG plant or method.

[0069] Further, coming from the end separation of a liquefied hydrocarbon process, such as LNG production, the gaseous stream **90** (which stream may also be termed a fuel gas stream) generally has a temperature between −150° C. and −170° C., usually about −160° C. to −162° C. The temperature of the gaseous stream **90** after passing through the heat exchanger **14** will preferably become above −40° C., for example above −30° C., preferably following any heat exchange with the second stream **40**.

[0070] Depending on the temperature of the second stream **40**, the gaseous stream **90** could be heated to a temperature >0° C., even between 30° C. and 50° C., for example between 35° C. and 45° C. by such further heat exchange. Where the exit stream **100** is to be used as a fuel gas, its temperature is not critical, such that a temperature of +40° C. is acceptable. The temperature of the second stream **40** may too low to reach such temperatures in the gaseous stream **90**, which may for instance occur if the feed stream has been pre-cooled while or before producing the first and second streams. In such a case, further cold recovery may be employed, either in the heat exchanger **14** or in other heat exchangers subsequently thereto, and/or against another stream or streams in order to be able to reach a temperature >0° C. in the gaseous stream **90**. The other stream may be selected to be warmer than the second stream **40**.

[0071] By being able to raise the temperature of the gaseous stream **90** beyond −40° C. or −30° C., there are two further benefits. Firstly, if this is done in the heat exchanger **14**, the heat exchanger **14**, in particular the cold recovery exchange area, can be smaller, possibly 20% or 30% smaller than the current usual design of heat exchanger for the reject gas from a gas/liquid separator. Thus, the heat exchange area in a typical heat exchanger could be less than 2500 $m^2$, preferably less than 2000 $m^2$.

[0072] Secondly, by being able to increase the resultant temperature of the gaseous stream **90** through one or more heat exchangers from the present maximum of −40° C. (based on refrigerants used) to a temperature of at least more than +20° C., possibly +30° C. or even +40° C. or more, this energy can be used to reduce the energy required for cooling or refrigeration elsewhere in the plant or system, such as the refrigerant compressor power used for one or more other feed streams or LNG streams in the plant. It is estimated that for an LNG plant having a capacity of approximately 5 Mtpa, the cold recovery exchanger duty of the usual heat exchanger for the gaseous stream from an end flash vessel can be doubled, leading to a reduction of the main refrigerant compressor power of 1% or more. A reduction of 1% in the main compression power is significant for industrial liquification plants, for example those of 1 Mtpa output or more.

[0073] FIG. **2** shows a similar arrangement to that in FIG. **1**, but wherein the cooled second stream **60** from the heat exchanger **40** is combined with the first liquefied stream **50** prior to the reduction in pressure of the first liquefied stream **50** through the expander **22**. As mentioned hereinbefore, the cooled second stream **60** and first liquefied stream **50** can be combined by a combiner **24** such as a junction or union to produce a combined stream **70*a***. The combined stream **70*a*** then passes through the expander **22** to provide an expanded stream **70*b***, which then inflows into the gas/liquid separator **12** as described above for FIG. **1**. A valve (not shown) is provided upstream of the gas/liquid separator **12** to reduce the pressure of the flow in line **70*b*** prior to entering the gas/liquid separator **12**.

[0074] It is known to the person skilled in the art how to balance the pressure of the streams prior to any combination. For example, any pressure letdown of the cooled second stream **60** prior to the combining could be controlled by a pressure or flow control valve.

[0075] The embodiments shown in FIGS. **1** and **2** confirm that the streams may be combined either prior to or after expansion of either stream, especially the first or major stream.

[0076] FIG. **3** shows a development of the general scheme for an LNG plant shown in FIG. **1**. In FIG. **3**, the initial feed stream **10** passes through a first cooling stage **8** to provide a pre-cooled stream **20** as hereinbefore described. The pre-cooled stream **20** is then divided by a stream splitter **16** into a second stream **40** which passes through a heat exchanger **14** as hereinbefore described, and a first stream, which is itself divided into two part-first streams **30*a***, **30*b***.

[0077] The first stream could be divided into any number of part-streams, and FIG. **3** shows the division into two part-first streams **30*a***, **30*b*** by way of example only. The division of the first stream into part-first streams could be carried out simultaneously with division of the pre-cooled stream **20** by the stream splitter **16**, or thereafter.

[0078] The first stream could be divided into two or more part-first streams based on any ratio of mass and/or volume and/or flow rate. The ratio may be based on the size or capacity of the subsequent parts of the liquefaction stages or systems or units, or due to other considerations. One example of the ratio is an equal division of, optionally pre-cooled, stream mass.

**[0079]** In FIG. **3**, the part first-streams **30***a*, **30***b* are separately liquefied by liquefaction systems, each generally including at least one heat exchanger **18***a*, **18***b* respectively, to provide separate liquefied streams **50***a*, **50***b* respectively. The liquefaction systems may use a shared common refrigerant cycle to provide their cooling power, but preferably each liquefaction system employs its own separate refrigerant cycle. Liquefaction systems and process conditions for liquefaction are well known in the art, and are not described further herein.

**[0080]** Each liquefied stream **50***a*, **50***b* is reduced in pressure by passage through respective expanders **22***a*, **22***b* to provide expanded liquefied streams **50***c* and **50***d* respectively.

**[0081]** Alternatively, and as shown in dotted line **50***f* in FIG. **3**, the separate liquefied streams **50***a* and **50***b* are combined prior to their expansion or reduction in pressure, to provide a combined stream **50***e* prior to a single expander, for example the expander **22***a* shown in FIG. **3**.

**[0082]** Further alternatively, the cooled second stream **60** can be combined with either of the separate liquefied streams **50***a* and **50***b* or the combined stream **50***e*, prior to any combined expansion as schematically shown in dotted line **60***a*.

**[0083]** The arrangement as to whether to combine the separate liquefied streams **50***a* and **50***b* prior to or after expansion may be influenced by the size or capacity of the expanders. For example, where an expander has a designed economic maximum size or capacity, which size is associated with the expected flow rate of each separate liquefied stream **50***a* and **50***b*, then separate expanders such as expanders **22***a* and **22***b* shown in FIG. **3**, are suitable.

**[0084]** One example of the scheme shown in FIG. **3** is a dual heat exchanger, dual refrigerant system, with the first cooling stage **8** serving two main, preferably cryogenic, refrigeration systems. Consequently, the temperature to which the feed stream **10**, which is preferably natural gas, is pre-cooled, may be increased. Moreover, the conditions of the first cooling stage **8** and for the liquefactions, for example the compositions of the refrigerant, can easily be adapted such that an efficient operation is achieved. Further, in case one of the main liquefying systems or one of its operations has to be reduced or taken out of operation, the conditions can be adapted to work efficiently with a single main liquefaction system. In this way, the liquefaction capacity can be increased without having to add a second first cooling stage, and this saves substantial costs. An example of a pre-cooled, dual heat exchanger, dual refrigerant system is shown in U.S. Pat. No. 6,389,844 B1.

**[0085]** The arrangement in FIG. **3** has the further advantage of carrying out certain operations in a combined manner to reduce capital and running costs, compared with the need for carrying out each operation individually, i.e. needing separate and duplicated liquefaction systems, sometimes also termed ‘trains’.

**[0086]** Returning to FIG. **3**, the second stream **40** created by the stream splitter **16** passes through the heat exchanger **14** and is cooled as described hereinabove to provide a cooled second stream **60**.

**[0087]** The cooled second stream **60** and the expanded liquefied streams **50***c*, **50***d* may be combined in any known manner, and in any known combination of steps. Two or more of any of the part-first streams which are liquefied may be recombined separately or simultaneously with other part-streams, and/or the cooled second stream **60**. Such combination of streams may be prior to or after any expansion of any of the liquefied streams, or some combination being prior to and some combination being after expansion of the liquefied streams. The arrangement required for any such combination would be known to the person skilled in the art.

**[0088]** The example arrangement shown in FIG. **3** is for the combination of the cooled second stream **60** and the expanded liquefied streams **50***c*, **50***d* at one point, using a combiner **24** known in the art. The combined stream **70***c* can then pass through a gas/liquid separator **12** to provide a liquefied hydrocarbon product stream **80** and a gaseous stream **90** as described hereinabove. A valve (not shown) is provided upstream of the gas/liquid separator **12** to reduce the pressure of the flow in line **70***c* prior to entering the gas/liquid separator **12**. The cold energy of the gaseous stream **90** can be recovered through the heat exchanger **14** and optional further heat exchangers optionally against one or more other, preferably warmer, streams, to provide an exit stream **100** as described hereinabove.

**[0089]** Table I gives an overview of various data including pressures and temperatures of streams at various parts in an example process of FIG. **3**.

TABLE 1

| Stream number | Temperature (° C.) | Pressure (bar) | Mass flow (kg/s) | Phase |
|---|---|---|---|---|
| 10 | 20.3 | 57.4 | 386.3 | Vapor |
| 20 | −27.2 | 71.3 | 310.9 | Vapor |
| 30a | −27.2 | 71.3 | 151.0 | Vapor |
| 30b | −27.2 | 71.3 | 151.0 | Vapor |
| 40 | −27.2 | 71.3 | 9.0 | Vapor |
| 50a | −149.9 | 65.8 | 151.0 | Liquid |
| 50b | −149.9 | 65.8 | 151.0 | Liquid |
| 50c | −150.7 | 5.3 | 151.0 | Liquid |
| 50d | −150.7 | 5.3 | 151.0 | Liquid |
| 60 | −157.3 | 70.8 | 9.0 | Liquid |
| 70c | −151.0 | 5.3 | 310.9 | Liquid |
| 80 | −160.3 | 1.1 | 290.2 | Liquid |
| 90 | −160.3 | 1.1 | 20.7 | Vapor |
| 100 | −30.2 | 0.8 | 20.7 | Vapor |

**[0090]** The arrangement in FIG. **3** has a number of advantages. One advantage is the reduction in the number of heat exchangers needed. Hitherto separate heat exchangers are used for the reject gas and the second stream, which involve additional installations and plant machinery, as well as additional energy requirements. In FIG. **3**, there is only one heat exchanger **14** for the direct interaction of the second stream **40** and gaseous stream **90**.

**[0091]** Another advantage is that the cold energy in the gaseous stream **90** could, if desired, be recovered up to a temperature of above +0°, possibly up to +20°, +30° or even +40° C. or above, as opposed to hitherto recovering cold only up to a maximum of −40° C. or only −50° C. from a reject gas stream against a standard liquid refrigerant. This could be carried out by involving another stream to be cooled in the heat exchanger **14**, or by using the cold energy still in the exit stream **100** in another heat exchange, for example for the first cooling stage **8**. The wider temperature approach can be used to decrease the cold recovery heat exchanger **14** in general, such as the heat exchanger area. The resultant fuel gas **100** from the heat exchanger **14** is useable at +0°, +20°, +30° or +40° C. or above as an energy source for the plant.

**[0092]** The efficiency (i.e. overall energy running requirement) of the overall LNG plant is therefore benefited by being able to achieve cold recovery from the gaseous stream **90** over its entire temperature range, and by being able to transfer cold directly from the gaseous stream **90** to an incoming stream **40**, rather than through one or more intermediate refrigerant streams (with the loss of energy recovery at each exchange).

**[0093]** In a first alternative, the stream **90** is passed to an alternative one or more heat exchangers to recover the cold energy therefrom, said heat exchanger(s) preferably being part of an LNG liquefaction system, such as the liquefaction heat exchanger **18** shown in FIG. **1**.

**[0094]** FIG. **4** shows a scheme for part of an alternative LNG plant arrangement, wherein the feed stream **10** is divided by a stream splitter **26** prior to the first cooling stage **8**, to provide a first stream **10***a* and a second stream **40***a*. The first stream **10***a* can then pass through a first cooling stage **8** as described hereinabove, to provide a pre-cooled stream **20**, which is then divided by a stream splitter **16** into two part-first streams **30***a*, **30***b*, which streams may then be liquefied and combined as described hereinabove for FIG. **3**. The second stream **40***a* can pass through a heat exchanger **14** to provide a cooled second stream **60** in a manner described hereinabove. The combination of the streams and their subsequent processing is also similar to that as described above.

**[0095]** The methods and apparatuses (plants) disclosed herein simplify the use of vapour from a flash tank, and thereby reduce the energy requirements of a liquefying plant or method. Moreover, the hydrocarbon streams are better stream-lined through the liquefying plant or method.

**[0096]** The person skilled in the art will understand that the present invention can be carried out in many various ways without departing from the scope of the appended claims.

**1**. A method of liquefying a hydrocarbon stream, the method at least comprising the steps of:

(a) providing a feed stream;

(b) producing, from the feed stream, a first stream and a second stream;

(c) liquefying the first stream in a main liquefaction system to provide a first liquefied stream;

(d) cooling the second stream of step (b) to provide a cooled second stream;

(e) combining the first liquefied stream of step (c) with the cooled second stream of step (d) to produce a combined stream;

(f) separating the combined stream in a gas/liquid separator into a liquefied hydrocarbon product stream and a gaseous stream;

(g) warming the gaseous stream of step (f) to a temperature above −40° C. by heat exchanging against any other stream used in the method, wherein the warming of the gaseous stream comprises at least heat exchanging against the second stream of step (b) to provide at least part of the cooling of step (d).

**2**. The method as claimed in claim **1**, wherein said producing of the first stream and the second stream comprises passing the feed stream through a first cooling stage.

**3**. The method as claimed in claim **2**, wherein the passing of the feed stream through the first cooling stage yields a pre-cooled stream, whereby the first stream and second stream are produced from the pre-cooled stream.

**4**. The method as claimed in claim **3**, wherein said producing of the first stream and the second stream comprises dividing the pre-cooled stream into at least the first stream and the second stream.

**5**. The method as claimed in claim **1**, wherein said producing of the first stream and the second stream comprises dividing the feed stream into at least the first stream and the second stream.

**6**. The method as claimed in claim **1**, said warming in step (g) is achieved by passing the gaseous stream through one or more heat exchangers.

**7**. The method as claimed in claim **6**, wherein the heat exchanging of the gaseous stream against the second stream of step (b) comprises passing the second stream through at least one of the one or more heat exchangers.

**8**. The method as claimed in claim **6**, wherein the temperature of the gaseous stream immediately after having passed through the one or more heat exchangers is above 0° C.

**9**. The method as claimed in claim **1**, further comprising the step of:

(h) using the warmed gaseous stream as a fuel gas stream.

**10**. The method as claimed in claim **1**, wherein after step (e) and before step (f) the combined stream is passed into the gas/liquid separator.

**11**. The method as claimed in claim **1**, wherein the pressure of the first liquefied stream is reduced prior to step (e).

**12**. The method as claimed in claim **1**, wherein the pressure of the combined stream of step (e) is reduced before said separating in step (f).

**13**. The method as claimed in claim **1**, wherein the first stream, in step (b), is produced in the form of two or more part-first streams.

**14**. The method as claimed in claim **13**, wherein the part-first streams are separately liquefied to produce separate liquefied part-streams, and wherein the separate liquefied part-streams are combined prior to, simultaneously with, or after step (e).

**15**. The method as claimed in claim **14**, wherein the pressure of each liquefied part-stream is reduced prior to step (e).

**16**. The method as claimed in claim **1**, wherein the first stream comprises at least a majority of the feed stream.

**17**. The method as claimed in claim **1**, wherein the second stream comprises preferably between 1-5 mass % of the feed stream.

**18**. The method as claimed in claim **1**, wherein the second stream is cooled to a temperature sufficiently low to provide a combined liquefied stream upon said combining in step (e) with the first liquefied stream.

**19**. The method as claimed in claim **1**, wherein the second stream is cooled in step (d) to a temperature of less than −100° C.

**20**. The method as claimed in claim **1**, wherein the second stream is cooled in step (d) to about the same temperature as that of the first liquefied stream.

**21**. The method as claimed in claim **1**, wherein the temperature above −40° C. in step (g) is a temperature above −30° C.

**22**. The method as claimed in claim **1**, wherein the temperature above −40° C. in step (g) is a temperature above 0° C.

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