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Abstract

Selection of the base load technology is a very important step on the way of future LNG project development. Therefore, it is necessary to pay attention to technology selection principle. In this paper, we considered literature according to the matter. Modernization of proposed in 2010 decision scale of assessment became the result of the literature review.

In the Thesis, the matrix was created for the project, which has not have final investment decision yet, Baltic LNG project. It will be located in Russia, on the territory of Leningrad region. Filled matrix showed three the most suitable technologies. These technologies are AP-C3MR, Shell DMR and Statoil-Linde MFC.

During simulation procedure, all necessary refrigerants compositions, which can guarantee adequate results of the simulation process, were revealed. Simulation results showed that AP-C3MR is not very suitable technology because it has a high value of relative to DMR specific work (1.08) and situated on the third place in the decision matrix, while DMR and MFC have values of 1.00 and 1.03 and stay on the second and first place respectively.

A detailed comparison of different parameters of other two processes concluded that DMR process has more chances to be implemented as a basis of Baltic LNG project.

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List of Abbreviations

ADJ	Adjuster
AP	Air Products and Chemicals Inc.
BOG	Boil-off gas
C3MR	Propane precooled mixed refrigerant
CAPEX	Capital expenditure
CPOCP	ConocoPhillips optimised cascade
CWHE	Coal-wound heat exchanger
DMR	Dual mixed refrigerant
E	Cooler
GHG	Greenhouse gas
HE	Heat exchanger
HHV	Higher heating value
IGU	International Gas Union
K	Compressor
LNG	Liquefied natural gas
MFC	Mixed fluid cascade
MIX	Mixer
MMSCFD	Million standard cubic feet per day
MR	Mixed refrigerant
MT	Million ton
MTPA	Million tonne per annum
NG	Natural gas
NGL	Natural gas liquids
OPEX	Operational expenditure
PFHE	Plate-fin heat exchanger
PMR	Parallel mixed refrigerant
S	Separator
SWHE	Spiral-wound heat exchanger
TEE	Splitter
TPA	Tonne per annum
TPD	Tonne per day
UGSS	Unified Gas Supply System of Russia
VLV	Valve

Introduction

Liquefied natural gas (LNG) attracts interest all over the world as a clean source of energy, which can successfully substitute other fossil fuels. It has a great impact on the reduction of GHGs problem due to its low carbon dioxide emissions. Recently the world has observed a sustainable growth of LNG demand, which will lead to steady increases. Many of the market players boost LNG production by developing new base load liquefaction capacities to meet growing demand. Such players are Australia, Malaysia, The United States, Cameroon, Indonesia and Russia.

Selection of liquefaction technology is a critical step in any LNG project. Many things depend on chosen variant of liquefaction. First of all, it is the efficiency of the process, the amount of energy consumption for production of certain LNG rates and many others. Different factors can influence the decision, starting from planning production volumes, project cost and possible partners, and ending by area availability and infrastructure. Criteria, which influence the decision, are described in Chapter 2.

A range of parameters, which must be considered during the selection process, may confuse by its variety. Therefore, general methodology should be created. This methodology will combine all factors for taking a “quick solution”, i.e. choosing two-three technologies for more detailed investigation and making the final decision. Many engineering articles describing selection principle do not give a clear selection procedure. It caused by large variations in different technologies and technological parameters. Therefore, in this Thesis, we state the purpose of finding the best methodology with the help of literature survey and developing it in accordance with the up-to-date situation.

A procedure that allows taking decision was described in the paper “Technology selection for liquefied natural gas (LNG) on baseload plants”, in XIX International Gas Convention [1]. The main idea of this article is to develop a suitable decision matrix and rate every process in accordance with it. Hence, for better understanding of the scale of assessment, the Thesis describes parameters included to the matrix and suggests possible improvements.

Based on developed principle, an example of the selection procedure must be provided. This procedure should narrow an appropriate technology search by presenting from 1 to 3 best liquefaction technologies for a special case.

Processes will be chosen with the help of matrix; then it will be described in details for better understanding of their flowsheets, equipment used and conditions during liquefaction to provide eventually accurate simulation results.

Steady state simulations have the aim to select the process with higher liquefaction rate for the same level of compressors work. The main problem appearing during simulation is the lack of available public information, especially about refrigerants composition and temperature/pressure levels. It means that all unknown information required for simulation will be guessed and further proved by simulations. Results of simulations then must be carefully investigated to recognise the most efficient process.

1 Refrigeration Fundamentals

All of the liquefaction technologies use the same principles. Natural gas flows into the main cryogenic heat exchanger. Inside this unit, natural gas is cooled, liquefied and subcooled until appropriate temperature -150°C with the help of refrigerant, which circulates within closed-loop. In base-load facilities cooling, liquefaction and subcooling can be divided into several stages with own heat exchangers. Refrigeration loop consists of a compressor, which increases the pressure of working fluid (1-2); condenser, which removes heat from compressed refrigerant (2-3); and throttle valve or expander responsible for temperature reduction due to pressure drop (3-4). According to Brayton Refrigeration Cycle, refrigerant after condenser should come into main heat exchanger for better heat rejection and more efficient expansion after that. The schematic principle of liquefaction is shown in Figure 1.1[2]. Driving force of compressor is usually gas turbine, the size of which depends on the compressor work. This work depends on the amount of LNG production and composition of natural gas and refrigerant.

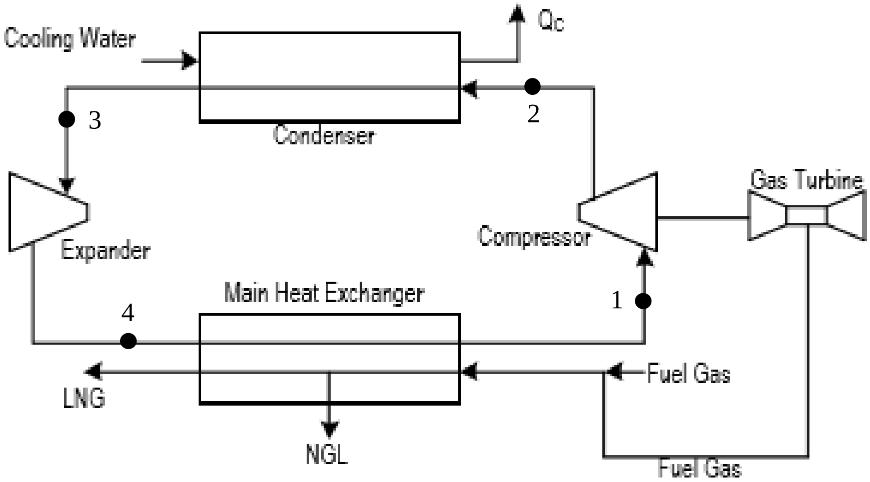


Figure 1.1 Simple refrigeration cycle. [3]

Figure 1.2 below presents Ts-diagram for a random natural gas. From ambient temperature 10°C natural gas is cooled, liquefied and subcooled along constant pressure curve. The upper part of this curve represents precooling stage; the lower part is subcooling, while intersection between diagram and isobar shows liquefaction. From this intersection, we see that liquefaction appears at temperature -50°C and ends at -70°C.

The area Q shows heat removed from the natural gas. The following thermodynamic relation (1) illustrates this process:

$$Tds = dh - vdp \tag{1}$$

If there is no pressure drop, enthalpy change corresponds to the area situated below 60 bar constant pressure curve and represents heat removing from the gas.

The work W is also shown in Figure 1.2. It is the area limited by constant pressure curve and ambient temperature line. Pressure, at which cooling runs, has a great influence on the work and heat. Therefore, there is very important to operate at such high pressure as possible to save the work and reduce heat, which should be removed from the fluid [4, 5].

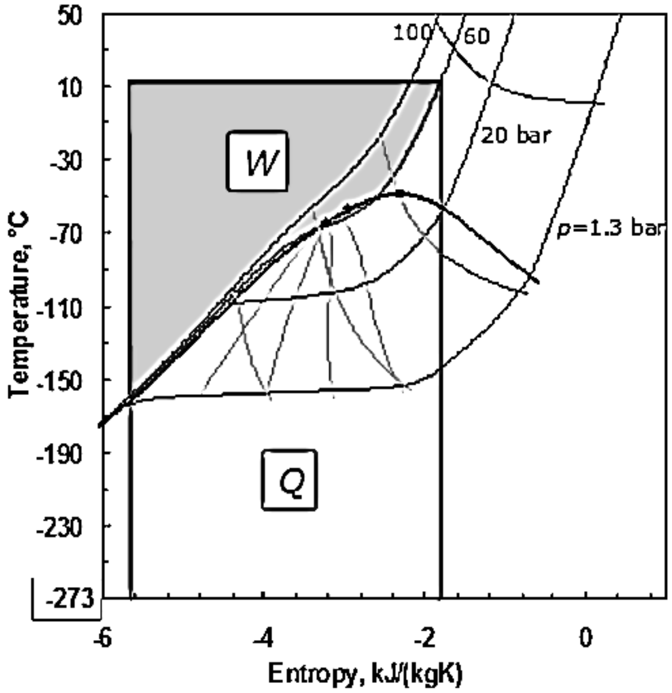


Figure 1.2 - Ts-diagram of random natural gas with ideal work of the process (W) and heat removed during liquefaction and subcooling (Q) [4]

The purpose of refrigerant is to remove heat from the natural gas. In Figure 1.1, flowing directions of natural gas and refrigerant are different because the “hot” and “cold” side of the natural gas should be from the “hot” and “cold” side of cooling gas respectively to provide more efficient heat exchange between fluids.

2 LNG Technology Selection Principle and Criteria Survey

Most of the papers about LNG technology choice principle are based on description processes itself, evaluation of advantages and disadvantages. It is well known, that technologies differ from each other and can be designed for special cases. Selection principle, however, noticed only in some of the articles.

For example, Amos Avidan [6] in 2003 in his paper “Natural gas liquefaction process designers look for larger, more efficient liquefaction plants” described four points which are important selection parameters. To his mind, these parameters are capital and operating costs, emissions, operability and “two-trains-in-one” concept. Also a lot of factors which affect technology choice are described in “Handbook of Liquefied Natural Gas” by S. Mokhatab [7].

The XIX International Gas Convention [1] in 2010 stated the procedure of base load LNG technology selection based on a ranking matrix. We brought up to date this matrix according to current prices stated in International Gas Union World LNG Report 2016 [8]. In addition, we considered strategic relations between countries and companies as an important parameter, which must be included in this matrix.

The ranking matrix must be created for the conditions of certain place and meet purposes of the future project. Thus, Table 2.1 have been compiled for the Leningrad Region (The Russian Federation), environment conditions of which will be described further. The process must stand for high temperature fluctuations within one year. Therefore, the flexibility of gas composition must have higher weight compared to the initially suggested matrix in order to evaluate better processes.

Assessment of appropriate technology focused on parameters, which directly affect the minimization of investment costs and maximising the efficiency of LNG production. The Table 2.1 below shows 21 parameters grouped into 9 primary sections. For simplicity, rating scale has values from 0 to 3, where 3 represents the best value.

After assigning weights to each parameter and sub-parameters, and defining the appropriate rating, the technologies can be ranked according to the resulting score from the weighted sum of the different parameters measured at the decision matrix. This technique is useful for quick viewing of the strengths and weaknesses of each technology while allowing comparisons between the options assessed.

From the total results, the best alternatives for the case study can be obtained. It is noteworthy that the selection technologies should be made based on the particular characteristics of each project or study case raised. As a general rule, it is possible to say that each project has individual priorities, where the selection criteria may change according to the design basis established for each case. Consequently, the weight assigned into the decision matrix can change depending on the case [1].

The decision scale can be filled after detailed consideration of parameters, which has influence on the process development. Such parameters are listed in Table 2.1 below and will be described further.

Table 2.1 Scale of assessment and assignment of weights to the parameters

№	PARAMETERS	WEIGHT	SCALE OF ASSESSMENT		
		(%)	1	2	3
1	Economics	15			
1.1	Investment costs	0.60	More than 1600 US$/TPA	Between 1200 and 1600 US$/TPA	Minor than 1200 US$/TPA
1.2	Operating costs[2010]	0.40	More than 8 US$/TPA	Between 7 and 8 US$/TPA	Minor than 7 US$/TPA
	Standardization	1.00			
2	Constructability	10			
2.1	Expandability plant	0.80	Low	Medium	High
2.2	Area required per train	0.20	More than 70000 m2	Between 60000 and 70000 m2	Minor than 60000 m2
	Standardization	1.00			
3	Maturity	15			
3.1	Years of operation	0.30	Less than 5	Between 5 and 10	More than 10
3.2	Maximum capacity per train set	0.20	Minor than 4 MTPA	Between 4 and 7 MTPA	More than 7 MTPA
3.3	Installed capacity	0.30	Minor than 10 MTPA	Between 10 and 50 MTPA	More than 50 MTPA
3.4	Maximum capacity per train planned	0.20	Minor than 4 MTPA	Between 4 and 8 MTPA	More than 8 MTPA
	Standardization	1.00			
4	Technical	15			
4.1	Cryogenic heat exchanger type	0.35	Only SWHE	Kettle or PFHE, combined with SWHE	Kettle or PFHE, or combinations
4.2	Compressor Type / actuator	0.30	Centrifugal/Frame5	Centrifugal/ Frame 6 or 7	Centrifugal or Axial/ Frame 6 or 7 or electric motor
4.3	Specific work	0.05	More than 14kW/TPD	between 12 and 14 kW/TPD	Minor than 12kW/TPD
4.4	Refrigerant type	0.15	Pure	Pure+mixed	Mixed
4.5	Number of refrigeration cycles	0.05	3	2	1
4.6	Availability of refrigerant	0.10	All require Import	Some require import	Available on site
	Standardization	1.00			
5	CO2 Emissions	5	More than 0,30 MT CO2/MT LNG	Between 0,30 and 0,28 MT CO2/MT LNG	Minor than 0,28 MT CO2/MT LNG
6	Flexibility gas composition	15	Pure	Pure+mixed	Mixed
7	Operability/Maintainability	5	Complex	Medium	Simple
8	Commercial flexibility of the licensor	5	Low	Medium	High
9	Domestic Preferences	15			
9.1	National Content	0.3	All will be imported	Some equipment can be manufactured in the country	All will be manufactured in the country
9.2	Sustainable Development	0.4	Not considered	Considered, but premise without	Included as premise
9.3	Partnership	0.3	No LNG projects	Same company - different projects	Same existing LNG projects
	Standardization	1.00			
	TOTAL	100			

2.1 Train size

There are three main types of onshore LNG plants all over the world:

- Base-load plants.

Such kind of plants is large, with a capacity of above 3 million tonnes per annum (MTPA) of LNG. The main purpose of these facilities to produce LNG for gas transportation from the field or for export LNG to other countries and continents. Many kinds of liquefaction technologies developed for base-load plants, but some of them have not been utilised yet (for example AXENS-Liquefin process [9] or self-refrigerated LNG process [10]).

- Peak-shaving plants.

These plants are smaller, with LNG productivity approximately 0.073 MTPA and vaporisation capacity of about 2.19 MTPA. Plants are connected to the domestic gas network and provide gas at the periods of high gas demand (winter period), while at periods with low demand they liquefy natural gas and storage until required.

- Small-scale plants.

The small-scale plants have a continuous production with a capacity below 0.5 MTPA of LNG. Transportation of liquefied gas to customers made by small LNG carriers, trains and even trucks [4].

Train sizes of baseload plants are combined in Table 2.2. It gives information for matrix parameter "Maturity". The data for this sub-parameter was collected from the IGU 2016 World LNG Report [8] and other articles [4, 9, 11-15]. Based on this table, we can fill following matrix parameters. Maximum capacity per train set (point 3.2) shows maximal existing technology size, while point 3.4 requires setting maximum possible capacities according to producer's statements. Installed capacity (3.3) summarises all existing plants capacities for the exact type of the process.

2.2 Technical risks

This parameter is often one of the most important for making an investment decision. It determines how experienced the process is, i.e. how long it is utilized in industry. Many of competitive technologies are not released now due to high technical risks. Such technologies are AP-DMR (double mixed refrigerant) of Air Products and Chemicals Inc., Liquefin created by AXENS and Shell's Parallel Mixed Refrigerant [16] technologies.

Capacities and years of experience for the most famous base load technologies are presented in Table 2.2.

As far as we can notice from this table, the most experienced processes are AP-C3MR (propane precooled mixed refrigerant) and ConocoPhillips optimised cascade, the previous version of which was used during a long time in the past. The largest existing capacity is among AP-C3MR, however, now the interest to the SplitMR process increases quite fast. It caused of similarity AP-C3MR/SplitMR technology with AP-C3MR. And how noticed by Meshcherin I.V., the main difference between them is in compressors and turbines configuration [17].

Some of the technologies such as Statoil-Linde MFC (mixed fluid cascade) and Shell-DMR are not very experienced and have small existing capacities. It can be explained by the fact that these processes were applied for exact facilities to withstand specific North conditions because other technologies had been firstly developed for locations with high ambient temperature conditions and could be less applicable for severe Arctic climate.

Table 2.2 - Train sizes.

Liquefaction technology	train size, MTPA			Total capacity, MTPA		Years in operation
	Claimed	Existing	Under construction	Existing	Under construction	
ConocoPhillips optimized cascade.	3.0 – 6.0	1.5 - 5.2	3.0 – 4.5	45.1	57.2	18
AP-C3MR	2.0 – 6.0	1.2-4.4	4.0 – 5.5	151.6	28.5	47
AP-C3MR/SplitMR	4.0 – 9.0	3.6-5.0	3.6 – 5.25	43.05	46.55	13
AP-X C3MR	6.5 - 11.0	7.8	-	46.8	-	8
Shell C3MR	3.0 – 6.0	4.3	-	4.3	-	5
Shell-DMR	3.0 - 9.0	4.8	-	9.6	-	8
Statoil-Linde MFC	4.0 - 8.0	4.2	-	4.2	-	10
AP-DMR	2.0 - 7.0	-	-	-	-	-
Liquefin (AXENS)	4.0 - 8.0	-	-	-	-	-
Shell PMR	6.5 - 12.0	-	-	-	-	-

Referring to International Gas Union annual report 2016 [8] the Figure 2.1 below has been created. It presents percentage sharing between existing base-load liquefaction processes. The most popular technology is a Propane-precooled mixed refrigerant process (C3MR), then the second place is shared by C3MR/splitMR, AP-X and ConocoPhillips Optimised Cascade.

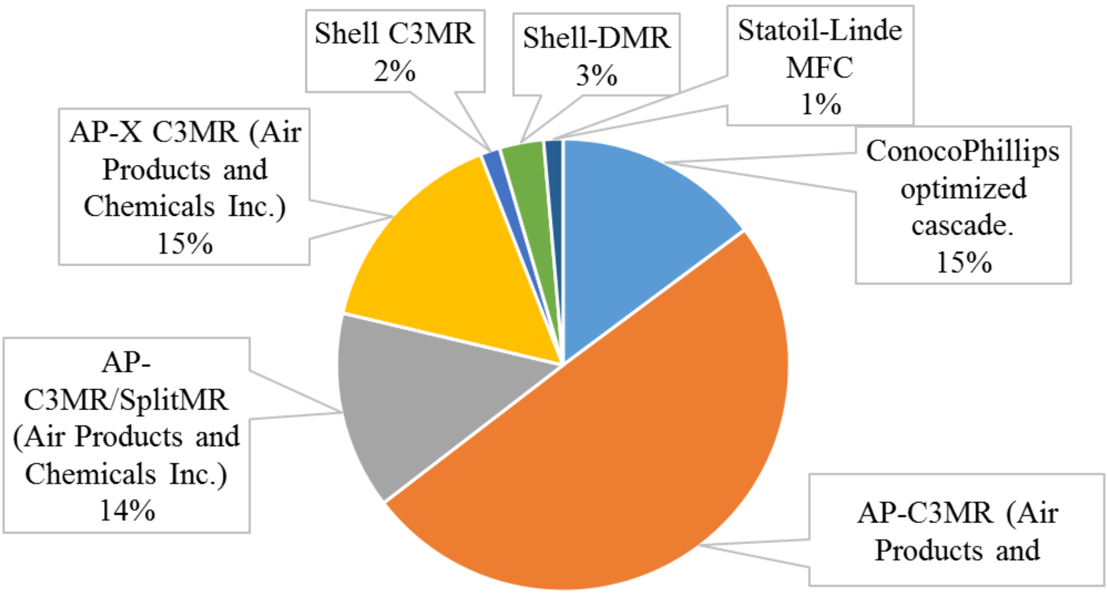


Figure 2.1- Total existing liquefaction base-load capacity by type of process based on IGU annual report 2016 data. [8]

Investors prefer to reduce technical risks by developing new plants with sustainable technologies, which proved themselves during more than 10 years. Such technologies can be less applicable than the others but have enough experience to attract attention.

2.3 Refrigerant selection

The primary objectives of liquefaction technologies innovations are increasing the volume of producing LNG and optimising the efficiency of the refrigeration process employed. The most thermodynamically efficient process is the one, which uses the refrigerant duplicating the shape of the natural gas cooling curve. One or several pure or mixed refrigerants can be used to repeat closely enough such shape [7]. The temperature of evaporating refrigerant must be as high as possible to reduce power need for heat pumping [5]. It can be achieved by utilising several stages of evaporation temperature or by using such refrigerant, which evaporates at gliding temperature.

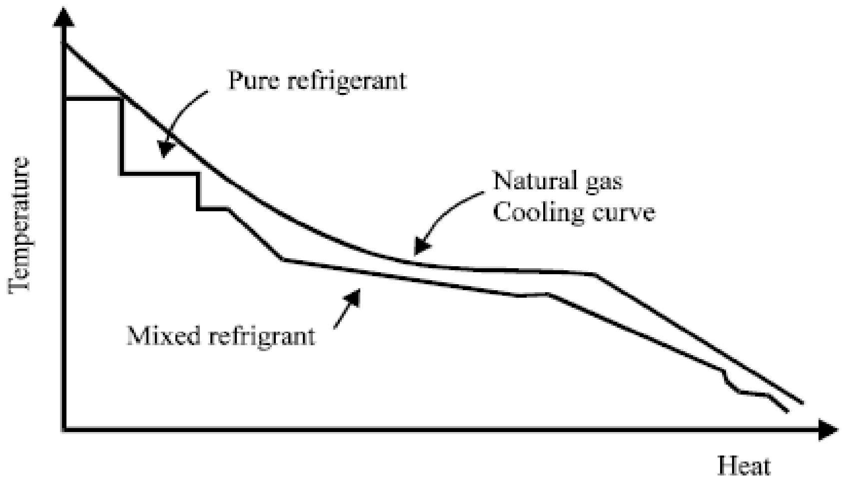


Figure 2.2 – Natural gas and refrigerant cooling curves. [18]

An example of cooling curves for natural gas and corresponding warming curves for the pure and mixed refrigerants is presented in Figure 2.2. The lower straight lines here represent the behaviour of pure refrigerant – propane, which evaporates at the constant temperature, while curved line shows mixed refrigerant heat flow.

As we can see, refrigerants curve is close enough to the natural gas cooling curve, which should provide high efficiency of the process. This is one of reasons, why propane precooled mixed process is so popular all over the world.

Refrigerants for base load liquefaction processes are presented in Table 2.3, which also show main components of these refrigerants.

2.4 Flexibility gas composition

One of the very important parameters is the gas composition flexibility. It describes how easy refrigerants can be improved to meet the cooling curve of changed feed gas. This parameter depends on the type of refrigerant. Pure refrigerants are not flexible, while mixed refrigerants have high flexibility.

Table 2.3 – Refrigerants and their flexibility

Liquefaction technology	Refrigerant	flexibility gas composition
ConocoPhillips optimized cascade.	3 pure: Propane/ethylene/methane	low
AP-C3MR	propane/MR (nitrogen, methane, ethane, propane)	moderate
AP-C3MR/SplitMR	propane/MR(nitrogen, methane, ethane, propane)	moderate
AP-X C3MR	propane/MR(nitrogen, methane, ethane, propane)/nitrogen	moderate
Shell C3MR	propane/MR (nitrogen, methane, ethane, and propane)	moderate
Shell-DMR	DMR: ethane, propane, butane/ nitrogen, methane, ethane, and propane	high
Statoil-Linde MFC	3 MR in cascade: ethane, propane, butane/ methane, ethane, propane/ nitrogen, methane, ethane, propane	high
AP-DMR	DMR: ethane, propane, butane/ nitrogen, methane, ethane, and propane	high
Axens Liquefin	DMR: ethane, propane, butane/ nitrogen, methane, ethane, and propane	high
Shell PMR	PMR: ethane, propane, butane/ nitrogen, methane, ethane, and propane	high

The flexibility of refrigerant composition is important in the case of high variations in natural gas composition entering the plant as well as in the case of location of liquefaction plant in cold climate conditions, characterised by great temperature variations within a year.

For example, Yamal LNG plant (C3MR) was built for the area with temperature difference between winter and summer of approximately 70°C: from -40° to +30° according to local weather forecasting web page [19].

2.5 Train efficiency

Train efficiency is expressed as the ratio of the total higher heating value (HHV) of the liquefied product to the total HHV of the feed gas. It is a common standard used for comparison competing processes for new projects [7, 20]. It is assumed here that liquefied product is not only LNG but also condensate from the internal fractionation; all energy used to run the facility is provided with fuel gas extracted from the plant itself. Vink [20] in his article also made an assumption that all power produced by gas turbines is consumed by compressors but not spent for steam/power generation.

Train efficiency depends on such factors as feed gas composition, inlet pressure and temperature of this gas, temperature and pressure of environment, and many others, for example, location of loading relative to the liquefaction process.

For evaluating thermal efficiency, all energy consumed by liquefaction process must be considered. It also includes a selection of gas turbine drivers, boil-off gas recovery, waste heat recovery, end-flash design, utility and even offsite system.[21].

All required work to produce 1 kg LNG is called specific work. For smaller specific work efficiency of the process is higher. The comparative analysis of technologies efficiency has been developed and described in several articles [20, 22-24]. Obtained results were collected by Wonsub Lim [25] and now mentioned in Table 2.4.

For meeting units in the matrix (Table 2.1) units of specific work in the Table 2.4 were converted from kJ/kg into kW/TPD (2).

$$\frac{kJ}{kg} = \frac{kW}{kg/s} = \frac{1}{86{,}4}\frac{kW}{TPD} \tag{2}$$

Table 2.4 - Efficiency comparisons of LNG processes based on specific work. [25]

Relative specific work (specific work, kJ/kg){specific work, kW/TPD}				
Process	Finn, A.J. [22] relative to cascade	Dam, W et al. [23] relative to DMR	Foerg, W et al. [24] relative to MFC	Vink et al. [20] relative to C3-MR
Cascade	1.00 (1188){13,75}	1.39 (1382){16,00}	1.155	1.156 (1218){14,10}
C3MR	1.15 (1366){15,81}	1.06 (1054){12,20}	1.033	1.000 (1054){12,20}
DMR		1.00 (994){11,50}		1.025 (1080){12,50}
MFC			1.000	

One of the possible reasons for the difference in results is the fact that comparisons were made under different conditions and for various designs. Besides, the use of different levels of optimisation, different equipment and efficiencies in each process could also explain the discrepancies among studies. Moreover, it is recommended to investigate specific work of processes in every special case.

From the Table 2.4 we can conclude the following: cascade process consume relatively much work for liquefaction of 1 tonne per day; other processes specific work various quite large and it is fair to assume equal number among them (score 2). In addition, many technologies are not included in the table above; such technologies will also get score 2. The best of them will be simulated further, and specific work will be clarified.

2.6 Equipment selection

The economics of liquefaction technologies is mainly dependent on equipment selection for the process [7]. The most expensive kinds of equipment are heat exchangers and drivers. This equipment has an influence on capital costs and efficiency. Such important parameters can affect the final decision. Table 2.7 generalises the equipment being utilised and proposed for main base load processes. It contains information about heat exchangers, drivers and compressors types, using in precooling and cryogenic cycles. The table was made based on several articles, and the basic one among them is “Comparison of Baseload Liquefaction Technologies” [20]. Created by K.J.Vink and R. Klein Nagelvoort table was extended and augmented to present equipment for larger number of technologies.

2.6.1 Heat exchangers

There are 3 types of heat exchangers used for liquefaction processes.

- Core-in-kettle type (Figure 2.3) proposed for propane precooling stage. This type of heat exchangers is considered as ideal in the case of pure refrigerants application because it decreases power consumption and has good reliability [1].

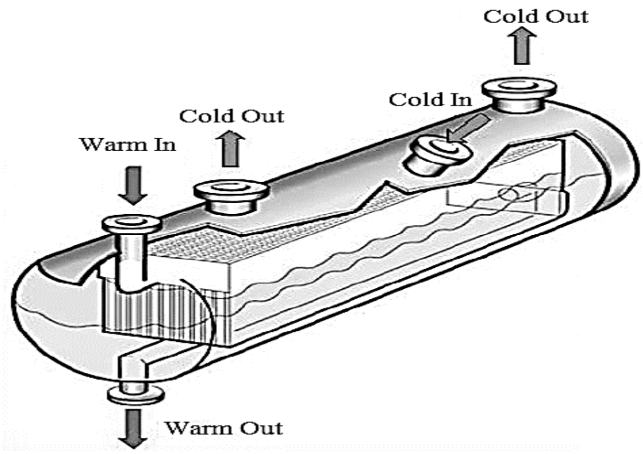


Figure 2.3- Core-in-kettle type heat exchanger [26]

Since core-in-kettle design use single-component refrigerants, it ensures excellent refrigerant flow through units. This also prevents thermal shock during start or stop of feed-gas flow [27]

- Plate-fin heat exchangers (PFHEs) consist of aluminium fins and can be configured by various ways: multi-pass, cross-current and parallel flow directions.

This kind of exchangers usually used as a precooling exchanger of main cryogenic, but in this case, for large capacity trains, PFHEs should be connected to series. The main advantages of PFHEs are relative low equipment weight and its compactness, as well as small foot-prints. Besides, in this case capital costs are low [7]. PFHEs imply the possibility to have competitive vendors. Also during exploitation, these heat exchangers provide low pressure drop and temperature difference [25].

Figure 2.4 below presents an example of the unit, as well as feed and refrigerant flow directions.

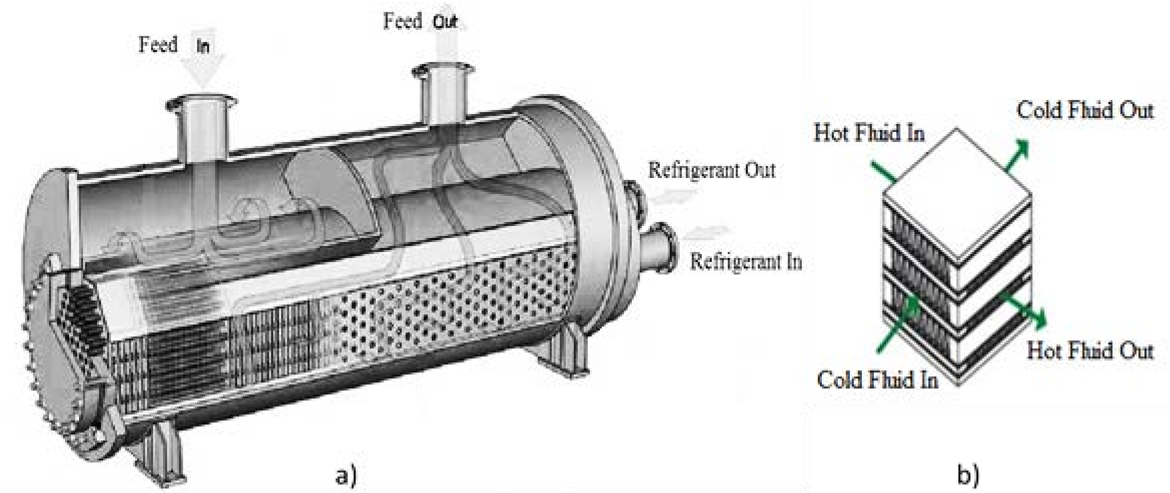


Figure 2.4 - Plate-fin heat exchanger, a) unit [28], b) work principle [29].

Despite many advantages of PFHEs, they have some cons. To avoid problems, they require careful design. Even with good design, it is vulnerable to upsets [25].

- Spiral-wound heat exchangers (SWHEs) or coal-wound heat exchangers (CWHEs) have a great heat transferring area, can operate with high temperature gradient and have tolerance to thermal shocks [7]. Moreover, this type of heat exchangers is robust. Therefore it has high operability. However, since all the SWHEs are proprietary, they are expensive [25].

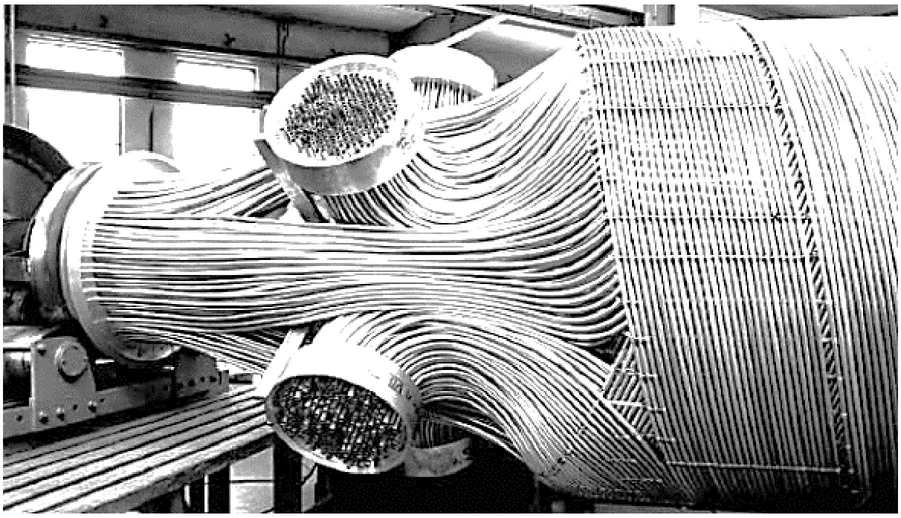


Figure 2.5 - Spiral-wound heat exchanger for multiple flows [24]

SWHEs are commonly utilised qua main cryogenic heat exchangers (MCHEs). This is the most important liquefaction unit which have bigger size, weight and therefore cost as compared to PFHEs [7].

2.6.2 Drivers

Equipment that provides power to the liquefaction system can be steam turbines, gas turbines or electric motors [30]. Steam turbine power plant more complicated in exploitation than gas turbines which are more compact and have shorter delivery and installation time. [22].

Avidan [6] categorised gas turbine types as heavy-duty industrial types and industrial aero-derivatives, which have high efficiency and light weight. Frame type gas turbines widely used in liquefaction plants, almost all compressors are driven by this type of turbines. [7]

There are several key driver selection criteria: driver power capability, reliability and availability experience, capital cost and technical issues. [30]. The total power required by compressors can be produced by several drivers (Figure 2.6), depending on the power of each, as mentioned below (Table 2.5).

Table 2.5 - Driver types and characteristics. [11]

Driver Type	ISO Power	Efficiency %	Relative specific cost
Heavy duty			
GE Frame 5D	32.6 MW	30.3	1.0
GE Frame 6	43.5 MW	33.2	0.82
GE Frame 7	85.4 MW	32.7	0.86
GE Frame 9E	123.4 MW	33.8	0.86
Aero-derivatives			
GE LM2500+	30	40.3	1.09
RR Cobera 6761	33.4	40.5	0.98
GE LM6000	44.6	42.6	1.04
RR Trent	52.55	42.5	0.98

Selection among above units is based on process technology. Some technologies may favour larger heavy duty machines while others can better utilise smaller machines. Both aero-derivatives and heavy duty types may be successfully used in LNG applications. However, the main difference is that heavy duty turbines tend to operate at lower pressure ratios and firing temperatures as indicated in Figure 2.6.

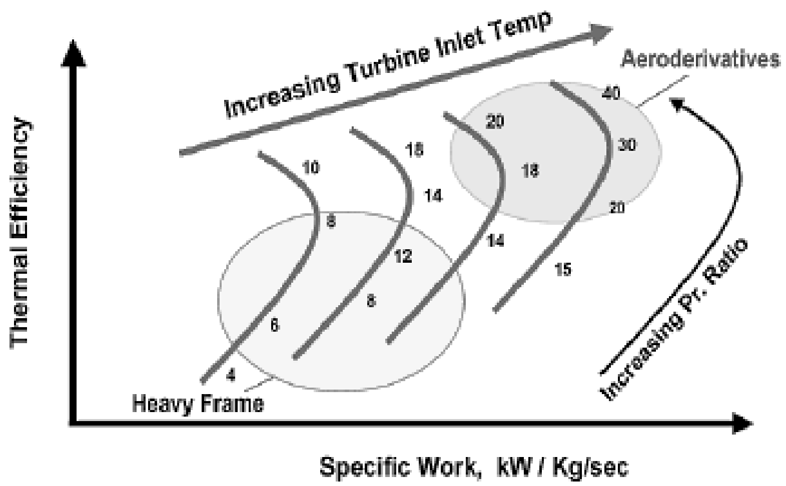


Figure 2.6 - Differentiation between heavy duty and aero-derivative [30]

Assume that for the production of 1 MTPA LNG required power is 35 MW. Then, based on numbers in Table 2.5, the graph of LNG production dependence on drivers can be built (Figure 2.7). Here gas turbines power was derated 80% from ISO for achieving "typical" tropical location [31].

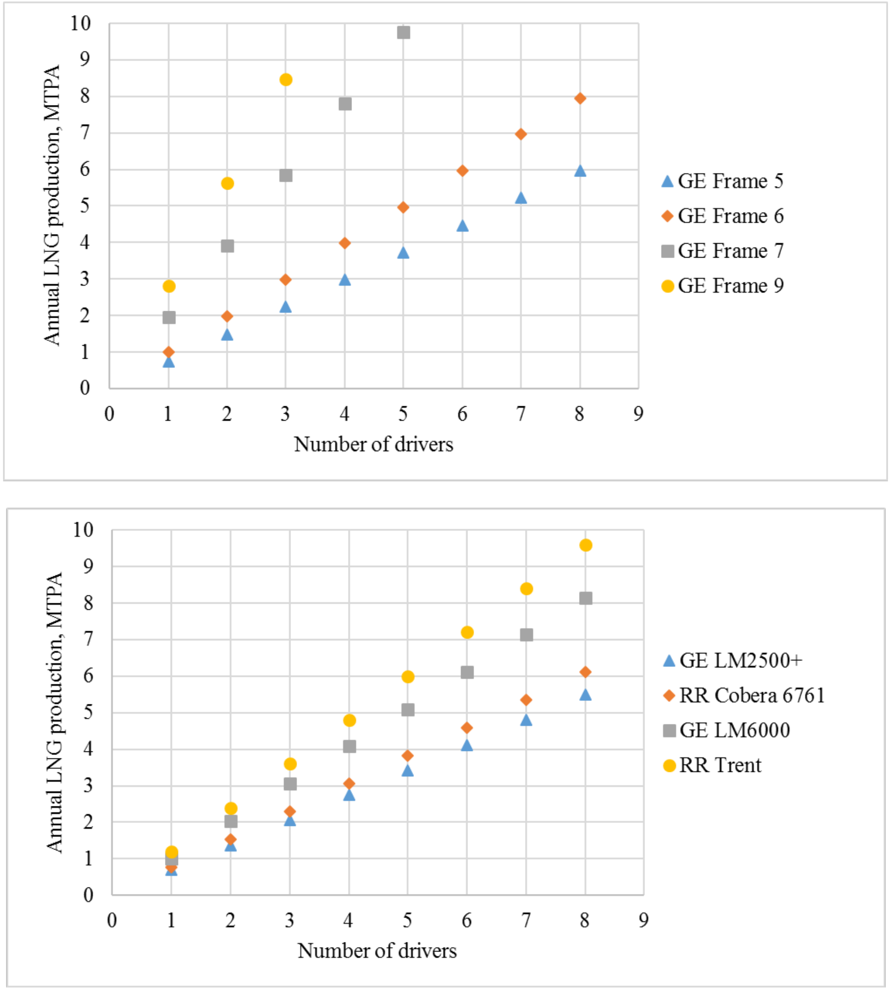


Figure 2.7 – LNG production depending on number of gas turbines [31]

From these plots we can estimate, how many turbines and which types can produce power required by the process. For example, production of 5.0 MTPA of liquefied natural gas can be achieved by different ways: applying of three "Frame 7" gas turbine types; five "Frame 6" or five "LM 6000", where frame-types are heavy duty turbines.

Heavy duty units are more rugged, while aero-derivatives demonstrate good availabilities in severe operating conditions. Maintenance of aero-derivatives is more complex procedure, but heavy duty maintenance requires more time. Aeroderivative engines have high power to weight ratio, which is important in the case of floating LNG facility [30].

In the matrix, all of the technologies, excluding CPOCP got score 3. Because producing of 5MTPA LNG consumes a large amount of power, which is easier to produce with the help of big heavy duty turbines or more efficient aero-derivatives.

2.6.3 Compressors

The typical compressor for large LNG plant is the centrifugal (beam type) they are robust with simple design and low manufacturing cost [25]. In the case of low-pressure mixed refrigerant, process applies to the axial compressor. The main advantage of axial compressors is the high efficiency, large flexibility [32] and high compression ratio [25]. Main characteristics and advantages centrifugal and axial compressors are presented in Table 2.6.

Table 2.6 - Axial and centrifugal compressors characteristics [32].

Parameters	Axial Compressor	Centrifugal compressor
Power required	16 – 28 MW	16 – 44 MW
Density/pressure	up to 5 kg/m3 inlet density	up to 60+ bar discharge pressure
Flow	up to 300000 m3/h inlet volumetric flow	up to 500000 m3/h (double flow)
Efficiency	90%	88%
Speed	fixed speed, VSV	IGV and speed variation
Advantages	flexibility for operation and start up, reliability	High reliability

Basically, centrifugal compressors are used for precooling, while liquefaction stage utilises both centrifugal and axial in tandem. Compressors also have some cons, such as low compression ratio and restriction to use at high flow rates for centrifugal compressors and the high price of axial with the possibility to utilise them only with large flow rates [25].

If the process uses axial compressors, driving by Frame 6, Frame 7 or aero-derivative type of turbines, it gets score 3 in the ranking matrix.

Table 2.7 generalises equipment decisions for base load liquefaction processes. For processes that are in utilisation, the most commonly used equipment was chosen. However, for technologies which have not found implementation in the industry yet, equipment stated like proposed by creators. Units mentioned in Table 2.7 can be replaced by other ones according to site preferences.

Table 2.7 – Equipment decisions/proposals.

Liquefaction technology	precooling			liquefaction		
	heat exchanger	driver	compressor	MCHE	driver	compressor
CPOCP	PFHE/core-in-kettle	2 GE-5C (+helper)	2 of 3-stage centrifugal	PFHE	4 GE-5C	Ethylene: 2 of 3-stage centrifugal. Methane: 2 of 4-stage centrifugal, 3 casings
AP-C3MR	core-in-kettle	GE-7EA (+generator)	4 stage centrifugal	SWHE/ CWHE	GE-7EA (+helper)	Axial + 2 stage centrifugal in tandem
AP-C3MR/SplitMR	core-in-kettle	GE-7EA (+generator)	4 stage centrifugal	SWHE/ CWHE	GE-7EA (+helper)	Axial + 2 stage centrifugal in tandem
AP-X C3MR	core-in-kettle	Frame 9	4 stage centrifugal	CWHEs, Nitrogen subcooling (PFHE)	2 Frame 9 (liquefaction and subcooling)	Axial + centrifugal in tandem
Shell-DMR	CWHE (Linder or APCI) or SWHE	GE-7EA (+helper)	3 stage centrifugal	SWHE/ CWHE	GE-7EA (+helper)	Axial + 2 stage centrifugal in tandem
Statoil-Linde MFC	PFHE	Snøvit: 5 of LM6000 in total	3centrifugal in total	CWHE	Snøvit: 5 of LM6000 in total	3 centrifugal in total
AP-DMR	CWHE (Linder or APCI)	GE-7EA (+helper)	3 stage centrifugal	CWHE (Linder or APCI)	GE-7EA (+helper)	Axial + 2 stage centrifugal in tandem
Liquefin	PFHE	3 Frame 7 - same set of drivers across different cycles	3 stage centrifugal	PFHE	3 Frame 7 - same set of drivers across different cycles	Axial + centrifugal in tandem
Shell PMR	CWHE/SWHE	3 GE-9E or 4 GE-7EA in total	3 stage centrifugal	CWHE/ SWHE	3 GE-9E or 4 GE-7EA in total	Axial + 2 stage centrifugal in tandem

2.7 CO2 emissions

Liquefaction of natural gas is a very energy intensive process. That is why liquefaction facilities are expected to be large producers of greenhouse gases (GHGs). Over 80% of GHGs emissions are released from cooling and electricity generation processes [33]. The following factors have an impact on emissions amount.

- Feed gas composition – if gas entering the liquefaction facility has CO2 in its composition it increases the risk of CO2 emissions. However usually exporting plants use pipeline-quality gas without CO2 content.

- Ambient temperature – low temperatures decrease emissions amount because under low temperatures turbines and compressors work more efficiently.

- Power resource – higher intensity of electricity utilising consumes much power production. Heavy duty gas turbines produce more emissions than aero-derivative types of drivers (Figure 2.8). Processes also can decrease emissions developing renewable energy utilisation.

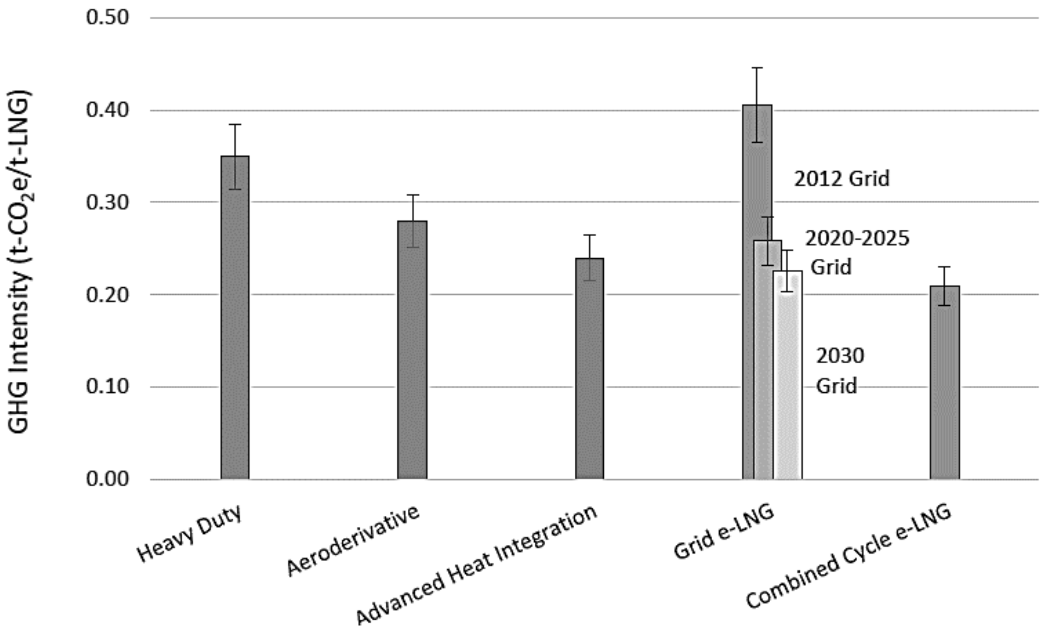


Figure 2.8- Greenhouse gases intensity, depending on kind of energy provider [33].

- Liquefaction process itself has a relatively low impact on CO2 emissions. It can influence only on the choice of equipment, especially drivers, as described above.

- Heat integration – new processes are already designed with the possibility to use waste heat for acid gas removal or gas dehydration, more advanced conceptions can use this heat for electricity production.

From the Figure 2.8, we clearly see advantages of aero-derivative gas turbines in comparison with heavy duty. According to the Table 2.7, aero-derivative type is used only in MFC, designed for Hammerfest area. That is why MFC get score 3 in the matrix, while other get score 1 (heavy duty). This estimation is true only if not a single process utilises cycles with advanced heat integration or combine cycles. Therefore, the score can be changed in dependence with changing equipment. For the comparative review it is assumed that all drivers are heavy duty turbines except MFC case with aero-derivatives.

2.8 Climate conditions

Generally, cold ambient temperatures increase the operating efficiencies and reduce energy consumption in the cryogenic facilities, independent on liquefaction technology. Temperature fluctuation is the most critical for liquefaction, especially when the cooling medium is the air. Due to temperature variations LNG production changes within a year [17].

The most popular liquefaction process, propane precooling technology, could be less suitable for locations with temperature fluctuations due to low degree of precooler flexibility.

In such cases, mixed refrigerant technologies could become a better choice. Here multicomponent refrigerant has a lower boiling point and can be regulated for changing environment temperature.

Parameter "climate conditions" is not in the matrix (Table 2.1). However, it adds weight to parameter "flexibility gas composition", making it very powerful in the process of making decision .

2.9 Constructability

The parameter which is called "area required per train" can be calculated for the desired amount of LNG production by Equation 3 below [34].

$$A(hectares) = 0.131F^{0,6} - 0.8 \qquad (3)$$

Where A – area per train,

F – throughput in MMSCFD.

LNG production is given in MTPA and 1000 MMSCFD = 6.972 MTPA [35]. Besides area is required in square meters and 1 hectare = 10000 m^2. According to this, Equation 3 becomes Equation 4.

$$A(m^2) = (0.131(Q * 1000/6.972)^{0,6} - 0.8) * 10000 \qquad (4)$$

Where Q – throughput in MTPA.

The above process area equation applies to a basic process area that includes LNG unloading, recondenser, boil-off gas (BOG) compressor, send-out pump, LNG vaporisation, odorizing natural gas, gas metering, electrical substation, and/or the control room and access roads.

Besides, during evaluation, the possible size of the train also must be considered. Therefore, such technologies as AP-X and PMR with the size of 7.5 MTPA per train will have a greater area than other with size 5 MTPA per train. Trains with 5MTPA of production will have size 59708m^2. While trains with 7.5 MTPA of production reserve 78356m^2. However, in the case of 15 MTPA plant two trains AP-X or PMR will save place – more than 20000 m^2.

These rough estimations allow setting score 3 to technologies with 5 MTPA of production.

2.10 Capital investments and operating costs

For the purpose of comparison costs of different liquefaction technologies liquefaction plant metric cost (5) should be evaluated.

$$Metric\ cost, \frac{\$}{TPA} = \frac{Cost\ of\ the\ plant, million\ US\$}{capacity, MTPA} \tag{5}$$

Figure 2.9 presents the trend of metric cost. Year points indicate project start-up dates. Liquefaction facility requires approximately 10 years for developing, including 4 years of construction.

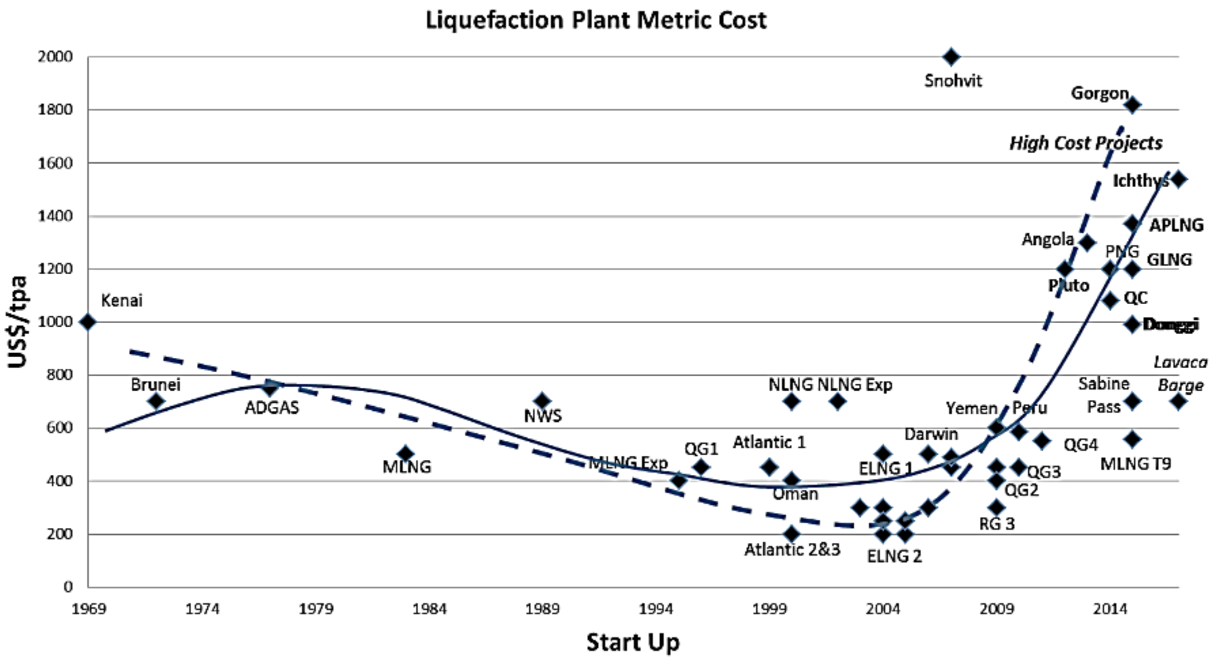


Figure 2.9 - Historical trend of liquefaction CAPEX in 2008 US$ [36]

Firstly, for the early plants, the cost was about 700$/TPA, then prices dropped to the level of 400$/TPA to the 2000 year [36]. Then, LNG projects have faced essential cost rise since the year 2000. Unit costs [8] for LNG facilities escalated from an average of \$397/tonne in the 2000-2007 period to \$807/tonne for the period from 2008 to 2015.

Besides, it is important to say that greenfield projects require more investments in comparison with brownfield and can achieve prices of \$1162/tonne for Atlantic-Medditerrian territories till period 2016-2021 [8]. Figure 2.10 presents the average liquefaction unit cost by basin and project type.

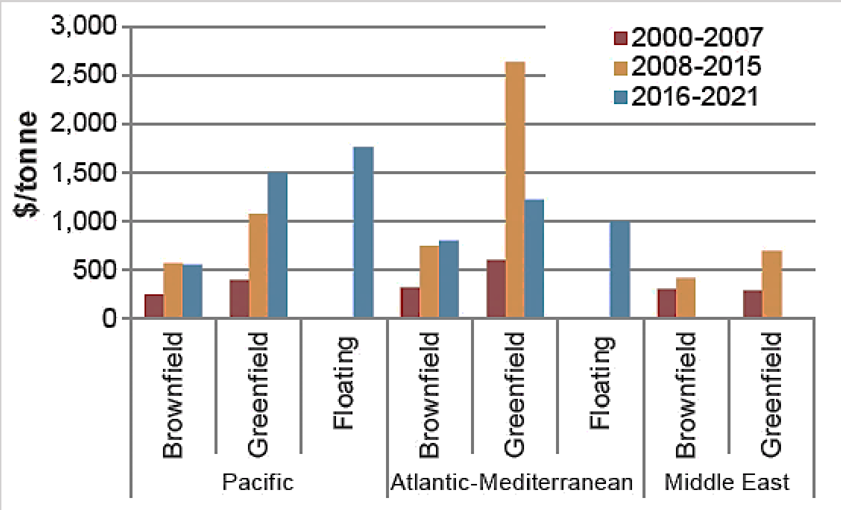


Figure 2.10 - Average liquefaction unit cost by basin and project type. [8]

Due to the lack of public information about operational costs of LNG production, in the ranking matrix the costs are mentioned for the year 2010 (taken from materials of XIX International Gas Convention [1]).

Confident information does not allow making any conclusion about CAPEX and OPEX of technologies realisation. Therefore, we set score 0 to all cells, excluding these parameters from research.

2.11 Operability/Maintainability

Operability parameter refers to design complexity of liquefaction technologies. Most of the baseload processes are complex. According to P.Y. Martin conference paper [9], Liquefin technology is considered as less complex in comparison with others. This note allows setting score 2 to the Liquefin cell.

All other technologies got score 1. However, it is also important to notice that ConocoPhillips optimised cascade is more complex than other technologies and propane precooled mixed refrigerant is easier in operation. Nevertheless, taking into account all of these, decision to set score 1 seems more reliable.

2.12 Commercial flexibility of the patentee

The majority of the technologies belongs to the concrete holder of a patent, which has own manufactory and interested in pushing the client placing the order for equipment with him. Such equipment will cost more expensive than units produced by other fabricators. However, there is impossible to refuse from such brand, because of patent restrictions. If the patent owner is flexible enough, it could reduce CAPEX significantly.

From the literature, we can conclude that ConocoPhillips Optimized Cascade and Liquefin technologies have the most flexible patentee (score 3), while AP-DMR proprietor is not flexible (score 1). Other processes allow ordering only part of the equipment from patent holder company, while another part can be purchased from somewhere else.

2.13 Domestic preferences

Domestic preferences is one of the criteria which gives the highest weight in scale of assessment. It includes such parameters as national content, sustainable development, and partnership.

National content is the parameter, which determines the possibility of manufacturing equipment and materials for the process in the country. In case all equipment can be manufactured in the country, evaluating process will get score 3, while in the case when everything is going to be imported, technology gets score 1.

Supposing that all technologies will require some import of equipment such as heat exchangers, score 2 seems correct value for chosen technologies.

The term sustainable development means the ability of technology to satisfy needs of the present generation without compromising the ability of future generations to meet their own needs [37]. Often sustainable development in LNG industry is connected with alternative energy sources utilisation as an environmentally friendly source, which satisfies energy needs and limits pollutions. For the project, we consider that sustainable development is possible and can be applied in the future. Therefore all technologies get score 2 for this parameter.

The partnership is one of the key parameters for LNG project development. Most of the companies prefer to continue sustainable cooperation with partners from other projects, especially if planning projects will be similar with existing. This fact simplifies many things from facility designing and financing to choosing of equipment manufacturers. That is why partnership has a huge weight in assessment scale. In Russia, at the present time, two liquefaction facilities work successfully.

First, it is Sakhalin-2 LNG, which was built with Shell-DMR technology as a basis for the plant to ensure maximum LNG production during severe winters. The facility has two parallel process trains with total capacity 9.6 million tonnes of LNG per year [38].

The second facility is Yamal LNG, which is located above the Arctic Circle. It started liquefaction in 2017 and utilises AP-C3MR. The facility consists of 3 liquefaction trains. Each plant's train is able to produce 5.5 MTPA [39].

Based on existing partnering relations, propane precooled mixed refrigerant technology and dual mixed refrigerant process got score 3 in the scale of decision. Other technologies developed by Air Products & Chemicals and Shell got score 2.

3 Decision Matrix Results

Technology selection is processed for the project, which has not have final investment decision yet. According to news from official Gazprom site [40] and Oil&Gas Journal [41], Gazprom stated that Baltic LNG plant with a capacity of 10 million tonne per annum would be built near Ust-Luga seaport, Leningrad Region, Russia. The project is primarily targeted to supply LNG to the European and Latin American markets. The plant will be able to increase production to 15MTPA in the future, according to LNG world news portal [42].

Such types of facilities are base load plants, and their sizes are mentioned in Table 2.2. The final result 15 MTPA can be achieved by building 3 trains of 5 MTPA each or 2 trains of 7.5 MTPA each. Then, baseload technologies are divided into two groups (Table 3.1).

Table 3.1 – Selected technologies which meet the size of future plant

Group 1 - 3 trains, 5MTPA each	Size, MTPA
ConocoPhillips optimized cascade	3.0 - 5.2
AP-C3MR	3.0 - 5.5
AP-C3MR/SplitMR	3.6 - 5.5
AP-DMR	3.0 - 7.0
Shell-DMR	3.0 - 7.0
Liquefin (AXENS)	4.0 - 8.0
MFC	4.0 - 10.0
Group 2 - 2 trains, 7.5MTPA each	Size. MTPA
Shell PMR	6.5 - 12.0
AP-X	6.5 – 11.0
Liquefin (AXENS)	4.0 - 8.0
MFC	4.0 - 10.0

However, Group 2 does not meet the initial conditions that plant must have interim 10 MTPA stage before expansion. Therefore, in the further work, we will pay attention only to technologies from group 1. It means that Shell PMR and AP-X due to restriction about size do not have the opportunity to continue selection procedure. However, they have great potential for similar projects but with an intermediate stage of 7.5 MTPA or without any.

When the size of the plant agreed and pretended technologies have been selected, the decision matrix should be filled. With the help of this ranking matrix, quick estimation can be done, and technologies with better score will be compared further.

The matrix form is presented in Table 2.1. Obtained results are in

Table 3.2 below. It must be noticed, that Leningrad Region has quite high temperature difference during one year. That is why flexibility of gas composition has weight of 15% in decision matrix.

The matrix shows that the most applicable technologies for Ust-Luga conditions are AP-C3MR, Shell DMR and Mixed Fluid Cascade. All of them have total grade higher than 180.

Table 3.2 – Decision matrix

№	PARAMETERS	WEIGHT	CPOCP		AP-C3MR		AP-C3MR/split		AP-DMR		Shell DMR		Liquefin		MFC	
		(%)	score	total	score	total	score	total	score	total	score	total	score	total	score	total
1	Economics	15		0.0		0.0		0.0		0.0		0.0		0.0		0.0
1.1	Investment costs	0.60		0.0		0.0		0.0		0.0		0.0		0.0		0.0
1.2	Operating costs	0.40		0.0		0.0		0.0		0.0		0.0		0.0		0.0
	Standardization	1.00														
2	Constructability	10		30.0		30.0		30.0		30.0		30.0		30.0		30.0
2.1	Expandability plant	0.80	3	24.0	3	24.0	3	24.0	3	24.0	3	24.0	3	24.0	3	24.0
2.2	Area required per train	0.20	3	6.0	3	6.0	3	6.0	3	6.0	3	6.0	3	6.0	3	6.0
	Standardization	1.00														
3	Maturity	15		34.5		39.0		37.5		6.0		28.5		6.0		25.5
3.1	Years of operation	0.30	3	13.5	3	13.5	3	13.5	0	0.0	2	9.0	0	0.0	2	9.0
3.2	Maximum capacity per train set	0.20	2	6.0	2	6.0	2	6.0	0	0.0	2	6.0	0	0.0	2	6.0
3.3	Installed capacity	0.30	2	9.0	3	13.5	2	9.0	0	0.0	1	4.5	0	0.0	1	4.5
3.4	Maximum capacity per train planned	0.20	2	6.0	2	6.0	3	9.0	2	6.0	3	9.0	2	6.0	2	6.0
	Standardization	1.00														
4	Technical	15		28.5		36.0		36.0		33.0		33.8		43.5		37.5
4.1	Cryogenic heat exchanger type	0.30	3	15.8	2	10.5	2	10.5	1	5.3	1	4.5	3	15.8	2	9.0
4.2	Compressor Type / actuator	0.30	1	4.5	3	13.5	3	13.5	3	13.5	3	13.5	3	13.5	3	13.5
4.3	Specific work	0.10	1	1.5	2	1.5	2	1.5	2	1.5	2	3.0	2	1.5	2	3.0
4.4	Refrigerant type	0.15	1	2.3	2	4.5	2	4.5	3	6.8	3	6.8	3	6.8	3	6.8
4.5	Number of refrigeration cycles	0.05	1	0.8	2	1.5	2	1.5	2	1.5	2	1.5	2	1.5	1	0.8
4.6	Availability of refrigerant	0.10	3	4.5	3	4.5	3	4.5	3	4.5	3	4.5	3	4.5	3	4.5
	Standardization	1.0														
5	CO2 Emissions	5	1	5.0	1	5.0	1	5.0	1	5.0	1	5.0	1	5.0	3	15.0
6	Flexibility gas composition	15	1	15.0	2	30.0	2	30.0	3	45.0	3	45.0	3	45.0	3	45.0
7	Operability/Maintainability	5	1	5.0	1	5.0	1	5.0	1	5.0	1	5.0	2	10.0	1	5.0
8	Commercial flexibility of the licensor	5	3	15.0	2	10.0	2	10.0	1	5.0	2	10.0	3	15.0	2	10.0
9	Domestic Preferences	15		19.5		28.5		24.0		24.0		28.5		19.5		19.5
9.1	National Content	0.30	2	9.0	2	9.0	2	9.0	2	9.0	2	9.0	2	9.0	2	9.0
9.2	Sustainable Development	0.40	1	6.0	1	6.0	1	6.0	1	6.0	1	6.0	1	6.0	1	6.0
9.3	Partnership	0.30	1	4.5	3	13.5	2	9.0	2	9.0	3	13.5	1	4.5	1	4.5
	Standardization	1.00														
	TOTAL	100.0		152.5		183.5		177.5		153.0		185.0		174.0		187.5

4 Appropriate Technologies Simulation in UniSim® Design

UniSim® Design is the simulation solution for engineers. It helps to create and optimise plant designs and monitor asset performance, thus enabling stable operations and plant safety. With comprehensive, first principles thermodynamics and unit operation models UniSim Design combines steady state and dynamic modelling to boost productivity and profitability across the project lifecycle [43].

The program also able to assist with LNG plant simulation. In reliance on natural gas composition, its pressure, flow rate and ambient conditions, it is possible to create liquefaction process and optimise it.

4.1 Initial conditions

Temperature conditions for Ust-Luga seaport presented in Table 4.1 and **Feil! Fant ikke referansekilden.**. Data in Table 4.1 is taken from the web pages, containing weather statistics [44, 45].

Table 4.1 – Temperatures, Ust-Luga seaport

Months	Temperature		
	Coldest average, °C	Average, °C	Warmest average, °C
January	-7	-4.5	-2
February	-8	-5	-2
March	-4	-1	2
April	1	5	9
May	7	11.5	16
June	11	15.5	20
July	15	19	23
August	14	17.5	21
September	9	12.5	16
October	4	6.5	9
November	-1	1	3
December	-5	-3	-1

Figure 4.1 shows us that temperature fluctuation can be high and difference within average temperature line is 24°C, while the difference between minimal winter and maximal summer temperature can be more than 60°C. Such meteorological observation proves the necessity of using the process, which can be flexible enough to achieve the best results.

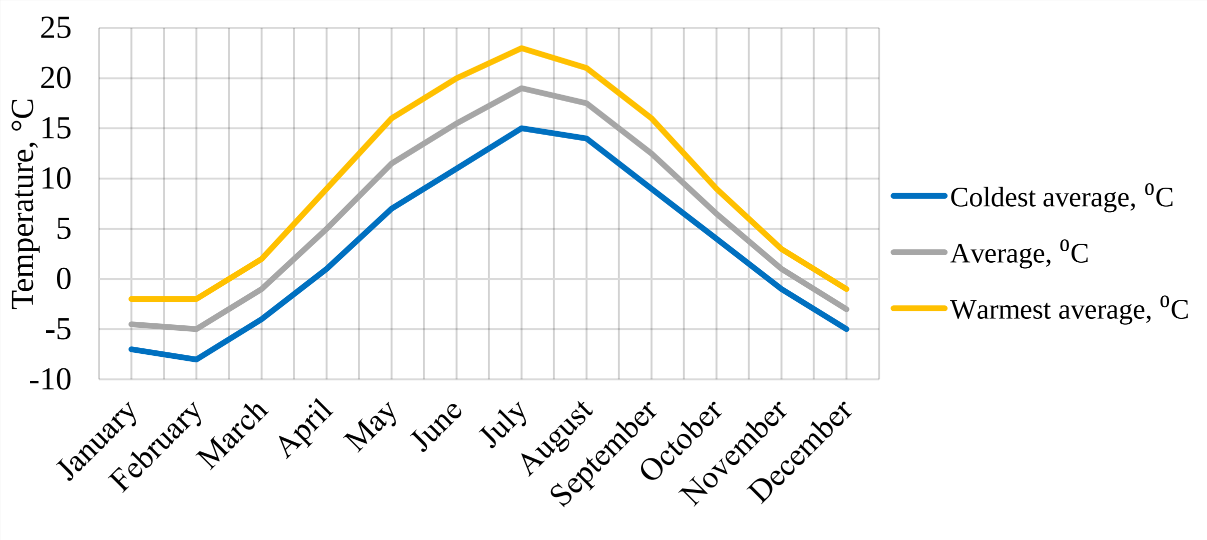


Figure 4.1– Average temperature diagram

The feed gas will enter the liquefaction plant from the Unified Gas Supply System of Russia (UGSS) [41]. The composition of this gas is approximately constant. However, it can contain such amount of heavy components, which could require NGL extraction units.

In the research, it is assumed that natural gas is prepared to the liquefaction and has the same composition liquefied natural gas composition of Sakhalin LNG plant. This assumption is necessary for simplification of simulation environment and can be applied in the real facility. For example, MFC has NGL extraction unit before the main liquefaction cycle. Avoiding fractionation step during liquefaction can increase the efficiency of the system. The Gas Composition Transition Agency Report 2013 [46] presented the values of Sakhalin natural gas composition (Table 4.2).

Table 4.2 – Natural gas composition

Parameter	Value	Units
Methane, CH4	92.54	%
Ethane, C2H6	4.47	%
Propane, C3H8	1.97	%
butane+, C4+	0.95	%
Nitrogen, N2	0.07	%
Carbon dioxide, CO2	0.00	%
Wobbe index	55.40	MJ/Nm3
Calorific value	43.30	MJ/Nm3
Relative density	0.61	-

Figure 4.2 shows a pressure-enthalpy diagram for the gas entering liquefaction plant (Table 4.2). The green line in this figure represent ambient temperature, and the blue line is isotherm, representing liquefied gas. In the case of 50 bars internal pressure, gas will start liquefying at the temperature -35°C (pink line) and end liquefaction at temperature -74°C (yellow line).

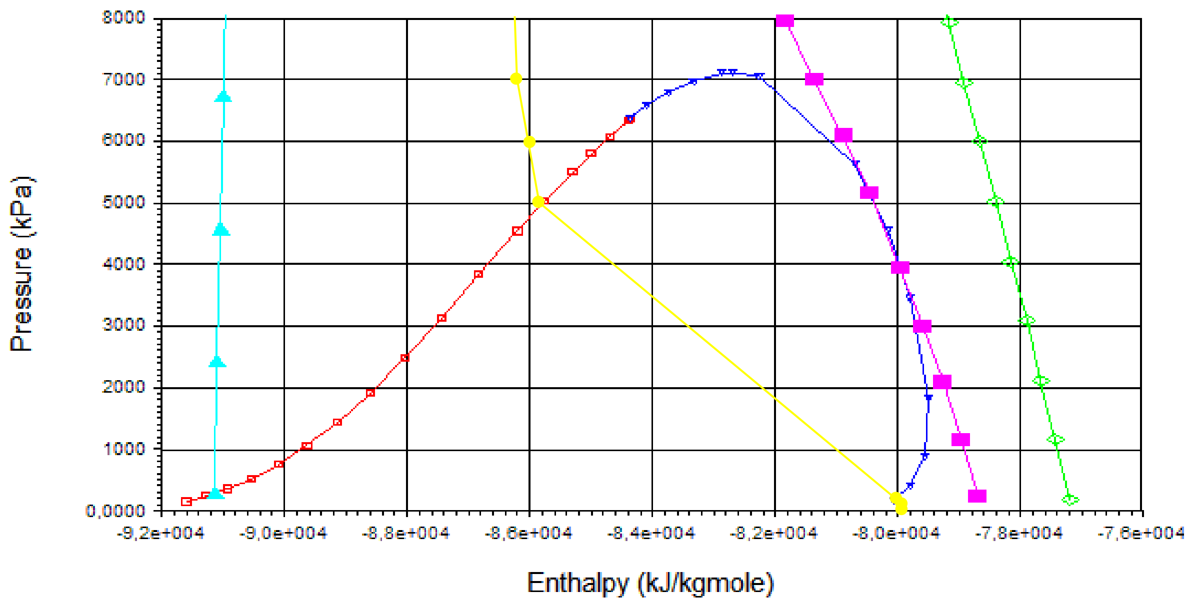


Figure 4.2 – Natural gas pressure-enthalpy diagram.

4.2 Processes flowsheets and descriptions

4.2.1 C3MR – Propane-precooling Mixed Refrigerant process

AP-C3MR (Figure 4.3) is the most popular process all over the world. It was created by Air Products and Chemicals Inc. to combine the best attributes of the cascade process with the MR process. There are two main stages. They are a precooling stage and liquefaction one. The precooling cycle uses pure component – propane. Chilling processed until natural gas achieves temperature “-40°C” in core-in-kettle type heat exchangers [47]. After heat exchangers, propane vapour flows through the four-stage centrifugal compressor, which is driving by GE Frame 7EA type turbine. According to Table 2.7 such type of turbine gives the power 85.4 MW.

After propane stage, natural gas cooling continues in a spiral- or coal-wound heat exchangers in which the cold duty is provided by the mixed refrigerant. This refrigerant consists of nitrogen, methane, ethane and propane. The tandem of axial and 2-stage centrifugal compressors operated by second GE Frame 7 turbine with a helper. After this stage, natural gas has a temperature -150°C.

Joule-Thompson valve creates the final subcooling by reducing pressure from 50 bars to the pressure, slightly higher than atmospheric.

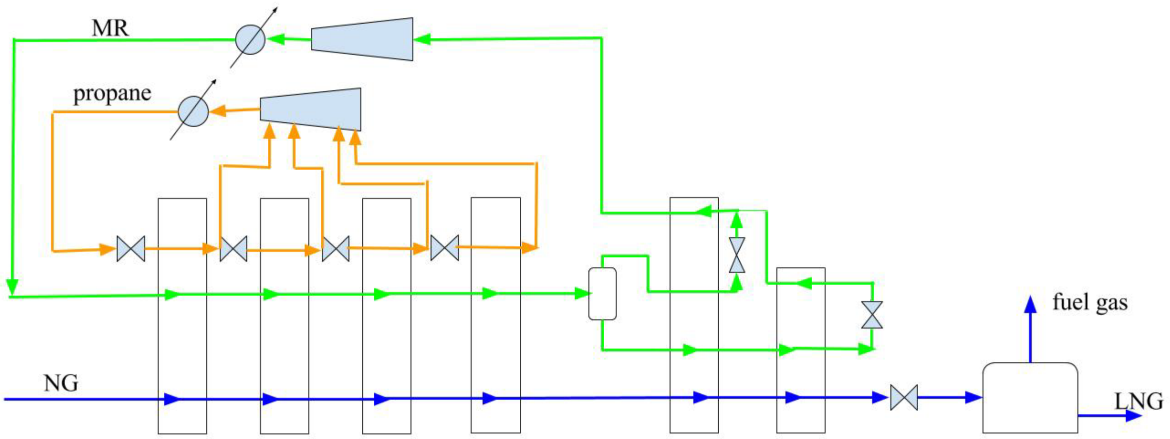


Figure 4.3– Propane precooled mixed refrigerant process

Traditional propane circle layout (Figure 4.4) include 4 stages of compression from LP to HHP. It joins all propane streams in sequence. J.J.B. Pek in his paper “Large capacity LNG plant development” [48] describes better compressors combination as shown in Figure 4.4 (SplitPropane Line-up) to increase possible natural gas flow rates for the same compressors power consumption.

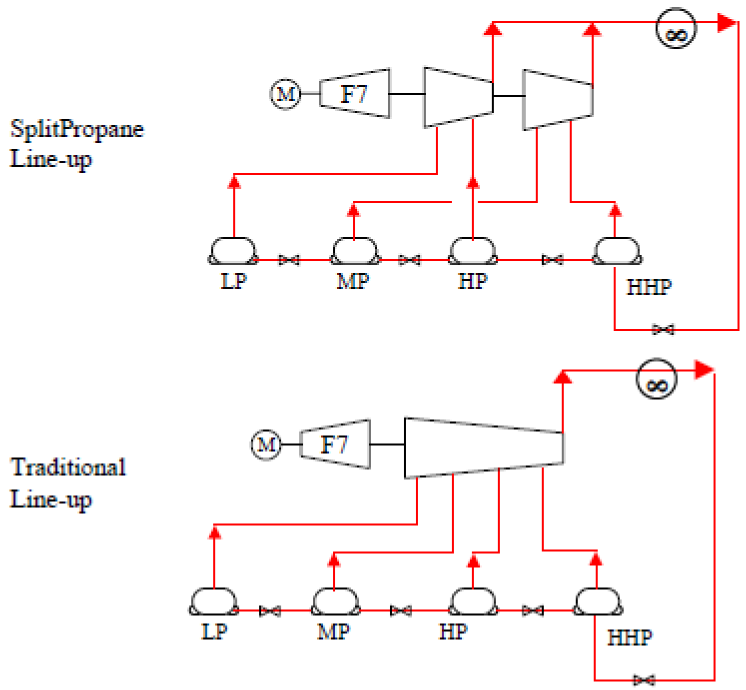


Figure 4.4 – Schematic process line-up is showing the SplitPropane technology [48].

Taking into consideration split propane line-up, flowsheet of propane precooled mixed refrigerant technology will look like presented in Figure 4.5 below.

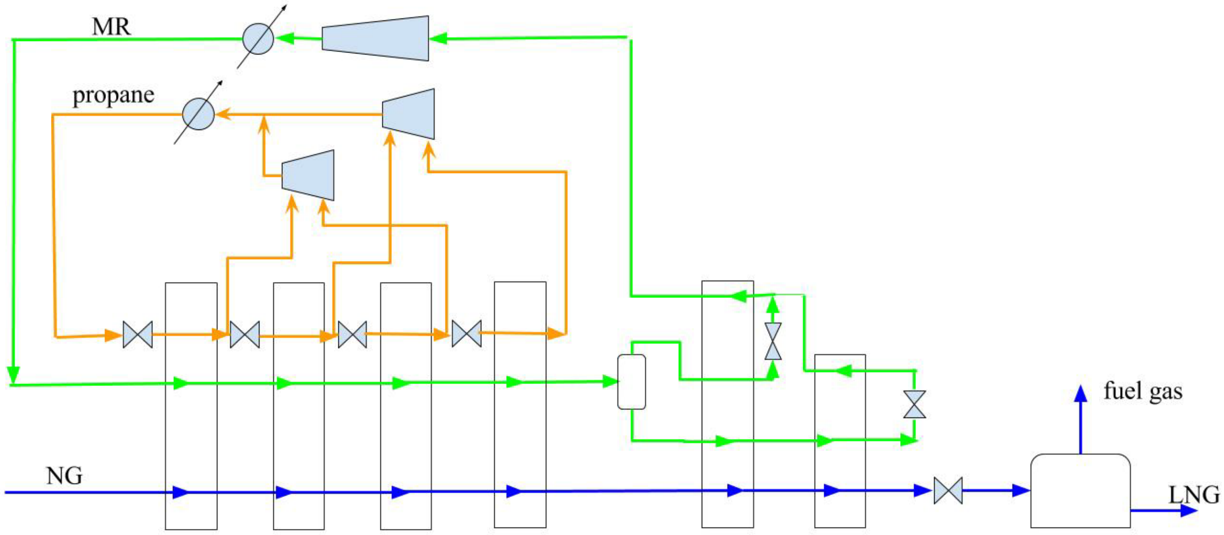


Figure 4.5 - SplitPropane precooled mixed refrigerant process.

For further simulation very important to know refrigerants composition. The part 2.4 of this paper presents components of mixed refrigerant. They are – nitrogen, methane, ethane, propane. Since these proportions are unknown, they are proposed in Table 4.3.

Table 4.3 – C3MR mixed refrigerant composition

Component	1.MR	2.MR	3.MR
Nitrogen	0.40	0.20	0.08
Methane	0.38	0.45	0.27
Ethane	0.18	0.20	0.50
Propane	0.04	0.15	0.15

Table 4.3 gives 3 probable variants of the mixed refrigerant composition. During simulation procedure, there will be chosen one of them and adjusted to achieve lower compressors power consumption.

4.2.2 Shell DMR – Double Mixed Refrigerant process

Shell Company had developed natural gas liquefaction process DMR in 2002 for middle and large-scale (2-5MTPA) liquefaction facilities [17]. The technology (Figure 4.6) uses two streams of mixed refrigerants, first (MR1) for the precooling purposes and second (MR2) for liquefaction and subcooling until -150°C.

Dry clean natural gas (NG) is coming into first and then second heat exchanger for precooling and partial liquefaction by the first stage refrigerant (MR1). In case if the natural gas contains heavy components, they must be separated after second heat exchanger with the help of fractionation column under temperature -38°C. However according to given conditions, gas does not have heavy components and can be liquefied without intermediate fractionation.

After precooling, NG flows to a main cryogenic heat exchanger where is liquefied and subcooled by MR2 with lower boiling temperature. The final pressure drop to atmospheric pressure provided by Joule-Thompson valve.

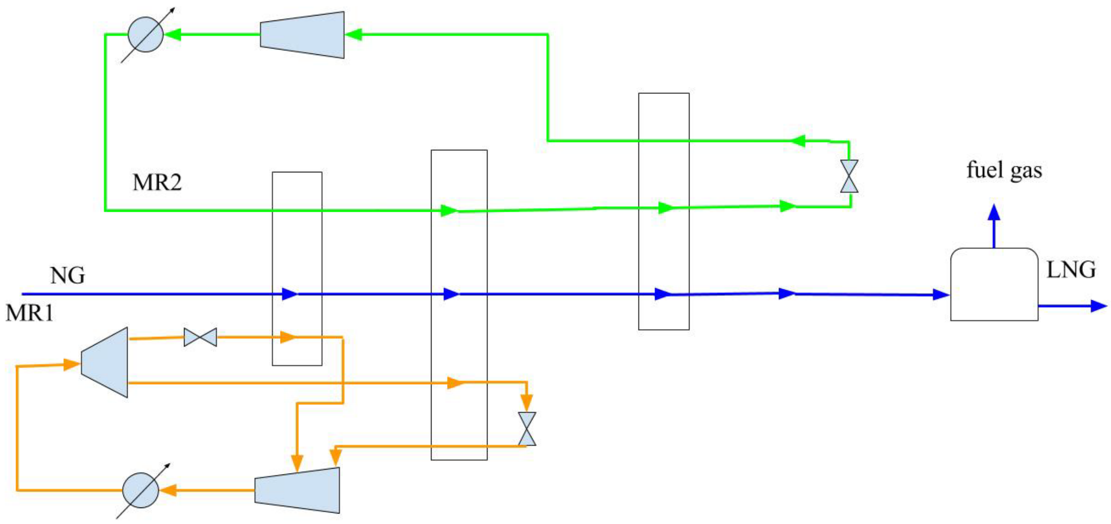


Figure 4.6 – Dual Mixed Refrigerant process

Figure 4.6 presents a simplified simulation model, which can be improved further to achieve better results.

DMR technology uses two different refrigerants, first (MR1) contain ethane, propane and butane, while the second (MR2) consists of nitrogen, methane, ethane and propane. Mixtures proportions are also assumed. For each MR case 3 variants of compositions are developed and listed in Table 4.4.

Table 4.4 – DMR refrigerants composition

Component	MR1			MR2		
	1.MR1	2.MR1	3.MR1	1.MR2	2.MR2	3.MR2
Nitrogen	0.00	0.00	0.00	0.22	0.10	0.05
Methane	0.00	0.00	0.00	0.40	0.40	0.50
Ethane	0.40	0.27	0.18	0.33	0.45	0.37
Propane	0.40	0.43	0.64	0.05	0.05	0.08
i-butane	0.15	0.19	0.10	0.00	0.00	0.00
n-butane	0.05	0.11	0.08	0.00	0.00	0.00

Refrigerants, which give less temperature cross in heat exchangers according to simulation, will be adjusted to rich minimum temperature approach 3°C and smooth cooling curve responsible for lower compressors work.

4.2.3 Statoil-Linde MFC – Mixed Fluid Cascade.

The Mixed fluid cascade was specially developed for natural gas liquefaction plant in Norway, Snøvit and can be applied to large LNG trains (>4 MTPA). At present time, Snøvit plant in Hammerfest is only one, which utilises MFC technology [7].

The process was created to withstand severe environment conditions. The main difference of this technology from the classic cascade is utilising of mixed refrigerants instead pure ones, which potentially improve thermodynamic efficiency and flexibility of the process.

Figure 4.7 presents the following: feed gas entering the process is cooled, liquefied and subcooled by three separate mixed refrigerant cycles. Precooling cycle transfer cold in plate-fin heat exchangers with the help of first mixed refrigerant (MR1) consisting of ethane, propane and butane.

Heat transferring in liquefaction and subcooling cycles flows in spiral-wound heat exchangers. Mixed refrigerant of the second stage (MR2) contain methane, ethane, propane while nitrogen is added to mixed refrigerant of the third cycle (MR3) to achieve lower temperatures (-150°C).

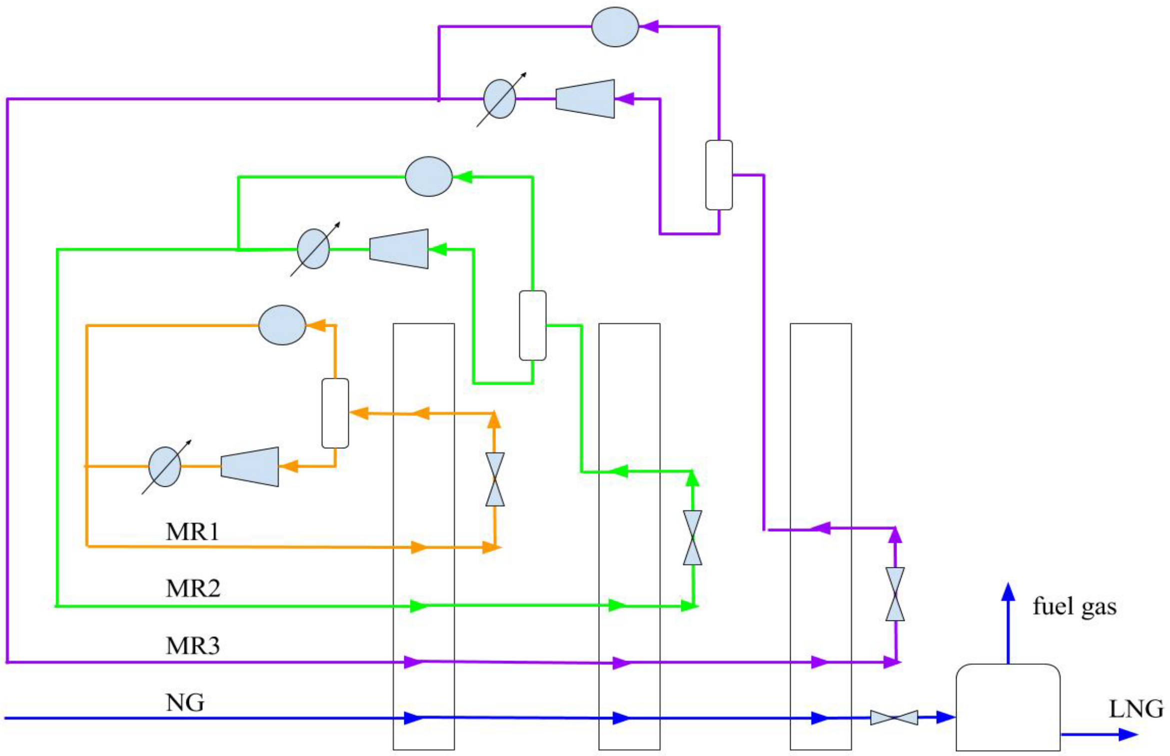


Figure 4.7 – Mixed Fluid Cascade

Three refrigerant compression systems can have separate drivers – GE Frame 7E or to be connected to the general electric driving system.

Three variants of each mixed refrigerant are presented below in Table 4.5. It is also important to notice that refrigerant adjustments must be started from MR3 to MR1 because with modifying the last refrigerant composition and, therefore, pressure, previous stages' curves also change, which requires refrigerant composition correction.

Table 4.5– MFC refrigerants variants

Component	MR1			MR2			MR3		
	1.MR1	2.MR1	3.MR1	1.MR2	2.MR2	3.MR2	1.MR3	2.MR3	3.MR3
Nitrogen	0.00	0.00	0.00	0.00	0.00	0.00	0.20	0.10	0.05
Methane	0.00	0.00	0.00	0.40	0.20	0.10	0.50	0.40	0.60
Ethane	0.70	0.50	0.35	0.50	0.60	0.60	0.30	0.50	0.35
Propane	0.30	0.40	0.55	0.10	0.20	0.30	0.00	0.00	0.00
i-butane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-butane	0.00	0.10	0.10	0.00	0.00	0.00	0.00	0.00	0.00

It is assumed that LNG production rate is constant and equal 5 MTPA, which is 5,708x10^5kg/s. Then during one year with temperature fluctuations compressors will be loaded by different ways. Compressors work will variate and require more or less power from gas turbines. Inlet pressure for all processes is 5000 kPa.

The purpose of the following simulations is to determine processes energy consumption within one year for average month's temperatures and show, which technology requires less power. Then we find average power consumption and assume it constant. It means that in reality, LNG production rate will vary.

4.3 Simulation procedure and results

4.3.1 C3MR

The program starts from opening Simulation Basis Manager. It requires entering following parameters:

- "Components" - from which natural gas and refrigerants consist of;
- "Fluid Package" – for liquefaction natural gas it is Peng-Robinson.

After completion of this procedure, there is becoming possible to create simulation environment, i.e. process flowsheet. The C3MR scheme is shown in Figure 4.5, while real simulation environment is presented in Appendix 1.

There is also assumed that natural gas and refrigerants come into the process with equal to ambient temperatures. To support such assumption condensers in refrigerants cycles are air-coolers, which can use the best attributes of cold winter air.

Natural gas is coming into the first heat exchanger (HE1) with the pressure 5000 kPa, which is constant during the process. This constant pressure assumption allows operating heat exchangers without any temperature drop, simplifying simulation.

The first part of the scheme is propane cooling cycle. It utilises four heat exchangers, i.e., own temperature and pressure level lies after each heat exchanger:

T_1=T_7=1°C;

T_2=T_8= -14°C;

T_{38}=T_{39}= -25°C;

T_3=T_9= -38°C.

For the brief pressure determination, there is important to look at pressure-enthalpy diagram for propane (Figure 4.8). The diagram was plotted for April temperature conditions (isotherm T=5°C).

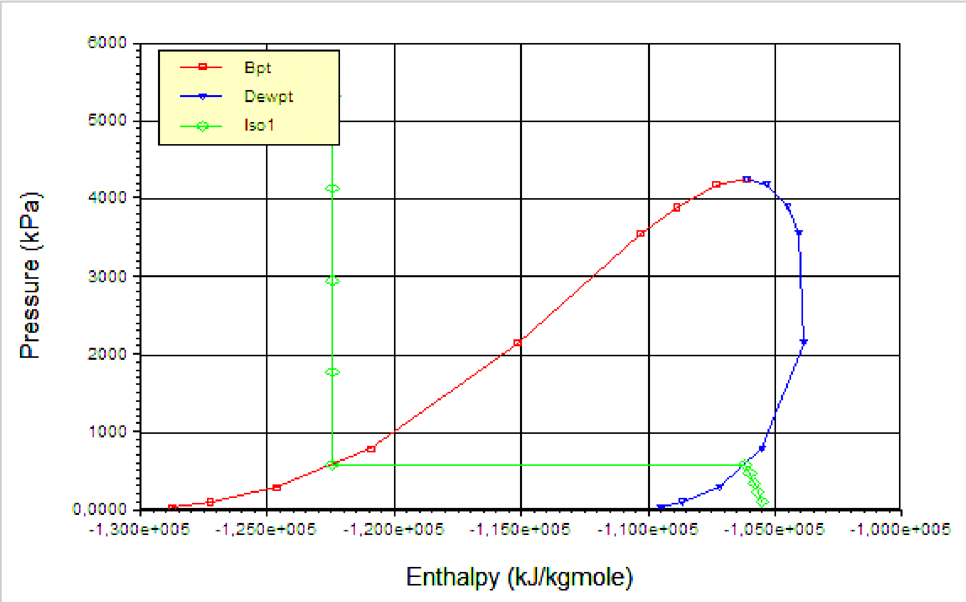


Figure 4.8 – Pressure-enthalpy diagram for propane. (UniSim: Envelope Utility).

Isotherm1 cross propane envelope in Figure 4.8 in the point $P_{propane}=P_0=595$ kPa. We choose this pressure as initial value. Then set a minimum pressure for stream 46. It should be slightly less than atmospheric pressure ($P_{46}= P_4=106$ kPa). The next step is calculation of pressure relation and pressure for each step (equations 6 and 7).

$$R = \sqrt[m]{\frac{P_i}{P_m}}, \qquad (6)$$

where i – initial conditions, m – number of steps.

$$R = \sqrt[4]{\frac{P_0}{P_4}} = \sqrt[4]{\frac{595}{106}} = 1{,}54.$$

$$R = \frac{P_0}{P_1} = \frac{P_1}{P_2} = \frac{P_2}{P_3} = \frac{P_3}{P_4}\,. \qquad (7)$$

Therefore, $P_1= 390$ kPa, $P_2= 253$ kPa, $P_3= 164$ kPa.

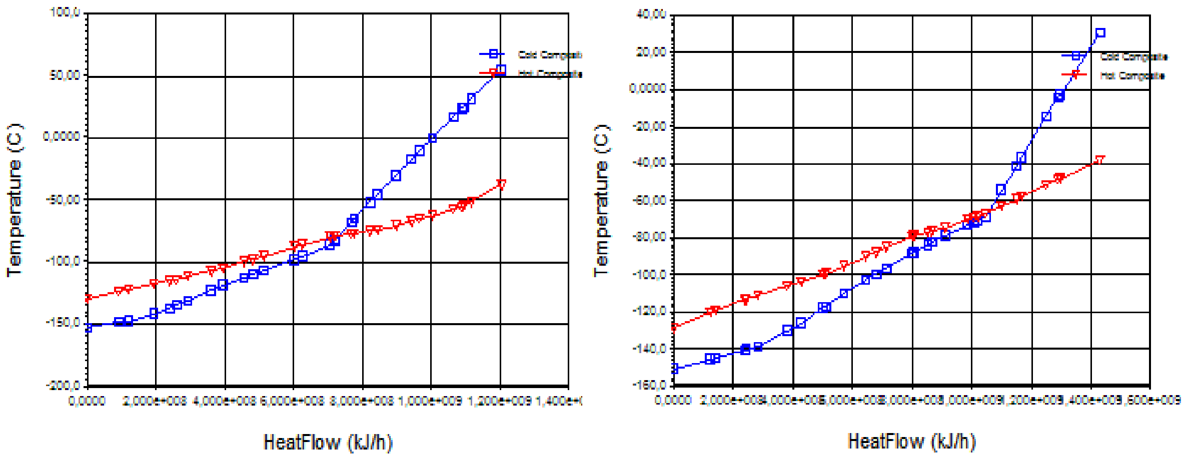
After pressure is set, program can calculate temperatures relevant to above pressures.

When propane circuit is ready, the mixed refrigerant loop must be prepared. As a first step, temperature after HE5 and HE6 is set like described in APCI patent [29] $T_4=T_{11.1}=T_{10.1}=$ -129°C and $T_5=T_{11.2}=$ -151°C.

After temperature identification, we start a determination of mixed refrigerant composition, flow rate and pressure. Proposed mixed refrigerant compositions presented in Table 4.3. Assumed mass flow rate is $q_{MR}=2.2x10^6$kg/s, pressure depends on fluid composition.

If refrigerant contains more light components then heavy ones, the pressure will be higher. Pressure p_{MR} is assumed equal to 2000 kPa.

The 1.MR refrigerant from table 3.5 with high nitrogen content gives huge temperature cross in HE5 (Figure 4.9a) and minimal approach in HE6 ΔT=18°C. The next refrigerant 2.MR also gives temperature cross, big enough to stop considering this composition.



a b

Figure 4.9 – Temperature cross in HE5 for a) 1.MR and b)2.MR

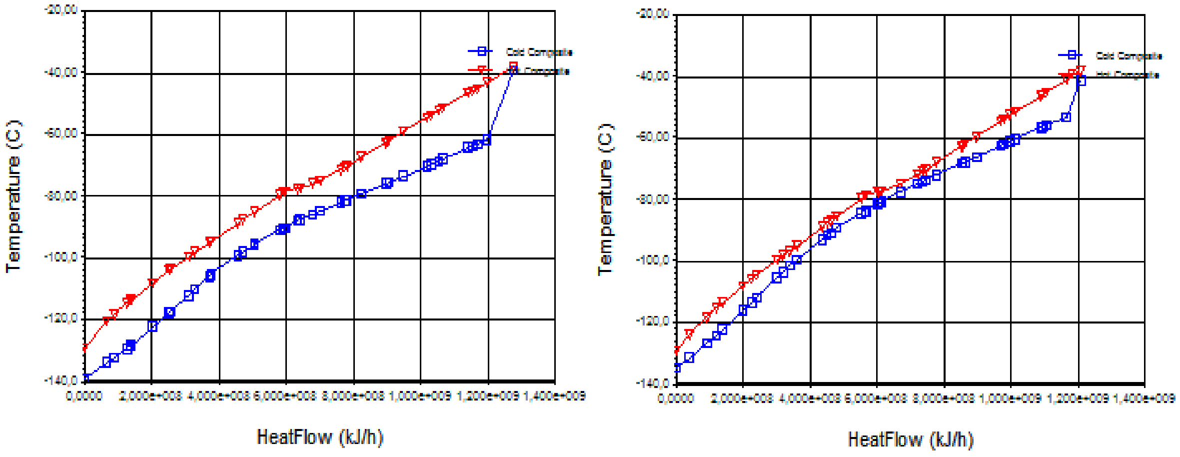
Figure 4.9 shows that high nitrogen and methane content for refrigerant is not a good decision. Light components will boil off fast, making cooling in this heat exchanger impossible.

The last refrigerant 3.MR contain much less nitrogen and methane. Figure 4.10a presents that this composition is suitable, gives minimal approach ΔT=1.2°C and can be used after adjustments. The new composition of refrigerant is mentioned in Table 4.6, new pressure p=1850kPa, and mass flow rate $q=2.1x10^6$kg/s.

Table 4.6 – C3MR mixed refrigerant composition after adjustment.

Component	MR
Nitrogen	0.076
Methane	0.248
Ethane	0.489
Propane	0.187

The cold composite curve on Figure 4.10b almost duplicates hot composite curve, which gives the best compressors power and coolers duty (Table 4.7).



a b

Figure 4.10 – Mixed refrigerant T-Q diagram a) before and b) after adjustments.

When mixed refrigerant loop works properly, there is time to adjust mass flow rate and pressure for propane loop. Assume mass flow rate for propane $q_{propane}=2x10^6$kg/s. Pressure for propane loop was determined before. Under these pressures, minimum approach in heat exchangers will vary between 7°C in HE1 to 3°C in HE4. For equalising minimal approach in all heat exchangers MA=3°C (3-4°C is recommended), we use function “adjust”. This function regulates pressures after all heat exchangers depending on changes in inlet conditions. Such pressure variations help to keep compressors work on the low level (Table 4.7) and avoid exceeding work. Mass flow rate can also be reduced to $q=1.7x10^6$kg/s, it will not change the process flow but reduce propane consumption.

Table 4.7 – C3MR power consumption results for 5°C ambient temperature.

Unit	Propane	Unit	MR	Total
Compressors power, MW				
K_1	11.81	K_{1MR}	25.60	130.40MW
K_2	9.66	K_{2MR}	29.32	
K_3	14.47	K_{3MR}	36.61	
K_4	2.93			
Coolers duty, MW				
E-100	126.20	E-102	6.87	256.18MW
E-103	77.72	E-101	45.39	
				∑= 386.58MW

The table above gives only results for example case with ambient temperature 5°C. The pressure, flow rates and power for every month are mentioned in Appendix 2.

4.3.2 DMR

The double mixed refrigerant process got the name because of the existence of two different mixed refrigerants loops. The DMR simulation procedure starts by the same way as C3MR. Technology flowsheet is modelled as described in paragraph 4.2.2 by the Figure 4.6. Actual scheme of the process, which was used for simulations, is in Appendix 1 (DMR).

It is assumed that natural gas (NG) coming into heat exchanger 1 (HE1) has a temperature equal to ambient conditions, the same as refrigerants after air coolers. An example of calculation procedure and simulation is described for April with T_{amb}=5°C. Pressure of natural gas is assumed p_{NG}=5000kPa. Mass flow rate is varying, because LNG production is proposed to be constant, and q_{LNG}=5.708x10^6kg/s.

Flow scheme used for simulation has four heat exchangers. The first and second heat exchangers (HE1 and HE2) utilise mixed refrigerant of the first stage as cooling fluid. It chills NG and the mixed refrigerant of the second stage from temperature 5°C to -25°C and -50°C in the first and second heat exchangers respectively.

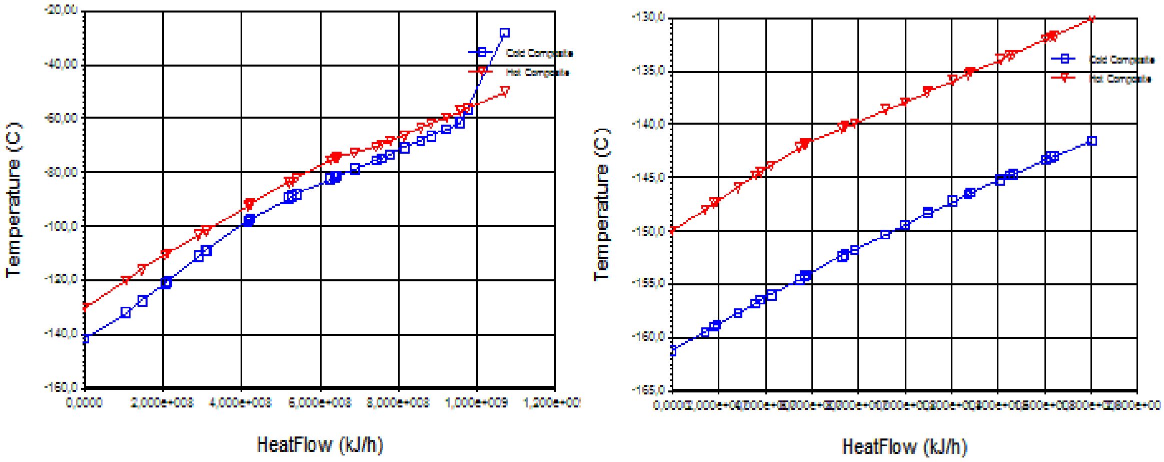
MR1 should contain different components to support a wide range of cooling temperature, i.e. ethane, to allow cooling till -50°C, and also heavier components like propane and butane to prevent fast boiling of the refrigerant and make refrigerant cooling curve slope as close as possible to hot components curve slope. It would make the process more efficient and will require less work of compressors than in case cooling curve does not duplicate hot components curve. However, for creation of appropriate composition, all parameters of the second mixed refrigerant (MR2) must be known.

Mixed refrigerant of the second stage (MR2) chills natural gas also in two heat exchangers (HE3 and HE4) to temperature -130°C and -150°C. The scheme allows having single heat exchanger for the second stage, but temperature difference between hot and cold side would be large. Hence, we took a decision that having two heat exchangers, joined in one cold box, for this stage would be better solution.

MR2 is responsible for liquefaction and subcooling of natural gas. It should contain such components as nitrogen and methane, responsible for low temperatures, and ethane with propane, which start working close to the hot side of the main cryogenic heat exchanger (MCHE). Pressure for MR2 depends on the composition of refrigerant. Therefore, a unique pressure is expected for each case.

Proposed variants for the second mixed refrigerant composition as well as for the first one are presented in Table 4.4. First of all, we check the following variants: 1.MR2 with high nitrogen content and 3.MR2 with much less amount of nitrogen.

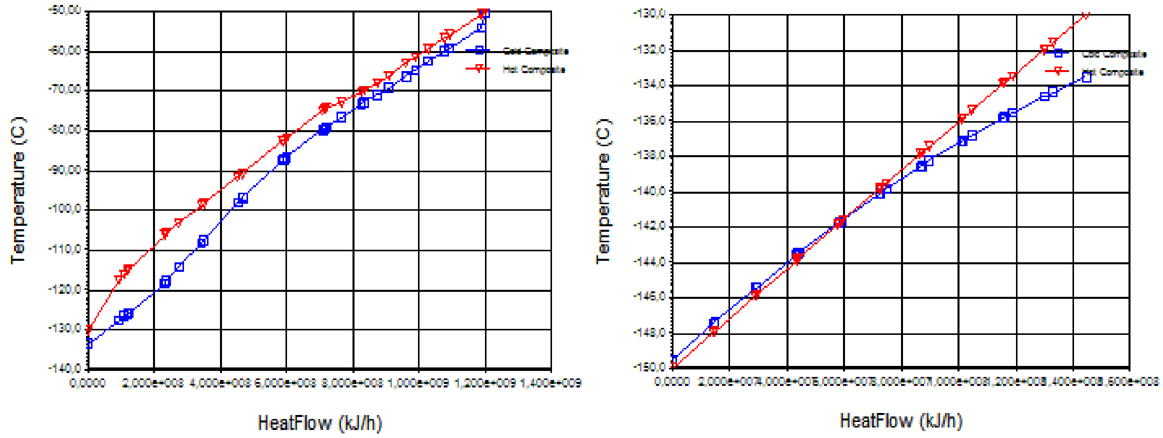
For refrigerant 1.MR2 we assume pressure 2000kPa, flow rate 2,1x10^6kg/h and pressure after expansion valve VLV-102 is 500kPa. Such conditions give temperature cross in HE3 and minimum approach more than 10°C in HE4 (Figure 4.11). Minimum approach is recommended to be between 3-10°C. Changing of the pressure or flow rate can not solve the problem. On the contrary, there expected an increase in compressors loading, which is undesirable.



a b

Figure 4.11 – Heat flow provided by 1.MR2 in a) HE3 and b) HE4.

It is apparently that here the content of light components is high and does not allow operating in appropriate way. That is why next composition 3.MR2 was chosen for analysis. Due to the low content of nitrogen (5%), pressure was assumed 1500kPa, while the flow rate is the same as for the previous composition $q=2.1x10^6$kg/h. The pressure of stream 5 is 500kPa. Refrigerant 3.MR2 under these conditions shows quite good results. However it still gives pressure cross in HE4 and situation close to temperature cross in HE3 (Figure 4.12).

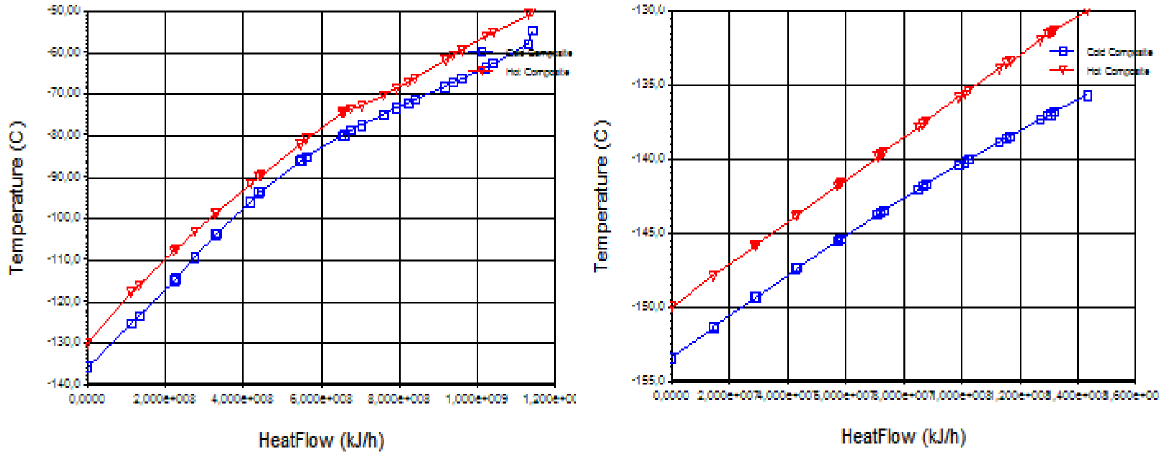


a b

Figure 4.12 – Heat flow provided by 3.MR2 in a) HE3 and b) HE4.

Analysis of the above depicted graphs gives following results: the content of nitrogen must be slightly higher to downgrade left side of the cold composite curve in HE4; percentage of ethane better to be raised to demote right side of the cold composite curve in HE3. That is why the middle composition (variant 2.MR2) might be the best choice.

The flow rate for 2.MR2 is the same as for two previous cases. Pressure increases due to nitrogen content increase, so p=1550kPa. The pressure of expanded stream coming into HE4 is 500kPa. Results of this simulation are presented below.



a b

Figure 4.13 - Heat flow provided by 2.MR2 in a) HE3 and b) HE4.

From the Figure 4.13, one can see that cold composite curve almost duplicate hot composite curve. This result is good for subsequent simulation. However, for reduction of the energy consumption of compressors, the adjuster can be applied to reduce minimum approach in the heat exchanger to the level of 3°C. This adjuster changes the pressure of stream 5 until the best result is acquired. So, the pressure of cold MR2 coming into HE4 becomes 529.5kPa.

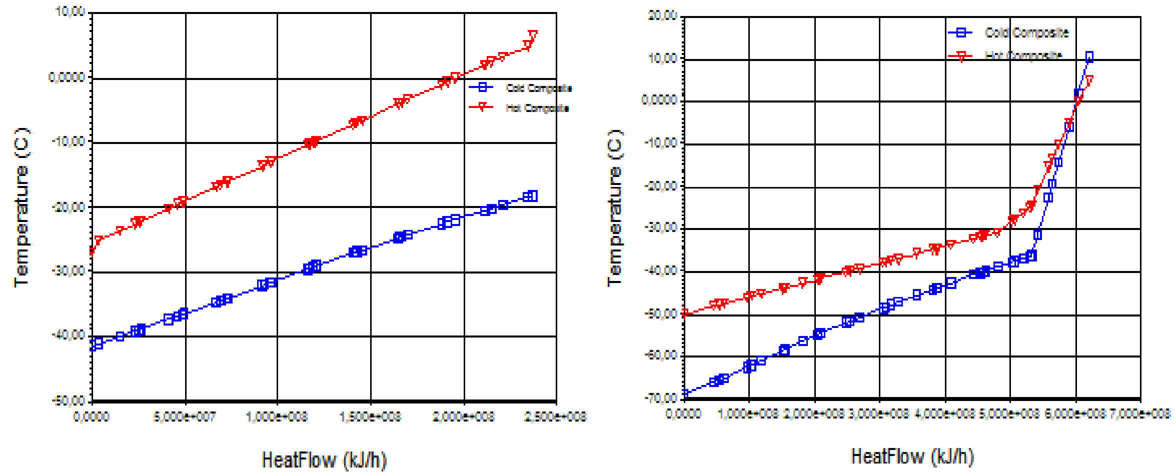
Hot MR2, outgoing from HE3 is compressed then by tree-stage compressor until approaching an initial pressure level of 1550kPa. Then the heat of compression is removed by air cooler, and the refrigeration loop is closed.

It is also important to keep the temperature of NG after HE4 constant to prevent changing the flow rate of natural gas because the flow rate of LNG is assumed constant.

After the procedure of setting MR2 loop parameters is completed, data of the first refrigeration stage must be specified. This step has two heat exchangers, which utilise MR1 under own pressures and flow rates.

The loop starts from the unit called splitter. For chilling in the second heat exchanger down to colder temperatures, refrigerant should have larger flow rates and lower pressure than those in the first heat exchanger. That is why, as a first assumption, TEE-100 splitting parameters are set in amounts 40% and 60% for stream 11 coming into HE1 and stream 15 entering HE2 respectively. Pressures after expansion valves are also assumed. For stream 17 coming into HE2 pressure is 100kPa, slightly less than atmospheric pressure. For stream 13 it should be higher, for example, 300kPa. The first compressor increases pressure from 100 to 300kPa. The function “set” helps to make pressure after the first compressor K1MR1 the same as the pressure of cold stream outgoing from HE1. It is necessary to mix the flow compressed by K1MR1 with the flow after HE1 without any pressure loss.

After preparations are over, the procedure of MR1 selections is started. From the Table 4.4, where variants of refrigerant were guessed, we take 1.MR1 and check its behaviour. Flow rate should be in 3-4 times higher than one of natural gas, so $q=2x10^6$kg/h. The pressure of such composition is required to be on the level of 1220kPa, which is taken from the envelope utility for this composition.



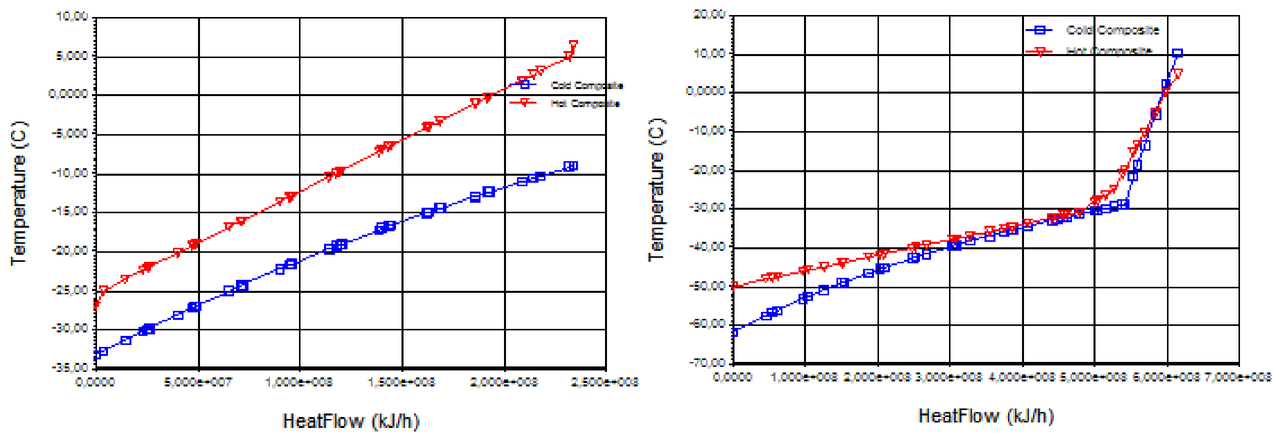
a b

Figure 4.14 - Heat flow provided by 1.MR1 in a) HE1 and b) HE2.

The above figure describes heat flow in heat exchangers. Big difference between curves in Figure 4.14a (more than 14°C) and fast evaporation giving temperature cross in Figure 4.14b tell about the necessity of reduction of light components content.

Other two variants 2.MR1 and 3.MR1 consist of approximately twice less amount of ethane and almost the same amount of propane. The further step is to test these two compositions and choose the better one.

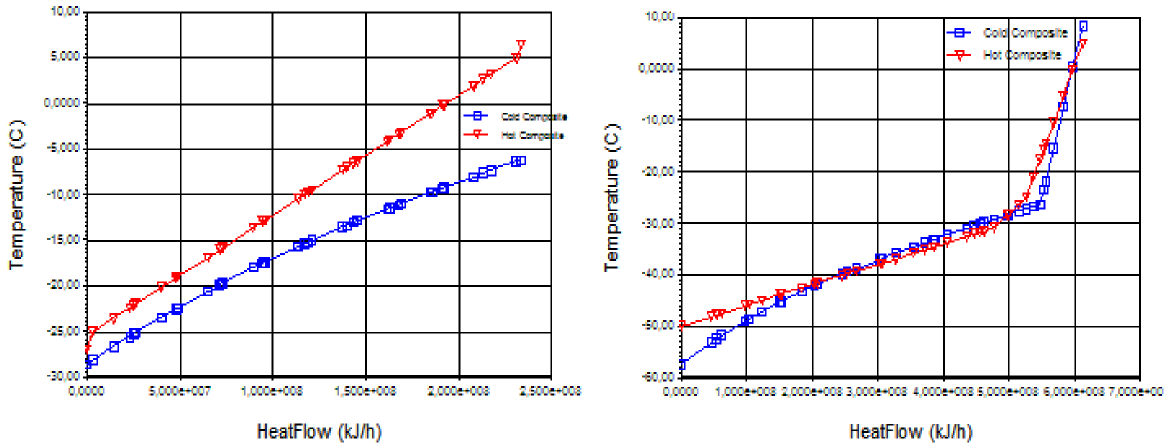
Flow rates are left constant, while pressures for 2.MR1 is 920kPa and for 3.MR is 780kPa, which are shown by phase envelopes at ambient temperature.



a b

Figure 4.15 - Heat flow provided by 2.MR1 in a) HE1 and b) HE2.

Above result shows that cold curve on the Figure 4.15a almost parallel to the hot line. The minimum approach here is 6°C with maximum temperature difference of 15.5°C for top points (hot side). Temperature cross still exists in heat exchanger HE2.

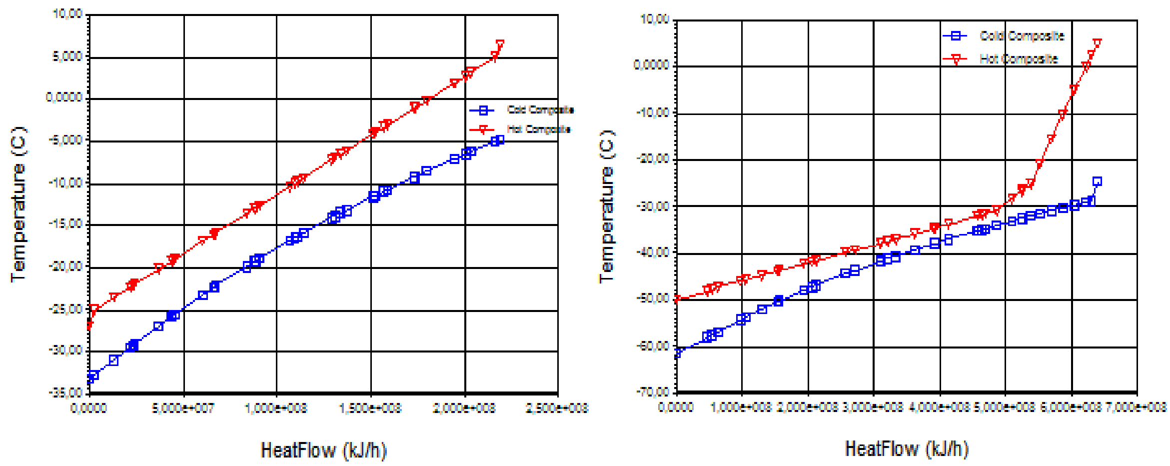


a b

Figure 4.16 - Heat flow provided by 3.MR1 in a) HE1 and b) HE2.

From the Figure 4.16a, it can be observed that the shape of the cold curve was changed in comparison with 2.MR1. Minimum approach decreased, which means that heavy components content better for the variant 2.MR1. Heat flow in second heat exchangers HE2 (Figure 4.15b and Figure 4.16b) looks almost the same, cold flow duplicate hot flow, but for 2.MR1 cold curve lies on the lower level, which is more preferable.

The abovementioned results are not satisfactory. It means that problem is not about composition, it is in flow rates. The most promising variant is 2.MR1. In this case, changes in flow rates can be created by reducing flow, coming in HE1, and enhancement correspondingly cold flow in HE2. Therefore splitting parameters will be other, i.e. 30% and 70% to HE1 and HE2 respectively. Renewed result of heat flow is mentioned below in Figure 4.17.



a b

Figure 4.17 – Resulted heat flow provided by 2.MR1 in a) HE1 and b) HE2.

The abovementioned results do not give temperature cross and curves lie almost in parallel, which is good enough to get appropriate compressors loading. However, the minimal approach in heat exchanger HE1 is still 6°C, which can be reduced manually or with the help of adjuster to 3°C.

After heat exchangers, the flow has four stage of compression, which increase the pressure of the refrigerant to 920 kPa. Coolers are located after each compression step to reduce temperature down to ambient, which prevents exceeding of compressors work and appearing of overheated vapour. After the last air cooler, the loop is closed.

For calculation of total process power consumption, every unit power should be considered. For this purpose Table 4.8 below is created.

Table 4.8 – DMR power consumption results for 5°C ambient temperature.

Unit	MR1	Unit	MR2	Total
Compressors power, MW				
K_{1MR1}	30.37	K_{1MR2}	22.89	118.92 MW
K_{2MR1}	13.04	K_{2MR2}	24.95	
K_{3MR1}	7.55	K_{3MR2}	17.17	
K_{4MR1}	2.61			
P_{1MR1}	0.01			
P_{2MR1}	0.11			
P_{3MR1}	0.23			
Coolers duty, MW				
E-104	15.22	E-103	15.74	244.02 MW
E-102	83.91			
E-100	80.01			
E-101	50.76			
				∑=362.94 MW

The above table shows how power consumption was calculated for ambient temperature 5°C. Results for other months are in Appendix 2. From comparison of Table 4.7 for C3MR process and Table 4.8 for DMR, it is clear that DMR technology consumes less power.

4.3.3 MFC

Mixed fluid cascade is the process created especially for Hammerfest area to withstand severe climate conditions with the help of optimal refrigerant composition and the possibility to operate refrigerants loops independently from each other.

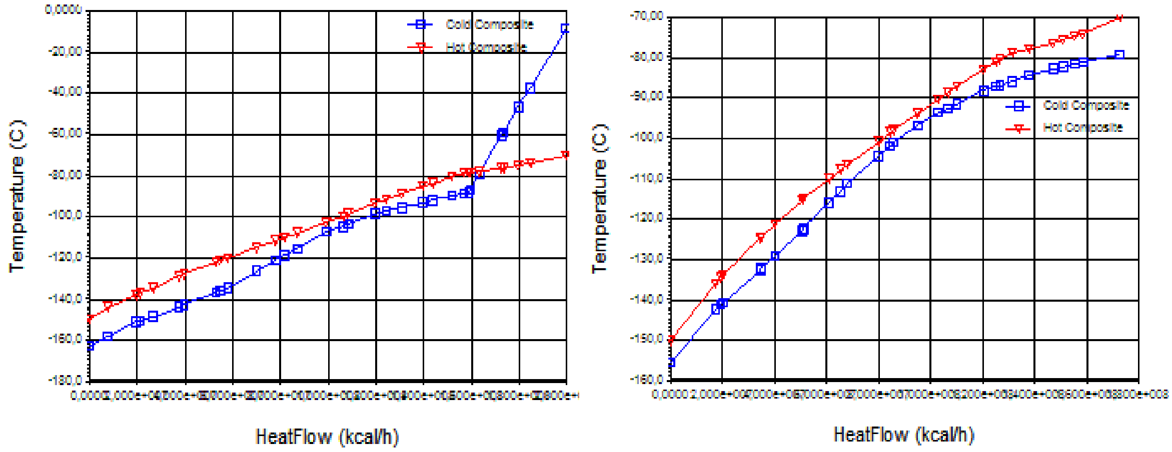
There are three refrigeration loops in the process. Each loop operates on its own temperature and pressure level. The refrigerant of the first stage is MR1. It consists of ethane, propane and butane. During this stage, MR1 cools natural gas and refrigerants of higher levels in two heat exchangers from ambient temperature (for example case “April”, T_{amb}=5°C) to temperature -35°C.

Mixed refrigerant of the second stage liquefies natural gas and cools mixed refrigerant of the next stage at the temperatures from -35°C to -70°C in single heat exchanger (HE3). Second refrigerant (MR2) in contradistinction from the first stage refrigerant (MR1) contain lighter component – methane and does not contain butane.

For the third stage, MR3 is responsible. It contains ethane and methane as well as nitrogen. Subcooling of natural gas in HE4 is processed from -70°C to -150°C.

The problem with refrigerants composition is the same as for the previous two cases of C3MR and DMR, i.e. components are well known, but rates must be settled during the simulation process. Adjustments are started from refrigerant for the highest stage, because any changes in it would cause variations in hot composite cooling curves for previous two stages. Variants of mixed refrigerants are specified in Table 4.5.

We start our search for refrigerant from composition 1.MR3. It contains a huge amount of light components, 20% of nitrogen and 50% of methane, which can probably allow cooling down to -150°C. This idea requires simulation support. Pressure is set to 1500 kPa and mass flow rate to 1.3×10^6kg/h. After expansion valve, pressure is assumed to be 350kPa. These conditions give heat flow like presented in Figure 4.18a below.



a b

Figure 4.18 – Heat flow provided by a)1.MR3 and b)2.MR3

Detailed analysis of Figure 4.18 concerning 1.MR3 refrigerant cooling curve, indicates that composition can be changed towards adding heavier components and reducing the percentage of nitrogen and methane. Such a shift would prevent temperature cross.

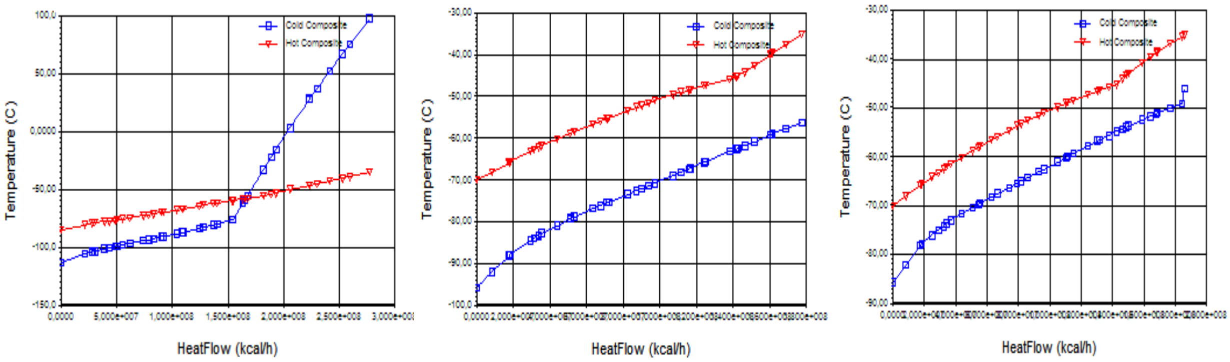
Considering a 2.MR3 variant of refrigerant composition, we preserve the flow rate, but reduce pressure down to 1400 kPa because of less nitrogen content. The same pressure after expansion, gives a heat flow diagram presented in Figure 4.18b. From this plot, one can notice that cold curve is almost the same as a hot composite curve, which indicates that compressor work can be low enough to apply 1.MR3 for the process.

The further diminishing of nitrogen content can lead to undesirable temperature cross. That is why the case 2.MR3 should be deemed as final.

Adjuster 1 (ADJ-1) regulates the pressure of stream 3 to get minimal approach in heat exchanger equal to 3°C. This alteration does not change pressure after the valve (VLV-100) in this temperature case and remain pressure equal to 350 kPa.

Outlet flow from HE4 is two-phase flow, and before compression, it must be separated to send vapour into compressors K1MR3 and K2MR3 and liquid into pump P1MR3. Compressors power is 47.47 MW, and pump requires 0.05 MW. Than fluids are mixed, and loop 3 closes.

Next circuit operated by MR2. The selection procedure is described below. The flow rate for every case is q=1.6x10^6kg/h, pressure after expansion is 200kPa. Inlet pressure varies in accordance with composition.



a b c

Figure 4.19 – Heat flow provided by a) 1.MR2, b)2.MR2 and c)3.MR2

For the first variant 1.MR2 (Figure 4.19a) seems like any pressure leads to huge temperature cross. Therefore, 40% of methane in the refrigerant composition is too much, and this percentage must be reduced. Next composition is 2.MR2 requires pressure approximately 1500kPa to avoid temperature cross in the heat exchanger. From Figure 4.19b, we see that cold curve lies far from hot curve leading to overwork of compressors (more than 70MW for the loop). Third variant 3.MR2 seems more promising for further adjustments, as cooling curve lies at the less distance than in case b. However, power consumption is still high because cold cooling curves lie far away from each other. For compression of this refrigerant to the pressure 820kPa, more than 50MW is spent in the process.

Changing slightly the pressure of inlet and compressed streams, and also composition, we get following parameters (Table 4.9). Heat flow graph is presented in Figure 4.20.

Table 4.9 – MR2 composition and operation parameters.

Component	Amount	Stream	Pressure
Methane	8%	MR2	760kPa
Ethane	61%	13	270kPa
Propane	31%		

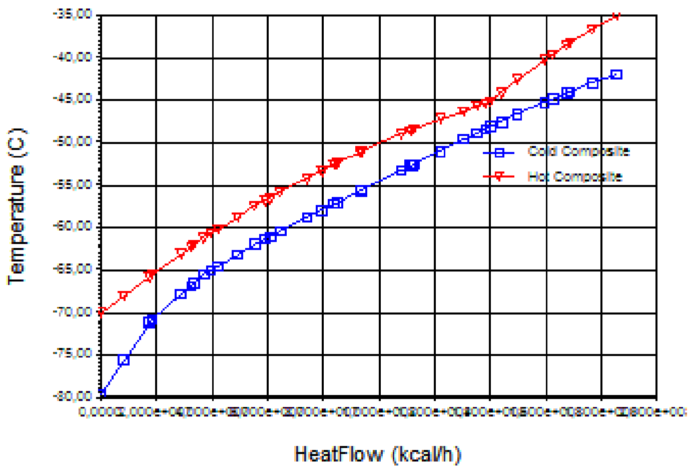


Figure 4.20 – Heat flow provided by MR2.

Figure above shows that cold components curve is close enough to the hot fluids cooling curve. This loop requires only 36 MW for increasing pressure in compressors, which is much less than was observed in the previous cases. Further adjustments, such as adding propane and reducing ethane content would cause cross in the left section of the graph and linearization in the right section, at the same time adding lighter components and reducing propane percentage would cause temperature cross in the right section. Composition from Table 4.9 is considered as the best choice.

After heat exchanger, separator S2 is set. It is used for preventing liquid coming into compressors. Mixer MIX-100 combines vapour and liquid streams (in the case of two-phase flow from HE3) and sends this flow to cooler E-100, where the air under ambient temperature captures heat from MR2. Then circuit is closed.

When third and second loops are ready, we can start searching for refrigerant composition for the first circuit. We already know that MR1 contain ethane, propane and butane. Let us set following parameters: mass flow rate $q=2.4x10^6$kg/h; pressures after expansion for streams 30 and 25 are p_{30}=600kPa and p_{25}=300kPa respectively; there is no need to send much flow to HE1, therefore flow 23 should be higher than flow 29 (70% to 30%) after TEE-100.

There are three variants of mixed refrigerant 1 (MR1) presented in Table 4.5. The first variant consists of only ethane and propane at the rate 70% to 30%. For such composition isotherm for 5°C crosses a boiling point curve in the point p=1944 kPa (Figure 4.21). Therefore, it is better to rectify compressors on this outlet pressure.

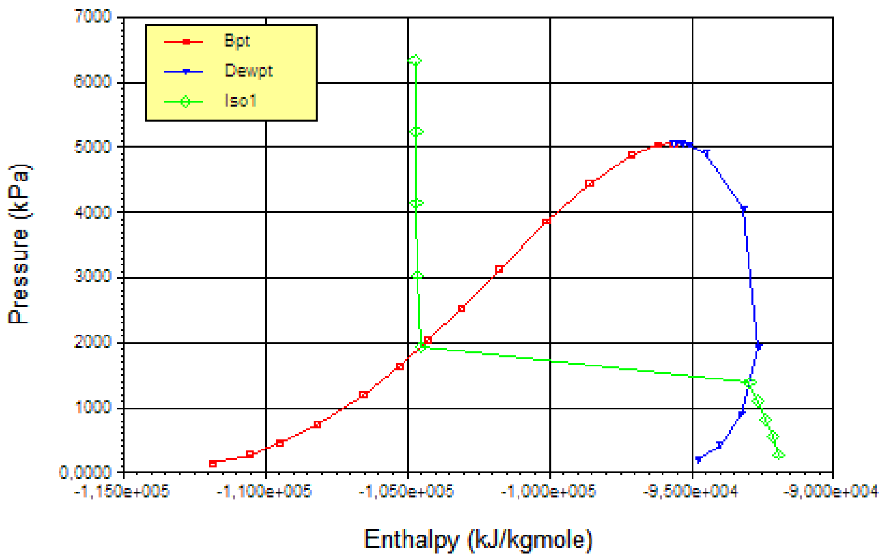
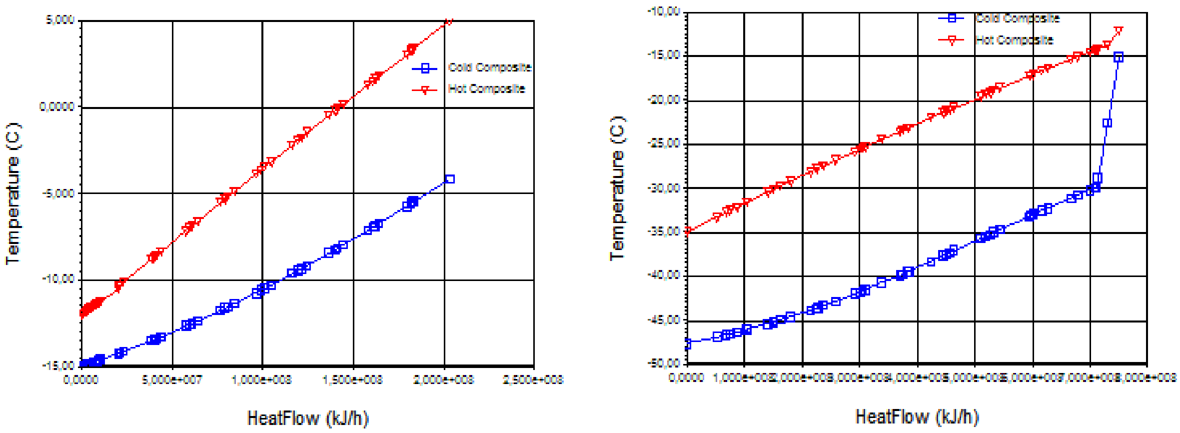


Figure 4.21– Pressure-enthalpy diagram for 1.MR1.

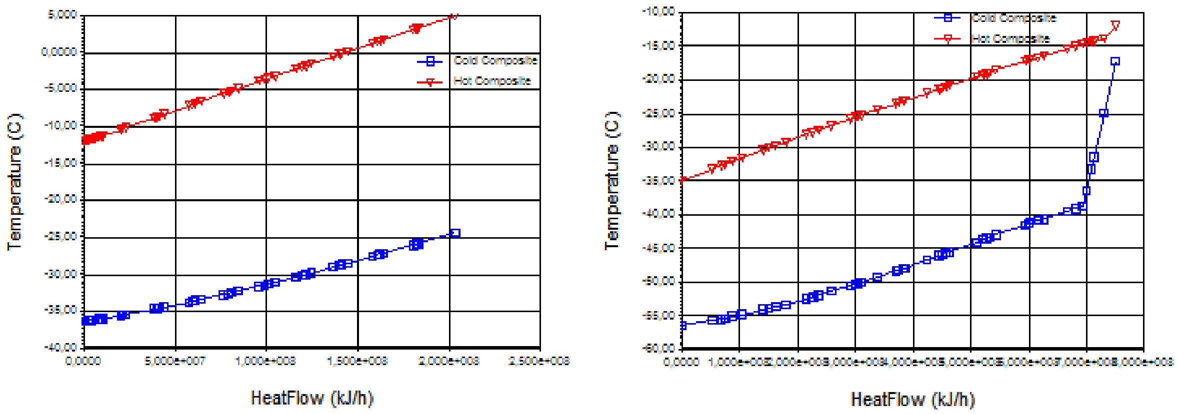
Above composition, without tuning of pressure after valves VLV-103 and VLV-102 gives a heat flow described by Figure 4.22.

We see that pressure after valves must be changed in order to reduce compressors power consumption. For this purpose, we set unit "adjuster". ADJ-4 and ADJ-3 are responsible for pressure settlement. Target variable for this operation is Minimum Approach in heat exchangers with the value 3°C. It gives us pressures 423kPa and 1174kPa. Heat flow results are shown in Figure 4.23.



a b

Figure 4.22 – Heat flow provided by 1.MR1 in a)HE1 and b)HE2 for original conditions.

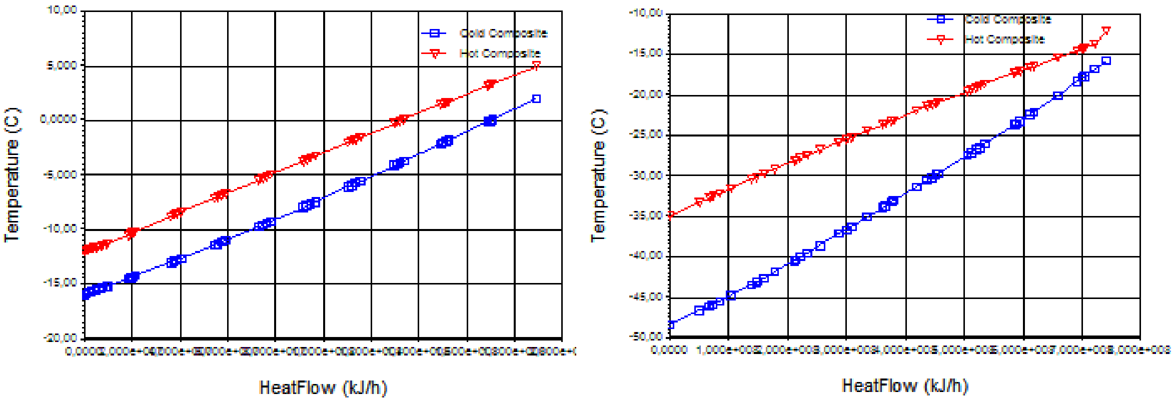


a b

Figure 4.23 – Heat flow provided by 1.MR1 in a)HE1 and b)HE2 after pressure adjustments.

From Figure 4.23 we can see that blue lines are little curved. It is better to make them linear in order to get the best numbers of compressors power. For this case, power consumption constitutes 62 MW. Forthcoming of right ends of blue lines can be achieved by adding butane to the system; therefore, case 2.MR1 should be examined next.

For 2.MR1 pressure was determined by the same way as for 1.MR1 with the help of envelope utility. The pressure is 1480kPa. Adjusters are left on, which gives pressures 845kPa and 296kPa for HE1 and HE2, respectively. Heat flow is presented in Figure 4.24.

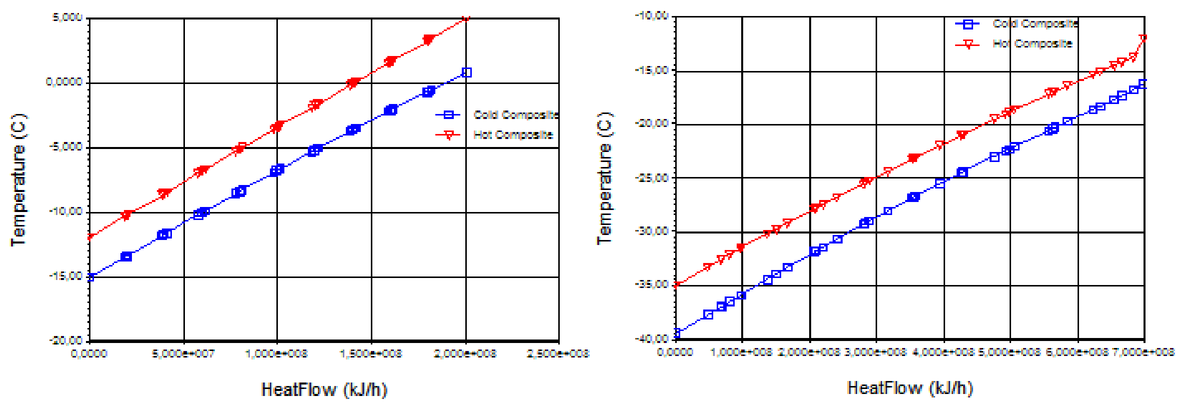


a b

Figure 4.24 – Heat flow provided by 2.MR1 in a)HE1 and b)HE2

Above Figure 4.24b shows that ethane contain can be reduced even more in order to reach up to the left end of the blue line. For this composition power of compressors achieved 55MW, which is lower than that in the previous case, but still can be decreased.

Last variant 3.MR1 requires initial pressure 1160kPa and higher flow rate q=2.8x10^6kg/h. Pressures after valves are regulated by adjusters and are as follows: 673kPa and 300kPa for streams 30 and 25, respectively. Figure 4.25 presents heat flow in heat exchangers HE1 and HE2. Power consumption for this case is reduced to the value 49 MW.



a b

Figure 4.25 - Heat flow provided by 3.MR1 in a)HE1 and b)HE2

Performance lines in Figure 4.25 look almost like parallel lines, which is very good for compressors. Small regulation of flow rates after TEE-100 (increasing stream 23 until 72% from the total flow 22) and, therefore, pressures of streams 30 and 25 can reduce compressors power to the value 48MW.

There are two stages of compression in this circuit; each stage has a separator, which prevents liquid drops from falling to compressors blades. Intermediate and final cooling of working fluid is a very important factor. It prevents overheated vapour coming into the compressor, which reduces its work.

For calculation, the total power producing by the process, each unit power consumption is to be taken into account (Table 4.10).

Table 4.10 – MFC power consumption results for 5°C ambient temperature.

Unit	MR1	Unit	MR2	Unit	MR3	Total
Compressors and pumps power, MW						
K_{1MR1}	20.25	K_{1MR2}	11.53	K_{1MR3}	35.07	125.39 MW
K_{2MR1}	24.75	K_{2MR2}	21.36	K_{2MR3}	12.08	
P_{1MR1}	0.09	P_{1MR2}	0.01	P_{1MR3}	0.01	
P_{2MR1}	0.24					
Coolers duty, MW						
E-102	8.00	E-100	1.78	E-103	8.74	251.52 MW
E-101	233.00					
						∑ = 376.91MW

4.4 Results comparison and discussion

Simulation of technologies has been proceeded for chosen refrigerant compositions (Table 4.11) with varying average ambient conditions during one year (Table 4.1) and corresponded pressures and flow rates. Detailed simulation results are collected in Appendix 2.

Table 4.11 - Actual simulation refrigerants compositions

Component	C3MR	DMR		MFC		
	MR	MR1	MR2	MR1	MR2	MR3
Nitrogen	0.0761	0.0000	0.1000	0.0000	0.0000	0.1000
Methane	0.2482	0.0000	0.4000	0.0000	0.0800	0.4000
Ethane	0.4885	0.2700	0.4500	0.3400	0.6100	0.5000
Propane	0.1872	0.4300	0.0500	0.5800	0.3100	0.0000
i-butane	0.0000	0.1900	0.0000	0.0000	0.0000	0.0000
n-butane	0.0000	0.1100	0.0000	0.0800	0.0000	0.0000

The Table 4.12 presents extraction of power requirements for successful working processes from Appendix 2.

Table 4.12 – Compressors average power consumption, MW

Month	C3MR	DMR	MFC
January	118.12	107.57	109.37
February	117.31	106.92	109.09
March	123.24	112.33	115.55
April	130.39	118.92	125.39
May	138.37	128.81	135.33
June	146.16	138.12	142.36
July	151.82	146.53	148.68
August	149.88	142.44	145.75
September	142.68	131.50	137.11
October	132.90	121.00	127.59
November	125.43	113.53	119.00
December	120.49	109.64	111.97

For clear understanding of obtained power consumption results, the diagram in Figure 4.26 was plotted.

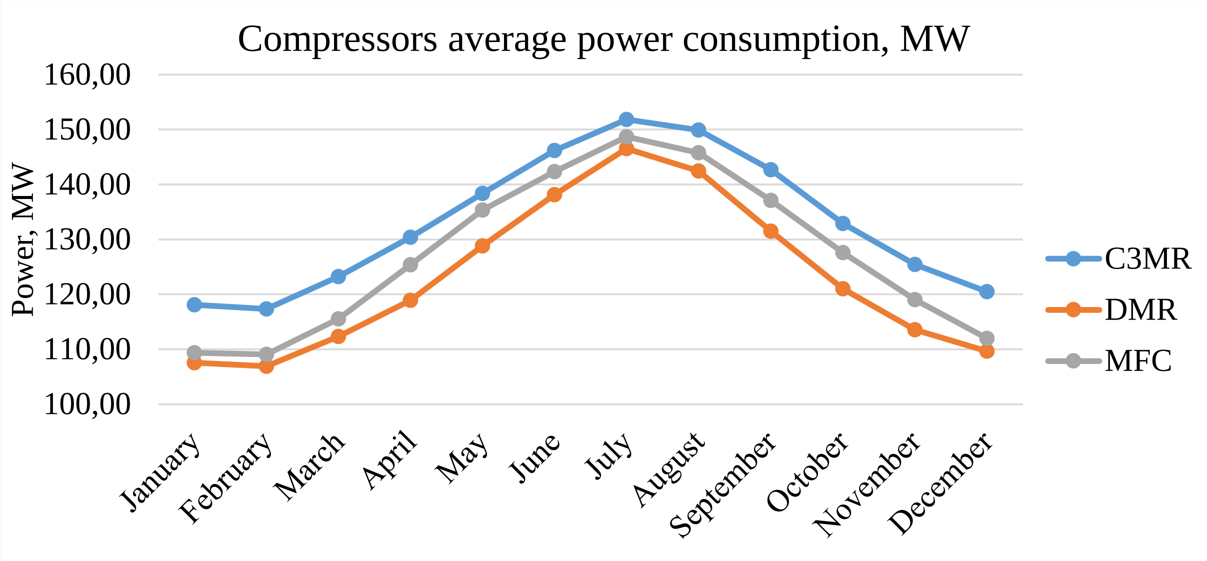


Figure 4.26 - Compressors power variations with temperature fluctuations.

From the Figure 4.26, one can see that power consumption of technologies varies during one year quite intense. Fluctuation for DMR and MFC (ΔP=40MW) are higher than that for C3MR (ΔP=35MW), because of the presence of one-component refrigerant (propane) in it. It can be also observed, that dual mixed refrigerant process presents better results since its diagram lies below the other two. It can be explained by optimal refrigerant composition and correct values of pressure and temperature levels.

Moreover, temperatures between heat exchangers were mostly created on the basis of equal distribution if there were no tips in technologies description [13, 49]. This assumption could cause serious variations in values. However, obtained results belong to expected ranges. The Figure 4.27 below represents the trends of compressors power consumption.

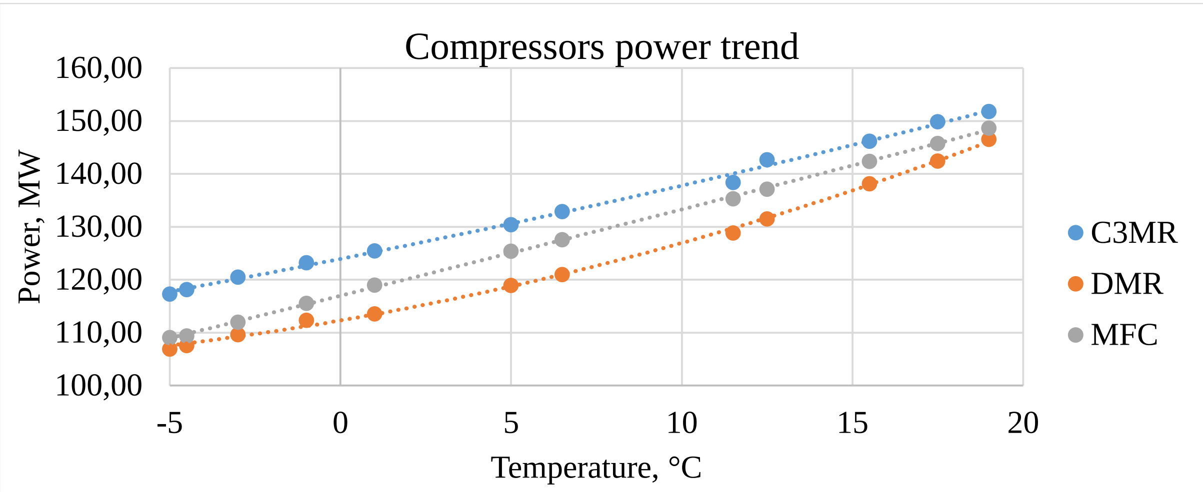


Figure 4.27 – Compressors power consumption trend

From the figure above, we can conclude that more linear relation of power consumption is for mixed fluid cascade and propane precooled mixed refrigerant, while dual mixed refrigerant technology tends to parabolize the trend. Such observation shows that even if DMR and MFC have close values on the borders of simulation temperatures range, midpoints of curves lie quite far from each other, which influences the total energy consumption of the system during the year and makes DMR more attractive process for future development.

Average specific work for each liquefaction technology is mentioned in Table 4.13 below. From this table, we can see, that DMR is the less power consuming process, while C3MR requires more energy. From the literature review, we concluded that average processes specific work belongs to the interval from 12 to 14 kW/TPD. However, from decision matrix we remember, that specific work less than 12kW/TPD assigns score 3, which is the best one. Therefore, each technology considered in this part belongs to this range.

Table 4.13 - Average specific work by the type of process

Process	C3MR	DMR	MFC
Specific work, kW/TPD	9.71	8.99	9.29
Relative specific work	1.08	1.00	1.03

The difference between results from Table 2.4 and Table 4.13 can be explained by the fact that for simulation we assumed natural gas, which does not require fractionation column for NGL extraction. Therefore, it seems that statement "removing fractionation column from the main scheme will increase efficiency" is true.

Table 4.13 also shows numbers of the relative to DMR specific work, which is based on simulation results. It shows that C3MR and MFC are less efficient than DMR process on 8 % and 3%, respectively.

For simulation purposes, LNG production rate during one year was assumed as the constant value, while compressors work was a varying parameter. However, real facility compressors should have constant work, while LNG production varies depending on ambient conditions. Prediction of production rate variation for average specific work of every process is presented in Figure 4.28.

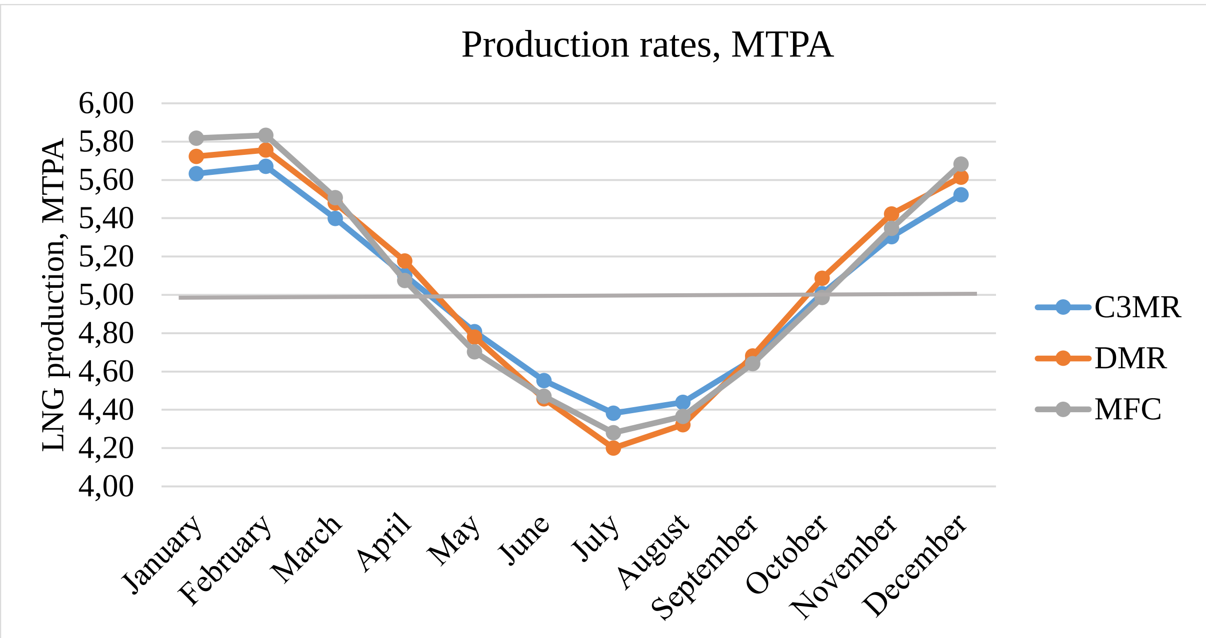


Figure 4.28 - Production rates variation for average specific work of every process

From the figure above, we also can notice that C3MR is quite stable process in comparison with DMR and MFC, giving more LNG production in summer days and less during winter. However, this observation is true in case if we compare processes with their average specific work. If we choose the best process and compare technologies for exact specific work value (8.99kW/TPD for DMR), we will see large variations of production values (Figure 4.29).

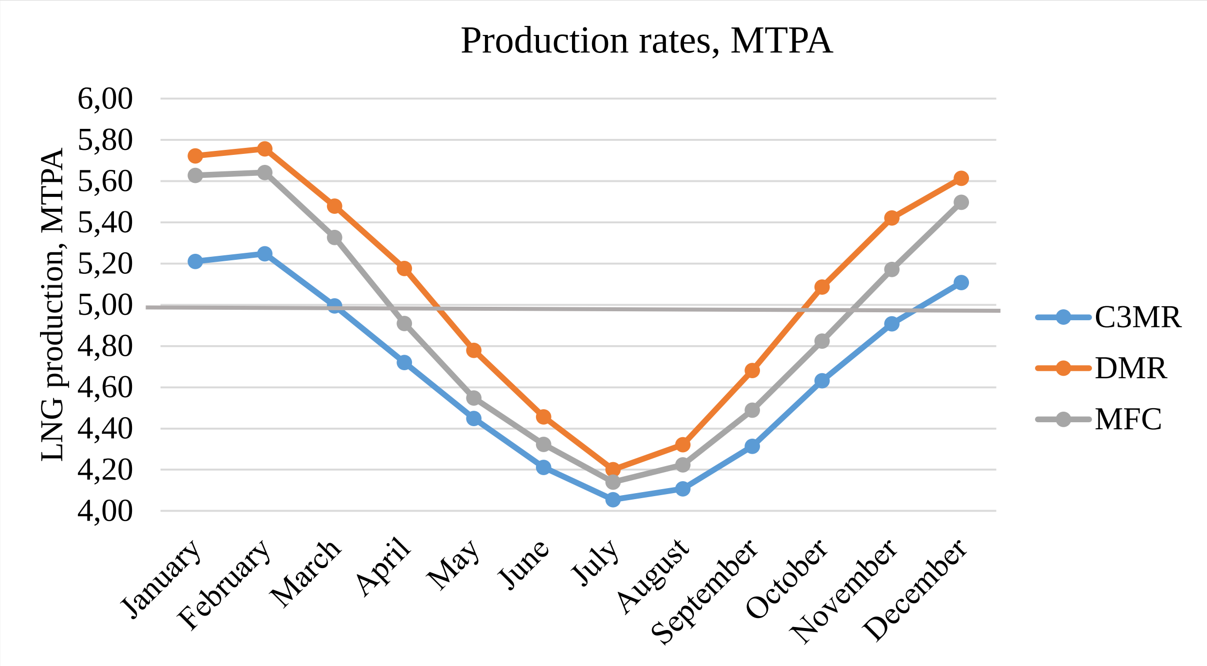


Figure 4.29 - Production values variations for specific work 8.99kW/TPD

The main purpose of simulation procedure was to find less power consuming process, i.e. technology that would produce liquefied natural gas with higher rates. The most power consuming process is C3MR with specific work 9.71kW/TPD, MFC has lower specific work 9.29kW/TPD, while DMR requires the only 8.99kW for the production of one tonne per day. Setting of specific work for every process on the lowest level (DMR specific work result) allowed observing the difference in year LNG production. In this case, average LNG production rate for C3MR differs from DMR on the value 0.4MTPA while MFC rate differs on 0.1MTPA.

Taking into account abovementioned values of production, we can conclude that C3MR technology is not the best decision for chosen conditions. Therefore, appropriate technology must be chosen from two other (Table 4.14). Both processes, MFC and DMR, showed high results in initial decision matrix (Table 3.2) as well as during simulation. It caused mostly by avoiding of pure refrigerants utilisation.

Since the difference between results about DMR and MFC is not clear enough, CAPEX and OPEX were not considered, the decision can be made based on following conclusions.

Taking into consideration technical parameters of the technologies, we can notice that MFC has three refrigerant cycles. DMR, in turn, has two cycles. Additional loop requires additional attention and equipment, such as heat exchanger, technological pipelines and other. Auxiliary equipment requires more space and high investments which reduces the attractiveness of MFC technology. According to Figure 2.9 (Historical trend of liquefaction CAPEX), Snøvit facility has a very high level of CAPEX. This observation can get investor attention.

In additional to above, less power consumption of DMR for the same with MFC LNG production requires fewer expenses for process operation. Summing economics up, we can set the highest score to DMR technology (score 3) and score 2 to MFC (Table 4.14).

Moreover, in accordance with announcing from Gazprom, Baltic LNG facility should have the aim to produce comparatively small amounts of GHGs and other pollutions. This aim cannot be achieved with the utilisation of heavy duty gas turbines only, which means that probably, there will be taken a decision about the application of aero-derivative gas turbines or combine cycles or other options (score 3).

Table 4.14 – Scale of DMR and MFC assessment

№	PARAMETERS	WEIGHT	Shell DMR		MFC	
		(%)	score	total	score	total
1	Economics	15		45.00		30.00
1.1	Investment costs	0.60	3	27.00	2	18.00
1.2	Operating costs	0.40	3	18.00	2	12.00
	Standardization	1.00				
2	Constructability	10		30.00		30.00
2.1	Expandability plant	0.80	3	24.00	3	24.00
2.2	Area required per train	0.20	3	6.00	3	6.00
	Standardization	1.00				
3	Maturity	15		28.50		25.50
3.1	Years of operation	0.30	2	9.00	2	9.00
3.2	Maximum capacity per train set	0.20	2	6.00	2	6.00
3.3	Installed capacity	0.30	1	4.50	1	4.50
3.4	Maximum capacity per train planned	0.20	3	9.00	2	6.00
	Standardization	1.00				
4	Technical	15		33.75		37.50
4.1	Cryogenic heat exchanger type	0.30	1	4.50	2	9.00
4.2	Compressor Type / actuator	0.30	3	13.50	3	13.50
4.3	Specific work	0.10	2	3.00	2	3.00
4.4	Refrigerant type	0.15	3	6.75	3	6.75
4.5	Number of refrigeration cycles	0.05	2	1.50	1	0.75
4.6	Availability of refrigerant	0.10	3	4.50	3	4.50
	Standardization	1.0				
5	CO2 Emissions	5	3	15.00	3	15.00
6	Flexibility gas composition	15	3	45.00	3	45.00
7	Operability/Maintainability	5	1	5.00	1	5.00
8	Commercial flexibility of the licensor	5	2	10.00	2	10.00
9	Domestic Preferences	15		15.00		15.00
9.1	National Content	0.30	2	9.00	2	9.00
9.2	Sustainable Development	0.40	1	6.00	1	6.00
9.3	Partnership	0.30	3	13.50	1	4.50
	Standardization	1.00				
	TOTAL	100.0		227.25		213.00

From the final results in Table 4.14, we observe that evaluation of process specific work, which allowed proposing of economic parameters, shifts the initial balance of matrix parameters (Table 3.2) in favour of DMR technology.

One of the very important parameters is a partnership between countries and companies. Shells DMR project had been realised in Russia in 2009, which means that country has experience in building and operation of this kind of facility, has qualified personnel and requires less spending of money for training [50]. Besides, Gazprom and Shell have strategic cooperation in LNG sector; this fact plays a big role in the selection process.

To sum everything up, the most efficient process, dual mixed refrigerant technology, can be recommended for Leningrad Region Ust-Luga seaport Baltic LNG facility.

4.5 DMR end flash gas utilisation in cold boxes

End flash gas (EFG) is the gas boiled off from LNG tank. It has a very low temperature (-160.7°C). EFG can be compressed from atmospheric pressure to turbine inlet pressure and used directly as fuel, or it can be sent to the cold box to utilise low temperature for increasing efficiency of liquefaction and only after being compressed for becoming turbine fuel. Different turbines have different inlet pressure and fuel rate requirements, which do not belong the research.

With the constant mass flow of LNG q_{LNG}=5.708x10^5kg/h, flash gas flow rate is q_{fg}=0.486x10^5kg/h, which is 8.5% of the amount of LNG production. The flow rate also can be expressed in volumetric units and q_{fg}=161.3m^3/h=2,69m^3/min. The composition of end flash gas is as follows: methane 99.39%, nitrogen 0.6% and ethane 0.01%.

There was a suggestion that EFG would chill SWHE of the second stage, i.e. streams 5, 6 and 7 on the Figure 4.30.

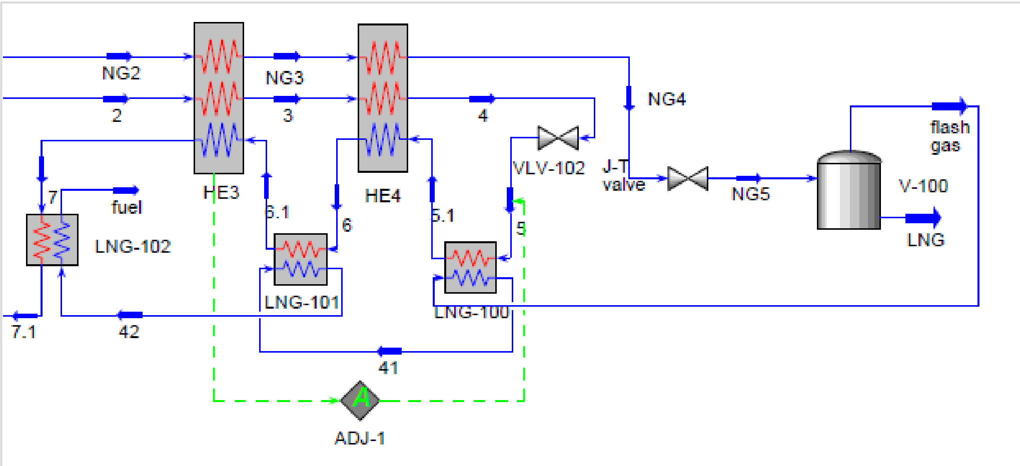


Figure 4.30 – Cold box arrangement in simulation environment.

The temperature difference between inlet and outlet streams from LNG-100, 101 and 102 is mentioned in Table 4.15. Despite small temperature drop provided by cold box, the pressure of the stream 5 increased to rich minimum temperature approach in heat exchanger (HE3). Rising of the stream 5 pressure reduce compressors power consumption and raise system efficiency.

Table 4.15 – Temperature difference within cold box streams

	LNG-100		LNG-101		LNG-102	
Stream#	5	5.1	6	6.1	7	7.1
Temperature	-152.9	-153.0	-134.9	-135.1	-54.7	-56.4

The inlet temperature of the cold phase to heat exchanger HE3 should be -153°C for appropriate refrigeration procedure. The pressure of stream 5 for cold box case is higher on the value Δp=7.5kPa than that for case without cold box, and it equals p_5=536.8kPa. This pressure raises, and also lower temperature on the compressor inlet provides the following compressors power consumption results (Table 4.16).

Table 4.16 - Comparison of cases with and without cold box

Month	DMR compressors work, MW (without cold box)	DMR compressors work, MW (with cold box)
January	107.57	105.68
February	106.92	105.00
March	112.33	110.38
April	118.92	116.99
May	128.81	126.86
June	138.12	135.82
July	146.53	144.71
August	142.44	140.39
September	131.50	129.58
October	121.00	119.05
November	113.53	111.65
December	109.64	107.69

According to table above, the average reduction of compressors, work is 1.96MW. This improvement results in 1.59% reduction of compressors power consumption. Thus with the same power provided to the system, the process with cold box shows better results and higher LNG production rates can be expected.

Conclusion

The purpose of the current study was to develop a methodology of the appropriate LNG technology selection procedure. This aim was achieved by the creation of the decision matrix, which establishes the scale of assessment for base load processes. The literature review showed that weight of parameters in the matrix is not stable value and can vary depending on country preferences and climate conditions. Therefore, the real project was required for filling of weights and scores during scale settlement. Baltic LNG project was chosen because it had not had final investment decision and technology chosen for it had not been declared. This project attracted attention because it could not influence results of research.

Dispelling base load technologies that have minimal LNG production rate more than 5MTPA, the following technologies were taken for comparison: ConocoPhillips optimized cascade (CPOCP); Air Products dual mixed refrigerant (AP-DMR), propane precooled mixed refrigerant (C3MR) and its analogue C3MR/SplitMR; Dual mixed refrigerant process created by Shell (DMR); Axens Liquefin and mixed fluid cascade (MFC) proposed by Statoil-Linde cooperation.

Three technologies have a total score higher than 180 out of the maximum possible 300 points of the ranking matrix (Table 3.2). These technologies are C3MR, Shell DMR and MFC with scores 183.5, 185.0 and 187.5 respectively. Due to the low difference in score results, the chosen processes were subjected to detailed investigation.

Based on data described in many articles, simplified flowsheets with the assumption of deficiency NGL extraction unit were created. The problem connected with the indeterminacy of refrigerants composition was solved with the help of principle of trying and errors in UniSim program. The best refrigerant compositions for liquefaction processes are mentioned in the Table 4.11.

The simulation showed that C3MR and MFC have relative specific work higher than DMR (1.08 and 1.03, respectively). C3MR was rejected from further consideration due to the lowest simulation results and only the third place of the total score in the scale. For the rest two technologies, the new scale of evaluation was created. Economic scores in this matrix were based on specific work results and amount of required equipment estimation. These findings enabled to evaluate investment and operational costs. Also, we considered the possibility to apply advanced gas turbine cycles for DMR. Listed factors affected the total score, shifting it in favour of dual mixed refrigerant technology.

The impact of cold boxes for the DMR process was considered as a factor, improving specific work. The simulation showed that cold box could reduce second stage compressors power consumption on 1.59%.

Summarising everything up, it is possible to conclude that DMR technology has more chances to become the basis of Baltic LNG facility. The following fact appeared very recently, confirms rectitude of the conclusion.

Press release from Gazprom website from 3rd June 2017 [51] states that “Gazprom and Shell have signed two agreements on Baltic LNG project”. Heads of companies signed an agreement about setting up a joint venture at St. Petersburg International Economic Forum 2017. The Joint Study Framework Agreement on the Baltic LNG project was also signed. These documents allow starting the process of designing, construction and operation as well as securing financing and starting developing preliminary project documentation in the near future.

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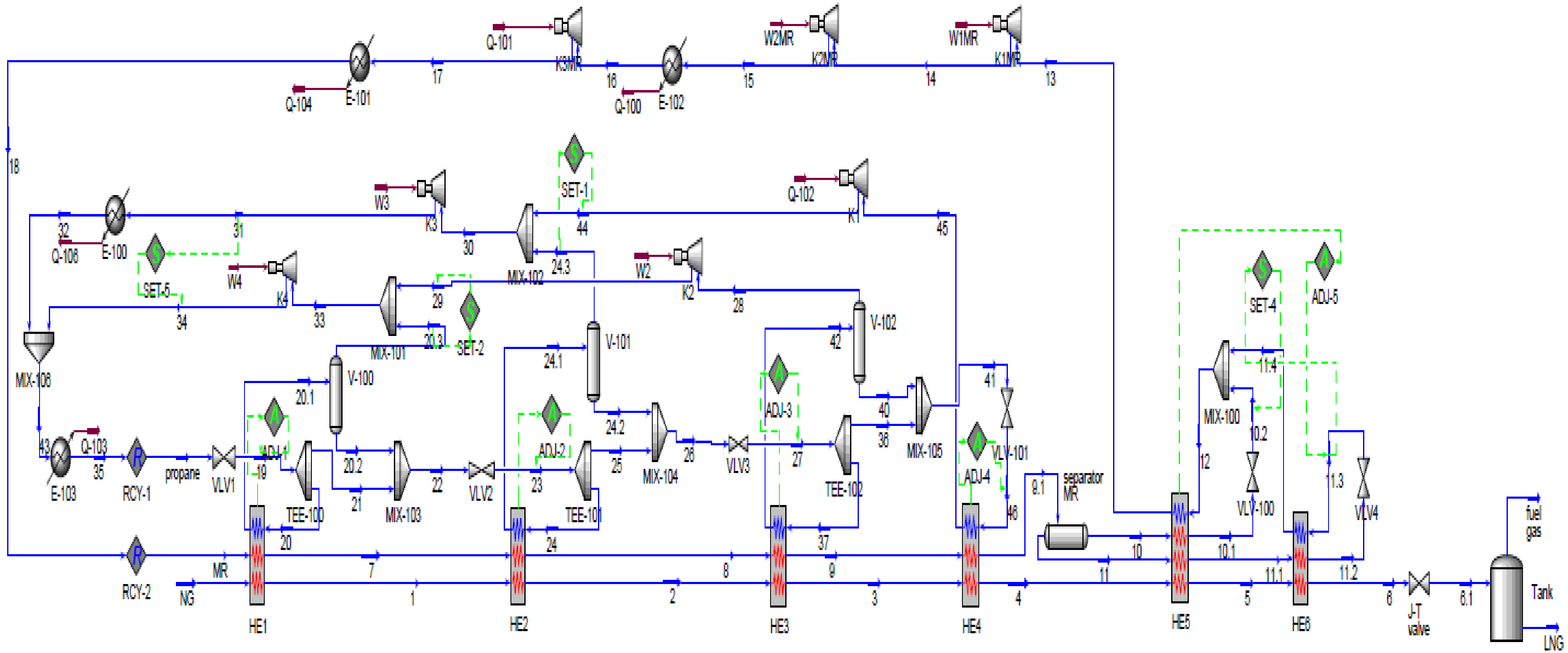
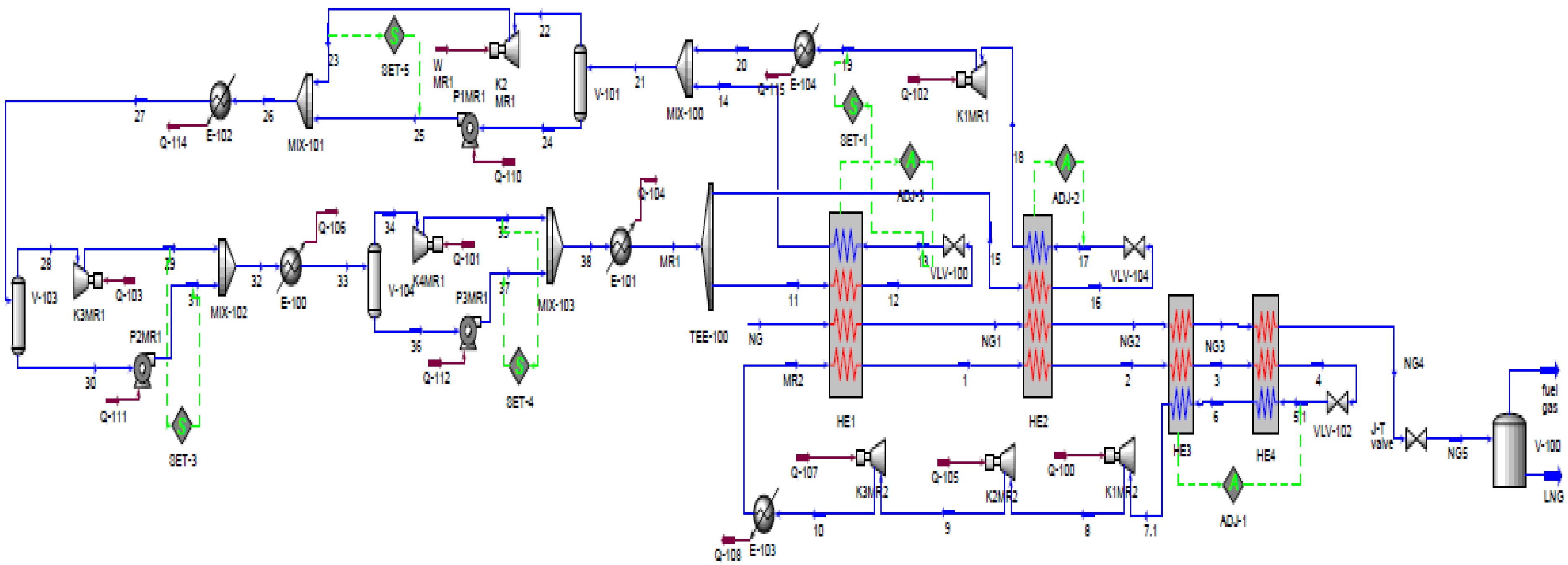
Appendix 1 – Simulation flowsheets

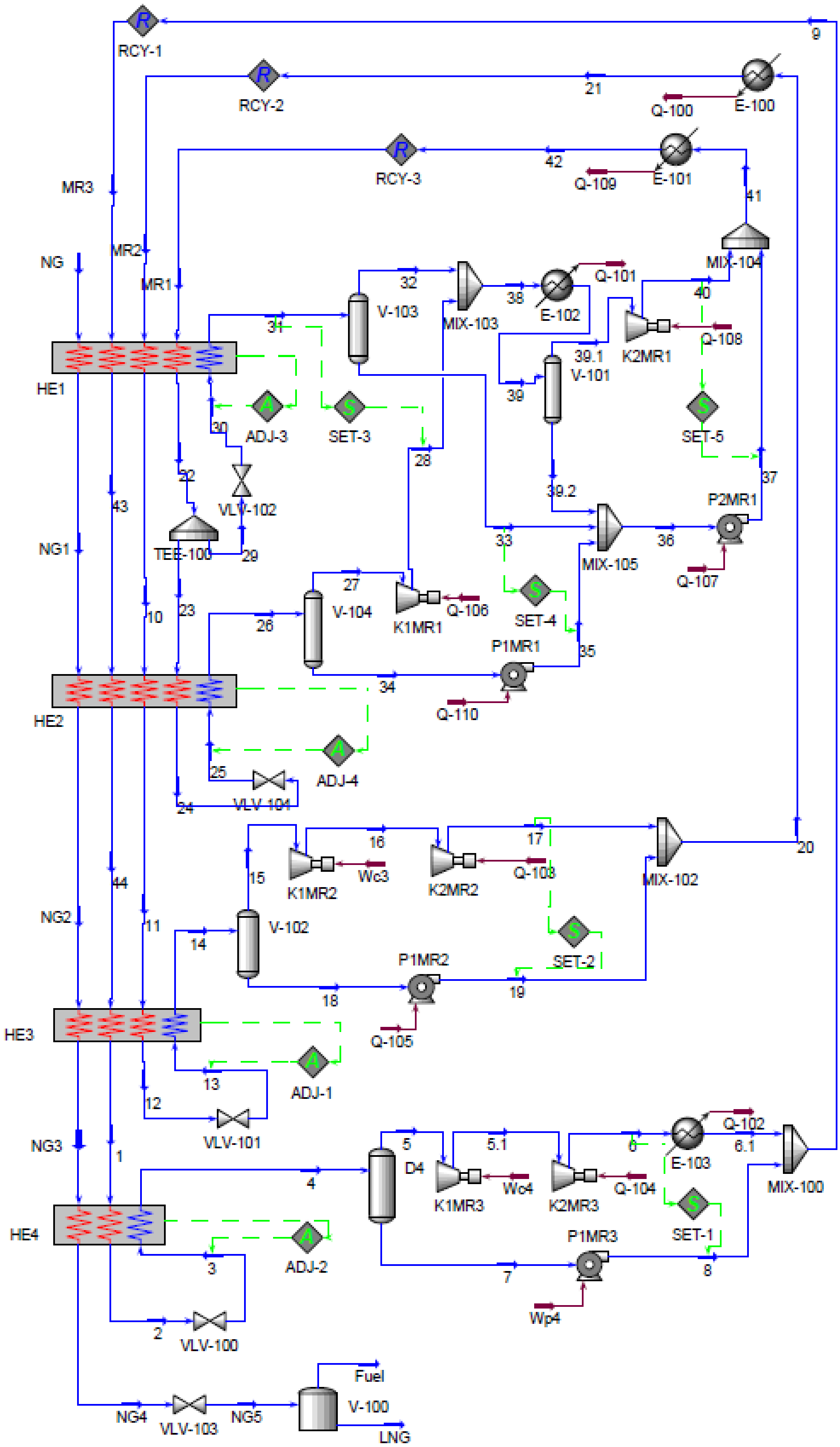
C3MR



DMR



MFC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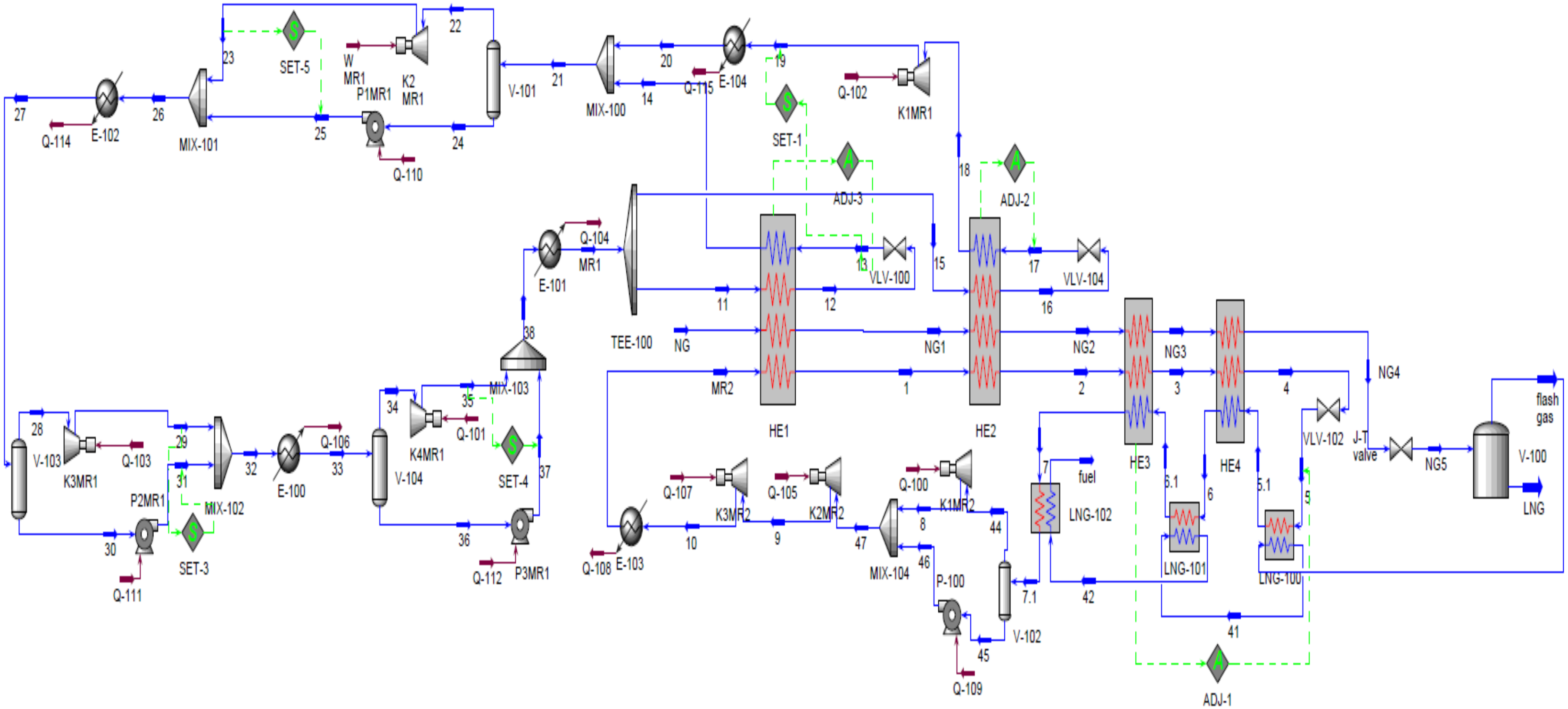
9
RCY-2
21
Q-100
E-100
RCY-3
42
Q-109
E-101
41
MR3
MIX-104
NG
MR2
MR1
32
38
Q-101
40
V-103
MIX-103
E-102
Q-108
31
39.1
K2MR1
HE1
V-101
39
ADJ-3
SET-3
SET-5
30
28
22
37
43
VLV-102
39.2
P2MR1
NG1
TEE-100
29
33
36
MIX-105
Q-107
27
10
23
26
V-104
Q-106
SET-4
K1MR1
P1MR1
35
34
Q-110
HE2
ADJ-4
25
VLV-104
24
17
16
20
K1MR2
Wc3
K2MR2
Q-103
MIX-102
15
44
NG2
11
V-102
14
P1MR2
SET-2
18
19
HE3
Q-105
13
ADJ-1
12
VLV-101
NG3
1
5
5.1
6
Q-102
6.1
D4
E-103
MIX-100
4
K1MR3
Wc4
K2MR3
Q-104
SET-1
HE4
P1MR3
ADJ-2
3
7
8
Wp4
2
VLV-100
Fuel
V-100
NG4
VLV-103
NG5
LNG

DMR with cold box on stage 2



Appendix 2 – Simulation parameters and results

C3MR		Propane				Mixed Refrigerant				Total compressors power, MW	Total power, MW
Month	Temperature, °C	Flow rate, kg/h	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW	Flow rate, kg/h	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW		
January	-4.5	$1.60\cdot10^6$	420	27.06	176.19	$2.2\cdot10^6$	1850	91.07	63.10	118.13	357.42
February	-5	$1.60\cdot10^6$	410	26.32	174.67	$2.2\cdot10^6$	1850	90.99	63.63	117.31	355.61
March	-1	$1.65\cdot10^6$	480	31.69	186.64	$2.2\cdot10^6$	1850	91.55	59.35	123.24	369.23
April	5	$1.80\cdot10^6$	560	38.87	203.92	$2.2\cdot10^6$	1850	91.53	52.26	130.40	386.58
May	11.5	$2.00\cdot10^6$	670	44.71	220.40	$2.2\cdot10^6$	1850	93.66	46.38	138.37	405.15
June	15.5	$2.15\cdot10^6$	750	52.53	234.77	$2.2\cdot10^6$	1850	93.63	41.51	146.16	422.44
July	19	$2.25\cdot10^6$	820	58.19	246.20	$2.2\cdot10^6$	1850	93.63	37.27	151.82	435.29
August	17.5	$2.20\cdot10^6$	800	56.25	241.80	$2.2\cdot10^6$	1850	93.63	39.09	149.88	430.77
September	12.5	$2.05\cdot10^6$	710	48.94	226.22	$2.2\cdot10^6$	1850	93.74	45.25	142.68	414.15
October	6.5	$1.85\cdot10^6$	590	40.12	207.55	$2.2\cdot10^6$	1850	92.78	51.53	132.9	391.98
November	1	$1.70\cdot10^6$	500	33.55	193.63	$2.2\cdot10^6$	1850	91.88	57.26	125.43	376.32
December	-3	$1.60\cdot10^6$	450	29.24	180.88	$2.2\cdot10^6$	1850	91.25	61.48	120.49	362.85

DMR		Mixed Refrigerant 1				Mixed Refrigerant 2				Total compressors power, MW	Total power, MW
Month	Temperature, °C	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW		
January	-4.5	$2.00\cdot10^6$	730	42.56	203.02	$2.10\cdot10^6$	1550	65.01	26.23	107.57	336.82
February	-5	$2.00\cdot10^6$	720	41.91	201.60	$2.10\cdot10^6$	1550	65.01	26.78	106.92	335.30
March	-1	$2.00\cdot10^6$	790	47.28	212.29	$2.10\cdot10^6$	1550	65.05	22.40	112.33	347.02
April	5	$2.00\cdot10^6$	920	53.92	228.28	$2.10\cdot10^6$	1550	65.00	15.74	118.92	362.94
May	11.5	$2.00\cdot10^6$	1070	63.78	249.20	$2.10\cdot10^6$	1550	65.03	8.55	128.81	386.56
June	15.5	$2.20\cdot10^6$	1170	73.06	264.70	$2.10\cdot10^6$	1550	65.06	4.13	138.12	406.95
July	19	$2.40\cdot10^6$	1260	81.57	278.87	$2.10\cdot10^6$	1550	64.96	0.00	146.53	425.40
August	17.5	$2.30\cdot10^6$	1220	77.48	272.27	$2.10\cdot10^6$	1550	64.96	1.81	142.44	416.52
September	12.5	$2.10\cdot10^6$	1090	66.47	253.47	$2.10\cdot10^6$	1550	65.03	7.45	131.50	392.42
October	6.5	$2.00\cdot10^6$	960	55.99	233.64	$2.10\cdot10^6$	1550	65.01	14.08	121.00	368.72
November	1	$2.00\cdot10^6$	830	48.55	217.63	$2.10\cdot10^6$	1550	64.98	20.14	113.53	351.30
December	-3	$2.00\cdot10^6$	760	44.59	207.41	$2.10\cdot10^6$	1550	65.05	24.61	109.64	341.66

MFC		MR 1				MR 2				MR3				Total compressors power, MW	Total power, MW
Month	Temperature, °C	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW		
January	-4.5	$2.8\cdot10^6$	930	29.26	208.01	$1.5\cdot10^6$	760	32.92	8.83	$1.1\cdot10^6$	1400	47.19	14.13	109.37	340.34
February	-5	$2.8\cdot10^6$	930	28.98	207.01	$1.5\cdot10^6$	760	32.92	9.19	$1.1\cdot10^6$	1400	47.19	14.42	109.09	339.71
March	-1	$2.8\cdot10^6$	1030	35.44	220.58	$1.5\cdot10^6$	760	32.92	6.25	$1.1\cdot10^6$	1400	47.19	12.16	115.55	354.54
April	5	$2.8\cdot10^6$	1190	45.33	241.00	$1.5\cdot10^6$	760	32.90	1.78	$1.1\cdot10^6$	1400	47.16	8.74	125.39	376.91
May	11.5	$2.8\cdot10^6$	1380	55.24	259.38	$1.5\cdot10^6$	760	32.92	0.00	$1.1\cdot10^6$	1400	47.16	5.04	135.32	399.74
June	15.5	$3.0\cdot10^6$	1520	62.27	270.50	$1.5\cdot10^6$	760	32.93	0.00	$1.1\cdot10^6$	1400	47.16	2.76	142.36	415.62
July	19	$2.8\cdot10^6$	1640	68.27	280.00	$1.5\cdot10^6$	760	32.93	0.00	$1.1\cdot10^6$	1400	47.48	1.08	148.68	429.76
August	17.5	$2.8\cdot10^6$	1580	65.68	275.90	$1.5\cdot10^6$	760	32.91	0.00	$1.1\cdot10^6$	1400	47.16	1.61	145.75	423.26
September	12.5	$2.8\cdot10^6$	1420	56.79	261.89	$1.5\cdot10^6$	760	32.91	0.00	$1.1\cdot10^6$	1400	47.16	4.47	136.86	403.22
October	6.5	$2.8\cdot10^6$	1230	47.51	245.86	$1.5\cdot10^6$	760	32.92	0.68	$1.1\cdot10^6$	1400	47.16	7.89	127.59	382.02
November	1	$2.8\cdot10^6$	1080	38.93	227.16	$1.5\cdot10^6$	760	32.91	4.75	$1.1\cdot10^6$	1400	47.16	11.00	119.00	361.91
December	-3	$2.8\cdot10^6$	980	31.91	213.45	$1.5\cdot10^6$	760	32.91	7.71	$1.1\cdot10^6$	1400	47.16	13.26	111.98	346.40

DMR with cold box		Mixed Refrigerant 1				Mixed Refrigerant 2				Total compressors power, MW	Total power, MW
Month	Temperature, °C	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW	Flow rate, kg/s	Inlet pressure, kPa	Compressors power, MW	Coolers duty, MW		
January	-4.5	$2.00 \cdot 10^6$	730	42.56	203.02	$2.10 \cdot 10^6$	1550	63.12	21.46	105.68	330.16
February	-5	$2.00 \cdot 10^6$	720	41.91	201.60	$2.10 \cdot 10^6$	1550	63.09	21.97	105.00	328.57
March	-1	$2.00 \cdot 10^6$	790	47.28	212.29	$2.10 \cdot 10^6$	1550	63.10	17.57	110.38	340.24
April	5	$2.00 \cdot 10^6$	920	53.92	228.28	$2.10 \cdot 10^6$	1550	63.07	10.91	116.99	356.18
May	11.5	$2.00 \cdot 10^6$	1070	63.78	249.20	$2.10 \cdot 10^6$	1550	63.08	3.72	126.86	379.78
June	15.5	$2.20 \cdot 10^6$	1170	72.71	263.95	$2.10 \cdot 10^6$	1550	63.11	0.00	135.82	399.77
July	19	$2.40 \cdot 10^6$	1260	81.57	278.87	$2.10 \cdot 10^6$	1550	63.14	0.00	144.71	423.58
August	17.5	$2.30 \cdot 10^6$	1220	77.29	269.13	$2.10 \cdot 10^6$	1550	63.10	0.00	140.39	409.52
September	12.5	$2.10 \cdot 10^6$	1090	66.47	253.47	$2.10 \cdot 10^6$	1550	63.09	2.61	129.58	385.64
October	6.5	$2.00 \cdot 10^6$	960	55.99	233.64	$2.10 \cdot 10^6$	1550	63.06	9.24	119.05	361.93
November	1	$2.00 \cdot 10^6$	830	48.55	217.63	$2.10 \cdot 10^6$	1550	63.10	15.36	111.65	344.64
December	-3	$2.00 \cdot 10^6$	760	44.59	207.41	$2.10 \cdot 10^6$	1550	63.10	19.78	107.69	334.88