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Cost optimization and flexibility analysis for the liquefaction of an associated natural gas stream

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Abstract

Liquefaction and then transportation to the market is one of the promising options for the utilization of associated natural gas resources which are produced in oil fields. However, the flow of such resources is normally unsteady. Additionally, the associated gas in one oil field may exhaust in a few years and the liquefaction plant should be moved to another oil field with different specifications. In order to tackle such challenges, liquefaction systems not only must be optimally designed and operated, but also they should be flexible with respect to the gas flow fluctuations. The flexibility analysis of such processes is usually ignored in the optimization studies. In this research, first, the economic performance of two small-scale liquefaction processes (a Single Mixed-Refrigerant process, SMR, and a nitrogen expander process) were optimized and compared. The results showed that the SMR process is economically more attractive (49% lower lifecycle cost compared to the nitrogen expander process). As a post-optimization step, flexibility analysis was performed to investigate the ability of optimal designs in overcoming gas flow fluctuations. For this purpose, five-thousand feed samples with different flowrate and methane content were supposed which formed a feasibility-check region. The results showed that with respect to the design constraints, the optimal SMR process is more flexible and feasibly operates in the entire region. However, the nitrogen expander process cannot feasibly operate for the gas feed with high flowrate and low methane content.

Keywords: Associated Gas; LNG Process; Process Optimization; Equation-Oriented; Flexible process

1. Introduction

The improving living standards and growing population result in the increasing demand of energy. Natural gas, viewed as a cleaner alternative to other fossil fuels in terms of acidic and greenhouse gas emissions, provides an attractive option in order to fulfill this growing demand [1]. In comparison to the demand for coal and oil, the natural gas demand is increasing rapidly and it is also expected that the investment to find new natural gas reserves across the globe will increase in near future [2]. Associated natural gas (ANG) is one of the natural gas resources which is produced as a byproduct in oil fields. ANG is usually flared or vented, contributing a major part in global greenhouse gas emissions [3]. Therefore, the appropriate management of such resources has a very high priority not only with respect to environmental issues but also for monetizing them as energy sources.

Pipeline transmission is the most common transportation option for non-associated natural gases [4]. However, it is inflexible with respect to the transmission capacity and requires huge initial construction cost, and it is a large burden for small and offshore gas resources such as ANG [5]. Liquefied Natural Gas (LNG) is an energy carrier preferred over pipeline transmission for such resources [6,7]. The produced LNG can be transmitted and used in many industries including as the fuel of advanced combined power plants which have enhanced thermal efficiency [8] and as an alternative transportation fuel [9].

Among various types of liquefaction processes, nitrogen expander process and Single-stage Mixed-Refrigerant (SMR) process have been considered as the suitable liquefaction process for small-scale and offshore LNG platforms [10]. These processes have relatively simple configuration, require low capital investment, and provide quick start-up and shut-down [10,11]. The difference among these processes is mainly related to their construction and operating costs. Nitrogen expander processes are extremely compact and inherently safe to operate [11]. Compared to other technologies, they usually require small heat transfer area for the main heat exchanger (MHEX). However, they have a very high specific power demand (shaft work per produced amount of LNG). In SMR processes, a mixture

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Corresponding author's last name: Rashtchian

of nitrogen and light hydrocarbons are usually utilized as the refrigerant which evaporates over a similar range as natural gas. Therefore, they provide a close match between hot and cold composite curves in the MHEX and this ultimately reduces the power consumption [4]. It should be noted that in liquefaction processes, there is always a trade-off between power consumption and total heat exchanger surface area. Castillo and Dorao [11] have thoroughly compared pros and cons of different small-scale LNG technologies.

Qyyum et al. [2], He et al. [10], and Austbø et al. [12] have comprehensively reviewed the different aspects of the studies concerning LNG process design and optimization. Although LNG is a promising option for the management of ANG, the liquefaction process is energy intensive. The main contribution to energy consumption is the power consumption required by the compressor system. Due to this fact, most research activities have focused on the optimization of required power in liquefaction systems. In the following, several representative studies regarding the optimization of liquefaction process for associated natural gas and small-scale applications are reviewed.

Yin et al. [13] optimized two types of natural gas liquefaction processes (SMR and nitrogen expander) for small-scale applications. In their study the SMR process showed better performance in terms of both energy consumption and economics [13].

Shirazi and Mowla [14] optimized the power consumption of an SMR process for a peak shaving plant. Additionally, they evaluated the exergy efficiency of the equipment in the process. They concluded that the majority part of exergy loss occurs in the compression stages [14].

He and Lu [15] presented and optimized a novel process for small-scale natural gas liquefaction. The proposed process utilizes the pressure exergy of the pipeline to liquefy a part of natural gas without any energy consumption. Although this process has zero energy consumption, it can liquefy a small portion of the natural gas stream, and the majority of the gas remains in the gaseous form [15].

81 He and Ju [16] proposed and optimized a nitrogen expansion liquefaction process for an associated
 82 natural gas stream. Then, they tested the performance of the process using two different gas streams
 83 while fixing the process parameters as the baseline case. They concluded that the optimized process
 84 can also liquefy the new gas streams. Moreover, compared with two other nitrogen expansion
 85 processes, they showed that the proposed configuration has better performance in term of energy
 86 efficiency [16].

87 Yuan et al. [17] proposed and optimized a nitrogen expansion process with carbon dioxide pre-
 88 cooling. They investigated the exergy loss of the equipment and concluded that the compressor
 89 system causes the majority part of the exergy losses. In addition, they studied the effect of the natural
 90 gas feed condition on the energy performance of the process. They showed that the specific energy
 91 consumption is reduced as the feed gas pressure increases. On the other hand, increasing the feed
 92 temperature and methane content result in higher specific energy consumption [17].

93 Aslambakhsh et al. [18] optimized the lifecycle cost of a mini-scale SMR process. They investigated
 94 the effects of the process parameters (such as the refrigerant flowrate and inlet/outlet pressure levels
 95 of the compressors) on the plant lifecycle cost [18].

96 The abovementioned studies have mainly focused on the energetic analysis of liquefaction processes.
 97 However, in addition to the energy efficiency, the flexibility of liquefaction processes should be
 98 evaluated and analyzed. It is a crucial issue specially for associated gas produced in oil fields, because
 99 the gas flow is normally unsteady in such fields. Moreover, the associated gas in one oil field may
 100 exhaust in a few years. Therefore, the LNG plant should be moved to another oil field to produce
 101 LNG with different gas conditions [16]. Therefore, it is essential to analyze whether the liquefaction
 102 process is flexible enough to operate under different feed conditions.

103 Process flexibility analysis was introduced with the goal of evaluating robust and flexible process
 104 design [19–21]. Flexibility analysis can be used to quantify the extent to which the uncertainty or

105 changes in parameters are tolerated by a particular process design [22]. In other words, if a process
 106 remains feasible over a pre-defined range of potential deviations in process parameters, it could be
 107 concluded that the process is flexible in that range. In the feasible operation, all the constraints (heat
 108 and material balance, capacity, and equipment specifications) must be satisfied.

109 This study aims at investigating a suitable process in order to optimally liquefy an associated natural
 110 gas stream. To this end, two types of liquefaction processes (an SMR process and a nitrogen expander
 111 process) were modelled and optimized, and then their economic performances were compared. An
 112 Equation-Oriented (EO) approach was adopted to model the processes. The total lifecycle cost of the
 113 plants, operating for 25 years, was considered as the objective function. This objective covers both
 114 operational and capital costs and establishes a reasonable trade-off between power consumption and
 115 total heat exchanger surface area. As a post-optimization study, the flexibility of the SMR and
 116 nitrogen expander processes was compared using the enumeration of the feasible region. For this
 117 goal, a range for the feed conditions (flowrate and methane content) was considered while fixing the
 118 design parameters of the processes.

119 This section presented the research context. The rest of this paper is organized as follows. In Section
 120 2, initially, the design problem is presented and the process modeling approach are discussed. Then,
 121 the optimization formulation and the procedure of process flexibility analysis are discussed.
 122 Afterward, in Section 3, the results of both the optimization and the flexibility analysis are presented
 123 and compared for the liquefaction processes. Finally, Section 4 presents the conclusion of this
 124 research.

125 **2. Problem statement and methodology**

126 **2.1. Process modeling**

127 This study investigated the potential of utilizing an associated natural gas stream by liquefying it and
 128 producing LNG. The gas conditions and the LNG specifications are presented in Table 1.

129

130

Table 1 Associated natural gas and LNG specifications

Parameter	Natural Gas	LNG
Temperature (°C)	35	-158.3
Pressure (bar)	40	1.05
Flowrate (kmol/h)	2673.0	2673.0
Composition (mole %)		
Carbon dioxide	0.47	0.47
Methane	85.27	85.27
Ethane	7.39	7.39
Propane	5.31	5.31
Butanes	1.01	1.01
Pentanes	0.46	0.46
Hexanes	0.09	0.09
Nitrogen	0.00	0.00
H ₂ S	0.00	0.00
Liquefaction cooling load (kW)	11987.5	-

131 This gas is produced as a byproduct from an offshore and remote oil field. Therefore, it is crucial to
132 select a small-scale and compact process in order to meet offshore requirements. As discussed in
133 Section 1, single-stage mixed-refrigerant (SMR) and nitrogen expander processes are two suitable
134 candidates for small-scale applications. These processes have relatively simple configuration, require
135 low capital investment, and provide quick start-up and shut-down [10,11].

136 PRICO (Poly-Refrigerated Integrated Cycle Operations) is one of the well-known SMR processes
137 which is widely used in small-scale liquefaction facilities [10,11,18,23]. Figure 1 shows the flow
138 diagram of this process.

139 In liquefaction processes, the number of compressor stages depends on the compression ratio which
140 is limited to about four for each compression stage [24]. In comparison to single-stage compressor
141 systems, two-stage systems with intermediate cooling will increase the efficiency [25,26]. In this
142 process, the natural gas stream (stream NG1) enters the MHEX where it is cooled (stream NG2).
143 Furthermore, it is expanded to a near atmospheric pressure (stream LNG) and then it is transported
144 as the LNG product. The cooling duty is provided by a refrigeration cycle using a multi-component
145 working fluid (mixed-refrigerant). Conventionally, the mixture consists of nitrogen, methane, ethane,
146 propane and butanes [27]. In this process, a constraint on the degree of superheating for stream 4

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Assigned paper number: JERT-19-1723
Corresponding author's last name: Rashtchian

Table 2 PRICO process variable constraints

Process design variables	Lower Limit	Upper Limit
Suction pressure (bar)	1.50	8.00
1 st stage discharge pressure (bar)	1.50	90.00
2 nd stage discharge pressure (bar)	1.50	90.00
Mixed-refrigerant component flowrate (kmol/h)	0.0	5,000
Minimum internal temperature approach (K)	3.0	-
Degree of superheating (K)	10.0	-

Compared to SMR processes, nitrogen expander processes require smaller heat transfer area and are more compact while they consume more energy in the compressor system. A conventional nitrogen expander process consists of a multi-stage compressor system, an after cooler, a main heat exchanger, and an expander [29]. Figure 2 illustrates the process flow diagram of a nitrogen single-expander cycle.

Fig. 2 Nitrogen single-expander process

The nitrogen expander process which is shown in has totally five decision variables (suction pressure, 1st stage discharge pressure, 2nd stage discharge pressure, 3rd stage discharge pressure, and nitrogen flowrate). Table 3 presents the lower and upper bounds for these variables which were considered in this study.

Assigned paper number: JERT-19-1723
Corresponding author's last name: Rashtchian

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Table 3 Nitrogen expander process variable constraints

Process design variables	Lower Limit	Upper Limit
Suction pressure (bar)	1.50	10.0
1 st stage discharge pressure (bar)	1.50	90.0
2 nd stage discharge pressure (bar)	1.50	90.0
3 rd stage discharge pressure (bar)	1.50	90.0
Nitrogen flowrate (kmol/h)	0.0	20,000
Minimum internal temperature approach (K)	3.0	-
Degree of superheating (K)	10.0	-

176 In this study, an equation-oriented (EO) approach was adopted to model the PRICO and expander
177 processes [30]. EO approach enables simultaneous optimization and convergence of the whole
178 flowsheet, using an appropriate optimization algorithm. Rigorous thermodynamic calculations were
179 conducted using Peng-Robinson equation of states. For two-phase calculations, the formulation
180 proposed by Kamath et al. [31] was implemented. For the main heat exchanger, a detailed modeling
181 was conducted using energy balance equations with internal temperature feasibility constraints. The
182 temperature feasibility condition implies that the temperature difference between the hot and cold
183 composite curves at any internal point must be greater than the minimum internal temperature
184 approach value [32].

185 In the following section the formulation of the objective function is explained.

186 **2.2. Objective function**

187 In order to design a liquefaction process, there is always a trade-off between power consumption and
188 heat transfer area. In order to balance this situation, one of the most reliable objective functions is the
189 plant total lifecycle cost which includes both operational and capital costs [18]. Complete liquefaction
190 of the feed gas stream was supposed in this study, therefore, the LNG sale profit was fixed and was
191 not considered in the objective function. The plants were considered to operate for a period of twenty-
192 five years and 330 days per year. The electricity price was supposed to be 6.95 ¢/kWh [18]. The rise
193 in the electricity price during this period was considered to be 3% per annum [18].

194 For capital cost, only the purchase costs of compressors, turbine, and MHEX were considered. The
 195 costs of other equipment are negligible, compared to these elements [18]. The capital cost is finally
 196 multiplied by three in order to consider the installation, piping, set up and other supplementary prices
 197 [18].

198 The purchase cost of a centrifugal compressor was calculated using the following equation [33]:

$$C_B = \exp\{9.1553 + 0.63[\ln(P_c)]\} \quad (1)$$

199 Here C_B and P_c are the FOB (free on board) price (US dollars in 2013) and power consumption of a
 200 compressor (horsepower, hp), respectively.

201 For the estimation of the purchase cost of an expander, the following correlation was used [33]:

$$C_B = 600 P_E^{0.81} \quad (2)$$

202 In this equation C_B is the FOB price of an expander in US dollars (price in 2013) and P_E is the
 203 generated power in horsepower (hp).

204 The main heat exchanger price was calculated based on 425 $US\$/m^2$ in 2011 and considering an
 205 overall heat transfer coefficient, U , of 700 $kJ/m^2.h.K$ [18]:

$$C_B = 425A \quad (3)$$

206 In this formula C_B is the FOB price of the MHEX in US dollars (price in 2011) and A is the heat
 207 transfer area of the MHEX (in m^2).

208 The objective function was calculated based on the prices in 2018, therefore chemical engineering's
 209 plant cost index (CEPCI) was considered to adjust the prices ($CEPCI_{2011} = 585.7$, $CEPCI_{2013} = 567.3$,
 210 and $CEPCI_{2018} = 603.1$) [34].

2.3. Flexibility analysis

In this study, flexibility analysis (using feasible region enumeration) was used to identify the extent to which the uncertainty or changes in the feed condition are tolerated by the optimal designs. In other words, after the process optimization using the baseline feed condition (Table 1), both processes were optimized considering a range of feed overall flowrate and methane content (as the dominant component of the natural gas).

For this analysis, two more constraints were added to the optimization problem. First, the heat transfer area (UA) was supposed to be less or equal than the optimized value (the value obtained for the baseline feed conditions). This constraint limited the available heat transfer area to the optimized value. Second, the refrigerant flowrate inside the liquefaction cycle was limited to vary in a range of 90% to 110% of the optimized flowrate (the value obtained for the baseline feed conditions). This ensures a reasonable flowrate with respect to the optimized capacity of the cycles.

In order to form a feasibility-check region, a range between -10% and +10% of the baseline feed flowrate was supposed for the natural gas flowrate, and for the methane content a range between 50% to 95% was considered. In order to generate feed sample points (totally 5000 samples) in such a way that uniformly covers the above-mentioned range, Latin Hypercube Sampling (LHS) was used [35,36]. LSH is an efficient method for generating random samples of uncertain parameters in a multi-dimensional design space [36]. The distribution of the samples is shown in Fig. 3. The star symbol in the figure shows the baseline feed condition.

Fig. 3 The sample region for the flexibility analysis

As the methane content of the natural gas increases, the temperature of the natural gas which leaves the MHEX (stream NG2 in Fig. 1 and Fig. 2) decreases. Therefore, the refrigerant must provide such a low temperature. On the other hand, as the methane content decreases the feed becomes heavier. Hence, the cooling load that must be absorbed by the cycle increases. By increasing the feed flowrate, the cooling load also increases. Figure 4 shows the gas leaving temperature as well as the required cooling load for the feed samples generated in this section. As this figure shows, the bottom-right part of the region has the highest cooling load.

Fig. 4 (a) The temperature of the natural gas after the MHEX, and (b) The required cooling load for each feed sample

Considering the overall region shown in Fig. 3, the process which feasibly operates in more points (feed samples) is considered to be more flexible. The next section presents the optimization results as well as the results of the flexibility analysis.

3. Results and discussion

The PRICO and nitrogen expander processes were modeled in GAMS software (version 24.7.4) [37]. These models were formulated as NLP (nonlinear programming) problems and solved using CONOPT (CONtinuous OPTimization) solver [38] in GAMS. In order to well initialize the optimization process, the processes were first simulated in Aspen HYSYS (version 8.8) software and the initial value of the variables was extracted from the simulation case.

3.1. Optimization results

Considering the lifecycle costs as the objective function, the PRICO and nitrogen expander processes were optimized. Table 4 and Table 5 show the optimization results for the PRICO process and nitrogen expander process, respectively.

Table 4 The optimal solution of the PRICO process

Variable	Optimal value
Suction pressure (bar)	1.52
1 st stage discharge pressure (bar)	6.08
2 nd stage discharge pressure (bar)	24.30
Mixed-refrigerant flowrate (kmol/h)	5930.1
Mixed-refrigerant composition (%)	
nitrogen	6.46
methane	21.42
ethane	35.34
propane	0.00
i-butane	0.00
n-butane	36.78
Power consumption (MW)	14.91
UA (kJ/K.h)	3.04×10^7
Capital cost (\$)	77×10^6
Operating cost (\$), 25 years operation	298×10^6
Plant total cost (\$), 25 years operation	375×10^6

Table 5 The optimal solution of the nitrogen expander process

Variable	Optimal value
Suction pressure (bar)	1.82
1 st stage discharge pressure (bar)	6.96
2 nd stage discharge pressure (bar)	25.44
3 rd stage discharge pressure (bar)	90.00
Nitrogen flowrate (kmol/h)	10646.1
Power consumption (MW)	34.06
UA (kJ/K.h)	3.73×10^6
Capital cost (\$)	58×10^6
Operating cost (\$), 25 years operation	680×10^6
Plant total cost (\$), 25 years operation	738×10^6

As Table 4 and Table 5 show, the PRICO process is economically more attractive than the nitrogen expander process for the liquefaction of the associated natural gas stream. Although nitrogen expander process has a significantly smaller UA, but it requires more shaft work (128% higher) than that of the PRICO process.

3.1. Flexibility analysis

As it was discussed in Section 2.3, considering a range between 2406 kmol/h (-10% of the baseline feed flowrate) to 2940 kmol/h (+10% of the baseline feed flowrate) for the natural gas flowrate and a range between 50% to 95% for the natural gas methane content, 5000 feed samples were generated using LHS method (as shown in Fig. 3). Both PRICO and nitrogen expander processes were optimized for each sample. In order to fix the process designs, two more constraints were added to the process models with respect to the optimal design parameters in Table 4 and Table 5; one constraint was considered to limit the process flowrate (a representative measure of the cycle capacity) and another to provide an upper limit on the available heat transfer area. These constraints are as follows:

For the PRICO process:

$$\left[0.9 \times 5930.1 \frac{\text{kmol}}{\text{h}} = 5337.1 \frac{\text{kmol}}{\text{h}} \right] \leq MR \text{ Flowrate} \leq \left[1.1 \times 5930.1 \frac{\text{kmol}}{\text{h}} = 6523.1 \frac{\text{kmol}}{\text{h}} \right] \quad (4)$$

$$UA \leq 3.04 \times 10^7 \frac{\text{kJ}}{\text{K.h}} \quad (5)$$

For the nitrogen expander process:

$$\left[0.9 \times 10646.1 \frac{\text{kmol}}{\text{h}} = 9581.5 \frac{\text{kmol}}{\text{h}}\right] \leq N_2 \text{ flow rate} \leq \left[1.1 \times 10646.1 \frac{\text{kmol}}{\text{h}} = 11710.7 \frac{\text{kmol}}{\text{h}}\right] \quad (6)$$
$$UA \leq 3.73 \times 10^6 \frac{\text{kJ}}{\text{K} \cdot \text{h}} \quad (7)$$

Figure 5 presents the results (feasible samples) for the PRICO process. As this figure shows the PRICO process is feasible for the entire region.

Fig. 5 the flexible region for the optimal PRICO process

The contour plot of the refrigerant flowrate and power consumption for the feasible points of the PRICO process are shown in Fig. 6.

Fig. 6 (a)The refrigerant flowrate and (b) power consumption contours for the PRICO flexible region

As Fig. 6 (a) shows the upper limit of the mixed-refrigerant flowrate is touched for the feed samples with high natural gas flowrate and high methane content. Also the figure shows that such feed samples require more power consumption (Fig. 6 (b)).

Table 6 presents the feed conditions as well as the optimal design parameters for the points A, B, C, and D which are shown in Fig. 6.

Table 6 Feed condition and optimal design for the points shown in Fig. 6

Parameter	Point A	Point B	Point C	Point D
Temperature (°C)	35	35	35	35
Pressure (bar)	40	40	40	40
Flowrate (kmol/h)	2409.4	2410.2	2934.4	2934.7
Composition (%)				
Carbon dioxide	1.59	0.16	0.16	1.58
Methane	50.05	94.84	94.85	50.41
Ethane	25.07	2.59	2.59	24.89
Propane	18.00	1.86	1.86	17.87
Butanes	3.40	0.35	0.35	3.37
Pentanes	1.58	0.16	0.16	1.57
Hexanes	0.31	0.03	0.03	0.31
Cooling load (kW)	13222.1	9993.9	12166.3	16094.8
Suction pressure (bar)	1.60	1.50	1.50	1.50
Intermediate pressure (bar)	5.07	5.76	6.00	5.84
Discharge pressure (bar)	20.30	23.03	24.0	23.34
Mixed-refrigerant flowrate (kmol/h)	5590.6	5786.0	6523.1	6523.1
Mixed-refrigerant composition (%)				
nitrogen	5.37	6.47	6.56	5.28
methane	22.58	21.39	20.88	23.28
ethane	26.83	38.64	41.41	26.74
n-butane	45.21	33.50	31.15	44.71
Power consumption (MW)	12.71	14.30	16.40	16.17
UA (kJ/K.h)	3.04×10^7	2.37×10^7	2.84×10^7	3.04×10^7

As Table 6 shows, when the methane content decreases (heavier feed samples, such as points A and D), the mixed-refrigerant becomes heavier by increasing the flowrate of the heaviest component (n-butane) in order to provide the cooling load. This fact is also shown in Fig. 7.

Fig. 7 n-butane content in the mixed-refrigerant for the feasible region of the PRICO process

Figure 8 present the feasible region for the nitrogen expander processes. As this figure shows the nitrogen expander process is not feasible for the bottom-right part of the region. It means that the optimal design which obtained for the baseline feed condition, cannot operate for the feed samples with very high cooling load. This is due to the fact that for such feed samples, the process requires higher refrigerant (nitrogen) flow rate. However the refrigerant flowrate is limited by a constraint (equation 6). In the PRICO process this issue is managed by making the mixed-refrigerant heavier (i.e. increasing n-butane content). But this is not possible in the nitrogen expander process; because the refrigerant in this process is pure nitrogen. Figure 9 shows the nitrogen flowrate as well as the power consumption for the feasible region of the nitrogen expander process. As Fig. 9 shows, by increasing the required cooling load of the feed (low methane content and high flowrate) the power consumption of the process increases. It happens because the refrigerant flowrate is increased for such conditions.

Fig. 8 the flexible region for the optimal nitrogen expander process

Fig. 9 (a) The refrigerant flowrate and (b) power consumption contours for the nitrogen expander flexible region

The feed conditions at points A, B, C, D, and E (shown in Fig. 9) as well as the optimal parameters of the nitrogen expander process for these points are presented in Table 7. As this table shows, the nitrogen flowrate limit is touched for the points C, D, and E, because these points have higher cooling load than the others.

The results presented in this section showed that the PRICO process is economically more attractive in order to liquefy the associated natural gas stream; its power consumption (which is the dominant factor in the lifecycle costs of a liquefaction process) is lower than the nitrogen expander process. The nitrogen expander process requires smaller heat transfer area and consequently it has lower capital investment in comparison to the PRICO process.

The results of the flexibility analysis showed that the nitrogen expander process cannot feasibly operate at high feed flowrate and low methane content (higher cooling load) while the PRICO process is more flexible and operates in the entire feasibility-check region.

Table 7 Feed condition and optimal design for the points shown in Fig. 9

Parameter		Point A	Point B	Point C	Point D	Point E
Feed conditions	Temperature (°C)	35	35	35	35	35
	Pressure (bar)	40	40	40	40	40
	Flowrate (kmol/h)	2409.4	2410.2	2934.4	2929.6	2694.2
	Composition (%)					
	Carbon dioxide	1.59	0.16	0.16	0.83	1.56
	Methane	50.05	94.84	94.85	73.83	50.90
	Ethane	25.07	2.59	2.59	13.14	24.65
	Propane	18.00	1.86	1.86	9.43	17.69
	Butanes	3.40	0.35	0.35	1.78	3.34
Optimal process variables	Pentanes	1.58	0.16	0.16	0.83	1.56
	Hexanes	0.31	0.03	0.03	0.16	0.31
	Cooling Load (kW)	13222.1	9993.9	12166.3	14280.8	14763.0
	Suction pressure (bar)	2.03	2.27	2.11	1.51	1.51
	1st stage discharge pressure (bar)	7.48	8.06	7.89	5.64	5.65
	2nd stage discharge pressure (bar)	26.37	27.37	27.08	22.57	22.60
	3rd stage discharge pressure (bar)	90.00	90.00	90.00	90.00	90.00
	Nitrogen flowrate (kmol/h)	11624.1	9581.5	11710.7	11710.7	11710.7
	Power consumption (MW)	35.43	28.73	35.72	40.17	39.74
	UA (kJ/K.h)	2.92×10 ⁶	3.62×10 ⁶	3.73×10 ⁶	3.73×10 ⁶	3.09×10 ⁶

4. Conclusion

Associated natural gas is one of the natural gas resources which is produced as a byproduct in oil fields. Such gas resources are usually flared or vented, contributing a major part in global greenhouse gas emissions. This research focused on the optimization of liquefaction processes in order to liquefy an associated natural gas stream. To this end, two small-scale liquefaction processes (PRICO process and nitrogen expander process) were modeled and optimized using an equation-oriented approach. The objective function was the plant lifecycle costs in which both capital and operation costs were included for a period of twenty-five years. This objective function was selected because it can establish a reasonable trade-off between shaft work and total heat exchanger surface area of the liquefaction processes. The results showed that the PRICO process is economically more attractive (49% lower lifecycle cost compared to the nitrogen expander process). Although it requires a considerably larger heat transfer area and consequently higher capital cost, it consumes 56% lower shaft work which has the dominant effects on the lifecycle cost.

Besides the economic performance, the flexibility of liquefaction processes is an important issue which is not usually considered in the optimization studies. This is an important issue specially when the feed is an associated gas stream. In such cases, the gas flow is normally unsteady. Moreover, the associated gas in one oil field may exhaust in a few years and the liquefaction plant should be moved to another oil field with different gas specifications. The results of the flexibility analysis provide an in-depth insight regarding the behavior of the processes in case of gas flow fluctuations which has not been well studied before. In this research, the flexibility of the aforementioned liquefaction processes was also studied. For this purpose, it was supposed that the feed gas flowrate can vary in a range of $\pm 10\%$ of the baseline value and the methane content of the feed gas varies between 50% to 95%. Using Latin Hypercube Sampling (LHS), five-thousand feed samples were generated which formed a feasibility-check region. The processes were optimized again for each feed sample considering two more constraints which made the designs to be fixed (refrigerant flowrate and overall heat transfer coefficient) and the feasible region of each design was identified. The results showed that with respect to the design constraints, the PRICO process is more flexible and operates in the entire feed condition region. However, the nitrogen process could not feasibly operate for the feed samples with high flowrate and low methane content.

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375 **5. Nomenclature**

Abbreviations

ANG	Associated Natural Gas
CEPCI	Chemical Engineering's Plant Cost Index
EO	Equation-Oriented
FOB	Free On Board
LHS	Latin Hypercube Sampling
LNG	Liquefied Natural Gas
MHEX	Main Heat Exchanger
MR	Mixed-Refrigerant
PRICO	Poly-Refrigerated Integrated Cycle Operations
SM	Sequential Modular
SMR	Single-stage Mixed-Refrigerant

Symbol

A	Heat transfer area of the MHEX (m ²)
C _B	Free on board price (US dollars)
P _c	Power consumption of a compressor (horsepower, hp)
P _E	Expander generated power (horsepower, hp)
U	Overall heat transfer coefficient (kJ/m ² .h.K)
UA	The product of overall heat transfer coefficient and heat transfer area (kJ/h.K)

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379 6. References

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477 **List of table captions**

478	Table 1	Associated natural gas and LNG specifications
479	Table 2	PRICO process variable constraints
480	Table 3	Nitrogen expander process variable constraints
481	Table 4	The optimal solution of the PRICO process
482	Table 5	The optimal solution of the nitrogen expander process
483	Table 6	Feed condition and optimal design for the points shown in Fig. 6
484	Table 7	Feed condition and optimal design for the points shown in Fig. 9

485
486 **List of figure captions**

487	Fig. 1	PRICO process flow diagram
488	Fig. 2	Nitrogen single-expander process
489	Fig. 3	The sample region for the flexibility analysis
490	Fig. 4	(a) The temperature of the natural gas after the MHEX, and (b) The required cooling load for
491		each feed sample
492	Fig. 5	the flexible region for the optimal PRICO process
493	Fig. 6	(a)The refrigerant flowrate and (b) power consumption contours for the PRICO flexible region
494	Fig. 7	n-butane content in the mixed-refrigerant for the feasible region of the PRICO process
495	Fig. 8	the flexible region for the optimal nitrogen expander process
496	Fig. 9	(a) The refrigerant flowrate and (b) power consumption contours for the nitrogen expander
497		flexible region