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Skid-Mounted SMR Packages for LNG Production: Configuration Selection and Sensitivity Analysis

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ABSTRACT

On review of skid-mounted LNG technology providers, single mixed refrigerant (SMR), ni-trogen expander and self-refrigerated processes have been used for LNG production in skid scale. However, SMR processes are more efficient and have lower rotating equipment. By RIPI comparative study on commercialized SMR processes and more than 100 patents in this topic, the SMR process with one phase separator (by 43% sharing in SMR processes), has been selected for skid LNG plant. Regarding to process complexity of multi-phase separators in SMR loop, these types of cycles were not selected. Otherwise SMR process without phase separator was not selected for skid LNG plant because of the freezing possibility of heavy hydrocarbon refrigerants in this configuration.

Several single-phase separator SMR processes can be used based on arrangement of equipment in liquefaction and refrigeration sections. By extensive study and according to skid design limitations (e.g., the minimum number of fixed and rotating equipment, minimum process complexity and dimension and etc.), two process arrangements has been selected, simulated and optimized. Also, a sensitivity analysis on the feed pressure and temperature as well as the composition of MR and feed was done. Energy consumption of these two configurations was calculated and the complexity of them was compared. According to the results obtained in this study and considering lower total annualized cost of LNG unit and the necessity of pro-cess simplicity in the skid scales, the best case was recommended for LNG skid-mounted packages.

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1. Introduction

The increasing of LNG production volume in the world supports the significant contribution of growing natural gas demand by developing conventional & unconventional gas sources (Barcly and Shukri, 2007). Gas pipeline and Liquefied natural gas (LNG) are two major op-tions for transferring natural gas from sources to end users. Lower operating cost of gas pipe-line in comparison with LNG is the main reason which it is generally used for transporting natural gas; although it needs high capital investment cost for long distances gas transmission in addition to limitation of gas transfer capacity (Moein et al., 2015).

LNG is natural gas (predominantly methane) that has been cooled down to liquid, at tempera-ture below -161°C and pressure 101.325 kPa with its volume reduced by a factor of 600. LNG production, storage and safe transfer need costly and high-tech equipment (Cranmore and Stanton, 2000). Liquefied natural gas is considered as efficient, clean and economical energy sources. It is a fuel for the future; because its low NO_x and SO_x emissions combined with re-duced CO_2 permanently improve the ecological balance (He et al. 2018). Using LNG for gas transferring from the remote gas resources to consumers has been done (Pfoser et al., 2018). According to IGU World Gas LNG Report in 2018 Edition, the nominal LNG production capacity has been increased up to 359.5 MTPA at the end of 2017.

There are two different LNG production capacities: The first one is base-load liquefaction plants with capacity more than 3.4 MTPA, the second one is small scale liquefaction plants with capacity lower than 1 MTPA per train (Yin et al., 2008). In the last decade, skid mount-ed or containerized LNG plant is suggested to provide a cleaner, abundant fuel source that is ideal for use with stranded gas sources not connected to a network. Small scale LNG plants are developed in recent decades regarding to new applications such as using natural gas in heavy vehicles and utilization of small gas resources (Yin et al., 2008; Kohler et al., 2014). Be-side the environmental and economic benefits of small-scale LNG development in the global supply chain, the capital investment cost per ton of LNG production is more than base-load LNG plants (Nguyen et al., 2016).

Skid mounted LNG plant is the intellectual choice for virtual pipelines, producing a replacea-ble fuel for diesel, high horsepower fuel applications including marine, rail, mining, drilling and other oil industrial fuel applications. The facilities are modular, compact, quick installa-tion, commissioning and running. The other benefits include lower construction costs, im-proved quality and safety, faster project execution, and easily moved and re-deployed in the future if desire. Skid mounted solutions for LNG processing mainly used in pipeline gas, oil-associated gas, flare gas, bio gas and other small scale conventional and unconventional gas sources.

According to the review of more than 20 skid-mounted LNG technology providers, single mixed refrigerant (SMR), nitrogen expander and self-refrigerated processes have been used for liquefaction process of skid scale and Iso-Container type LNG.

These liquefaction processes are different due to their equipment and operation costs and min-imum approach temperature in composite curves (Hatcher et al., 2012). Generally, the opera-tion cost (due to high efficiency) and the number of rotary equipment of SMR processes are lower than N_2-expansion cycles (Moein et al., 2016) and self-refrigerated processes. Process simplicity, single phase and non-toxic refrigerant, against high energy consumption per unit of LNG production and the more number of rotating equipment are the pros and cons of N_2 ex-pander process (Moein et al., 2015). Self-refrigerated processes use a part of inlet feed to liq-uefaction unit (treated gas) as refrigerant for LNG production. Typically, these processes work at high operating pressures (e.g., 250 barg) and they need an auxiliary cooling cycle like pro-pane or ammonia for LNG production. One advantage of this cycle is using (or even not being used) low volume of pure refrigerant (e.g., propane) in liquefaction cycle and there will be no a lot of supply, storage, leakage and make up problems of refrigerants as a main issue in this process.

Among the mixed refrigerant processes, the single mixed refrigerant process is the simplest one (Cao et al., 2006) with simple multi stage compressors and one main cold box. This process can be used for all ranges of liquefaction production rates from small scale to base-load (Swenson, 1977). SMR processes can be divided into two categories. The first group process-es contain phase separators to separate the vapor and liquid phases of mixed refrigerant and the second one does not use any separator at mixed refrigerant loop (Venkatarathnam, 2008). The SMR processes with phase separator can be organized into single phase separator to more than four phase separators. Although increasing the number of separators increases process efficiency, but process complexities and foot print also increase (Yin et al., 2008).

The enthalpy of natural gas changes nonlinearly during liquefaction process because of com-plex nature of mixed hydrocarbons. The efficiency and entropy of liquefaction processes can be increased and decreased respectively by decreasing the temperature difference between hot and cold composite curves, in cryogenic heat exchangers. MR cycle requires no additional equipment such as turbo expander for reducing refrigerant temperature (Venkatarathnam, 2008). Also, power consumption in MR processes is very significant, so the optimization method is needed to minimize energy consumption. Some references in the open literature review different methods of MR process optimization. Gong et al. (2000) applied the BOX optimization tool and Wahl et al. (2013) used sequential quadratic programming (SQP) to op-timize the PRICO process. Lee et al. (2002) optimized a multi-stage MR system and mini-mized shaft work requirement by a graphical targeting technique. Aspelund et al. (2010) opti-mized the PRICO process by using of Tabu Search algorithm (TS) and Nelder-Mead Downhill Simplex (NMDS) method. Genetic algorithm used for optimization of SMR process by Cam-marata et al. (2001), Mokarizadeh and Mowla (2010), Taleshbahrami, H. and Saffari, H. (2010), Alabdulkarem et al. (2011), Li et al. (2012), Xu et al. (2013), He and Ju (2014), Moein et al. (2015), Ding et al. (2017) and Abdelhamid et al. (2017). Combination of the genetic algorithm and sequential quadratic programming used by Hwang et al. (2013) for optimizing a DMR LNG process. Particle-swarm paradigm coded in MATLAB connected to UniSim simu-lation software used by Khan and Lee (2013) and Khan et al. (2013) and MATLAB's built-in fmincon solver used by Jacobsen and Skogestad (2013) to optimize a SMR process. Austbo et al. (2014) applied Sequential optimization for SMR LNG process. Ngoc Pham et al. (2016) optimized the modified SMR LNG process using by multivariate Coggin's algorithm com-bined with process knowledge. Ali et al. (2018) optimized energy for SMR LNG process us-ing the metaheuristic vortex search algorithm. Qyyum et al. (2018) applied a hybrid modified coordinate descent (HMCD) algorithm to optimize the SMR LNG process.

Because of the nature of nonlinearity and thermodynamic complexity of the LNG processes, optimization of these processes should be performed by global search methods such as GA, Tabu Search (TS), or Simulated Annealing (SA). Due to the fact that GA does not require derivatives and initial points, it was chosen as the optimization method for the current research.

In this study, SMR process was selected for liquefaction of natural gas due to its more effi-cient; lower operating pressure and lower number of rotary equipment rather than the other refrigeration cycles. By comprehensive study on various SMR configurations in the liquefaction and refrigeration section arrangements, concerning skid design limitations (e.g. the minimum fixed and rotating equipment, minimum process complexity, minimum dimension and etc.), two process arrangements has been selected, simulated in Aspen HYSYS and optimized by Genetic Algorithm to minimize required work for LNG production. The sensitivity analysis by feed specification and MR composition changes on the total required work were investi-gated. Energy consumption of these two configurations was calculated and compared with each other to select the best configuration.

2. SINGLE MIXED REFRIGERANT PROCESS CONFIGURATION SELECTION

As mentioned before, single mixed refrigerant processes can be classified into those that use phase separator and those that do not (Venkatarathnam, 2008). The most famous SMR pro-cess without phase separator (in refrigeration loop) is PRICO process by Black & Veatch. As a definition, when we count a phase separator that after separation of liquid and gas phases, each one enters to heat exchanger separately. So, in PRICO process, we do not have any phase separator given the above definition.

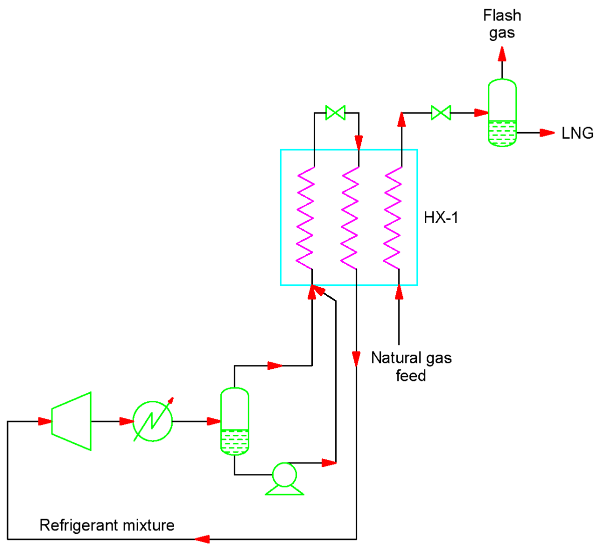


Figure 1. PRICO Liquefaction Process

In the PRICO process freezing possibility of high boilers in the refrigerant can be solved using phase separators that return heavy refrigerant components to the compressor at higher temper-atures, much above the freezing point of the heavy refrigerants (Venkatarathnam, 2008).

The SMR processes with phase separator can be organized into single to five phase separators. This categorization is based on patent analysis and not necessarily all of them in-dustrialized. Depending on different factors such as investment cost, operating costs and pro-cess complexity & flexibility; the optimal number of separators will be determined.

Although increasing the number of separators increases process efficiency, but process com-plexities also increase (Yin et al., 2008).

According to RIPI study on commercialized SMR processes and more than 100 patents in this topic, the order of SMR processes based on their contribution is as follows:

- Without phase separator: 21%
- 1-phase separator: 43%
- 2-phase separators: 23%
- 3-phase separators: 9%
- 4-phase separators: 3%
- 5-phase separators: 1%

Figure 2 shows a typical 1-phase separator SMR loop. In this configuration, the 2-phase mixed refrigerant stream, is separated to liquid and vapour streams PS-1 and enter to the cold box separately. There is no any separator in the flow path of the refrigerant and low-pressure re-frigerant enter to compressor package.

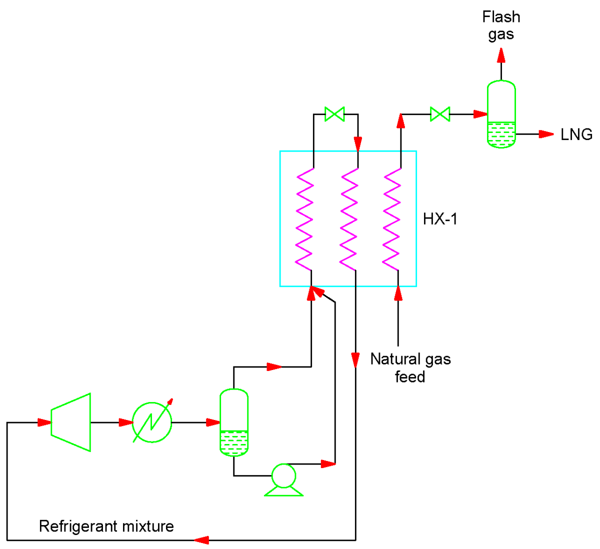


Figure 2. Typical 1-phase separator SMR loop

Figure 3 shows a typical 2-phase separators SMR loop. In this configuration, the 2-phase mixed refrigerant stream, is separated to liquid and vapour streams in PS-1 and enter to the cold box. After pre-cooling the vapour MR, this 2-phase stream is separated in the second phase separator (PS-2).

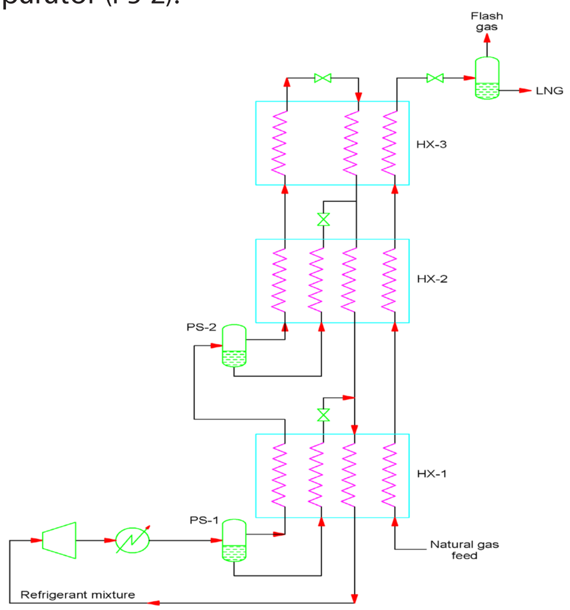


Figure 3. Typical 2-phase separators SMR loop

Figures 4, 5 and 6 show a typical three, four and five phase separators in the flow path of the mixed refrigerant. As seen, by increasing the number of phase separators, the complexity of the refrigerant cycle and cooling stages in cryogenic heat exchanger is greatly increased. Ap-proximately, the cooling stages are equal to the number of phase separator plus one.

Due to the freezing possibility of refrigerant high boilers in the PRICO process, a phase sepa-rator should be considered in the selected process. On the other hand, as shown above, the process complexity of more than 1 phase separator in SMR loop and also the considerations of small foot print in skid design, the SMR process with one phase separator (by 43% sharing in SMR processes) has been selected in this paper. The commercialized liquefaction units by companies such as Linde, Black & Veatch, Air Products and so on, also contains only the 0-phase, 1-phase and 2-phase separators in their refrigeration cycles.

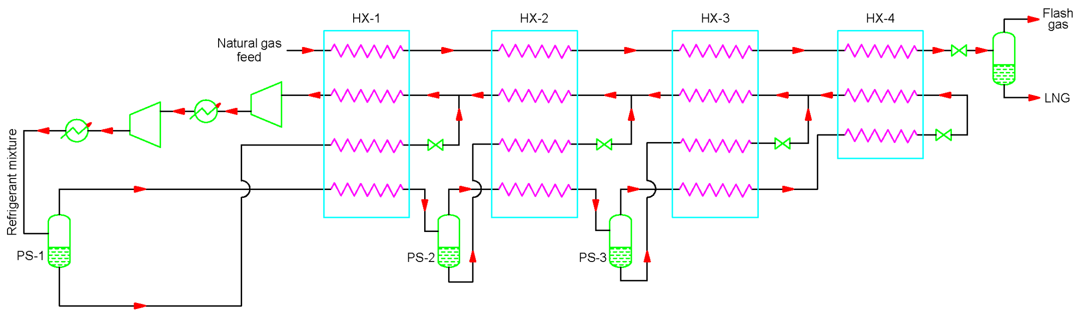


Figure 4. Typical 3-phase separators SMR loop

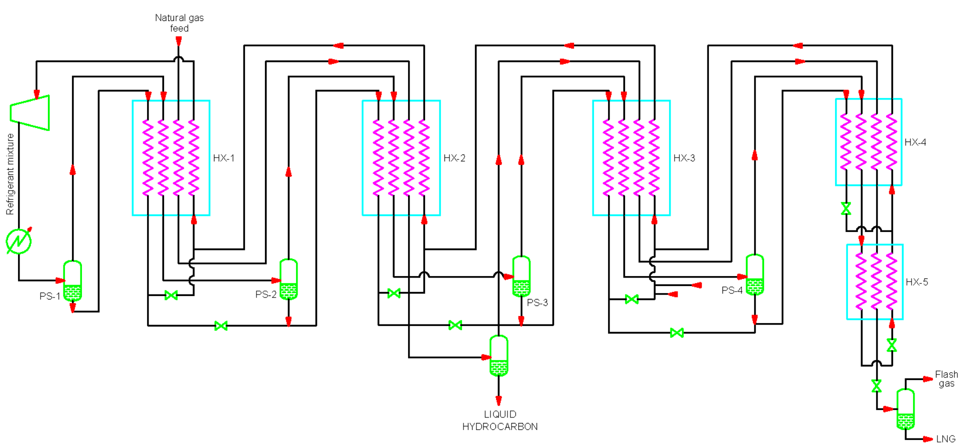


Figure 5. Typical 4-phase separators SMR loop

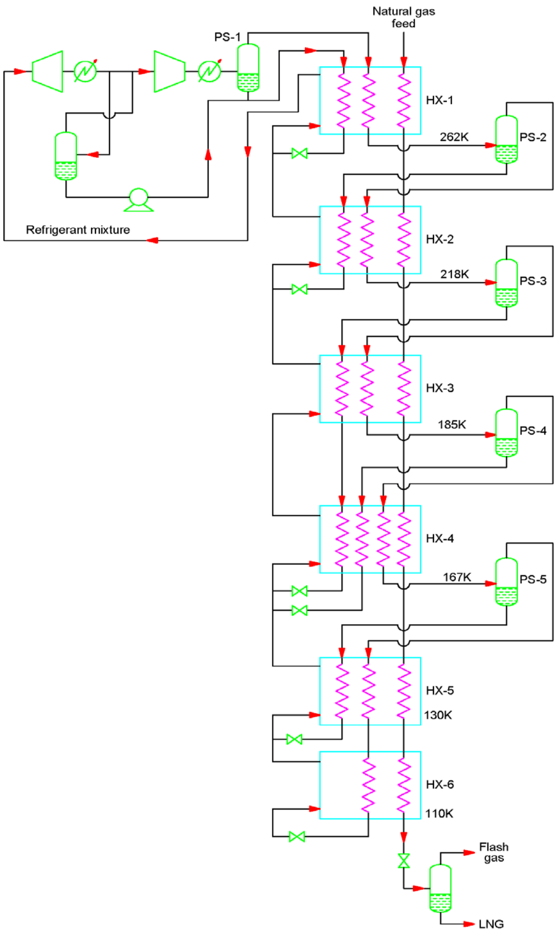


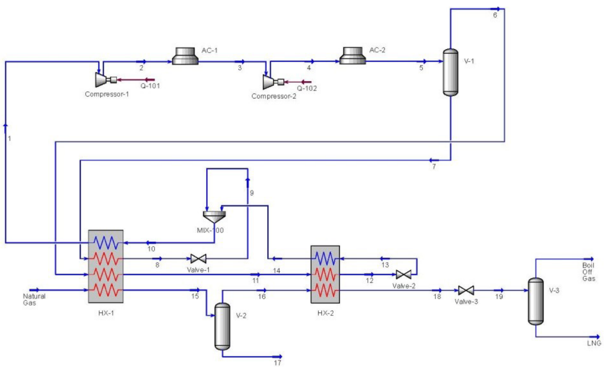
Figure 6. Typical 5-phase separators SMR loop

2.1. SMR PROCESS FOR SKID DESIGN

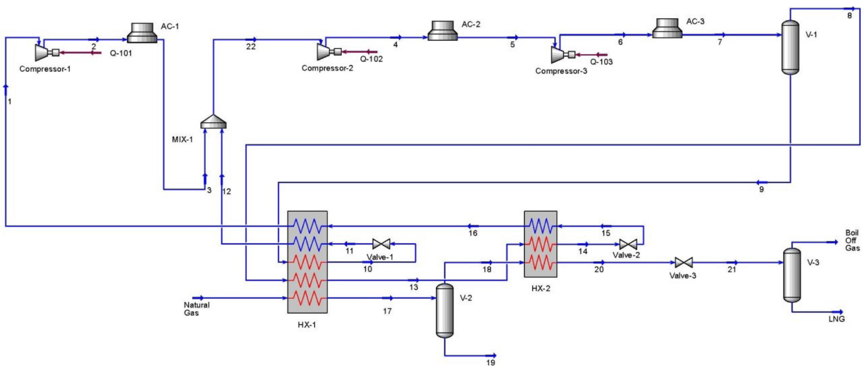
There are various types of SMR process with single phase separator in refrigeration cycle. These processes are different in the location of phase separator, the number of outlet low pres-sure refrigerant stream from cold box, the configuration of compression section, the presence of pump in compression cycle and etc. According to comprehensive study (simulation, optimi-zation, process analysis and etc.) of these configurations in RIPI and with respect to skid de-sign limitations (e.g., the minimum number of fixed and rotating equipment, minimum number of streams which has exchanged heat in cold box, minimum process complexity and etc.), two process arrangements has been selected (Schmidt, 2009; Heng and Wenhua, 2014).

2.1. SMR PROCESS SIMULATION (TWO CONFIGURATIONS)

According to section 2.1, two SMR process configurations have been selected for liquefaction unit. The flow sheets of these two processes modeled by Aspen HYSYS are shown in Figure 7.



(a)



(b)

Figure 7. SMR process simulation of (a) Configuration Case I, (b) Configuration Case II.

As shown in Figure 7 (a), high pressure stream (stream No. 4) after leaving the last stage of the compressor and cooling in air cooler (AC-2), is divided into two phases (liquid and vapor) in a separator (V-1). These two streams enter the first heat exchanger (HX-1) individually and their temperature decreases. The cooled liquid stream passes through a throttling valve (stream No. 9). The cooled vapor phase goes to the second heat exchanger (HX-2) and after cooling, passes through a throttling valve and returns to the heat exchanger as a cold stream and after heat exchanging (stream No. 14), mixed with stream No. 9. The mixed stream enters to the first heat exchanger (HX-1) as a cold stream and after heat exchanging, enters the first stage of compression. In this process, the temperature of natural gas after two steps decreases to about -150 ºC and after throttling valve, LNG is produced at -161 ºC.

In Figure 7 (b), high pressure stream (stream No. 6) after leaving the third stage of the com-pressor and cooling in air cooler (AC-3), is separated into two phases (liquid and vapor) in a separator (V-1). Each of these streams enters to the first heat exchanger (HX-1) separately as hot streams. The cooled liquid stream after passing through a throttling valve returns to the heat exchanger as a cold stream and after heat exchanging (stream No. 12), mixes with steam No.3 and enters the second stage of compression. The cooled vapor phase goes to the second heat exchanger. The exit cold stream after passing through a throttling valve returns to the second heat exchanger (HX-2) as a cold stream and after heat exchanging goes to the first heat exchanger (HX-1). The outlet stream enters the first stage of compression. The tempera-ture of natural gas after two steps decreases to about -150 ºC and after throttling valve, LNG is produced at -161 ºC

Heavy hydrocarbons in natural gas can freeze at cryogenic temperatures in the liquefaction process. So, the amount of C_5^+ hydrocarbons should be decreased to less than 1000 ppmv be-fore LNG production. The separation of heavy hydrocarbons is done by condensation & grav-ity separation in the Heavies Separator (V-2) after pre-cooling of gas (typically @ -30 to -40 ºC) in the cold box. For this project, according to the feed gas composition the proper tem-perature for effective separation of C_5^+ hydrocarbons is -36 ºC based on simulation studies.

In this study, natural Gas (@ 25 ºC and 33 barg) including 86% methane, 5% nitrogen, 4% ethane and 5% C_3^+ has been used for 15 tons per day LNG production.

2.3. OPTIMIZATION PROBLEM

Power consumption in LNG processes is very significant, so the optimization methods are needed for minimizing operation costs. In order to optimize the energy consumption of these two liquefaction processes, GA method was used by connecting MATLAB to Aspen HYSYS process simulator. A MATLAB program code adapted from previous work (Moein et al., 2015; Moein et al., 2016) was developed to use GA optimization procedure on simulated pro-cesses in Aspen HYSYS Simulator. This MATLAB code calls the Aspen HYSYS simulation and transfers the data produced by GA to Aspen HYSYS. In this situation GA method acts as a controller and the Aspen HYSYS model is a server. The total required work of the pro-cess was considered as fitness function of GA method which should be minimized. Aspen HYSYS calculated the value of total required work in each generation and sent back to MATLAB to evaluate the fitness function value. So, there is a continuous linking between MATLAB optimization and HYSYS simulation.

As previously mentioned, the total required work of the process was defined as an objective function of the GA optimization method. The value of this objective function was calculated by Aspen HYSYS process simulator. Therefore, this objective function is a black box that contains an Aspen HYSYS model. The black box cannot provide gradient information or a reliable initial value. On the other hand, because of nonlinear property of mixed refrigerant processes, the objective function has multiple local minimum points. So, a global search meth-od needless derivative and an initial value is required to prevail these challenges and thermo-dynamic complexity of the process. In this research genetic algorithm as a global search meth-od is used to optimize the SMR processes. GA is a search heuristic that mimics the process of natural evolution, which was invented by Holland (1975) and further developed by his students and colleagues.

2.4. OPTIMIZATION FORMULATION

In this research minimizing of total required work of the process was defined as an objective function of GA method which expressed as:

$$Minimize\ f(X)=\sum W_{Compressors} \qquad (1)$$

In the above equation X is an adjusted variable vector including MR molar flow rates of components, the outlet pressure of MR compressor stages and the MR pressure after each throttling valve. Nitrogen, methane, ethane, propane and butane are used as the components of the mixed refrigerant. Other parameters such as natural gas composition and outlet LNG pressure and temperature is fixed during optimization. Thermodynamic properties of MR and NG were calculated by Peng-Robinson equation of state.

In this investigation three constraints were used for optimization of SMR processes. The first one is minimum approach temperature between hot and cold streams in plate-fin heat exchangers which should be greater than 2 ºC to satisfy reliability and feasibility of the process (ALPEMA, 2010). The second one is the temperature of MR in the inlet of each stages of compressor which should be greater than the dew temperature of the fluid in that pressure to prevent formation of liquid in the suction part of the compressors. The third one is pressure of those streams which will be mixed in the mixer that should be in the same pressure. These constraints are formulated as follow:

$$\Delta T_{min.HX-i} \geq 2\ ^{\circ}C$$

$$T_i \geq T_{dew,i}$$

$$P_{i,in} = P_{j,in} \qquad (for\ Mixers)$$

Where $\Delta T_{min.HX-i}$ represents the minimum approach temperature in heat exchanger i, T_i and $T_{dew,i}$ refer to operating temperature and dew point temperature of stream i and $P_{i,in}$ and $P_{j,in}$ illustrate the mixer pressure of inlet streams (i and j).

The last limitation was adjusted in Aspen HYSYS simulator while the first and second constraints are defined as the penalties for the objective function as shown below:

$$Minimize\ P(X.r) = f(x) + r\left(\sum_i [max\{0.g_i(x)\}]^2\right)$$

$$g_1(x) = 2 - \Delta T_{min,HX-i} \qquad (2)$$

$$g_2(x) = T_{dew,i} - T_i$$

where r is the penalty factor, assumed here as 10+14. If all the constraints are satisfied, the second term in the right-hand side of Eq. (2) would be zero ($P(X,r) = f(x)$). Otherwise, the second term will be a large value that GA modifies the penalty function in the next generation. The tuning parameters of GA are listed in Table 1.

Table 1. Tuning parameters of GA

Tuning parameter	Value
Population size	300
Selection method	Tournament
Tournament size	4
Mutation method	Adaptive feasible
Crossover fraction	0.8
Crossover function	Two-point
Stopping criteria:	
Maximum number of generations	200
Objective function tolerance	10^{-6}

3. RESULTS AND DISCUSSIONS

The result of optimized cases is shown in Table 2.

Table 2. Result of process optimization

Description	Case I	Case II
MR Pressure (kPa)	3870	3810
MR Flow Rate (kg/h)	5787	5223
	Methane: 32	Methane: 34
	Ethane: 29	Ethane: 27
MR composition (mole %)	Propane: 3	Propane: 2
	Butane: 26	Butane: 28
	Nitrogen: 10	Nitrogen: 9
Net Compression Power (kW)	319	309
Specific Power Consumption (SPC) (kWh/kg LNG)	0.51	0.49

The composite curve of optimized cases is shown in Figure 8 (a) & (b).

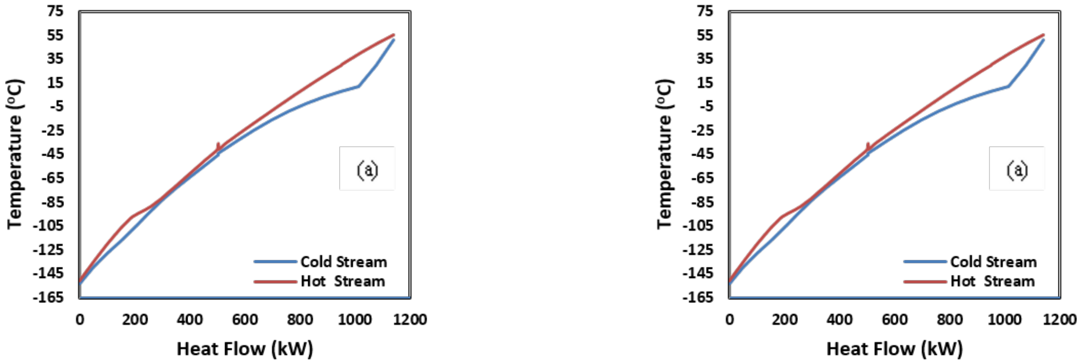


Figure 8. Composite curves of two heat exchangers in the cold box (a) Optimized Case I, (b) Opti-mized Case II

The results show that the "net compression power" of the Case I is only about 3.2% more than the Case II. However, given that the amount of LNG production in skid-mounted projects is low, the energy consumption is not important and the process simplicity has a higher priority. Furthermore, the fixed equipment cost of Case II is more than Case I due to the addition of a compression stage and air cooler in refrigeration cycle. Also, in commercial small scale lique-faction processes such as CB&I (Don Henry Coers & Jackie Wayne Sudduth, 1975) and Linde LIMUM-3 (Dr.-Carl-von, 2018) SMR processes, the mixing of low pressure MR streams has been done similar to process pattern of Case I.

3.1. SENSITIVITY ANALYSIS

Changing some operating parameters can affect the liquefaction process especially minimum approach temperature in PFHE. Some of these parameters are as follows:

1) Feed temperature
2) Feed pressure
3) Feed composition
4) MR composition

Changes in the feed temperature can be eliminated in treatment section or adjusted by chang-ing the amount of refrigerant flow rate. The feed pressure also can be set at inlet facility. However, the effect of unwanted feed gas temperature and pressure changes on minimum approach temperature of LNG heat exchanger is shown for Cases I and II in Figure 9 and Figure 10 in the range of 10 to 35 ºC and 30 to 40 barg, respectively.

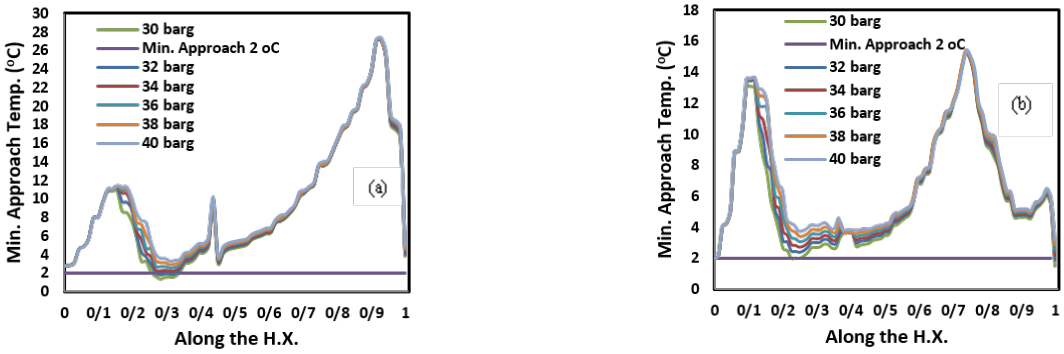


Figure 9. Effect of Feed Pressure on Min. Approach Temp. of (a) Case I, (b) Case II.

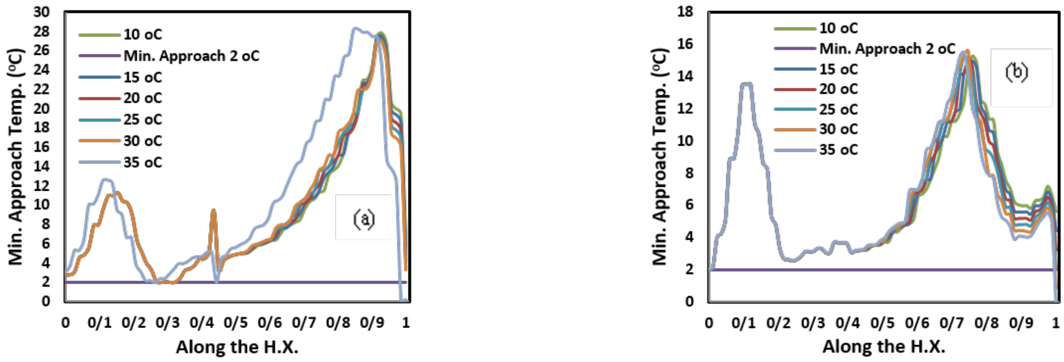


Figure 10. Effect of Feed Temp. on Min. Approach Temp. of (a) Case I, (b) Case II

As shown in Figure 9 (a), in Case I, by changing the pressure in the range of 30 to 40 barg, minimum approach temperature is less than 2 ºC at feed pressure between 30 to 33 barg. So Case I is sensitive to changes in feed pressure. But in Case II, by changing the pressure in the range of 30 to 40 barg, minimum approach temperature in all pressures, is equal or higher than 2. Therefore, Case II is not sensitive to changes in feed pressure.

Also, as shown in Figure 10 (b), Case II is more sensible to feed temperature changes, due to decrease the minimum approach temperature less than 2 ºC in the hot section of heat exchang-er.

Due to the different sources of gas supply in Iran, one of the most likely changes in specifica-tion of feed is the change in composition over the year. The effect of feed composition chang-es (methane content from 90% to 93%) on. minimum approach temperature of LNG heat ex-changer is shown in Figure 11 for Cases I and II.

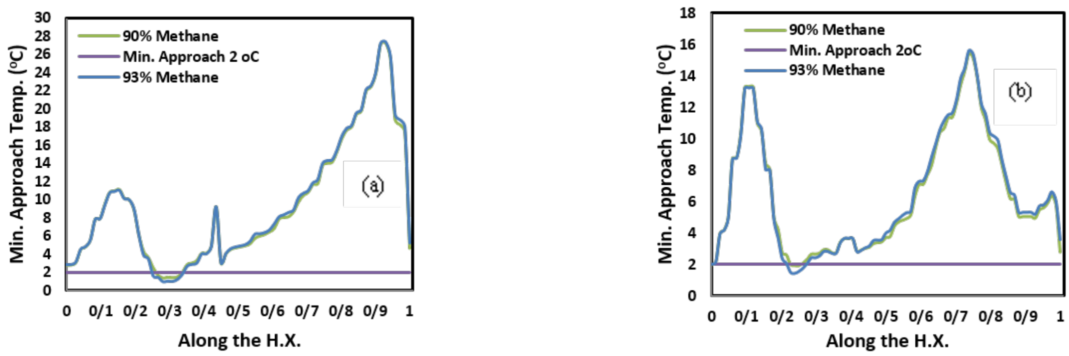


Figure 11. Effect of Feed Composition on Min. Approach Temp. of (a) Case I, (b) Case II

The initial mole fraction of methane in the feed gas is considered 86%. The variation range of methane content is considered from 86 to 93% mole according to maximum methane content in Iran gas pipelines. In Case I, by changing the concentration of methane in the mentioned range, minimum approach temperature is less than 2 ºC which indicates the sensitivity of this case to changes of methane content. In Case II by, changing the concentration of methane up to 90%, minimum approach temperature is equal or higher than 2 ºC and above 90%, minimum approach temperature is less than 2 ºC. So, Case I is more sensible to feed composition chang-es.

Due to leakage probability in the MR compressor, there is the possibility of changing the com-position of refrigerant during operation. This variation can be eliminated by make-up system to control MR composition. However, the effect of unwanted MR composition changes on min-imum approach temperature of LNG heat exchanger is shown in Figure 12, 13, 14, 15 and 16 for Cases I and II.

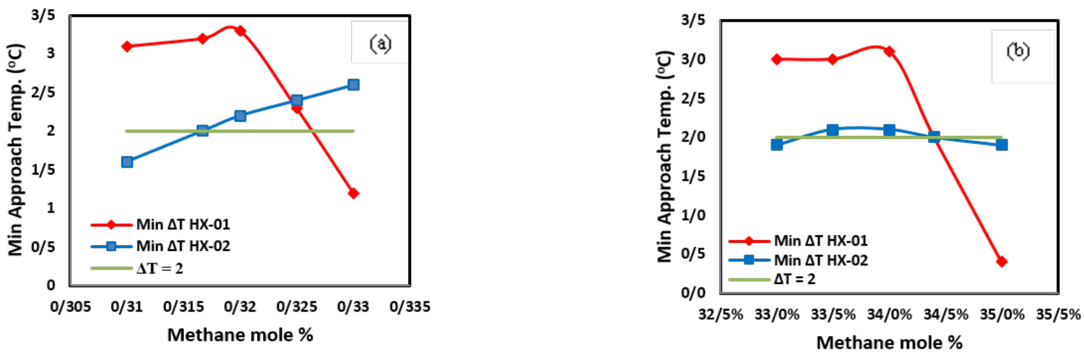


Figure 12. Min. Approach Temp. Sensitivity vs. Methane mole% of (a) Case I, (b) Case II.

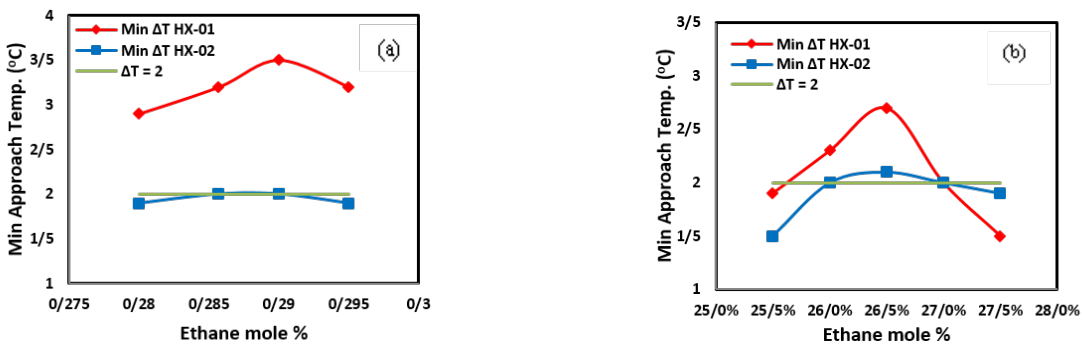


Figure 13. Min. Approach Temp. Sensitivity vs. Ethane mole% of (a) Case I, (b) Case II.

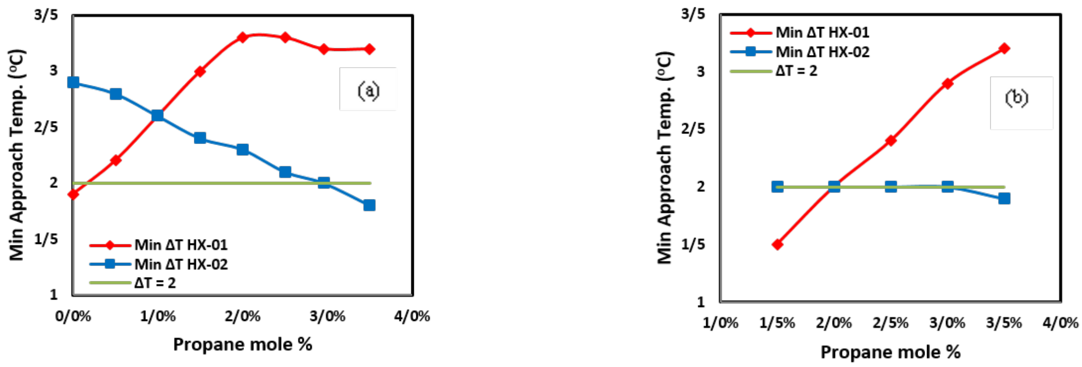


Figure 14. Min. Approach Temp. Sensitivity vs. Propane mole% of (a) Case I, (b) Case II.

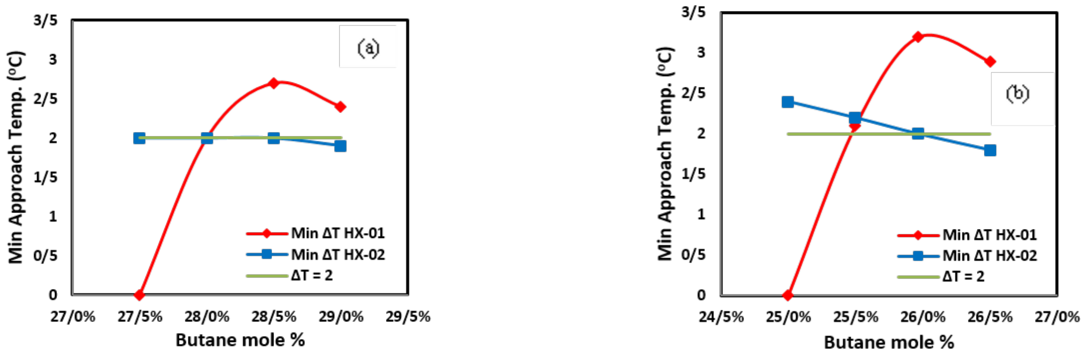


Figure 15. Min. Approach Temp. Sensitivity vs. Butane mole% of (a) Case I, (b) Case II.

Changes in the feed temperature can be eliminated in treatment section or adjusted by chang-ing the amount of refrigerant flow rate. The feed pressure also can be set at inlet facility. However, the effect of unwanted feed gas temperature and pressure changes on minimum approach temperature of LNG heat exchanger is shown for Cases I and II in Figure 9 and Figure 10 in the range of 10 to 35 oC and 30 to 40 barg, respectively.

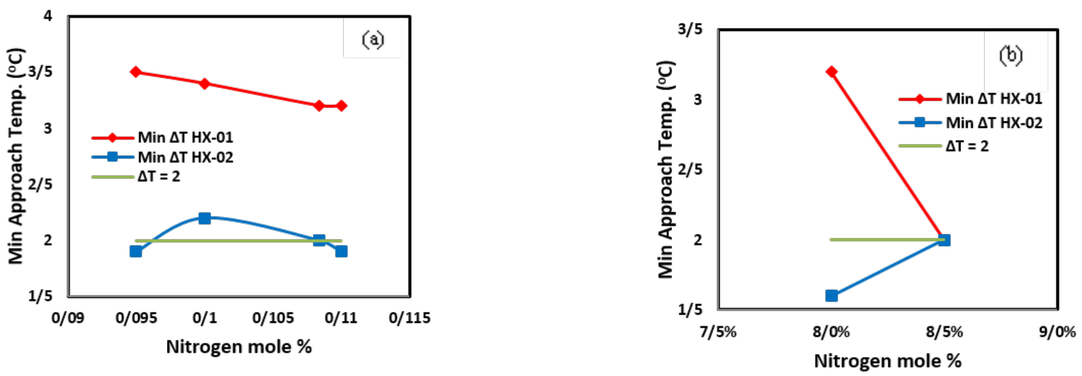


Figure 16. Min. Approach Temp. Sensitivity vs. Nitrogen mole% of (a) Case I, (b) Case II.

The results of MR sensitivity analysis is shown in Table 3. According to this table and Figures 12-16, Case II is more sensible to MR composition due to limited range of allowable changes in propane and nitrogen mole percent.

Table 3: MR composition Sensitivity

MR Component	Case I (mole %)	Case II (mole %)
Methane	31.0-33.0 (31.7)	33.0-35.0 (34.4)
Ethane	28.0-29.5 (28.6)	25.5-27.5 (27.0)
Propane	0.0-3.5 (3.0)	1.5-3.5 (2.0)
Butane	25.0-26.5 (26.0)	27.5-29.0 (28.0)
Nitrogen	9.5-11.0 (10.8)	8.0-8.5 (8.5)

The results of above sensitivity analysis are as follows:

- Case I is more sensible to feed pressure and composition changes
- Case II is more sensible to feed temperature and MR composition changes
- Propane content of MR fluid can be decreased to zero and 1% in Case I and II, re-spectively.

The comparison of two scenarios (Case I and II) shows that:

1) Energy consumption of Case I is slightly more than the Case II
2) Case II is more complicated than the Case I

According to above results, sensitivity analysis, and the small production rate of LNG in this project, the process simplicity is preferred to energy efficiency and the recommended process configuration is Case I.

4. ECONOMIC ANALYSIS

The preliminary cost estimation has been done for Case I and Case II. The equations which have been used for costs of purchased equipment are shown in table 4. The operating cost is calculated and compared by multiplying the power consumption (kW) of the compressors by working hours of a year by 3 cents per kWh as the electricity expense in Iran.

The cost data available from different sources are related to the past years and thus modification factors from Chemical Engineering Cost Index are used to update the costs of the reference year according to the following equation (Peters et al. 2003):

$$Cost_{\ reference\ year} = Cost_{\ original} \times \frac{Cost\ index_{\ reference\ year}}{Cost\ index_{\ original\ year}}$$

The comparison of economic results show that the purchased cost estimated for Case I is about 1,428,000 U$ whilst the purchased cost of Case II is estimated 1,716,000 U$. Also, the annual electricity cost for Case I & Case II is 66,000 U$ and 64,000 U$, respectively. i.e., the purchased cost for the Case II is 16.8% higher than the Case I. However, the electricity cost of for Case I is 3.1 % higher than the Case II.

Also, the total annualized cost of Case I & Case II is estimated by using of following equation (Towler et al., 2013):

$$Total\ Annualized\ Cost\ (TAC) = Capital\ cost \times \frac{i\ (i+1)^{n}}{(i+1)^{n\text{-}1}} + Operation\ cost$$

$$i = Discount\ Rate\ (8\%)$$

$$n = Plant\ Lifetime\ (20\ year)$$

The comparison of results show that the total annualized cost estimated for Case I is about 212,000 U$ whilst the total annualized cost of Case II is estimated 239,000 U$. So, the total annualized cost for Case II is 11.4% higher than the Case I.

Table 4. Purchased Cost of Equipment

Component	Purchased equipment cost expression
Plate fin heat exchanger	$C_{HX}=A_f \times \{C_a+C_b\ A_t^{c}\}$ (Sanaye et al., 2019, Mishra et al., 2004) C_{HX} : Heat exchanger cost ($) A_t : Heat transfer surface C_a = 30000 C_b = 750 A_f = 0.322 c = 0.8
Two-phase separator	$log_{10}^{Cspe} = k_1 + k_2 \log_{10}^{V} + k_3 [\log_{10}^{V}]^2$ (Turton et al., 2008) C_{spe} : Separator cost ($) V : Separator volume ($m^3$) k_1 = 3.4974 k_2 = 0.4485 k_3 = 0.074
Compressor	$C = a + bs^{n}$ (Towler et al., 2013) C : Cost of compressor ($) a = 580,000 b = 20,000 n = 0.6 s = Power of compressor (kW)
Cooler	$C_C=1.218k(1 + f_d + f_p)Q^{0.86}$ (Couper et al., 2005, Ghorbani et al., 2017) C_C : Cost of cooler (k$) Q : Duty (MBTU/h) k : Tube material f_d : Design type f_p : Design pressure

Abbreviations

GA	Genetic Algorithm
LNG	Liquefied Natural Gas
MR	Mixed Refrigerant
NG	Natural Gas
MRC	Mixed-Refrigerant Cycles
SMR	Single Mixed Refrigerant
TS	Tabu Search
NMDS	Nelder-Mead Downhill Simplex
SA	Simulated Annealing

5. CONCLUSION

Single mixed refrigerant process was studied in skid LNG technology. Because of process complexity of more than 1 phase separator in SMR loop, these types of processes are not suitable for skid-mounted packages. On the other hand, freezing possibility of heavy hydrocarbons content of refrigerant in the SMR process without phase separator and also the consideration of small foot print of these processes in skid-mounted packages, one phase separator SMR process has been selected for this design. With respect to skid design limitations (e.g. the minimum fixed and rotating equipment, minimum process complexity, minimum dimension and etc.), two process arrangements has been selected, simulated and optimized by coupling of Aspen HYSYS as the process simulator and MATLAB as the optimizing software. GA optimization method was chosen in order to make sure a global optimum condition would be reached. Key parameters were optimized and the unit energy consumption was minimized as an objective function. Also, the effect of pressure, temperature and composition of feed and MR composition changes on the process performance (minimum approach temperature of LNG heat exchangers) was investigated. The results showed that case I is more sensible to feed pressure and feed gas composition changes. Case II is more sensible to feed temperature and MR composition changes. Energy consumption of case I is slightly more than Case II. According to optimization results, sensitivity analysis, total annualized cost estimation & small production rate of LNG, the recommended process configuration is Case I.

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تولید LNG به روش مبرد آمیخته تک‌مرحله‌ای در مقیاس قابل‌حمل: انتخاب پیکربندی و آنالیز حساسیت

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چکیـــــــده

بررسی صاحبان فناوری مایع سازی گاز طبیعی در مقیاس قابل‌حمل، نشان می‌دهد که فرآیندهای مبرد آمیخته تک‌مرحله‌ای، سیکل انبساطی نیتروژن و فرآیندهای خود سرمایش، برای تولید گاز طبیعی مایع‌شده، استفاده‌شده است. در بین فرایندهای نامبرده شده، فرآیندهای مبرد آمیخته تک‌مرحله‌ای از بازده انرژی بالاتری برخوردار بوده و از تجهیزات دوار کمتری استفاده می‌کند. با توجه به مطالعات پژوهشگاه صنعت نفت در مورد فرآیندهای تجاری سیکل مبرد آمیخته تک‌مرحله‌ای و مطالعه بیش از ۱۰۰ اختراع در این رابطه، فرایند مبرد آمیخته تک‌مرحله‌ای با یک جداکننده فازی (با ۳۴٪ اشتراک در فرآیندهای مبرد آمیخته تک‌مرحله‌ای)، برای تولید گاز طبیعی مایع‌شده در مقیاس قابل‌حمل، انتخاب‌شده است. با توجه به پیچیدگی فرآیندهای مبرد آمیخته تک‌مرحله‌ای با چند جداکننده فازی، این نوع چرخه‌ها انتخاب‌نشده‌اند. از طرفی فرآیند مبرد آمیخته تک‌مرحله‌ای بدون جداکننده فازی به دلیل احتمال یخ‌زدگی هیدروکربن‌های سنگین موجود در مبرد، انتخاب‌نشده است.

چندین فرآیند مبرد آمیخته تک‌مرحله‌ای با یک جداکننده فازی بر اساس چیدمان تجهیزات در بخش‌های مایع سازی و سردسازی قابل‌استفاده است. بامطالعه گسترده و با توجه به محدودیت‌های طراحی در مقیاس قابل‌حمل (به‌عنوان‌مثال حداقل تعداد تجهیزات ثابت و دوار، حداقل پیچیدگی فرآیند و ابعاد و غیره)، دو آرایش فرآیندی در این مقاله انتخاب، شبیه‌سازی و بهینه‌سازی شده است. آنالیز حساسیت بر روی فشار و دمای خوراک و همچنین اجزاء مبرد آمیخته و ترکیب درصد خوراک انجام‌شده است. انرژی مصرفی این دو آرایش محاسبه‌شده و پیچیدگی آن‌ها با یکدیگر مقایسه شده است. با توجه به پایین بودن میزان تولید LNG و لزوم سادگی فرآیند در مقیاس قابل‌حمل و با توجه به نتایج اقتصادی بهترین گزینه توصیه‌شده است.

واژگان کلیدی: LNG، مبرد آمیخته تک‌مرحله‌ای، بهینه‌سازی، آنالیز حساسیت و اقتصادی، طراحی قابل‌حمل