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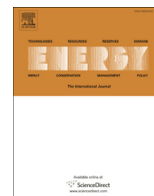
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# Technical and economic optimization of expander-based small-scale natural gas liquefaction processes with absorption precooling cycle

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## ABSTRACT

The objective of this study is to investigate potential technical and economic performance improvement for expander-based natural gas liquefaction processes in small-scale applications. Four expander-based processes were optimized and compared in this study, including conventional single nitrogen expansion process without (SN) and with ammonia absorption precooling (SNA), and single methane expansion process without (SM) and with ammonia absorption precooling (SMA). A two-phase expander is utilized in the methane expansion process to enable liquid generation at the expander outlet. The optimization was done with two objective functions: minimization of specific energy consumption and minimization of production cost. The energy and cost analyses were performed for the four processes by comparing optimization results. Lastly, exergy losses in the main equipment were analyzed. The results show that the ammonia precooling cycle reduces energy consumption and production cost by 26–35% and 13–17%, respectively. The single methane process with precooling is the most promising process, which has 28–48% lower energy consumption and 13–43% lower production cost compared to those of the other three processes. Results also indicate that the best techno-economic performance is obtained with objective of minimizing production cost and not with the commonly used energy-related objective.

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## 1. Introduction

According to the Shell LNG Outlook 2019, more than 40% of the energy demand growth between 2019 and 2035 will be covered by natural gas, with liquefied natural gas (LNG) being the fastest-growing supply source (annual growth rate of 4%) [1]. Driven by increasing demand for LNG, demand for small-scale LNG plants (from 5 to 500 tonnes of LNG per day) are showing strong growth. This is because they are suitable for the exploitation of abundant small size and satellite stranded gas resources [2,3]. An expander-based natural gas liquefaction process is suitable for small-scale LNG plants because of its advantages in simplicity, start-stop convenience, insensitivity to motion, and strong mobility [4]. The main disadvantage of the expander-based liquefaction process is low energy efficiency compared to that of the mixed-refrigerant liquefaction process and the pure-refrigerant cascade liquefaction process [5]. Hence, research focus has mainly been on improving the energy efficiency of the expander-based liquefaction process

[6–10]. However, focusing on only energy saving will not always lead to the lowest production cost [3,11] as the increase in capital and maintenance costs could exceed the energy costs saving. Therefore, maintaining low capital costs and maintenance costs, as well as improving the energy efficiency of the expander-based liquefaction process, are the key to a successful optimization.

Several studies focused on improving the energy efficiency of the expander-based natural gas liquefaction process. Their efforts include utilizing mixed refrigerant, two-phase expander, and adding precooling cycle. Cao et al. [12] designed and optimized the expander-based liquefaction process by using a N<sub>2</sub>–CH<sub>4</sub> mixed refrigerant. Their results showed that the efficiency of the N<sub>2</sub>–CH<sub>4</sub> expander process exceeded that of a single mixed refrigerant (SMR) process. Ding et al. [13] proposed and optimized a N<sub>2</sub>–CH<sub>4</sub> expander process with propane precooling. This process could reduce unit power consumption by 36% compared to that of a conventional nitrogen expander process [13]. Remelje and Hoadley [14] evaluated four liquefaction processes, including an SMR process, a two-stage N<sub>2</sub> expander process, and two open-loop natural gas expander processes. Their exergy analysis showed that the SMR process was the most efficient with the lowest total shaft work requirements. However, the two-stage nitrogen expander process

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and one of the open-loop expander process were more suitable for small-scale and offshore LNG production because of compactness and relatively high efficiency. Qyyum et al. [9] designed a two-phase expander-based liquefaction process adopting  $N_2$ – $C_3H_8$  mixed refrigerant. They investigated the potential of using a two-phase cryogenic expander to generate a cooling effect. Their results showed that energy consumption and refrigerant flow rate reduced up to 46% and 28%, respectively, compared to those of a conventional nitrogen expander process. He and Ju [15] proposed four modification strategies for the expander-based liquefaction process to improve its energy efficiency: 1) using multistage expanders; 2) adding a single precooling cycle; 3) using an additional heat exchanger to subcool the refrigerant before expansion; and 4) replacing the single working fluid refrigerant with a mixture of working fluid refrigerant. They determined that a parallel nitrogen expander process with a R410A (a mixture of difluoromethane and pentafluoroethane) precooling cycle had lowest energy consumption compared to that of other processes.

Adding a vapor compression precooling cycle, with propane, propylene, or carbon dioxide as refrigerant, to the expander-based liquefaction process can efficiently reduce energy consumption [4]. Adding an ammonia absorption precooling cycle, which is driven by available waste heat, would further reduce energy consumption. Mehrpooya et al. [16] proposed a novel mixed-fluid cascade process with ammonia absorption precooling. The ammonia absorption precooling cycle was driven by waste heat (180 °C and 550 kPa) from the liquefaction plant. The simulation results showed that the precooling system helped to reduce power consumption and the required heat transfer area by 30% and 31%, respectively. Mortazavi et al. [17] improved the energy efficiency of a propane pre-cooled mixed refrigerant process by absorption chillers. They used the gas turbine exhaust, which drives the compressors, to provide the required waste heat (180 °C) for the absorption chillers (demand is 97% of gas turbine waste heat). This improvement reduced energy consumption by 21%. Rodgers et al. [18] evaluated a propane pre-cooled mixed refrigerant process enhanced by three types of waste heat driven absorption chillers: single-effect, double-effect, and cascaded single- and double-effect chillers. They used actual operating data to determine the thermodynamic performance of each improvement. Their results showed that the required waste heat for absorption chillers could be recovered from a single gas turbine, and the coefficient of performance (ratio of cooling capacity to compressor work) and cooling capacity were increased by 13% and 23%, respectively. The studies above clearly show the benefit of adding an absorption cycle to the liquefaction process. However, they all focus on the mixed refrigerant or cascade processes, which are both large-scale liquefaction processes. The techno-economic performance for adopting an absorption pre-cooling cycle to small-scale expander-based processes is still unclear. This paper aims to fill that gap.

This study will investigate potential improvement options with the goal of improving the techno-economic performance of expander-based liquefaction processes. Based on the literature review above, three strategies were investigated: 1) use of mixed refrigerant; 2) use of two-phase expander; and 3) adding a pre-cooling cycle. This paper applied these strategies by incorporating an ammonia absorption precooling cycle to a conventional nitrogen expander process and to a novel open-loop expander process. The conventional nitrogen expander process was set up for model validation and comparison purpose. The novel open-loop expander process uses natural gas as refrigerant and replaces the gas expander with a two-phase expander. The processes were simulated and optimized in Aspen Plus. The optimization of the proposed expander-based processes was done by two objective functions: minimizing specific energy consumption and

minimizing production cost. Based on the results, the optimal small-scale expander-based LNG plant is proposed.

## 2. Process design and description

Based on improvement options found in the literature review for expander-based process, four processes are designed in order to investigate the potential improvement. The description of the process design starts from ammonia absorption precooling cycle followed by nitrogen expansion process and methane expansion process.

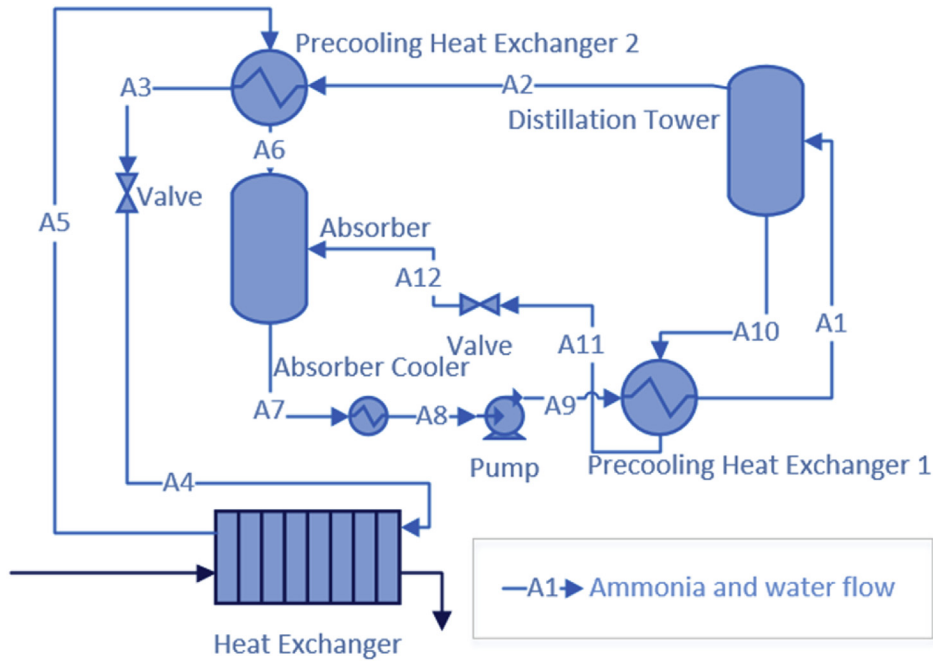
### 2.1. Ammonia absorption precooling cycle

Ammonia absorption refrigeration is a vapor refrigeration process. It uses a pump instead of a conventional compressor, thereby significantly reducing compression work [19,20]. This is possible by dissolving the ammonia refrigerant in water before the pressurization step.

The ammonia absorption precooling cycle used in this study is a single effect absorption cycle, which was developed based on [16] (Scheme 1). A concentrated ammonia-water solution (A1: 25% mass fraction) is boiled at a distillation tower, and separated in a high-pressure ammonia gas flow (A2: 1300 kPa) and a diluted ammonia-water solution (A10: 0.01% mass fraction). Stream A2 is cooled in a heat exchanger (Precooling Heat Exchanger 2), and expanded in a valve to a low-pressure and low-temperature flow (A4: 120 kPa and  $-29.5$  °C). Next, stream A4 is evaporated in a heat exchanger (Heat Exchanger) to provide cooling duty. Stream A6 is absorbed by the diluted ammonia-water solution (A12) in the absorber. Lastly, a concentrated ammonia-water solution (A8) is pumped to a high-pressure stream (1300 kPa) and feeds back to the distillation tower. The cycle is powered by low-pressure steam ( $T = 180$  °C and  $p = 550$  kPa), which is produced from available waste heat of the gas turbine exhaust [16,17]. The coefficient of performance (ratio of cooling duty to waste heat needed) of this cycle is 0.485. Mortazavi et al. [17] point out that the available amount of waste heat of the gas turbine is higher than the compressor power provided by the gas turbine. Therefore, if the required waste heat for absorption precooling cycle is less than the power required by the compressor, the process does not need additional heat input and can be self-sufficient.

### 2.2. Nitrogen expansion process

In this study, the nitrogen expansion processes include single nitrogen expansion process (SN) and single nitrogen expansion process with absorption precooling (SNA). At present, SN (see Scheme 2) is the most commonly used process in small-scale liquefaction plants [21]. In the nitrogen cycle of SN, a nitrogen flow (N1) is first compressed through two compressors (Compressor 1 and Compressor 2), and then cooled by a heat exchanger (Heat Exchanger 1) to become a high-pressure, medium-temperature flow (N6). Next, stream N6 goes through an expander to become a low-pressure, low-temperature flow (N7). Then, stream N7 provides cooling duty in two heat exchangers (Heat Exchanger 1 and Heat Exchanger 2). Lastly, stream N1 goes back to the compressors. In the natural gas cycle of SN, a natural gas flow (1) is first cooled in two heat exchangers, and then expanded in a valve. Lastly, stream 4 is separated in a separator as flash gas (5) and LNG. The flash gas is returned to the heat exchanger to recover the cold energy. This SN process is used as a base case. The difference between SN and SNA is that SNA adds a heat exchanger for the absorption precooling cycle (see Scheme 3).



Scheme 1. Ammonia absorption precooling cycle.

2.3. Methane expansion process

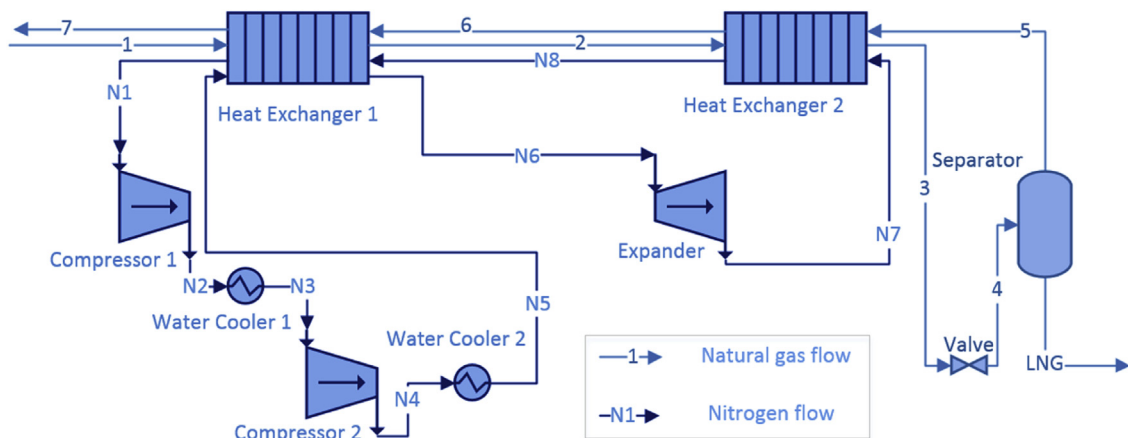
The methane expansion processes proposed in this study were inspired by the ZR-LNG process [22], which is an open-loop process using the natural gas (mainly methane) itself as refrigerant. The methane expansion processes include single methane expansion process (SM) and single methane expansion process with absorption precooling (SMA). In the SM process (see Scheme 4), a feed natural gas (NG) and a refrigerant natural gas flow (15) are first mixed in a mixer, and then the mixed flow (1) is compressed and cooled to a high-pressure, medium-temperature flow (4). Next, stream 4 goes to a two-phase expander to expand to a low-temperature, low-pressure flow (5). Then, stream 5 is separated as a liquid flow (6) and a gas flow (8) in a separator (Separator 1). Stream 8 goes back to the heat exchanger to provide cooling duty. Next, stream 9 is compressed to the same pressure as the feed natural gas. Lastly, stream 7 is expanded in a valve and separated as flash gas (14) and LNG in the separator (Separator 2). In addition,

the flash gas goes back to the heat exchanger to recover the cold energy. The difference between SM and SMA is that SMA adds a heat exchanger for the absorption precooling cycle (see Scheme 5).

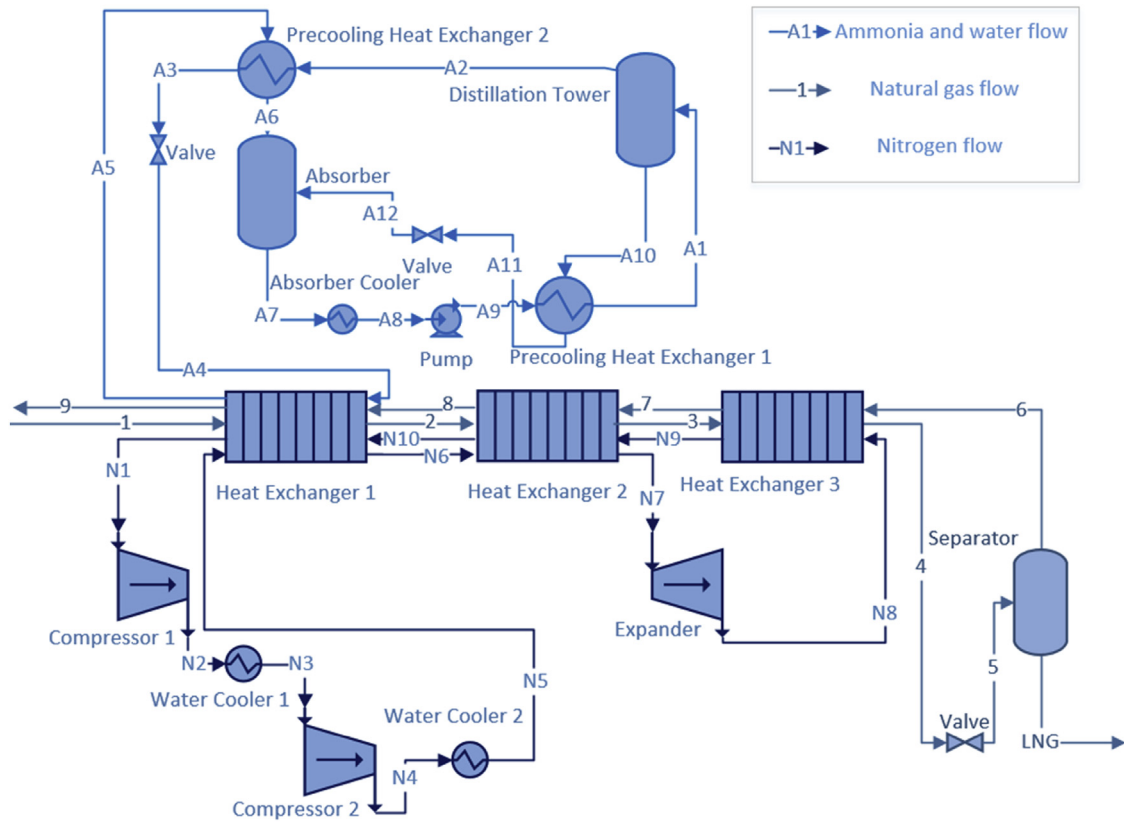
2.4. Feed gas conditions and simulation assumptions

In this study, the four processes (SN, SNA, SM, and SMA) are intended for a small-scale LNG plant located in a remote area without pipeline infrastructure. The feed gas conditions are listed in Table 1, which is obtained from Yuan et al. [4]. In order to compare the performance of the four processes, the liquefaction ratio is kept the same at 85% [23]. The liquefaction capacity of the small-scale LNG plant is set as 0.85 kg/s (0.025 million tonne per annual with an availability of 93.2% [24]).

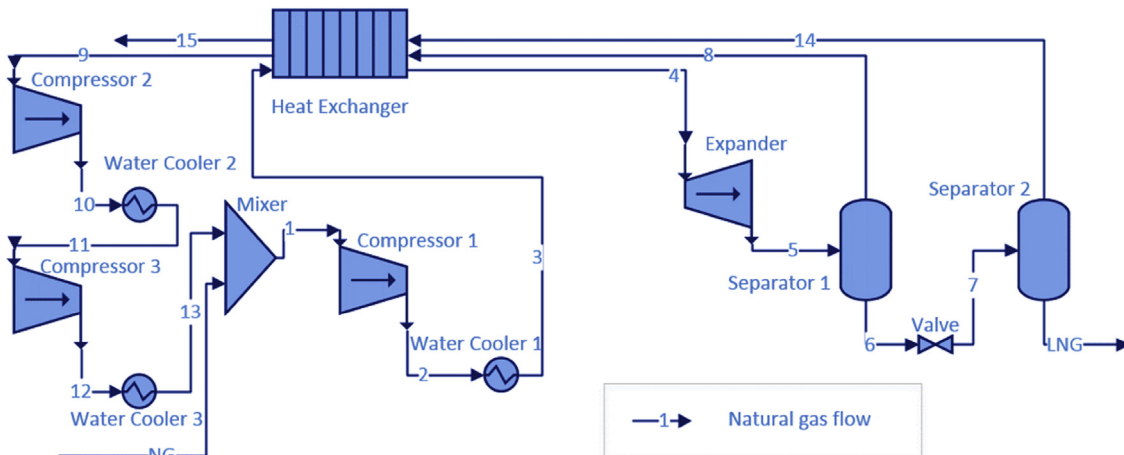
The four processes were simulated in Aspen Plus V8.6 with steady-state conditions. Peng-Robinson Equation of State was selected as the phase equilibrium equation, because it is suitable for gas, refinery, and petrochemical applications [3,4]. The steam for



Scheme 2. SN flowsheet.



Scheme 3. SNA flowsheet.



Scheme 4. SM flowsheet.

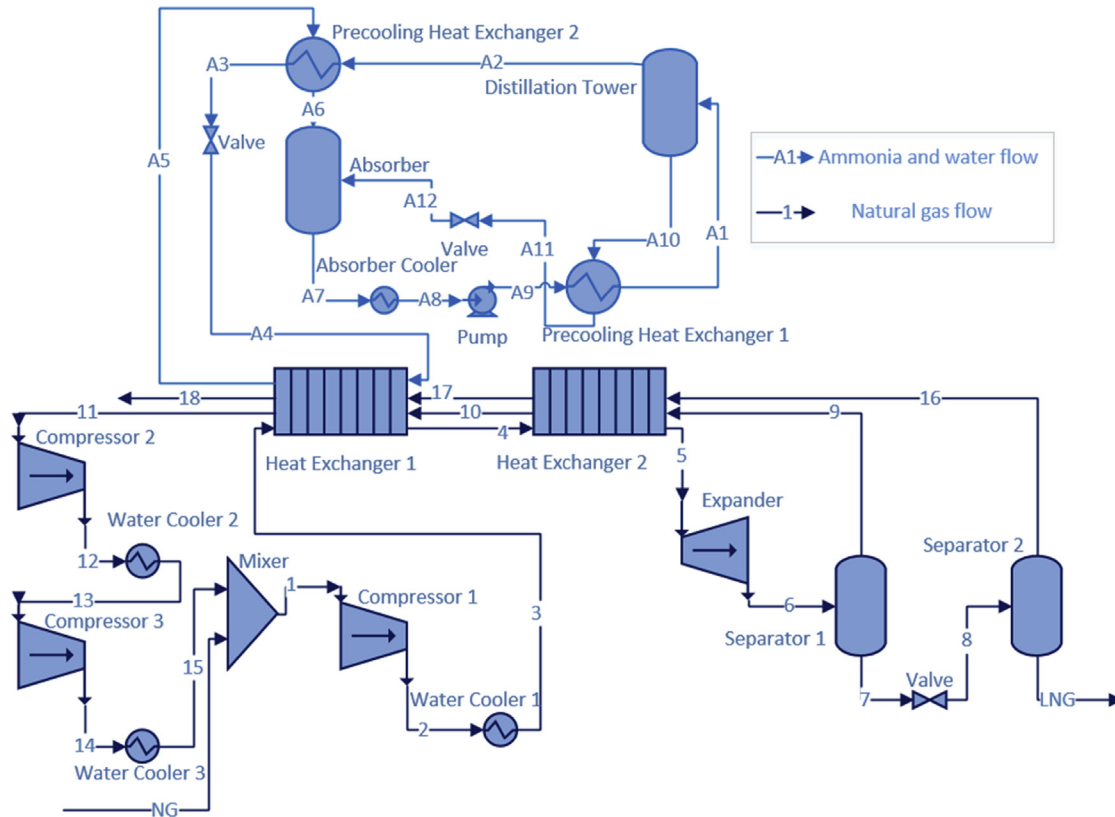
the absorption precooling cycle will be provided by the available waste heat of the gas turbine [16,17]. Two-phase expanders were used in SM and SMA process to allow liquid formulation at the expander outlet. It is assumed that the two-phase expander has the same efficiency and cost compared to those of a gas expander (as reported in Refs. [9,25,26]). Furthermore, it is assumed that the energy recovered by expander is used to drive the compressors without any losses. Based on [27], the mechanical efficiency of gas turbine for small-scale is much lower than that of large-scale ones. The assumptions listed in Table 2 are widely used in small-scale liquefaction process simulation, which makes the simulation

results of this study comparable to those of other studies.

### 3. Optimization and analysis methods

#### 3.1. Optimization methods

The steady-state optimizer embedded in Aspen Plus was used to conduct optimization for the four processes. The well-known Complex algorithm was selected as the optimization algorithm to give a global optimum [32]. The Complex algorithm is a pattern search algorithm, and its descriptive search routine is shown in



Scheme 5. SMA flowsheet.

Table 1  
Feed gas conditions.

Parameters	Value
NG composition (mole fraction)	[CH <sub>4</sub> , C <sub>2</sub> H <sub>6</sub> , C <sub>3</sub> H <sub>8</sub> , i-C <sub>4</sub> H <sub>10</sub> , n-C <sub>4</sub> H <sub>10</sub> , N <sub>2</sub> ] = [82.0%, 11.2%, 4.0%, 1.2%, 0.9%, 0.7%]
Feed NG temperature	32 °C
Feed NG pressure	4800 kPa
Feed NG mass flow rate	1 kg/s

Fig. 1 [33]. The pattern search involves two moves: exploratory move and pattern move [34]. The algorithm starts with an initial guess of the relevant parameters and calculates the objective function value (Base Point). The parameters are then varied

according to the step size (Exploratory move) and the objective function is recalculated. If the outcome is better than the previous result then the changed parameters are altered in a larger step accordingly (Pattern move). The parameters are then changed again according to the step size (Exploratory move) and the cycle repeats. When after an Exploratory move the objective function value has not improved then the step size is reduced and the cycle is repeated. This continues until the step size drops below the minimum step size at which point the optimization finishes. An LNG process is an energy-intensive process, so it is important to optimize the process with low energy consumption. However, it is recommended by several studies [3,11,35–37] that cost-related optimization is as important as energy-related optimization. Therefore, two different objective functions were used in this study

Table 2  
Simulation assumptions.

Parameters	Value	Reference
Heat loss and pressure drop in pipeline and heat exchanger	None	[15,28]
Adiabatic efficiency for compressors	85%	[4,10]
Adiabatic efficiency for gas expanders and two-phase expanders	80%	[4,9,10]
Refrigerant temperature after the water cooler	35 °C	[4]
Precooling cycle evaporation temperature	−29.6 °C	[16]
LNG mass flow rate	0.85 kg/s	–
Liquefaction ratio	85%	[23]
LNG storage pressure	200 kPa	[3,4,13,29]
Discount rate ( <i>r</i> )	12%	[30]
Plant life ( <i>n</i> )	20 year	[30]
Installation factor ( <i>F</i> )	4.74	[31]
Operation and maintenance cost factor ( <i>i</i> )	2%	[3]
Mechanical efficiency of gas turbine ( $\eta_{GT}$ )	25%	[27]
Availability ( <i>t</i> )	8160 h/yr	[24]
Unit fuel natural gas cost ( $C_{fuel}$ )	2.75 \$/GJ	[30]

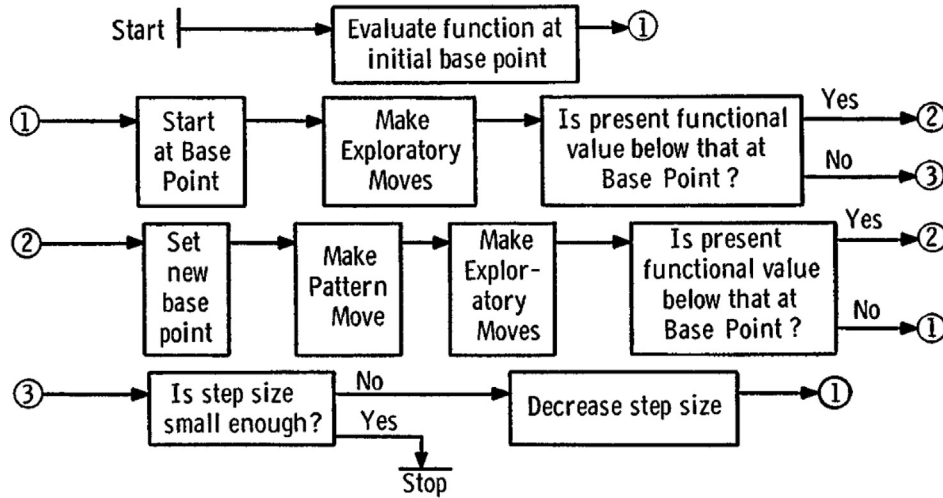


Fig. 1. Descriptive search routine of Complex algorithm [33].

for optimization. Firstly, the processes were optimized to minimize the specific energy consumption. Secondly, the processes were optimized to minimize production cost. Costs are indexed to \$<sub>2018</sub> using the Chemical Engineering Plant Cost Index (CEPCI).

The variables for the optimization include the mass flows of the precooling and liquefaction cycles, the inlet temperature of the expander, the outlet pressure of the expander, and the outlet pressure of specific compressors. Table 3 shows the variables with their lower and upper bounds for the four processes and step size for optimization.

### 3.2. Specific energy consumption

The first objective function is minimization of specific energy consumption (OBJ1: minimum energy). *SEC* is the specific energy consumption (kJ/kg), which is calculated via Eq. (1):

$$SEC = \frac{W_{compressor} - W_{expander}}{m_{LNG}} \quad (1)$$

where  $W_{compressor}$  is the energy consumption of the compressors (kW),  $W_{expander}$  is the energy recovered by the expander (kW), and  $m_{LNG}$  is the mass flow of the produced LNG (kg/s).

### 3.3. Production cost

The second objective function is minimization of production cost (OBJ2: minimum cost) by optimizing the variables in Table 3. *SPC* is the production cost (\$/kg), which is calculated as Eq. (2):

$$SPC = C_{amortized\ capex} + C_{amortized\ opex} \quad (2)$$

where  $C_{amortized\ capex}$  is the amortized capital cost, which is the capital cost of the plant to produce 1 kg LNG (\$/kg) considering discount rate and plant life.  $C_{amortized\ opex}$  is the amortized operating cost (\$/kg).

The amortized capital cost is calculated via Eq. (3):

$$C_{amortized\ capex} = C_{specific\ capex} * \left( \frac{r*(1+r)^n}{(1+r)^n - 1} \right) \quad (3)$$

The discount rate ( $r$ ) and plant life ( $n$ ) are assumed to be 12% and 20 years [30], respectively.  $C_{specific\ capex}$  (\$/(kg/year)) is the specific capital cost (Eq. (4)), which is calculated as  $C_{capex}$  (\$) divided by capacity  $Y$  (kg/year).  $C_{capex}$  (\$) is the total capital cost, which is the total purchased-equipment cost multiplied by an installation factor ( $F$ ) [31]:

$$C_{specific\ capex} = C_{capex} / Y = F * \sum PEC / Y \quad (4)$$

$PEC$  (\$) is the purchased-equipment cost for individual equipment [3,38], which is estimated based on a factorial costing technique and calculated using Eq. (5):

$$PEC = a + b * S^m \quad (5)$$

where  $a$ ,  $b$ ,  $m$  are constants for individual equipment, and  $S$  is the capacity of the individual equipment (see Table 4).

The amortized operating cost (Eq. (6)) is calculated as the operating cost (\$/year) divided by capacity  $Y$  (kg/year). The operating cost includes operation and maintenance cost (\$/year) and fuel cost (\$/year):

Table 3  
Optimized variables of four processes with lower and upper bounds including step size.

Process	SN	Range	SNA	Range	SM	Range	SMA	Range
Mass flow refrigerant in of precooling cycle (kg/s)	—	—	$m_{A1}$	1–10	—	—	$m_{A1}$	1–10
Mass flow refrigerant in of liquefaction cycle (kg/s)	$m_{N1}$	1–10	$m_{N1}$	1–10	$m_1$	1–10	$m_1$	1–10
Expander inlet temperature (°C)	$T_{N6}$	–70 to –30	$T_{N7}$	–70 to –30	$T_4$	–70 to –30	$T_5$	–70 to –30
Expander outlet pressure (kPa)	$P_{N7}$	200–1000	$P_{N8}$	200–1000	$P_5$	300–2000	$P_6$	300–2000
Intermediate compressor outlet pressure (kPa)	$P_{N2}$	600–1600	$P_{N2}$	600–1600	$P_{10}$	1200–3200	$P_{12}$	1200–3200
High compressor outlet pressure (kPa)	—	—	—	—	$P_2$	6000–9000	$P_2$	6000–9000
Step size		0.01%–1% of the range						

$$C_{amortized\ opex} = C_{opex} / Y = (C_{OMC} + C_{FC}) / Y \quad (6)$$

$$C_{OMC} = i * \sum PEC \quad (7)$$

$$C_{FC} = \frac{W_{total} * t * C_{fuel} * 3600}{\eta_{GT} * 1000000} \quad (8)$$

The operation and maintenance cost ( $C_{OMC}$ ) is assumed to be the total purchased equipment cost multiplied by an operation and maintenance cost factor ( $i$ ) (Eq. (7)) [3]. Fuel cost ( $C_{FC}$ ) is calculated using Eq. (8).  $W_{total}$  is the total compressor power minus expander power (kW),  $\eta_{GT}$  is the mechanical efficiency of gas turbine,  $t$  is availability (hr/yr), and  $C_{fuel}$  is unit fuel natural gas cost (\$/GJ), using a lower heating value is 53.18 GJ/t [42–44].

The net present value (NPV) is used to determine the discounted profitability of the processes in this study. It is calculated using Eq. (9) [45,46].  $x$  is the related year of cash flow,  $C_{LNG}$  is the freight on board price of the produced LNG (0.25 \$/kg) [47], and  $C_{NG}$  is feed price of used NG (0.16 \$/kg) [30].

$$NPV = -C_{capex} + \sum_{x=1}^n \frac{((C_{LNG} - C_{NG}) * Y - C_{FC} - C_{OMC})}{(1 + r)^x} \quad (9)$$

### 3.4. Exergy analysis

Exergy is a measure of the maximum available energy that can be used when bringing a process stream to its thermodynamic equilibrium state [14,48]. Exergy analysis can evaluate the magnitude of exergetic losses of each component and points out the direction for potential improvement. In this study, only the physical exergy is taken into account, which is defined by stream enthalpy ( $H$ ) and entropy ( $S$ ). Potential and kinetic exergy is ignored as is common in steady-state systems [4]. Chemical exergy is also ignored because no chemical reactions take place. The equilibrium state is set to the environmental conditions, which are 20 °C and 101.325 kPa. The exergy  $Ex$  (kW) is expressed in Eq. (10).

$$Ex = me = m \left( (h - T_0s)_{T,p} - (h - T_0s)_{T_0,p_0} \right) \quad (10)$$

where the  $T_0$  and  $p_0$  are the equilibrium state temperature and pressure,  $h$ ,  $s$  and  $e$  are specific stream enthalpy (kJ/kg), entropy (kJ/kg\*K) and exergy (kJ/kg), respectively.  $m$  is the mass flow (kg/s) of the stream. The exergy is calculated based on 1 kg LNG production.

In this study, the exergy losses are calculated by performing an exergy balance equation over the compressor, heat exchanger, expander, and valve [49,50]. The energy and exergy equations for the different equipment are listed in Table 5 [49,51].

The exergy efficiency is the ratio of exergy recovered to exergy supplied [52]. For the proposed four processes, the exergy efficiency is calculated as the ratio of the minimum work required to produce 1 kg LNG (exergy recovered) to the actual work required

and exergy supplied by waste heat (exergy supplied). It is expressed in the following equations: Eqs. (11), (12) and (13).

$$\eta_{Ex} = \frac{Ex_r}{Ex_s} \quad (11)$$

$$\eta_{Ex} = \frac{W_{min}}{SEC + Ex_{s,waste\ heat}} \quad (12)$$

$$W_{min} = (h_{LNG} - h_{NG}) - T_0(s_{LNG} - s_{NG}) \quad (13)$$

### 3.5. Constraints

The following constraints were used during the optimization:

- The minimum pinch temperature in the heat exchangers is 5 °C
- The compression ratio of all compressors must be between 1.5 and 4.0 [35].
- When multiple cold streams exit the same multi-stream heat exchanger, they must have the same outlet temperature. The same holds for hot streams [35].
- The maximum system pressure of SN and SNA is set at 2500 kPa based on [4], and the maximum system pressure of SM and SMA is set at 9000 kPa based on [28,53].
- Different streams entering the same mixer must have the same inlet pressure

For the SN and SNA process, constraints a) – d) were used, while constraints a) – e) were used for the SM and SMA process.

## 4. Results and discussions

To analyze the technical and economic performance, the four processes were optimized with two objective functions: minimization of specific energy consumption (OBJ1: minimum work) and minimization of production cost (OBJ2: minimum cost). The optimized variables of each process are shown in Tables 6 and 7. Detailed flow information of the optimized processes can be found in the Appendix A. The comparison between processes without precooling and processes with precooling shows that an absorption precooling cycle reduces the refrigerant mass flow as well as the expander inlet temperature. The most significant difference in the optimization variables between the two objective functions is that the intermediate compressor outlet pressure in OBJ2 is higher than in OBJ1.

### 4.1. Specific energy consumption

The energy consumption of compressors and expander, and required waste heat are shown in Table 8, and the specific energy consumption for each process is shown in Fig. 2. The required heat for the absorption precooling cycle is less than the total work for

**Table 4**  
Purchased-equipment cost constants.

Equipment	a	b	m	S	Reference
Compressor	–	430360	0.69	Volumetric flow rate (m <sup>3</sup> /s)	[11,39]
Expander	–	554	0.81	Expansion work (kW)	[40]
Pump	9054	272	0.90	Volumetric flow rate (L/s)	[38]
Heat exchanger	14783	370	0.80	heat transfer area (m <sup>2</sup> )	[41]
Vessel (tower)	2077	2481	0.85	Volume (m <sup>3</sup> )	[31,38]



**Table 5**  
Energy and exergy balance of each equipment.

Equipment	Energy balance	Exergy balance
Compressor	$W_C = m(h_o - h_i)$	$Ex_{C, loss} = Ex_i - Ex_o = \sum (me)_i + W - \sum (me)_o$
Heat exchanger	$Q_H = \sum (mh)_o - \sum (mh)_i$	$Ex_{H, loss} = Ex_i - Ex_o = \sum (me)_i - \sum (me)_o$
Expander	$W_E = m(h_o - h_i)$	$Ex_{E, loss} = Ex_i - Ex_o = \sum (me)_i - W - \sum (me)_o$
Valve	$h_i = h_o$	$Ex_{V, loss} = Ex_i - Ex_o = \sum (me)_i - \sum (me)_o$

**Table 6**  
The optimized variables of SN and SNA with two objective functions.

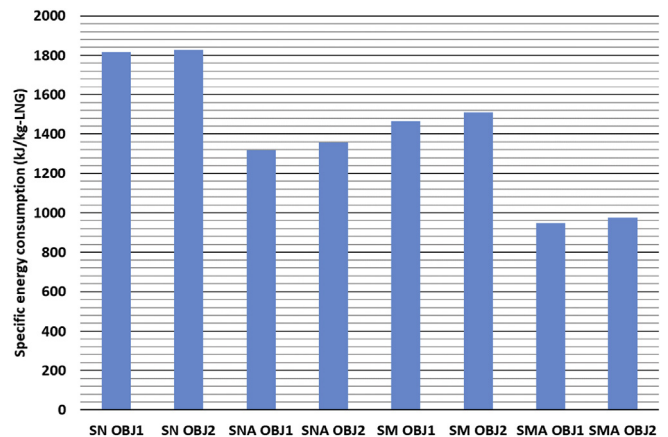
Process	SN			SNA		
		OBJ1	OBJ2		OBJ1	OBJ2
Objective function						
Mass flow of refrigerant in precooling cycle (kg/s)	–	–	–	$m_{A1}$	1.57	1.56
Mass flow of refrigerant in liquefaction cycle (kg/s)	$m_{N1}$	7.77	7.77	$m_{N1}$	6.49	6.44
Expander inlet temperature (°C)	$T_{N6}$	–41.22	–41.16	$T_{N7}$	–53.28	–53.19
Expander outlet pressure (kPa)	$P_{N7}$	255.56	255.21	$P_{N8}$	332.78	325.75
Intermediate compressor outlet pressure (kPa)	$P_{N2}$	821.57	977.02	$P_{N2}$	984.09	1303.00

**Table 7**  
The optimized variables of SM and SMA with two objective functions.

Process	SM			SMA		
		OBJ1	OBJ2		OBJ1	OBJ2
Objective function						
Mass flow of refrigerant in precooling cycle (kg/s)	–	–	–	$m_{A1}$	1.99	2.00
Mass flow of refrigerant in liquefaction cycle (kg/s)	$m_1$	4.85	4.83	$m_1$	3.40	3.42
Expander inlet temperature (°C)	$T_4$	–52.87	–52.74	$T_5$	–58.90	–58.92
Expander outlet pressure (kPa)	$P_5$	766.81	756.30	$P_6$	764.98	773.73
Intermediate compressor outlet pressure (kPa)	$P_{10}$	1972.45	3063.67	$P_{12}$	1978.26	2982.43
Last compressor outlet pressure (kPa)	$P_2$	7200.35	7218.68	$P_2$	7200.05	7204.93

the compressor in SNA and SMA, indicating that for both processes the gas turbine exhaust has enough waste heat available. It is clear that the specific energy consumption of the processes with precooling is much lower than the processes without precooling. Under the two objectives, energy consumption is 26–27% lower for SNA compared to SN and 35% lower for SMA compared to SM. It is also clear that the specific energy consumption of methane expansion process is much lower than nitrogen expansion process. Under the two objectives, energy consumption is 18–19% lower for SM compared to SN and 28% lower for SMA compared to SNA. For each process, the specific energy consumption is slightly higher under OBJ2 than under OBJ1 (0.7–3.1%).

The results above can be explained by the cold and hot composite curves shown in Figs. 3 and 4. The processes with precooling cycle have a smaller difference in temperature between the cold and hot composite curves compared to that of the process without precooling cycle, resulting in lower energy consumption. The difference in temperature between the cold and hot composite curves for methane expansion processes is similar to the nitrogen expansion processes. However, because of the utilization of a two-phase expander, the energy consumption of the methane



**Fig. 2.** Specific energy consumption for each process.

expansion processes is lower than the nitrogen processes. The two-phase expander can recover the pressure exergy within the feed

**Table 8**  
The energy consumption for SN, SNA, SM and SMA with two objective functions for 0.85 kg/s LNG production.

Process	SN		SNA		SM		SMA	
	OBJ1	OBJ2	OBJ1	OBJ2	OBJ1	OBJ2	OBJ1	OBJ2
Compressor 1 (kW)	1130	1357	865	1148	325	326	223	225
Compressor 2 (kW)	1098	881	747	496	701	1088	438	658
Compressor 3 (kW)	–	–	–	–	654	306	405	206
Expander (kW)	–684	–684	–491	–491	–436	–437	–259	–260
Total work (kW)	1544	1554	1121	1153	1245	1282	806	828
Required waste heat (kW)	–	–	438	435	–	–	556	558

natural gas in SM and SMA, which is wasted at the valve in SN and SNA. The two-phase expander can also provide cooling for natural gas. The natural gas is cooled to only around  $-60\text{ }^{\circ}\text{C}$  at a multi-stream heat exchanger in SM and SMA. It then expands in the two-phase expander to around  $-130\text{ }^{\circ}\text{C}$ . This isentropic expansion in the two-phase expander results in a higher cooling capacity and provides additional power [9]. In contrast, in SN and SNA the natural gas needs to be cooled to around  $-130\text{ }^{\circ}\text{C}$  at the multi-stream heat exchanger. Therefore, the energy consumption for the methane expansion processes is much lower than the nitrogen expansion processes. Lastly, the slightly higher energy consumption of OBJ2 compared to OBJ1 is caused by an increase in the intermediate compressor outlet pressure, which makes the increase in energy consumption in the intermediate compressor exceed the decrease in energy consumption in the successive compressor.

4.2. Production cost

The production cost for each process is shown in Fig. 5. It is clear that the processes with precooling have not only lower specific energy consumption but also lower production cost than those of the processes without precooling. The absorption precooling cycle in SNA helps to reduce both capital cost and operating cost compared to those of SN. SNA is 16–17% lower in production cost than that of SN under two objectives. The absorption precooling cycle in SMA slightly increases the capital cost, but it reduces operating cost significantly compared to those of SM. SMA has 13–14% lower in production cost than that of SM under two objectives. It is also clear that the methane expansion processes have lower production cost than that of the nitrogen expansion processes: SM is 32–33% lower than SN, and SMA is 30% lower than SNA under two objectives. Despite the higher energy consumption of OBJ2 for all process, the production cost of OBJ2 is 1.7–2.3% lower

compared to that of OBJ1. This means that minimization of specific energy consumption may not lead to the lowest production cost.

The annual operating cost breakdown is shown in Fig. 6. Specific energy consumption significantly influences operating cost, because a large amount of the operating cost is the fuel cost (92–97% of operating cost). Therefore, the operating cost is dominated by energy efficiency.

The total capital cost breakdown is shown in Fig. 7. The capital cost of compressors dominates in total capital cost (51%–85%). It can be seen that SNA has a lower capital cost than that of SN, and SMA has almost the same capital cost compared to that of SM. Although an absorption precooling cycle adds additional capital cost for the heat exchanger and tower, it reduces the capital cost of the compressors and expander because adding an absorption precooling cycle provides cooling duty to reduce the volumetric flow rate of refrigerant in the liquefaction cycle. This flowrate determines the compressor size and therefore affects compressor capital cost [11]. It is clear that the methane expansion processes have lower capital cost than nitrogen expansion processes, especially for the compressor and heat exchanger. The utilization of two-phase expander in the methane expansion processes provides additional cooling effect to reduce the volumetric flow rate of refrigerant compared to that of the nitrogen expansion processes, thereby reducing the capital cost of compressor and heat exchanger. It can also be seen that the total capital cost for OBJ2 is lower than OBJ1 for each process. This is mainly caused by cost reduction in the compressor. The increase of the intermediate compressor outlet pressure in OBJ2 compared to that of OBJ1 reduces the volumetric flowrate in the successive compressor, thereby reducing its cost. However, it will also increase energy consumption as discussed before. A trade-off between energy consumption and capital cost is optimized in OBJ2 to find the minimum production cost. Although OBJ2 has higher specific

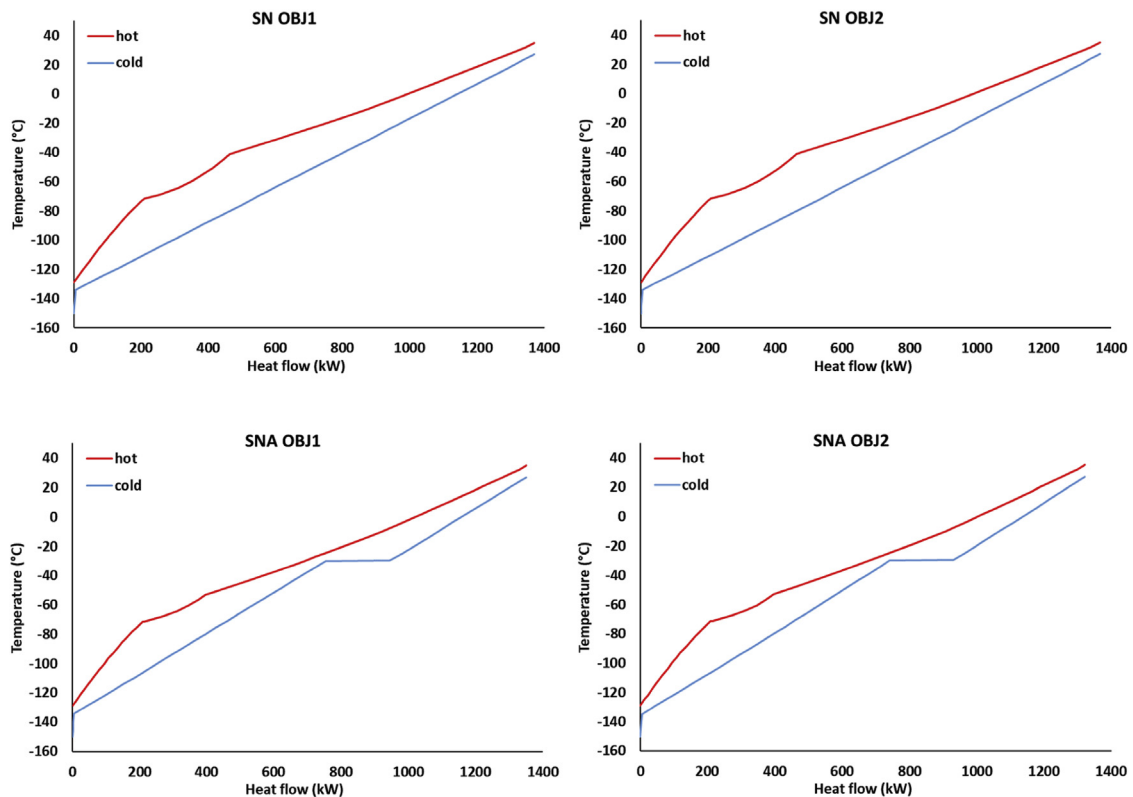


Fig. 3. Cold and hot composite curves of multi-stream heat exchangers in SN and SNA.

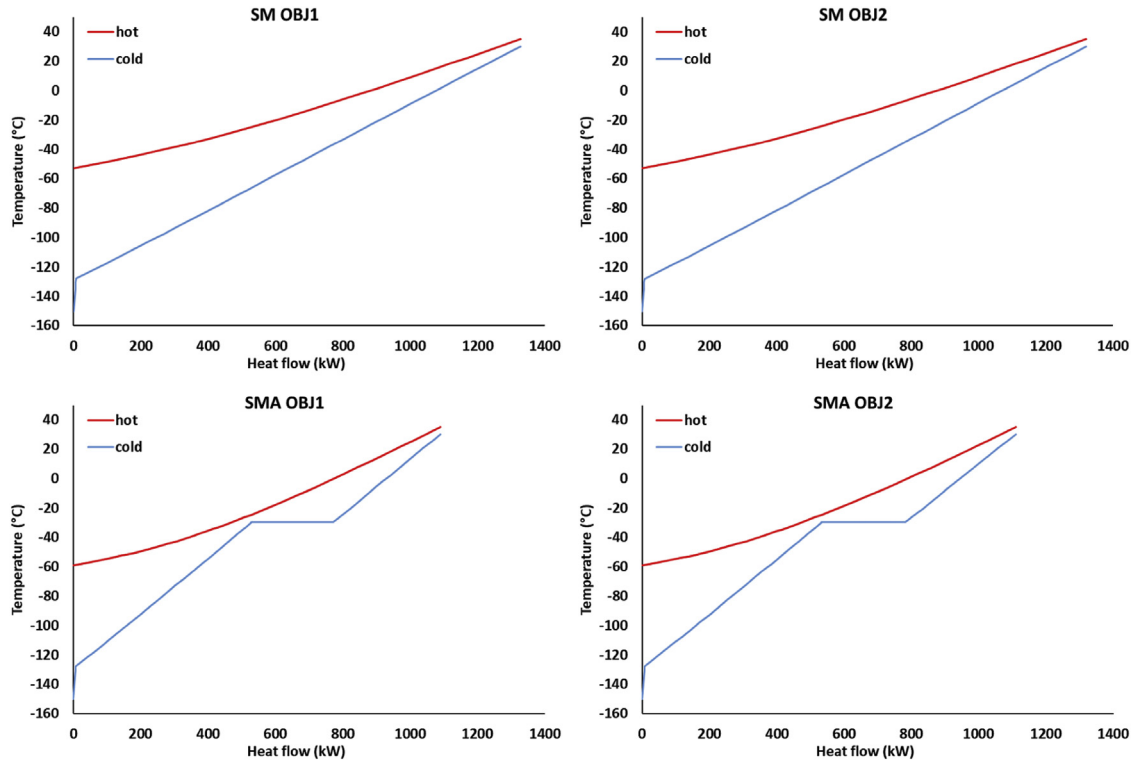


Fig. 4. Cold and hot composite curves of multi-stream heat exchangers in SM and SMA.

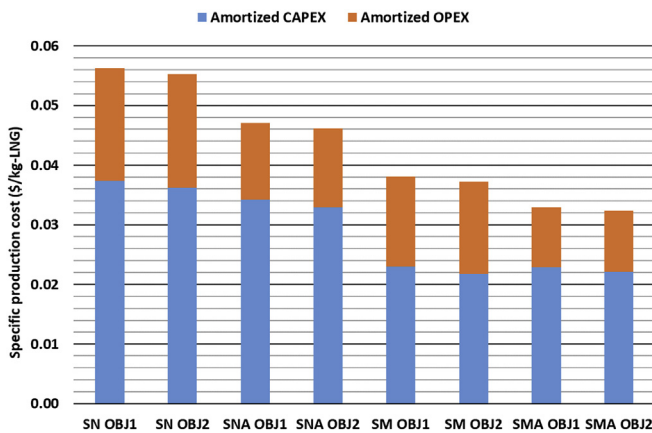


Fig. 5. Specific production cost for each process.

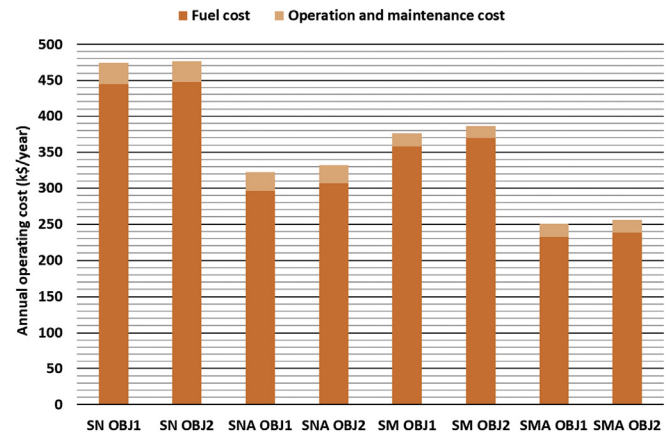


Fig. 6. Annual operating cost break down for each process.

energy consumption than OBJ1, OBJ2 can still achieve lower production cost by reducing capital cost.

The NPV of the four processes under OBJ1 and OBJ2 is shown in Table 9. Results show that adding an ammonia precooling cycle increases the NPV. As expected the minimization of production cost (OBJ2) results in a higher NPV than minimization of specific work (OBJ1). Lastly, the methane expansion processes have a higher NPV compared to the nitrogen expansion processes. These results are in line with the results from the energy analysis.

### 4.3. Exergy analysis

The exergy losses of major equipment and exergy efficiency for each process is shown in Fig. 8. Adding precooling cycle reduces exergy losses in the compressor, heat exchanger and expander

because of flowrate reduction of refrigerant. Adding precooling cycle also increases the exergy efficiency of the process, because exergy supplied by waste heat is less than the reduction in specific energy consumption. The methane expansion processes have lower exergy losses and higher exergy efficiency than that of the nitrogen expansion processes, mainly because utilization of the two-phase expander recovers the pressure exergy within natural gas. The exergy losses in the heat exchanger in the methane expansion processes are higher than that of the nitrogen expansion processes because of the higher temperature difference between cold and hot stream. The exergy losses in the valve of the methane expansion processes are lower than that of the nitrogen expansion processes because the methane expansion process already partly expanded the natural gas in its expander. It should be noted that although SNA has lower specific energy consumption than SM, its exergy

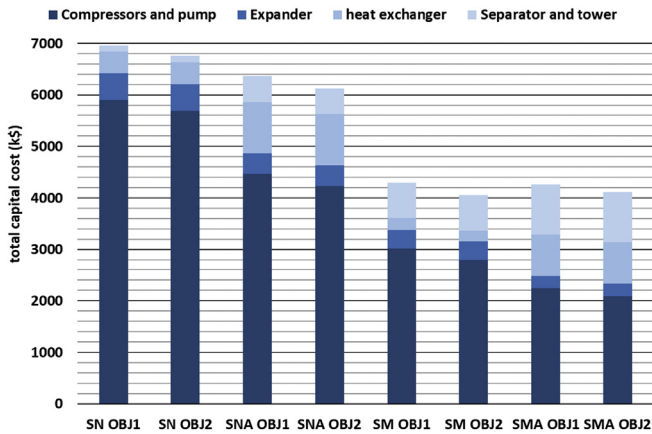


Fig. 7. Total capital cost break down for each process.

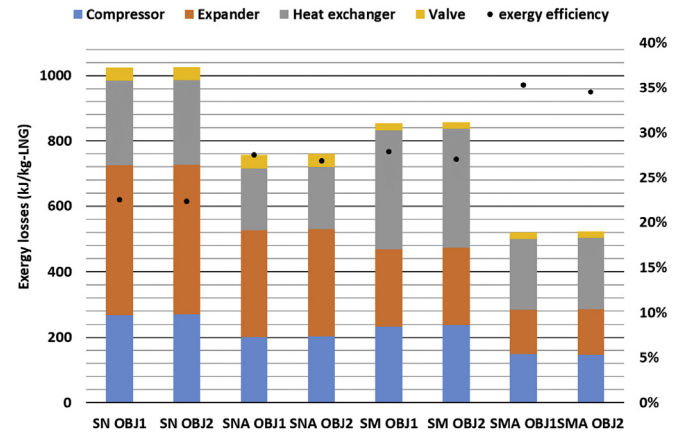


Fig. 8. Exergy losses of major equipment and exergy efficiency for each process.

efficiency is lower than SM because of additional exergy supplied by waste heat.

4.4. Study limitations

The results of this study have to be seen in the light of some limitations. The primary limitation is the limited accuracy of equipment capital cost estimation because of the difficulty of capital cost estimation of equipment [54–57]. According to Symister [58], capital cost estimation using different estimating methods, including the method used in this study, can greatly vary by –30 to +50% for most equipment. The factorial costing method is chosen for this study, because it provides the available cost data of all the equipment used in this study and its average scale factor is medium among that of other methods [58]. Future work in capital cost estimation may include updating the vendor cost data and improving cost curve accuracy. The second limitation is that the efficiency and cost of a two-phase expander are assumed identical to a gas expander based on [9]. However, data on two-phase expander are limited because it is a relatively novel equipment. The application of two-phase expander in Poland [25] showed that it could significantly improve the energy efficiency of an LNG process resulting in a payback period of under six months. The third limitation is that the processes are simulated with no heat loss and pressure drop. Heat loss and pressure drop will definitely increase the energy consumption of LNG process. For example, the SN process with 2% heat loss or 2% pressure drop in the heat exchanger results in around 4% increase in specific energy consumption for each. Because most of the studies that focus on LNG process simulation neglect heat loss and pressure drops, the assumption with no heat loss and pressure drop is chosen in this study to make the results comparable with those of previous studies.

5. Conclusion

The technical and economic performance of four small-scale

LNG processes was determined in this study. A conventional nitrogen expansion process and a proposed methane expansion process, both with and without ammonia absorption precooling cycle, were optimized. Two different optimization objectives were used, namely minimization of specific energy consumption and minimization of production cost. Lastly, an exergy analysis of the main processes was performed. From the results the following can be concluded:

- The ammonia absorption precooling cycle reduces not only the specific energy consumption by 26–35%, but also the production cost by 13–17%. This shows that adding an ammonia absorption precooling cycle is a promising improvement for small-scale expander-based process.
- The methane expansion processes have 17–28% lower specific energy consumption and 21–32% lower production cost compared to those of the nitrogen expansion processes under two optimization objectives.
- The waste heat from the gas turbine exhaust can provide all the required heat for the ammonia precooling cycle.
- Although ammonia absorption precooling cycle needs additional exergy supply, the exergy efficiency of process with precooling is still higher than the process without precooling.
- The comparison between optimization with two objective functions shows the trade-off between specific energy consumption and capital cost. Although minimization of production cost increases specific energy consumption by 0.7–3.1% compared to that of minimization of specific energy consumption, it decreases the capital cost by 3.0–5.7%. This results in 1.7–2.3% reduction in the production cost. The results indicate that the commonly used energy-related objective function may not lead to the best economic performance.

In conclusion, the methane expansion processes (SM and SMA) have promising techno-economic performance for small-scale LNG

Table 9  
Net present value for the four processes.

Process	SN		SNA		SM		SMA	
Objective function	OBJ1	OBJ2	OBJ1	OBJ2	OBJ1	OBJ2	OBJ1	OBJ2
Sales (k\$)	14921	14921	14921	14921	14921	14921	14921	14921
Capital cost (k\$)	6961	6753	6364	6128	4295	4052	4267	4118
Fuel cost (k\$)	3320	3344	2210	2288	2676	2759	1734	1782
Operation and maintenance cost (k\$)	219	213	201	193	135	128	134	130
Net present value (k\$)	4420	4611	6146	6312	7814	7982	8786	8891

plant compared to that of nitrogen expansion processes (SN and SNA). Adding an ammonia absorption precooling cycle is a promising improvement for small-scale expander-based process both for energy saving and cost saving. Minimization of production cost can lead to better techno-economic performance than that of minimization of specific energy consumption.

The Optimal operation condition can be found in Supporting Information.

### CRedit authorship contribution statement

**Jinrui Zhang:** Data curation, Formal analysis, Investigation, Methodology, Visualization, Writing - original draft. **Hans Meerman:** Supervision, Writing - review & editing. **René Benders:** Supervision, Writing - review & editing. **André Faaij:** Supervision, Writing - review & editing.

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### Appendix A. Supplementary data

Supplementary data to this article can be found online at <https://doi.org/10.1016/j.energy.2019.116592>.

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