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Research Paper

Gas–liquid dual-expander natural gas liquefaction process with confirmation of biogeography-based energy and cost savings

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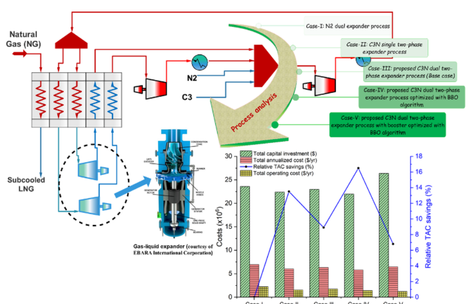
HIGHLIGHTS

- Two-phase dual expander refrigeration cycle for LNG production.
- Propane-nitrogen binary mixed refrigerant is used.
- Biogeography algorithm is used for optimization of proposed process.
- Energy savings up to 38.12% compared with the conventional N_2 dual-expander process.
- Payback period based economic evaluation is performed.

GRAPHICAL ABSTRACT



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ABSTRACT

The use of nitrogen (N_2) expander-based liquefaction processes is prevalent in offshore sites for the production of floating liquefied natural gas. It is safe, has simple operability, and has portable design with a small deck space requirement. However, the high operating cost that mainly accounts for the shaft work requirement in the compression units of the refrigeration cycle, is still a major ongoing issue associated with nitrogen expander liquefaction processes. This high operating cost increases the total annualized costs of the N_2 expander liquefaction technology, and this ultimately reduces its global competitiveness of the process. Recent developments in expansion devices pave the way toward the handling of gas–liquid (two-phase) refrigerant in an isentropic manner instead of an isenthalpic one. This study presents the propane-nitrogen two-phase dual expander liquefaction process for offshore applications. A bio-inspired strategy named "biogeography" is used to confirm the overall energy savings with minimal total annualized costs. The results show that the proposed liquefaction technology gives 36.6% operating cost savings, and this confirms a 16.5% saving in total annualized costs compared with the conventional nitrogen dual gas-phase expander liquefaction process. This study is an extension of our previous study "Innovative propane-nitrogen two-phase expander refrigeration cycle for energy-efficient and low-global warming potential LNG production".

1. Introduction

The exponential increase in the release of CO_2 into the atmosphere significantly contributes to global warming. By November 2018, the concentration of CO_2 in the atmosphere increased drastically to a critical level of 408.46 ppm for the first time in the human history [1]. It is projected that, by the end of this century, the global temperature will rise by 1.4 °C to 5.0 °C, and the concentration of CO_2 will rise to a level

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Nomenclature and Abbreviations

E_p	equipment purchase cost
k_1, k_2, k_3	constants
A	capacity parameter
C_{BM}	bare module cost
F_{BM}	bare module factor
FLNG	floating liquefied natural gas
LNG	liquefied natural gas
NG	natural gas
TDCC	temperature- approach (delta) temperature composite curves
THCC	temperature-heat flow composite curves
TAC	total annualized costs
C3N	propane-nitrogen
CPPs	critical process parameters
BBO	biogeography-based optimization
MITA	minimum internal temperature approach
EFG	end flash gas
MCT	module costing technique
SCP	specific compression power
TCI	total capital investment
OC	operating cost

between 478 and 1099 ppm [2,3]. Green energy technologies are in the early stages of their development and cannot fulfill the increasing energy demands by themselves as a substitute for fossil fuels. Among all available fossil fuels, natural gas (NG) is the cleanest one so far [4,5]. NG is an efficient fuel that gives high energy density with significantly low air pollutants [6] and meets the strict environmental regulations in comparison with other fossil fuels, such as coal and oil [7]. Therefore, the demand for NG is increasing rapidly as an after-effect of world economic crisis. A 60% rise in demand for NG is projected from 2010 to 2030, and it is expected that its share in the global energy consumption will increase from the current share of 23% to 26% by 2040 [8,9].

The transportation and storage economics of NG depend on the location of its reservoirs, which are generally located in remote and offshore areas. The NG is transported via pipelines (in the form of a pressurized gas at distance < 2000 km) or in the form of liquefied NG (LNG) by means of large cargo vessels [10]. NG transportation in liquid form has been approved and commercialized as an economically advantageous strategy. However, the refrigeration and liquefaction step in the LNG value chain is considered energy- and cost intensive, costing approximately half of the total expenditure of the value chain [7,11]. Meanwhile, the energy demand for liquefaction varies depending on the liquefaction technology and the environmental plant site conditions [12].

The most commonly used offshore liquefaction technologies are single mixed refrigerant processes and nitrogen expander cycles [13]. The N_2 expander LNG process has been found to have high occupational safety, low capital investment, easy operation, and portability, which make it the best choice for offshore operations. Nevertheless, LNG production adopting nitrogen expander refrigeration cycles is more energy intensive than mixed refrigerant liquefaction processes [14]; therefore, these are only favorable for small-scale and offshore applications [15]. Many investigations have been attempted to analyze and improve the energy efficiency of N_2 expander liquefaction processes. For instance, Gu et al. [16], suggested the use of binary mixed refrigerant by adding methane in the nitrogen and reduced the operating costs in terms of the shaft work requirement for N_2 expander LNG processes. Further improvement to the N_2–CH_4 expander refrigeration cycle through optimization solely considering the shaft work requirement as an objective function was done by Cao et al. [17]. Due to the relatively high energy efficiency with minimal TAC, recently, Haider et al. [18] used N_2–CH_4 expander process for the liquefaction of biomethane. He et al. [19] investigated propane and R410a as precooling refrigerants in precooling refrigeration cycle to enhance the performance of the nitrogen expander process for offshore LNG production. In another investigation, He et al. [20] modified the process configuration by adding a parallel N_2 expander process with the main refrigeration unit and optimized the proposed configuration using the genetic algorithm to get the maximum benefits of the modifications. Moreover, Song et al. [21] enhanced the energy efficiency of the N_2 expander process by using the empirical model technique. Later, Khan et al. [22] reduced the operating costs of nitrogen single- and dual-expander liquefaction technologies adopting optimization of operating parameters. Recently, Qyyum et al. [23] proposed a self-cooling recuperative nitrogen single-expander liquefaction technology and concluded that the overall energy efficiency of the N_2 expander-based technology can be improved significantly without involving any external precooling cycle. In other studies [24,25], Qyyum et al. integrated a vortex tube with a nitrogen expander refrigeration cycle and improved the energy efficiency of the liquefaction process for small-scale and offshore applications. Most recently, Palizdar et al. [26] performed an advanced exergoeconomic analysis to evaluate a gas-phase dual-expander refrigeration cycle for small-scale LNG production.

The aforementioned literature demonstrates that the overall competitiveness of nitrogen expander liquefaction processes has been improved to some extent by modifying the existing configuration, adding an expander and/or through optimization. It has also been reported [22,26,27] that the overall energy efficiency of the nitrogen single-expander liquefaction process can be enhanced by adding another gas-phase expander at the expense of capital investment and degree of complexity to some extent. Nevertheless, the total annualized cost (TAC) associated with the nitrogen dual-expander liquefaction process is still not reasonable, mainly because of the significant high operating costs (in terms of compression power) as compared to mixed refrigerant-based liquefaction processes [14]. In this context, there is a need to improve the dual-expander LNG processes in order to make it the best competitive liquefaction process especially for offshore applications.

In a previous study [28], a propane–nitrogen binary mixed refrigerant adopting a gas–liquid single expander followed by particle swarm optimization was studied. In that study, it was concluded that gas–liquid isentropic expansion provides a remarkable energy efficiency enhancement in comparison with a nitrogen single gas-phase expander. During the revision of that study [28], a detailed economic analysis in terms of the TAC of the C3N gas–liquid single expander was suggested. Analysis of the dual gas–liquid expander configuration in comparison with a C3N single gas–liquid expander, as well as a gas-phase dual nitrogen expander, was also recommended. The previous study [28] also includes details about gas–liquid expanders and detailed exergy analysis; therefore, in that study, to avoid the dilution of the main contribution it was not possible to consider all these issues and recommendations.

Therefore, this study presents a dual gas–liquid expander liquefaction process adopting propane–nitrogen as a binary mixed refrigerant. A bio-inspired approach, “biogeography,” was used to confirm the energy savings with minimal TAC as compared to the classical gas-phase dual nitrogen expander and C3N gas–liquid single-expander liquefaction processes. Furthermore, the impact of boosted feed (put under high pressure through booster) NG on the overall operating and maintenance costs and the TAC of the C3N dual gas–liquid expander liquefaction process was also investigated. Finally, this study proposes an energy- and cost-efficient gas–liquid isentropic-expansion-based LNG process that can be one of the most suitable candidates for FLNG projects.

2. Process description

The classical N_2 dual-expander process consists of a main cryogenic heat exchanger, a refrigeration cycle composed of high-pressure and low-pressure compression sections involving two compressors with an inter-stage cooling system for each compression section and two expanders for the expansion of the refrigerant to generate a cooling effect for the liquefaction of NG. The detailed explanation about the typical nitrogen dual-expander liquefaction process can be found in [22,29].

In the proposed processes, the conventional refrigerant is replaced with a new binary mixed refrigerant (C3N), and the structure is modified by replacing the gas-phase expanders with gas–liquid expanders that provide isentropic expansion of the refrigerant with a shaft work as an extra benefit. To lower the pressure of the subcooled LNG, the Joule–Thomson valve is replaced with a cryogenic liquid turbine. The proposed gas–liquid dual-expander liquefaction process is schematized in Fig. 1. To avoid any confusion in the process description of the proposed liquefaction technology, the refrigerant streams are labeled "stream-*x*" (*x* = 1, 2, 3…) and the NG streams are labelled "stream-*y*" (*y* = A, B, C…).

Stream-1 is compressed in the low-pressure two-stage compression section equipped with interstage coolers. Stream-5 is maintained at a pressure of 17.94 bar and mixed with the recycled intermediate pressure stream-6. Fresh makeup streams of nitrogen and propane are also added at this point to maintain the optimal composition of the C3N mixed refrigerant. Stream-7 at an intermediate pressure value of 17.94 bar is then introduced to the high-pressure two-stage compression section. This section is also equipped with inter-stage coolers. The high-pressure compression unit increases the pressure of the refrigerant to the optimal value, i.e. 83.75 bar (stream-11). The pressure ratio for the low-pressure and high-pressure sections is kept 1:3 considering the capital investment and irreversibilities in the compressors. The high-pressure stream is split into two streams (12 and 13) before it is introduced to the cryogenic heat exchanger (LNG-100), where it passes parallel to the feed NG, as shown in Fig. 1. The streams (12, 13, and C) exchange their heat with incoming stream-16 and stream-17, which are coming from gas–liquid expanders. Stream-16 is obtained as a result of the isentropic expansion of stream-14 to 3.5 bar through gas–liquid expander (Ex-1), whereas stream-17 is obtained after expansion of stream-15 to 18.04 bar through gas–liquid expander (Ex-2). The exiting superheated vapor streams (1 and 6) are recycled, and the refrigeration loop is accomplished. The high-pressure subcooled LNG is obtained as stream-D, which is introduced in the cryogenic liquid turbine, and the pressure is reduced to just above the atmospheric pressure of 1.20 bar to make an economic transportation. This pressure reduction of subcooled LNG gives a final product at a temperature of –158 °C with 8% end flash gas (EFG).

Table 1
Feed NG conditions and composition.

Feed natural gas	Value
Temperature (°C)	30.0
Pressure (bar)	50.0
Flow rate (kg/h)	1.0
Composition	Mole %
Methane	91.30
Ethane	5.40
Propane	2.10
i-Butane	0.50
n-Butane	0.50
i-Pentane	0.01
n-Pentane	0.01
Nitrogen	0.20

3. Aspen Hysys-based process simulation

The simulation of the proposed liquefaction process was conducted using the well-established commercial software Aspen Hysys v.10. A Peng-Robinson fluid package with the option of Lee-Kesler (for enthalpy calculations) equation was chosen for the simulation of proposed LNG process. The feed NG composition and conditions are given in Table 1, and the fundamental simulation assumptions are listed in Table 2. Tables 1 and 2 were adopted and arranged from prior research [22,28].

4. Process optimization

Process optimization (either design or operation) is considered as one of the most important and critical steps during the scaling up or improvement of any process with the aim of achieving maximum performance at minimal cost [33]. During design optimization, critical process parameters (CPPs) and key decision variables affecting the performance of the process are identified in a rigorous and robust manner. Process and operations engineers focus on increasing overall profit by increasing the production rate or minimizing operating costs

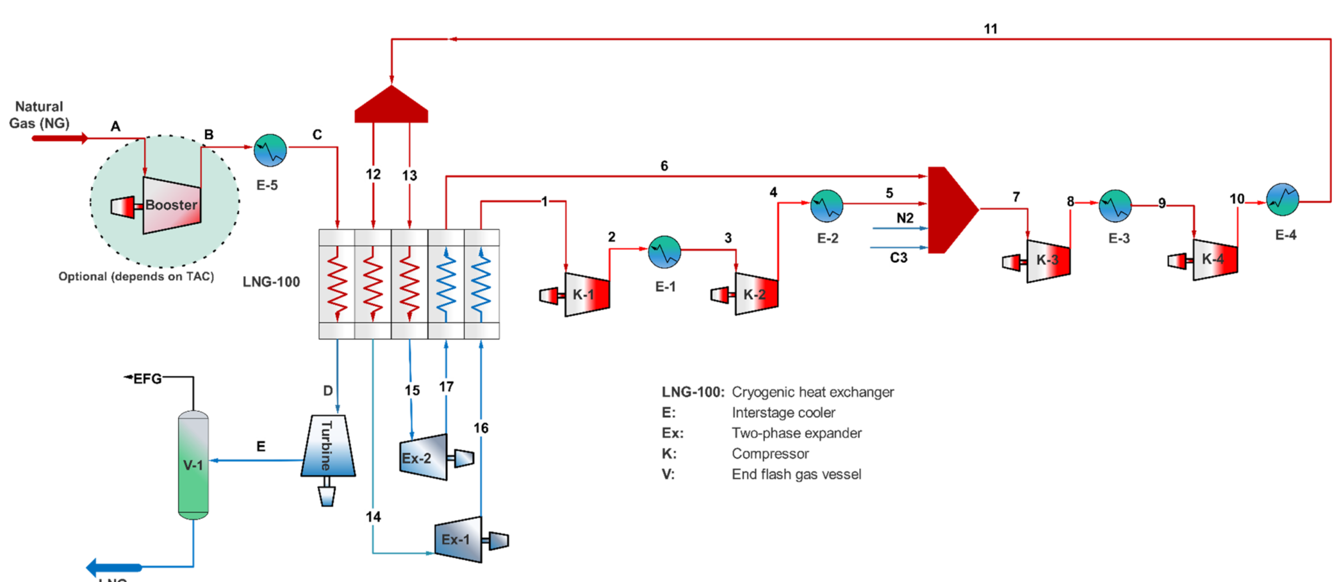

Fig. 1. Process flow diagram of the proposed C3N gas–liquid dual-expander LNG process.

Table 2
Fundamental assumptions for the simulation of proposed liquefaction process.

Simulation assumptions	Value
ΔP across the inter-stage coolers	0.25 bar
Inter-stage cooling medium temperature	20.0 °C
Heat losses	Negligible
Minimum internal temperature approach (MITA)	3.0 °C
End flash gas (EFG) pressure	1.20 bar
LNG temperature	–158.5 °C
LNG liquid fraction (by mole)	0.92
Compressor isentropic efficiency	75%
Gas-liquid expander isentropic efficiency	80%
LNG turbine isentropic efficiency	90.0% [30–32]
cooler outlet temperature	30.0 °C
ΔP across the LNG exchanger	
Hot stream (bar)	1.0
Cold Stream (bar)	0.1

through fine-tuning the CPPs. This means that even if the initial design optimization is successfully accomplished, the design work is not completed until maximum possible process performance attained at minimal TAC. Hence, process optimization using dedicated algorithms is always an important step in achieving improved performance at minimal operating costs. Furthermore, replacing new equipment, such as a gas–liquid expander with C3N binary mixed refrigerant in this study meant that while the overall configuration of the gas-phase dual-expander liquefaction process remained the same, changes to the optimal design variables can be made. These changes can reduce the exergy efficiency significantly as a result of the non-optimal execution of key decision variables. This could lead to energy wastage. Therefore, it is important for a rigorous optimization review to follow any process design modification to ensure that the maximum potential benefits from installing that process are realized. In this context, the proposed gas–liquid dual-expander LNG process was optimized through an evolutionary algorithm named "biogeography-based optimization" (BBO) [34–36] by interfacing Aspen Hysys v.10 with MATLAB 2018a. This interface between Hysys and MATLAB was built using the component object model. The objective function, constraint, and decision variables of the proposed processes are listed in Table 3.

Table 3
Objective function, constraint, and decision variables with their lower and upper limits.

Objective function:		
Operating costs in terms of overall compression power (kW)	$Minimize f(X) = Min.\left(\frac{\sum_{i=0}^{n}\dot{W}_i}{\dot{m}_{NG}}\right)$	
Constraint:		
Minimum internal approach temperature (°C)	$\Delta T_{\min}(X) \geq 3.0$ $X_{Lower} < X < X_{upper}$ where, X is a vector of the decision variables	
Decision Variables	Lower limit	Upper limit
Mass flow rate of nitrogen, $\dot{m}_{N2}$ (kg/hr)	2.5	5.5
Mass flow rate of propane, $\dot{m}_{C3}$ (kg/hr)	2.0	4.5
Refrigerant (C3N) low pressure (bar)	1.5	7.5
C3N medium pressure (bar)	15.0	35.0
C3N high pressure (bar)	45.0	110.0
C3N split ratio	0.5	0.85
C3N subcooling temperature (°C)	–160.0	–140.0

Biogeography-based optimization can be applied to optimization problems from any domain. It has already been applied in continuous optimization, multi-objective optimization, combinatorial optimization, constrained optimization, and noisy optimization. In addition to its application in optimization problems, it has many applications in the field of engineering. For example, power system problems, parameter estimation problems, data analysis, network and antenna problems, image processing, and scheduling problems have been solved using different variants of the BBO algorithm [37]. As far as the advantages of BBO are concerned, it is the fastest-growing nature algorithm and efficiently solves problems related to practical optimization. Furthermore, its qualities of being simple, flexible, and computationally efficient make it more versatile, because it does not perform any derivative operation on the objective functions, which confirms its stochastic nature.

The working principle of BBO is motivated by the natural events of biogeography. The basic concepts of BBO are analogous to the traits of biogeography. A general framework of the BBO algorithm is shown in Fig. 2. Further detailed explanation about the BBO can be found in [34–36].

5. Results and discussions: Process analysis

After the introduction of the dual gas–liquid expander with BBO-based optimal execution, the performance of the liquefaction process was analyzed according to the sequence shown in Fig. 3.

First, N_2 the dual-expander process [22] (case-I) and the C3N gas–liquid single expander process [28] (case-II) were chosen to create a standard for the confirmation of energy and cost savings by the proposed liquefaction process. The base case (case-III) of the proposed C3N dual gas–liquid expander process was modeled and optimized by adopting design parameters and variables from case-I and case-II. Subsequently, case-III was optimized using the BBO algorithm considering the overall compression power as an objective function constrained by a minimum internal approach temperature (MITA) value of 3.0 °C. As a result of BBO optimization, case-IV was obtained. It has been reported [32,38] that the feed NG pressure is also significantly affected by the performance parameters (such as overall compression energy and approach temperature) of the LNG process. Therefore, a feed NG booster was also applied (case-V) with a C3N dual gas–liquid expansion refrigeration cycle.

Table 4 summarizes the design parameters and variables of the proposed liquefaction process (case-IV) in comparison with the case-I, case-II, case-III, and case-V. According to Table 4, the total circulating refrigerant mass flow is 13.66, 5.97, 8.48, 7.46, and 7.5 kg/h for case-I, case-II, case-III, case-IV, and case-V, respectively. Among them, the C3N two-phase single expander LNG process has a lower mass flow, and the conventional gas-phase dual-expander process uses a high refrigerant mass flow, i.e. 13.66, which is 45.4% higher than for the C3N gas–liquid dual-expander process without a booster and 45% higher with a booster. Here, one thing is interesting: the total refrigerant mass flows for case-IV and refrigerant V have a negligible difference, but their relative energy savings significantly differ. For case-IV, 29.8% of energy can be saved in comparison with case-I, whereas, for case-V, net energy savings can be as much as 39.4%. The operating pressures (condensation and evaporation) of the refrigeration cycle also affect the overall performance of the liquefaction process. In the conventional process, case-I, the high (condensation) and low (evaporation) pressure of the refrigerant in the loop has values of 100.0 and 14.0 bar, respectively.

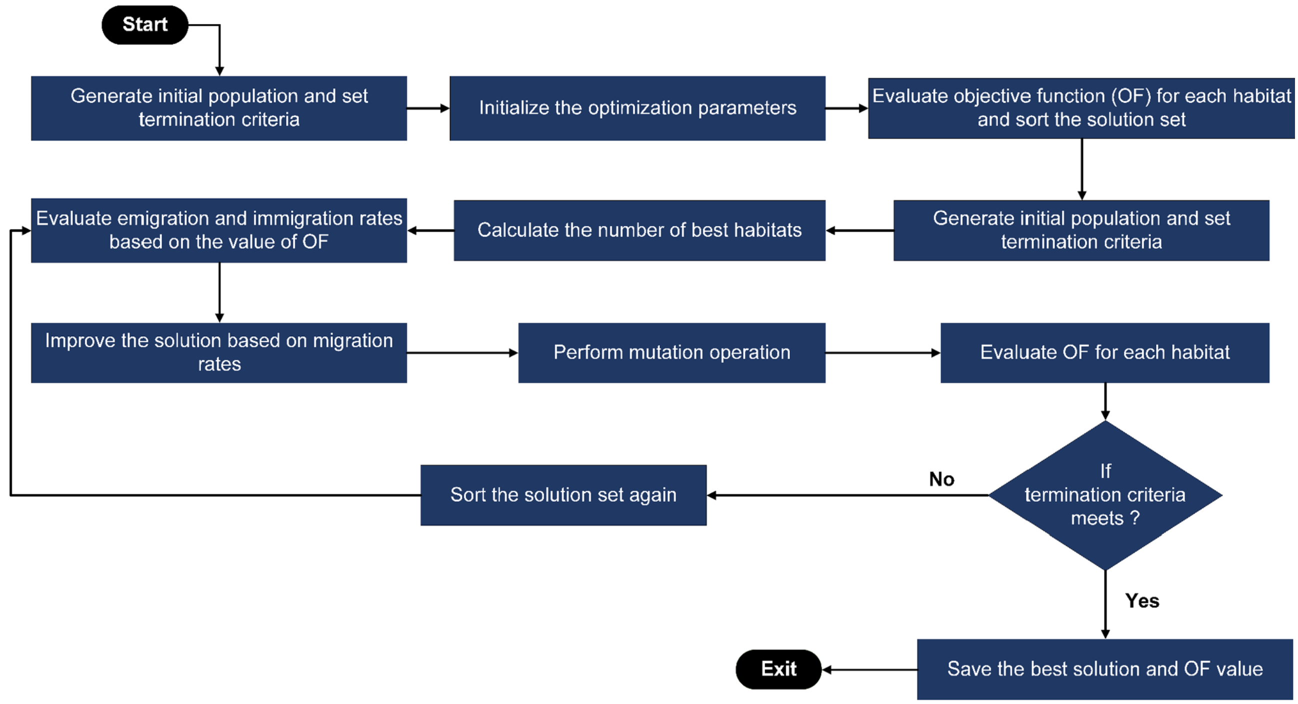

Fig. 2. Working (searching) flowchart of BBO algorithm.

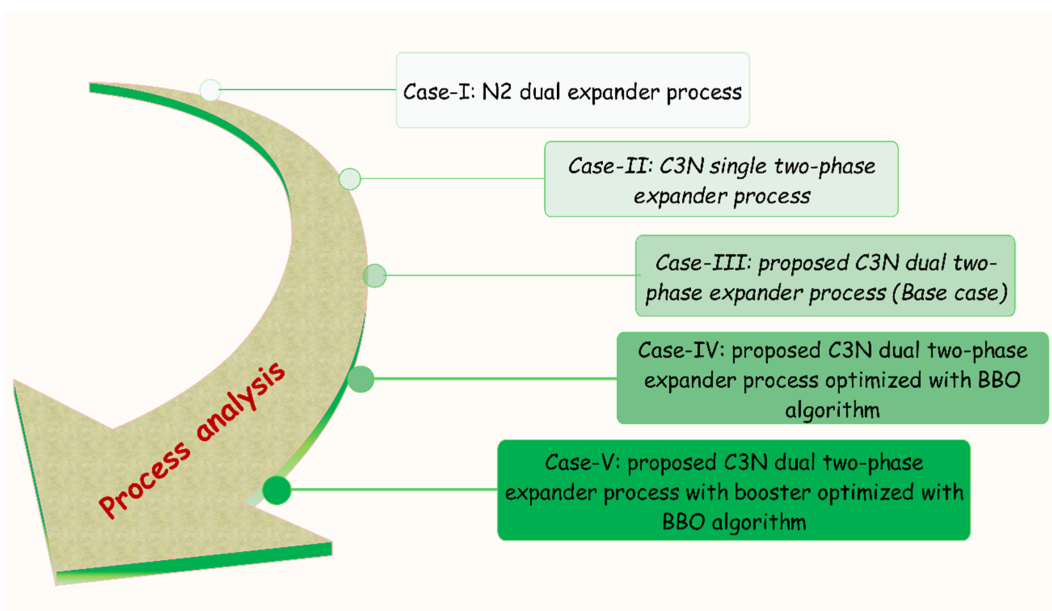

Fig. 3. Process analysis sequence for dual C3N gas–liquid expander liquefaction process.

After BBO optimization, the high pressure (condensation pressure) for case-IV and case-V were 84.0 and 76.0 bar, respectively, which are significantly lower than that (100.0 bar) of the conventional gas-phase dual-expander liquefaction process. This difference is mainly due to the result of the presence of propane, which is considered a high-boiling point component and has a higher molar mass than nitrogen. This ultimately leads to a higher specific refrigeration effect with a relatively low compression power. Therefore, the net compression power requirement for the conventional process (case-I) is 0.5010 kW, whereas the proposed optimal schemes (case-IV and case-V) require 0.3521 kW and 0.3039 kW, respectively. Even the single expander process adopting propane–nitrogen also has a low net power requirement, i.e. 0.3989 kW, which is 20.5% lower than that of case-I. In fact, the reduction of the net power requirement leads to reduced operating costs, which ultimately improves the overall profit of the liquefaction process.

Table 4
Optimal findings of the proposed liquefaction process in comparison with the base case and previously published processes.

Parameters	Case-I [22]	Case-II [28]	Case-III	Case-IV	Case-V
EFG (vapor fraction)	0.08	0.08	0.08	0.08	0.08
MITA (°C)	3.0	3.0	3.0	3.0	3.0
Flow rate of N_2[$\dot{m}_{N2}$ (kg/h), $\dot{n}_{N2}$ (kgmol/h)]	13.66, 0.4878	3.16, 0.1128	5.30, 0.1892	4.06, 0.1449	4.0, 0.1428
Flow rate of C_3[$\dot{m}_{C3}$ (kg/h), $\dot{n}_{C3}$ (kgmol/h)]	–	2.81, 0.06372	3.18, 0.07211	3.396, 0.07702	3.50,0.07937
Total refrigerant (kg/h)	13.66	5.97	8.48	7.46	7.50
Refrigerant low pressure (bar)	14.00	4.80	2.70	3.40	4.70
Refrigerant medium pressure (bar)	30.0	–	18.00	17.94	22.0
Refrigerant high pressure (bar)	100.00	80.00	85.00	84.00	76.00
Refrigerant split ratio	0.78	–	0.62	0.6694	0.6959
Refrigerant subcooling temperature (°C)	−153.0	−149.0	−150.3	−154.5	−148.9 (stream-16)
Compression power (kW)	0.6876	0.4734	0.5825	0.4344	0.3723
Generated power (kW)	0.1866	0.0745	0.1179	0.0822	0.0684
Net power requirement (kW)	0.5010	0.3989	0.4647	0.3521	0.3039
Specific power consumption (kW/kmol)	8.94	7.11	8.29	6.28	5.42
Relative net energy savings (%)	–	20.5	7.3	29.8	39.4

5.1. Composite curves analysis

Fig. 4 presents the composite curves analysis of the proposed liquefaction processes case-III, case-IV, and case-V, respectively. Fig. 4(a)–(c) visualize the composite curves between temperature and approach temperature (ΔT) also known as (TDCCs); whereas, Fig. 4(d)–(f) are the elaborated graphical form of temperature and heat-flow (THCCs). In the THCCs, heat flow is given at abscissa and the temperature at ordinate for the hot and cold streams. Whereas, in TDCCs, temperature is given at the abscissa and the approach temperature is at the ordinate.

the TDCCs, as illustrated in Fig. 4(a), the approach temperature between the composite curves remains higher (i.e., > 25 °C) than the defined MITA value of 3.0 °C, which shows the non-optimal execution of decision variables. After applying the BBO strategy, the TDDC shown in Fig. 4(b) was obtained, clearly showing that the peak of the TDCC is lower than that of the base case. When the feed NG was boosted through a booster and was also optimized using BBO, the peak of the TDCC, as in Fig. 4(c), was reduced further, which shows significant energy savings in comparison with case-III and case-IV. The peaks of the TDCC for case-IV and case-V are 23 °C and 16 °C, respectively, which are still higher than that of the specified MITA value of 3.0 °C. In real-

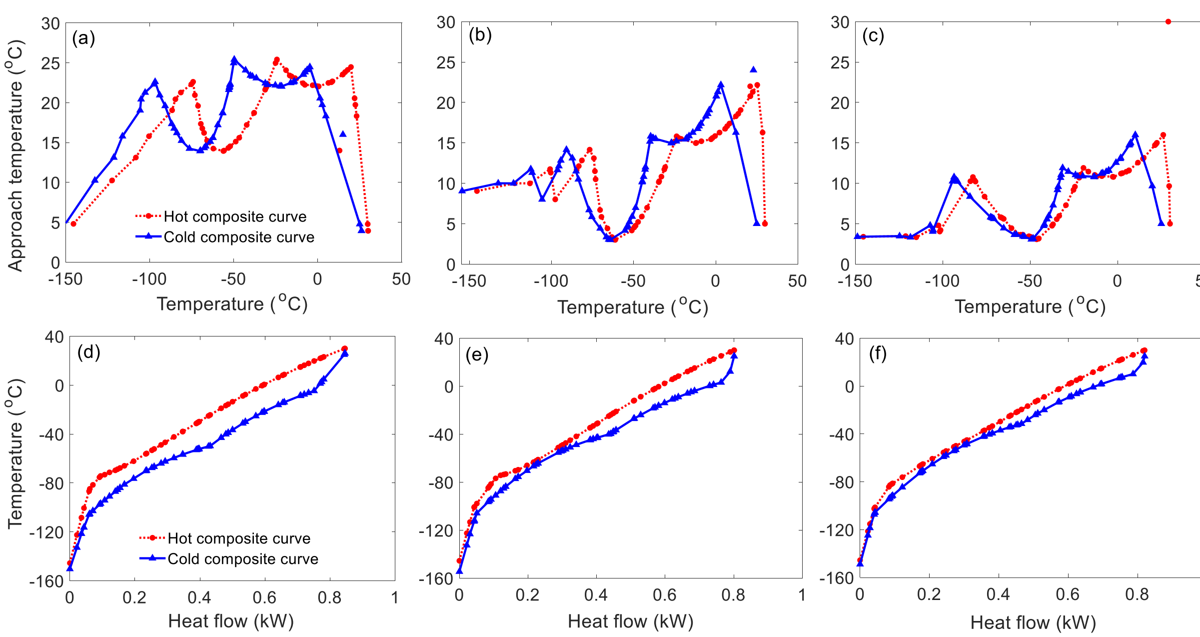

Fig. 4. TDCC curves (a)–(c) for case-III, IV, and V, respectively; THCC curves (d)–(f) for case-III, IV, and V, respectively.

The composite curve analysis is a birds-eye assessment of the exergy loss (or entropy generation) within the LNG cold box. This loss of exergy is generally analyzed by observing the gap margin of hot and cold composite curves. The gap margin in the hot and cold composite curve (THCC) for the base case in Fig. 4(d) is larger than that of Fig. 4(e) and (f), which is why the relative net energy saving for case-III is 7.3% lower than for case-IV and case-V. Whereas, the TDCC guide to analyze the peak of the MITA value that should be followed is the defined feasible MITA value of 3.0 °C throughout the cryogenic heat exchanger for an efficient heat transfer between the refrigerant and the feed NG. In

life LNG operations, it is a challenging task to follow the recommended approach temperature (i.e., 1.0–3.0 °C) throughout the cryogenic exchanger length, mainly because of the feed NG ingredients, such as methane, nitrogen, and hydrocarbons, which have different boiling points. Therefore, it is considered a challenging task to find optimal flow rates of refrigerants with optimal operating conditions to match the cold composite curve with the hot composite curve. Furthermore, the composite curve analysis of case-IV and case-V also shows that the energy savings corresponding to minimal capital investment can be better achieved through further rigorous optimization using either a

Table 5
Type of cost along with equations for the liquefaction process cost estimation.

Cost	Equation
Equipment Purchase Cost [40]	$log_{10}(E_p) = k_1 + k_2 log_{10}A + k_3(log_{10}A)^2$
Bare Module Cost [40]	$C_{BM} = E_p F_{BM}$
Total Capital Investment [40]	$TCI = 1.18 \sum_i^n C_{BM,i}$
Grass Root Cost [40]	$GRC = TCI + 0.5 \times \sum_i^n C_{BM,i}$
Operating Cost [39]	$OC = Costofelectricity\left(\frac{\$}{kW.yr}\right) \times (SCP)$
Total Annualized Cost [39]	$TAC = \left(\frac{Capitalcost}{Paybackperiod}\right) + Operatingcost$

deterministic or a stochastic approach.

5.2. Economic evaluation

There are several possible ways to calculate the TAC in the conceptual design stage. One of most popular methods is the ACCR (Annual Capital Charge Ratio) based on the interest and plant life time, which considers the time value of money. However, because of difficulties in choosing proper assumptions for interest and plant life time which largely affect the TAC result, the authors adopted a simpler method from [39] that is based on the payback period as seen in Table 5. The payback period for the return on the investment was assumed to be 5 years. Further details about the economic analysis method adopted in the proposed LNG processes can be found in the handbooks of the Turton and Luyben [39,40]. Moreover, the equations incorporated for the cost estimation are given in Table 5.

To carry out the cost estimation, the capacity of the proposed processes was set to 6480 kg/h (just assumed value) of LNG to analyze the commercial viability of the proposed processes. Using the equations provided in Table 5, the cost for the process equipment, i.e., compressors, gas–liquid expander, inter-coolers, heat exchanger, and liquid turbine, was estimated. To analyze the cost of compressors, gas–liquid expanders, and liquid turbines, the capacity factor, i.e., fluid power (kW), was obtained from the Aspen Hysys. However, to calculate the cost of the cryogenic heat exchanger, the capacity factor was the area (A) of the heat exchanger, which could not be obtained from the Aspen Hysys. In fact, Aspen Hysys v.10 provides the value of UA rather than separate area (A) value [41]. Table 6 lists the UA (product of overall heat transfer coefficient and area) values obtained from Aspen Hysys.

Table 6
UA values of LNG heat exchanger for all cases (I, II, III, IV, and V).

Cases	UA (kJ/°C.hr)
Case-I	266.4
Case-II	254.2
Case-III	176.3
Case-IV	319.5
Case-V	433.4

To find the heat exchanger area A, the value 3600 W/m^{2}K of the heat transfer coefficient (U) was taken from a recent study [42] relevant to LNG process costing. The cost of intercoolers was estimated using the method devised by Luyben et al.[39]. To evaluate the operating costs, especially for compression units, the electricity cost for the proposed processes was taken as $16.80/GJ [39]. The comparison of the economic analysis of the proposed optimized LNG processes (case-IV and case-V) with the base case (case-III) and previously published cases (case-I and case-II) is given in Table 7.

According to Table 7, for case-V (boosted C3N dual two-phase expander process), the relative operating costs can be reduced by as much as 45.5% compared with case-I. The high operating cost savings are mainly due to the result of the high energy efficiency because of the direct relation between the process energy efficiency and the operating costs. However, the complexity and footprint of the liquefaction process also increase upon introducing any additional equipment, such as a booster in case-V. The high degree of complexity and large footprint make the process less attractive for offshore applications. Furthermore, there is always a tradeoff between operating costs and capital investment/increment. This tradeoff clearly can be seen in Table 7 for case-V, which has the least operating cost, (i.e. $1.22 million/yr), but the highest total capital investment (i.e. $26.38 million) in comparison with all other cases. The highest total capital investment of case-V leads to the higher total annualized costs (TAC) compared with case-IV (without a booster). The proposed C3N dual-expander process with a booster (case-V) gives the best results in terms of the highest energy and operating cost savings, 39.4% and 45.5% respectively, but the major drawback associated with this process is the high capital investment (i.e. $26.38 million), which ultimately lowers the TAC savings to 6.8% compared with the conventional nitrogen dual-expander process. However, the proposed C3N dual-expander process (case-IV) gives the lowest total capital investment requirement of $21.98 million and the highest TAC savings of as much as 16.5% compared with the other liquefaction processes. Therefore, the proposed case-IV can be the most

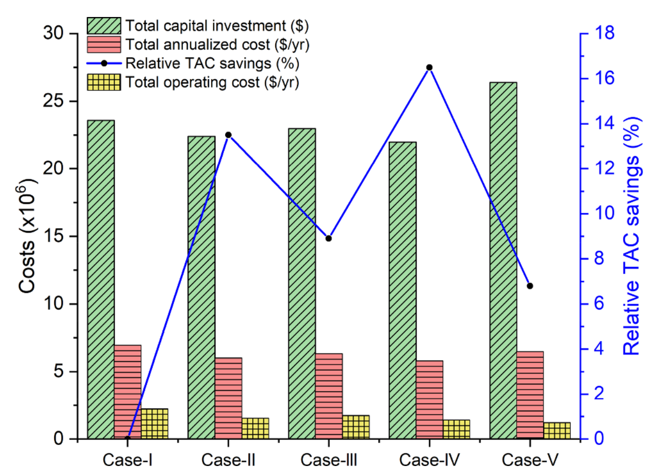

Fig. 5. Economic analysis of the proposed C3N dual two-phase expander process in comparison with other available cases.

Table 7
Economic evaluation of the proposed LNG processes.

Cost	Case-I [22]	Case-II [28]	Case-III	Case-IV	Case-V
Total equipment purchase cost (10^6 $)	4.84	4.62	4.70	4.50	5.42
Total base module cost (10^6 $)	19.99	18.98	19.48	18.63	22.36
Total capital investment (10^6 $)	23.59	22.40	22.98	21.98	26.38
Total operating cost (10^6 $/yr)	2.24	1.55	1.75	1.42	1.22
Total annualized cost or TAC (10^6 $/yr)	6.96	6.02	6.34	5.81	6.49
Relative operating cost savings (%)	_	30.8	21.9	36.6	**45.5**
Relative TAC cost savings (%)	_	13.5	8.9	**16.5**	6.8

suitable and promising candidate for FLNG projects, primarily because of the highest TAC savings. Furthermore, Fig. 5 demonstrates the birds-eye economic evaluation (TCI, TAC, operating costs, and TAC savings) of the proposed LNG processes in comparison with others.

6. Conclusions

The conventional N_2 dual-expander process has upgraded successfully by employing the dual gas–liquid expanders adopting a binary mixed refrigerant consisting of propane and nitrogen. A feed NG booster has also investigated for further energy efficiency enhancement of the proposed liquefaction process. The proposed processes have optimized using a BBO technique, which contributed to obtaining the maximum potential benefits of the proposed process modifications. To provide a clear and easy understanding of the proposed contribution, the process analysis has categorized into five cases. The major conclusions from the proposed study are as follows.

- The amount of the refrigerant gradually decreases when the capital and operating cost are reduced simultaneously.
- The proposed case-IV and case-V can result in relative energy savings of 29.8% and 39.4%, respectively, compared with the conventional N_2 dual-expander process (case-I).
- However, the economic analysis gives a superior edge to the proposed C3N dual gas–liquid expander process (case-IV) because of its low total capital investment of $21.98 million, which ultimately leads to reducing the TAC to $5.81 million/yr — equivalent to a 10.5% saving — compared with case-V.
- The 16.5% relative TAC savings of the proposed C3N dual gas–liquid expander process against the conventional N_2 dual-expander process demonstrates its potential as a competitive and promising candidate for offshore LNG production processes.

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.

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