THERMODYNAMIC EVALUATION OF FLOATING PRODUCTION STORAGE AND OFFLOADING FACILITIES WITH LIQUEFACTION PROCESSES

Tuong-Van Nguyen, tungu@mek.dtu.dk[1,2]
Yamid Alberto Carranza-Sánchez, yamidcarranza@usp.br[1,3]
Silvio de Oliveira Júnior, soj@usp.br[1]

[1]Laboratory of Environmental and Thermal Engineering, Polytechnic School, University of São Paulo, 05508-900 São Paulo, Brazil
[2]Department of Mechanical Engineering, Technical University of Denmark, Nils Koppels Allé, 2800 Kongens Lyngby, Denmark
[3]School of Mechanical Technology, Technological University of Pereira, La Julita, 660003 Pereira, Colombia

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 four possible petroleum compositions, which differ in their contents of carbon dioxide, light and heavy hydrocarbons. 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. However, the produced gas can be exported and sold, which results in higher economic benefits than in the conventional situation with gas injection.*
Keywords: *FPSO, Gas liquefaction, Energy, Exergy*

1. INTRODUCTION

Floating production, storage and offloading (FPSO) facilities have attracted more attention in Brazil, especially in the Campos Basin. On the opposite of fixed platforms, FPSOs are ship-shaped and floating vessels that can be used for the production of petroleum from deep- to ultra-deep fields, with water depths exceeding 1500 m. They can be designed to receive and process the oil and gas produced on-site or from nearby fields, and the hydrocarbons can be stored and offloaded into tankers, or transported through pipelines. The latter is less likely, as one advantage of FPSOs compared to conventional facilities is that FPSOs can be operated in remote locations, where no local infrastructures are readily available or installed.

The processes installed on FPSOs do not fundamentally differ from fixed platforms (Bothamley (2004)) and include basic operations such as water and hydrocarbon separation, oil pumping, gas compression and treatment. The energy demands in terms of heating and power are similar and can range from a few to several hundreds MW, depending on the field conditions (e.g. temperature and pressure), feed composition (e.g. water and impurities' content) as well as operating strategy (e.g. oil offloading, gas injection, gas lift).

FPSOs may be operated in remote locations, meaning that there may not be any infrastructure to export the produced gas. In such cases, gas may be injected into the exploited field, resulting in a loss of hydrocarbons. Offshore liquefaction of natural gas creates the possibility to avoid flaring or reinjection, increasing meanwhile the monetary value of small- and medium-scale fields. However, the requirements for offshore applications are different from onshore facilities (Barclay and Denton (2005); Bukowski (2013); Kusmaya, Maya (2014)) - for example, it is desirable to avoid too large and complex processes (space and volume limitations) and that are sensitive to motion (maldistribution of the working fluid). While cascade (Andress (1996)) and propane-precooled mixed-refrigerant (Shah *et al.* (2009)) processes are preferred onshore because of their high performance, they may not be suitable for offshore implementation. Expander-based processes (He and Ju (2014)), on the other hand, use nitrogen in gas form, which makes these systems more relevant because of the inherent cycle safety.

The literature discusses the performance of small-scale liquefaction processes: for example, Cao *et al.* (2006) assess the performance of two types of processes and claim that expansion-based systems may be more efficient than mixture-based ones, without precooling. The opposite conclusion is drawn by Remeljej and Hoadley (2006), as well as by Chang (2015), who assess the maximum theoretical efficiency. These works investigate the efficiency of LNG processes but do not discuss their integration within the offshore facility, and how the performance of the combined LNG-FPSO plant is affected.

An energy analysis indicates the changes from one form of energy to another, which allows the tracing of energy flows throughout a given process (Szargut *et al.* (1988)). Unlike energy, exergy is destroyed in real processes because of irreversible phenomena such as heat transfer across finite temperature differences, pressure drops, and chemical reactions. An exergy analysis shows the locations and extents of these thermodynamic imperfections, and indicates therefore the additional fuel use required to compensate these imperfections.

A few works on the thermodynamic performance of the offshore processing (without the inclusion of the power generation facility) can be found in the literature, such as the studies of Oliveira Jr. and Van Hombeeck (1997) on a Brazilian plant, as well as Voldsund *et al.* (2014) and Nguyen *et al.* (2014b) on North Sea facilities. These show that the gas compression process is responsible for the greatest share of the plant power consumption and exergy destruction. Efforts should therefore focus on improving this section of the offshore facility. Other works, such as the studies of Nguyen *et al.* (2014a) and Sánchez *et al.* (2015) assess the performance of the entire offshore plant. The latter deals with the analysis of the performance of a FPSO plant, and their results show that the gas turbines are responsible for most exergy destruction. It is nevertheless underlined that these irreversibilites are unavoidable, because of the non-ideal behaviour of the combustion reactions. More recently, the work of Nguyen et al. (2014) deals with the comparison and integration of LNG processes on FPSOs where gas export is a possibility. The performance of the combined FPSO-LNG process is evaluated based on thermodynamic criteria. The findings show that the gas liquefaction process requires a greater power consumption than the gas injection one, leading to a greater power consumption. The additional exergy destruction taking place in the cryogenic system is actually as high as the exergy destruction within the remaining processes of the offshore plant, for all the modelled refrigeration cycles. This work was only conducted for one petroleum composition, and the present work aims at addressing these gaps. The first purpose is to investigate whether the same conclusions can be drawn for platforms processing a petroleum feed with high carbon dioxide content. The second one is to evaluate more in details the sources and locations of performance losses, per component and sub-system.

The present paper is divided as follows. Section 2 consists of a description of a typical FPSO system and presents the gas liquefaction processes investigated in this work, along with the modelling assumptions and analysis methods. Section 3 goes through the results of the process and thermodynamic analysis, and draws suggestions for improving FPSO-LNG plants, when relevant. Section 4 concludes the present study and gives hints for future work in that field.

2. METHODS

2.1 System description

2.1.1 Floating, production, storage and offloading facilities

Oil and gas wells produce mixtures of oil, condensate, and gas (petroleum) extracted together with water, and possibly other gases such as carbon dioxide and impurities such as salt and sand. The purpose of offshore processing is to recover hydrocarbons for further export, treatment and sales: the oil is treated to meet the transport specifications (e.g. salt content and vapour pressure), while the associated gas is either exported or injected back into the field. If significant amounts of carbon dioxide are extracted, which is the case for numerous Brazilian fields, additional compression trains may be installed to support field injection and geological storage. The produced water and impurities are removed and discharged into the environment. The main operations of offshore processing (Fig. 1) are therefore phase separation, temperature and pressure changes (heat exchange, compression, expansion). The power and heat required to drive these operations are produced by consuming a fraction of the extracted gas in gas turbines.

The decision of treating associated gas depends on the associated gas composition and therefore on the initial petroleum properties. Purifying from carbon dioxide, most often with membranes, is required if gas should be exported to prevent corrosion issues in pipelines (transport in gaseous form) or heat exchanger plugging (transport in liquid form). Low contents of carbon dioxide in the fuel gas are also preferable to avoid the operation of the gas turbines far from their design conditions. In general, gas export takes place, if possible, in the peak production period of oil and gas, while gas injection is more common when the hydrocarbon extraction is declining.

The process simulations were carried out with Aspen Plus version 7.2 using the Peng-Robinson equation of state, with the exception of the carbon dioxide treatment system that is modelled using the predictive Soave-Redlich-Kwong model. *Two petroleum compositions* were considered, one with a low (less than 1 % on a molar basis) and one with a high (more than 25 %) content of carbon dioxide. The following assumptions are taken:

- a crude production of 150,000 barrels per day is assumed;
- for the first composition, 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 less than 1 % of carbon dioxide;
- for the second composition, the well-fluid consists of app. 26 % of carbon dioxide, 41 % of methane, 10 % of medium-weight hydrocarbons, and the remaining is heavy-weight hydrocarbons (C5+);
- dilution water is not considered;
- processes such as oil desalting, gas dehydration with glycol, and carbon dioxide separation with membranes are not investigated in details - ideal separation is assumed in these cases;
- an adiabatic efficiency of 75 % is assumed for all centrifugal compressors present in the compression and recompression trains;

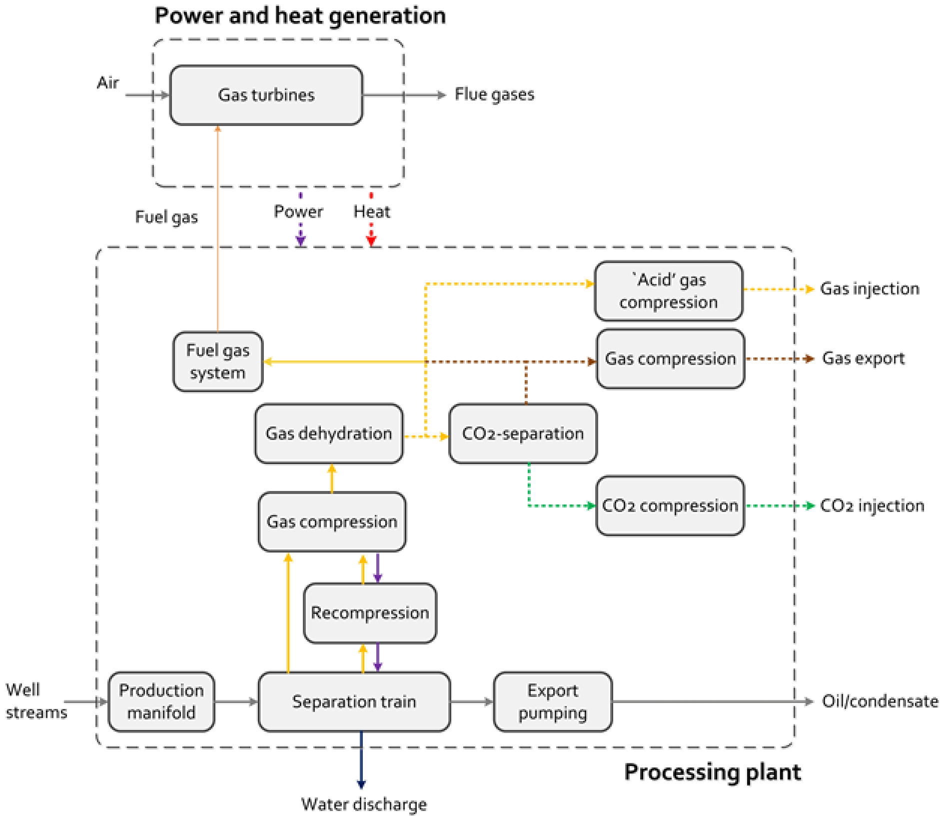

Figure 1: General system overview of an oil and gas platform. Arrows represent one to several streams while block represent different subsystems. Solid lines indicate that the corresponding stream or process is present for all the studied platforms and can generally be found on all typical oil and gas facilities, while dotted ones denote flows that exist only in specific operating modes or facilities. Grey denotes liquid hydrocarbons (oil), dark blue produced water, orange untreated gas, green carbon dioxide, and brown treated gas.

- the auxiliary power demand (e.g. secondary utilities such as HVAC) is not considered in the calculations;
- 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

Gas liquefaction is a well-known technology, with clean natural gas being cooled at high pressure to temperatures as low as -160 °C. Three main types of refrigeration cycles are used for this purpose, namely the cascade systems, the mixed-refrigerant processes and the expander-based cycles. They differ in the selection of the working fluid (e.g. pure fluid or zeotropic mixture), temperature and pressure levels (e.g. pressures above 50 bar expected for nitrogen expander cycles), component inventory (e.g. turbines or expansion valves), performance and field of application. The processes investigated in this work are the single mixed-refrigerant (SMR) and single-expander systems. They are considered suitable for offshore applications, because they present a smaller equipment count and higher compactness than state-of-the-art liquefaction systems such as the cascade process.

The SMR (Fig. 2) process belongs to the category of mixed-refrigerant processes, in which the working fluid consists of a mixture of hydrocarbons (e.g. methane and ethane) and nitrogen. The cooling effect is generated by an adiabatic expansion through a valve device (Joule-Thomson effect). The refrigerant undergoes phase change along a temperature glide, implying that a better match of the hot and cold temperature profiles can be reached. This results in a greater system performance of mixed-refrigerant systems compared to expander-based ones.

The single (Fig. 3) expander system is based on gaseous nitrogen as working fluid. The refrigerant does not undergo phase change, and the cooling effect is generated by expanding the working fluid through a turbo-expander operating over a large pressure ratio. Lower performances than mixed-refrigerant processes are expected, but the working fluid is constantly on a gaseous phase, which avoids flow maldistribution and instabilities in the cryogenic heat exchangers.

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 1.7 bar;
- the recovered off-gas after subcooling and expansion (boil-off gas) is not re-liquefied but used directly in the gas turbines, together with the fuel gas, after compression;

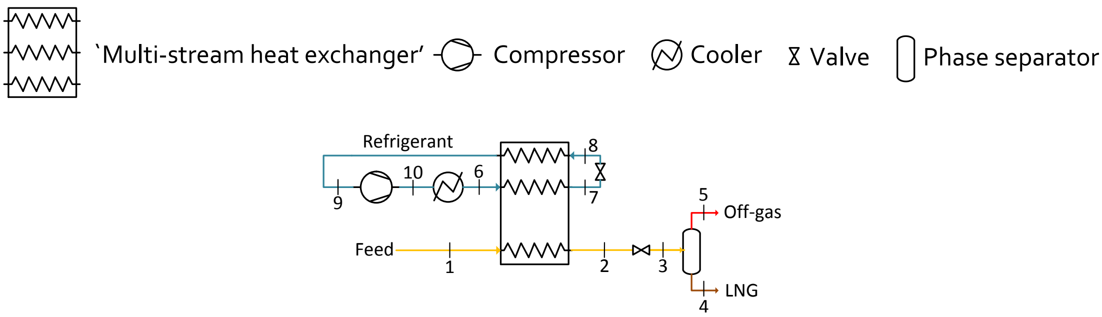

Figure 2: Process flow diagram of the single-mixed refrigerant system (SMR).

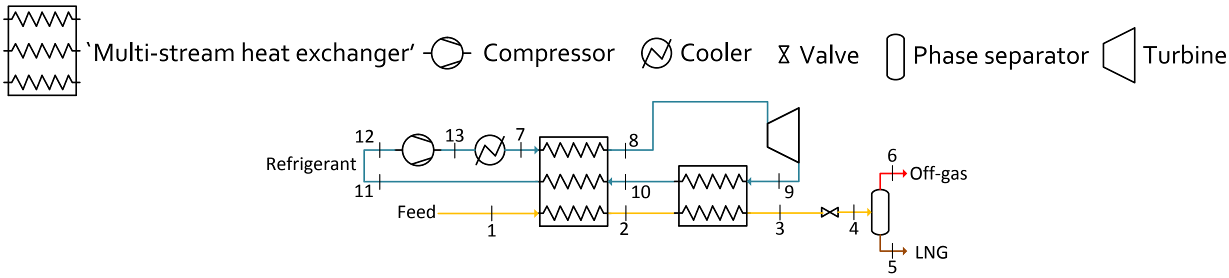

Figure 3: Process flow diagram of the expansion-based process.

- a minimum temperature difference of 3 °C is assumed for the cryogenic heat exchangers;
- the compressors have a polytropic efficiency of 85 %;
- inter- and after-cooling to a temperature of 40 °C can be achieved;
- the turbines have an isentropic efficiency of 80 %;
- for the single mixed-refrigerant process, the refrigerant composition is taken to 11.7 % nitrogen, 28.4 % ethane, 14 % propane, 5.7 % n-butane and 9.5 % isopentane (molar basis);
- the pressure levels are assumed to be 4.2 and 24.4 bar;
- for the single expander process, the pressure levels of nitrogen are 8 and 106.5 bar, and the compression takes place over two stages;
- the precooling temperature of nitrogen is taken to -45 °C;

2.2 Thermodynamic analysis

2.2.1 Energy analysis

Energy can be stored, transformed from one form to another (e.g. heat to power) and transferred between systems, but can neither be created nor destroyed (conservation law). In open systems, it can be transferred by streams of matter, heat and work. Neglecting the changes in kinetic and potential energies, which are negligible, in our case, against the chemical energy of hydrocarbons, the energy rate balance at steady-state is expressed as:

$$0 = \sum_k \dot{Q}_k - \dot{W} + \sum \dot{H}_{\text{in}} - \sum \dot{H}_{\text{out}} \tag{1}$$

where $\dot{Q}_k$ and $\dot{W}$ are the time rates of energy transfer by heat and work ($\dot{Q} \geq 0$ indicates heat transfer *to* the system and $\dot{W} \geq 0$ work done *by* the system) and $\dot{H}$ is the enthalpy flow of material streams. The subscripts in and out denote the inlet and outlet of the system of study and k the boundary of the component of interest. The energy balance for the processing plant of the oil and gas facility can be expressed as:

$$\dot{H}_{\text{feed}} + \dot{Q}_{\text{heat}} + \dot{W}_{\text{UT}} = \sum_k \dot{H}_k + \dot{Q}_{\text{cool}} \tag{2}$$

The left-hand side term represents, on a time rate basis, the energy associated with the feed entering the processing plant $\dot{H}_{\text{feed}}$ and the energy transfers with the heating medium $\dot{Q}_{\text{heat}}$ and power $\dot{W}_{\text{UT}}$ from the utility plant that are consumed within the separation and treatment modules. The right-hand side term denotes the energy of the outlet streams of the processing plant (e.g. oil, gas, carbon dioxide) and $\sum_k \dot{H}_k$ and the energy discharged with the cooling medium flow that is discharged to the sea $\dot{Q}_{\text{cool}}$.

2.2.2 Exergy analysis

Exergy may be defined as the 'maximum theoretical useful work (shaft work or electrical work) as the system is brought into complete thermodynamic equilibrium with the thermodynamic environment while the system interacts with it only'. Unlike energy, exergy is not conserved but some is destroyed because of the irreversible phenomena taking place in real processes (e.g. chemical reactions, heat transfer):

$$\dot{E}_d = \sum \dot{E}_{\text{in}} - \sum \dot{E}_{\text{out}} = \sum_k \left(1 - \frac{T_0}{T_k}\right) \dot{Q}_k - \dot{W} + \sum \dot{m}_{\text{in}} e_{\text{in}} - \sum \dot{m}_{\text{out}} e_{\text{out}} \tag{3}$$

where $\dot{E}_d$, $\dot{E}_{\text{in}}$ and $\dot{E}_{\text{out}}$ are the destroyed, inflowing and outflowing exergy rates. e is the specific exergy of material stream, T_0 and T_k are the environmental and instantaneous temperatures. The exergy destruction rate can also be calculated from the Gouy-Stodola theorem, which is, on a time rate form, expressed as (Bejan, 2006):

$$\dot{E}_d = T_0 \dot{S}_{\text{gen}} \tag{4}$$

where $\dot{S}_{\text{gen}}$ is the entropy generation rate.

In the absence of nuclear, magnetic and electrical interactions, and neglecting as well the kinetic and potential contributions, the exergy of a material stream is a function of its physical e^{ph}, chemical e^{ch} components, as given by:

$$e = e^{\text{ph}} + e^{\text{ch}} \tag{5}$$

Physical exergy accounts for temperature and pressure differences from the environmental state, while the chemical exergy accounts for deviations in chemical composition from reference substances present in the environment. Exergy transferred with work is equal to its energy content, while it depends on the system boundary and environmental temperatures when associated with heat.

The exergy balance for the processing plant of the oil and gas facility can be expressed as:

$$\dot{E}_{\text{feed}} + \dot{E}^{Q}_{\text{heat}} + \dot{E}^{W}_{\text{UT}} = \sum_k \dot{E}_k + \dot{E}^{Q}_{\text{cool}} + \dot{E}_{d,\text{PP}} \tag{6}$$

The left-hand side term represents, on a time rate basis, the exergy associated with the feed entering the processing plant (i.e. reservoir fluid) $\dot{E}_{\text{feed}}$ and the exergy transfers with the heating medium $\dot{E}^{Q}_{\text{heat}}$ and power $\dot{E}^{W}_{\text{UT}}$ from the utility plant that are consumed within the separation and treatment modules. The right-hand side term denotes the exergy of the outlet streams of the processing plant (i.e. oil, gas, condensate, flared gas, fuel gas, produced water) $\sum_k \dot{E}_k$ and the exergy gain discarded with the cooling medium discharged to the environment $\dot{E}^{Q}_{\text{cool}}$. $\dot{E}_{d,\text{PP}}$ is the exergy destroyed in the processing plant. In this work, the dead state is taken to be 25 °C and 1 atmosphere, and the reference environment of Szargut is selected for the chemical exergy calculations.

3. RESULTS AND DISCUSSION

3.1 Gas liquefaction process

A comparison of the two processes under study suggests that, as stated in the literature, mixed-refrigerant processes can achieve smaller net power consumptions because of the improved match of the temperature profiles within the cryogenic heat exchangers (Fig. 4). The specific power consumption of the SMR process is as low as 1500 kJ/kg, while it is in the range of 2500-3000 kJ/kg for the expander-based process. These numbers correspond to an expense of respectively 4 and 6 % of the total energy content of the gas to liquefy, on a higher heating value basis. These figures are similar to the numbers given in the literature, where a power consumption of about 4 to 15 % of the LNG energy content is expected. The SMR process presents a higher performance than the reverse Brayton cycle (single expander process). This results from a better temperature match between the hot and cold sides in the cryogenic heat exchanger, which in turn leads to a smaller power consumption. The trends are similar for both compositions of the incoming feed (CO_2-rich and CO_2-lean) - this is expected as the gas to liquefy has a low carbon dioxide content because of the integration of the membrane separation system.

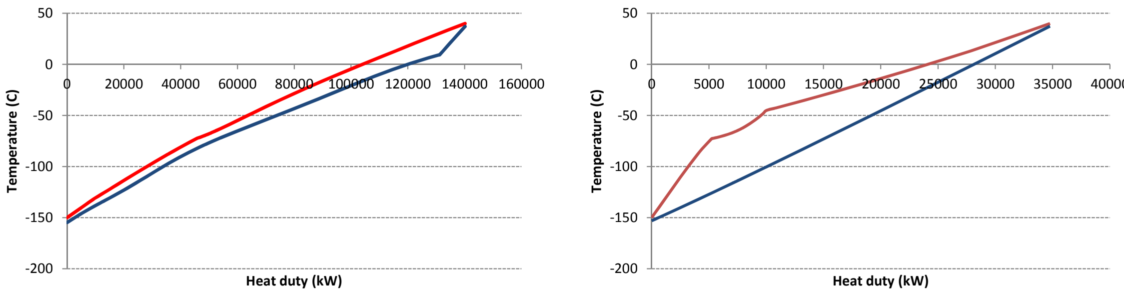

Figure 4: Temperature profiles in the cryogenic heat exchangers, in the single mixed refrigerant process (left) and reverse Brayton cycle (right).

The cooling demand for the mixed-refrigerant process is significantly higher than for the reverse Brayton cycle by a factor 1.5. The greatest share is found for the cooling of the high-pressure refrigerant in the cryogenic heat exchanger, while the cooling of the natural gas on itself represents only a small share, lower than 10 %. On the contrary, for the expander-based cycle, the highest cooling demand is associated with the inter- and aftercooling processes after the nitrogen compressors. A more detailed investigation of the heat transfer within the cryogenic heat exchangers shows that the amount of heat transferred is higher for the mixed-refrigerant process, exceeding 4000 kJ per kg of LNG, while it does not exceed 1500 kJ per kg of LNG for the reverse Brayton cycle.

The space required for the liquefaction process can be partly assessed by the conductance of the heat exchanger network (UA). When normalised to 1 kg of LNG, the thermal conductance of the reverse Brayton cycle is about 80 kW/K, while it reaches 600 kW/K for the single mixed-refrigerant process. These findings illustrate that the temperature profiles when employing mixtures are much closer. This indicates smaller driving forces in the heat transfer process, and therefore involve the need for larger heat exchangers.

3.2 FPSO system

In the baseline case, i.e. without any liquefaction process, the power consumption is divided between (i) the vapour recovery unit, where the gas recovered from the separation process is recompressed to the initial feed pressure; (ii) the gas compression section, where the gas recovered after the separation system is compressed further to the desired pressure at the inlet of the treament process; (iii) the gas export process, to reach the pressure required for export to the shore, (iv) the carbon dioxide compression and injection, where the highest system pressures are reached, with an upper limit of 55 bar. The latter exists obviously only for platforms processing feeds with high CO_2-contents. The power demands of the other sub-systems, for e.g. pumping the oil to a higher pressure for export, recirculating the oil or injecting water into the reservoir are negligible in comparison.

For the case with the highest CO_2-content (about 26 %), the CO_2-compression and injection process represents about 25 % of the total power consumption, followed by the produced gas compression section and the gas export system. The other demands, associated with the vapour recovery unit or the oil and water pumping, are negligible in comparison. For the case with the lowest CO_2-content, the demand of the CO_2-compression process is negligible, the gas compression and export systems representing the lion's share of the total power consumption. It is worth noticing that these figures are highly dependent on the operation mode of the FPSO, whether part or all the produced gas is to be injected, and whether it is treated in a CO_2-membrane system.

The heating demand is significant in the separation process to enhance the separation between the oil, gas and water phases. In particular, if the feed has a low extraction temperature or is particularly viscous, preheating to about 90 °C is essential. It ensures that the light hydrocarbons such as methane, ethane and propane are removed from the crude oil and that the desired vapour pressure is achieved. Other heating demands, for example for preheating the fuel gas at the inlet of the gas turbines, are negligible in comparison, and are generally ensured by electric heating. The cooling demand is mainly associated with the gas cooling operations prior to the gas compressors, to remove medium-weight hydrocarbons as liquid droplets in the scrubbers, and to lower the power demand of the compression operations.

In all cases, whether the feed petroleum has a high or low content of carbon dioxide, the integration of the liquefaction plant results in a significantly greater power consumption (nearly twice as much as the initial value, without liquefaction) of the processing plant. This illustrates that the benefits of compressing the associated gas to a lower pressure, of only 50 bar instead of 250 bar, are largely outweighed by the power demand of the compressors of the refrigeration cycles.

The integration of the gas liquefaction plant does not affect sensibly the power consumption and exergy destruction of the separation, oil pumping, produced water, recompression and gas compression sections. This process is indeed integrated downstream, and there are no direct interactions between these subsystems. However, as the power consumption of this process is greater than the gas export process, the demand for fuel gas is higher, and the amount of gas to process decreases. These trends are valid for both cases, which suggests that these results can be generalised to all types of FPSOs.

The exergy assessment shows that the power generation plant is responsible for the highest exergy destruction, being responsible for about 60 % of the total exergy destruction on-site. However, most is actually associated with the combustion processes taking place in the gas turbine chamber. This is on itself a highly non-ideal process that can only be improved marginally, by e.g. adjusting the air-to-fuel ratio. Focusing on the processing plant (Fig. 5), it can be seen that the most exergy-destroying processes is the CO_2-compression and injection process, if the feed has a high CO_2-content, and the export gas system if the feed has a low CO_2-content.

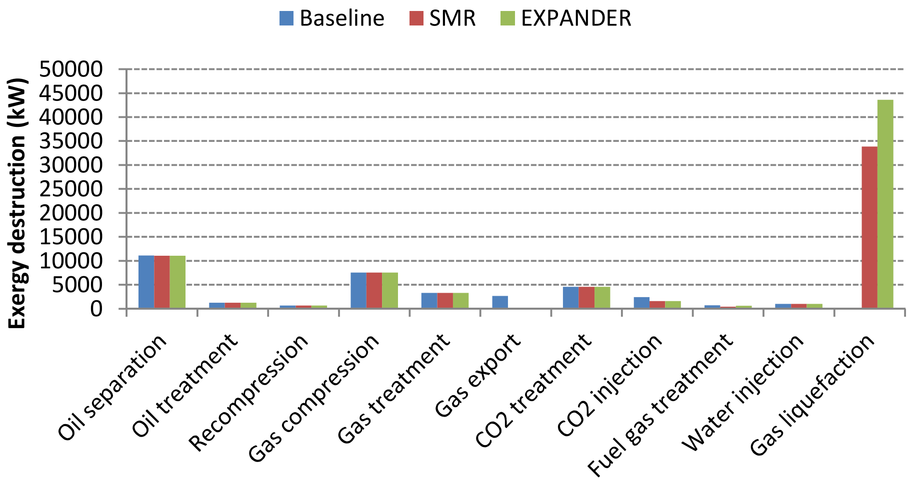

Figure 5: Exergy destruction in the *processing plant*, sorted by process, for the petroleum case with the high CO_2-content.

A more thorough investigation of the entire platform (Fig. 6), splitting the exergy destruction share per type of component, shows as well that the combustion process is responsible for most irreversibilities on-site. It is followed by the compressors, turbines, coolers and heaters, which suggests that efforts should focus on improving the performance of the turbomachinery, and possibly on a better use of the waste heat from the exhaust gases. If possible, it may be promising to recover and convert low-grade heat, as the demand for cooling is increased significantly when a gas liquefaction process is implemented. This results from the intercooling and aftercooling processes since the refrigerant is compressed over large pressure ratios.

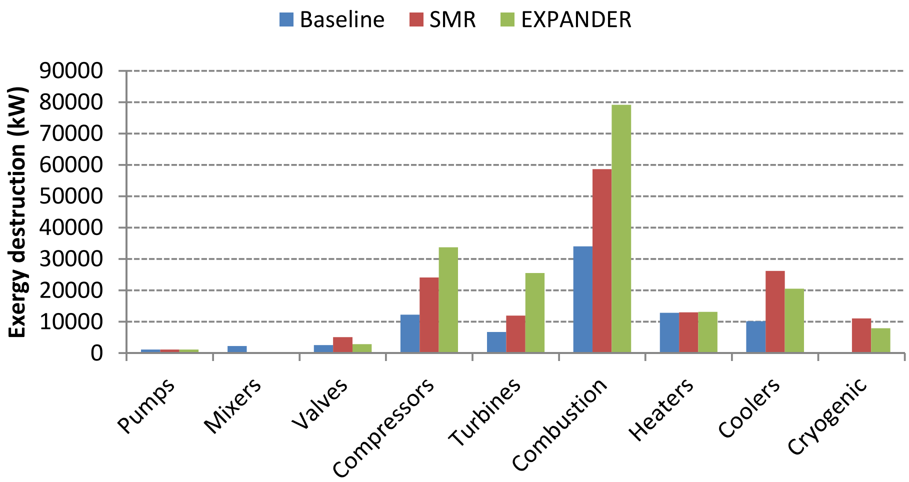

Figure 6: Exergy destruction in the *entire plant*, sorted by type of component, for the petroleum case with the high CO_2-content.

4. CONCLUSION

The integration of two gas liquefaction processes on floating, production, storage and offloading facilities was analysed and evaluated by means of process simulation and thermodynamic assessments. The combination of the offshore processing of oil and gas together with the liquefaction processes was evaluated for two feed compositions, a CO_2-rich (app. 26 %-molar) and CO_2-lean (app. 0.8 %-molar), considering the case that the oil and gas production is at its maximum. In all cases, the integration of gas liquefaction systems results in greater power consumption, cooling demand and exergy destruction on-site. This results from the large irreversibilities taking place in the compression process and cryogenic heat exchange, and those may be reduced by investing into more-efficient turbomachineries or by utilising the low-grade heat. Despite these drawbacks, implementing a LNG process allows for valorising and selling gas that would otherwise be injected into the reservoir. A cautious techno-economic analysis, carried on for each offshore site, could complement the present results and indicates whether it is feasible and desirable to integrate liquefaction systems on floating plants.

5. ACKNOWLEDGEMENTS

The present work was supported by the National Agency of Petroleum, Natural Gas and Biofuels (ANP) and BG E&P Brasil Ltda, through Investment in Research, Development and Innovation Clause, contained in contracts for Exploration, Development and Production of Oil and Natural Gas. The authors acknowledge the financial support from the Fundação de Amparo à Pesquisa do Estado de São Paulo (São Paulo research foundation, FAPESP), through the grant 2015/09157-1, from the Technological University of Pereira and COLCIENCIA, and from the Conselho Nacional de Desenvolvimento Científico e Tecnológico (Brazilian National Council for Scientific and Technological Development, CNPq), through the grant 306033/2012-7.

6. REFERENCES

Andress, D., 1996. *The Phillips Optimized Cascade LNG Process: A Quarter-of-a-Century of Improvements*. American Institute of Chemical Engineers.

Barclay, M. and Denton, N., 2005. “Selecting offshore lng processes”. *LNG journal*, Vol. 10, No. 1, pp. 34–36.

Bothamley, M., 2004. “Offshore Processing Options for Oil Platforms”. In *Proceedings of the SPE Annual Technical Conference and Exhibition*. Society of Petroleum Engineers, Houston, USA, pp. 1–17 (Paper SPE 90325).

Bukowski, J.D., 2013. “Natural gas liquefaction technology for floating LNG facilities”. In *Proceedings of the 17th International Conference & Exhibition on Liquefied Natural Gas*. Houston, United States.

Cao, W.s., Lu, X.s., Lin, W.s. and Gu, A.z., 2006. “Parameter comparison of two small-scale natural gas liquefaction processes in skid-mounted packages”. *Applied Thermal Engineering*, Vol. 26, No. 8, pp. 898–904.

Chang, H.M., 2015. “A thermodynamic review of cryogenic refrigeration cycles for liquefaction of natural gas”. *Cryogenics*, Vol. 72, pp. 127–147.

He, T. and Ju, Y., 2014. “Performance improvement of nitrogen expansion liquefaction process for small-scale lng plant”. *Cryogenics*, Vol. 61, pp. 111–119.

Kusmaya, Maya, 2014. “Liquefaction process evaluation for floating LNG”. Technical report, SINTEF, NTNU, Trondheim, Norway.

Nguyen, T.V., Fülöp, T., Breuhaus, P. and Elmegaard, B., 2014a. “Life performance of oil and gas offshore platforms –site-scale integration and exergy-based assessment”. *Energy*, Vol. 73, pp. 282–301.

Nguyen, T.V., Jacyno, T., Breuhaus, P., Voldsund, M. and Elmegaard, B., 2014b. “Thermodynamic analysis of an upstream petroleum plant operated on a mature field”. *Energy*, Vol. 68, pp. 454–469.

Oliveira Jr., S.D. and Van Hombeeck, M., 1997. “Exergy Analysis of Petroleum Separation Processes in Offshore Platforms”. *Energy Conversion and Management*, Vol. 38, No. 15-17, pp. 1577–1584.

Remeljej, C. and Hoadley, A., 2006. “An exergy analysis of small-scale liquefied natural gas (lng) liquefaction processes”. *Energy*, Vol. 31, No. 12, pp. 2005–2019.

Sánchez, Y.A.C., de Oliveira Júnior, S., da Silva, J.A.M. and Nguyen, T.V., 2015. “Energy and Exergy Performance of three FPSO Operational Modes”. In *Proceedings of the 23rd ABCM International Congress of Mechanical Engineering*.

Shah, N.M., Hoadley, A.F. and Rangaiah, G., 2009. “Inherent safety analysis of a propane precooled gas-phase liquefied natural gas process”. *Industrial & Engineering Chemistry Research*, Vol. 48, No. 10, pp. 4917–4927.

Szargut, J., Morris, D. and Steward, F., 1988. *Exergy analysis of thermal, chemical, and metallurgical processes*. Hemisphere, New York, USA.

Voldsund, M., Nguyen, T.V., Elmegaard, B., Ertesvåg, I., Røsjorde, A., Jøssang, K. and Kjelstrup, S., 2014. “Exergy destruction and losses on four North Sea offshore platforms: A comparative study of the oil and gas processing plants”. *Energy*, Vol. 74, pp. 45–58.

7. RESPONSIBILITY NOTICE

The authors are the only responsible for the printed material included in this paper.