The developing role of gas in decarbonizing China's energy system

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Comprehensive Review of Current Natural Gas Liquefaction Processes on Technical and Economic Performance

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Abstract

This paper provides a quantitative technical and economic overview of the status of natural-gas liquefaction (LNG) processes. Data is based on industrial practices in technical reports and optimization results in academic literature, which are harmonized to primary energy input and production cost. The LNG processes reviewed are classified into three categories: onshore large-scale, onshore small-scale and offshore. These categories each have a different optimization focus in academic literature. Besides minimizing energy consumption, the focus is also on: coproduction for large-scale; simplicity and ease of operation for small-scale; and low space requirement, safety and insensitivity to motion for offshore. The review on academic literature also indicated that optimization for lowest energy consumption may not lead to the lowest production cost. The review on technical reports shows that the mixed-refrigerant process dominates the LNG industry, but has competitions from the cascade process in large-scale applications and from the expander-based process in small-scale and offshore applications. This study also found that there is a potential improvement in adopting new optimization algorithms for efficiently solving complex optimization problems. The technical performance overview shows that the primary energy input for large-scale processes (0.031–0.102 GJ/GJ LNG) is lower than for small-scale processes (0.049–0.362 GJ/GJ LNG). However, the primary energy input for identical processes do not necessarily decrease with increasing capacity and the performance of major equipment shows low correlation with scale. The economic performance overview shows specific capital costs varying significantly from 124–2,255 $/TPA LNG. The variation could be, among others, caused by the different complexities of the facility and different local circumstances. Production cost, excluding feed costs, varies between 0.69–4.10 $/GJ LNG, with capital costs being the dominant contributor. The feed cost itself could be 1.51–4.01 $/GJ LNG, depending on the location. Lastly, the quantitative harmonization results on technical and economic performance in this study can function as a baseline for the purpose of comparison.

Keywords

LNG; optimization; energy efficiency; CAPEX; OPEX; harmonization

1. Introduction

With the expected global population growth and economic development, energy demand is projected to grow rapidly. To meet this demand, and because of economic and environmental pressure, natural gas (NG) demand is expected to grow by 1.6% p.a. in the coming decades, providing a quarter of the global energy demand in 2030 [1][2][3]. By 2035, natural gas could overtake coal as the second-largest fuel source of primary energy [4]. Relatively cheap natural gas, which is now accessible because of the development of horizontal drilling and hydraulic fracturing technologies for shale, also drives the growth of natural gas production [5]. Furthermore, natural gas is often seen as a transition fuel in the move toward a low greenhouse gas (GHG) economy because it is the cleanest fossil fuel, emitting about 29% to 44% less CO_2 per unit of energy compared to oil and coal [6]. In addition, combustion of natural gas emits relatively small amounts of pollutants compared to oil and coal : 20% more and 81% less CO; 79% less and 80% less NO_x; 99.9% less and 99.996% less SO_2; 92% less and 99.7 % less particulates [6].

Natural gas can be transported mainly via two options: gaseous or liquefied natural gas (LNG). Currently, natural gas is transported mainly in gaseous form via pipelines. Pipelines are suitable for short- to medium-length overland transport distance. At these distances, a pipeline is less costly than an LNG chain, because there is no need for a capital-intensive liquefaction plant and regasification terminal. The typical amount of energy consumed to deliver gas via pipeline is 10–15% of the energy delivered, whereas for LNG this is about 25% of the energy delivered [7]. Transport via a pipeline also emits less GHG compared to LNG. However, the advantages of the pipeline disappear with increasing transport distance. Onshore pipelines longer than 4800 km and offshore pipelines longer than 1600 km are not economical compared to LNG [8]. Energy consumed and GHG emissions are equal for onshore pipelines and LNG with transport distances of 13000 km and 7500 km, respectively [9]. The drawbacks of pipeline gas include: lack of flexibility in the transport route; dependence of the supply mainly on long-term contracts; and the supply capacity being fixed by the pipeline pressure differential.

For alternative LNG transport, natural gas is condensed by cooling it to below –162°C, thereby reducing its volume by a factor of 600 [10]. LNG is transported cryogenically by truck, train or ship. One benefit of LNG is that one liquefaction plant can serve several regasification plants and vice versa. Furthermore, LNG can easily adjust its supply capacity and destination, making it more adaptable than pipeline gas [8]. Another advantage of LNG is that small-scale LNG and offshore LNG allow the exploitation of remote small gas resources and offshore gas reserves, for which it is not economical to build a pipeline [11][12]. To meet the increasing demand for natural gas, research institutes and major energy companies are trying to develop small-scale LNG plants to allow exploitation of the abundant smaller-sized stranded gas resources (Stranded gas fields are fields that are not commercially exploited for physical or economic reasons [13]). The demand for small-scale LNG is mainly driven by the need for environmentally friendly fuels for marine and heavy road transport, and by end users in remote supply areas or places with insufficient pipeline gas availability [2][14][15]. Offshore LNG plants are also gaining attention because of abundant offshore gas resources. An offshore liquefaction plant is even more costly than an onshore plant because of harsh conditions and space constraints. However, transporting gas from an offshore extraction platform to an onshore liquefaction plant is also costly because of the low density of the gas, costs of the subsea pipeline, gas separation equipment, and so on [16]. The construction of such an infrastructure onshore could also be more time-intensive than an offshore liquefaction plant, because offshore liquefaction plants can be easily modularized in labor-rich areas [16]. The best solution could be LNG floating production, storage, and offloading (LNG FPSO) and floating LNG (FLNG). LNG is expected to play a vital role in meeting the projected increase in energy demand. This is because the costs of all segments of LNG chain have reduced substantially compared to in the last decades [17], as well as the fact that undeveloped or unconventional gas fields are often located far from the gas market or are too small for a pipeline. According to the BP energy outlook [18], LNG will surpass pipeline gas as the main form of internationally traded gas by 2035.

There are several recent review papers on LNG processes, like by Lim et al. (2013) [19] who described several liquefaction processes that are currently commercially available. Khan et al. (2017) [20] presented an overview of LNG liquefaction technologies and summarized key parameters for technology selection of onshore processes. Qyyum et al. (2018) [21] provided a comprehensive review focusing on optimization of LNG processes. He et al. (2018) [22] provided a state-of-the-art review on recent developments of onshore and offshore LNG processes and potential developments of LNG process optimization. However, a quantitative overview of the technical and economic performance of each

LNG process is missing. It is unclear how the capacity of the LNG process could affect its technical and economic performance. It is also interesting to investigate the difference in improvements between LNG processes. This review paper aims to first compare the different improvements made in studies for each LNG process, and then to provide a harmonized quantitative overview of technical and economic performance related to capacity.

This paper starts with an overview of the state-of-the-art in LNG processes regarding industrial application and academic research. The LNG processes are divided into three categories: onshore large-scale, onshore small-scale, and offshore. The academic literature is organized according to the different improvements made for each process. Then, a quantitative overview of the technical and economic performance of LNG technologies is given with respect to harmonizing capacity, primary energy input, capital costs, and total production cost of each process. The data are obtained from technical reposts and academic literature. Lastly, the harmonized results are discussed and recommendations for future research are given.

2. Basics and principles

LNG technologies are based on refrigeration cycles. In this study, the focus is on vapor compression cycle and gas expansion cycle. The main difference of two cycles is that: the refrigerant experiences phase change in vapor compression cycle and the refrigerant remains gaseous in gas expansion cycle. The two cycles involve four main steps (see Figure 1): 1) compression of the refrigerant to a high-pressure, hot stream (compressor); 2) heat released from compressed refrigerant (condenser or cooler and heat exchanger); 3) expansion of the compressed refrigerant to a low-pressure, cold stream (valve or expander); and 4) heat absorbed by the cold refrigerant (heat exchanger). The last step is where the cooling duty is provided to the natural gas. By repeating these four steps, natural gas can be cooled continuously.

LNG technologies can be categorized into three main types: cascade technology (Cascade), mixed refrigerant technology (MR), and expander-based technology (EXP) [16] (see Figure 2). Cascade normally has three refrigeration cycles, each at a different temperature level and containing pure propane, ethylene, and methane, respectively, as refrigerant. In MR, there is only a single refrigeration cycle. This single cycle requires a refrigerant that is composed of a mixture of light hydrocarbons. In EXP, pure nitrogen or methane is used as the refrigerant. These refrigerants can reach the low temperatures needed for the liquefaction of NG in a single loop, but at the cost of a lower efficiency compared to those of Cascade and MR. To reduce energy consumption, the EXP process recovers part of the compressor work by replacing the throttling valve with an expander. The advantages and disadvantages of the three LNG technologies are summarized in Table 1, which is modified from Lim et al. (2013) [19]. The differences between the technologies are mainly caused by the inherent complexity of them: three separate cycles for Cascade, a single cycle with a mixed refrigerant for MR, and a single cycle with pure refrigerant for EXP.

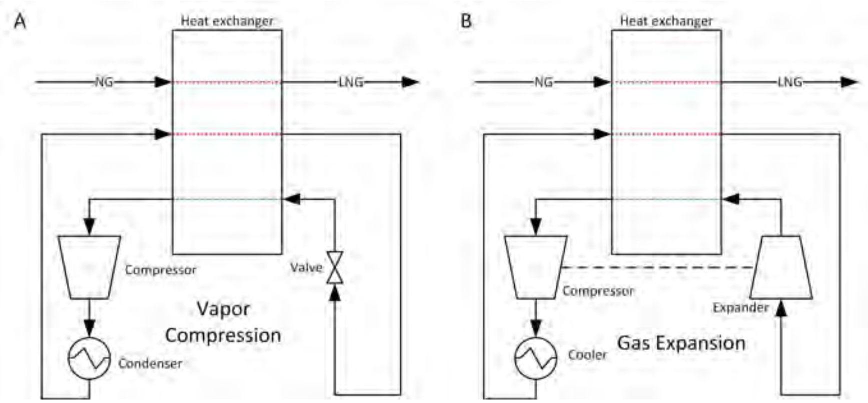

Figure 1. Schematic of vapor compression and gas expansion cycle

The energy consumption of liquefaction is closely related to the cooling curve of the process. Figure 3 shows the cooling curves of Cascade, MR, and EXP. Because Cascade uses multiple refrigerants, it has several cooling temperature levels. This allows for small temperature differences between the hot and cold sides in the heat exchangers [23]. MR can mimic the natural-gas cooling curve by using a refrigerant consisting of a carefully selected mixture of hydrocarbons. It has an even smaller temperature difference than that of Cascade, but it also requires more heat-exchange surface area [16]. The pure refrigerant in EXP remains in a gaseous state throughout the process, resulting in a constant specific-heat value for the cooling curve. EXP has a relatively large temperature difference between the refrigerant and natural gas, especially on the high-temperature end, resulting in high energy consumption [16]. Although the large temperature difference can reduce the heat-exchanger area, this is countered by the much lower heat-transfer coefficient of nitrogen compared to that of hydrocarbons [16].

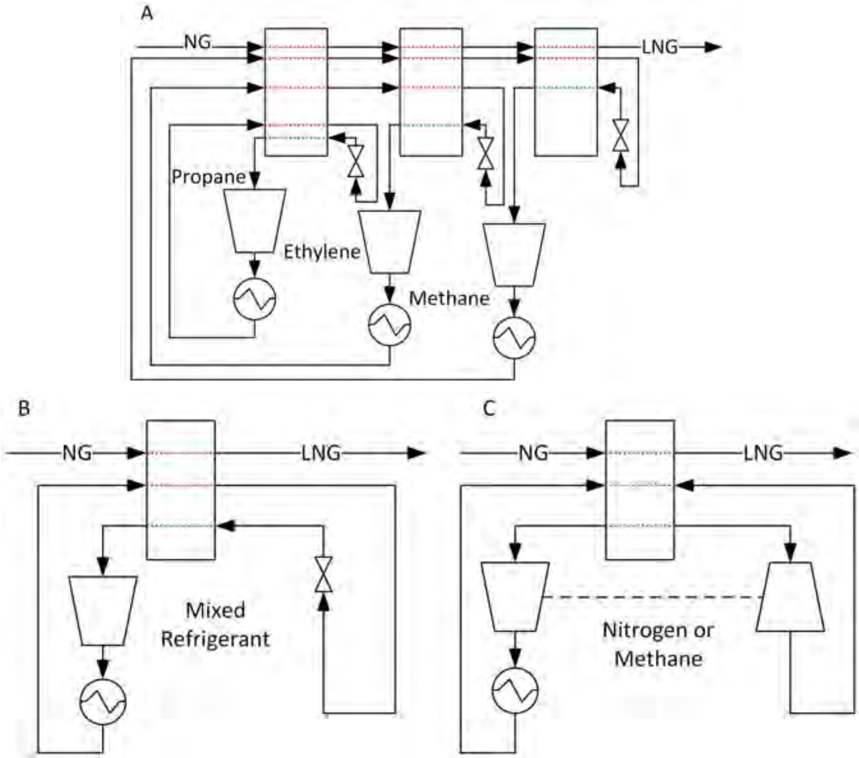

Figure 2. Schematic overview of three liquefaction technologies (A=Cascade, B=MR, C=EXP)

Table 1. Evaluation criteria for three LNG technologies (based on [19][8][16][20])

Criteria	Cascade	MR	EXP
Application	Onshore large-scale	Onshore large-scale, small-scale and offshore	Onshore small-scale and offshore
Energy efficiency	High	Medium to high	Low
Equipment count	High	Low to medium	Low
Heat-transfer surface area	Medium	High	Low
Simplicity of operation	Low	Low to medium	High
Ease of start-up and line-up	Medium	Low	High
Adaptability of feed-gas compositions	High	Medium	High
Sensitivity to ship motion	High	Medium to high	Low
Space requirement	High	Medium	Low
Hydrocarbon-refrigerant storage	High	Medium to high	None
Capital costs	High	Low to medium	Low

The evaluation of criteria for three LNG technologies are based on relative comparison.

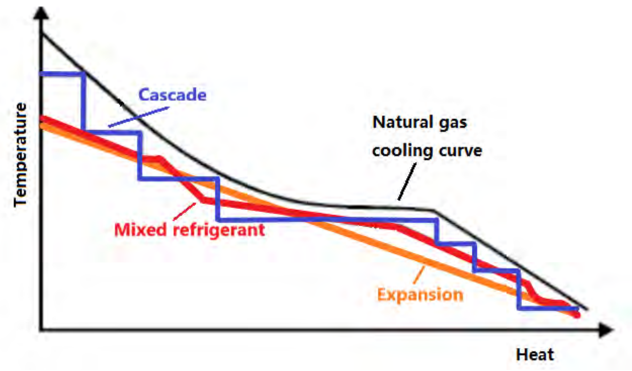

Figure 3. The cooling curve of Cascade, MR and EXP (modified from[23])

3. Natural-gas liquefaction processes

The LNG processes are divided into three categories: onshore large-scale (capacity > 1 million tonnes LNG per annum (MTPA)), onshore small-scale (capacity < 1 MTPA), and offshore processes. The industrial practices for LNG processes are obtained mainly from the Handbook of Liquefied Natural Gas [16], supplemented by [19][20][21][22]. The description and diagram of LNG processes can be found in previous reviews [19][20][21][22]. The academic studies on LNG processes are based on publications from 1998 to 2018 (which are provided by Qyyum et al. (2018) [21] and He et al. (2018) [22]), and are organized according to the improvements to the processes. In addition, the comparison between the LNG processes and developments in optimization algorithms are summarized.

3.1. Onshore large-scale natural-gas liquefaction processes

As mentioned above, the LNG processes used in onshore large-scale applications are Cascade and MR processes. Different variations of Cascade and MR are summarized in Table 2, while their applications, along with start years and capacities, are summarized in Figure 4. Several other commercial processes designed for onshore large-scale plants are not included in this review because they are considered unproven by industrial standards. The large-scale LNG industry is dominated by AP-C3MR, AP-X, and CPOC [24]. AP-C3MR is the most utilized process, and CPOC has become widely used since 2000 (Figure 4). AP-X technology is specially designed to use the advantages of both MR and EXP.

Table 2. Large-scale processes and specific features

Technology	Process name and supplier	Abbreviation	Specific features
Cascade	ConocoPhillips Optimized Cascade	CPOC	Evolved Cascade technology
	Statoil/Linde Mixed Fluid Cascade	MFC	A closer matching NG cooling curve
MR	APCI Propane Precooled Mixed Refrigerant	AP-C3MR	Most utilized process
	APCI AP-X	AP-X	Nitrogen expander sub-cooling cycle
	Shell Dual Mixed Refrigerant	DMR	MR precooling cycle

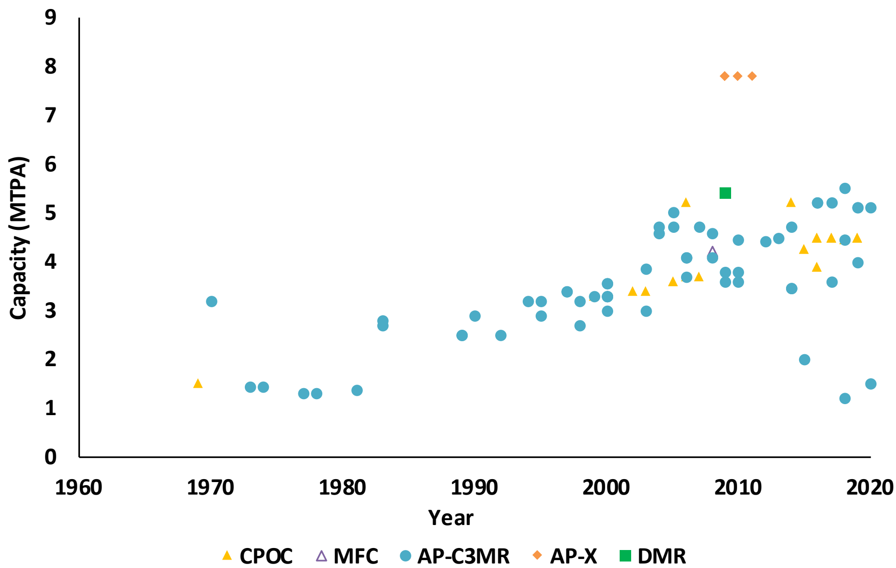

Figure 4. Applications of onshore large-scale LNG plants [25][26][27]

3.1.1. Onshore large-scale cascade processes

ConocoPhillips Optimized Cascade (CPOC) [28].

The original CPOC was developed by the Phillips Petroleum Company in the 1960s and first used in 1969 in the Kenai LNG plant [16]. An improved version of CPOC (Figure 5) became available in 1999 and was utilized in the Trinidad LNG plant [19]. It uses pure propane, ethylene, and methane as refrigerants in the three cycles. The methane cycle is an open cycle and avoids the need for a fuel compressor [19]. The maximum CPOC train capacity built to date is 5.2 MTPA. The number of plants using the CPOC process increased after 2000, partly because of its suitability for the recently developed coal-bed methane projects in Australia and the US.

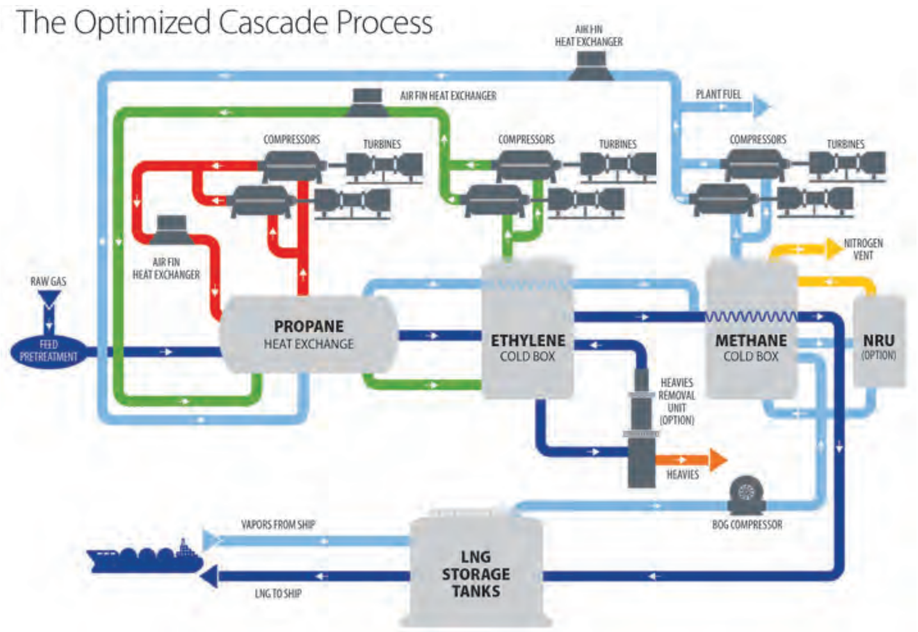

Figure 5. ConocoPhillips Optimized Cascade process flow diagram [29]

The improvements in academic studies on the CPOC process can be categorized into: A). Adopting a new refrigerant, B). Configuration adjustment, and C). Optimization according to ambient temperature (Table 3).

Table 3. Academic studies on CPOC process

Improvement	Measure	Reference
A new refrigerant combination: C_3H_8, N_2O, and N_2	Optimization of first- and second-stage pressure of the three cycles adopting a new refrigerant combination	[30]
Configuration adjustment	Optimization of operating temperatures of the propane precooling cycle, pressurized-LNG , and replacement of expansion valves with expanders	[31][32][33]
Optimization according to ambient temperature	Influence of different sea-surface temperatures	[34]

Statoil/Linde Mixed Fluid Cascade process (MFC) [35].

The MFC process (see Figure 6) is developed by the Statoil/Linde LNG technology alliance as an alternative to LNG baseload plants that is suitable for harsh environments [16]. The only industrial application is the Snøhvit plant located in Norway, with a capacity of 4.2 MTPA. The difference between the MFC process and a conventional cascade process is that MFC uses mixed refrigerant instead of pure refrigerant for the three cycles (see Figure 6). Using mixed refrigerants reduces energy consumption because a closer temperature match between the refrigerant and NG in the heat exchanger can be achieved [19]. However, performance and reliability problems with the MFC process in Snøhvit from 2008 to 2016 resulted in multiple unplanned shutdowns [20].

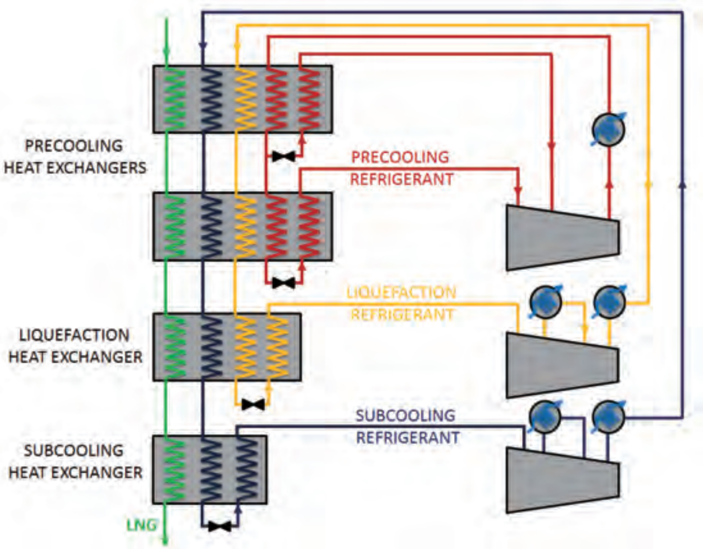

Figure 6. Mixed Fluid Cascade process flow diagram [36]

The improvements in academic studies on the MFC process can be categorized into: A). Optimization of decision variables, B). Integration of LNG process, natural gas liquids recovery (NGL) and nitrogen removal, C). Configuration adjustment, and D). Optimization depending on ambient temperature (Table 4).

Table 4. Academic studies on MFC process

Improvement	Measure	Reference
Determine the active constraints and optimize the decision variables	Optimization of three NG cooling temperatures and one compressor outlet pressure	[37]
Integration of LNG and NGL coproduction and/or nitrogen removal	Optimization of composition, mass flow rate, and pressure levels of refrigerant in heat exchanger; analysis on methane content of feed, and cold recycle temperature and ratio	[38][39]
Configuration adjustment	Optimization of precooling cycle with three pressure levels	[40]
Optimization depending on ambient temperature	Influence of different sea-surface temperatures	[34]

3.1.2. Onshore large-scale mixed-refrigerant processes

APCI Propane Precooled Mixed Refrigerant process (AP-C3MR) [41].

Air Products & Chemicals Inc. (APCI) developed the AP-C3MR process. As shown in Figure 7, the AP-C3MR process has two main cycles: propane precooling cycle and mixed-refrigerant liquefaction cycle. NG is precooled by propane to an intermediate temperature and then liquefied by the mixed refrigerant. The mixed refrigerant is also precooled by the propane before entering the main heat exchanger [16]. A new design called SplitMR is now being utilized in several new plants [19]. The SplitMR technology separates the propane and MR compressors equally between two similar-capacity gas turbines. The maximum AP-C3MR single-train capacity currently being used is 5.5 MTPA.

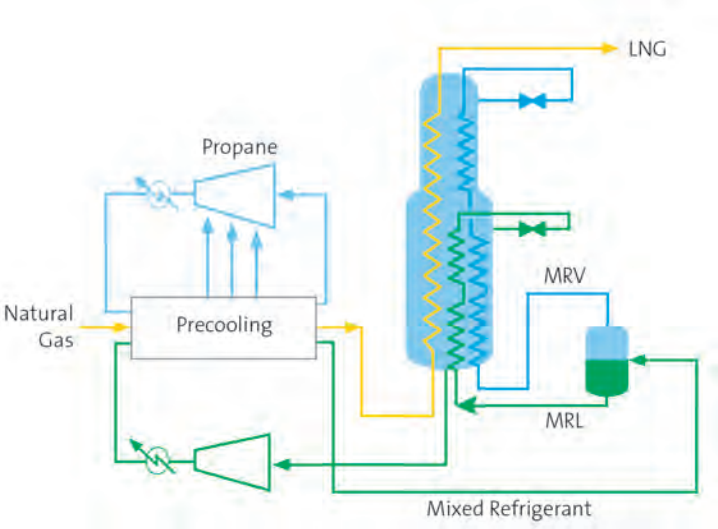

Figure 7. AP-C3MR process flow diagram [41]

APCI AP-X (AP-X) [42].

AP-X process is a modified process based on AP-C3MR to increase the train capacity. It has a third nitrogen expander cycle providing subcooling [16], as shown in Figure 8. This nitrogen expander cycle allows the flow of propane and mixed refrigerant to be reduced by 20% and 40% [20], respectively. Therefore, the process can reach a higher capacity than that of the AP-C3MR process. This process is used in Qatar at a capacity of 7.8 MTPA per train.

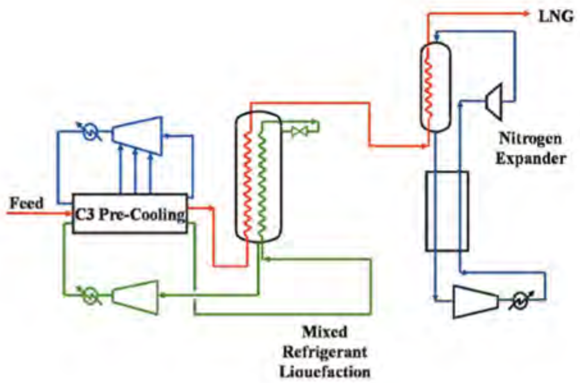

Figure 8. AP-X process flow diagram [42]

The improvements in academic studies on the C3MR process can be categorized into: A). Optimization of decision variables, B). Adopting new optimization objectives, C). Integration of LNG process, NGL process and/or a power plant, D). Heat integration, and E). Configuration adjustment (Table 5).

Table 5. Academic studies on C3MR process

Improvement	Measure	Reference
Optimization of decision variables	Optimization of pressure levels, temperature levels, mass flow, and mole composition of mixed refrigerant, and precooling temperature	[43][44][45][46][47][48][49]
Adopting new objective functions	Maximization of Exergy efficiency, and minimization of capital costs and operating costs	[50][51][52]
Integration of LNG, NGL, or a power plant	Analysis on methane content of feed, and cold recycle temperature and ratio.	[38][53]
Heat integration	Enhancement of precooling cycle with wasted heat-powered absorption cycle, and cold recovery of flash gas	[54][55][56]
Configuration adjustment	Replacement of expansion valves by two-phase expanders and liquid expanders	[57]

Shell Dual Mixed Refrigerant process (DMR) [58].

The configuration of the DMR cycle is similar to that of C3MR, except that DMR uses a mixed refrigerant instead of propane for precooling cycle in order to overcome the limitations of compressor size, as shown in Figure 9. The DMR process is very suitable for cold climates because, compared to propane, the mixed-refrigerant precooling cycle can better handle temperature and storage limitations [16]. The DMR cycle is used in the Sakhalin LNG project with a single train capacity of 5.2 MTPA. Further development in DMR is in utilizing an electrical driver instead of a mechanical driver. Electrical drivers are more costly than mechanical ones in daily operation, but have a higher availability [58]. The electrical driver version of DMR is suitable for a capacity of 5 – 8 MTPA [58].

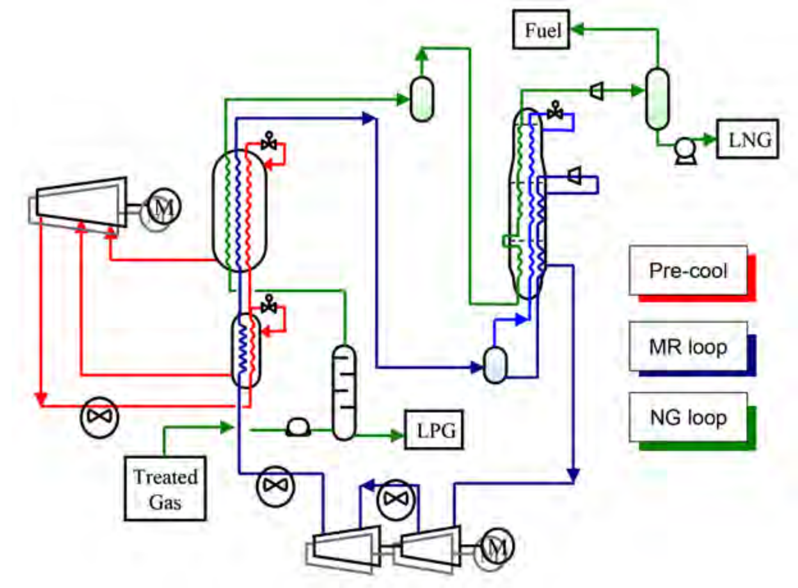

Figure 9. Shell Dual Mixed Refrigerant process flow diagram [58]

The improvements in academic studies on the DMR process can be categorized into: A). Optimization of decision variables, B). Adopting new optimization objectives, and C). Integration with NGL process (Table 6).

Table 6. Academic studies on DMR process

Improvement	Measure	Reference
Optimization of decision variables	Optimization of pressure levels, temperature levels, mass flow, and composition of mixed refrigerant	[59]
Adopting new optimization objectives	Maximization of exergy efficiency, and minimization of capital costs, operating costs, total production cost, and heat-exchanger surface area	[51][60][61]
Integration of LNG and NGL process	Analysis on exergy efficiency, methane content of feed and different operating conditions	[38][62]

3.2. Onshore small-scale natural-gas liquefaction processes

A summary of commercially available LNG processes for small-scale is given in Table 7. There are numerous small-scale LNG plants all over the world with a total capacity of 11.9 MTPA [63]. For some of these plants, covering 77% of installed capacity, detailed information is available and given in Figure 10 [2][16][26][35][64][65][66][67][68][69][70][71]. Based on this data, the AP and Linde processes are dominant in the small-scale LNG liquefaction market. The PRICO and AP-N process are the processes also used for capacities exceeding 1 MTPA.

Table 7. Small-scale processes and specific features

Technology	Process name and supplier	Abbreviation	Specific features
MR	Black & Veatch Pritchard PRICO Process	PRICO	Simple single MR cycle
	Technip/Air Liquide TEALARC	TEALARC	MR precooled MR cycle
	APCI Single Mixed Refrigerant Process	AP-SMR	First single MR cycle
	Linde Multistage Mixed Refrigerant process	LIMUM	Three-stage single MR cycle
	Kryopak Precooled Mixed Refrigerant Process	PCMR	Precooled MR cycle
EXP	Single Expander process	SE	Simplest expander cycle
	Air Product AP-N process	AP-N	Optimized from AP-X

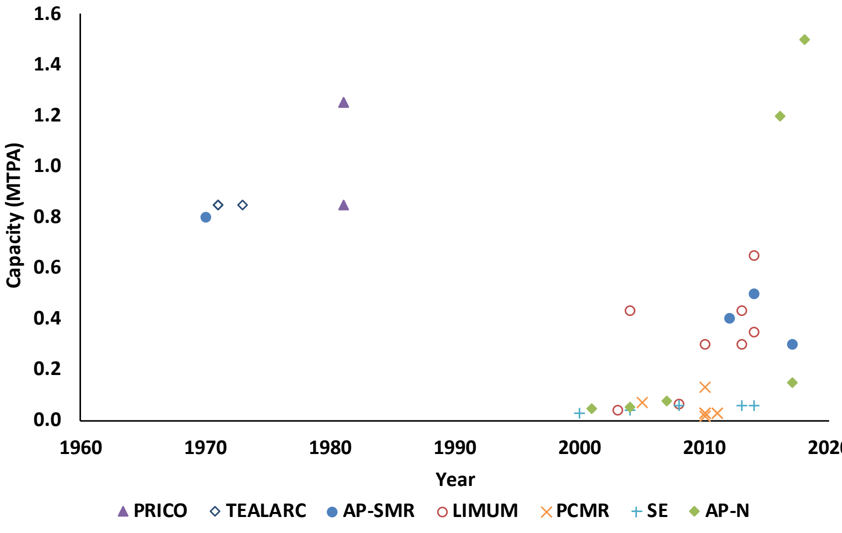

Figure 10. Application of small-scale LNG plants

PRICO: Black & Veatch Pritchard single MR Process [2][69], TEALARC: Technip/Air Liquide TEALARC [2][69], AP-SMR: Air Product single MR process [2][64], LIMUM: Linde Multistage MR process [2][35], PCMR: Kryopak Precooled Mixed Refrigerant Process [2][16][70], SE: single expander process [2][66][70], AP-N: Air Product dual nitrogen expander process [2][68].

3.2.1. Onshore small-scale mixed refrigerant processes

Black & Veatch Pritchard PRICO Process (PRICO) [72].

PRICO is a single mixed refrigerant (SMR) process developed by Black & Veatch Prichard, as shown in Figure 11. In this process, natural gas is first cooled to –35 °C, to recover the heavy hydrocarbons [16]. The PRICO process is simple to operate with a few pieces of equipment and therefore suitable for small-scale applications. This process is used in the Skikda plant in the 1970s and is used in the US and China for peak shaving and vehicle fuel applications [16][19].

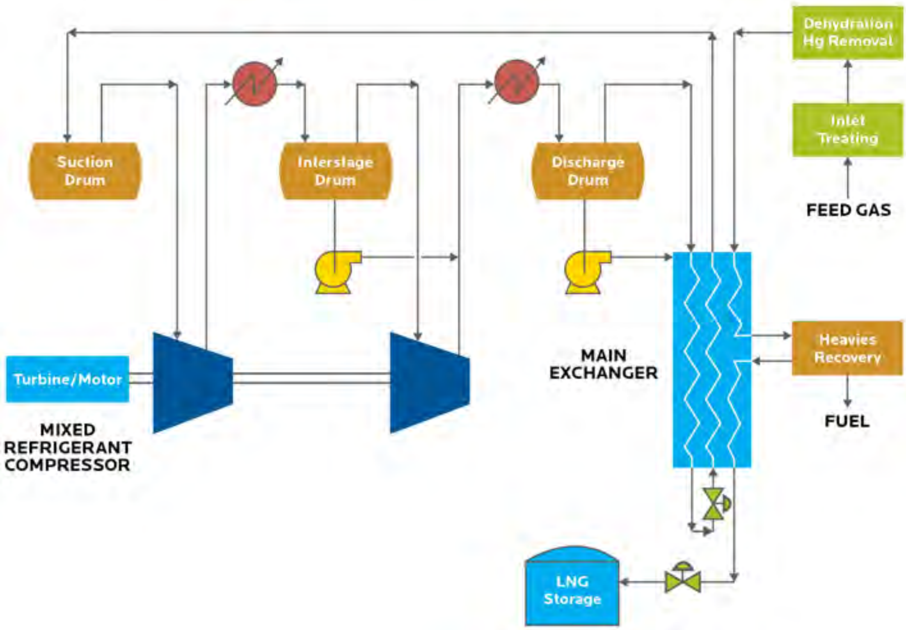

Figure 11. Black & Veatch Pritchard PRICO Process flow diagram [72]

Technip/Air Liquide TEALARC process (TEALARC) [73].

TEALARC is an MR process and consists of two mixed-refrigerant cycles. The first MR cycle precools the second MR cycle, and the second cycle cools the NG flow [74], as shown in Figure 12. This process is utilized in the Sonatrach Skikda LNG plant in 1971 and 1973, and the capacity for each train is 0.85 MTPA [69].

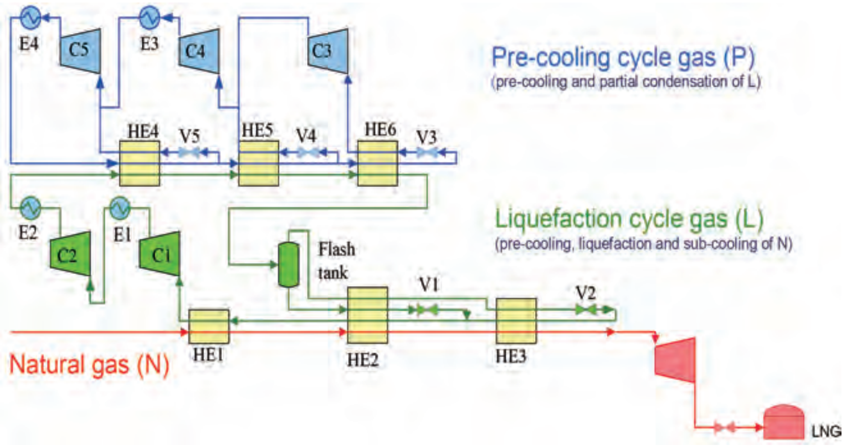

Figure 12 Technip/Air Liquide TEALARC process flow diagram [73]

APCI Single Mixed Refrigerant Process (AP-SMR) [75].

AP-SMR is the first SMR process developed by APCI, which only uses one MR cycle for pre-cooling, liquefaction, and sub-cooling. It has a lower equipment count but also a lower efficiency compared to those of AP-C3MR and DMR [75]. In this process, MR is split into warm MR and cold MR in order to improve the efficiency [20]. There are four 0.8 MTPA AP-SMR trains in Libya and two 0.4 MTPA trains in China [75].

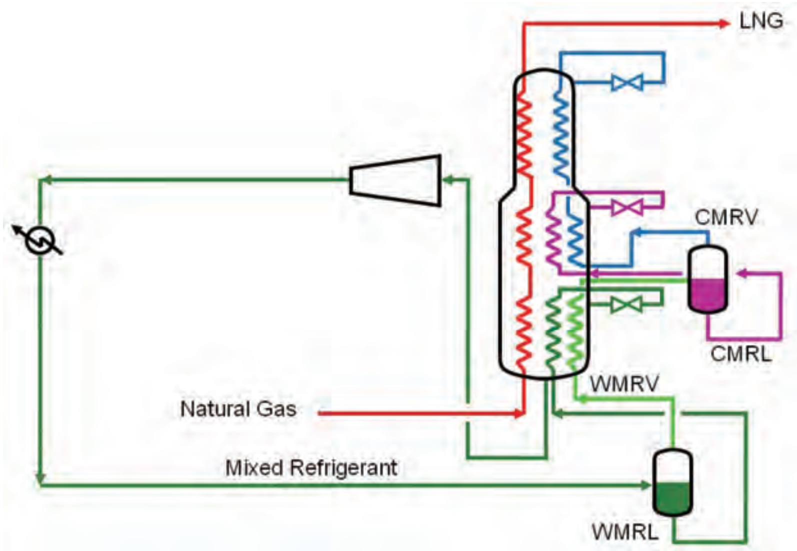

Figure 13. APCI Single Mixed Refrigerant Process flow diagram [75]

Linde Multistage Mixed Refrigerant process (LIMUM) [35].

LIMUM was a multi-stage SMR process developed by Linde. There are two utilizations of LIMUM: a two-stage mixed-refrigerant cycle using a plate-fin heat exchanger for capacities smaller than 0.5 MTPA LNG and a three-stage mixed-refrigerant cycle using a coil-wound heat exchanger for trains of 0.2 to 1.0 MTPA LNG [35]. There are several plants using this process since 2000 in Norway, China, Australia, and Malaysia.

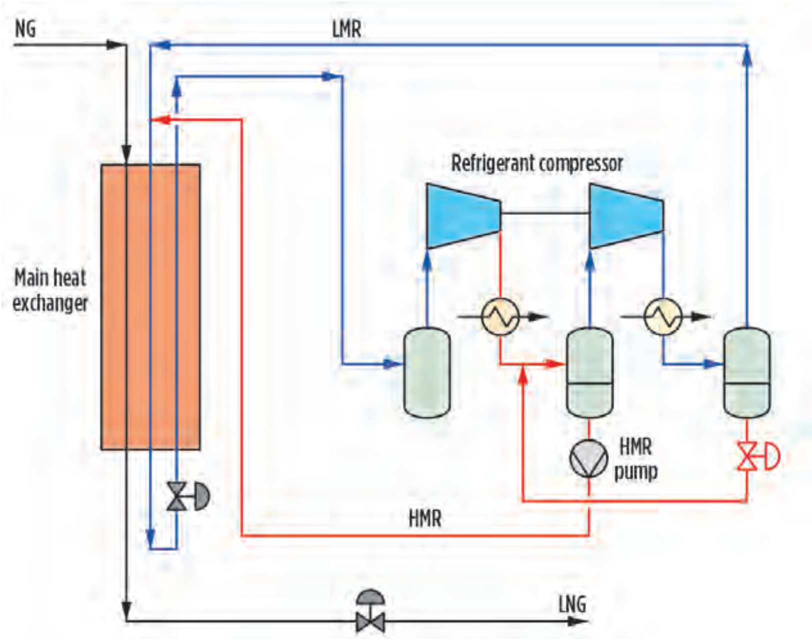

Figure 14. Linde Multistage Mixed Refrigerant process flow diagram [76]

Kryopak Precooled Mixed Refrigerant Process (PCMR) [70].

The PCMR process was developed by Kryopak, and consists of a pre-cooling cycle (propane or ammonia) and an MR cycle [16], as shown in Figure 15. The pre-cooling cycle cools the MR cycle, and then the MR cycle provides cooling duty to liquefy the NG. The PCMR cycle has been used in several small-scale plants located in China and the US with capacities of up to 0.1 MTPA [70].

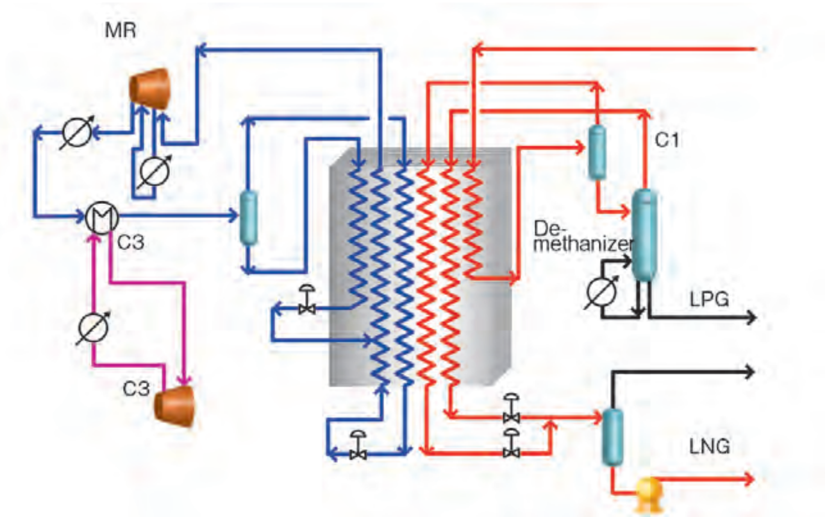

Figure 15. Kryopak Precooled Mixed Refrigerant Process flow diagram [77]

Many studies focus on SMR process optimization because it is a research hot spot in the LNG process. Their improvements can be categorized into: A). Optimization of decision variables, B). Adopting new objective functions, C). Optimization according to ambient temperature and relative humidity, D). Heat integration, E). Configuration adjustment, and F). Optimization of the operating control system (Table 8).

Table 8. Academic studies on SMR process

Improvement	Measure	Reference
Optimization of decision variables	Optimization of composition of mixed refrigerant, pressure levels of condensation, and evaporation	[43][78][79][80] [81][82][83][84]
Adopting new objective functions	Minimization of capital costs, operating costs, total production cost, and heat-exchanger surface area	[85][86][87][88]
Optimization according to ambient temperature and relative humidity	Influence of sea-surface temperatures and relative humidity in different areas	[89][90][91]
Heat integration	Recovery of the cold energy of flash gas	[56]
Configuration adjustment	Pump added three-stage compression, and pressurized LNG	[32][92]
Optimization of the operating control system	Utilization of a real-time steady-state optimization system to minimize the setpoint energy loss	[93][94]

3.2.2. Onshore small-scale expander-based processes

Single Expander process (SE) [95].

The SE process is the simplest expander process and has only one cycle with one expander, as shown in Figure 16. It works by compressing and expanding a pure refrigerant to generate the cooling duty. The refrigerant used in this process is normally nitrogen or methane [16]. Despite its relatively low efficiency, its simplicity makes it suitable for small-scale plants [19]. The main advantages of this process are its use of a non-flammable refrigerant (nitrogen), its possible lack of need for any refrigerant storage and make-up (methane), and its flexibility with different feed-gas compositions. The nitrogen single expander process has been used in several peak-shaving plants and liquefaction of boil-offs in LNG carriers [2][66][70].

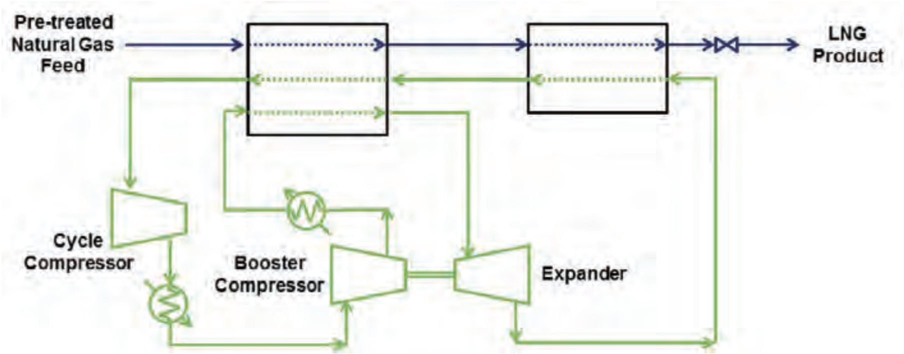

Figure 16. Single Expander process flow diagram [95]

Air Product AP-N process (AP-N) [75].

The AP-N process was developed from the AP-X process with an optimized number of expanders, pressure levels, and temperature levels [20]. It has one pressure level and two expander temperature levels. Part of the nitrogen provides pre-cooling for the rest of the nitrogen stream to improve energy efficiency. The compression is done in several stages: the expanders drive the last stage and the external drivers drive the other stages. This process is being used in several plants in Japan and the US with capacities under 0.2 MTPA [68].

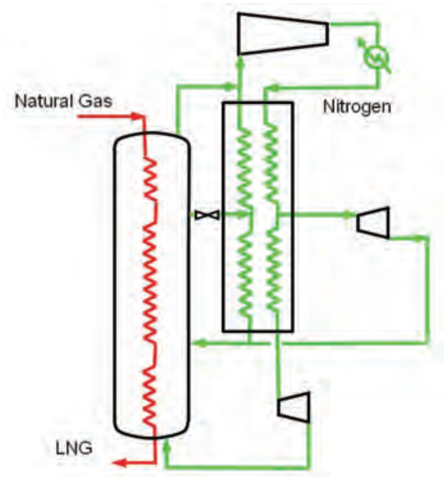

Figure 17. Air Product AP-N process flow diagram [75]

There is an increased attention on the EXP process, because it is simple and suitable for small-scale applications. The improvements can be categorized into: A). Adopting a new refrigerant, B). Adopting new objective functions, C). Heat integration, and D). Configuration adjustment (Table 9).

Table 9. Academic studies on EXP process

Improvement	Measure	Reference
Adopting a new refrigerant	Nitrogen-methane, propane-nitrogen, feed gas as a refrigerant, and liquid nitrogen and carbon dioxide	[43][79][80][96][97][98]
Adopting new objective functions	Minimization of capital costs, operating costs and total production cost; and safety-related objective	[82][88][99]
Heat integration	Heat integration between heat exchanger and regenerator, utilization of pressure exergy of pipeline gas to providing cooling duty, and recovering the cold energy of the flash gas	[56][100][101][102][103]
Configuration adjustment	Adding a precooling cycle, multistage expansion, utilization of two-phase expander, pressurized LNG concept, and open loop concept	[11][32][79][97][104][105][106]

3.3. Offshore natural-gas liquefaction processes

The criteria for process selection for offshore are different from those for onshore applications. For an offshore application, the small footprint of equipment, ease of maintenance, sensitivity to motion, and safety are more important than efficiency and maximum capacity, because of a lack of deck space and ocean environment [107]. The characteristics of offshore applications make MR and EXP processes more suitable than the Cascade process [108]. Currently operating offshore LNG plants are shown in Figure 18.

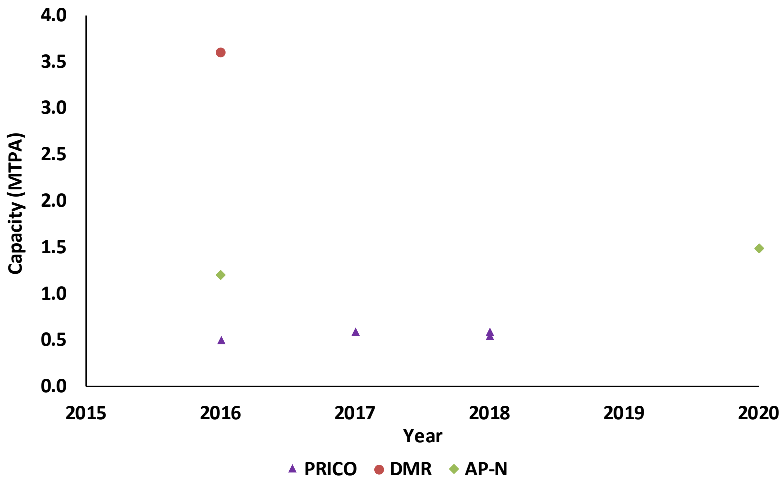

Figure 18. Application of offshore LNG plants [24][107]

3.3.1. Offshore mixed-refrigerant processes

MR technology has been applied on offshore liquefaction plants for single mixed refrigerants (PRICO) and dual mixed refrigerants (DMR). PRICO is utilized for small train capacities (below 1 MTPA) and DMR is utilized for large train capacities (beyond 1 MTPA) [107]. MR technology in the offshore application has the advantage of a relatively high thermodynamic efficiency and low refrigerant volume (as the refrigerant is in a liquid form) compared to those of the EXP process. The space used by the MR process is only half of that the EXP process [107]. However, the drawbacks for MR are: use of flammable refrigerant with safety and pipeline arrangement issues [108]; and slower start-up and shutdowns [108]. There is increasing attention on the offshore SMR process. The improvements for SMR and DMR are summarized in Table 10 and Table 11, respectively.

Table 10. Academic studies on offshore SMR process

Improvement	Measure	Reference
Optimization of decision variables	Optimization of composition of mixed refrigerant, pressure levels of condensation, and evaporation	[109][110]
Adopting new objective functions	Minimization of capital costs, operating costs, and total production cost; layout, sensitivity to motion, flexibility for gas composition, safety, and operability related objective	[111]
Configuration adjustment	Separating the mixed refrigerant in different ways, replacement of expansion valves by two-phase expanders, and pressurized LNG	[32][112][113]
Optimization of the operating control system	Development of a control structure to control the flow-rate ratio of heavy and light mixed refrigerant	[114]

Table 11. Academic studies on offshore DMR process

Improvement	Measure	Reference
Optimization of decision variables	Optimization of composition and mass flow of mixed refrigerant, pressure levels and temperature levels of process	[115][116]
Adopting new objective functions	Analysis on explosion risks for different refrigerant compositions	[116]
Configuration adjustment	Single cycle with regeneration, multi-stage compression with inter-cooling, multi-stage refrigeration, and multi-stage compression refrigeration	[117]

3.3.2. Offshore expander-based processes

The only EXP technology utilized in offshore applications is the AP-N process. The reason is that only proven onshore liquefaction processes are considered for offshore applications to minimize the risk [107]. There are two natural-gas liquefaction projects using the AP-N process: PFLNG1 and PFLNG2 in Malaysia. Compared to the MR process, the EXP process has the advantage of simplicity and low equipment count. In addition, the EXP refrigerant remains gaseous and is not sensitive to ship motions. Moreover, nitrogen is not flammable and safer than MR. The EXP process is also more flexible to gas composition, easier for operation and quicker to start-up compared to MR. The major disadvantages of the EXP process are low energy efficiency and a large space requirement. The improvements from academic studies on the offshore EXP process are summarized in Table 12.

Table 12. Academic studies on offshore EXP process

Improvement	Measure	Reference
Adopting a new refrigerant	Feed gas as a refrigerant and nitrogen-carbon dioxide	[118][119]
Optimization of decision variables	Optimization of the refrigerant flow rate, and pressure and temperature levels	[120]
Adopting new objective functions	Analysis on economic performance, sensitivity to motion, flexibility for gas composition, quick start-up, ease of operation, reliability, low space requirement, and safety	[108][111][121]
Configuration adjustment	Dual expansion and pressurized LNG	[32][120][121]

3.4. Comparison between liquefaction processes

The comparison between liquefaction processes is made based on the type of refrigerant, heat exchanger, driver, and compressor (Table 13). The heat exchanger, driver, and compressor are the most capital-intensive equipment in the process.

- The refrigerant used in liquefaction processes can be classified as either mixed or pure. MR uses a mix of specially selected light hydrocarbons, which can be adjusted to mimic the cooling curve of NG. Cascade uses several different pure refrigerants with a cascade of boiling temperature, over the cycles. EXP uses nitrogen or methane, which has a very low boiling temperature allowing this process to liquefy NG in one cycle. When the processes are ranked according to the temperature difference between the refrigerant and NG, the sequence is MR < Cascade < EXP. A smaller temperature difference can reduce energy consumption. However, it also requires a greater heat exchange area, which increases capital costs. Therefore, the liquefaction process can be optimized between the refrigerant and heat exchanger areas [16].

- Currently, there are two main types of heat exchanger in use in the LNG industry: 1) plate-fin or brazed aluminum type (PFHE) and 2) coil-wound or spiral-wound type (SWHE). PFHE has the advantage of competitive vendors, a low-pressure drop, and variability in low-temperature differences. However, it needs to be carefully designed and is very vulnerable to physical damage and thermal shocks, because it is made of aluminum [19]. SWHE is very robust and can be easily operated, but it is more expensive and proprietary to only a few companies. SWHE also has limited flexibility with feed-gas composition, and higher capital costs, footprint, and weight. The upper limit capacity for a single PFHE is 1.5 MTPA, and for a single SWHE it is 4 MTPA. These pros and cons explain why PFHE is normally used for plants using Cascade and EXP and why SWHE is normally used for large-scale MR plants [95]. For precooling, the core-in-kettle type (CKHE) heat exchanger is often used.

- There are five types of drivers and two types of compressors to be considered here. For a liquefaction process, the driver and compressor can be tailored to a specific need. Most of the processes in Table 13 do not have a fixed driver and compressor type, except for AP-X, CPOC, and DMR. They are equipped with a high-efficiency GE 9E frame-type gas turbine (maximum capacity of 3.8 MTPA LNG per turbine), an aero-derivative gas turbine, and an electric motor, respectively.

- The centrifugal and axial compressors are the most utilized compressors in the liquefaction industry. The centrifugal compressor is usually used in the precooling system, because of its low capital costs and simple design [19]. The axial compressor is usually used in the main cooling system because of its high efficiency and high compression ratio [19].

Table 13. Comparison of key components used in each process [16][20][122][123] (modified from [19])

Process		Precooling	Liquefaction	Subcooling	Driver and compressor type
CPOC	R	Propane	Ethylene	Methane	Aero-derivative gas turbine
	H	PFHE/CKHE	PFHE	PFHE	Centrifugal compressor
MFC	R	MR	MR	MR	Various turbine type
	H	PFHE	SWHE	SWHE	Axial compressor
AP-C3MR	R	Propane	MR	-	Various turbine type
	H	CKHE/PFHE	SWHE	-	Centrifugal and axial compressor
AP-X	R	Propane	MR	N_2	GE 9E
	H	CKHE	SWHE	SWHE & PFHE	Centrifugal and axial compressor
DMR	R	MR	MR	-	Electric motor
	H	SWHE	SWHE	-	Axial compressor
PRICO	R	-	MR	-	Various turbine type
	H	-	PFHE	-	Axial compressor
TEALARC	R	MR	MR	-	Various turbine type
	H	PFHE	PFHE	-	Centrifugal and axial compressor
AP-SMR	R	-	MR	-	Various turbine type
	H	-	SWHE	-	Centrifugal compressor
LIMUM	R	-	MR	-	Various turbine type
	H	-	SWHE&PFHE	-	Axial compressor
PCMR	R	Ammonia /Propane	MR	-	Various turbine type
	H	PFHE	PFHE	-	Centrifugal and axial compressor
SE	R	-	N_2 or Methane	-	Various turbine type
	H	-	PFHE	-	Axial compressor
AP-N	R	-	N_2	-	Various turbine type
	H	-	SWHE & PFHE	-	SWHE

R = refrigerant, H = heat exchanger

3.5. Process-modeling optimization algorithms

Optimization of LNG processes helps to reduce energy consumption significantly. However, this optimization is a challenge because it is a multi-variable, multi-objective, and highly non-linear problem [20]. The algorithms used in the reviewed studies are summarized in Table 14. There are three types of optimization approaches: deterministic, stochastic and hybrid approaches. The advantages of the deterministic approach are its easy-to-handle constraint, short calculation time, and small number of tuning parameters. The disadvantages are its inability to handle multiple-objective problems, need for a good initial estimate, and possibility for its convergence to end up in one of many local optimal results. The advantages and disadvantages of the stochastic approach are opposite to those of the deterministic approach. From Table 14, it is clear that the most utilized algorithms are the deterministic non-linear programming and the stochastic genetic algorithm. The hybrid approach knowledge-based optimization combines the advantages of both deterministic and stochastic approaches. It is a promising optimization algorithm because of its easy-to-handle constraint, medium calculation time, small number of tuning parameters, capability of handling multiple-objectives, independence of initial estimate, and robust convergence [20].

Table 14 Optimization algorithms

Type	Algorithm	Reference
Deterministic	Aspen Hysys non-linear programming	[30] [40] [44] [48] [49] [50] [51] [60] [78] [80] [104] [102] [103] [11] [124] [125] [126]
	Mixed-integer non-linear programming model	[127]
	Sequential quadratic programming	[46] [62] [85] [128]
	Successive reduced quadratic programming	[47] [61] [92] [87]
	Gradient assisted robust optimization algorithm	[129]
	gPROMS self-optimizing controls of active constraints	[37] [81] [130]
	Modified Dividing a hyper-RECTangle algorithm	[131]
Stochastic	Genetic algorithm	[32] [34] [39] [45] [52] [53] [59] [62] [83] [84] [86] [88] [89] [106] [96] [132] [133] [134] [135] [136]
	Non-dominated sorting genetic algorithm	[99]
	Particle swarm paradigm	[90] [97] [137]
	Sequential coordinate random search	[138]
	Evolutionary gradient free searches	[139] [140]
	Tabu Search	[139]
	Adaptive simulated annealing algorithm	[141]
	Modified coordinate descent methodology	[142]
Hybrid	Knowledge-based optimization	[15] [110] [119] [143]

4. Technical and economic performance of liquefaction processes

This section presents a harmonized quantitative overview of the technical and economic performance of LNG processes. Primary energy input, specific capital costs, and total production cost were analyzed as indicators.

4.1. Technical performance of liquefaction processes

As the previous review highlighted, there are numerous processes to liquefy NG. Comparing these processes, however, is difficult. Although the majority of studies that optimized LNG processes have the same objective, i.e., minimization of energy consumption [21], the indicators and units that they use differ. The indicators and units include: unit power consumption (kJ/kg LNG, kJ/kmol LNG, kW/(t/d) LNG, kWh/Nm^3 LNG), total shaft work (kW), coefficient of performance, and exergy efficiency. Therefore, in this study, the technical performance of the LNG processes is harmonized and expressed as the primary energy input (the primary energy (GJ HHV) needed to produce 1 GJ LNG (HHV)). For LNG, the gross energy used for calculation is 53.4 mmBtu/metric tonne, equal to 56.4 GJ HHV/metric tonne [144][145][146]. The cold energy of LNG is ignored, because it is usually not recovered at the regasification terminal. The primary energy input is calculated as specific work (kJ/kg LNG) divided by driver efficiency (Equation 1).

Equation 1

$$Primary\ energy\ input = \frac{Specific\ work}{Driver\ efficiency} \times \frac{1}{56.4 \times 1000}$$

In this section, the EXP process is divided into single expander process (SE) and other expander processes (OE), which include dual expander process and precooled expander process. The merit of using primary energy input as an indicator is that it represents the percentage of energy used to produce LNG. The capacity of the LNG plant is harmonized to the same unit (MTPA LNG), and the availability is assumed to be 340 days per year (93.2%) [147]. An assumption about gas turbines is made for technical performance harmonization. It is assumed that all other large-scale processes are equipped with a GE 7E frame-type gas turbine (maximum capacity of 2.5 MTPA LNG per turbine) and that the small-scale processes are equipped with a GE 5C frame-type gas turbine (maximum capacity of 1.1 MTPA LNG per turbine) [148]. The data of each liquefaction process were obtained from technical reports and academic literature between 1998 – 2019, and then harmonized to the same units (GJ/GJ LNG and MTPA).

Table 15. Liquefaction processes with harmonized technical performance

Process	Technical reports				Academic literature			
	Capacity range (MTPA)	Specific work (kJ/kg LNG)	Primary energy input (GJ/GJ LNG)	Reference	Capacity range (MTPA)	Specific work (kJ/kg LNG)	Primary energy input (GJ/GJ LNG)	Reference
CPOC	0.3 – 5.2	1166.4 – 1382.4	0.049 – 0.058	[42] [95] [149] [150] [151] [152]	1.3E-04 – 6.3	738.0 – 1226.5	0.031 – 0.051	[30] [31] [32] [33] [34]
MFC	3.0 – 6.0	907.2 – 1019.5	0.049 – 0.055	[152] [153]	0.47 – 7.2	612.3 – 1238.4	0.033 – 0.067	[34] [37] [38] [39] [40]
C3MR	1.3 – 7.8	1054.1 – 1080.0	0.057 –0.058	[42] [95] [122][149] [150] [151] [152] [154]	7.5E-06 – 7.5	903.9 – 1543.3	0.048 – 0.102	[38] [43] [44] [45] [46] [47] [48] [49] [50] [51] [52] [53] [56] [57] [59] [112] [127] [129] [138] [143] [155]
DMR	1.5 – 5.4	993.6 –1080.0	0.046 – 0.050	[42] [150] [151] [156]	6.1E-06 – 5.0	854.1 – 1490.4	0.040 – 0.070	[38] [51] [59] [62] [60] [61] [112] [115] [155]
SMR	0.013 – 2.4	1080.0 – 1451.5	0.066 –0.089	[42] [95] [122] [149] [150] [152] [156]	7.5E-06 – 3.8	792.7 – 5874.2	0.049 – 0.362	[15] [32] [43] [56] [59] [78] [79] [80] [81] [82] [83] [84] [85] [86] [87] [88] [89] [90] [91] [92] [93] [94] [112] [113] [110] [125] [126] [128] [130] [131] [132] [133] [136] [137] [138] [139] [140] [141] [142] [143] [157] [158]
SE	0.026 – 0.061	1425.6 – 3499.2	0.088 –0.215	[42] [95] [149] [154]	1.4E-04 – 0.93	1486.7 – 5414.1	0.092 – 0.333	[56] [59] [82] [84] [100] [105] [96] [159]
OE	0.045 – 1.5	1123.2 – 2350.1	0.069 –0.145	[95] [122] [149] [150] [154]	7.5E-06 – 1.0	1128.3 – 5112.0	0.069 – 0.315	[32] [43] [59] [79] [80] [84] [88] [104] [105] [11] [106] [96] [124] [97] [101] [99] [119] [120] [134] [135]

The assumptions for driver efficiencies are: gas turbine efficiency is 0.426 for CPOC (GE LM6000); 0.288 for SMR, SE and OE (GE 5C); 0.330 for MFC and C3MR (GE 7E) [148]; The electric motor efficiency for DMR is 0.95 and the conversion factor of primary energy to electricity is 0.40144.

The availability is assumed to be 340 days per year (93.2%) [147].

The results for harmonized technical performance are shown in Table 15 and Figure 19. The processes arranged by primary energy input, from lowest to highest, in technical reports are: DMR, MFC, CPOC, C3MR, SMR, OE, and SE. Meanwhile, the processes arranged by primary energy input, from lowest to highest, in academic studies are: CPOC, MFC, DMR, C3MR, SMR, OE, and SE. However, it should be noted that the primary energy input for large-scale processes (CPOC, MFC, DMR, and C3MR) is significantly lower than that for small-scale processes (SMR, OE, and SE) in both technical reports and academic studies. Figure 19 shows that the primary energy input drops dramatically with increasing capacity, especially after 1 MTPA. However, primary energy input for identical processes, either large-scale or small-scale, does not show a clear relationship with respect to different capacities. Based on the number of studies on each process, the C3MR process is the research hot spot for large-scale applications, and SMR and OE are gaining great attention in small-scale applications.

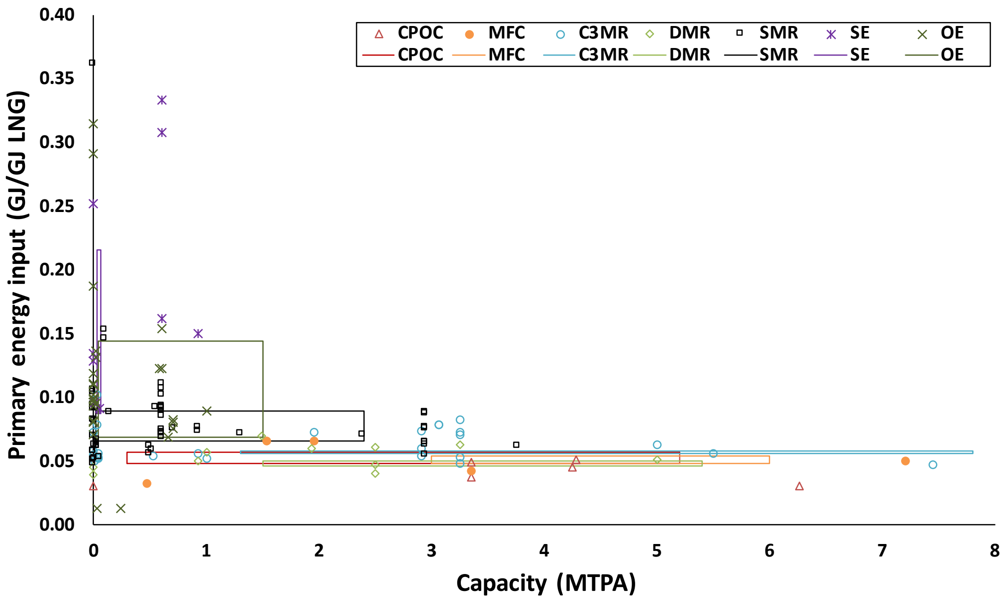

Figure 19. Technical performance by primary energy input and capacity range (dot represents for academic literature and rectangle represents for technical reports)

4.2. Economic performance of liquefaction processes

Compared to the numerous studies focusing on technical performance, only a few studies focus on the economic performance of LNG processes. In this section, specific capital costs are harmonized first, followed by the total production cost.

4.2.1. Capital costs

The capital-cost data from academic literature were gathered from eight studies [51][61][82][87][88][160][161][162] and include the C3MR, DMR, SMR, and EXP processes. Three capital-cost calculation methods were used in these studies: A). lump sum of investment method (top-down), B). six-tenth-factor rule method (bottom-up), and C). bare module cost method (bottom-up). In the studies [51] and [82], in which the lump sum of investment method was used, the total capital costs of a base capacity plant and the percentage of costs distribution of each main equipment were first obtained. Then, the target capacity plant capital costs were calculated using a scaling factor of 1. Lastly, the installed costs of each main equipment were calculated using the previously determined percentage. However, this lump sum of investment method is a non-rigorous approach because the inclusion of equipment for each plant may be different and a scaling factor of 1 is often conservative. Other studies utilized the six-tenth-factor rule method [61][87] and the bare module cost method [88][160][161][162] to estimate capital costs. The six-tenth-factor rule method and the bare module cost method are both bottom-up approaches based on cost estimation of major equipment. The six-tenth-factor rule method uses different scaling factors (normally 0.6) for each equipment to calculate the purchased equipment costs and installation costs as a whole from a base capacity to the target capacity, and then sums up the costs of major equipment to the total plant cost [87]. The bare module cost method uses a different scaling factor for each equipment to calculate the purchased equipment costs at base condition (base material and base operating pressure) first. Then, the purchased equipment costs are multiplied by a bare module cost factor (depending on direct costs, indirect costs, specific material and pressure) to the installed equipment costs. Finally, the installed equipment costs are summed up to the total plant cost [161]. Most of the studies include only the liquefaction system in the total plant capital costs; the exceptions are in [160][161]. Lee et al. (2016) [160] includes a storage system and Raj et al. (2016) [161] includes a pretreatment system and a storage system.

The industrial capital-cost data were obtained from two technical reports by Songhurst [163][164]. The technical reports include only large-scale plant capital costs in the period 2000 – 2018. The LNG plants are classified as MR (C3MR, DMR, and SMR) and Cascade (CPOC and MFC). Technical reports and academic literature show different definitions of total plant costs. The capital-cost definition used in the studies above includes total plant costs, which are the sum of individual installed equipment costs. Meanwhile, the capital-cost definition used in the technical reports is the total capital requirement. The definition includes total plant costs, allowance for funds used during construction, and owner's costs. There are two plant scopes in the technical reports: liquefaction train and complete plant. The liquefaction train includes the liquefaction system, pretreatment system, and storage system. The complete plant also requires all the necessary infrastructure besides the liquefaction train, such as a construction camp, township, dock, and breakwater [163]. The costs of the liquefaction train are roughly 66% of the cost of a complete plant [163].

The capital costs of the LNG processes are harmonized to specific capital costs ($/TPA), which are calculated as capital costs (millions of US dollar) divided by capacity (million tonnes per annum) in Equation 2.

Equation 2

$$C_{Specific\ capex} = \frac{C_{capex}}{L_{capacity}}$$

The capacity is the single-train capacity which is expressed in MTPA LNG, with an availability of 340 days per year [147] (93.2%). All costs mentioned in this paper are indexed to $\$_{2018\ Q2}$ using the IHS Upstream Capital Costs Index (UCCI). The harmonized results are shown in Figure 20. The capital-cost estimations in academic literature (125 – 1285 $/TPA) are much lower than in the technical reports (220 – 2255 $/TPA). The majority of plants in the technical reports are in a small capacity range (3.0 – 5.5 MTPA) because of the standard size of the industrial gas turbine and heat exchanger, but their specific capital costs vary significantly. Therefore, it appears that capacity is not a major factor that affects the specific capital costs, at least not at these capacities.

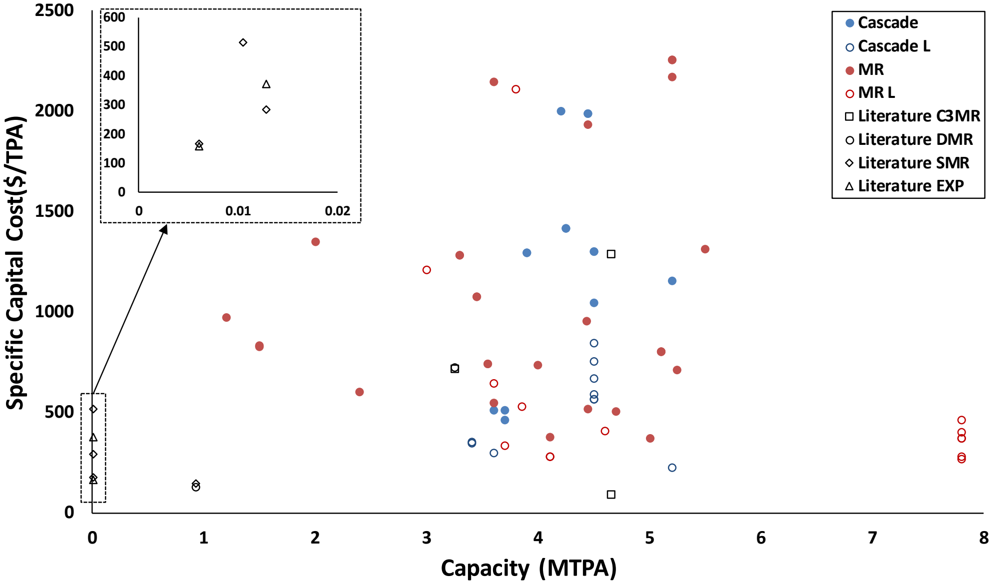

Figure 20. Specific capital costs comparison between technical reports and academic literature (L: liquefaction train only)

4.2.2. Total production cost

The total production cost is harmonized only for academic literature because of the lack of operating data in the technical reports. The total production cost is the cost to produce 1 GJ LNG ($/GJ HHV LNG), which includes two parts: amortized capital costs and amortized operation and maintenance (O&M) costs (Equation 3). The amortized capital costs are the capital costs of the plant to produce 1 GJ LNG ($/GJ LNG) by considering the discount rate and plant life (Equation 4). The discount rate (r) and plant life (n) are assumed to be 12% and 20 years, respectively [161]. The high heat value (e) of LNG is 56.4 GJ/t.

Equation 3

$$C_{total\ production\ cost} = C_{amortized\ capex} + C_{O\&M\ cost}$$

Equation 4

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

There are six main types of O&M costs considered in studies [51][61][82][87][88][160][161][162]: energy, equipment maintenance, feed natural gas, cooling water, labor, refrigerant. Most of these studies define O&M cost as energy, maintenance, and feed natural gas costs. The missing cooling water, labor, and refrigerant costs only make up a small part of the total O&M cost. Therefore, the O&M costs in this study include three parts: energy costs, maintenance costs, and feed natural gas costs (Equation 5). The costs for energy are due to the electricity required for compressors and pumps, and are harmonized to 8.73 $/GJ [51]. The maintenance costs per year are set at 4% of total capital costs [161].

Equation 5

$$C_{O\&M\ cost} = C_{energy\ cost} + C_{maintenance\ cost} + C_{feed}$$

The feed natural-gas costs (1.51 – 4.01 $/GJ) are set at 2.97 $/GJ [161]. Production cost excluding feed natural gas is between 0.69 – 4.10 $/GJ. Production cost breakdown results are shown in Figure 21, and the processes are listed in order of low capacity to high capacity. The energy costs of small-scale processes are higher than those of large-scale processes. The feed natural-gas costs represent 42% – 87% of total production cost. The amortized capital costs vary between 0.22 – 3.05 $/GJ, or 6% – 43% of total production cost, while the energy costs are only 0.14 – 1.43 $/GJ, or 2% – 24% of total production cost. It is also clear that the O&M costs (energy costs, maintenance costs, and feed costs) are higher than the amortized capital cost in Figure 21. To conclude, there is still a large uncertainty in the economic performance because industrial production cost data are lacking and the academic literature shows a large variation.

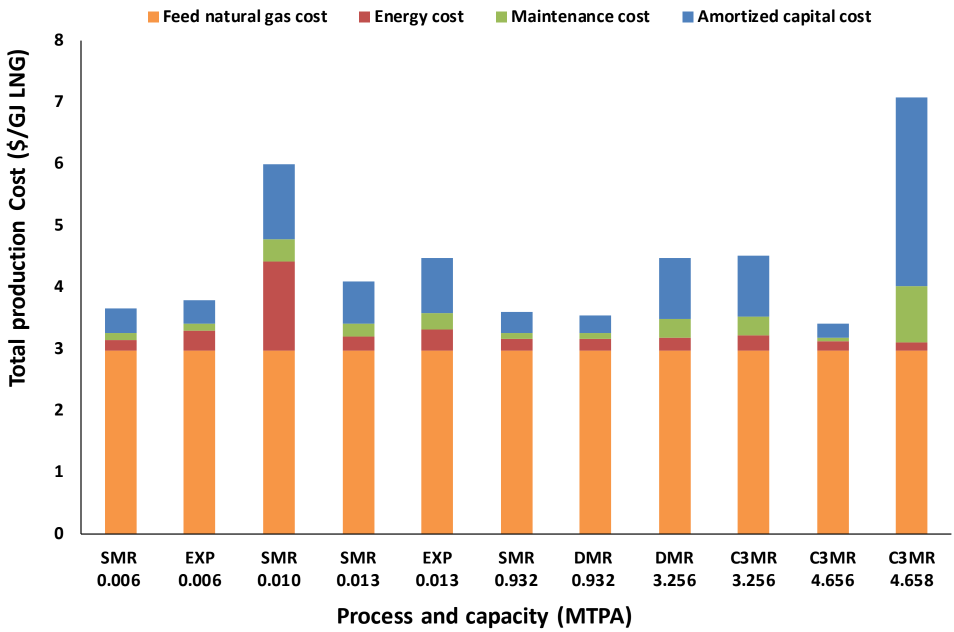

Figure 21. Harmonized total production cost break down for academic literature

Maintenance costs per year are set to be 4% of total capital costs [161].

5. Discussion and future research directions

There are three findings in improvements to LNG processes in academic studies (Section 3), which are discussed below:

- Some of the improvements from academic studies differ between LNG processes. The integration of LNG process, NGL process, N_2 removal process, or power plant applies only to large-scale processes. The potential reason could be that such an integration will add complexity and increase capital costs, which is not suitable for small-scale and offshore processes. The improvements of pure refrigerant (CPOC and EXP) and mixed refrigerant (MFC, C3MR, DMR, and SMR) processes differ in that new refrigerant is used in the former and optimization of the composition of mixed refrigerant is applied in the latter. Adopting new objective functions, such as minimizing total production cost, applies to almost all LNG processes, except CPOC and MFC processes, highlighting a research gap in the economic optimization of CPOC and MFC processes. There are a few studies focusing on the improvement of the operating control system, which appears only in the SMR process. The results of these studies point out the need for dynamic simulation of the LNG process to design a robust control structure, because operating parameters are varying with time.

- The optimization objective of most studies is the minimization of power consumption. However, in several studies [51][52][87][88], using minimization of power consumption as the only objective was found to possibly lead to non-optimal results. The selection of optimization objective should be according to the specific situation of each LNG plant. For example, besides minimization of power consumption, maximization of exergy efficiency and minimization of production cost could be of interest for large-scale plants, minimization of capital costs and simplicity related objective are important for small-scale plants, and safety-related objective and space-related objective are a key for offshore plants.

- Although non-linear programming and genetic algorithm have important drawbacks, they are still the most utilized algorithms for solving optimization problems of LNG processes. The reason may be that the non-linear programming is embedded in Aspen Hysys (energy simulation software) and the genetic algorithm can search globally to avoid getting stranded at one of many local optima. However, the hybrid knowledge-based optimization algorithm can overcome the drawbacks of deterministic and stochastic approach. In addition, there are also several new and efficient metaheuristic algorithms [21] which could be used to solve optimization problems. Researchers should stay open-minded by adopting new optimization algorithms, which could be beneficial to LNG process optimization.

There are several key findings in harmonizing results of technical and economic performance (Section 4), which are discussed as follows:

- Although large-scale processes have much lower primary energy input than that of small-scale processes in technical reports, the optimization work as performed in academic literature could reduce the gap. The primary energy input of small-scale processes reduces significantly in several optimization studies, including, but not limited to, the following efforts. Pressurized LNG process diminishes the need for CO_2 removal and reduces energy input by around 50% [32]; heat integration of heat exchanger reduces almost 50% of the energy consumption [157]; and adding a precooling cycle, e.g., using propane or CO_2, reduces the energy consumption by around 20% [11][105][165]. However, it is not clear whether these optimization efforts are promising from an economic point of view, because of the trade-off between energy efficiency and capital costs [84]. Therefore, it is important to perform the technical optimization with economic analysis, which is absent for most of the previous studies.

- From academic literature, it can be concluded that even identical processes with approximately the same capacity can have a wide range of primary energy input. A potential explanation for this could be the different simulation parameters [21][34]:

  - LNG storage pressure (1 – 10 bar)
  - Liquefaction rate (73% – 100%)
  - Minimum temperature approach (MITA) in heat exchanger (>0 – 5.36 °C)
  - Feed natural-gas temperature (11 – 40 °C), pressure (5 – 90 bar), and composition
  - Compressor and expander efficiency (70% – 90%) and number of stages (1 – 3)
  - Process simulation software and thermodynamic model
  - Ambient temperature

  For example, increasing LNG storage pressure from 1 to 10 bar results in 30% decrease of primary energy input [160]; 10% increase in liquefaction rate results in 10% increase in primary energy input [47]; a hot region with high ambient temperature increases primary energy input by around 25% [34].

- The primary energy input does not show a clear relationship with respect to capacity for either large-scale or small-scale processes. The reason could be that the simulation parameters discussed in previous findings show low correlation with scale; this is especially true for the parameters related to equipment efficiency (compressor, gas turbine, and heat exchanger). Therefore, the authors recommend that the selection of equipment efficiency could be related to capacity [22]. It also highlights the need for investigation on the efficiency with respect to scale for major equipment of the LNG plant.

- The specific capital costs of an LNG plant are much lower in academic literature than in technical reports. A potential reason could be that the definition of capital costs differs between academic literature and technical reports, with the technical reports also including allowance for funds used during construction and the owner's costs. Therefore, the capital costs in academic literature is only a part of that in technical reports, resulting in up to 38% lower capital costs estimates [163]. Another potential reason could be that most of the academic studies consider only the liquefaction system, while the industrial plant in technical reports also includes gas treatment system, storage system, power generation system, cooling water system, etc. The costs of a liquefaction system represent roughly 28% that of a liquefaction train [163]. The combined effect could result in technical reports having 5- to 6-fold higher capital costs than those of academic literature.

- Between technical reports, there is also a large variation in capital costs. This could be caused by specific situations for each plant:

    - Greenfield plant or duplication of a liquefaction train
    - Gas pretreatment system
    - Availability of existing infrastructure
    - Environmental regulation
    - Safety standards
    - Labor costs for installation

Building a greenfield plant could increase capital costs 2- to 3-fold compared to duplicating an existing liquefaction train at the same site [163]. The different impurities in feed gas could also add complexity to the facility, e.g., feed gas with sulfur needs an additional sulfur recovery pretreatment system. The availability of existing infrastructure, such as road, rail, and shipping connections, could significantly affect the capital costs. Strict environmental regulations and safety standards in the recent decade could result in adding additional facilities, which will increase costs. For example, the plants in Gorgon and Snøhvit equipped with carbon capture and storage to reduce carbon emissions will increase the costs [163], and recently built plants are willing to pay more for process safety management systems to ensure public security [166]. Differences in labor costs could be a major reason for high plant capital costs. For example in Australia, the worker's salary is double the global average [163]. Most plants with high specific capital costs (> 1000 $/TPA) were built after 2010 and in Australia.

- The production cost harmonization results show that the energy costs represent 2–24% of the total production cost, while the amortized capital costs and feed natural-gas costs represent 8–43% and 42–87%, respectively. Not only are the feed natural-gas costs the largest contributor to the total production cost, but there are also highly variable, ranging between 1.51–4.01 $/GJ depending on the location. Most of the studies focus only on reducing energy consumption (energy costs). However, it might be a misleading objective for minimizing production cost because the energy costs represent only a small part of production cost, and the decrease in energy costs could increase capital costs. Two studies [87][88] observed that energy-related objectives do not lead to the lowest production cost. Therefore, it is recommended that future studies should also consider capital costs and feed costs besides energy costs in LNG process optimization.

6. Conclusion

From the reviews of LNG processes, it is shown that the CPOC, MFC, C3MR, and DMR processes have low energy consumption and are well optimized for large-scale plants. The SMR and EXP processes are suitable for small-scale and offshore liquefaction plants because of their simplicity, low capital costs, and ease of operation. The improvements from academic studies for each process are different. Process integration applies only to large-scale processes, while configuration adjustment widely applies to small-scale processes. Improvement on operating control system appears only in the SMR process. There is also a lack of studies focusing on economic optimization of CPOC and MFC process.

The optimization objective for most studies is minimization of energy consumption. The other objectives used in the reviewed studies are: maximization of exergy efficiency and minimization of production cost for large-scale; minimization of production cost, simplicity-related and flexibility-related objectives for small-scale; and minimization of production cost, low space requirement, and safety-related objective for offshore processes. This study also highlights the potential improvement of adopting new optimization algorithms to solve complex optimization problems in LNG processes.

The harmonized technical performance provides a quantitative overview of energy consumption from small-scale to large-scale. It shows that large-scale processes (CPOC, MFC, C3MR, and DMR) have lower primary energy input than that of small-scale processes (SMR, SE, and OE). The improvements from academic studies reduce the primary energy input difference between large-scale and small-scale. However, it is not clear if these improvements are also promising in terms of economic performance because of the lack of economic analysis. The primary energy input for an identical process with similar capacity has a wide range and does not necessarily decrease with increasing capacity. The potential reason could be that the key simulation parameters are different and show low correlation with scale. In addition, there is a need for research on the relationship between efficiency and scale for major equipment of the LNG plant.

The harmonized economic performance provides a quantitative overview of specific capital costs and total production cost for LNG processes. The data from the technical reports include only large-scale plant capital costs because of the lack of information for small-scale and O&M costs. The data from academic literature are limited to data from eight studies. Several observations were made based on limited data. The specific capital costs in academic literature are much lower than those in technical reports, and the potential reason could be the different definition of capital costs. An explanation on the large variation of specific capital costs in the large-scale plant could be related to the complexity of the facility and local circumstances: a repetition train or a complete plant, need for gas pretreatment, need for infrastructure, and difference in environmental regulation, safety standard and labor costs. The capital costs and feed natural-gas costs are found as two major contributors that affect the total production cost. It is also indicated that there are only a few studies focusing on economic analysis for the LNG process.

Although there are review papers focusing on the design and optimization of LNG processes, a quantitative overview of the technical and economic performance is missing. This paper filled that gap by harmonizing key indicators of technical and economic performance, including primary energy input, capital costs, and total production cost. The quantitative overview of the technical and economic performance of LNG processes can function as a baseline for future studies for the purpose of comparison.

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