Korean J. Chem. Eng., **31**(9), 1522-1531 (2014)
DOI: 10.1007/s11814-014-0098-z

pISSN: 0256-1115
eISSN: 1975-7220

Design and analysis of multi-stage expander processes for liquefying natural gas

Wonsub Lim, Inkyu Lee*, Kwanghee Lee*, Byeonggil Lyu*, Junghwan Kim***, and Il Moon*,†**

*Department of Chemical and Biomolecular Engineering, Yonsei University,
50, Yonsei-ro, Seodaemun-gu, Seoul 120-749, Korea
**Hyundai Heavy Industries, 17-10, Mabuk-ro, 240 Beon-gil, Ciheung-gu, Yongin-si, Gyeonggi-do 446-912, Korea
***SK Innovation, 325, Expo-ro, Yuseong-gu, Daejeon 305-712, Korea

(*Received 4 September 2013 • accepted 3 April 2014*)

Abstract−Multi-stage expander refrigeration cycles were proposed and analyzed in order to develop an efficient natural gas liquefaction process. The proposed dual and cascade expander processes have high efficiency and the potential for larger liquefaction capacity and are suitable for small-scale and offshore natural gas liquefaction systems. While refrigeration cycles of conventional expander processes use pure nitrogen or methane as a refrigerant, the proposed refrigeration cycles use one or more mixtures as refrigerants. Since mixed refrigerants are used, the efficiency of the proposed multi-stage expander processes becomes higher than that of conventional expander processes. However, the proposed liquefaction processes are different from the single mixed refrigerant (SMR) and dual mixed refrigerant (DMR) processes. The proposed processes use mixed refrigerants as a form of gas, while the SMR and DMR processes use mixed refrigerants as a form of gas, liquid- or two-phase flow. Thus, expanders can be employed instead of Joule-Thomson (J-T) valves for refrigerant expansion. Expanders generate useful work, which is supplied to the compressor, while the high-pressure refrigerant is expanded in expanders to reduce its temperature. Various expander refrigeration cycles are presented to confirm their feasibility and estimate the performance of the proposed process. The specific work, composite curves and exergy analysis data are investigated to evaluate the performance of the proposed processes. A lower specific work was achieved to 1,590 kJ/kg in the dual expander process, and 1,460 kJ/kg in the cascade expander process. In addition, the results of exergy analysis revealed that cycle compressors with associated after-coolers and companders are main contributors to total exergy losses in proposed expander processes.

Keywords: Liquefied Natural Gas, Natural Gas Liquefaction, Refrigeration Cycle, Multi-stage Expander, Exergy Analysis

INTRODUCTION

Global energy demand is projected to be about 30% higher in 2040 compared to 2010 [1]. Natural gas is favored because it is the cleanest energy source among fossil fuels. For long-distance transport, liquefied natural gas (LNG) is widely used because its volume is smaller than the volume of natural gas, approximately 1/600 [2]. In addition, the growth rate of LNG trade is projected to have a higher value than the growth rate of pipeline trade [3]. To meet the growth in LNG demand, several countries have commercialized natural gas resources and constructed LNG plants. At the same time, new gas fields have been explored, and vendors are competing to develop new high efficiency and low cost technologies and processes [4]. As a result, a variety of natural gas liquefaction processes have been developed. Natural gas liquefaction processes can be classified into three categories based on the type of refrigeration cycle: a cascade process, a mixed refrigerant process, and an expander process (Fig. 1).

Most base load LNG plants applied the cascade, single mixed refrigerant (SMR), and propane pre-cooled mixed refrigerant (C3MR) processes from the 1970s to the 1990s [4]. The APCI's C3MR process achieved relatively high thermodynamic efficiency, and thus has remained the dominant liquefaction process in the LNG plant market, being utilized in over 60% of the currently operating base load LNG plants. Recent liquefaction processes have been developed in two main directions. Many liquefaction processes have focused on increasing the capacity of a single train to strengthen price competitiveness, with the goal of using fewer trains to achieve the same capacity. At the same time, other liquefaction processes have been developed with the aim of increasing the profitability of stranded gas fields. Specifically, liquefaction processes for offshore LNG plants have received great interest.

Many mixed refrigerant processes were developed to increase single train capacity. The AP-X process was developed from the C3MR process by APCI. A characteristic of the process is that the sub-cooling of LNG is performed by a nitrogen expander cycle [5]. The dual mixed refrigerant (DMR) process was developed by Shell [6]. The process uses a mixed refrigerant instead of pure propane in the pre-cooling cycle. Shell also developed the parallel mixed refrigerant (PMR) process, which consists of pre-cooling and liquefaction cycles [7]. A unique feature of the process is that two mixed refrigerant cycles are configured in parallel. Axens-IFP developed a liquefaction process with two mixed refrigerant cycles, pre-cooling and liquefaction [8]. Both cycles are designed to use the same amount of power so that the same set of drivers can be used for compressors across different cycles. The mixed fluid cascade (MFC) process was developed by Statoil-Linde. The process is similar to the cascade process, but uses mixed refrigerants. The cascade process

†To whom correspondence should be addressed.
E-mail: ilmoon@yonsei.ac.kr

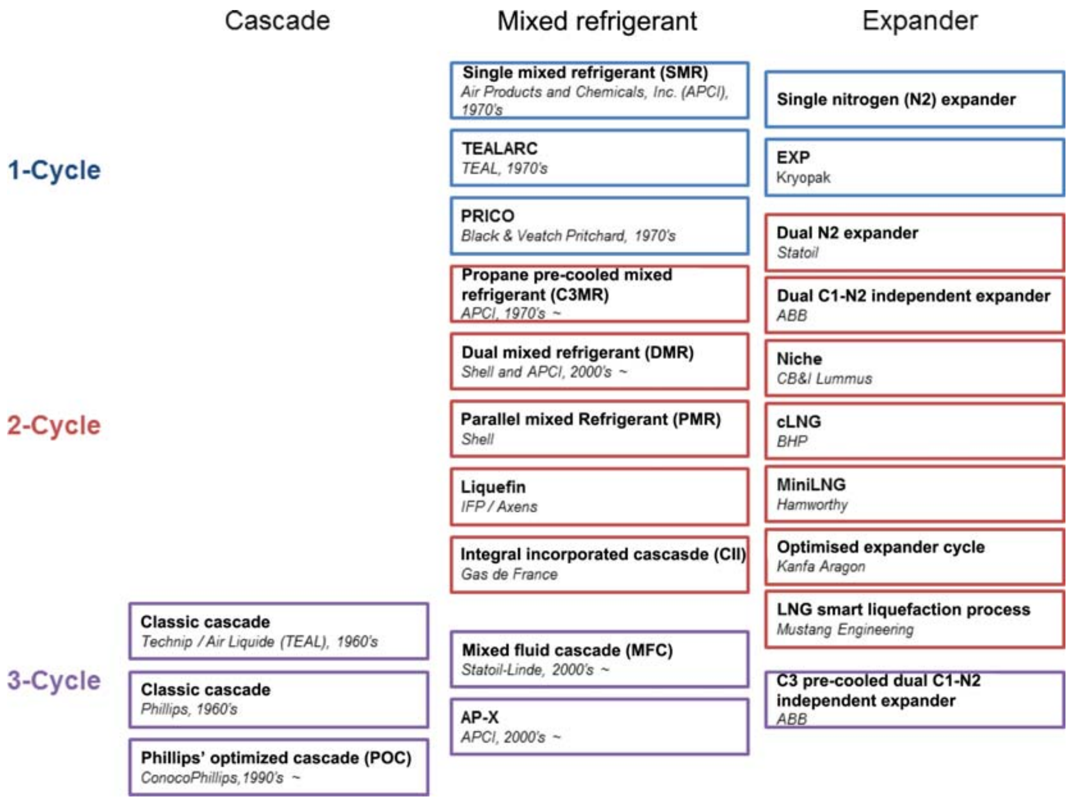

Fig. 1. Natural gas liquefaction processes.

was also innovated to increase the single train capacity. ConocoPhillips developed a new version of the cascade process, Phillips' optimized cascade (POC) process [9]. The process successfully achieved a reduction in power consumption.

Many expander processes have been developed for small-scale and offshore plants. Nitrogen expander processes, which are based on reverse-Brayton and Claude cycles, have been widely used for cryogenic liquefaction such as boil-off gas (BOG) re-liquefaction and LNG peak-shaving plants [10]. The single nitrogen expander process has a simple configuration, but its efficiency is relatively low because a single pure gas refrigerant is used over a wide temperature range [11]. The process uses expanders instead of Joule-Thomson valves for refrigerant expansion. Expanders generate useful work, which is usually supplied to compressors. The dual nitrogen expander process was developed by modifying the single nitrogen expander process to reduce energy consumption [12]. This process consists of warm and cold expander cycles. A compact LNG (cLNG) process was developed by BHP. This process configuration is similar to SMR process, except pure nitrogen refrigerant is adopted. To enhance energy efficiency, the cLNG process is operated under two pressure levels of nitrogen expansion. These expander processes can employ the propane pre-cooling cycle to improve their performance. In case of using propane pre-cooling cycle, power consumption of compressors is reduced by approximately 20%. Furthermore, many other expander processes have also been developed by various vendors, e.g., dual C1-N2 independent process by ABB, Niche by CB&I Lummus, EXP by Kryopak, miniLNG by Hamworthy, Optimized expander cycle by Kanfa Aragon, and LNG smart liquefaction processes by Mustang Engineering [13-20]. Most of these expander processes consist of one or two cycles and use pure refrigerants as a form of gas.

We focused on design and analysis of multi-stage expander liquefaction processes and used pure and mixed refrigerants as a form of gas. The proposed liquefaction processes are different from the SMR and DMR processes. The proposed process uses mixed refrigerants as a form of gas because liquid-phase or two-phase mixed refrigerants require large liquid storage space, which may induce the safety-related problem. As the proposed multi-stage expander liquefaction processes use mixed refrigerants as a form of gas, expanders can be used instead of Joule-Thomson (J-T) valves for refrigerant expansion. Various sets of refrigerants were proposed and investigated in dual and cascade expander processes. The single N2 expander, dual N2 expander, and SMR processes were also investigated under given conditions for comparison with the proposed processes.

PROPOSED LIQUEFACTION PROCESSES

1. Process Design Basis

The efficiency of liquefaction processes strongly depends on feed gas conditions and LNG product specifications. Proposed and existing liquefaction processes were designed under given conditions. The pressure of the natural gas feed was 5.0 MPa and the temperature was 37 °C. The mole fraction of feed gas was defined as nitrogen (N2) 0.2%, methane (C1) 91.3%, ethane (C2) 5.4%, propane (C3) 2.1%, i-butane (i-C4) 0.50%, and n-butane 0.50%. The pressure of the LNG was 150 kPa at the end of the liquefaction process. The vapor fraction of LNG was fixed at 0 so that the natural gas feed was totally changed to LNG. All compressors used were assumed

Table 1. Assumptions for natural gas liquefaction process design

Ambient temperature	20 °C
Centrifugal compressor adiabatic efficiency	0.82
Expander adiabatic efficiency	0.85
Pinch temperature	3 °C
Refrigerant temperature at after-coolers	37 °C
LNG expansion valve exit pressure	150 kPa
Pressure drop of feed gas in each part of LNG heat exchanger	200 kPa
Pressure drop of high-temperature refrigerant in each part of LNG heat exchanger	150 kPa
Pressure drop of low-temperature refrigerant in each part of LNG heat exchanger	20 kPa
Pressure drop in other heat exchangers	20 kPa
Maximum compression ratio of each compressor	3.0

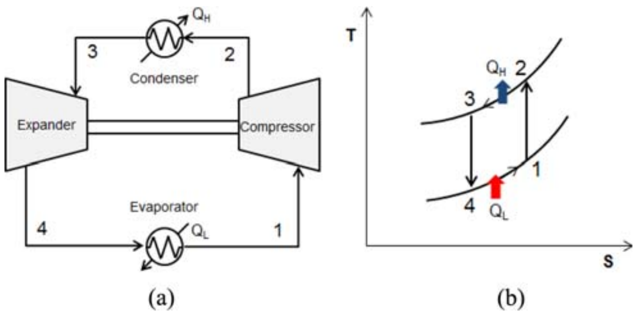

Fig. 2. Reverse-Brayton cycle. (a) Basic schematic, (b) T-S diagram.

to be centrifugal compressors. All after-coolers and condensers were assumed to be cooled by sea water. The LNG expansion was carried out using a Joule-Thomson (J-T) valve. Detailed design specifications are listed in Table 1.

2. Basic Cycle and Refrigerant Selection

We focused on the expander process based on the reverse-Brayton cycle, as shown in Fig. 2. Fig. 3 represents the single nitrogen expander process, which is the simplest expander process. Since the process uses an expander, it allows the use of pure or mixed refrigerants as a form of gas. Fig. 4 illustrates the dual nitrogen expander process, which was comprised of warm and cold expander cycles to reduce energy consumption compared with single N2 expander process. Conventional expander processes use pure nitrogen or methane as a form of gas refrigerant, but their efficiency is relatively low because the pure gas refrigerant is used over a wide temperature range. To improve the thermodynamic efficiency, we considered a variety of combinations of pure and mixed gas refrigerants. The proposed cycles included nitrogen (N2), methane (C1), ethane (C2), or carbon dioxide (CO2) as pure or mixed refrigerants.

The majority of expander processes consist of one or two refrigeration cycles. A few studies discussed the potential for the expander process to comprise three refrigeration cycles. These studies simply investigated the potential use of a propane pre-cooling cycle in single and dual expander processes. The power consumption of the conventional expander process was found to be reduced by approximately 20% through the application of a propane pre-cooling cycle. However, the propane pre-cooling cycle increases design complexity and requires a larger refrigerant flow rate. To overcome these disadvantages, this study proposes other expander processes employing three-stage refrigeration cycles. The proposed cascade expander process is structurally similar to the conventional cascade process.

3. Proposed Dual Expander Liquefaction Process (DEP)

Fig. 5 is a basic schematic of the proposed dual expander liquefaction process and Fig. 6 is a T-S diagram. The proposed process consists of two stages of refrigeration cycles, in which two cycles are operated separately at multiple pressure levels. Each refrigeration cycle is designed to have the same equipment configuration. However, the two cycles use different refrigerants and require a different amount of energy. The first cycle cools the natural gas and

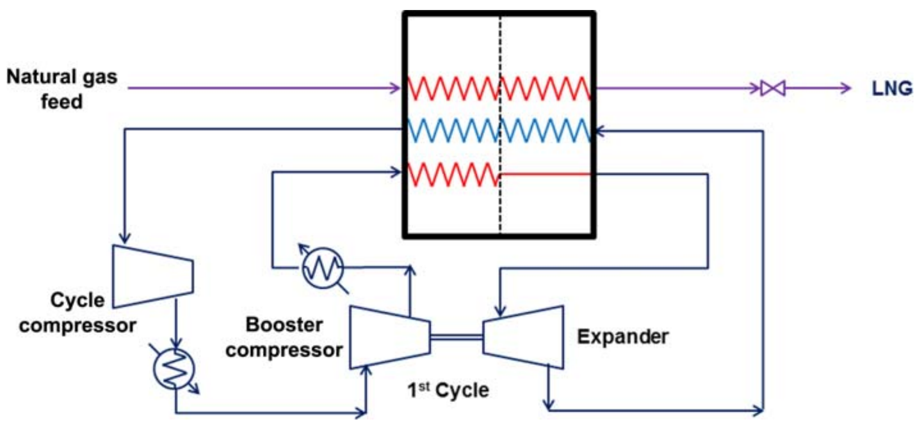

Fig. 3. Single nitrogen (N2) expander process.

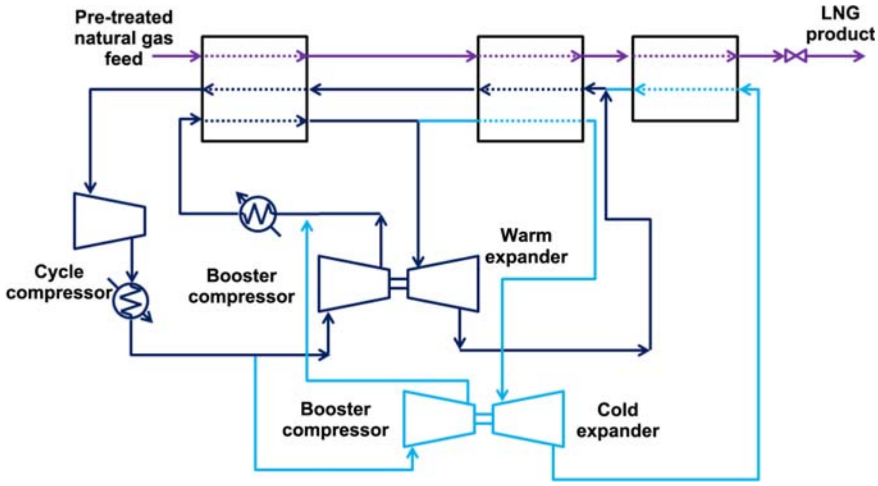

Fig. 4. Dual nitrogen (N2) expander process.

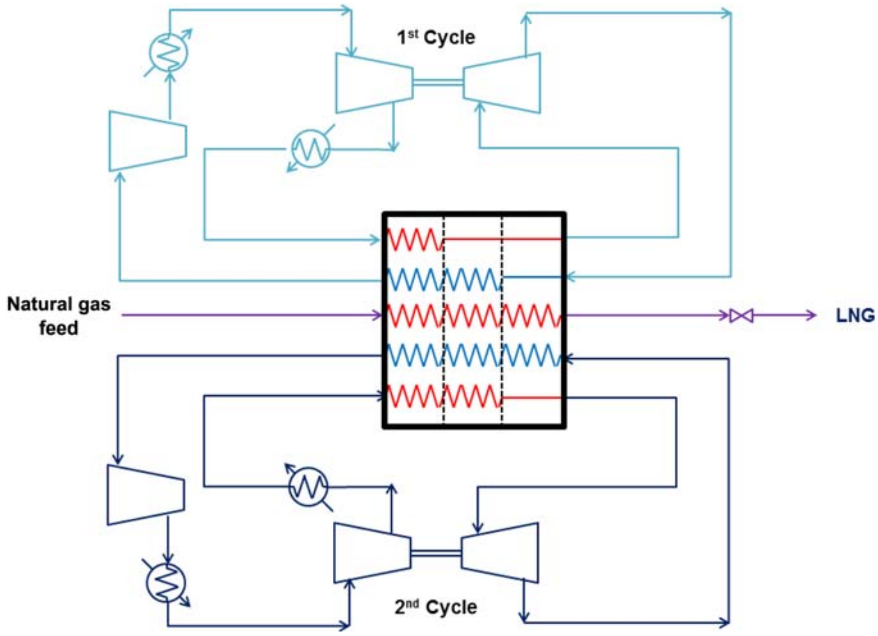

Fig. 5. Proposed dual expander process.

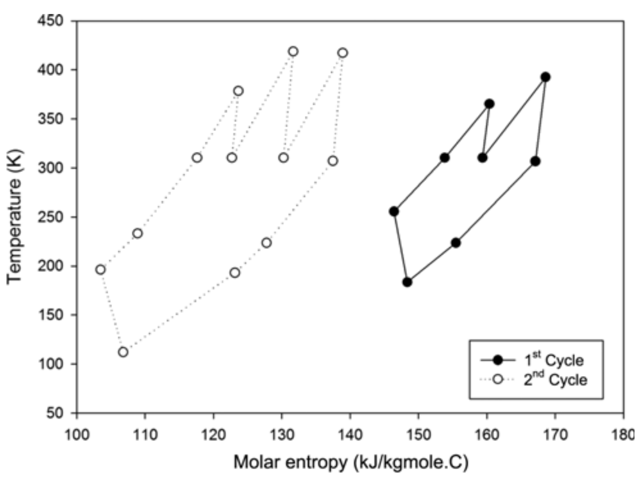

Fig. 6. T-S diagram of proposed dual expander process.

the refrigerant for the second cycle to around −80 °C. Operating conditions of DEP-7 is a little different because the DEP-7 is used carbon dioxide as a mixed refrigerant. The first cycle of DEP-7 cools the natural gas and the refrigerant for the second cycle to around −72 °C in order to avoid freezing of CO2. The freezing temperature of CO2 is −78 °C, and thus the minimum temperature of MR-4 which consists of CO2 and N2 is cooled down to approximately −75 °C. The freezing of CO2 is checked by "CO2 freeze out" utility provided by aspen HYSYS.

The refrigerant for the first cycle is compressed in a cycle compressor and then cooled by sea water in an after-cooler. A compander, combined expanders with compressors in a single machine, compresses the refrigerant to a higher pressure. The higher pressure refrigerant is then cooled by passing through the after-cooler and an LNG heat exchanger. Finally, the refrigerant is expanded to a lower temperature in an expander in the compander to cool the

Table 2. Refrigerants used in the proposed dual expander process and specific works

Process	Refrigerant	1st Expander cycle (mole fractions)	2nd Expander cycle (mole fractions)	Specific work (kJ/kg LNG)
DEP-1	C1-N2	Pure C1	Pure N2	1645.74
DEP-2	MR1-N2	C1 : C2 (0.9 : 0.1)	Pure N2	1620.95
DEP-3	MR2-N2	N2 : C1 (0.5 : 0.95)	Pure N2	1643.93
DEP-4	MR3-N2	N2 : C1 : C2 (0.5 : 0.85 : 0.10)	Pure N2	1623.73
DEP-5	C1-MR2	Pure C1	N2 : C1 (0.8 : 0.2)	1619.13
DEP-6	MR1-MR2	C1 : C2 (0.9 : 0.1)	N2 : C1 (0.8 : 0.2)	1596.53
DEP-7	MR4-N2	N2 : CO2 (0.85 : 0.15)	Pure N2	1818.93

natural gas and the second-cycle refrigerant. The second cycle is operated in the same way.

To generate useful work to drive the refrigerant compressor in the expander, the refrigerant is used as a form of gas. Nitrogen and hydrocarbons were used as the pure refrigerant or mixed refrigerant in this study because these refrigerants are widely used in LNG plants. In addition, carbon dioxide was investigated as a refrigerant due to its non-flammability. To improve energy efficiency, seven combinations of available refrigerants were simulated and analyzed (see Table 2). Optimization of the mixed refrigerant composition is not included in this work. Nevertheless, the simulation results indicated sufficient improvements. The proposed process requires a complicated heat exchanger, but it requires a lower specific work than the single nitrogen expander process.

4. Proposed Cascade Expander Liquefaction Process (CEP)

Fig. 7 is a basic schematic of the proposed cascade expander liquefaction process and Fig. 8 is a T-S diagram. The proposed cascade expander process consists of a three-stage refrigeration cycle. These three cycles are operated separately at multiple pressure levels. The proposed process employs a pre-cooling cycle to achieve a higher efficiency compared to the proposed dual expander process. The proposed pre-cooling cycle uses an expander, while the C3 pre-cooling cycle uses J-T valves. The proposed pre-cooling cycle cools the natural gas and the refrigerants for the other cycles to around –30 °C.

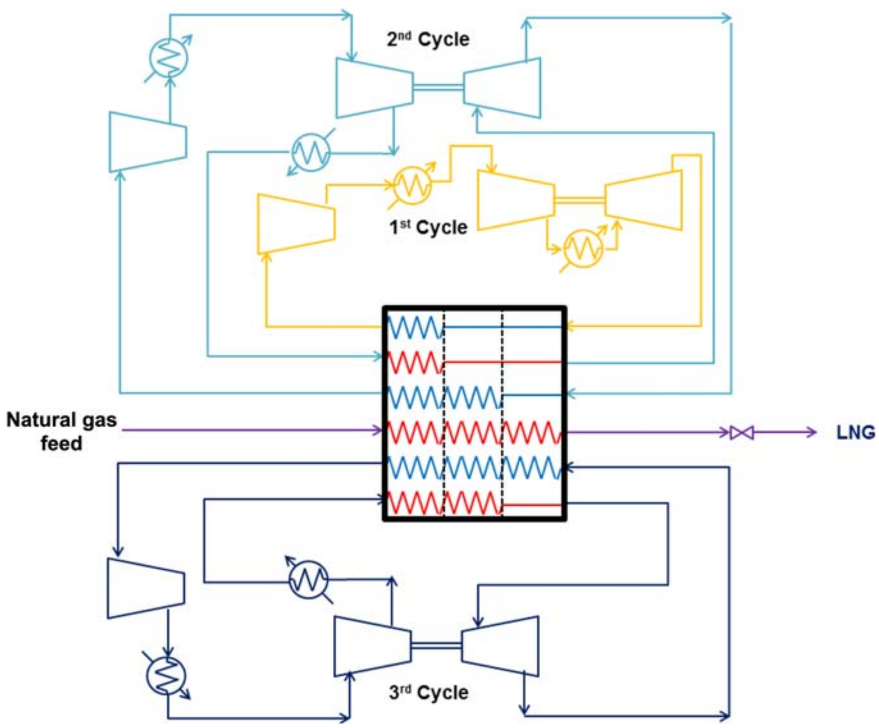

Fig. 7. Proposed cascade expander process.

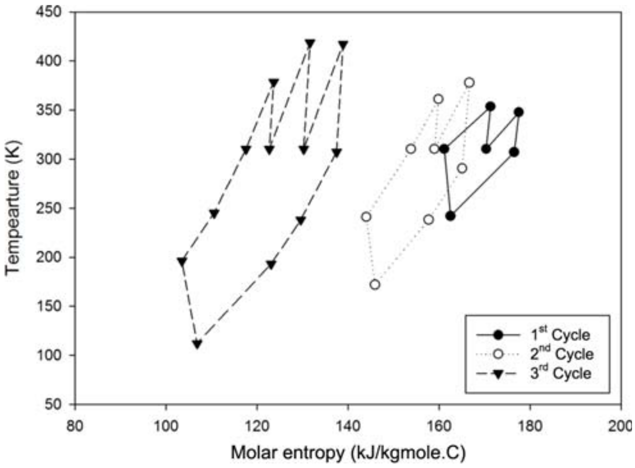

Fig. 8. T-S diagram of proposed cascade expander process.

The proposed cascade refrigeration cycle is operated in the same way as the dual expander process, except for the absence of self-cooling in the pre-cooling stage. Five combinations of available refrigerants were proposed and evaluated. These five combinations are summarized in Table 3. The proposed cascade expander process requires a lower specific work than the proposed dual expander process, but its design is more complicated.

EXERGY ANALYSIS

Exergy analysis is a useful method in studies on the effectiveness of an energy system, and its fundamental principles and methodology can be found in the literature [21,22]. A variety of studies used exergy analysis to evaluate thermodynamic efficiency in the liquefaction process [15,23-25]. We also used exergy analysis to evaluate the thermodynamic efficiency of each proposed process.

Exergy is defined as the useful available work in a gas, fluid, or mass as a result of its non-equilibrium condition relative to some reference condition. According to the second law of thermodynamics, the entropy of an isolated system which is not in equilibrium will continue to increase, and work must be done to reduce it. Exergy analysis is a technique that measures the loss of exergy or the work required to restore the system to the equilibrium state. Exergy of a stream under specified conditions can be measured at various points according to the following equation:

Table 4. Expressions for exergy loss in equipment

Equipment	Expression for the exergy loss
Heat exchanger	$Ex_{loss}=\Sigma Ex_{in}-\Sigma Ex_{out}$
Compressor	$Ex_{loss}=Ex_{in}-Ex_{out}+W_{input}$
Expander	$Ex_{loss}=Ex_{in}-Ex_{out}-W_{output}$
Cooler	$Ex_{loss}=(Ex_{R,in}+Ex_{W,in})-(Ex_{R,out}+Ex_{W,out})$
Valve	$Ex_{loss}=T_0(S_{out}+S_{in})$
Mixer	$Ex_{loss}=\Sigma Ex_{in}-\Sigma Ex_{out}$

- Ex_{loss} : exergy loss of equipment
- Ex_{in} : exergy of input stream
- Ex_{out} : exergy of output stream
- W_{input} : input power for compressor
- W_{output} : output power of expander
- $Ex_{R,in}$: exergy of refrigerant input stream
- $Ex_{R,out}$: exergy of refrigerant output stream
- $Ex_{W,in}$: exergy of water input stream
- $Ex_{W,out}$: exergy of water output stream
- T_0 : ambient temperature
- S_{out} : entropy of output stream
- S_{in} : entropy of input stream

$$Ex=(H-H_o)-T_0(S-S_0) \tag{1}$$

where H is enthalpy, T is temperature, S is entropy, and subscript 0 indicates the equilibrium condition, normally considered the ambient condition.

When matter is taken from its initial state to its final state in a hypothetical reversible process, the reference terms cancel out. Therefore, the change in exergy from the initial state to the final state can be expressed as follows:

$$\Delta Ex=(H_f-H_i)-T_0(S_f-S_i) \tag{2}$$

where H_f and S_f represent the specific enthalpy and entropy of the final state, respectively, and H_i and S_i represent the specific enthalpy and entropy of the initial state, respectively.

The exergy loss can be calculated based on the exergy balance from inlet to outlet streams. For example, using Eq. (2), the change in exergy of a stream due to heat exchange is calculated for all streams passing the heat exchangers. Finally, the exergy loss (Ex_{loss}) is calculated by subtracting the sum of exergy for output stream (ΣEx_{out}) from the sum of exergy for input stream (ΣEx_{in}) according to the

Table 3. Refrigerants used in the proposed cascade expander process and specific works

Process	Refrigerant	1st Expander cycle	2nd Expander cycle (mole fractions)	3rd Expander cycle (mole fractions)	Specific work (kJ/kg LNG)
CEP-1	C2-C1-N2	Pure C2	C1	N2	1538.42
CEP-2	C2-MR1-N2	Pure C2	C1 : C2 (0.9 : 0.1)	N2	1490.49
CEP-3	C2-C1-MR2	Pure C2	C1	N2 : C1 (0.8 : 0.2)	1490.98
CEP-4	C2-MR1-MR2	Pure C2	C1 : C2 (0.9 : 0.1)	N2 : C1 (0.8 : 0.2)	1466.05
CEP-5	CO2-C1-N2	Pure CO2	C1	N2	1569.41

following equation:

$$Ex_{loss}=\Sigma Ex_{in}-\Sigma Ex_{out} \tag{3}$$

Expressions for exergy loss in equipment are summarized in Table 4.

For a given natural gas feed condition and LNG specification, the minimum work required to produce the LNG is determined by the difference in the exergy of the LNG and the feed. This can be expressed as

$$W_{rev}=\Sigma(H-T_0S)_{LNG}-\Sigma(H-T_0S)_{feed} \tag{4}$$

For a natural gas feed pressure of 5.0 MPa absolute, a natural feed temperature of 37 °C, a LNG pressure of 150 kPa, a LNG temperature of −158.1 °C and an ambient temperature of 20 °C, the minimum reversible specific work requirement is 397.9 kJ/kg LNG. The actual required specific work is greater than the minimum reversible specific work because all of the main equipment in the natural gas liquefaction process involves irreversible operations.

RESULTS AND DISCUSSION

1. Specific Work

All proposed processes were simulated using commercial software, aspen HYSYS, with the Peng-Robinson equation of state. To compare process efficiency, three conventional processes were also simulated.

The specific work of each process was calculated under the same given conditions. Specific work requirements for all processes are summarized in Fig. 9. In the proposed liquefaction processes, the net liquefaction work per unit mass was estimated at 1,590-1,819 kJ/kg with dual expander processes, and 1,460-1,570 kJ/kg with cascade expander processes. In conventional processes, the liquefaction performance was calculated at 1,249 kJ/kg for the SMR process, 1,872 kJ/kg for the dual nitrogen expander process, and 2,492 kJ/kg for the single nitrogen expander process. The proposed expander processes do not seem to reach the level of performance of the mixed refrigerant processes. However, the proposed processes have potential for performance improvements because they are early versions and have not undergone optimization for specified site conditions.

Among the dual expander refrigeration cycles, the MR1(C1+C2)-MR2(N2+C1) cycle had the highest performance, but it can be dangerous due to the use of mixed refrigerants containing flammable hydrocarbons. The C3-N2 cycle employs two pure refrigerants, so

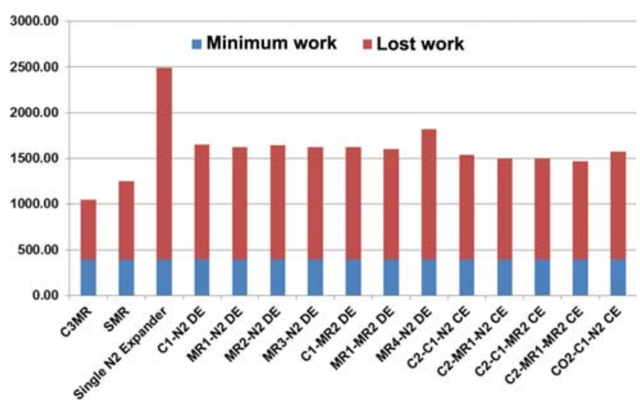

Fig. 9. Specific work requirements.

it is safer than the MR1-MR2 cycle. Nevertheless, this cycle also uses a hydrocarbon refrigerant; thus it is still more dangerous than the N2 expander processes. To improve the process safety, the MR4 (N2+CO2)-N2 cycle using non-flammable refrigerants was proposed and investigated. The MR4-N2 dual expander process seems to reach the safety level of the nitrogen expander processes, but its performance was the lowest level among proposed dual expander processes.

The proposed cascade expander process has a higher performance than the proposed dual expander process, likely due to the use of a

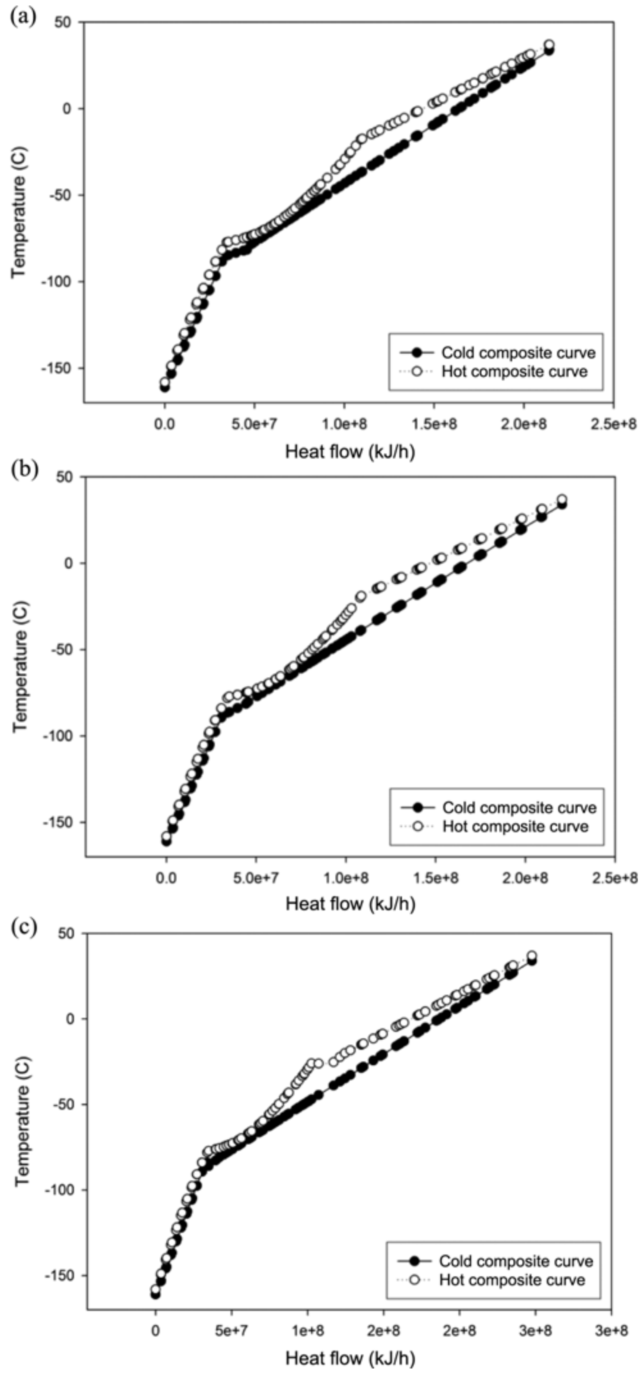

Fig. 10. Composite curves for the proposed dual expander processes. (a) DEP-1, (b) DEP-6, (c) DEP-7.

pre-cooling cycle. According to the simulation results, the C2-MR1 (C1+C2)-MR2(N2+C1) cycle had the highest efficiency, and the CO2-C1-N2 cycle had the lowest efficiency. The CO2-C1-N2 cycle adopted a CO2 pre-cooling cycle instead of a C2 pre-cooling cycle to increase the process safety, but its efficiency decreased. These results show a trade-off relationship between safety and performance. Therefore, the appropriate refrigerant cycle should be selected based on site conditions and process restrictions.

2. Composite Curves

The natural gas liquefaction process commonly involves the use of multi-stream heat exchanger (MSHE) as an LNG heat exchanger. Composite curves are employed to analyze the MSHE. All of the heat and cold flows in the MSHE are expressed by hot and cold composite curves, respectively. The exergy loss of MSHE is reflected by the area between the hot and cold composite curves. Large temperatures differences and heat exchanger load are primary reasons for exergy loss in LNG heat exchangers. A small area between curves would indicate an effective reduction in the specific work. As the temperature difference between hot and cold composite curves decreases, heat transfer area increases.

Representative processes were selected from among those proposed based on specific work, and the hot and cold composite curves for these processes are shown in Figs. 10 and 11. Figs. 10(a)-(c) show the composite curves for LNG heat exchangers in the DEP-1, DEP-6, and DEP-7, respectively. The two sections of the cold curve are linear and match the hot composite curve because the refrigerants do not change phase. Fig. 10 shows that the area between the hot and cold composite curves in the temperature range below −70 °C is very small. However, the area for temperatures above −70 °C is much larger than that for temperatures below −70 °C. Therefore, a higher degree of optimization may be possible at temperatures above −70 °C. Figs. 11(a)-(c) show the composite curves for the LNG heat

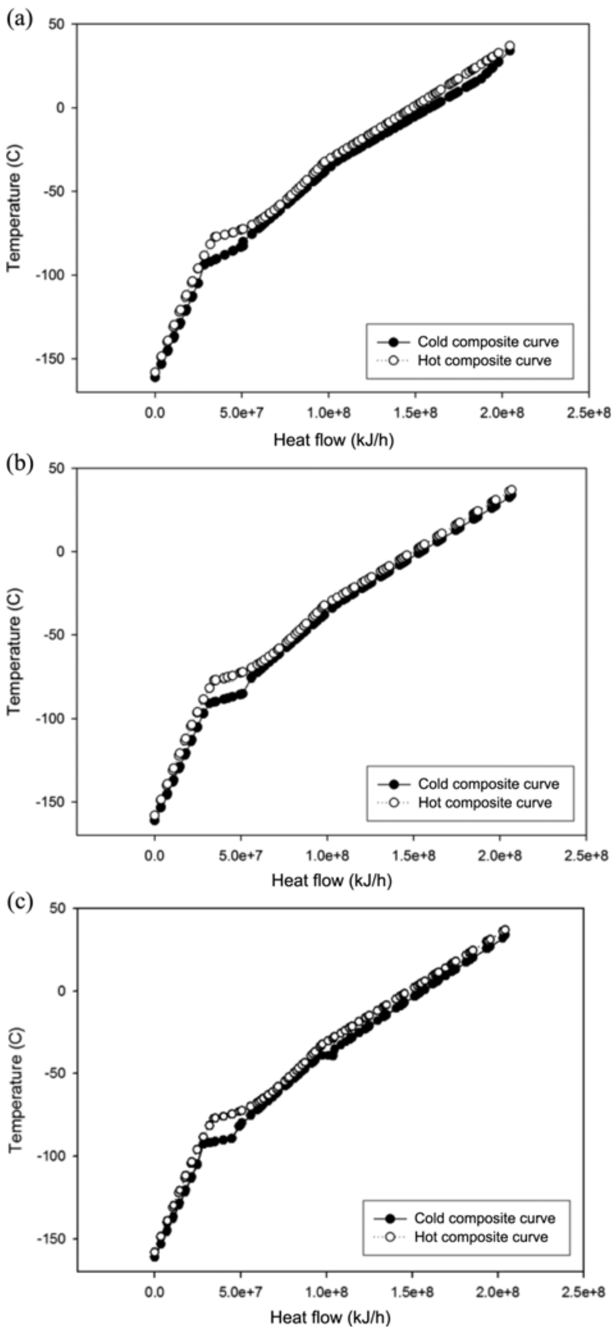

Fig. 11. Composite curves for the proposed cascade expander processes. (a) CEP-1, (b) CEP-4, (c) CEP5.

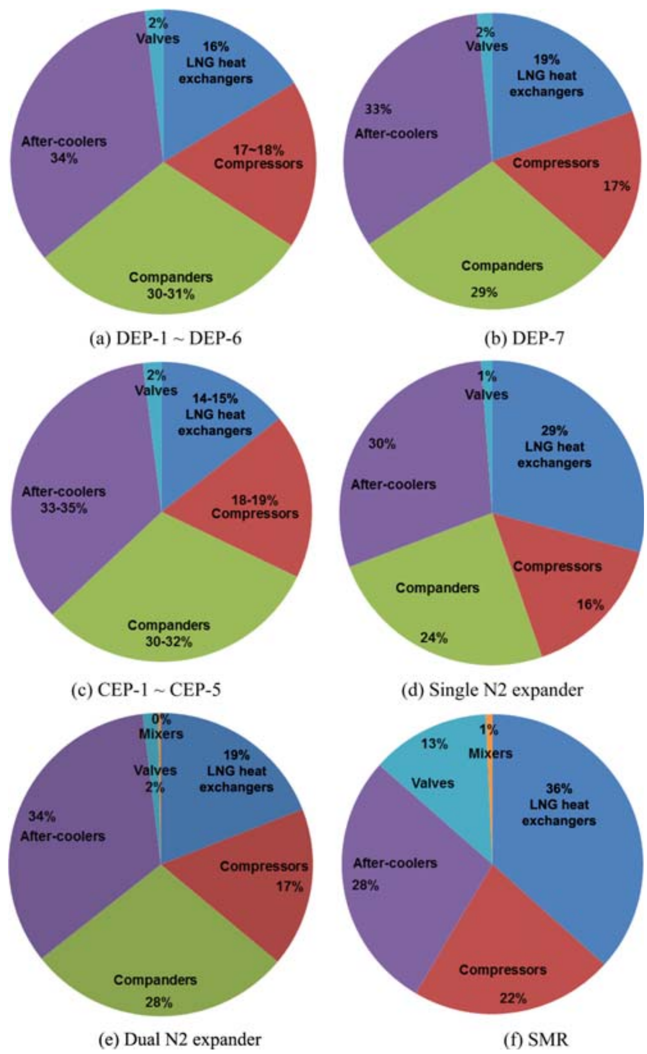

Fig. 12. Patterns of lost work.

Table 5. Number of equipment in each process

Equipment	Number of equipment				
	SMR	Single N2 expander	Dual N2 expander	Proposed DEPs	Proposed CEPs
LNG heat exchanger	1	1	1	1	1
Heat exchanger	0	0	0	0	0
Compressor	3	3	4	5	7
Pump	1	0	0	0	0
Expander	0	1	2	2	3
After-cooler	3	3	3	5	7
Valve	2	1	1	1	1
Mixer	1	0	2	0	0
Splitter	0	0	2	0	0
Separator	1	0	0	0	0
Total count	12	9	15	14	19

exchanger in CEP-1, CEP-4, and CEP-5, respectively. Employing a pre-cooling cycle in the proposed cascade expander process decreased the area at temperatures above −70 °C compared to that for the proposed dual expander process. As the area between composite curves was decreased, the refrigerant flow rate was also reduced. When the refrigerant flow rate was reduced, the power requirement for refrigerant compression was also reduced.

3. Exergy Analysis

The exergy losses for all equipment were calculated based on the exergy equilibrium equations (see Fig. 12). In the figure, LNG heat exchangers represent multi-stream heat exchangers, and heat exchangers represent all others. The compressors represent compression equipment except compressors equipped in companders, and companders represent compressors and the associated expanders. The after-coolers are associated with all compression equipment, including compressors or companders. As shown in Fig. 12, the order of contribution to exergy losses changed according to the liquefaction process because each process uses different equipment and configurations. The combined losses for the compression system, including the compressors and associated after-coolers, were the most significant contributor to total exergy loss. This finding indicates that a reduction in total exergy losses can be effectively achieved by decreasing the losses in the compression system. A decrease in the exergy losses in the compression system can be achieved by adopting multi-stage compression and optimizing the compression ratio and operating pressure range.

With regard to the proposed expander processes, the two other important pieces of equipment with irreversible operations are the companders and the LNG heat exchangers. The exergy loss of the J-T valves is lowest. The compander consists of a compressor and an expander on the same shaft without a driver. The expander generates power, and this power is recovered and employed to drive the compressor. Decreasing the exergy losses of companders is important in the proposed expander processes because they are the second largest contributor to these losses. Effective ways to reduce the losses of companders are improving their efficiencies and optimizing parameters related to expansion and compression devices.

Fig. 12 shows that the LNG heat exchangers contributed 29% of the exergy loss in the single N2 expander process, 19% for the dual N2 expander process, 16% for the dual expander processes except DEP-7, and 14-15% in the cascade expander processes. As mentioned in the previous section, the exergy loss in LNG exchangers is reflected in the hot and cold composite curves. The proposed expander processes effectively reduced the exergy loss in the LNG exchangers compared to the single and dual N2 expander processes as a result of the pre-cooling and cascade process concepts. The DEP-7 had a higher exergy loss in the compressors compared to the other proposed dual expander processes. This is because the refrigerant flow rate in the first cycle was increased due to the use of a mixed refrigerant comprised of N2 and CO2.

4. Other Considerations

Natural gas liquefaction processes were evaluated based on thermodynamic efficiency in previous sections. However, to achieve a successful LNG project, the liquefaction processes should be compared by taking technical and economic parameters into consideration. Table 5 summarizes the amount of equipment for the basic configuration used for the liquefaction processes in this study. The simplicity of both the single N2 expander and proposed expander processes is demonstrated by their low equipment counts. Compared to the single N2 expander process, both the capital cost and maintenance cost of the proposed expander processes would increase as equipment count increases. The relative equipment size is expected to decrease as specific work decreases because of lower refrigerant flow rates.

The safety of the liquefaction process is related to the kind of refrigerant used. Nitrogen and carbon dioxide refrigerants are safer than hydrocarbon refrigerants. However, the proposed processes that used N2 and CO2 as refrigerants required a higher energy. The cascade expander process is more suitable for achieving a larger capacity in LNG plants. When comparing LNG plants of the same capacity, the dual expander process requires a smaller plot area. These results indicate that each process has its own advantages and disadvantages.

The proposed processes have the potential to achieve a larger liquefaction capacity because these require lower refrigerant flow rate than the conventional expander processes. In general, there is limitation of equipment used in LNG systems. Because the refrigerant flow rate in each refrigeration cycle was reduced, the proposed processes would be reaching a higher capacity under the same limitation of equipment compared with conventional processes.

CONCLUSION

Dual and cascade expander processes were proposed and evaluated. Unlike conventional expander processes, the proposed processes employ pure and mixed refrigerant as a form of gas. In addition, use of carbon dioxide as a refrigerant was analyzed. The specific work values, composite curves, and exergy analyses were carried out for the proposed and conventional liquefaction processes. All of the proposed process showed enhanced performance compared with the conventional single and dual N2 expander processes in terms of thermodynamic efficiency. In particular, the MR1-MR2 dual expander and C2-MR1-MR2 cascade expander processes were most suitable for LNG plants due to their higher thermodynamic efficiency. However, when taking process safety into consideration, the MR4-N2 dual expander would be preferred because it employs non-flammable refrigerants. Each process has its own advantages and disadvantages; thus the appropriate liquefaction process can be altered based on primary considerations specific to the plant site.

The proposed processes are expected to be appropriate for small-scale and offshore LNG systems due to their simplified design. These processes generally use mixed refrigerant which consists of two components and is relatively simple compared to conventional mixed refrigerant processes such as SMR, C3MR. Therefore, the control of the make-up system for refrigerant in the proposed processes is simpler than of conventional MR processes. From the perspective of design simplicity and ease of operation, the proposed processes are suitable for LNG peak-shaving and BOG reliquefaction systems because the load of these systems is frequently changed. Furthermore, the results indicated that the proposed expander processes have a reasonably high efficiency and the potential to achieve a larger liquefaction capacity.

ACKNOWLEDGEMENTS

This research was supported by a grant from the GAS Plant R&D Center funded by the Ministry of Land, Transportation and Maritime Affairs (MLTM) of the Korean government and also respectfully supported by BK 21 Program funded by the Ministry of Education (MOE) of Korea.

REFERENCES

1. The outlook for energy: A view to 2040, Exxon Mobil (2012).
2. S. Kumar, H. T. Kwon, K. H. Choi, J. H. Cho, W. Lim and I. Moon, *Energy Policy*, **39**, 4097 (2011).
3. S. Kumar, H. T. Kwon, K. H. Choi, W. Lim, J. H. Cho, K. Tak and I. Moon, *Appl. Energy*, **88**, 4264 (2011).
4. W. Lim, K. Choi and I. Moon, *Ind. Eng. Chem. Res.*, **52**, 3065 (2013).
5. M. J. Roberts and R. Agrawal, US Patent, 6,308,531 (2001).
6. P. Boutelant, OAPEC-IFP Joint Seminar, Rueil-Malmaison, France (2008).
7. J. M. v. d. Graaf and B. Pek, Business Briefing: LNG Review (2005).
8. P.-Y. Martin, J. Pigourier and B. Fischer, Hydrocarb. Engineering (2004).
9. D. L. Andress, Phillips Petroleum Company (1996).
10. K. J. Vink and R. K. Nagelvoort, *12th International Conference and Exhibition on Liquefied Natural Gas (LNG-12)*, Perth, Australia (1998).
11. A. J. Finn, G L. Johnson and T. R. Tomlinson, *79th Annual GPA Convention*, GPA, Atlanta, USA (2000).
12. A. J. Finn, *81st Annual GPA Convention*, Dallas, USA (2002).
13. J. H. Foglietta, *AIChE Spring National Meeting*, New Orleans, USA (2002).
14. J. H. Foglietta, Hydrocarb. Process (2004).
15. C. W. Remeljej and A. F. A. Hoadley, *Energy*, **31**, 2005 (2006).
16. M. Barclay and N. Denton, *LNG Journal* (2005).
17. *The Kryopak EXP LNG process*, Kryopak (2012).
18. *Small scale and MiniLNG systems for LNG production and emission recovery*, Hamworthy (2012).
19. I. B. Waldmann, *Tekna Conference* (2008).
20. P. E. Susan Walther, *International Gas Union Research Conference*, Paris, France (2008).
21. R. A. Gaggioli, *Int. J. Appl. Thermodynam.*, **1**, 1 (1998).
22. I. b. Dincer and M. Rosen, Elsevier, Boston (2007).
23. M. Kanoglu, I. Dincer and M. A. Rosen, *Int. J. Energy Res.*, **32**, 35 (2008).
24. M. Kanoglu, *Int. J. Energy Res.*, **26**, 763 (2002).
25. G. Venkatarathnam and K. D. Timmerhaus, Springer, New York (2008).