Two-phase expander refrigeration cycles with ethane–nitrogen: A cost-efficient alternative LNG processes for offshore applications

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ABSTRACT

Nitrogen expander-based natural gas (NG) liquefaction processes are considered to be the most feasible and economic practices for offshore applications. The nitrogen single expander process is simple owing to its single-phase operation, and it employs high occupational safety and environment-friendly refrigerants. However, this process has low energy efficiency. Recent advancements in expansion devices show the possibility of handling two-phase refrigerant flow in an economical and simple manner. Accordingly, this paper proposes two-phase expansion using an innovative binary mixed refrigerant (MR) composed of ethane and nitrogen (C2N). Furthermore, a propane-precooling reflux cycle is also implemented and evaluated to analyze greater potential benefits of ethane–nitrogen refrigerant with two-phase expansion. To assess the technical and commercial feasibility of the proposed liquefaction process, energy, exergy, and economic (3E) analysis is performed. Using the C2N two-phase expander LNG process, 47.83% energy can be saved with 55.25% exergy destruction minimization and 24.12% total annualized costs (TAC) savings as compared to previously published nitrogen single expander process. Whereas, the C3-precooled C2N process gives higher energy savings i.e., 52.45% but low TAC savings i.e., 1.6% as compared to nitrogen single expander LNG process. Considering TAC savings, the C2N process (without propane-precooling) can be a promising candidate for offshore applications. © 2019 Elsevier Ltd. All rights reserved.

1. Introduction

The Industrial Revolution is responsible for a dramatic increase in air pollution. Since then, rising levels of air pollutants such as CO2, SOx, and particulates in the atmosphere have promoted climate change. In particular, large amounts of CO2, which leads to global warming, are emitted primarily from coal and oil-based power-generation industries (Qyyum et al., 2019a). During the past 100 years, the CO2 concentration levels in the atmosphere have increased rapidly, reaching approximately 404.34 ppm by the end of 2017 (Scripps Institution of Ocenography, 2017). It is estimated that by the end of this century, the global temperature will rise 2–5°C if CO2 emissions are not controlled (Ghaedi et al., 2017; Pachauri and Reisinger, 2007). Considering the present energy scenario along with environmental challenges, natural gas (NG) is acknowledged as one of the most suitable fuels among the various types of fossil fuels used to produce energy owing mainly to its low carbon emissions and its ability to meet strict environmental regulations. Shell has predicted that the demand for NG will increase 60% from 2010 to 2030 (Press Release December 2012, 2012); NG currently accounts for 23% of the total global energy consumption; this figure is expected to reach 26% by 2040 (ExxonMobil, 2014; Kuwahara et al., 2000).

However, NG reservoirs are usually found in offshore and remote areas, from which the product must be transported for trade in the global market; therefore, the locations of natural gas reserves play a vital role in the economics of transportation and storage (Strantzali et al., 2019). Natural gas can be either be transported through pipelines or liquefied and transported through cargo ships from remote areas. The transportation of natural gas in liquid form is considered to be more safe, feasible, and economical over long distances, particularly those >3500 km (Dutta et al., 2018; Horvath et al., 2018; Mehrpooya and Ansarinasab, 2015;
using the Extended Pinch Analysis and Design (ExPAnD) technique. Recently they used single-phase expander to produce LNG with lower power requirement costs depend on various parameters such as the refrigerant used in the refrigeration cycle and the feed NG, which leads to rapidly liquefaction and subcooling of NG to result in entropy generation. Most recently (Qyyum et al., 2018d), introduced two-phase single expander refrigeration cycle with propane and nitrogen (C3N) as a binary mixed refrigerant. They optimized the process using the particle swarm optimization technique to reduce the specific energy requirements.

It can be deduced from the literature that the major issue associated with the N2 expander LNG processes is the low energy efficiency (high exergy destruction), which ultimately leads to high total annualized costs (TAC), mainly due to the high operating costs. The high exergy destruction is attributed mainly due to the large temperature difference between the cooling temperature of the N2 refrigerant used in the refrigeration cycle and the feed NG, which leads to rapidly liquefaction and subcooling of NG to result in entropy generation. It is believed that the cooling of the NG before going to liquefaction and subcooling can be reduced the overall entropy generation, which ultimately will improve the overall energy efficiency. As reported previously (Gao et al., 2010; He and Ju, 2014; Mortazavi et al., 2012; Shah et al., 2009), exergy destruction can be reduced by modifying the N2 expander process through the involvement a precooling refrigeration cycle (e.g., C3/CO2/R410a precooling cycle) or through the addition of hydrocarbons (e.g., methane or propane) in the main refrigeration cycle. Previous research has shown that methane addition in the main refrigeration cycle (N2–CH4) increases the performance of the N2 expander process and C3-precooling cycle helped to further enhance the performance of the N2–CH4 cycle when a single-phase expander was used (Ding et al., 2016). However, recent innovations and improvements in expander technology facilitate an efficient handling of two-phase mixture. Furthermore, heat transfer characteristics of two-phase mixed hydrocarbon refrigerants flow boiling in shell and tube heat exchangers (Hu et al., 2018) as well as LNG heat exchangers (Hu et al., 2019) have also been investigated. Therefore, the high boiling hydrocarbon such as propane can be added to the main refrigeration cycle of nitrogen (C3N) as reported by (Qyyum et al., 2018d). They retrofitted a two-phase expander with a single-phase (gas) expander to handle a single mixed two-phase refrigerant, thus enhancing the process efficiency. Nevertheless, C3-precooling cycle is not a suitable option to integrate with C3N two-phase expander process, mainly due to the temperature distribution range of the precooling refrigerant (propane) that is already available with nitrogen in the main refrigeration loop. Therefore, new binary MRs are required to realize the benefits of the propane precooling refrigeration cycle integrated with two-phase expander LNG process. Ethane, with a normal boiling point (NBP) of ~89.0 °C, can be one of the possible candidates to make combination with nitrogen in two-phase expander single loop.

Hence, in this study, we propose a binary MR of ethane–nitrogen that can work with newly developed cryogenic two-phase expander. To achieve the maximum benefits of the proposed ethane–nitrogen refrigerant, a propane precooling

**Mehroooya and Chorbani, 2018).** However, NG liquefaction is an energy- and cost-intensive process; the liquefaction sections alone take up approximately 40%–50% of the total expenditure of the liquefied natural gas (LNG) value chain (Lim et al., 2013; Mehrpooya and Ansarinasa, 2015; Vatani et al., 2014). Liquefaction and energy requirement costs depend on various parameters such as the liquefaction technology used (Qyyum et al., 2018c) and the environmental conditions at the plant site (Qyyum et al., 2018b).

The most widely used liquefaction technologies are nitrogen (N2) expander-based LNG and mixed refrigerant (MR)-based processes, the latter of which is more economically feasible owing to its lower energy consumption and therefore its lower operating costs (Yin et al., 2008). However, the major drawbacks of the MR process are its high degree of complexity, high capital investment, and involvement of flammable hydrocarbons as a refrigerant, making them less attractive for offshore applications (Qyyum et al., 2018c). On the contrary, the N2 expander-based LNG process ensures high occupational safety, simple and user-friendly operability, and easy availability and portability. Further, its low capital investment makes it the best choice for offshore operations. However, high energy consumption is the main limitation associated with nitrogen expander-based liquefaction technologies (Qyyum et al., 2019b).

Therefore, many researchers have proposed different approaches using either refrigeration cycle retrofitting and optimization or only optimization to minimize the energy consumption of nitrogen expander liquefaction processes. For example (Cao et al., 2006), introduced a binary mixed refrigerant N2–CH4, however, they used single-phase expander to produce LNG with lower power consumption than conventional N2 expander process. Recently (Haider et al., 2019), also used N2–CH4 expander liquefaction process to produce liquefied biomethane followed by biogas upgrading. Gao et al. (2010) employed the propane pre-cooling refrigeration cycle to improve the energy efficiency of the N2 expander process for the liquefaction of coalbed methane. He and Ju (2014) reduced the operating costs of an N2 expander process by using R410a and propane for precooling, which reduced the energy consumption by 20% and 22.74%, respectively, compared with the conventional nitrogen expander process. (Aspelund et al., 2007) improved the operational performance and reduced the overall operational cost of the CO2 precooled N2 expander process using the Extended Pinch Analysis and Design (ExPAnD) technique. Yuan et al. (2014) also used CO2 as a precooling refrigerant to enhance the N2 expander process. They optimized the process to achieve a liquefaction rate of 77% with a unit energy consumption of 9.90 kWh/kmol. Shah et al. (2009) used non-dominated sorting genetic algorithm (NSGA-II) to optimize the total shaft work, capital cost, and annualized cost for propane precooled dual N2 expander LNG process. Khan et al. (2014) enhanced the performance of the N2 expander process using the CO2 precooling cycle for offshore applications. They optimized the process using knowledge-inspired approach and reduced the compression power by 15.8%. Ding et al. (2016) developed a model of the N2 expander process in Aspen Hysys® simulation environment and optimized the model using the genetic algorithm (GA). In addition, they also studied the effect of propane precooling corresponding to overall energy savings. Therefore, many researchers have proposed different approaches using either refrigeration cycle retrofitting and optimization or only optimization to minimize the energy consumption of nitrogen expander liquefaction processes. For example (Cao et al., 2006), introduced a binary mixed refrigerant N2–CH4, however, they used single-phase expander to produce LNG with lower power consumption than conventional N2 expander process. Recently (Haider et al., 2019), also used N2–CH4 expander liquefaction process to produce liquefied biomethane followed by biogas upgrading. Gao et al. (2010) employed the propane pre-cooling refrigeration cycle to improve the energy efficiency of the N2 expander process for the liquefaction of coalbed methane. He and Ju (2014) reduced the operating costs of an N2 expander process by using R410a and propane for precooling, which reduced the energy consumption by 20% and 22.74%, respectively, compared with the conventional nitrogen expander process. (Aspelund et al., 2007) improved the operational performance and reduced the overall operational cost of the CO2 precooled N2 expander process using the Extended Pinch Analysis and Design (ExPAnD) technique. Yuan et al. (2014) also used CO2 as a precooling refrigerant to enhance the N2 expander process. They optimized the process to achieve a liquefaction rate of 77% with a unit energy consumption of 9.90 kWh/kmol. Shah et al. (2009) used non-dominated sorting genetic algorithm (NSGA-II) to optimize the total shaft work, capital cost, and annualized cost for propane precooled dual N2 expander LNG process. Khan et al. (2014) enhanced the performance of the N2 expander process using the CO2 precooling cycle for offshore applications. They optimized the process using knowledge-inspired approach and reduced the compression power by 15.8%. Ding et al. (2016) developed a model of the N2 expander process in Aspen Hysys® simulation environment and optimized the model using the genetic algorithm (GA). In addition, they also studied the effect of propane precooling corresponding to overall energy savings. Most recently (Qyyum et al., 2018d), introduced two-phase single expander refrigeration cycle with propane and nitrogen (C3N) as a binary mixed refrigerant. They optimized the process using the particle swarm optimization technique to reduce the specific energy requirements.

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refrigeration cycle is also integrated with ethane–nitrogen two-phase single expander liquefaction process. The proposed processes are optimized using the modified coordinate descent (MCD) optimization technique to reduce the overall energy consumption. To evaluate the technical and commercial feasibility of the proposed processes, a detailed exergy and economic analysis is also performed. Further, the performance of the proposed optimized processes are compared with previously published nitrogen expander-based liquefaction processes. This article is organized as follows. Section 2 describes the main motivation and theory about the proposed ethane–nitrogen refrigeration cycles and propane precooling-assisted ethane–nitrogen refrigeration cycles for FLNG projects. Section 2.1 provides the detailed process description of the proposed liquefaction processes. Section 2.2 includes the feed NG composition, conditions, and simulation assumptions. Section 2.3 consists of process optimization. Section 3 includes the process analysis in terms of energy, exergy, and economic. Finally, the conclusions drawn from this study are presented in Section 4.

2. Propane precooling-assisted ethane–nitrogen refrigeration cycles

For efficient NG liquefaction, it is important to follow the thermodynamic pattern of gradual cooling, liquefaction, and subcooling. A high temperature gradient between the selected refrigerant and feed gas gives a rapid liquefaction and subcooling of the feed gas, resulting exergy destruction. This exergy destruction is the main reason for the low performance of the N2 expander process when compared with the MR liquefaction process. For example, the expander cycle in the N2 expander liquefaction process uses N2, which has a boiling point of −195 °C (1 kJ/mol specific refrigeration effect against 246.0 bar pressure), in the main refrigeration cycle, and feed NG is normally introduced at 30 °C (Khan et al., 2015a, 2014). In this case, a large difference is present between the temperatures of the introduced feed gas and the boiling point of the refrigerant in the refrigeration cycle. Therefore, the process undergoes liquefaction and subcooling rapidly by exchanging the only sensible heat of nitrogen with feed NG, which results in high exergy destruction. To avoid this exergy destruction, (Qyyum et al., 2018d) added an optimal flow rate of propane in the nitrogen refrigeration cycle to incorporate the precooling phenomenon. However, when propane was mixed with nitrogen, this mixture was appeared as two-phase, thus, two-phase expander was introduced. In the current investigation, ethane is mixed with nitrogen (C2N) to form a single MR for the main refrigeration cycle. The boiling point of ethane is significantly lower than that of propane (i.e., −89.0 °C) and it gives 1 kJ/mol specific refrigeration effect with significant lower pressure i.e., 16.5 bar. Therefore, ethane is preferred for the main loop, whereas propane is used inside the precooling loop owing to its strong performance as a precooling refrigerant (Khan et al., 2015a; Qyyum et al., 2018d). The reason for using propane inside the precooling refrigeration cycle can be attributed to its effective precooling properties. If used inside the main loop, propane provides a precooling effect inside the loop. However, it cannot enhance the cycle efficiency because its distribution range differs significantly from that of nitrogen; thus, C3-precooling for the C3N cycle would not have a significant effect. Therefore, ethane was mixed with nitrogen rather than propane. The optimal parameters (e.g., temperature, pressure, and flow rate) for each stream are used to explain the process description of the proposed processes.

In the proposed C2N process, the refrigerant (C2N) mixture (stream 1) is compressed in a four-stage compression cycle with interstage cooling in the pressure range of 4.78 bar–38.39 bar (stream 9), which is found by the optimization algorithm. The interstage coolers help to maintain the temperature of the propane and C2N mixture at 30 °C. A compression ratio of 1.3 is maintained in the compressors to favor reversibility. This is because any irreversibility generated in the process causes an overall increase in exergy destruction (Tsatsaronis and Morosuk, 2012). The streams of feed NG (stream A) and C2N (stream 10) are then passed through the LNG exchanger (LNG-100); the two-phase expander expands stream 10 through a high-pressure two-phase C2N to 4.88 bar (stream 11). This produces sufficient cooling for LNG-100 by promoting heat exchange between the warm NG and C2N streams and provides adequate cooling to liquefy the NG and to partially liquefy the C2N refrigerant. The exiting stream 12 from the heat exchangers is in the form of superheated vapors that are recycled to complete the refrigeration cycle. Similarly, in the second proposed process, the feed NG and refrigerant (C2N) are cooled to about 16 °C in a C3-precooling cycle. This is in addition to a C2N refrigeration cycle that condenses and subcools NG.

In the precooling refrigeration cycle, propane (stream 1-1) is compressed to an optimum pressure of 11.05 bar (stream 1–6) using three compression stages occurring in the interstage cooling system. Similarly, the C2N mixture (stream 1) is compressed in a four-stage compression cycle that includes interstage cooling from a pressure of 4.78 bar–38.39 bar (stream 9). The interstage coolers maintain the temperatures of the propane and C2N mixture at 30 °C. The compressed propane is expanded through a Joule–Thomson (JT) valve to reduce its pressure to 2.48 bar and temperature to −19.48 °C before it is introduced into the LNG exchanger, LNG-101, where it exchanges heat with the feed NG and compressed C2N and precools them both to −16.48 °C.

The precooled streams of feed NG (stream B) and C2N (stream 11) are then passed through the LNG-100 exchanger; the high-pressure two-phase stream of C2N is expanded in the two-phase expander to 4.88 bar to produce sufficient cooling for the LNG-100. This occurs through the heat exchange from the warm NG and C2N streams, thus providing adequate cooling to liquefy the NG and to partially liquefy the C2N refrigerant. The exiting streams (stream 1-1) and (stream 13) from the heat exchangers (LNG-101 and LNG-100) are in the form of superheated vapors that are recycled to complete the refrigeration cycle. Stream C exits LNG-100 as a subcooled liquid at a pressure of 50 bar. For economically feasible as well as safe storage and transportation of LNG this high pressure is not suitable. Therefore, stream C is passed through LNG turbine (Turb-1) to obtain LNG at −158.5 °C and a pressure which is slightly higher than the atmospheric pressure (1.2 bar was fixed in this study). LNG is produced at a liquefaction rate of 92% with 8% end flash gas. The produced LNG is then stored in storage tanks.

2.2. Process simulation

The proposed LNG processes were simulated using Aspen Hysys® version 10 software. The Peng–Robinson equation of state was chosen along with the Lee–Kesler equation for rigorous calculations of thermodynamic properties (Qadeer et al., 2018). It should be noted that the Lee–Kesler model is the precise standard used to calculate the enthalpy of gases at high pressures (Li et al., 2012; Yuan et al., 2015). The essential process feed conditions are shown in Table 1. Additionally, for process simulation, the following presumptions were incorporated:
The pressure drop (ΔP) in all inter-stage coolers was set at 0.25 bar.

The isentropic efficiencies of the compressor, two-phase expander, and LNG turbine were selected as 75%, 80%, and 90%, respectively.

The temperature of cooling medium (water/air) for inter-stage cooling was taken to be 20°C.

The pressure of the end flash gas was set at 1.2 bar.

The minimum internal temperature approach (MITA) was selected as 3.0°C in the main cryogenic heat exchanger.

All assumptions were taken from the recent research (Abdul Qyyum et al., 2018; Khan et al., 2015b; Qyyum et al., 2018d) of nitrogen expander-based LNG processes.

2.3. Process optimization

The non-optimal execution of design variables contribute toward significant exergy destruction, which results in low energy efficiency of the process. Suitable optimization techniques can result in large amounts of energy savings for a given LNG process. Alteration by adding or removing new devices in existing LNG processes can also change the optimal operation parameters of a given liquefaction process.

The key design parameters influencing process efficiency such as the refrigerant flow rate in refrigeration cycles, refrigerant

Table 1
Feed composition, conditions, and basis for process simulation.

<table>
<thead>
<tr>
<th>Feed composition</th>
<th>Mole (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>91.28</td>
</tr>
<tr>
<td>Ethane</td>
<td>5.40</td>
</tr>
<tr>
<td>Propane</td>
<td>2.10</td>
</tr>
<tr>
<td>i-Butane</td>
<td>0.50</td>
</tr>
<tr>
<td>n-Butane</td>
<td>0.50</td>
</tr>
<tr>
<td>i-Pentane</td>
<td>0.01</td>
</tr>
<tr>
<td>n-Pentane</td>
<td>0.01</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.20</td>
</tr>
<tr>
<td>After-cooler outlet temperature (°C)</td>
<td>30.00</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Natural gas feed conditions</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature (°C)</td>
<td>30.00</td>
</tr>
<tr>
<td>Pressure (bar)</td>
<td>50.00</td>
</tr>
<tr>
<td>Flow rate (kg/h)</td>
<td>1.00</td>
</tr>
<tr>
<td>Pressure drop (ΔP) across the main LNG cryogenic heat exchanger</td>
<td></td>
</tr>
<tr>
<td>“Stream-5” to “Stream-10”</td>
<td>1.0 bar (hot stream)</td>
</tr>
<tr>
<td>“Stream-11” to “Stream-12”</td>
<td>0.1 bar (cold stream)</td>
</tr>
<tr>
<td>“Stream-15” to “Stream-16”</td>
<td>1.0 bar (hot stream)</td>
</tr>
</tbody>
</table>

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evaporation and condensation pressures in both cycles, and precooling and subcooling temperatures need to be optimized to obtain the optimum design variables. For the design optimization of the proposed LNG processes, the design variables were chosen based on a deep knowledge-inspired based procedure as followed by (Khan et al., 2013; Pham et al., 2017) and are shown in Table 2 with the corresponding upper and lower limits for the proposed processes.

The optimization objective was to minimize the total power input of the compression cycles (C2N and C3) while constraining the MITA value to 3°C in the LNG-100 and LNG-101 heat exchangers, respectively. The objective function is represented by Eq. (1):

$$\text{Min} \{ X \} = \text{Min} \left( \sum_{j=1}^{n} \frac{W_j}{m_{\text{NG}}^j} \right)$$

(1)

which is subject to the conditions

$$\Delta T_{\text{min}}(X) \geq 3$$

(2)

and

$$X_{lb} < X < X_{ub}$$

(3)

where X is a vector of decision variables given as

$$X = \left( P_1, P_2, T, C_3P_1, C_3P_2, m_{C2}, m_{C3}, m_{N2} \right)$$

The constrained objective function and design variables undergo mutual non-linear interaction, which makes it difficult to utilize commonly available simulation software for optimization. Consequently, the process model in Aspen Hysys® should be linked with an externally available optimization technique. The MCD algorithm (Park et al., 2015; Qyyum et al., 2018a) was employed to optimize the proposed processes. Microsoft Visual Studio was utilized to develop MCD algorithm based optimizer and was then linked to Aspen Hysys®. The MCD optimization method has been used successfully to optimize highly non-linear and complex systems (Long et al., 2016; Qyyum et al., 2018a). Therefore, it is suitable for interactive and highly non-linear optimization problems such as LNG processes. Fig. 3 presents the working mechanism of the MCD optimization strategy can be found in previous research (Park et al., 2015).

3. Energy, exergy, and economic analysis

This section provides a detailed analysis of the proposed liquefaction processes with respect to energy, exergy, and economic evaluation (3Es). Furthermore, the optimal pressures and temperatures for all streams of the proposed C2N two-phase expander process and the C3-precooled C2N two-phase expander process are included in Supporting Information as Table S1 and Table S2, respectively.

3.1. Energy analysis

Energy analysis was conducted with the optimal design variables for the proposed C2N and C3-precooled C2N two-phase expander processes, and the results were compared with those of previously published nitrogen single, dual, and C3N two-phase expander processes (Table 3).

According to Table 3, the proposed C2N and C3-precooled C2N processes yielded 33.30% and 15.61% lower mass flow rates, respectively, in the refrigeration cycle. In the case of N2 only, the mass flow rate in the single expander process was 8.25 kg·h⁻¹, whereas those of the proposed C2N and C3-precooled C2N processes were 1.515 kg·h⁻¹ and 1.502 kg·h⁻¹, respectively. The evaporation pressure of the refrigerant was also lower for the proposed C2N and C3-precooled C2N processes, at 85.77 bar and 38.14 bar, respectively. This advantage regarding reduced energy consumption is not possible in conventional processes. In particular, the reduction in the amount of N2 and evaporation pressure of the refrigerant controls the capital cost as well as the operating costs as it affects the compressors’ power consumption. The compression ratio in the N2 single expander process was 2.12, whereas the compression ratios in the proposed C2N and C3-precooled C2N processes were 1.857 and 1.680, respectively. A low compression ratio is crucial for all refrigeration cycles because it increases the refrigerant performance in NG liquefaction.

Previously reported N2 expander processes have shown low energy efficiency with a 92% liquefaction rate, whereas the proposed processes exhibit a lower power demand at the same liquefaction rate. Table 4 compares the specific power consumption values of the N2 expander-based LNG production processes against the proposed C2N and C3-precooled C2N two-phase expander-based processes while neglecting their different feed conditions but keeping a constant liquefaction rate of 92% respectively. The comprehensive specific power consumptions of the proposed C2N and C3-precooled C2N processes were 79.72% and 81.52%, when compared to the conventional N2 single expander-based LNG process demonstrated by (Du et al., 2010). When compared to the N2 single expander-based LNG process developed by (Austbe and Gundersen, 2015), the proposed processes lowered the energy consumption by 51.16% and 55.49%, respectively. Although, the energy savings were relatively less when compared to the dual expander process (Khan et al., 2013), it still resulted in major reductions i.e. 20.83% and 27.85% respectively. A comparison with the C3N two-phase expander process proposed in previous research (Qyyum et al., 2018d) showed that the proposed processes yielded energy savings of 2.46% and 11.11%, respectively. In these cases, the power consumption depended specifically on the NG feed composition, liquefaction rate, and operating conditions.

3.2. Exergy analysis

Exergy analysis is used to identify the irreversibilities within

### Table 2

<table>
<thead>
<tr>
<th>Decision Variables</th>
<th>Lower Limit</th>
<th>Upper Limit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Refrigerant high pressure, $P_1$ (bar) (stream-9)</td>
<td>25.00</td>
<td>87.00</td>
</tr>
<tr>
<td>Refrigerant low pressure, $P_2$ (bar) (stream-1)</td>
<td>3.0</td>
<td>10.00</td>
</tr>
<tr>
<td>Nitrogen mass flow rate, $m_{N2}$ (kg/h)</td>
<td>1.50</td>
<td>10.00</td>
</tr>
<tr>
<td>Ethane mass flow rate, $m_{C2}$ (kg/h)</td>
<td>0.50</td>
<td>4.50</td>
</tr>
<tr>
<td>Precooling refrigerant high pressure, $C_3P_1$ (bar) (stream-1-6)</td>
<td>8.00</td>
<td>40.0</td>
</tr>
<tr>
<td>Precooling refrigerant low pressure, $C_3P_2$ (bar) (stream-1-1)</td>
<td>2.00</td>
<td>15.00</td>
</tr>
<tr>
<td>Precooling Temperature, $T$ (°C)</td>
<td>–60.0</td>
<td>–20.0</td>
</tr>
<tr>
<td>Propane flow rate, $m_{C3}$ (kg/h)</td>
<td>3.00</td>
<td>10.0</td>
</tr>
</tbody>
</table>

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each system component. By performing an exergy analysis, we can find the location, magnitude and causes of process inefficiencies. This technique introduces single unit operations, which represent a large amount of lost work; it thus helps to identify valuable information for process improvement from the design and operations perspectives. Generally, the exergy of any stream constitutes of physical, chemical and mechanical; whereby the mechanical exergy can be neglected due to its small involvement (Qyyum et al., 2018a). Moreover, due to the absence of any chemical reaction in the process, the chemical exergy is also taken as zero (Qyyum et al., 2018a; Tsatsaronis and Morosuk, 2012). The physical exergy can then be expressed as:

\[
Ex = (h - h_0) - T_0(s - s_0).
\]  

In this study, exergy analysis was performed for each equipment associated with previous published nitrogen expander-based and proposed LNG processes. The equations used to calculate exergy destruction were sourced from previous studies (Venkatarathnam and Timmerhaus, 2008) and are given below as Eq. (5) to (8). For equations (5) to (8), the values for \(Ex_{in}\) and \(Ex_{out}\) were taken directly from the stream properties calculated by Aspen Hysys v10. The Aspen Hysys v10 estimate the exergy of each stream assuming standard conditions as; \(T_0 = 25.0^\circ C\) and \(P_0 = 1.013\) bar.

The equation for exergy destruction calculation in the compressor is given in Eq. (5).
The exergy destruction analysis were performed for the proposed LNG liquefaction processes in comparison with previous published N2 single expander (Khan et al., 2015a), N2 dual expander (Qyyum et al., 2018d) and proposed C3n single expander (Qyyum et al., 2018d). The equation for exergy destruction calculation in the after-coolers is given in Eq. (7).

\[ \text{Ex}_{\text{destruction, after-cooler}} = (m)(\text{Ex}_{\text{in}} - \text{Ex}_{\text{out}}) + W \]  

The equation for exergy destruction calculation in the LNG heat exchanger is given in Eq. (8).

\[ \text{Ex}_{\text{destruction, LNG exchanger}} = \sum (m)(\text{Ex}_{\text{in}} - \text{Ex}_{\text{out}}) \]  

The exergy destruction analysis were performed for the proposed liquefaction processes in comparison with previous published N2 single expander (Khan et al., 2015a), N2 dual expander (Khan et al., 2015a), and C3N single expander (Qyyum et al., 2018d). As shown in the figure, approximately 75% exergy destruction occurred in the compression section of the N2 single expander process, whereas that in the compression section of all other processes was below 40%. In the inter-stage coolers and expander, the N2 single expander and C3N single expander processes exhibited major exergy destruction of more than 20%, whereas the expander section in the N2 dual expander process also exhibited exergy destruction of more than 20%. With respect to the expander section, the second-highest exergy destruction was observed in the C3N single expander process, whereas other processes exhibited less than 20% exergy destruction. However, in the LNG exchanger section, the highest exergy destruction occurred in the proposed processes, whereas the commercial processes exhibited lower exergy destruction. Similarly, Fig. 5 displays the overall exergy destruction occurring in various processes. The conventional N2 single expander process (Khan et al., 2015a) shows a total exergy destruction of 2298.52 kW, whereas the proposed C3N two-phase expander and C3-precooled C2N two-phase expander-based LNG processes exhibited total exergy destruction values of 1028.52 kW and 946.23 kW with 55.25% and 58.83% individual decrease in energy efficiency, respectively. These values suggest that energy efficiency can be improved in the proposed and commercially available processes either by optimization or by retrofitting the existing processes.
the refrigeration cost. Hence, the THCC composite curves play an important role in the energy efficiency analysis of any liquefaction process. The high energy consumption in the N₂ single expander process is attributed to the large temperature difference (>3°C) between −149°C and 30°C in the composite curves as shown in Fig. 6(a). Comparatively, the CₐN process depicts a smaller gap between the hot and cold composite curves between −116°C and −67°C representing a lower degree of irreversibility corresponding to the beneficial addition of ethane as a refrigerant. However, since the peak approach temperature for CₐN is comparatively higher than the N₂ expander based process TDCC, it presents opportunity for further energy enhancement. For this, the C₃-precooled C₂N TDCC shows smaller gap along with a peak approach temperature lower than both Fig. 6(a) and (b). Therefore, Fig. 6(c) represents a case with lowest entropy generation and highest energy savings. From a thermodynamic perspective, the liquefaction process is referred to as an economical process as long as the THCC curves are kept at the lowest gap possible. Comparatively, the approach temperature (TDCC) must be minimal throughout the cryogenic heat exchanger. The value of MITA must be within 1°C–3°C for effective and modest heat transfer (Hasan et al., 2009). In this context, the value of MITA was kept at 3°C, and the value of the approach temperature in the heat exchanger for the N₂ expander process met the desired values at the inlets of the heat exchanger. Fig. 6(a) illustrates the changes in the MITA values of this process; the highest MITA value was 50°C. Comparatively, in the proposed C₂N and C₃-precooled C₂N processes, as presented in Fig. 6(c) and (e), respectively, the approach temperature was satisfied at both ends of the exchanger and for a wide range of temperatures inside the exchanger, at −70°C to −130°C. This phenomenon is attributed to the presence of ethane in the refrigeration cycle. Furthermore, the TDCC curves of the proposed liquefaction processes in comparison with conventional N₂ single expander process are shown in Fig. 7. Accordingly, for C₃-precooled C₂N process, the height of TDCC throughout the heat exchanger’s length is lower than that of N₂ single expander process and the proposed C₂N process. The significant lowering (especially in the encircled region of Fig. 7) in heights of TDCC curves proves the overall high energy-efficiency of the proposed liquefaction processes as compared to N₂ single expander LNG process.

In this study, the gap between the hot and cold composite curves of the proposed C₂N and C₃-precooled C₂N processes (Fig. 6(d) and (f)) was reduced by MCD optimization, resulting in an overall reduction in power consumption of up to 47.83% and 52.45%, respectively. Moreover, the THCC composite curves of the C₂N system, as depicted in Fig. 6(d), indicate that the process still contains gaps, which implies a further potential for energy savings by applying highly efficient and rigorous optimization methods.

Fig. 8 shows the THCC and TDCC curves of the precooling exchanger in the propane refrigeration cycle and the main cryogenic heat exchanger. The TDCC curves in Fig. 8(a) show that in the precooiling exchanger, ΔT is large between the hot and cold composite curves when compared with the composite curves of the main LNG exchanger in Fig. 8(c). The gap between the THCC hot and cold composite lines in Fig. 8(b) and (d) indicate that the process has further potential for energy savings not only in the precooling exchanger but also in the main cryogenic exchanger.

### 3.3. Energy and exergy analysis through composite curves

Fig. 6 compares the conventional N₂ expander and proposed processes in terms of composite curve analysis. Fig. 6(a), (b), and (c) show the temperature difference between composite curves (TDCC) in the main LNG exchanger for the N₂ single expander process and the proposed C₂N and C₃-precooled C₂N processes. Fig. 6(d), (e), and (f) demonstrate the temperature–heat flow composite curves (THCC) in the main LNG exchanger of the N₂ single expander and the proposed C₂N and C₃-precooled C₂N processes.

The gap between the THCC curves is a measure of the entropy generation in the cryogenic heat exchanger, which in turn increases the refrigeration cost. Hence, the THCC composite curves play an important role in the energy efficiency analysis of any liquefaction process. The high energy consumption in the N₂ single expander process is attributed to the large temperature difference (>3°C) between −149°C and 30°C in the composite curves as shown in Fig. 6(a). Comparatively, the CₐN process depicts a smaller gap between the hot and cold composite curves between −116°C and −67°C representing a lower degree of irreversibility corresponding to the beneficial addition of ethane as a refrigerant. However, since the peak approach temperature for CₐN is comparatively higher than the N₂ expander based process TDCC, it presents opportunity for further energy enhancement. For this, the C₃-precooled C₂N TDCC shows smaller gap along with a peak approach temperature lower than both Fig. 6(a) and (b). Therefore, Fig. 6(c) represents a case with lowest entropy generation and highest energy savings. From a thermodynamic perspective, the liquefaction process is referred to as an economical process as long as the THCC curves are kept at the lowest gap possible. Comparatively, the approach temperature (TDCC) must be minimal throughout the cryogenic heat exchanger. The value of MITA must be within 1°C–3°C for effective and modest heat transfer (Hasan et al., 2009). In this context, the value of MITA was kept at 3°C, and the value of the approach temperature in the heat exchanger for the N₂ expander process met the desired values at the inlets of the heat exchanger. Fig. 6(a) illustrates the changes in the MITA values of this process; the highest MITA value was 50°C. Comparatively, in the proposed C₂N and C₃-precooled C₂N processes, as presented in Fig. 6(c) and (e), respectively, the approach temperature was satisfied at both ends of the exchanger and for a wide range of temperatures inside the exchanger, at −70°C to −130°C. This phenomenon is attributed to the presence of ethane in the refrigeration cycle. Furthermore, the TDCC curves of the proposed liquefaction processes in comparison with conventional N₂ single expander process are shown in Fig. 7. Accordingly, for C₃-precooled C₂N process, the height of TDCC throughout the heat exchanger’s length is lower than that of N₂ single expander process and the proposed C₂N process. The significant lowering (especially in the encircled region of Fig. 7) in heights of TDCC curves proves the overall high energy-efficiency of the proposed liquefaction processes as compared to N₂ single expander LNG process.

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### 3.4. Economic evaluation

The module costing technique was incorporated for economic evaluation of the proposed configurations for NG liquefaction. This technique was broadly incorporated for the preliminary cost estimation of chemical plants (Nagy, 2016; Turton et al., 2008). On the basis of this approach, the equipment purchase cost was calculated...
using Eq. (9):

$$
\log_{10}(E_p) = k_1 + k_2 \log_{10} A + k_3 (\log_{10} A)^2
$$

where $E_p$ is the equipment purchase cost, and $A$ is the capacity parameter, i.e., fluid power (kW) for the compressors, expanders, and turbines, area (m²) for heat exchangers and inter-coolers. In addition, $k_1$, $k_2$, and $k_3$ are capacity-based constants specific for each equipment piece and are given in Table 5 (Turton et al., 2008).

The capacity constants for the two-phase expander are assumed to be the same as that of the single expander because of the unavailability of cost constants in open literature for this relatively new technology.

After calculating $E_p$, all other direct and indirect costs related to the equipment are calculated by using the bare module factor $F_{BM}$ as shown in Eq. (10) (Nagy, 2016; Turton et al., 2008).

$$
C_{BM} = E_p F_{BM}
$$

The total capital investment is given by Eq. (11) (Turton et al., 2008).

$$
TCI = 1.18 \sum_{i}^{n} C_{BM,i}
$$

The grassroots cost for establishing a plant from scratch is given by Eq. (12) (Turton et al., 2008).

$$
GRC = TCI + 0.5 \times \sum_{i}^{n} C_{BM,i}
$$

In this context, the total cost of the compressors, two-phase expanders, cryogenic heat exchangers, intercoolers, and liquids turbines are estimated by using the relations of Turton given above (Turton et al., 2008). To find the commercial feasibility of the proposed processes, the LNG production capacity was set at 6480 kg/h. The capacity factors of the compressors and expanders can be obtained from Aspen Hysys®. It should be noted that Aspen Hysys® does not provide the area $A$ of the exchanger, although it provides the overall heat transfer coefficient $UA$. Therefore, to find the area $A$, a value of $U = 3600$ W/m²K was adopted from the literature (Coulson et al., 1985; Kakac et al., 2002; Luyben and Chien, 2011; Turton et al., 2008), which is the mean value of 1200–6000 W/m²K, as used by (Qyyum and Lee, 2018). The inter-cooler area $A$ is required to estimate costs according to the method prescribed by Luyben and Chien (2011). As discussed in the previous section, the LNG processes are energy intensive owing to the high power consumption in the compression section. Therefore, the operating cost of the compressors is affected by electricity charges. Electricity consumption is calculated by using the total power required by the compressors while ignoring the power produced by the expander.
The cost of electricity, USD 16.8/GJ (Turton et al., 2008), was taken from the literature to calculate the operating cost (Eq. (13)).

\[
OC = \frac{\text{Cost of electricity}}{\text{Specific compression power}} \times \text{(Specific compression power)}
\]  
(13)

The plant maintenance cost was fixed at 2% of the total capital investment along with a five-year payback period to calculate the annualized cost. The total annualized cost was estimated by Eq. (14), as described in previous research (Luyben and Chien, 2011).

\[
\text{TAC} = \left( \frac{\text{Capital cost}}{\text{Payback period}} \right) + \text{Operating cost}
\]  
(14)

Table 5
Equipment purchase cost constants.

<table>
<thead>
<tr>
<th>Equipment Type</th>
<th>Equipment Description</th>
<th>( K_1 )</th>
<th>( K_2 )</th>
<th>( K_3 )</th>
</tr>
</thead>
<tbody>
<tr>
<td>Compressor</td>
<td>Centrifugal, axial, reciprocating</td>
<td>2.2897</td>
<td>1.3604</td>
<td>-0.1027</td>
</tr>
<tr>
<td>Turbine</td>
<td>Axial gas turbine</td>
<td>2.7051</td>
<td>1.4398</td>
<td>-0.1776</td>
</tr>
<tr>
<td>Turbine</td>
<td>Liquid expander</td>
<td>2.2476</td>
<td>1.4965</td>
<td>-0.1618</td>
</tr>
<tr>
<td>Heat exchanger</td>
<td>Flat plate</td>
<td>4.6656</td>
<td>-0.1557</td>
<td>0.1547</td>
</tr>
<tr>
<td>Heat exchanger</td>
<td>Air cooler</td>
<td>4.0336</td>
<td>0.2341</td>
<td>0.0497</td>
</tr>
</tbody>
</table>

The results of economic analysis of the proposed processes are given in Table 6.

Table 6

<table>
<thead>
<tr>
<th>Cost</th>
<th>( N_2 ) single expander (Khan et al., 2015a)</th>
<th>( N_2 ) dual expander (Khan et al., 2015a)</th>
<th>( C_3 )N two-phase expander (Qyyum et al., 2018a)</th>
<th>( C_3 )N two-phase expander</th>
<th>( C_3 )-precooled ( C_2 )N two-phase expander</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total Equipment Purchase Cost (( 10^6 ) $)</td>
<td>4.76</td>
<td>4.84</td>
<td>4.62</td>
<td>4.46</td>
<td>6.35</td>
</tr>
<tr>
<td>Total Base Module Cost (( 10^6 ) $)</td>
<td>19.69</td>
<td>19.99</td>
<td>18.98</td>
<td>18.25</td>
<td>25.90</td>
</tr>
<tr>
<td>Total Capital Investment (( 10^6 ) $)</td>
<td>23.23</td>
<td>23.59</td>
<td>22.40</td>
<td>21.53</td>
<td>30.56</td>
</tr>
<tr>
<td>Grass root Cost (( 10^6 ) $)</td>
<td>33.08</td>
<td>33.58</td>
<td>31.90</td>
<td>30.66</td>
<td>43.51</td>
</tr>
<tr>
<td>Total Operating cost (( 10^6 ) $)</td>
<td>2.31</td>
<td>2.24</td>
<td>1.54</td>
<td>1.35</td>
<td>1.23</td>
</tr>
<tr>
<td>Maintenance cost (( 10^6 ) $)</td>
<td>0.46</td>
<td>0.47</td>
<td>0.44</td>
<td>0.43</td>
<td>0.61</td>
</tr>
<tr>
<td>Total Annualized Cost (( 10^6 ) $)</td>
<td>7.46</td>
<td>6.96</td>
<td>6.02</td>
<td>5.66</td>
<td>7.34</td>
</tr>
<tr>
<td>Relative operating cost savings (%)</td>
<td>20.28</td>
<td>45.19</td>
<td>51.95</td>
<td>56.22</td>
<td></td>
</tr>
<tr>
<td>Relative TAC cost savings (%)</td>
<td>6.70</td>
<td>19.30</td>
<td>24.12</td>
<td>1.60</td>
<td></td>
</tr>
</tbody>
</table>

Fig. 8. TDCC and THCC curves of the (a) and (c) \( C_3 \)-precooling exchanger and (b) and (d) \( C_2 \)N main cryogenic heat exchanger.
The results of the economic analysis revealed that as the process complexity increased, the capital cost required also increased. Furthermore, as the energy efficiency of a process increases, its capital cost and the overall operating cost reduce. The TAC is affected by the total capital investment and operating costs. The N2 single expander process exhibited the highest operating and capital costs among all single-cycle processes, with a TAC of US $7.46 million. Moreover, the proposed C2N process has a TAC of US $5.66 million owing to its low capital investment and operating costs. Although the relative operating cost savings show that the proposed C3-precooling process results in 56.22% savings when compared with the conventional N2 single expander process, the relative TAC savings show that the proposed C2N process is the most economical process. For this reason, apart from applying C3-precooling and increasing the overall TAC, the proposed C2N process is the most economical and feasible process.

4. Conclusions

A new binary MR, i.e., ethane–nitrogen, was applied in a two-phase expander process to reduce the overall operating costs of nitrogen expander-based liquefaction processes for LNG-FPSO projects. A C2N-precooling refrigeration cycle was also integrated in the two-phase expander process to further reduce the operating costs at the expense of an increase in capital costs. The MCD algorithm was applied to realize the maximum benefits from the proposed improvements. The proposed enhancements significantly reduced the energy consumption for LNG production, which ultimately led to reductions in the operating cost and the total annualized cost. The following conclusions were drawn from the current study:

- When compared with the N2 single expander process, the proposed C2N-based and C3-precooled C2N processes can produce LNG with significant energy savings (79.72% and 81.52%), depending on the feed conditions, composition, and design parameters.
- The relative TAC savings were 24.12% for the proposed C2N process and 1.60% for the proposed C3-precooled C2N process. Hence, it is concluded that the proposed C2N process is the most optimal, in terms of economic efficiency, because its TAC savings were much higher than those realized by the propane-precooled C2N two-phase expander and the single and dual nitrogen-based and C3N two-phase expander liquefaction processes.
- The exergy analysis revealed that the process performance can be further enhanced by using a rigorous optimization technique or by improving the refrigeration cycle.

The major limitation of the proposed liquefaction processes is the stability and reliability of the two-phase expander. Therefore, it is mandatory to evaluate the robustness of the proposed processes. Furthermore, detailed thermodynamic evaluation in terms of advanced exergy analysis can also be performed to evaluate the real-potential improvements to make the feasible process for commercialization. The proposed LNG processes can be further improved by investigating and employing other precooling refrigerants (such as CO2, HFO-1234yf, and NH3) or introducing intermediate cooling step in order to reduce the overall entropy generation.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Acknowledgements

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Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.jclepro.2019.119189.

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