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A preliminary analysis of floating production storage and offloading facilities with gas liquefaction processes

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Abstract:

Floating, production, storage and offloading (FPSO) plants are facilities used in upstream petroleum processing. They have gained interest because they are more flexible than conventional plants and can be used for producing oil and gas in deep-water fields. In general, gas export is challenging because of the lack of infrastructure in remote locations. The present work investigates the possibility of integrating liquefaction processes on such facilities, considering two mixed-refrigerant and two expansion-based processes suitable for offshore applications. Two FPSO configurations are considered in this work, and they were suggested by Brazilian operators for fields processing natural gas with moderate to high content of carbon dioxide. The performance of the combined systems is analysed by conducting energy and exergy analyses. The integration of gas liquefaction results in greater power consumption and exergy destruction than in the baseline cases with compression and injection. Gas liquefaction systems with expansion-based processes are the least performant: the exergy destruction reaches up to 50 MW, and may represent up to 120% of the exergy destruction in the processing plant.

Keywords:

Gas liquefaction, Oil platforms, Thermodynamic analysis, Process integration

1. Introduction

Oil and gas platforms are energy-intensive systems, consuming from a few to several hundred MW of electrical – mechanical power, and using as well from a few to several tens MW of thermal energy. The exact demands of these facilities depend on the field conditions (e.g. temperature and pressure), feed properties (e.g. oil, gas, water and carbon dioxide contents), and operating strategies (e.g. gas injection or export). As discussed in the work of Bothamley [1], these differences lead to different platform designs, with regards to e.g. the number of separation and compression stages, and the implementation of complex processes such as natural gas liquids recovery. However, as pinpointed in Voldsund et al. [2], the structural design of an oil and gas plant stays similar across different fields, and operations such as separation, compression, pumping, and power generation are present on *all* facilities.

In the case of Brazil, floating production, storage and offloading (FPSO) facilities have gained more and more attention, especially in the Campos Basin. At the difference of `fixed' facilities, FPSOs are ship-shaped vessels which may be used for producing oil and gas from deep- to ultra-deep water fields, i.e. with water depths that exceed 1500 m. They may be operated at large distances from the shore, meaning that gas exportation through pipelines is impractical. Offshore liquefaction of natural gas gives the opportunity to avoid flaring or reinjection, while monetising small- and medium-scale fields in remote locations. The interest for floating liquefied natural gas facilities has therefore gained interest in the last years, since several fields could be exploited consecutively. The

first well-known FLNG is the Prelude plant, which was launched by Shell in 2013, but these technologies are still scarce [3].

Natural gas liquefaction processes can be divided into three sub-categories depending on the configuration of the refrigeration cycle and on the type of refrigerant [4]. These groups are the cascade, mixed-refrigerant and expansion-based systems. At present, most of the existing base-load liquefaction plants use large mixed-refrigerant or cascade processes because of their higher performance. However, requirement for small-scale or offshore facilities are different, as pinpointed by Barclay and Denton [5]. The performance of the LNG process (e.g. power consumption and efficiency) is of key importance in all cases, but other factors such as the equipment count, dynamic behaviour, compactness, safety, etc., are important as well. For example, the pure-refrigerant cascade process includes up to ten pressure levels and is considered as a reference plant efficiency-wise [6], but it is not suitable for offshore applications because of the high equipment count and large required space. In general, mixed-refrigerant systems are harder to implement in an offshore environment, because the refrigerant is generally two-phase. They are therefore more sensible to the plant motion, as this can cause maldistribution issues in the heat exchangers. On the contrary, expansion-based processes are interesting for offshore implementation because of (i) the system simplicity and compactness because of the high pressure levels, (ii) the inherent cycle safety as nitrogen is non-flammable, (iii) the insensitivity to motion as the refrigerant is always in gas phase, (iv) good start-up characteristics compared to cascade and mixed-refrigerant processes.

The literature presents some contradictory findings. For instance, on the one hand, Cao et al. [7] evaluate two types of small-scale liquefaction processes and conclude that the expansion-based system under study is more efficient than a single mixed-refrigerant process without precooling. On the other hand, Remeljej and Hoadley [8] analyse four liquefaction configurations and claim that processes with mixture are the most efficient ones. Chang et al. [9] assess the maximum exergy efficiency for several types of gas liquefaction systems and demonstrate, for their particular application, that reverse Brayton cycles can achieve higher efficiencies than basic mixed-refrigerant processes. These studies investigate the performance of LNG processes, but do not discuss the integration of such systems with the petroleum processing plant. It is therefore challenging to evaluate how the performance of the entire LNG-FPSO system can be improved.

Several analyses of primary petroleum processing plants were conducted in the last decades, and a survey of the existing literature shows that the interest in this field has grown in the last years. Oliveira and Van Hombeeck [10] conducted an exergy analysis of a Brazilian oil platform, from the separation to the compression modules. They show that the most resource-demanding processes were the petroleum heating and gas compression. The works of Voldsund et al. [11] and Nguyen et al. [12] on Norwegian oil platforms also pinpoint the gas compression section, in addition to the production manifolds. A further work of Nguyen et al. [13] deals with the thermodynamic performance of oil and gas systems at different stages of a Norwegian field lifetime. The authors find that (i) most exergy destroyed on an offshore plant is destroyed in the power plant (gas turbines) but is mostly unavoidable, and that (ii) the gas treatment process and compression is always responsible for most of the thermodynamic irreversibility taking place in the processing plant. More recently, Sánchez et al. [14] assess the exergetic performance of a FPSO plant, and show that the exergy destruction is the greatest in the gas turbines, followed by the main compressors used to inject carbon dioxide and gas. They also show that the FPSO consumes and destroys the highest amount of exergy, on a specific oil basis, when operating in end-life conditions. The analysed platforms present valuable findings, but none actually investigates the integration of LNG processes on such facilities.

The present work aims at addressing these gaps and has two main objectives. The first purpose is to compare on an identical and consistent basis LNG processes that have attracted interest for offshore implementation. The second goal is to investigate the integration of these systems into two different FPSO configurations where gas export is a possibility. The performance of the FPSO and LNG processes is evaluated based on thermodynamic criteria (energy and exergy).

2. Methods

2.1. System description

2.1.1. Floating, production, storage and offloading facilities (FPSOs)

The main features of the typical oil and gas processing on floating, production, storage and offloading facilities is shown in Fig. 1. The reservoir fluids from the producing wells enter one or several production manifolds, where all streams are mixed and the pressure is reduced. The feed streams enter then the separation section, in which oil, gas and water are separated in several stages, with a decreasing pressure. Petroleum heating may be necessary if the feed streams enter at temperatures lower than 60°C or if they have a high viscosity. In general, heating is *always* required in Brazilian fields, whist it is generally *not* essential for Norwegian ones. The separated oil or condensate is treated in a separate section and pumped to the desired pressure before further export. The produced water is discharged into the sea. The recovered gas from the final separation stages is first recompressed in a vapour recompression unit, back to the initial feed pressure. It is then mixed with the gas recovered at the first stage, and compressed further for dehydration, where water is removed by glycol absorption. Part of the purified gas is sent to the fuel gas section and is then fed to the gas turbines, where it is consumed, and power and heat are generated.
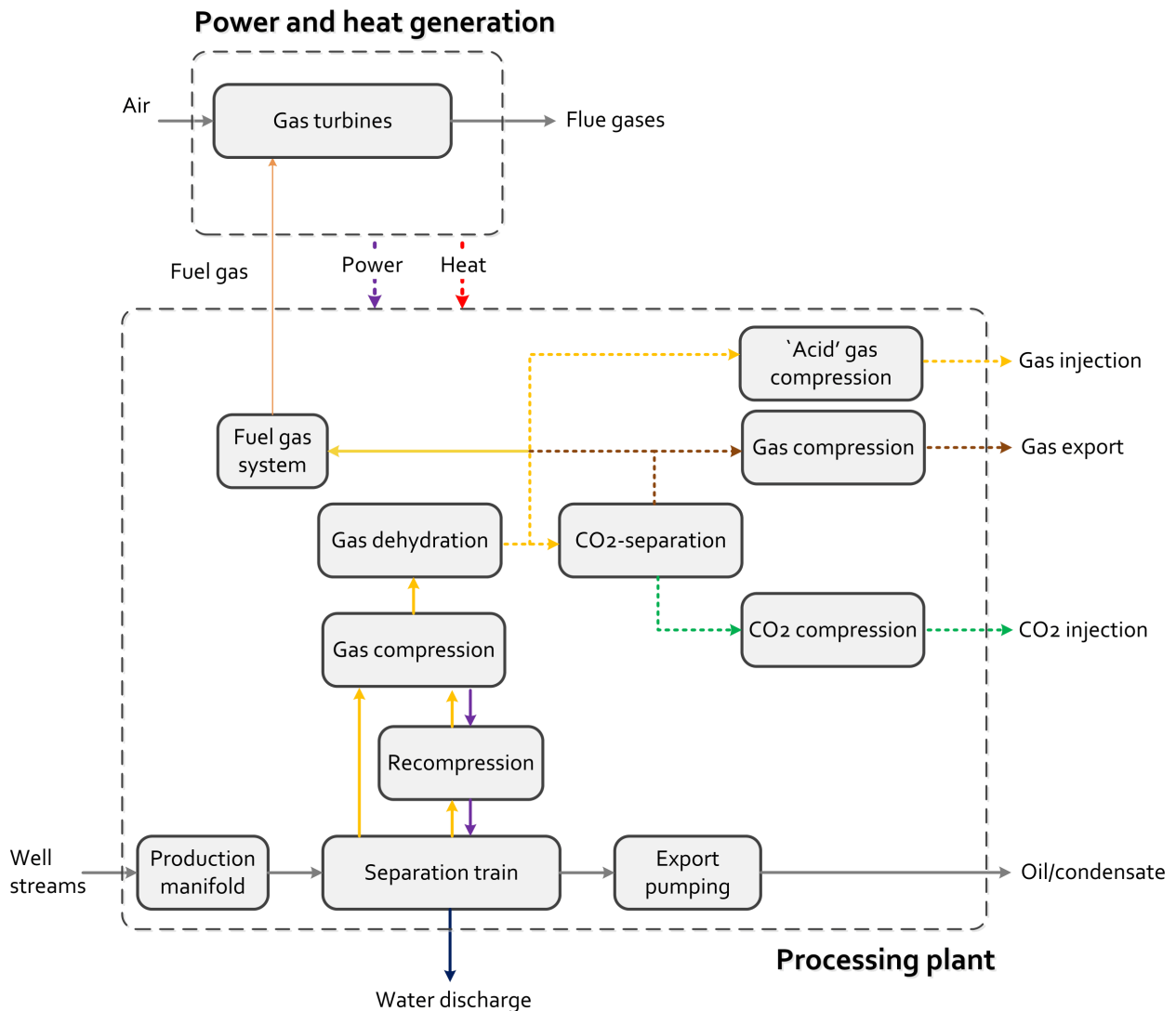

Fig. 1. Block diagram of a typical Brazilian floating production storage offloading (FPSO) platform.

The next process operations are specific to Brazilian FPSOs. For example, for most Norwegian fields, gas is further compressed for export to the shore, or injected into the reservoir or the wells for lifting oil. In the present case, the recovered gas in the fields of the `pre-salt' Brazilian region has a significantly high content of carbon dioxide, which can be as low as 5% to as high as 60%.

Purification of the acid gas is therefore required if the gas should be exported, and this is generally achieved by use of membranes. High contents of carbon dioxide are not desired, because those would result in pipeline corrosion if the gas is exported as a liquid, or in heat exchanger plugging if the gas is first liquefied. The recovered carbon dioxide is then further compressed to app. 55 MPa, while the exported gas is then compressed to app. 25 MPa. Carbon dioxide treatment is not required in the case that all gas should be injected, and the gas injection pressure is app. 55 MPa.

The choice of purifying, injecting or exporting gas depends on the initial petroleum composition. According to the FPSO operators, it is preferable to export gas in the first stage of a field lifetime, when the ratio oil – gas to water is at its maximum, and the oil and gas production are near their peaks. In the declining phase, when the crude oil contains about 50% of water (50% BSW), part of the gas is treated and then exported, while the remaining gas is directly injected. Finally, in end-life conditions, when the water and carbon dioxide contents represent more than 80 – 90% of the total feed content, the acid gas is directly injected, and clean gas is imported from other facilities to fuel the gas turbines. The present works focuses on the two first stages and operating conditions, i.e. when gas is either partly or completely exported.

The process simulations were carried out with Aspen HYSYS ® version 8.6 based on the Peng-Robinson [15] equation of state. The following assumptions are considered:

- a crude production of 150 000 barrels per day is assumed, which corresponds to a feed input of about 627 tons per hour;
- the well-fluid (excluding water) consists of app. 60% of methane, 15% of medium-weight hydrocarbons (e.g. ethane, propane and butanes), and the remaining fraction is composed of heavy-weight hydrocarbons (C5+) and impurities such as carbon dioxide and nitrogen;
- dilution water is required to enhance separation, and it amounts to 4% for the case with maximum oil and gas production;
- two case studies are considered for the scaling of the gas liquefaction process, a first case with maximum gas production and full exportation of gas, and a second case with maximum gas production, but 50% of the produced gas is injected for enhanced oil recovery
- in the 50% BSW mode, 50% of the gas bypasses the carbon dioxide separation system and is injected into the reservoir, while the remaining 50% is treated. Most is actually exported, as only a small fraction is used for power generation purposes;
- processes such as oil desalting, gas dehydration with glycol and carbon dioxide separation with membranes are *not* investigated in details, and perfect separation is assumed in these cases;
- an adiabatic efficiency of 75% is assumed for all centrifugal compressors present in the compression and recompression trains;
- an efficiency of 96% is assumed for the generators;
- the auxiliary power demand (e.g. secondary utilities) represents an additional 42% of the process power consumption;
- the maximum discharge pressure of the `acid' gas is app. 31 MPa, while the export pressure is 25 MPa and the injection pressure 55 MPa;

2.1.2. Gas liquefaction processes

Natural gas liquefaction processes can be divided into three sub-categories: the cascade, mixed-refrigerant and expansion-based systems. Cascade systems generally use three pure refrigerants with different boiling temperatures, such as propane, ethylene and methane, in three closed refrigeration cycles. Natural gas is first precooled to app. -30°C with propane, then liquefied with ethylene, and subcooled down to app. -162°C with methane. Each step may actually consist of three to four stages to reach a higher performance of the cascade process. About nine to ten heat exchangers are therefore required, which results in high equipment count and large facilities.

Mixed-refrigerant systems use, as their name suggests, mixtures of refrigerants (generally methane, ethane, ethylene, propane, butanes, pentanes and nitrogen), and the cooling effect is produced by the Joule-Thomson effect (i.e. adiabatic expansion through a valve). The most well-known processes of this type are (i) the single mixed refrigerant process (Figure 2), with only one refrigeration cycle; (ii) the propane pre-cooled mixed-refrigerant system (C3MR), which consists of two refrigeration cycles: a precooling stage with propane, and a liquefaction – subcooling stage with a mixed refrigerant; and (iii) the dual mixed refrigerant process (Figure 3), which consists of two refrigeration cycles, each using a different mixture.
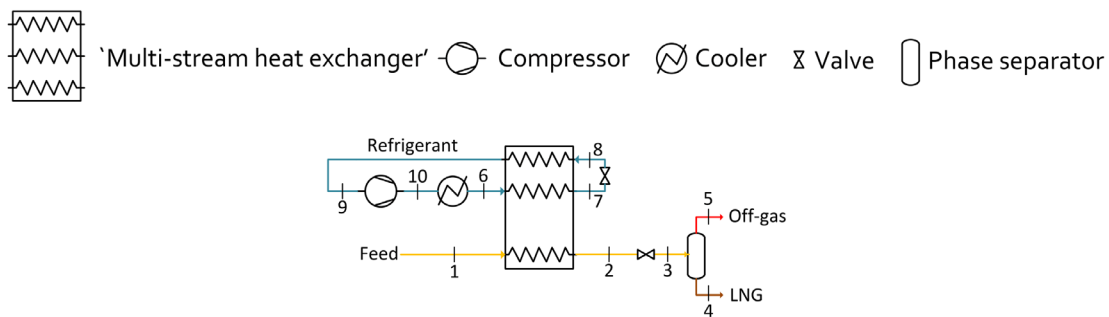

Fig. 2. *Process flow diagram of the single-mixed refrigerant process (SMR).*

All the mixed-refrigerant cycles are characterised by the following advantages. First, a better heat transfer can be expected in the heat exchangers, as the latent heat of vaporisation of the refrigerant(s) is used to cool and condense the natural gas. Secondly, the mixture composition can be tuned to the case study to obtain a better match of the temperature – heat profiles.
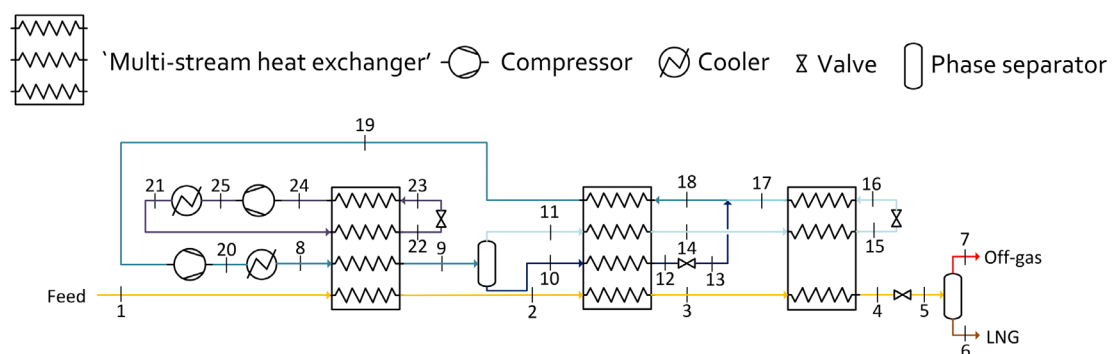

Fig. 3. *Process flow diagram of the dual-mixed refrigerant process (DMR).*

Finally, expansion-based processes are based on the use of a reverse Brayton cycle, and the cooling effect is generated by vapour expansion through a turbo-expander from which power is extracted. In general, the refrigerant medium is nitrogen or methane, and the use of mixtures is uncommon. The simplest configuration is the single reverse Brayton process (Figure 4).
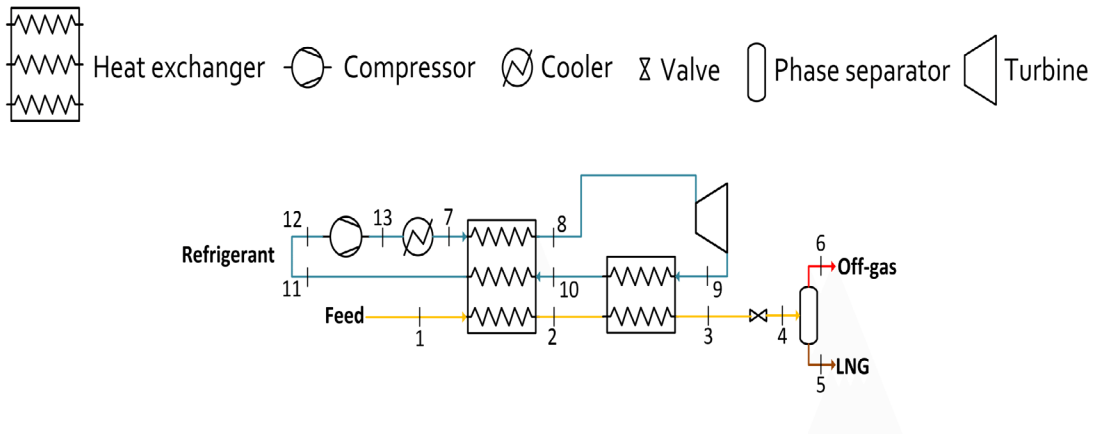
Fig. 4. Process flow diagram of the single reverse Brayton process.

The present work deals with the integration of gas liquefaction processes for offshore applications on FPSO facilities. Cascade processes are discarded because of their high complexity, as well as mixed-refrigerant processes with more than two stages. The processes that are studied are the following: the single mixed-refrigerant (SMR, Figure 2), the dual mixed-refrigerant (DMR, Figure 3), and the single- (Figure 4) and dual-expander (Figure 5) systems.
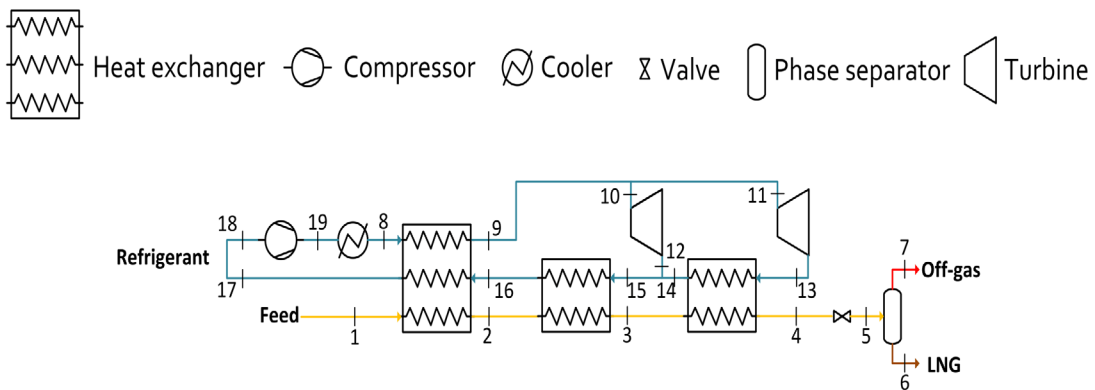

Fig. 5. Process flow diagram of the dual reverse Brayton process.

The process simulations were carried out based on the following assumptions:

- the produced LNG has a temperature of -155°C after subcooling and is delivered at atmospheric conditions;
- the recovered off-gas after subcooling and expansion is not re-liquefied and has a negligible flow rate (less than 0.1% of the feed gas flow) compared to the produced LNG;
- heat losses and pressure drops inside the heat exchangers are neglected;
- a minimum temperature difference of 3°C is required in the cryogenic heat exchangers, and it is set to 6°C for gas coolers;
- the compressors have a polytropic efficiency of 85%;
- intercooling to a temperature of 20°C within the compression process can be achieved;
- the turbines have an isentropic efficiency of 80%;
- the maximum pressure allowed in the refrigeration system is 90 bar;
- the cooling water has a temperature of 20°C.

2.2. System evaluation

The performance of a given system can be evaluated by carrying out energy and exergy accountings. An energy analysis illustrates the conversion of one form of energy (e.g. thermal, chemical, electrical) to another one. An exergy analysis is used to identify the sources, magnitudes and locations of the thermodynamic inefficiencies of the oil and gas system. The deviations from an ideal – reversible – behaviour can be divided into two categories of irreversibility:

- internal irreversibility – it is caused by internal dissipative effects within the operation units, e.g. heat transfer across a finite temperature difference in a heat exchanger, friction effects in a compressor, etc.
- external irreversibility – it is related to the interactions between the system under study and its surroundings, e.g. heat losses from a component, discharge of a given stream to the environment, etc.

The former is denominated `exergy destruction', while the latter is called `exergy losses. The aim of an exergy analysis is to quantify this irreversibility and to suggest system improvements in function. This field is well-established, and only the main equations are given as follows. The reader is referred to the textbook of e.g. Kotas [16] for a more thorough introduction.

For an open control volume in steady-state conditions, the exergy destruction is defined as the difference between the exergy input and output, and is given by, on a rate form, as:

$$\dot{E}_d = \sum \dot{E}_{in} - \sum \dot{E}_{out} = \dot{E}_W + \dot{E}_Q + \sum_k \dot{m}_k e_k$$

where:

- $\dot{E}_W$ and $\dot{E}_Q$ are the exergy rates associated with the work and heat transfers;
- $\dot{m}_k$ and e_k are the mass flowrate and corresponding specific exergy of the material stream k;

The exergy of a stream of matter, neglecting nuclear interactions, is expressed as the sum of its kinetic e^{kn}, potential e^{pt}, physical e^{ph} and chemical components e^{ch}:

$$e = e^{kn} + e^{pt} + e^{ph} + e^{ch}$$

The specific physical exergy accounts for the differences in temperature and pressure in reference to the environmental conditions (T_0, p_0) without considering changes in chemical composition, and is defined as:

$$e^{ph} = (h - h_0) - T_0(s - s_0)$$

where h and s are the enthalpy and entropy calculated at the stream conditions and h_0 and s_0 at the ambient temperature T_o and pressure p_o. The chemical exergy accounts for the differences in chemical composition with the reference environment, and is given, on a rate form, by:

$$e^{ch} = \underbrace{\sum_j x_j \bar{e}_j}_{I} + \underbrace{\left[\left(h_0 - \sum_j x_j h_{j,0} \right) - T_0 \left(s_0 - \sum_j x_j s_{j,0} \right) \right]}_{II} = \underbrace{\sum_j x_j \bar{\bar{e}}_j}_{III}$$

where the term *I* represents the chemical exergy of the pure components, with x_j the mass fraction and $\bar{e}_j$ the specific chemical exergy. The term *II* corresponds to the decrease of chemical exergy because of the mixing effects, with $h_{j,0}$ the chemical enthalpy of the pure component j at ambient conditions, and $s_{j,0}$ the corresponding entropy, on a mass basis. This term is therefore always negative, and is named in the literature `compositional exergy'. The term *III* denotes the chemical exergy of the components in the mixture, with $\bar{\bar{e}}_j$ the specific chemical exergy of the component j in the mixture. The specific potential and kinetic exergies are equal to the potential and kinetic energies, respectively, and are assumed negligible in comparison with the chemical and physical exergies.

The performance of these plants can be based on the following indicators:

- the specific power consumption per unit of liquefied natural gas for the liquefaction process, which is expressed as:

$$w = \frac{\dot{W}_{\text{liquefaction}}}{\dot{m}_{\text{LNG}}}$$

- the overall thermal conductance of the cryogenic heat exchangers (UA), which is determined by estimating the differential energy transfer ($d\dot{q}$) between the hot and cold streams within a control volume characterised by a segment length dx and a local temperature difference $T_h - T_c$.

$$UA = \frac{d\dot{q}}{(T_h - T_c)} \frac{L}{dx}$$

- the exergy destruction ratio y_k^*, which is defined as the ratio of the exergy destruction taking place in a given process k to the exergy destruction taking place in the complete system:

$$y_k^* = \frac{\dot{E}_{\text{d},k}}{\dot{E}_{\text{d}}}$$

3. Results

3.1. Gas liquefaction

A comparison of the processes under study shows that, as stated in the literature, mixed-refrigerant processes have smaller net power consumption. The specific power consumption w is in the range of 800 – 900 kJ/kg$_{\text{LNG}}$ for the dual case, while they reach 1000 kJ/kg$_{\text{LNG}}$ for the configuration with a single stage. Higher figures are found for the expansion-based systems, of 1100 – 1200 kJ/kg$_{\text{LNG}}$ for the dual set-up and 1500 – 1600 kJ/kg$_{\text{LNG}}$. These numbers can be compared against the energy content of the produced LNG, which is 53.4 MJ/kg (higher heating value) or 48.4 MJ/kg (lower heating value) at 15°C and 1.013 bar. The previous figures are thus equivalent to a power consumption of 0.22, 0.33, 0.30 and 0.41 kWh/kg$_{\text{LNG}}$, and to 1.7%, 2.5%, 2.2% and 3% of the LNG energy content (LHV basis). The cooling demand is the highest for the mixed-refrigerant processes, and it can be seen that most of the heat transfer in the cryogenic heat exchangers is associated with the cooling of the refrigerant at high pressure.
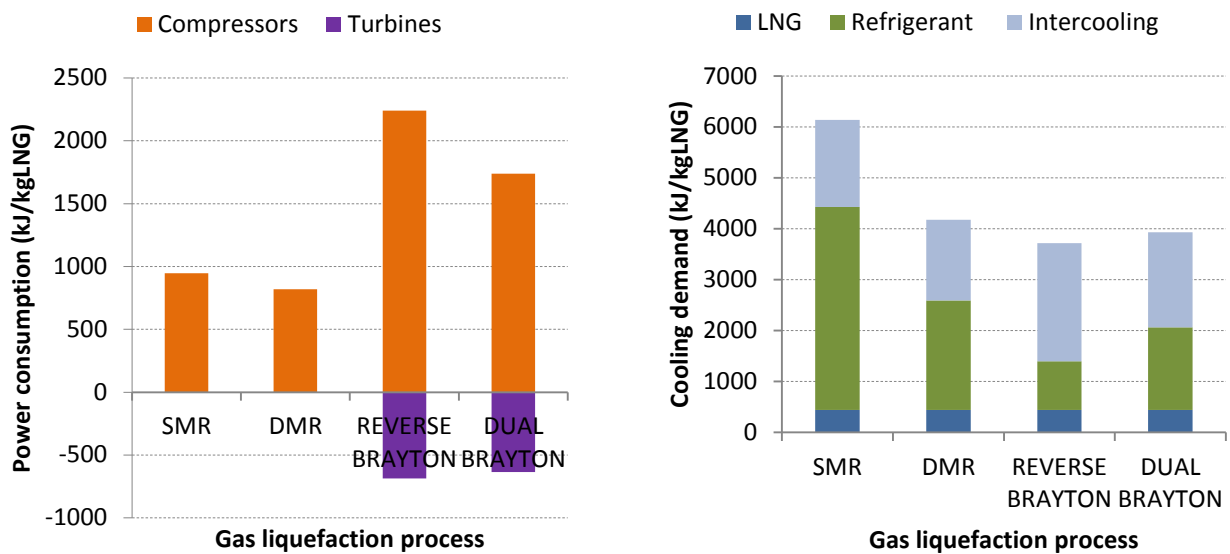

Fig. 6. *Power consumption and cooling demand sorted per process and type of component.*

A comparison of the temperature-heat profiles in the four cases illustrates significant differences in the heat transfer process in the cryogenic heat exchangers. The amount of heat transferred is as low as 1300 kW in the conventional reverse Brayton cycle, while it exceeds 4000 kW in the single mixed-refrigerant process. As expected, the temperature-heat profile of the mixed refrigerants follows closely the temperature-heat profile of the natural gas mixture because of the phase changes on both sides, while large temperature differences are found in the case of expansion-based systems. These findings are confirmed by the values of the conductance of the cryogenic heat exchangers (UA). When normalised for 1 kg of LNG, the thermal conductance is app. 80 kW/K (reverse Brayton), 210 kW/K (dual reverse Brayton), 590 kW/K (single mixed refrigerant) and 410 kW/K (dual mixed refrigerant).
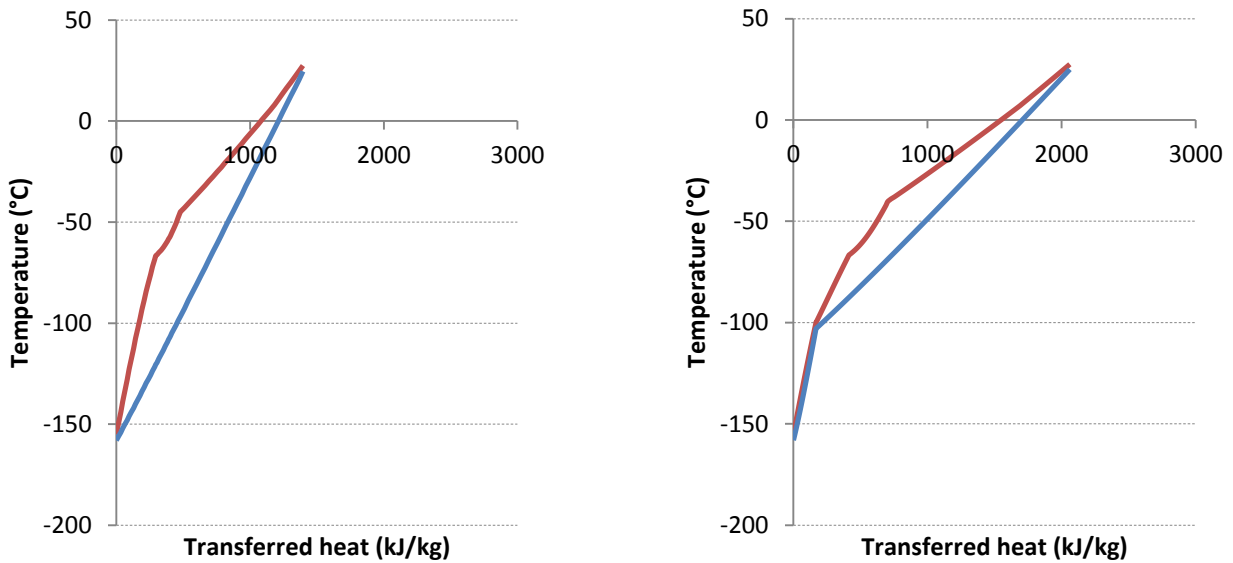

Fig. 7. *Temperature-heat profiles of the cryogenic heat exchangers for the single (left) and dual (right) reverse Brayton cycles.*
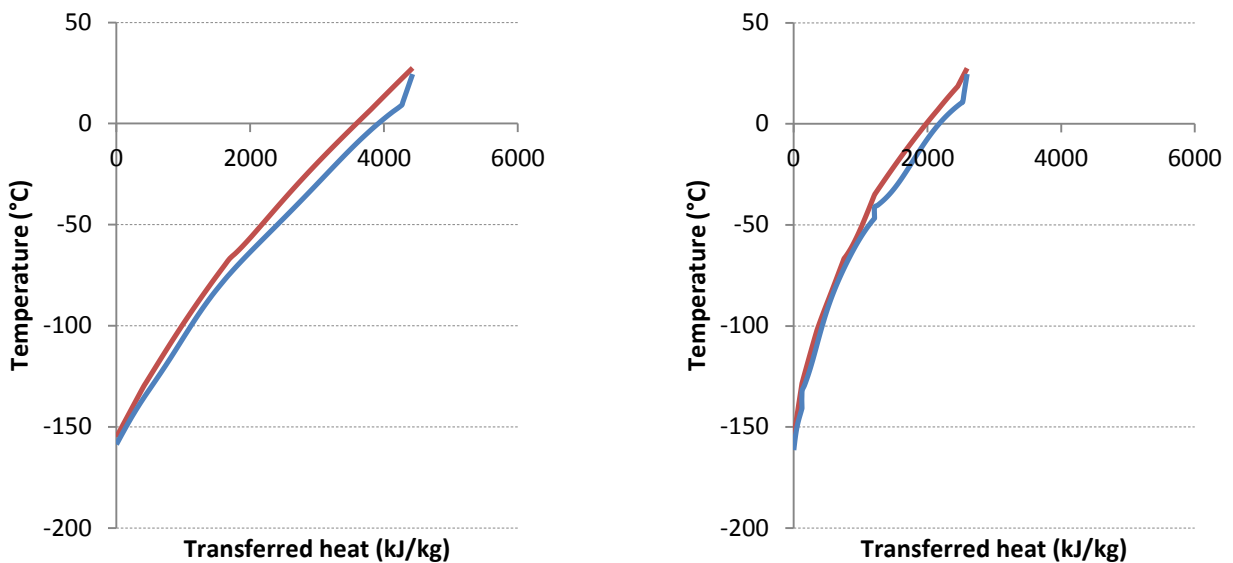

Fig. 8. *Temperature-heat profiles of the cryogenic heat exchangers for the single (left) and dual (right) mixed-refrigerant cycles.*

The exergy analysis shows that the exergy destruction is the lowest in the case of the DMR process. Most exergy is destroyed in the turbomachinery components (more than 60%) in the case of the expansion-based systems, which can be expected because of the large flowrate of nitrogen and high pressure ratios. The heat transfer process in the cryogenic heat exchangers is responsible for about 30% to 55% of the total exergy destruction in the mixed-refrigerant processes, as a result of the significant amount of heat transferred (more than 2800 kJ/kg$_{LNG}$).
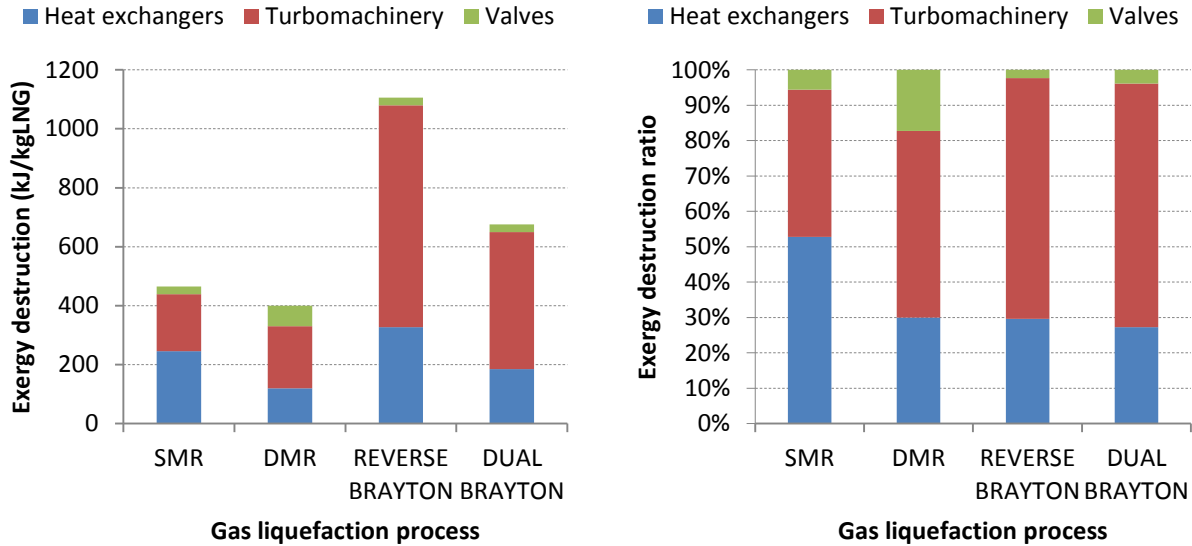

Fig. 9. Exergy destruction expressed per kg of LNG for the sole gas liquefaction process, sorted by type of components (note that the exergy destruction taking place in the intercooling of the gas compression is accounted within the turbomachinery category).

As mentioned previously, aspects such as the equipment count, weight and safety gain importance when gas liquefaction systems are to be implemented in offshore conditions. Although `simple' mixed-refrigerant processes present a lower equipment count than `basic' expansion-based ones (Table 1), the former are more complex to control in start-up and dynamic conditions. They are indeed more sensitive to the natural gas composition and rate, as the temperature profiles are closer. The actual flow rates at the first compressor may give an insight into the piping dimensions, and the dual Brayton process seems to be the most compact process. It may, however, be one of the heaviest, because it operates under high pressure and requires therefore special materials.

Table 1. Equipment count and actual flow rates per train, for each gas liquefaction process.

	SMR	DMR	Reverse Brayton	Dual Brayton
Equipment count				
• Cryogenic heat exchangers	1	3	2	3
• Gas coolers	2	2	5	10
• Compressors	2	2	4	8
• Companders	0	0	1	2
• Separator	1	2	1	1
Refrigerant flowrate (total)				
• Overall (kg/s)	225	101+76	265	390+90
• Suction first compressor (m^3/s)	37	14+13	55	12+5.5

3.2. FPSO system

3.2.1. Energy demands

An overview of the whole system shows that the highest pressures are found for the streams of gas and carbon dioxide after compression. The lowest pressure is reached at the outlet of the separation process, since the crude is depressurised to enhance separation between the oil, water and gas phases. In the baseline case, i.e. when no liquefaction process is implemented, the power consumption is divided between: (i) the separation process, for the water and oil recirculation; (ii) the vapour recovery unit or re-compression, to increase the pressure of the recovered gas to the initial feed pressure; (iii) the associated gas compression (1st stage, before treatment, and further stages, after dehydration and removal); and (iv) the carbon dioxide compression (if separated). The power demands of the gas compression steps are the major power consumers, while the power required in the separation step is negligible. Heating takes place in two processes: the separation unit, to preheat the crude at the inlet of the separation stages, and to a minor extent (less than 15% of the total heating demand), in the fuel gas system, to preheat the fuel before combustion in the gas turbine.

The integration of gas liquefaction processes results in a significant increase of the total power demand of the processing plant (Figure 10). As seen in Section 3.1, expansion-based processes are the most energy-intensive processes, but all systems have a specific power consumption of at least 800 kJ/kg$_{LNG}$. In the *first case* (maximum hydrocarbon production), all produced gas, with the exception of the fuel gas, is routed to the gas liquefaction process. The power demands are therefore 50 MW (single reverse Brayton cycle), 35.4 MW (dual reverse Brayton cycle), 30.4 MW (single mixed-refrigerant cycle) and 26.4 MW (dual mixed-refrigerant cycle).
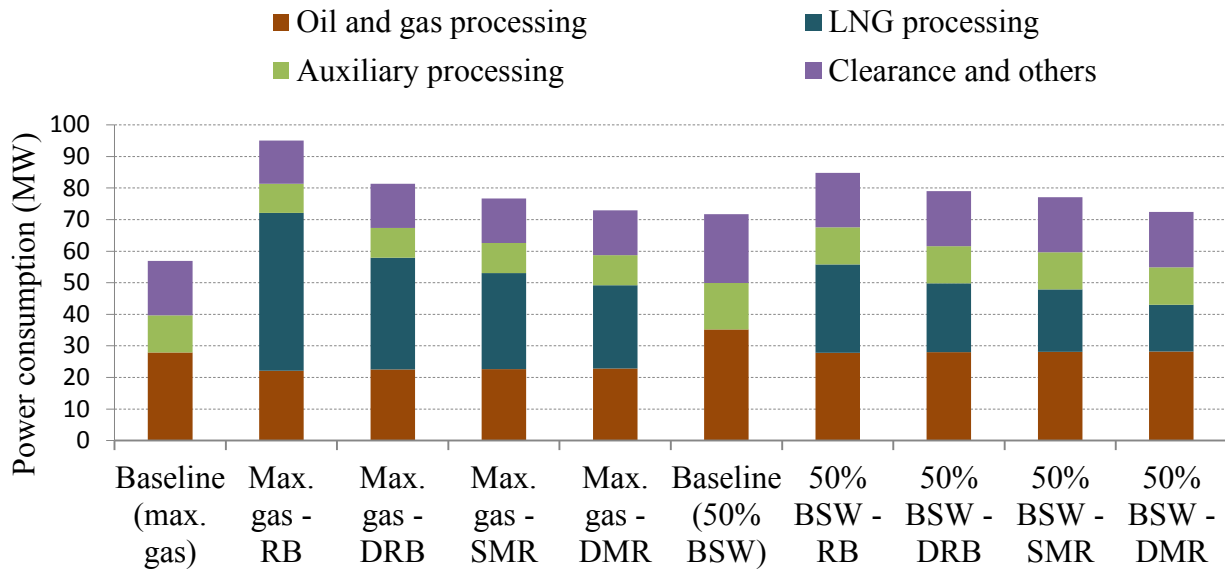

Fig.10. Power consumption sorted by case (50% export/injection or maximum gas export) and per type of gas liquefaction process (RB = Reverse Brayton, DRB = Dual Reverse Brayton, SMR = Single-Mixed Refrigerant, DMR = Dual Mixed-Refrigerant).

In the *second case* (*50% export*), the resulting power demands are, in the same order of cycle, 28 MW, 21.8 MW, 19.8 MW and 14.8 MW. Depending on the type of cycle, the gas liquefaction process represents either the first or second main power consumer, after the gas compression step. The integration of gas liquefaction does not have any impact on the power consumption of the upstream processes (e.g. separation) but on the downstream ones (e.g. second-stage compression), because less, if any, gas will be compressed up to 55 MPa.

3.2.2. Exergy accounting

The exergy analysis shows that the gas liquefaction process is, in all cases, the *greatest exergy-destroying process*, within the processing plant. It amounts, in the maximum gas production situation, to a maximum of 36 MW with the reverse Brayton cycle, and to a minimum of 13 MW with the dual mixed-refrigerant process. The same figures are of 20 MW and 7.2 MW in the 50% export case. These findings illustrate the low exergetic efficiency of these cryogenic processes (30-45%) compared to the gas compression systems (50-60%), and the difficulties of increasing the temperature-based exergy at low temperature levels (Figure 11).

However, put in perspective (Figure 12), the gas liquefaction process represents only between 4% and 12% of the exergy destruction taking place on the whole platform, which shows that the irreversibility associated with the combustion, heat transfer and transmission losses in the power generation system are dominant. This irreversibility is mostly unavoidable and cannot be reduced unless significant system designs (e.g. electrification) occur.
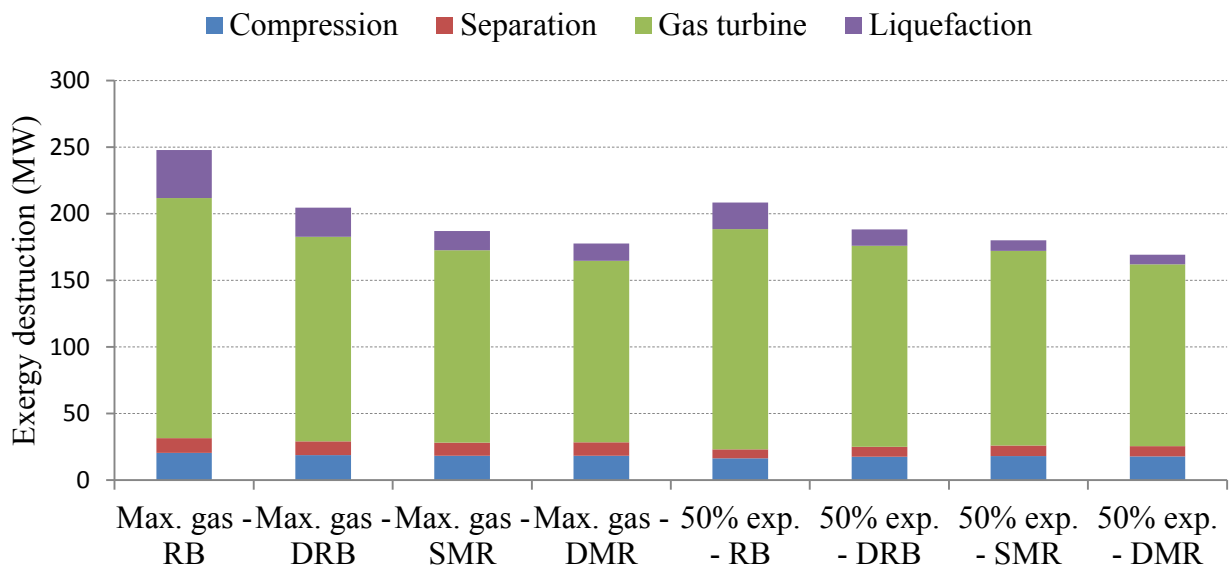

Fig.11. Exergy destruction sorted by case (50% export/injection or maximum gas export) and per type of gas liquefaction process (RB = Reverse Brayton, DRB = Dual Reverse Brayton, SMR = Single-Mixed Refrigerant, DMR = Dual Mixed-Refrigerant).
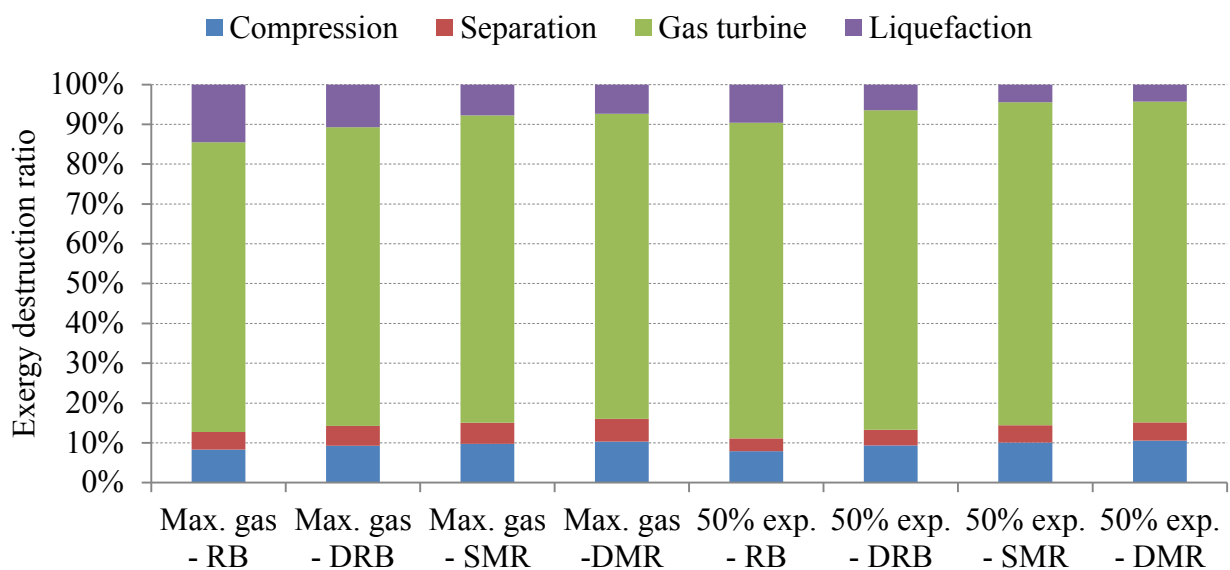

Fig.12. Share of exergy destruction sorted by case (50% export/injection or maximum gas export) and per type of gas liquefaction process (RB = Reverse Brayton, DRB = Dual Reverse Brayton, SMR = Single-Mixed Refrigerant, DMR = Dual Mixed-Refrigerant).

4. Discussion

Comparison with the scientific literature

A direct comparison with the literature on the numbers mentioned previously is made difficult because of the different assumptions (temperature, pressure, etc.). Remeljej and Hoadley [8] compared the thermodynamic performance of the SMR process with other expansion systems, which are not presented in the present work. They find that the SMR process is preferable against expansion-based processes, as it presents the smallest specific power requirement. Their analysis shows the same findings as in the present work. The exergy destruction in the valves is negligible, and that the compression and aftercooling processes represent the highest share of irreversibility.

A comparison with the work of Kusmaya shows similar trends for the four investigated processes, and similar power consumption was found for the four processes. The figures obtained in the present work are lower by about 5%, and this difference can be explained by the discrepancies in the compressor and turbine efficiencies.

Future work

Future work within this topic will include performance comparisons for a higher number offshore liquefaction processes, such as the expansion-based processes with CO_2-precooling, or with dual cycles with nitrogen and methane. All the investigated cycles should then be optimised with respect to the power consumption, required space and capital costs, in order to propose solutions that are technically and economically acceptable. An economic analysis coupled with a multi-objective optimisation routine may be a next step, as it would show clearly the trade-offs between (i) the additional costs of the liquefaction cycle against the gains in sales export, (ii) the overall performance and power consumption against the weight and footprint of the facility.

A more thorough analysis of the combined FPSO-LNG system may as well be performed, by investing opportunities for process integration and simulating system improvements. The use of process integration tools aims at identifying the heating and cooling needs, and would help promoting energy savings. As the separation and compression sections interact with the LNG subsystem, any improvement would affect the gas composition and flow rate at the inlet of the liquefaction process.

Finally, caution should be exercised when extending or extrapolating these results – only two configurations and feed compositions, with moderate carbon dioxide contents, were considered. Additional simulations should then be performed to assess whether the same trends are found for other petroleum properties, and to optimise the resulting system if needed.

5. Conclusion

The integration of gas liquefaction processes on floating, production, storage and offloading facilities was analysed. Four cycles were assessed with regards to their power consumption and thermodynamic performance. Although the single- and dual-mixed refrigerant processes are the most efficient, they are as well the largest ones, which may create challenges when implementing them in an offshore environment. Adding these cryogenic processes on a FPSO is feasible after gas treatment and may be interesting when the field produces large quantities of hydrocarbons (start- and peak periods). However, liquefying gas seems to be more energy-intensive and exergy-consuming than injecting or exporting gas. The total power consumption increases by up to 30%, while the exergy destruction within the processing plant increases by up to 100%. With regards to these findings, and considering the costs of installing a LNG plant, it is essential to conduct a careful analysis of the trade-off between the capital costs and operating revenues for such options.

Acknowledgments

Nomenclature

A heat transfer area, m^2

$\dot{E}$ exergy flow rate, W

U overall heat transfer coefficient, $W/(m^2\,K)$

T temperature, °C or K

$\dot{W}$ power consumption, W

$d\dot{q}$ differential heat transfer

dx differential segment length

e specific exergy, J/kg

h specific enthalpy, J/kg

$\dot{m}$ mass flow rate, kg/s

s specific entropy, J/kgK

w specific power consumption, J/kg

y^* exergy destruction ratio

Abbreviations

BSW   basic sediment and water

DMR   dual mixed refrigerant

FPSO   floating, production, storage and offloading

LNG   liquefied natural gas

SMR   single mixed-refrigerant

Subscripts

c   cold stream

d   destruction

h   hot stream

k k-th component or stream

in   input

out   output

0   reference state

Superscripts

ch   chemical

kn   kinetic

ph   physical

pt   potential

Q heat transfer

W work transfer

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