Techno-economic Analysis of Acid Gas Removal and Liquefaction for Pressurized LNG

S H Lee¹, Y K Seo² and D J Chang²
¹ Graduate School of Ocean Systems Engineering, Department of Mechanical Engineering, Korea Advanced Institute of Science and Technology, Republic of Korea
² Department of Mechanical Engineering, Korea Advanced Institute of Science and Technology, Republic of Korea

computer10@kaist.ac.kr, djchang@kaist.ac.kr

Abstract. This study estimated the life cycle cost (LCC) of an acid gas removal and a liquefaction processes for Pressurized LNG (PLNG) production and compared the results with the cost of normal LNG production. PLNG is pressurized LNG that is liquefied at a higher pressure and temperature than normal LNG. Due to the high temperature, the energy for liquefaction is reduced. The allowable CO₂ concentration in PLNG is increased up to 3 mol% when the product pressure 25 bar. An amine process with 35 wt% of diethanolamine (DEA) aqueous solution and a nitrogen expansion cycle were selected for the acid gas removal and the liquefaction processes, respectively. Two types of CO₂ concentration in the feed gas were investigated to analyze their impacts on the acid gas removal unit. When the CO₂ concentration was 5 mol%, the acid gas removal unit was required for both LNG and PLNG production. However, the acid gas removal unit was not necessary in PLNG when the concentration was 0.5 mol% and the pressure was higher than 15 bar. The results showed that the LCC of PLNG was reduced by almost 35% relative to that of LNG when the PLNG pressure was higher than 15 bar.

1. Introduction
Small floating liquefied natural gas (FLNG) projects are under consideration to monetize small-scale or stranded gas fields located in near-shore areas [1]. If FLNG plants are installed at near-shore sites, the emission of greenhouse gases such as CO₂ from these plants will become a new environmental issue. Generally, a large amount of CO₂ is captured during the acid gas removal process at FLNG plants and vented to the atmosphere, as the CO₂ concentration in LNG should be maintained below 50 ppm to avoid the formation of solids. Pressurized Liquefied Natural Gas (PLNG) can provide a new economical solution for reducing the emissions of greenhouse gases in FLNG. PLNG is produced at a higher pressure and temperature than normal LNG. Due to its high temperature condition, CO₂ can be present in PLNG at rates as high as 1 mol% without the formation of solids when the pressure of PLNG is 15 bar; moreover, the level can reach 3 mol% when the pressure is 25 bar [2]. It is also possible to reduce the energy required and cost of liquefaction. The previous study have shown that when the product pressure is 10 bar compared to 1 bar, almost 35% of the cost for the liquefaction can be saved [3]. Several studies have suggested novel liquefaction processes for PLNG with a cryogenic CO₂ removal unit in which the CO₂ is removed by crystallization during the liquefaction process [2, 4]. However, a strategy for handling solid particles after the cryogenic removal unit in FLNG is still required. The purpose of this study is to estimate the lifecycle cost (LCC) of these separate two processes (acid gas removal and liquefaction) for PLNG and to investigate the economic feasibility of PLNG production by comparing it with the LCC of LNG production.
2. Simulation methodology and economic evaluation

Before simulating the processes, the feed flow rate capacity is assumed to be 0.5 million tons per annum (MTPA) to satisfy the capacity of a small FLNG installation. Additionally, rich feed (with a CO\textsubscript{2} concentration in the feed gas of 5 mol\%) and lean feed (with a concentration of 0.5 mol\%) gases are considered for the feed gas. In the rich feed case, an acid gas removal process should be included for both LNG and PLNG liquefaction. In contrast, in the lean feed case, the acid gas removal process can be omitted for PLNG when the PLNG pressure is higher than 15 bar. Because the solubility of CO\textsubscript{2} in PLNG (higher than 15 bar) is higher than 0.5 mol\%, no solidification occurs during the liquefaction process. The solubility of CO\textsubscript{2} depending on the product pressure is tabulated in Table 1.

| Product pressure (bar) | CO\textsubscript{2} solubility (mol\%) |
|------------------------|--------------------------------------|
| 1                      | 50 ppm                               |
| 5                      | 0.1                                  |
| 15                     | 1                                    |
| 25                     | 3                                    |

2.1. Simulation of the Acid gas removal process

An amine process with chemical solvent is considered for the acid gas removal process, and 35 wt\% of diethanolamine (DEA) in the form of an aqueous solution is used as a chemical solvent to meet the maximum concentration requirement of CO\textsubscript{2} for LNG production [5]. Figure 1 shows a process flow diagram (PFD) of this process. The amine process involves the use of an absorption unit and an amine regeneration unit. In the absorption tower, the feed gas enters the bottom of the tower and comes into contact with the solvent. The CO\textsubscript{2} in the feed gas is then dissolved into the solvent and the sweetened gas leaves the top of the tower. To maintain the pressure of the feed gas, the pressure at which the absorption tower operates is identical to that of the feed gas. After absorption, the rich amine solution which contains the CO\textsubscript{2} is flashed to 2 bar and temporarily stored in a flash tank to eliminate the dissolved hydrocarbons in the solution. In the regeneration column, the CO\textsubscript{2} in the rich amine solution is removed as vapor at the top of the column and the lean amine solution is generated at the bottom. After the regeneration step, the temperature of the lean amine solution is decreased and the pressure of the solution is increased to the operating condition of the absorption tower. The lean amine solution is then recycled to the tower. The simulation is performed using Aspen HYSYS (Aspen Technology, Inc., USA). The detailed parameters used in the simulation are given described in Table 2.

The circulation rate of the amine solution is determined by the CO\textsubscript{2} concentration in the sweet gas. Because PLNG can contain more CO\textsubscript{2} than LNG, the amount of CO\textsubscript{2} that should be removed in the tower for PLNG production is less than that for LNG production. Generally, the amount of the CO\textsubscript{2} removed in the absorption column is proportional to the circulation rate of the amine solution. The circulation rate is calculated by adjusting the flow rate of the lean amine solution to the absorption tower until the CO\textsubscript{2} concentration in the sweet gas reaches a retainable concentration. The retainable concentration of CO\textsubscript{2} in the sweet gas is assumed to be identical to the solubility of CO\textsubscript{2}, as tabulated in Table 1.

2.2. Simulation of the Liquefaction process

A nitrogen expansion cycle, known as suitable for small-scale liquefaction, is considered for the liquefaction process [6]. Figure 2 illustrates the PFD of the liquefaction process. The nitrogen expansion cycle utilizes a cryogenic heat exchanger, a multi-stage compressor, and a turbo-expander. The nitrogen refrigerant is compressed at the multi-stage compressor up to 120 bar. The refrigerant is then pre-cooled in the heat exchanger. The pre-cooled refrigerant is expanded at the turbo–expander and become a cold refrigerant, which liquefies the feed gas. However, before the feed gas enters the liquefaction process, the water in the feed gas is eliminated by a dehydration process. In this study, the sweet gas which enters the liquefaction process is assumed to have been dehydrated earlier. The dehydrated sweet gas is liquefied by the cold refrigerant at the heat exchanger. After liquefaction, the sweet gas is expanded to the product pressure and sent to the storage tank. The simulation of the
liquefaction process is also performed using Aspen HYSYS, and the detailed parameters for this simulation are described in Table 3.

| Item                                       | Value             |
|--------------------------------------------|-------------------|
| Operating pressure in the absorbtion tower | 70 bar            |
| Inlet temperature of the feed gas and the lean amine | 35 °C            |
| Number of stages in the tower              | 10                |
| Operating pressure in the regeneration column | 2 bar             |
| Number of stages in the column             | 10                |
| Minimum temperature approach in the Heat exchanger | 10 °C            |
| Inlet temperature to the regeneration column | 90 ~ 100 °C       |

Figure 1. Process flow diagram of the acid gas removal process.

2.3. Methodology for the economic evaluation

To estimate the economic feasibility of PLNG, a LCC analysis is conducted. LCC is usually divided into capital expenditures (CAPEX) and operational expenditures (OPEX). CAPEX is calculated using what is known as the bare module cost method [8]. As shown in equation 1, this method estimates the capital investment cost of the process \((C_{BM})\) by multiplying the purchase cost of the equipment \((C_p)\), a material factor \((F_M)\), a pressure factor \((F_p)\), and cost factors \((B_1\) and \(B_2\)). The purchase cost of the equipment is obtained from the Aspen process economic analyser (Aspen Technology, Inc. USA), and the material and pressure factors are estimated by considering the operating conditions of the equipment. Finally, the CAPEX is scaled to current prices using the chemical engineering plant cost index (CEPCI). OPEX is calculated by equation 2 [3]. OPEX usually expressed in a typical range of the utility \((C_{utility})\), labor \((C_{OL})\) and capital costs \((C_{CAPEX})\). The required utilities and the corresponding costs are estimated from the duty of the cooler, the compressor, the reboiler, and the condenser used in the processes. The labor cost is calculated by considering the number of pieces of equipment used in the processes. The lifecycle of the processes is assumed to be 15 years. After the estimation step, CAPEX and OPEX are expressed as normalized values.

\[
CAPEX \text{(USD)} = C_{BM} = C_p F_{BM} = C_p [B_1 + B_2 F_p F_M] \quad (1)
\]

\[
OPEX \text{(USD/yr)} = 0.18 C_{capex} + 2.73 C_{OL} + 1.23 C_{utility} \quad (2)
\]
Table 3. Simulation parameters for the liquefaction process [7].

| Item                                                      | Value               |
|-----------------------------------------------------------|---------------------|
| Number of stages in the multi-stage compressor            | 3                   |
| Inlet pressure at the compressor                          | 6.3 bar             |
| Outlet pressure at the compressor                         | 120 bar             |
| Minimum temperature approach in the heat exchanger        | 3 °C                |
| Pressure drop in the heat exchanger                       | 0 bar               |
| Product pressure                                          | 1, 5, 15, 25 bar    |
| Product temperature                                       | -167 °C~ -102 °C    |

Figure 2. Process Flow Diagram of the liquefaction process.

3. Results and discussion

3.1 Simulation results of the processes

Figures 3 and 4 present the simulation results of the acid gas removal process. Figure 3 presents the circulation rate of the amine solution per unit of kgmole of the feed gas. For the rich feed gas, when the final product pressure is lower than 5 bar, a larger amount of solution is added to the absorption tower as compared to the feed gas to remove a large amount of CO₂ in the feed gas. However, when the final product pressure is higher than 15 bar, a smaller amount of solution than the feed gas is enough to remove the CO₂ in the feed gas. Figure 4 presents the total heating and cooling duty in the equipment per unit of kgmole for the feed gas. The duty in the equipment decreases considerably as the final product pressure increases. When the final product pressure is 15 bar, the equipment consumes only 65% of the energy relative to that at 1 bar. In the case of the lean feed gas, similar results are obtained.

Figure 5 presents the simulation results of the liquefaction process. As shown in this figure, when the product pressure is lower than 5 bar, the net power consumed in the compressor is identical in the two cases. However, when the product pressure exceeds 15 bar, the net power levels differ slightly because, in the rich feed case, the sweet gas, which enters the liquefaction process, contains the maximum amount of CO₂ in the stream. Hence, the compressor in the rich feed case consumes more energy than that in the lean feed case. However, the CO₂ containment in the sweet gas does not have much of an effect on the net power of the compressor.
Figure 3. Circulation rate of the amine solution per unit of kgmole for the feed gas.

Figure 4. Heating and cooling duty per unit of kgmole for the feed gas.

Figure 5. Net power in the compressor per unit of kgmole for sweet gas.

Figure 6. Total LCC for acid gas removal and liquefaction.
3.2 Results of the economic evaluation

Figure 6 presents the LCC for the acid gas removal and liquefaction processes with the product pressure. The estimation results show that the total LCC for the rich feed is decreased by nearly 35% when the liquefaction pressure exceeds 15 bar relative to that at 1 bar. Furthermore, 40% of the total LCC is saved when the acid gas removal unit is not required.

4. Conclusion

In this study, an economic analysis of acid gas removal and liquefaction processes for LNG and PLNG production were conducted and the results were compared to the cost for LNG production. When the CO₂ solubility in the feed gas exceeded the CO₂ concentration in the feed, the acid gas removal unit could be omitted and CO₂ venting to the atmosphere was not required. The evaluation results showed that the PLNG production was much more economical than LNG production. Consequently, PLNG can be an economical and eco-friendly technology for near-shore FLNG projects.

Acknowledgements

This research was supported by a grant from the LNG Plant R&D Centre funded by the Ministry of Land, Infrastructure and Transport of the Korean government.

References

[1] Jeanneau E, Chretien D, and Fuchs V 2013 17th Int. Conf. & exhibition on liquefied natural gas (Houston)
[2] Xiong X, Lin W, and Gu A 2015 Energy 93 1-9
[3] Lee S, Seo Y, Lee J, and Chang D 2016 J. of Natural Gas Science and Engineering 33 405-418
[4] Lee J, Jung J, and Kim K 2012 Offshore Technology Conf. (Houston)
[5] Kidnay A J, Parrish W R, and McCartney D G 2011 Fundamentals of natural gas processing Vol. 218 (Boca raton: CRC Press.)
[6] Barclay M A, Gongaware D F, Dalton K, and Skrzypkowski M P 2004 Cryogenic Engineering Conf. (Anchorage) Vol. 710 No. 1 pp 75-82 (Melville: AIP Publishing)
[7] Venkatathram G 2008 Cryogenic mixed refrigerant processes (New York: Springer) 149-220
[8] Turton R, Bailie R C, Whiting W B, and Shaewitz J A 2008 Analysis, synthesis and design of chemical processes (London: Pearson Education)