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LNG PRODUCTION TECHNOLOGIES AND PROCESS SIMULATION.

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"The MSc Program in Petroleum Engineering of the Technical University of Crete was attended and completed by Mrs. Efsevia Fragkou due to the Hellenic Petroleum Group Scholarship award."

Abstract

The main goal of this master thesis is to present the refrigeration cycles used in the industry of LNG and simulate the most widely used for the liquefaction of natural gas. In this case, a comparison between the different refrigeration cycles was contacted and C3MR liquefaction process was selected to be modelled. The simulation was based on ***Air Products and Chemicals Inc.*** (APCI) technology as well as Swenson's patent from 1977, 'SINGLE MIXED REFRIGERANT, CLOSED LOOP PROCESS FOR LIQUEFYING NATURAL GAS'.

Simulation was designed for a large scale project of 3 MTPA production of LNG and was conducted in Aspen HYSYS, a software widely used in the oil and gas industry, in collaboration with ASPROFOS Engineering that provided a license for the software. The outcome of the simulation was used to conduct a cost analysis of the liquefaction unit. The cost of each unit was compared to the total equipment cost and the conditions of each unit are given in detail, with compressors turning out to be the most expensive part of equipment, accounting for almost 81% of the total equipment cost.

For concluding on the sustainability of the project, a financial analysis was also conducted over a thirty-year period and the results were very promising. A project of this magnitude turned out to be very efficient and profitable, with a payback period, only for the liquefaction unit and not the whole liquefaction plant, being less than a year.

Περίληψη

Ο κύριος σκοπός της παρούσας μεταπτυχιακής διατριβής είναι να παρουσιάσει όλες τις μεθόδους υγροποίησης που χρησιμοποιούνται στη βιομηχανία για την παραγωγή υγροποιημένου φυσικού αερίου και να προσομοιώσει τη μέθοδο που χρησιμοποιείται στην πλειοψηφία των περιπτώσεων. Στη συγκεκριμένη περίπτωση, η μέθοδος που επιλέχθηκε για προσομοίωση ήταν η διαδικασία κλειστού βρόχου με τη χρήση ενός μικτού ψυκτικού (πατέντα US4033735A) και πιο συγκεκριμένα η τεχνολογία που αναπτύχθηκε από την ***Air Products and Chemicals Inc.*** (APCI), C3MR, που χρησιμοποιεί προπάνιο για την προκαταρκτική ψύξη του ψυκτικού υγρού αλλά και του φυσικού αερίου.

Η προσομοίωση βασίστηκε στο σχεδιασμό μίας μεγάλης μονάδας παραγωγής υγροποιημένου αερίου, συνολικής χωρητικότητας τριών εκατομμυρίων τόνων ετησίως και διεξήχθη στο λογισμικό της Aspen, HYSYS, ένα λογισμικό που χρησιμοποιείται ευρέως στη βιομηχανία για το σχεδιασμό εργοστασίων. Η άδεια χρήσης του προγράμματος δόθηκε από την ASPROFOS Engineering A.E., μία εταιρεία του ομίλου ΕΛΛΗΝΙΚΑ ΠΕΤΡΕΛΑΙΑ (ΕΛΠΕ), η οποία παρέχει εξειδικευμένες υπηρεσίες σε έργα Πετρελαίου, Φυσικού Αερίου, Πετροχημικών και Υποδομών. Το αποτέλεσμα της προσομοίωσης χρησιμοποιήθηκε για την εκτίμηση του κόστους της μονάδας υγροποίησης. Το κόστος και οι συνθήκες του επιμέρους εξοπλισμού παρατίθενται αναλυτικά και γίνεται μία σχετική σύγκριση μεταξύ των μονάδων από την οποία προκύπτει ότι το μεγαλύτερο κόστος σε μία μονάδα υγροποίησης αφορά την εγκατάσταση και λειτουργία των συμπιεστών (σχεδόν το 81% του συνολικού κόστους του εξοπλισμού).

Προκειμένου να υπάρξει μία καταληκτική απόφαση για τη βιωσιμότητα της μονάδας, μία βασική τεχνοοικονομική ανάλυση διεξήχθη, για λειτουργία τουλάχιστον τριάντα χρόνων και τα αποτελέσματα ήταν ιδιαίτερα ικανοποιητικά. Το μέγεθος του έργου αποδείχθηκε επικερδές, με χρόνο αποπληρωμής λιγότερο από ένα χρόνο μόνο για τη μονάδα υγροποίησης και όχι για ολόκληρη την εγκατάσταση εργοστασίου υγροποίησης φυσικού αερίου.

Preface

This Master thesis was completed at the Mineral Resources Engineering Department of the Technical University of Crete (TUC), from June 2018 to January 2019, and was part of the Petroleum Engineering MSc Program due to Hellenic Petroleum Group Scholarship award.

First, I would like to thank my supervisor, Dr. Dimitris Marinakis, for the subject he commissioned and for his guidance and patience throughout this whole.

A special thank you is to the Hellenic Petroleum Group for the Scholarship they awarder me and the chance they gave me to come in contact with my object of interest, the Oil and Gas Industry. I would also like to thank Asprofos Engineering, for granting me access to the Aspen HYSYS software program and their guidelines on the simulation part.

Finally, I would like to thank my friends and family for their support all this time, which kept me strong until the completion of the MSc Program.

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Nomenclature table

APCI: Air Products and Chemicals Inc.

BAHX: Brazed Aluminum Heat Exchanger

bcm/y: billion cubic meters of gas per year

BOG: Boil-off Gas

Btu/m^3: British thermal units per cubic meters

Btu/scf: British thermal units per standard cubic feet

GCU: Gas Combustion Unit

cf: cubic feet

CNG: Compressed Natural Gas

GSU: Gas Sweetening Unit

DFDE: Dual fuel diesel electric

DMR: Dual-Mixed Refrigerant

EU: European Union

FLNG: Floating Liquefied Natural Gas

FSRU: Floating Storage Regasification Unit

HP: High Pressure

HP-MR: High Pressure Mixed Refrigerant

ITG: Interconnector Turkey-Greece

J-T: Joule-Thomson

LMR: Liquefaction Mixed Refrigerant

LNG: Liquefied Natural Gas

LP: Low Pressure

LPG: Liquefied Petroleum Gas

MCHE: Main Cryogenic Heat Exchanger

MFCP: Mixed Fluid Cascade Process

MMBtu: Million Metric British Thermal Units

MMTPA: Million Metric Tonnes per Annum

MR: Mixed Refrigerant

Mt/a: mega/million tons per annum

MTPA: Mega Tonnes per Annum

MW: Molecular Weight

MW: Mega Watt

NG: Natural Gas

NGL: Natural Gas Liquids

Nm^3: Normal Cubic Meters. Natural Gas at conditions of absolute pressure 1.01325 bar and 0 ^oC, occupies volume of one cubic meter.

NPV: Net Present Value

PBP: Pay Back Period

PFHE: Plate Fin Heat Exchanger

PMR: Precooling Mixed Refrigerant

PNG: Papua New Guinea

POCLP: Phillips Optimized Cascade Liquefaction Process

POT: Pay-out time

PPMR or C3MR: Propane pre-cooled single mixed-refrigerant

ppmv: parts per million by volume

PRICO®: Poly Refrigerant Integrated Cycle Operation

ROR: Rate of Return

SMR: Single Mixed Refrigerant

SMR: Sub-cooling Mixed Refrigerant

SRU: Sulfur Recovery Unit

SWHE: Spiral-Wound Heat Exchanges

TGTU: Tail Gas Treating Unit

tpy: tonnes per year

UGS: Union of Greek Shipowners

US: United States

USD: US dollars

USD/MMBtu or **$/MMBtu**: US dollars per million British thermal units

USD/tpa or **$/tpa**: US dollars per tonnes per annum

WCHE: Wound Coil Heat Exchanger

WI: Wobbe Index

A. Liquefied Natural Gas (LNG)

A.1. Liquefied Natural Gas

The growing demand in energy worldwide, as well as the need to use sustainable and clean energy to protect the environment, have led to new era, where renewable energy sources and fuels with low emissions are desirable. One industry that is growing rapidly the past few decades to meet this era, is the Natural Gas Industry, and more specifically, the Liquefaction of Natural Gas (LNG) Industry. Natural gas is a mixture, which consists of gaseous hydrocarbons, primarily methane at varying concentrations (usually between 87-97%), ethane (1.5-9%), propane (0.1-1.53%), iso-butane and n-butane (0.01-0.03%), and at smaller amounts higher hydrocarbons (C_{5+} <0.06%) , carbon dioxide (0.05-1.0%), nitrogen (0.2-5.5%), hydrogen sulfide (0.0-0.3%) and even helium (0.0-0.05%) [1,2]. It occurred after millions of years of transformation of organic matter, which was buried over the time and was exposed to the right conditions of temperature and pressure. Organic matter is mainly composed of animal and plant matter. Plants are capable of storing solar energy (photosynthesis), which is recycled through herbivore animals to finally end up to the Earth as decomposing organic matter, which will eventually form fossil fuels. Out of all fossil fuels, natural gas is the cleanest due its low emissions when burnt. Combustion of natural gas primarily emits water vapors and carbon dioxide (CO_2). The CO_2 emissions are 30% lower than fuel oil, 45% less than coal and very low sulphur dioxide (SO_2) emissions [3].

LNG is natural gas that has been cooled down to cryogenic temperatures in order to take a liquid form. This is a desirable process, since LNG takes up less space when stored and is easier to be transported. As the natural gas, LNG has no odor or color, it is non-toxic and non-corrosive. However, it can be flammable if vaporized and even explosive if present in a mixture with oxygen and air to a 5.3–15% concentration.

The natural gas before enters the liquefaction process, needs to be brought at a certain composition which is essential for the quality of the LNG. This process involves the removal of impurities, acidic gases after passing through a sweetening unit, water and higher hydrocarbons from a fractionation unit, which are collected and sold as separate products. The natural gas is then condensed into a liquid at approximately −162 °C (−260 °F) and atmospheric pressure.

The natural gas is transported to the final consumers as LNG (30%) and the rest 70% through pipelines. Natural gas is cooled down to liquid form (LNG) for easy and safe transportation and storage at atmospheric pressure. The transformation of natural gas into LNG has great advantages for transportation overseas where pipeline transportation is not feasible or is economically unattractive. When compared to other methods, the volume reduction during the liquefaction is much greater than that of the compressed natural gas (CNG), the volumetric energy density of LNG is 2.4 times greater than that of CNG or 60% that of diesel fuel [4]. LNG takes up about 1/600th of the volume of natural gas in the gaseous state at standard conditions. This property makes LNG cost efficient in marine transportation over long distances. Once the LNG is transported to the desired station, it is regasified and distributed with pipelines or trucks to remote places where pipeline systems cannot reach.

The LNG industry is quite extensive: it involves upstream facilities, LNG production plants, shipping, storage and regasification, in order to be distributed to the consumers. LNG production which is the main focus of this diploma thesis, is capital intensive, hence, a plant overview and cost analysis need to be evaluated before proceeding with the construction of such a project.

A.2. LNG origins and history in the market place

A.2.1. LNG History Worldwide

The liquefaction process dates back almost 2 eons, to the early 1820s, when Michael Faraday first experimented and achieved the liquefaction of gases with high critical temperatures. Half a century later, his pioneering work was used to liquify gases with lower critical temperatures, and it was physicist Louis Paul Cailletet who first liquefied methane by discovering that marsh gas liquefied at 180 atm and 44.6 ^{o}F in 1878 and the chemist Raoul Pierre Pictet who developed a cascade cooling apparatus for the liquefaction of gases. By the end of the 19th century, Carl von Linde, a pioneer of refrigeration, built the first air liquefaction plant in 1895, utilizing counter current cooling. Godfrey Cabot in 1915 patented a method for storing liquid gases at very low temperatures. His patent consisted of a thermos like design; cold inner tank within a separated outer tank for better insulation. Cabot also founded the ***Liquid Fuel Company*** in West Virginia to use LNG for welding, but it was abandoned as soon as 1921. Following Cabot's footsteps, Lee Twomey in 1937 took LNG to large scale, aspiring to use it in peak shaving plants (plants that store excess LNG to save energy according to seasonal demand). ***East Ohio Gas Company*** in 1940 built the first full-scale LNG plant which operated for three years until it ceased due to a major spilling accident.

After this incident, LNG facilities didn't bloom for a while and attention was given in the research of new and improved ways of insulation and refrigeration methods. It was until 1951, when ***Union Stockyard Company***, imported LNG for power generation. The company was trendsetter when commissioned the first Floating Liquefied Natural Gas unit (FLNG); a floating unit which liquefies natural gas offshore. ***ConocoPhillips*** advised this endeavor which ended with the creation of a barge, the ***Methane*** and the establishment of a new corporation between the two companies (***Constock Liquid Methane Corporation)***, for trans-ocean shipments of LNG from the gas fields of South America to Europe. The first tanker to manage trans-ocean transportation of LNG was named ***Methane Pioneer*** and was of 5,000 m^3 total capacity.

Venezuela was a growing gas market in the 1960s. Algeria followed those steps in 1961 forming the ***Companie Algerienne du Methane*** (CAMEL) followed by the construction of Arzew LNG plant of 1 MTPA capacity. The total cost of the CAMEL LNG project was estimated to be 89×10^6 USD and the price of the delivered product was 0.76 USD/MMBtu, which was considered high priced for the time and indicated the large transportation cost, especially when compared to the present where transportation cost is much reduced due to evolved technologies.

Europe first entered the LNG market by importing LNG from Venezuela to Britain and later from Algeria to France in 1965. The same year, the Algerian government took a 20% stake of CAMEL and by 1977, the whole company was acquired by a government company, ***Sonatrach***.

In 1962, natural gas was discovered in Alaska by ***Phillips***, in the Kenai region and the following year, ***Shell*** discovered a gas field off Brunei. Both companies aspired to open the LNG market in Asia, by transporting LNG to Japan, which was finally achieved in 1969, at a delivered LNG price of 0.52 USD/MMBtu. ***Bechtel*** was the one to finally construct a two-train, 1.3 MTPA liquefaction plant in Kenai, at a total cost of 200×10^6 USD, while Brunei's LNG came on stream in 1972 and the plant originally was of 3.7 MTPA capacity.

Libya's gas reserves were exploited by ***Exxon***, which constructed an LNG plant, Marsa el Brega in 1970, that exported LNG to Italy and Spain. The plant operated with a new liquefaction process developed by ***Air Products***, which took advantage of the cooling curves of natural gas and made a Mixed Refrigerant with specific composition to approach these curves and maximize the heat transfer.

In 1972, a second Algerian LNG plant started operation (Skikda) as well as the Brunei LNG plant. The most important fact of the 1970s, was when four oil and gas industries (***Pertamina***, ***Huffco***, ***Mobil*** and ***Nissho Iwai***) joined to develop the Indonesian LNG production, by constructing two plants (Arun and Bontang) that would have the ability to export. This venture was successful and put Indonesia on the top of LNG exporters for at least two decades to follow. The total cost of both plants, constructed by ***Bechtel***, raised to 1.64×10^9 USD. Bontang LNG plant was of 3.3 MTPA capacity and Arun was of 5.1 MTPA.

Middle East's first LNG plant, *ADGAS*, was built in Abu Dhabi in 1973. By that time, there were nine operating LNG plants worldwide of around 33 Mtpa total capacity. By 1979, Arzew plant had reached a total of 8.9 MTPA and Algeria exported 8.5×10^6 tonnes of LNG. Global sales had reached 24.7×10^6 tonnes, out of which the 14 were exported to Japan and 5.3 were exported to the USA market. Marine transportation of LNG was carried out by 52 ships until the late 1970s.

America did not permit power generation by fuel gas, a restriction that affected the gas production industry and the common belief that gas is in shortage. A lot of companies started importing LNG, with six long-term contracts for LNG supply being signed till the mid-1970s and three new receiving terminals built. As it turned out, LNG imports became uncompetitive due to price deregulation, stimulated domestic production and a fall in the gas prices. Imports of LNG fell from 5.3 to less than 2×10^6 tonnes in 1980 and finally reached zero by 1987. Algeria was affected significantly because a lot of US buyers pursued a termination of their contracts and caused the two LNG plants to operate at 50% of their capacity for more than ten years.

Besides Brunei, ***Shell*** discovered in 1969 substantial gas reserves in Malaysia. Bureaucratic issues led to exploitation of the field a decade later. It was in 1978, when ***Malaysia LNG Sdn Bhd*** was incorporated to develop the first Malaysian LNG project, ***MLNG Satu***, which included 5 LNG carriers. ***MLNG Satu's*** total design capacity was set at 6 Mtpa and the LNG produced was mainly exported to Japan.

Australia also faced bureaucratic problems when attempted to enter the LNG market for the first time, with the North Rankin gas in 1971. Japanese economy and high competition by that time, made it difficult to secure commitments for exploiting and exporting. The first SPA (sales and purchase agreement) was signed in 1982 when the North West Shelf venture begun. The plant at Dampier, started operation in 1989 of a total capacity of 7.5 MTPA. The project cost 25×10^9 USD. The extremely high cost of this project was due to the construction of the largest gas platform (North Rankin A) and the subsea infrastructure that draw gas from two different fields (N. Rankin and Perseus). By that time, this was the largest engineering project in the world. The plant reached a total capacity of 16.3 MTPA by 2008 with the addition of more trains in the liquefaction process.

By 1989, 10 liquefaction plants were operating in Algeria, Abu Dhabi, Australia, Brunei, Indonesia, Libya, Malaysia and the USA. Exports had reached 48×10^6 tonnes, Japan imported 32×10^6 tonnes and the second largest importer was France at 7×10^6 tonnes. LNG fleet had grown substantially, reaching a total of 70 vessels globally.

In the 1990's, the LNG market grew exponentially, as new entries joined the industry. Qatar, Oman, Nigeria and Trinidad set a new day for the LNG industry. In the 1980's it was very difficult to conclude and sign SPAs, yet in the 1990's, a greater demand for LNG made things easier. ***Qatar Liquefied Natural Gas Company Ltd*** **(*Qatargas*)** secured its first SPA in 1992 and started the construction of an LNG plant to Chiyoda in 1993. Qatargas plant had a capacity of 10 Mtpa and finally came on stream in 1997. Another LNG plant, the RasGas with 6.7 MTPA capacity, started operation at the end of the millennium and costed 3.26×10^9 USD including the upstream facilities. During 1990's Qatar bloomed extremely fast, and as a result, by 2006 it overtook Indonesia and became the largest LNG producer of total producing capacity 77 MTPA.

Nigeria started a two-train LNG plant in 1999 of 6.6 MTPA capacity, which by 2007 came to a total capacity of 21.1 MTPA after the addition of three more trains. The same year Trinidad LNG started operation and Oman LNG followed the next year. The total number of projects during the 1990's reached the ten, with an addition of 55 MTPA in the LNG production. This number was to be overtaken by the 2000's, where 17 projects took place in 14 countries, and blooming plant construction gave 160 MTPA of new capacity. US by that time has a great demand in natural gas and needs to import what is missing as LNG and countries like South America (Peru), Russia (Sakhalin), Norway (Snohvit) are now entering the scene.

The early 2000's, the total of LNG receiving terminals had reached 31 worldwide and the following years about ten new liquefaction projects were to start in Europe, Africa, Asia and South America, most of which would export LNG to America. Even though the US spent quite an amount of money during this period to open and build in total 14 LNG import terminals, most of them were never used since shale gas made its entrance to the gas market and America started exploiting it. By 2010 global sales of LNG had reached 193×10^6 tonnes from 24 projects in 18 countries. The largest importer in 2010 was Japan (69 MTPA), followed by Korea (33 MTPA), Spain (20 MTPA) and the UK (14 MTPA). India started importing LNG in 2003 and China in 2006 and by 2010 imports into China had reached 9.5 MTPA.

BP had discovered gas in the Hides Field in Papua New Guinea (PNG) in 1987 and proposed an LNG project which was developed only after twenty years. The PNG LNG plant was constructed and started operation in 2014 with a production capacity of 6.9 MTPA. About the same time, two more LNG plants, one in Australia and one in Angola, came on stream, each of about 5 MTPA capacity.

Relatively new technologies for extracting shale gas, led to an excess of gas production in the market. This growth in the produced gas caused a growth in the need of LNG receiving terminals and new projects counting to the day are in process of construction.

Floating storage has proved to be much cheaper than conventional onshore storage terminals and can be deployed in months rather than the 3-4 years it takes to build a conventional onshore terminal. 22 FSRUs (Floating Storage Regasification Units) are now in operation around the world, and one is expected to start operation by 2020 in Alexandroupolis, Greece. After floating storage, floating liquefaction plants (FLNG) follow, that are going to set a whole new era in the LNG industry. The first floating liquefaction project, *Prelude*, by ***Shell*** in 2011 and many more FLNG projects being under consideration or construction until the present. The *Prelude* FLNG unit is the largest vessel ever built and is designed to produce 3.6 MTPA of LNG, 1.3 MTPA of condensate and 0.4 MTPA of LPG [5].

At a worldwide scale, the LNG trade ranges between Europe that accounts for 1/4th of the global LNG market, Asia for about 1/3rd and the US for about 1/15th. Asia is a major power in the LNG market, with Japan and South Korea being in the center [6].

LNG has met great recognition because it offers alternatives when compared to other fuels, it is cheaper, it is cleaner and it can be used in power plant units, public and marine transportation. If used for transportation of this magnitude, LNG could reduce the energy cost as much as 30% comparing with the diesel that is widely used at the moment. Europe and North America have focused on LNG as a marine fuel to help reduce the sulphur levels, China aspires to use it on the truck market, Indonesia has focused on remote power and mining plants and Australia on mining and road train operators. The major drawback is natural gas's inelastic demand, which is due to the difficulty in altering to another form of energy, since natural gas does not have a convenient substitute. Most of the heavy machinery, means of transportation and industry use specific forms of energy and altering to another form demands lot of time and money. The inelastic demand also comes from the fact that countries and consumers that have contracts for importing LNG, are bound to use natural gas for specific periods of time and store the excess which is not always feasible.

LNG has become a global market, growing at a very fast pace and natural gas is expected to be the fuel of interest for the years to follow. LNG seems to be leading the way and every country has plans for building LNG terminals. The first exports from an FLNG project commenced in 2017 and several more FLNG projects are expected to follow with a discussed total capacity of 875 MTPA [5,7].

Global Liquefaction Capacity

Global liquefaction capacity has increased rapidly in the last two decades as discussed already and it is shown in the Figure below. From 2000 to the present, capacity has grown exponentially worldwide, exceeding a total of 350 MTPA.

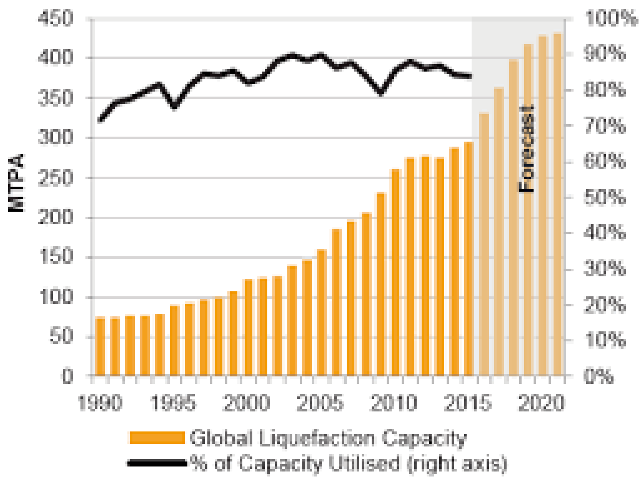

FIGURE 1: Global Liquefaction capacity Build-Out, 1990–2023 [7].

Up to date, there are 19 countries of significant liquefaction capacity, with Qatar being the major producer with 77 MTPA of liquefaction capacity (1/5th of the world's total). A distribution on liquefaction capacity per country is presented in Figure 2.

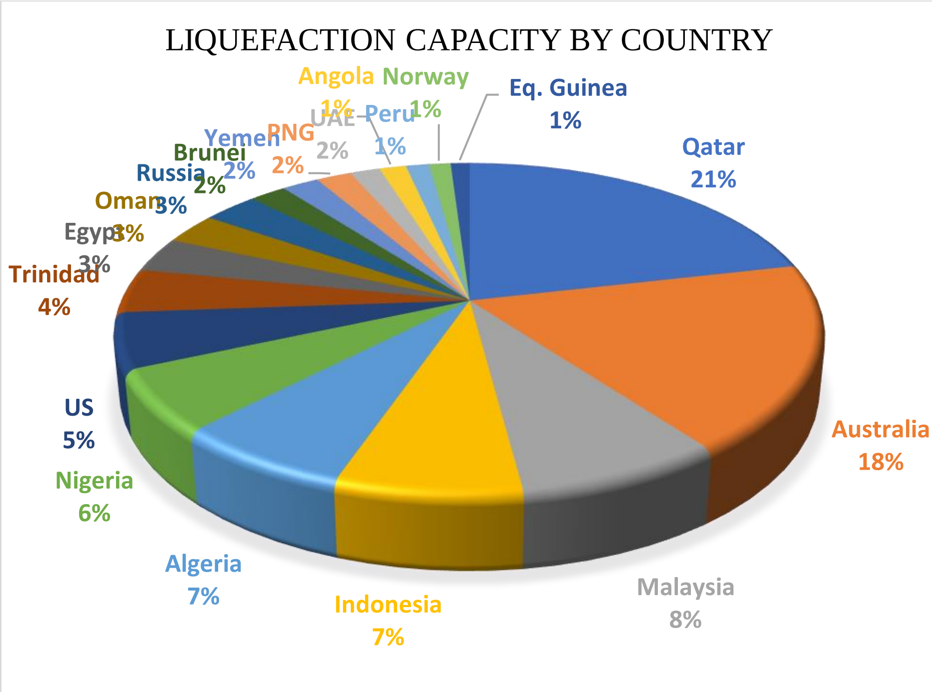

FIGURE 2: Liquefaction capacity by country as of March 2018.

Global liquefaction capacity grows dramatically the last couple of years. US has the biggest share in liquefaction capacity, followed by Australia and Russia. Australia is adding a number of trains to increase its liquefaction capacity, and when completed, is expected to surpass Qatar's total liquefaction capacity.

TABLE 1: Expected growth in liquefaction capacity by country as of March 2018.

Country	Expected Growth in Liquefaction Capacity (in MTPA)
US	48.6
Australia	17
Russia	13.7
Indonesia	4.3
Mozambique	3.4
Malaysia	2.7
Cameroon	2.4
Total	**92.1**

A.2.2. LNG History in Greece

Natural Gas met the Greek market for the first time at the late 1990's. Consumption of natural gas since then has grown dramatically. Due to the financial crisis in 2009, its demand slightly decreased, but the steady decommissioning of oil and coal-driven electricity plants for environmental reasons, made natural gas an upcoming player in the Greek market place.

Greece has no domestic gas production yet, even though it is estimated to have significant reserves that could exploit in the future. The main import source of natural gas is Russia, which is imported via pipelines. Geopolitical games have led Greece to consider other potential suppliers or even starting its own production to cover for the energy demands. In 2005 Russian imports accounted for 84.2%, whereas in 2008 Greek imports from Russia covered 66.3% of demand. Turkish gas exports channeled to the National Gas Transmission System through the ITG pipeline (Interconnector Turkey-Greece) accounted for 10.4% of demand in 2008. The remaining share of imports comes from LNG.

In 2017 Greece imported 1.62×10^9 m^3 of natural gas as LNG (net of re-exports) – an increase of 139.3% from 2016 – and was Europe's seventh largest importer of LNG (jointly with Poland). Between 2014 and 2017 there was a 68% increase in gas consumption in Greece, with a 21% increase in 2017 compared with 2016. Greece covers the energy demands at about 15% with natural gas, out of which, 25% is imported as LNG at Revythousa terminal.

The only active LNG terminal in Greece is the 8.25×10^9 m^3/y capacity, Revythousa terminal, located in Revythousa west of Athens. It came into operation in 2000 and it is comprised by two LNG tanks of combined storage capacity of 130,000 m^3. A third tank is to be added of 95,000 m^3 capacity and the terminal. This addition is of great importance, since larger vessels could reach the terminal, a great step for the Greek participation in the LNG world. The terminal also offers regasification and truck loading services. Future work will include reloading and loading of bunkering ships of very small capacity (~1,000 m^3). Revythousa is owned and controlled from ***DESFA***, the ***TSO*** (Transmission System Operator) and currently operates at 1/4th of its technical capacity [8].

Two more LNG terminals are expected to take place in Greece, from Greek-owned companies (***DESFA*** and ***PPC***), one in Crete (Astakos project) and one in Alexandroupolis, both of which are expected to operate with Qatari LNG; Crete could also import LNG from Egypt. Importing LNG to Crete from a nearby source could lead to reduction of oil dependence of the island for power generation.

Greece also aspires to open a Floating Storage Regasification Unit (FSRU) in northeastern Greece, to Alexandroupolis port. The ***Gastrade*** company is in charge of the project, with a 20% stake owned by ***Gaslog***. The FSRU is scheduled to begin operation by the end of 2020. This project is of great importance for Greece and Balkans in general, and it will promote healthy competition in benefit of the final consumer and safer supply of natural gas for the Balkans [9].

Greece is importing LNG from Algeria and Egypt; in 2005 LNG accounted for 26% of demand based on a long-term contract with Algeria's ***Sonatrach***, and in 2008, 4.9% of the LNG imports originated from

Egypt. In 2009, due to the increase of spot-market cargoes, LNG imports reached around 28% of demand. The first privately owned cargo (Mytilineos) arrived in Revythousa on 11 May 2010 [6].

As far as the Greek shipping is concerned, Greece controls almost 1/3rd of the global fleet of oil tankers, 1/6th of the chemical and petrochemical shipping [10], with the Greek fleet consisting of up to 4,746 ships, each with capacity of 365.45×10^6 tonnes. It has an indispensable role in world seaborne trade, including the EU's export - import trade and in particular in securing the EU's energy needs through the provision of reliable, efficient, green and safe sea transportation for the benefit of EU citizens and the EU economy.

A.2.3. LNG characteristics

A. *Properties of LNG*

As mentioned before, LNG is a liquid fuel at cryogenic temperatures and atmospheric pressure. It has no odor or color and does not corrode the material of the vessels when contained. Its combustion releases low particle emissions and in comparison to other fossil fuels, it generates less carbon dioxide emissions and rarely sulfur and nitrogen oxides, which contributes to being considered the cleanest hydrocarbon fuel with the least impact on the environment.

Its composition is based primarily on methane, with lower concentrations of ethane, propane, butane, nitrogen and traces of higher hydrocarbons, sulfur (<4ppmv) and carbon dioxide (~50ppmv). Composition of the natural gas is determined by the reservoir. However, the composition of LNG is different from the produced gas since before the liquefaction, natural gas follows a process of cleaning and sweetening where higher hydrocarbons are collected and sold separately. Composition primarily affects the properties of the LNG, such as the boiling point and density.

TABLE 2: Chemical Composition of various LNG plants in the world [11].

Terminal	**Methane**	**Ethane**	**Propane**	**Butane**	**Nitrogen**
Alaska	99.80	0.10	NA	NA	NA
Trinidad	92.26	6.39	0.91	0.43	0.00
Nigeria	91.60	4.60	2.40	1.30	0.10
Algeria	91.40	7.87	0.44	0.00	0.28
Malaysia	91.15	4.28	2.87	1.36	0.32
Indonesia	90.60	6.00	2.48	0.82	0.09
Qatar	89.87	6.65	2.30	0.98	0.19
Brunei	89.40	6.30	2.80	1.30	0.00
Australia	87.82	8.30	2.98	0.88	0.01
Oman	87.66	9.72	2.04	0.69	0.00
Abu Dhabi	87.07	11.41	1.27	0.14	0.11

The boiling point of LNG is typically around –162 ºC (–259 ºF) and its density ranges between 430-470 kg/m^3, meaning when spilt in water it will float since its density is less than half of water. In its gaseous form, LNG presents the same thermal characteristics as natural gas: high ignition energy and low laminar burning velocity.

B. LNG quality

The LNG quality is determined by specifications that are defined from end users and pipeline transmission companies. LNG must comply with environmental regulations and must be of specific quality to ensure safety. Since LNG composition varies from project to project, a common basis needs to be set and usually relies on the heating value and composition of the gas.

Gas composition is characterized by an index, Wobbe Index (WI), which indicates how similar a gas or a gas mixture is to others by comparing the heat that is released during combustion. WI is based on the combustion energy which corresponds to the density and heating value of a specific gas.

WI is calculated by dividing the higher heating value of a gas mixture by the square root of the mixture's density or MW (molecular weight) relative to air. This index is used to control the quality of LNG and is the most common factor to be examined in the industry. WI is controlled and regulated by the end users, hence the LNG receiving terminals must be able to adjust this factor to the required values. Variations in LNG composition and Wobbe Index lie risks on the gas interchangeability.

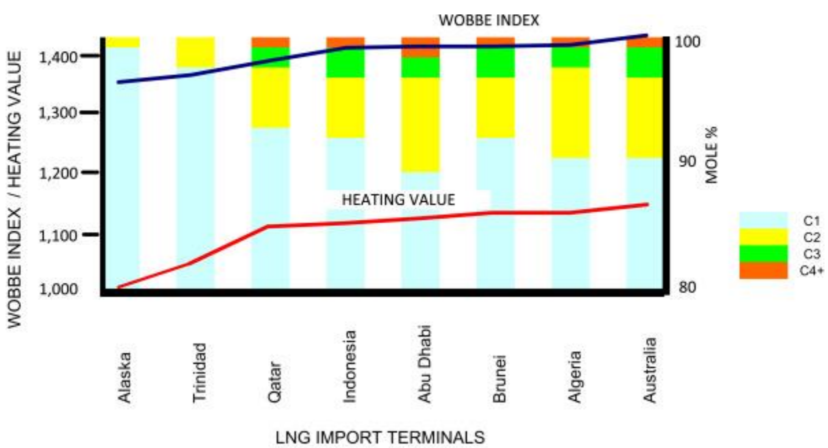

FIGURE 3: Wobbe Index and compositions for different LNG import terminals [12].

The calorific value of a gas (or its Wobbe Index), can be controlled by diluting the vaporized LNG with nitrogen. Nitrogen is an inert gas that is capable of lowering the heating value of gas and increasing its molecular weight when mixed with it. In order to meet pipeline limits, usually no more than 2 to 3 mole% (3 is the upper limit restriction) of nitrogen content is needed. Wobbe Index must be monitored in the regasification terminal, since a part of LNG would evaporate to form boil-off gas, resulting in an increase in the molecular weight of the gas and its calorific value. Some regasification terminals are also capable of injecting nitrogen or LPG (liquified petroleum gas) to the gas mixture, regulating its heating value (increasing or decreasing it). In most Asian countries, a higher calorific value of LNG is desired, hence instead of adding nitrogen, propane is added in the pipeline system to increase the Btu value of the final product.

In case nitrogen dilution cannot meet the regulations (there is an upper limit restriction), then the gas mixture needs to become lighter by removing part of its heavier compounds. If that option is not economically attractive, the only other way to meet the regulations is by adding to the LNG a leaner LNG source.

C. LNG Units

The LNG industry, due to the large quantities it deals with, most often uses million metric tonnes per year to express production. Other units that are commonly used are: MMTPA (million metric tonnes per annum), MTPA (mega tonnes per annum), tpy (tonnes per year), bcm/y (billion cubic meters per year) or Mt/a (mega/million tons per annum), MMSCF/y (million standard cubic feet per year) and others that can be derived from the previous (per day, per hour, etc.).

In order to have an estimation of the calorific value of the produced LNG the units that are used are usually Btu/scf (British thermal units per standard cubic feet), q (British thermal units per cubic meters). To compare prices of the products, the most common units are: USD/tpa or $/tpa (US dollars per tonnes per annum) to correlate cost with production and USD/MMBtu or $/MMBtu (US dollars per million British thermal units) to correlate cost with calorific value.

A.2.4. LNG hazards and safety precautions

LNG is nontoxic, however, when vaporized to natural gas, it can cause asphyxiation due to lack of oxygen in an unventilated, confined area, and can be ignited if mixed with the right concentrations of air.

Vapors released from LNG, if not contained, will mix with surrounding air, which may create a vapor cloud that may become flammable and explosive (unconfined vapor cloud explosions-UVCE). For an explosion like that to occur a number of factors need to be taken into consideration, such as the chemical structure of the vapor molecules, the size and concentration of the vapor cloud, the strength of the ignition source, and the degree of confinement of the vapor cloud. The flammability limits are 5% and 15% by volume in air. Outside of this range, the methane/air mixture is not flammable. When fuel concentration exceeds its upper flammability limit, it cannot burn because too little oxygen is present. When fuel concentration is below the lower flammability limit, it cannot burn because too little methane is present. These conditions can rarely occur in an LNG facility, so such explosions should not be considered as potential hazards.

If spilled on water, LNG floats on top and vaporizes rapidly because it has almost half the density of water. Initially LNG vapors are heavier than air and will stay near ground level. However, as LNG vapors begin to warm up by the surroundings and reach temperatures of approximately $-166\ ^oF$, the density of the vapors is lighter than air and the vapors become buoyant. Cold LNG vapors (below $-166\ ^oF$) are more likely to accumulate in low areas until the vapors warm up. A release of LNG in an enclosed space or low spot will tend to displace air, making the area hazardous for breathing. The LNG industry has an excellent safety record so far, as a result of good engineering, well understanding of the LNG properties and continuous evolving and improving.

LNG spills could be prevented by detection equipment, spill risk assessments on the equipment used for storage and transportation and protocols for prevention and containment for different scenarios.

During the LNG transportation cycle, leaks and vapor release could occur due to thermal differentiations. In extreme situations, stratification of the hydrocarbons inside a tank due to weathering could result in a rather sudden increase in the tank head pressure and lead to rollover of the tank. Rollover occurs in a weathered tank, where there is rapid mixing of stratified layers within an LNG tank causing sudden release of very high levels of boil-off gas within a short period.

Fire hazard also poses a threat if operations during storage, loading and unloading are misheld. In order to prevent such accidents, equipment must be designed to comply with international safety standards to prevent fires and explosion hazards and have flexibility at different operating modes. Safety precautions that could apply are the design of storing excess LNG, boil-off gas management control and frequent maintenance and inspection by experienced personnel.

A. LNG Emissions

Compared to petrochemical facilities, an LNG plant is considered cleaner since its emissions and wastes are considerably lower, yet it still produces wastes and emissions that are not environmentally friendly. An LNG supply chain has various different units, each generating different emissions.

The most common emissions from an LNG supply chain are CO_2, NO_x, SO_x, H_2S, mercury and other chemical waste (mostly used in catalysts), waste water (mostly used in boilers, heaters and condensers), cold water disposal (affects the marine ecosystem) and can arise from the liquefaction process, the regasification process, the feed gas conditioning (cleaning and processing) even from the transportation (release of boil-off gas, CO_2 emissions, NO_x, SO_2). The majority of the CO_2 emissions is generated by the gas processing and liquefaction step. A typical LNG liquefaction terminal exporting 4.5 million tonnes of LNG can be expected to produce in the order of 1.2 million tonnes equivalent carbon dioxide of direct emissions. If fuel gas is used to regasify LNG, the greenhouse gas emissions are increased further by an equivalent of about 0.2 million tonnes of carbon dioxide emissions.

Emissions must be estimated to comply with local and federal environmental regulations. Emissions are measured starting from the well head to the final consumers. The environmental impact statement is part of the overall safety and security assessment.

Measures to reduce emissions

- Air coolers should replace when possible the cooling water; leading to minimization of water consumption.
- Cold water produced from the regasification process should be discharged to the ocean within 3 ºC of the surrounding water at the edge of the mixing zone or within 100 meters of the discharge point, as noted in the International Finance Corporation (IFC) guideline.
- Chemical additives should be used carefully and after strict consideration of their impact to the marine ecosystem when disposed.
- Waste water containing chemicals (waste oils, hydraulic fluids, waste chemicals, filter elements, solvent wastes, spent molecular sieves and spent catalysts and chemicals) and solid waste must be properly disposed or recycled. Recycling is a desirable option but when not feasible, wastes must be disposed of with respect to the environment and regulations.
- Flaring or venting is used to avoid build-up of excess gas in the system. Flaring could be avoided by recovering the excess gas as fuel gas.
- BOG from the liquefaction and regasification processes could be compressed and used as fuel gas or reliquefied in the process. BOG venting is not allowed during normal operation.
- Air emissions of high temperature should be considered as a heat source for LNG regasification.
- Installation of noise-dampening devices should be used to reduce noise generated by pumps, compressors, generators, drivers, fan air coolers, LNG vaporizers and heaters.

B. *Environmental Impacts and Elimination of CO2 emissions*

Not every LNG project has the same impact on the environment. Apart from what was previously discussed, another aspect that should be taken into consideration is the location of an LNG infrastructure. The proper location and infrastructure should minimize the effects of natural hazards such as earthquakes, tidal waves, floods and fires from surrounding areas. Site selection should also minimize the project's impact (from construction to operation) on sensitive ecosystems such as marine life and agricultural fields and comply with local regulations.

After careful consideration and design, an LNG plant could reduce its CO_2 emissions to almost elimination. This could be achieved by improving the plant's efficiency on liquefaction cycles and equipment, by reinjecting CO_2 from the amine unit for sequestration, by utilization of LNG cold in the receiving terminals. In addition, shortening the distance between LNG plant and end users and using renewable energy to generate power for the different processes, CO_2 footprint could further be reduced.

A.2.5. LNG transport

When the final LNG product is formed, it needs to be transported to the final consumers. The transportation can occur via two ways: through the land by trucks or through the sea by ships.

A. *Transportation on Land-Trucks*

For the consumers that do not have access to seaborne LNG, trucks, delivery units and cryogenic containers can transfer the wanted product to the final inland destination. Trucks are used as an alternative to pipeline systems, since the latter is not always an economical means of fueling for remote areas, but LNG is still a commodity and needs to be transferred to the point.

Inland transfer of LNG is an industry with many years of experience. Truck use specialized, double-skinned vacuum insulated tanks and deliver LNG to refilling stations (insulated pressurized storage tanks). From the storage tanks, LNG is delivered via pipeline systems to the final consumers for domestic use, after being brought to its gaseous state, metered and odorized to be easily detectable if leaks occur.

Truck fleet is now trying to make a switch from diesel-operated engines to a cheaper and cleaner fuel such as LNG. LNG could also serve other transportation vehicles such as buses, automobiles, taxis, since it is a high efficiency fuel with higher heat release during combustion than other oil derived fuels. As a result, more and more LNG refilling stations are built and more technologies are developed to boost the market of LNG.

B. *Transfer by sea- Carriers*

For long distance destinations, LNG is transported through the sea in carriers (ships for the transportation of LNG) that are specially designed. A carrier consists of spherical tanks that are well insulated and transfer the cargo at a temperature of –259 ºF (–169 ºC) and pressures slightly above atmospheric (~30 kPa). Sometimes it is possible for heat to reach the LNG no matter the insulation used. This heat can lead to vaporization of some of the product during shipping. This process, called ageing or weathering, is not homogeneous, since LNG is a mixture of gases and the lighter components are the ones to evaporate first (such as methane and nitrogen-lowest boiling points). As a result, the LNG becomes heavier and its heating value and Wobbe Index increase over time. The boil-off gas that occurs, needs to be removed (usually at a rate of 0.10% to 0.15% of the ship volume per day) in order to maintain constant pressure of the tank, but it does not go to waste, since it can be used as a fuel for the ship, it can be burned to generate steam in boilers or it can be reliquefied and returned to the cargo tank, an option that is more preferable for long distance voyages in order to maintain the initial volume and composition of the cargo.

The first LNG ships had a standard capacity of 125,000 m^3 with five cargo tanks. Nowadays, there is a big variety in sizes, as there are LNG ships that are used for small distances and usually have small volume capacity (less or about 30,000 m^3), but also LNG ships designed for long-distance supply chains. Such ships are *Q-Flex* (~216,000 m^3 LNG capacity) and *Q-Max* (~265,000 m^3 LNG capacity), which use diesel as a fuel of transportation and are equipped with an on-board liquefaction system that re-liquefies the vaporized LNG. Compared to the size of a typical LNG carrier (~130,000 m^3), the *Q-Flex* and *Q-Max* are almost double the size. The only drawback of these carriers is that not every terminal station is capable of sustaining the amount of LNG they deliver.

TABLE 3: Typical Dimensions of LNG carriers [13].

	Cargo Volume (10^3 m^3)	Length (m)	Depth (m)	Number of tanks	Propeller system
Typical LNG Ship	138-173	277-290	26-26.5	4	Single or twin
Q-flex	210	315	27	5	Twin
Q-max	263	345	27	4	Twin

LNG carriers are distinguished in two main categories, depending on the type of the tanks, a membrane design and an independent solid type structure.

Solid type-Freestanding tanks: These particular tanks are either spherical or prismatic robust structures, that are autonomous from the rest of the tanks and the ship's structure. They are usually constructed with nickel steel or aluminum alloys and use a number of external layers in order to achieve maximum insulation. Figures 4 and 5 illustrate the two shapes of freestanding tanks.

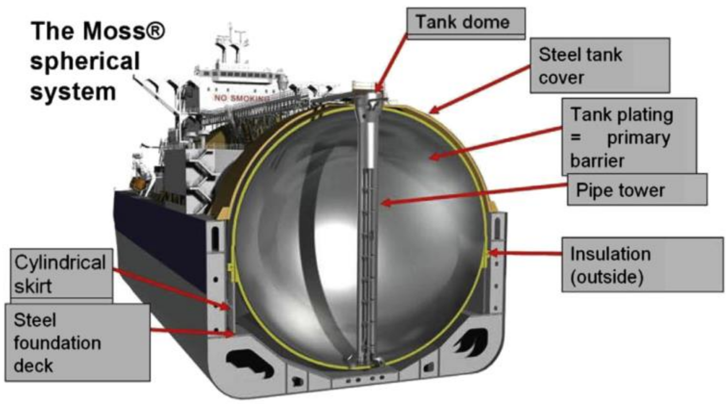

FIGURE 4: Free-standing spherical LNG tank. (Moss Maritime®)

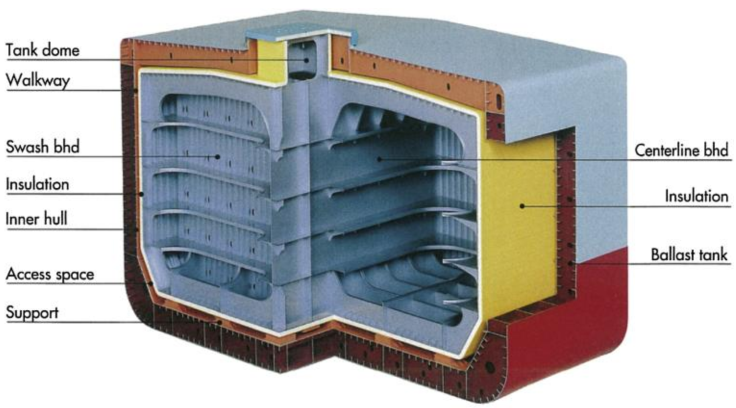

FIGURE 5: Typical illustration of free-standing prismatic type B LNG tank [14].

<u>*Membrane tank structure*</u>: The second type of LNG tanks used for marine transportation are the membrane type and compared to the freestanding tanks, they are not independent from the ship's hull. Instead, they are supported by the ship's structure. Their structure is made of successive layers of membrane barrier and insulation to reduce the heat transfer between the cargo and the environment. Membrane barriers have the ability to expand and contract in different temperature conditions without causing extensive stress to the material (usually steel and nickel alloys). The membrane material, as designed by ***Gaz Transport*** and ***Technigaz*** (TG MARK III), is a high nickel content steel (Invar) or a chrome/nickel stainless steel respectively. The first membrane tank of ***Gaz Transport*** (GT N_o 96), uses plywood and perlite as an insulation material, while the second type of ***Technigaz*** (TG MARK III), uses two layers of reinforced polyurethane foam, with triplex in between. The two companies have now joined together and have developed a new membrane type of tank, Combined System Number One (CS1), which is a combination of their two former designs.

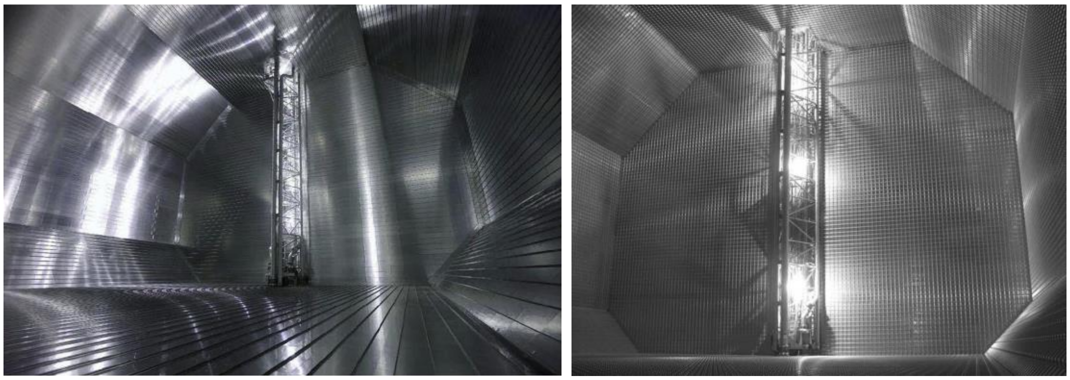

FIGURE 6: On the left, the GT No 96 membrane containment system and on the right, the TG Mark III membrane containment system.

LNG ships can also be categorized by the propulsion system they use. Even though the traditional way used to be a boiler and a steam propulsion system that used the boil-off gas, technology has moved forward to diesel engines and alternative propulsion systems to eliminate the cost of transportation fuels. The up to date propulsion systems also include a gas combustion unit (GCU) for the disposal of excess BOG to avoid pressure buildup in the tanks. The most common propulsion systems are:

Dual fuel diesel electric (DFDE) propulsion system: DFDE system can operate with both fuel oil and BOG, offering high efficiency. It is usually accompanied by electric propulsion systems that are of high maintenance due to the gear and motors they use, resulting to a higher cost compared to steam engines.

Gas turbine propulsion systems using onboard BOG: The gas turbine propulsion systems are well established in the LNG transportation world, due to their easy operation and low maintenance needed. They can operate using dual fuel and provide sufficient energy for the voyage. The power they generate, could be used, besides the propelling system, in compressors used on-board for liquefaction (FLNG vessels).

High pressure steam turbines: This type of turbines, offer better fuel efficiency over the conventional steam turbines but are of high maintenance. Boilers cannot operate with seawater so a different aquatic source is needed for their operation. This is the main reason why the high pressure steam turbines come with a high cost and are usually not preferred on LNG carriers.

C. LNG transportation in arctic climates

LNG could also be transported from and to an arctic environment of production. Transfer in this occasion is only feasible through sea, yet the conditions of transport are unique. A significant amount of oil and gas (gas hydrates) is present in these areas and need to be exploited, but the transportation to the consumers is very difficult and costly. A way to overcome transportation problems, was to build Arctic LNG ships. Ice-class LNG vessels first operated in the Sakhalin-II project in eastern Russia with a few more to follow. These ships are designed to withstand the ice and are equipped with ice breakers. They operate with very powerful engines (85 to 120 MW), narrower beams and strong propulsion equipment in order to break the ice as they move forward. A trip in an arctic or subarctic region could last a long period of time, with an ice-class LNG vessel moving at times as slow as two nautical miles per hour [15].

A.2.6. Economics of LNG

A. *Economics of the LNG chain*

The cost breakdown of the production and distribution of the LNG can be divided into five major categories: the gas production and treatment, the liquefaction process (including the nitrogen rejection unit), the fractionation which produces LPGs, the utilities and offsites used and last, the storage and loading/unloading of LNG.

The gas treatment accounts for almost 12% of the total cost of the LNG chain. Gas production cost typically varies from 0.25 USD/MMBtu to more than 1.0 USD/MMBtu. A production cost of less than 1.0 USD/MMBtu is desirable to make the LNG option economically viable and competitive.

LNG projects are capital intensive. The liquefaction plant is the largest cost component, accounting for approximately 50% of the total cost of the LNG chain. The capital cost of the liquefaction facilities depends on several factors such as plant location, size of plant, site conditions and quality of feed gas. The liquefaction process leads to a final LNG cost of delivery from 1.5 to 2.0 USD/MMBtu. The cost of a liquefaction plant is a significant component of the cost of the LNG chain; hence, cost reduction of the liquefaction facility is an important issue and can occur by improving the efficiency of the liquefaction processes.

Transportation of LNG is a significant portion of the total cost of the LNG chain. The cost of shipping depends primarily on the distance between the liquefaction facility and the market. A typical contribution of the shipping cost to the cost of delivered LNG is approximately 0.5 to 1.2 USD/MMBtu.

The receiving terminals with tanks, vaporization equipment and utilities contribute approximately 0.3 to 0.4 USD/MMBtu to the delivered price of LNG. These costs are highly dependent on the design of storage tanks and specific site conditions [66].

B. *Economics of LNG Projects*

Capital Expenditures (CAPEX)

The CAPEX of LNG projects is estimated to be between 1.32 USD/MMBtu and 7.21 USD/MMBtu depending on several factors, including type of technology used, shipping distance between the liquefaction plant and regasification facility, economies of scale, learning curve improvement and local infrastructure availability. Published data on the CAPEX of LNG projects shows a range of 200 USD/TPA (3.35 USD/MMBtu) to more than 850 USD/TPA (16.35 USD/MMBtu.)

Annual Operating Expenditures (OPEX)

Liquefaction fuel, re-gasification fuels and tanker boil-off contribute significantly to the operating cost of the LNG projects. It should be noted that value is added to the gas as it goes from one value chain to the other, another major influence is the location of the buyer and the seller, reported shipping costs to the US or Europe to range between 0.6 and 1.75 USD/MMBtu for existing and potential Latin American, African and Middle Eastern LNG projects.

Net Cash Flow (NCF) of LNG Projects

The two measures used to assess the economic viability and profitability of LNG projects are the rate of return (ROR) and undiscounted pay-out time (POT) or pay-back period (PBP). ROR is the discount rate at which the net present value (NPV) is equal to zero. NPV can also be a measure to conclude for the profitability of the project, as long as it presents values greater than zero. The undiscounted PBP is the time required in years, after the commissioning (start-up) of the project, to pay back the undiscounted initial investment [67].

A.3. Types of Liquefaction plants

Liquefaction plants can be categorized based on their size and function, into three main categories: large base load, peak-shaving and small- to medium-scale plants.

Baseload: Base load are plants of high capacity consisting of one or multiple trains. They are usually located at the point of large producing reservoirs due to their high capacity; as much as 7.8 MTPA single train capacity in Qatar. Base load plants supply natural gas in the form of LNG, from the source of production (natural gas producers) to consumer nations. Due to their size and cost, base load plants are mega projects and exist around the globe to the largest gas reservoirs in Asia, Australia, Middle East and West Africa.

Peak-shaving: Peak-shaving plants are smaller in size than base load plants, up to 0.1 MTPA, and are used to liquefy and store excess gas during peak demand periods. Therefore, their function is mainly to balance fluctuations in natural gas demand and supply during seasonal changes.

Medium-small scale: The last category of LNG plants, medium to small scale, have emerged from the exploitation of smaller gas reservoirs all over the world. The capacity of a mid-scale LNG plant varies from 0.3 to 1.5 MTPA, while small-scale LNG plants have capacities as low as 0.01 MTPA. The producing LNG from these plants is distributed inlands by trucks since offshore transportation is economically unattractive for such small capacities and when pipeline distribution systems are not available. Small-scale LNG units have recently been installed on large LNG carriers to reliquefy boil-off gas and reduce shrinkage of the LNG during the transportation period.

Apart from their size, LNG production plants are typically consisted of the same processing units. A typical scheme of an LNG production plant is presented below.

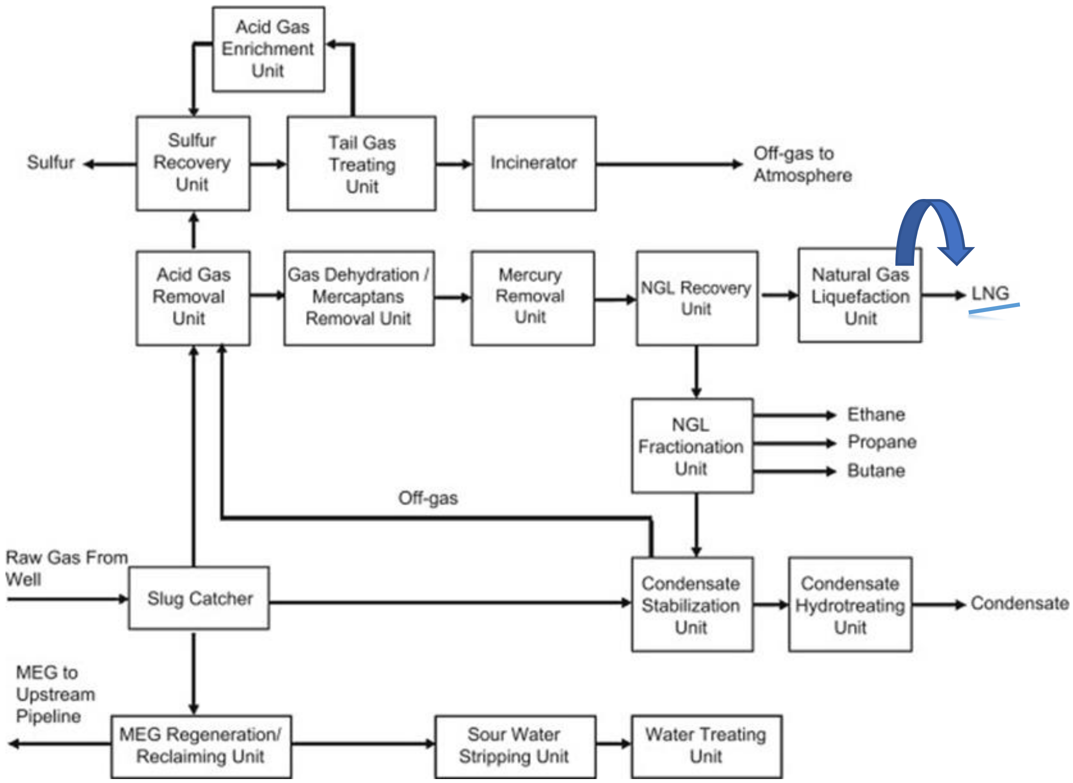

FIGURE 7: Typical scheme of an LNG production plant.

As seen on Figure 7, the raw gas from the well is firstly headed in a slug catcher, to separate the liquids from the gas. The gas is then led in a high pressure (HP) separator, while the liquids are separated in medium pressures. After the separation, the hydrocarbon liquids are fractionated in the NGL (Natural Gas Liquids) fractionation unit, where ethane, propane and butane are collected and sold as separate products. The remaining condensate enters a hydrotreating unit where any H_2S content is removed and pressure is controlled to meet specifications for safe transport and storage (82.7 kPa).

All the vapors produced during this process, are collected and sent back to the HP separator. The gas output enters a Gas Sweetening Unit, GSU (or Acid Gas Removal Unit, AGRU), in which H_2S and CO_2 are removed to meet product specifications. H_2S is removed by an amine solvent to meet the total sulfur product specification (typically 4 ppmv). CO_2 is removed to 50 ppmv to avoid CO_2 freezing in the main exchangers in the liquefaction plant. Carbonyl sulfide (COS) and mercaptans (R-SH) if present, contribute to the sulfur contaminants and must also be removed.

Acid gas exiting the GSU passes through a Sulfur Recovery Unit (SRU), which consists of a Claus unit and a Tail Gas Treating Unit (TGTU). The off-gas from the TGTU absorber is incinerated. The sweet gas exiting the GSU needs to be dehydrated in order to avoid hydrate formation in the NGL recovery unit.

Molecular sieves are used to remove water content and can also be designed to remove mercaptans. Traces of mercury could also be present in the feed gas and their removal can occur with special beds to less than 10 nanograms per cubic meter to avoid mercury corrosion in the MCHE.

The three major processes of an LNG plant, feed gas conditioning, NGL recovery and liquefaction, are briefly presented below.

A. *Feed gas conditioning*

Natural gas typically contains a mixture of methane and heavier hydrocarbon gases, as well as carbon dioxide, nitrogen, water, mercury and a range of other unwanted components. Unwanted components must be removed and the heavier hydrocarbons separated prior to liquefying and transporting natural gas. Therefore, the first step in an LNG production plant is to treat the feed gas composition in order to meet the final product's specifications and the legal regulations regarding the emissions and its overall environmental impact. This process, known as feed gas conditioning, is the first and most important step to provide the design specifications of the LNG plant and meet the final product requirements set by contractual agreements.

TABLE 4: Typical LNG Product Specifications.

	Rundown to Storage	**Unloaded at customer port**	**Units**
Properties			
Higher Heating Value	42-44	42-45	MJ/Sm^3 (15 ºC)
Wobbe Index	51-53	51-54	MJ/Sm^3 (15 ºC)
Compositions			
Methane	84-96	84-96	Mole %
Max C_4+	2.4	2.5	Mole %
Max C_5+	0.1	0.1	Mole %
N_2	1.4	1.0	Mole %

As seen on the Table above, there are two LNG product specifications: one for the LNG from the liquefaction plant and the other one for the LNG unloaded at customer port. The difference is due to the boil-off of the volatile components from heat leaks during storage, ship loading and unloading, and ship voyage.

TABLE 5: LNG Quality Specifications for the Revythousa terminal [16].

VALUE	SPECIFICATION	UNITS
WOBBE INDEX	13.066-16.328	kWh/Nm^3
GROSS CALORIFIC VALUE (GCV)	11.131-12.647	kWh/Nm^3
LNG DENSITY	430-478	Kg/m^3
MOLECULAR WEIGHT	16.52-18.88	Kg/Kmol
METHANE	85.0-97.0	% mole
I-BUTANE AND N-BUTANE	<4	% mole
I-PENTANE AND N-PENTANE	<2	% mole
NITROGEN	<1.24	%mole
HYDROGEN SULFIDE	<5.0	mg/Nm^3
TOTAL SULFUR	<30.0	mg/Nm^3
TEMPERATURE	< -158	^{o} C

B. Natural Gas Liquids (NGL) recovery

The gas that has been processed in the feed gas conditioning and the harmful components have been removed, is routed to a Natural Gas Liquid (NGL) recovery unit. The NGL recovery unit is designed to remove and recover the heavier hydrocarbons such as ethane, propane and butane and produce a lean gas, natural gas that contains a few or no liquefiable liquid hydrocarbons, that will eventually be converted into LNG.

This step is crucial since no further units for the removal of heavy hydrocarbons will be needed (heavier hydrocarbons and aromatics could cause waxing in the main exchanger and additional units to scrub the waxes produced would be needed). Furthermore, the recovered liquids can be processed and sold separately. Ethane can be used in petrochemical plants to produce ethylene, propane and butane are exported as separate products and C_{5+} or heavier hydrocarbons can be used in gasoline blending. Mercaptans, if present in the feed gas, make their appearance in C_{5+} liquids, so they would further need to be treated in order to be removed and meet the final product's specifications.

C. Liquefaction

The gas that leaves the NGL recovery unit is headed to the liquefaction unit in order to be converted into its liquid form. This process, called liquefaction, is the most important part of an LNG plant after which, the final product is ready to be transported. Liquefaction process has been based on refrigeration cycles. A refrigeration cycle uses a refrigerant to remove heat from the feed gas and finally bring it to cryogenic conditions. The refrigerant is recycled throughout the process and passes through stages of compression and expansion to control the heat transfer between the stages of cooling. The refrigerant could be part of the natural gas feed (open-cycle process) or a separate fluid continuously recirculated in the process (closed-cycle process).

During or after liquefaction, the Wobbe Index is determined and controlled by a nitrogen rejection unit that uses cryogenic separation process to remove nitrogen if its content exceeds the commercial specifications of LNG (~1 mole %). Nitrogen is flashed off and removed from a flash drum, being the lighter component of the LNG mixture. The nitrogen-rich vapor is compressed and recovered as fuel gas. When the nitrogen feed gas is higher than the specifications, an additional fractionation step is required. Nitrogen removal is essential to avoid low liquefaction temperatures and rollover of the storage tanks and so that the boil-off gas could be used as fuel gas. Nitrogen content could also be controlled before or during liquefaction.

A.4. Onshore natural gas liquefaction technology

As mentioned before, liquefaction is a process based on refrigeration cycles, where the pretreated feed gas is cooled down to liquid form. The refrigerant could be part of the gas feed (open-cycle process) or a separate refrigerant that circulates continuously in a heat exchanger (closed-cycle process). The extremely low temperatures that liquefaction of the gas feed requires, can be achieved through an extra work that is introduced to the refrigeration cycle from compressors and by the rejection of the heat through the air or by the utilization of air coolers.

Over the years numerous liquefaction processes have been developed and a lot of work has been put into the optimization of this process, that could not only lead in reduction of the operating cost and unit's investment, but also in the maximization of the LNG volume produced and overall better efficiency of the refrigeration cycle. The main components that can affect the design of the liquefaction process are: the refrigerant selection and composition, the design of the heat exchanger and the compressor's capacity that is determined by the refrigerant power consumption.

A typical natural gas liquefaction is divided into three main zones: a precooling zone, followed by a liquefaction zone and a sub-cooling zone. These zones differ in slopes, or specific heats that determine the heat transfer area, along the process. The cooling curve is used as a basis and every other refrigerant or refrigeration process tries to fit as much as possible the shape of the natural gas cooling curve at operating conditions, for each distinctive zone, in order to increase refrigeration efficiency and reduce energy consumption. The Figure below shows the typical natural gas cooling curve along with a pure and a mixed refrigerant curve.

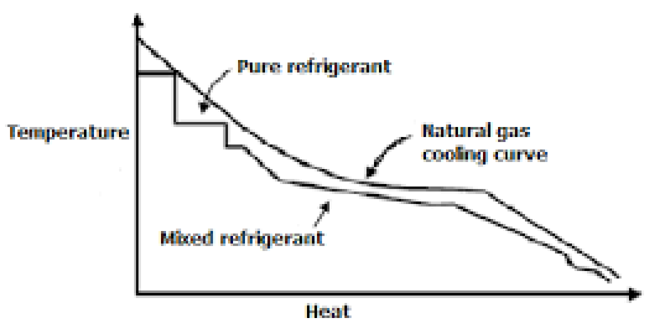

FIGURE 8: Typical natural gas cooling curve [17].

Among the different liquefaction processes that are already in use or have been suggested for onshore production, three broader categories can be distinguished; ***cascade liquefaction processes, mixed-refrigerant and expansion-based processes***. The comparison between these liquefaction processes lies in the energy consumption, economic performance, liquefaction efficiency and exergy analysis. Another proposed way to categorize the liquefaction methods, is based on their size. Small to mid-scale LNG plants rely mostly on single pure refrigerants while the large-scale LNG plants rely on mixed-refrigerant cycles. Small-scale LNG meets a lot of variations depending on their location, purpose (peak-shaving) and capacity and so different liquefactions processes may be selected for each scenario. Large-scale LNG plants on the other hand, due to their capacity, use liquefaction methods based on mixed-refrigerant (MR) technologies such as: Propane pre-cooled (C3MR) single mixed-refrigerant (about 77% of the world's LNG plants use this liquefaction method), dual-mixed refrigerant (DMR), parallel mixed refrigerant, mixed fluid cascade process (MFCP) and C3MR with a nitrogen refrigeration cycle (AP-X) process [18]. All of liquefaction processes mentioned above will be presented analytically in the next section.

A.4.1. Natural gas liquefaction technologies

A. Cascade Cycles

When it is not possible to use only one refrigerant due to high thermal differentiations between hot and cold zones, or between the gas and the refrigerant, the cascade cycle is used as a refrigeration method. This configuration uses more than one individual cycles with a common heat exchanger between cycles. The condenser of a cold cycle is the evaporator of the following hotter cycle, as shown in Figure 9 [19].

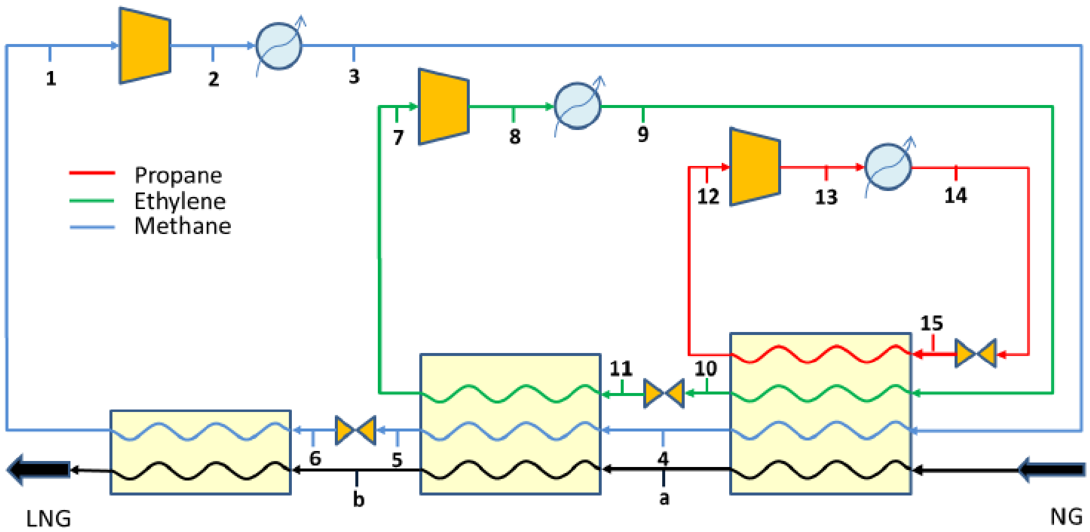

FIGURE 9: Refrigeration cycle using a triple cascade [19].

The classical cascade liquefaction process attempts to approximate the natural gas cooling curve by the use of a series of refrigerants (most commonly three) in separate operations. More than three refrigerants allow an optimized approach to the NG cooling curve but it comes with the disadvantage of additional equipment, higher cycle complexity higher operating costs and larger plant footprint [20].

The cascade cycle is capable of reducing the heat exchange losses due to the utilization of different refrigerants that vaporize at different but constant temperatures. Each refrigeration cycle dictated by the different refrigerants can be controlled separately, giving this method flexibility in operation. Compared to other methods, the cascade cycle has a shorter construction period due to the standard equipment utilized and presents lower technical risks since it is a well-established method. Another advantage of this method lies in the low heat exchange area that requires per unit of capacity. This property makes the method a better fit for large train operations, since the low heat exchange area and the low power requirement eliminate the need for multiple operational units. On the other hand, cascade technology requires high capital investments, it is a very composition-specific process since it doesn't offer flexibility in the feed gas composition and it comes with limitations regarding the LNG train capacity. Conoco-Phillips has proposed a modification of the typical cascade cycle, the Optimized Cascade Cycle and Linde and Statoil have come up with another modification, the Mixed-Fluid Cascade.

A.1 ConocoPhillips Optimized Cascade® Process

The Optimized Cascade process was first developed by Phillips Petroleum Company and came into operation with the Kenai LNG plant in 1969. The ConocoPhillips Optimized Cascade Liquefaction Process (POCLP) is based on three multi-staged, cascaded refrigerant circuits using pure refrigerants (propane, ethylene and methane), brazed aluminum heat exchangers and insulated cold box modules. ConocoPhillips has optimized the heat integration to closely approach the natural gas cooling curve. Pure refrigerants are selected due to their known and predictable properties, offering flexibility in the heat integration. The heat exchangers and coolers that are used in this process (brazed aluminum heat exchangers and cold boxes) can be scaled to match different LNG plant sizes at a high efficiency heat transfer. Figure 10 below illustrates the optimized cascade process [21].

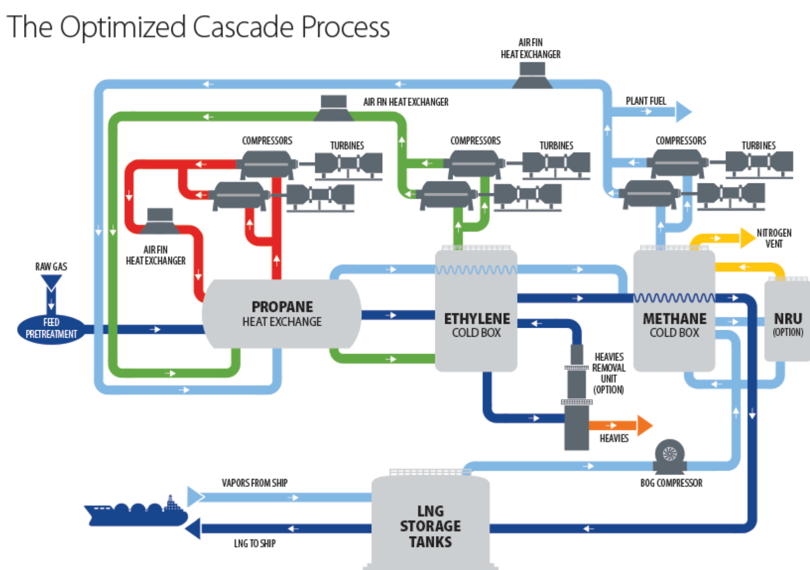

FIGURE 10: The optimized cascade process [21].

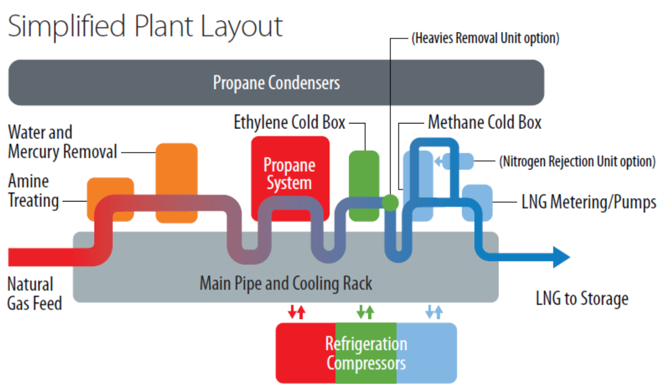

FIGURE 11: The optimized cascade process simplified [21].

Feed gas is routed successively through three interactive cooling stages of pure propane, ethylene and methane that are used as refrigerants. Each refrigerant cycle operates at different pressure conditions in order to improve the overall efficiency. Air or water is used to cool down and condense the propane. Propane then condenses ethylene and ethylene in turn, condenses methane. Methane is then flashed to meet product specifications. In the presence of nitrogen, a slip stream is drawn off to be used as fuel and avoid build-up of the inert.

Each refrigerant circuit utilizes two compressors (for each refrigerant) and for the propane cycle, core-in-kettle type exchangers are used, while the ethylene and methane cycles use brazed aluminum plate-fin exchangers (cold boxes). The cooling process takes place in the two cold boxes, with the exception of propane which is chilled by a water or air cooler. The heat exchanger designs used for the cascade cycles are less complex than the proprietary spiral-wound heat exchanges (SWHE) used in the mixed-refrigerant processes (such as C3MR).

The final LNG product is pumped into insulated storage tanks where it remains until shipment at approximately 7 kPa above atmospheric pressure and at -161 ^{o}C. Boil-off gas and ship return vapors are captured and recycled through the Optimized Cascade process for efficient reliquefaction [21].

The main difference of this method from the typical cascade cycle, is the fact that the optimized method can be used for a wide range of gas compositions and conditions by controlling the loads between the refrigerants. The POCLP could be used for large trains of LNG plants and is able to provide high thermal efficiency (net heat removal to energy input ratio); exceeding 93% in some cases like the optimized cascade facilities commissioned in Darwin Australia in 2006 [22]. *Phillips* and *Bechtel* in a collaboration provide services from designing, constructing and commissioning POCLP plants, with responsibilities been drawn to a single contractor, for the process execution and performance, thus lowering the operational and financial risk.

A.2 Statoil / Linde Mixed Fluid Cascade Process (MFCP)

A combination of cascade and mixed refrigerant technologies was developed by Statoil and Linde, addressing in baseload LNG plants operating in harsher environments. The method was used for the first time in Snøhvit LNG Project in Hammerfest, Norway, which remains to date the only baseload export gas liquefaction plant in Europe. The plant has 4.3 MTPA capacity and uses waste heat to generate electricity for the operation of the three compressors. Each MR is matched to the NG cooling curve as closely as possible to increase the overall efficiency. The plant also utilizes aero derivative turbines to reduce the energy requirements of the liquefaction process, plate-fin exchangers for the first two stages of cooling, spiral wound heat exchangers (SWHEs-a Linde product) and generators for a continuous operation. A flow schematic of the Snøhvit project is provided in Figure 12.

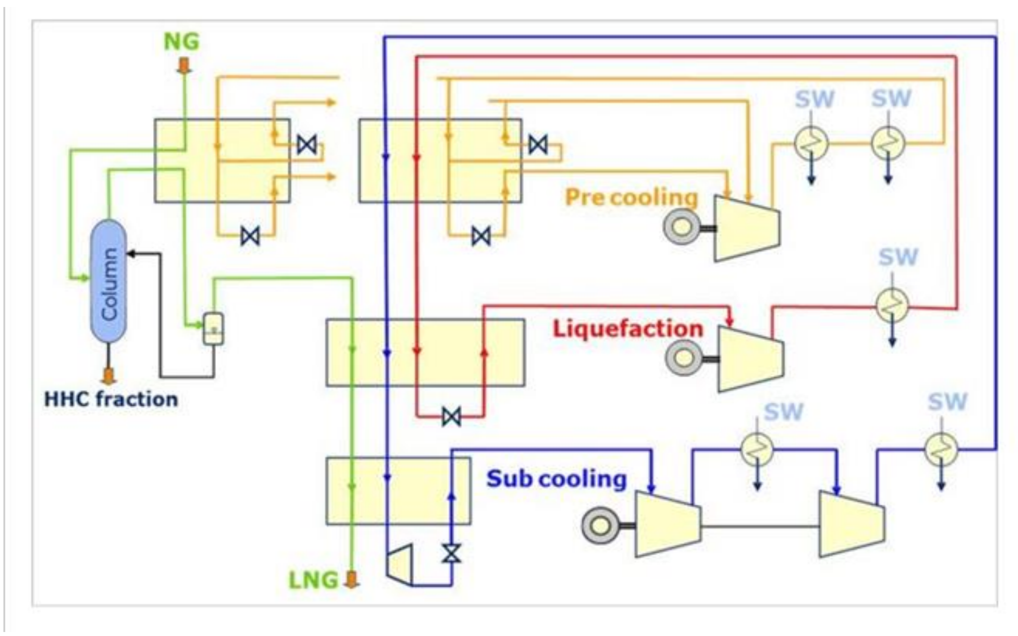

FIGURE 12: Mixed Fluid Cascade (MFC®) process Courtesy Statoil [23].

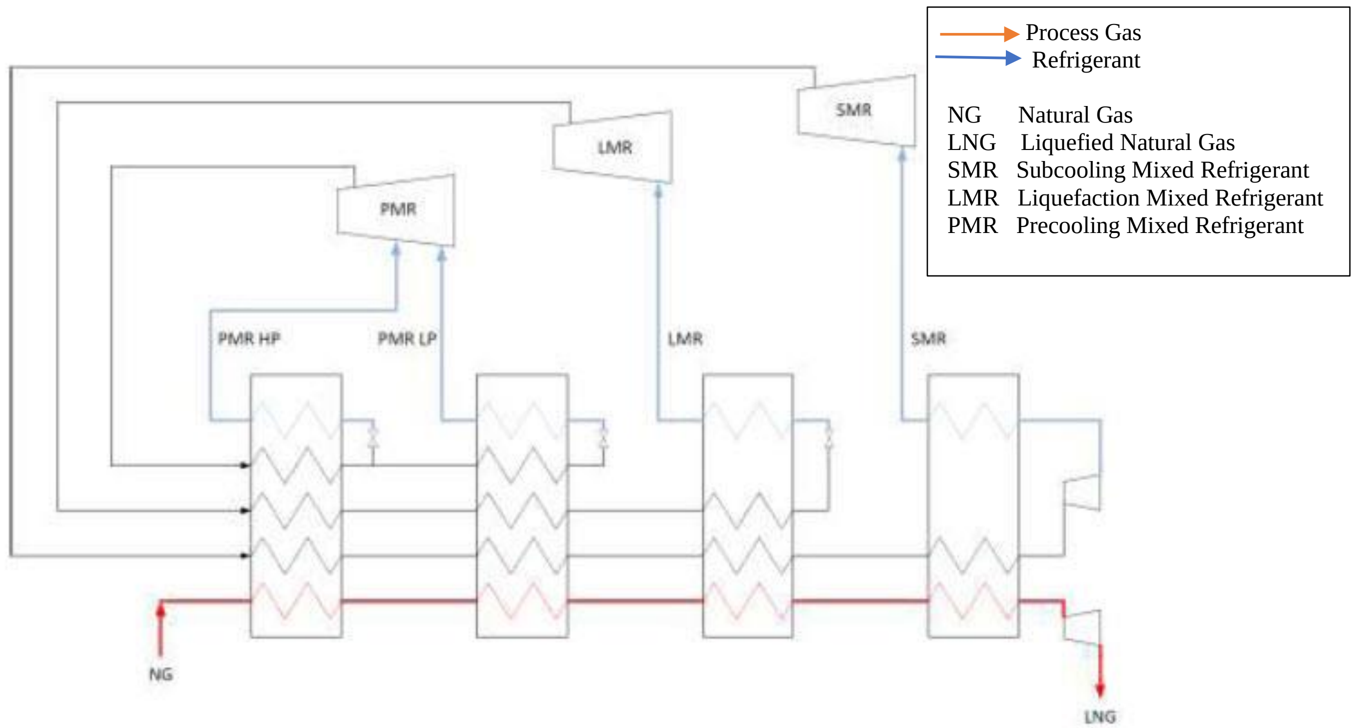

FIGURE 13: Linde Mixed Fluid Cascade LNG Liquefaction Process [18].

This process uses three separate mixed refrigerant (MR) systems to progressively cool the gas; Precooling Mixed Refrigerant (PMR), Liquefaction Mixed Refrigerant (LMR) and Sub-cooling Mixed Refrigerant (SMR). Natural gas, after being conditioned, flows through all three MR cycles. The precooling step could work with two plate fin heat exchangers (PFHEs) or with two spiral-wound heat exchangers (SWHEs), also used in liquefaction and sub-cooling cycles. Snøhvit plant so far is the only plant using this refrigeration method and it has dealt with heat exchangers' malfunction and overall poor performance.

B. Mixed Refrigerant Cycles

A mixed refrigerant cycle (MR) cools down a natural gas stream in a continuous process, by the usage of different refrigerants combined together. The blend that occurs is typically a mixture of light hydrocarbons and nitrogen that approximates the cooling curve of natural gas, in order to reduce energy consumption of the operation and optimize the size of the heat exchangers.

Mixed refrigerant cycle offers some advantages over the cascade process, mainly concerning the operation of the heat exchangers, the number of compressors it utilizes and the flexibility it offers in refrigerant compositions to compensate for changes in gas composition and operating conditions. These advantages don't address to a single MR cycle though, since it cannot optimally match the cooling curve of natural gas.

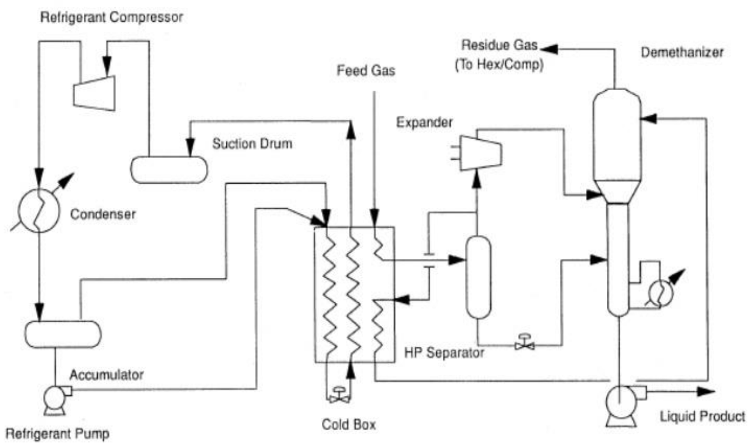

FIGURE 14: A typical scheme of a Mixed Refrigerant Process [24].

The MR process is very desired in the industry and has been applied in the majority of LNG plants. MR technology is provided by different companies, Air Products and Chemicals Inc. (APCI), Shell, Statoil / Linde and Axens, each with a patented MR technology. APCI has developed the C3MR process which is the dominant technology for the industry. AP-X is a modified version of C3MR. Pure refrigerant technology for large scale LNG projects is for the most part limited to the ConocoPhillips Optimized Cascade Process [18].

B.1. Single Mixed Refrigerant Cycle

The single mixed refrigerant cycle (SMR) is based on the reverse Rankine cycle, where the gas is cooled down and liquefied in just one heat exchanger. The physics behind the normal Rankine cycle is the conversion of heat into work, using steam or hydrocarbons. Hence, the reverse cycle does exactly the opposite; it uses work (power) to remove heat and chill the gas. The stages that a reverse Rankine cycle follows in a successive order are: compression→ cooling at ambient temperatures→ condensation→ expansion→ evaporation at low temperatures. The SMR process offers low thermal efficiency, therefore is not indicated for large LNG plants. When used for an LNG plant, the working fluid could be either a mixed refrigerant or propane.

B.1.1. APCI Propane Pre-Cooled Mixed Refrigerant Process (C3MR)

The Propane Pre-cooled Mixed Refrigerant Process (PPMR or C3MR), is designed by Air Products and Chemicals Inc. and is the process used by almost ¾ of the world's LNG plants. The technology is based on a two-stage refrigerant cooling, where the first cycle employs a single propane refrigerant and the second cycle a mixed refrigerant. The process is powered by two compressors, each of 85 MW power. The first stage is a precooling circuit using pure propane to cool the gas and to precool and partially liquefy the mixed refrigerant at about -30 to -35 ^{o}C. Propane cooling may use up to four flash stages to achieve the desired temperature and is normally conducted with shell and tube type exchangers (spiral wound heat exchangers-SWHEs or wound coil heat exchangers-WCHEs based on in ***Linde's*** technology) with the feed gas and MR. The following stage takes part in a heat exchanging tower and uses to cool and liquefy the natural gas. The process is illustrated in detail in Figures 15 and 16 that follow.

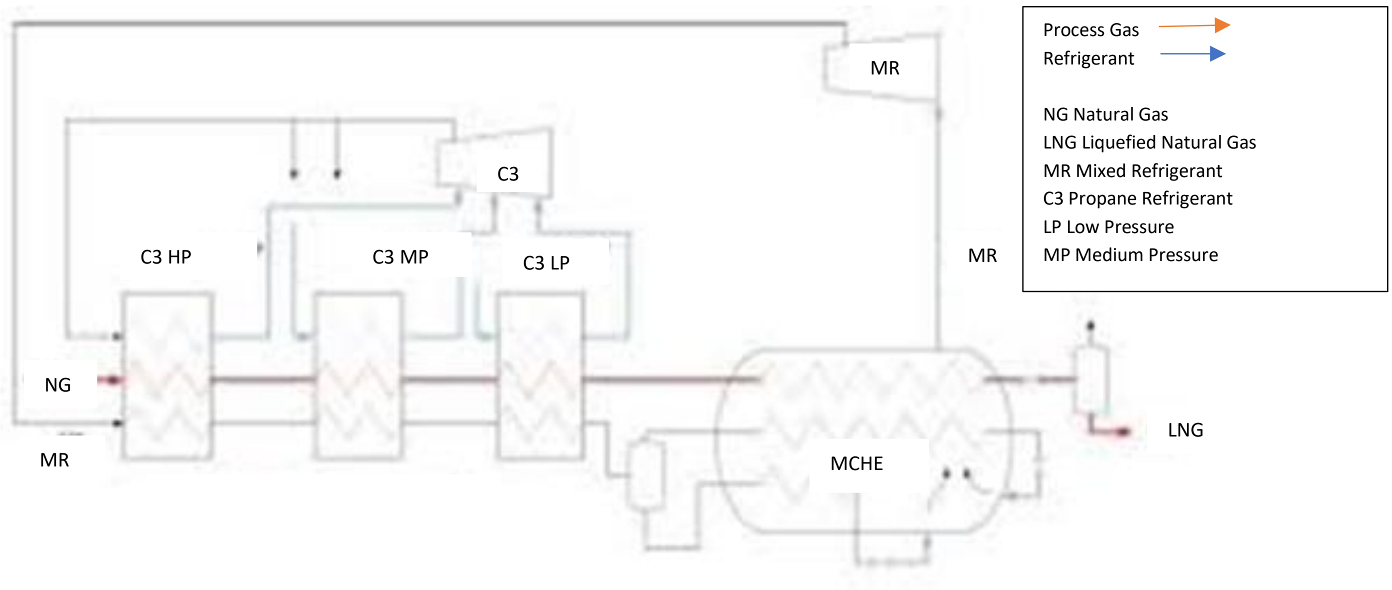

FIGURE 15: APCI C3MR LNG Liquefaction Process [18].

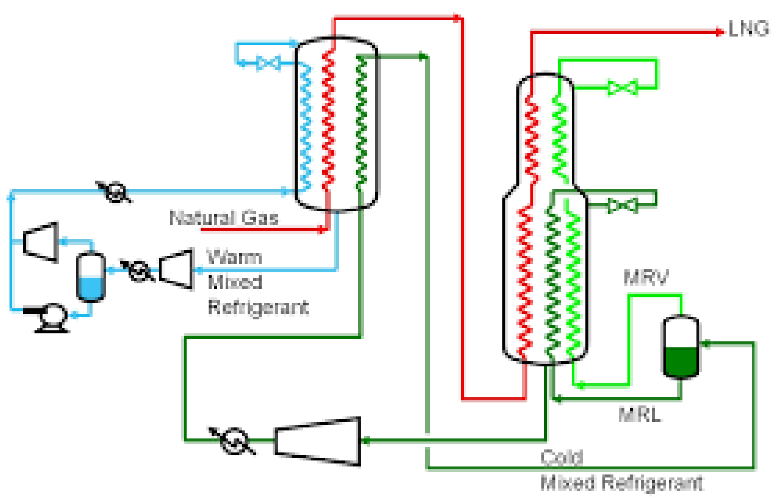

FIGURE 16: Air Products AP-C3MR™ LNG Process [25].

A typical mixed refrigerant consists of nitrogen, methane, ethane and propane. The MCHE is a spiral wound heat exchanger consisting of bundles with thousands of tubes to provide sufficient surface area needed for a close temperature approach between the inlet gas and the cooling medium. These bundles can be classified as warm and cold bundles and are arranged in a vertical shell with the warm bundle on the bottom and the cold on top.

The high pressure mixed refrigerant (HP-MR) is cooled by propane and then separated into two streams, a light and a heavy MR stream, by a flash tank. The light coolant consists of methane, ethane and nitrogen while the heavy coolant (red stream) consists of propane, butane and a small fraction of ethane. The HP-MR and the feed gas streams flow upward through the tube side of the MCHE where the HP-MR undergoes a series of flashes to reduce drastically the temperature. The cold MR produced, flows in the shell side of the MCHE to cool both the inlet gas and the inlet MR. The final cooling stage is accomplished through a J-T (Joule-Thomson) valve or hydraulic expander on top of the heat exchanger, where NG (blue stream) reaches temperatures up to -160 ºC and is liquefied and any excess of nitrogen is removed. The warm vaporized MR stream is taken off the shell side (bottom part) of the exchanger and enters a compressor and a cooler successively in order to get to its original conditions before returning to the LNG exchanger to repeat the cycle [26].

The C3MR process achieves high efficiency due to the ability to match the MR boiling curve to the feed condensation curve. This process is well established and is the one considered for FLNG development. However, the use of propane refrigerant in the precooling loop and the consequent large inventory of propane that is required, especially when kettle-type exchangers are used, are factors to be considered. Another inhibitory factor arises from the relatively large space that is required for the propane evaporators and the sloshing that occurs due to vessel motion, resulting in reduced process performance [25].

APCI has also invented another moderated C3MR process, the APCI-X technology. The difference between the two cycles is that the latter uses nitrogen for the third refrigerant loop instead of MR to cool down the natural gas. Nitrogen addition to the loop, allows for increased production since it reduces the work force of the pre-cooling compressor. The size of the main exchanger can be kept the same, with the subcooling duty shared by the nitrogen cycle. The nitrogen use after the MCHE also helps increasing the LNG capacity.

B.1.2. Black and Veatch Pritchard PRICO® process

A single refrigeration system is the most efficient way to reduce the equipment count. Black and Veatch Pritchard has developed a single mixed refrigerant process, PRICO® (poly Refrigerant Integrated Cycle Operation), which has been successfully used in both baseload and peak shaving LNG plants. The process is composed of a single mixed refrigerant loop and a single refrigeration compression system.

The process goes as follows. Natural gas enters the heat exchanger with a pressure of around 60 bars and temperature of about 12 ^{o}C and is initially cooled at about -35 ^{o}C. At the exit, it is then sent to a separator to remove the heavier components that will be further separated to a fractionation unit while the light components, mostly methane, will be cooled down to liquefaction temperature by the mixed refrigerant.

Natural gas is usually composed of methane, ethane, propane, n-butane and nitrogen. The MR is made up of about the same components as natural gas feed and more specifically by nitrogen, methane, ethane, propane and isopentane. The component ratio is selected in the best possible way to match the boiling point curve of the natural gas. The MR is initially compressed and partially condensed in order to enter the cold box, which consists of numerous platefin heat exchangers-PFHEs. The cold box is capable of cooling multiple streams to extremely low temperatures. When MR exits the cold box is fully condensed and then is flashed across an expansion valve (J-T effect) where the temperature is decreased even further to produce a low temperature two-phase mixture, which is vaporized in the main heat exchanger to cool the natural gas, and a high pressure hot refrigerant. The refrigerant needs to be superheated (by 5-10 ^{o}C) before it enters the compressor to avoid damaging it.

The MR is used to cool down the natural gas in the heat exchanger. When leaving the heat exchanger, the temperature of the natural gas has been reduced to about -155 ^{o}C. The temperature is further decreased (at about -163 ^{o}C) when pressure is lowered to near atmospheric.

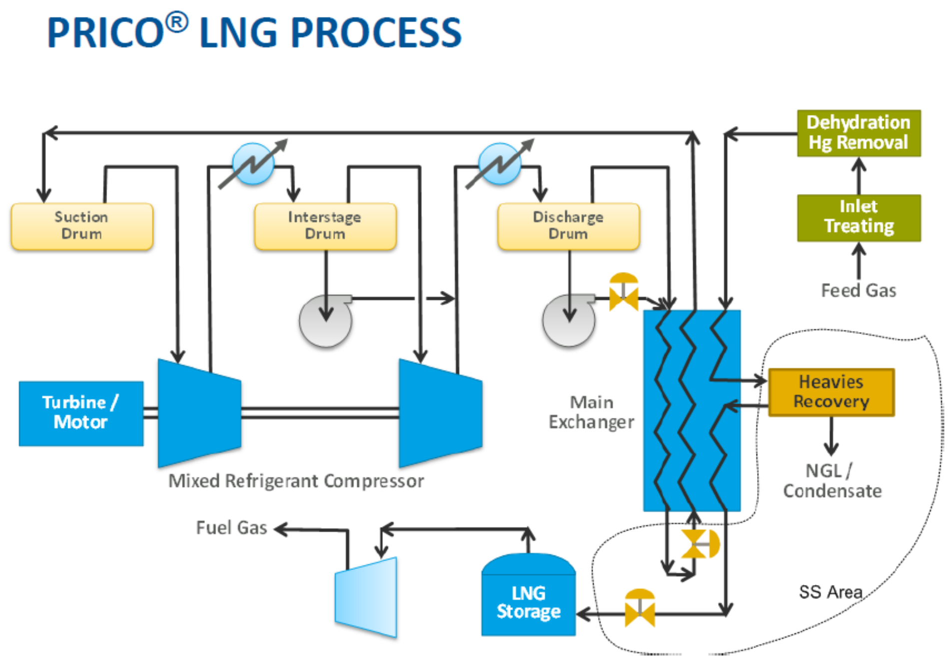

FIGURE 17: Schematic of the PRICO® LNG process [27].

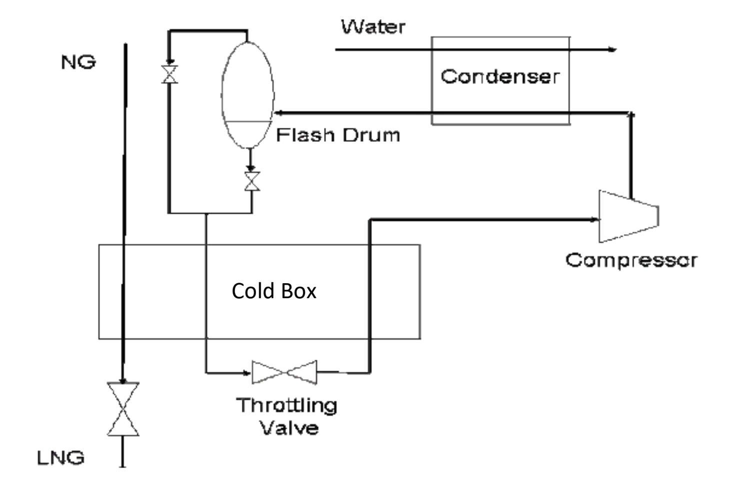

FIGURE 18: Schematic of the PRICO process for LNG production [28].

The advantages of using a single PRICO® refrigeration are mainly the flexibility it offers in handling a wide range of feed gas compositions and the reduction of the equipment necessary compared to the more complex cascade or multiple MR processes, which utilize a significant amount of heat exchangers. PRICO® also simplifies the piping, controls and equipment for the liquefaction unit that translates into capital cost savings of up to 30% [29]. However, the PRICO® single-cycle process is not as efficient as a multiple-cycle process, making it an unattractive option for large scale LNG plants but a desirable method in peak shaving applications, due to its lower capital and operating cost.

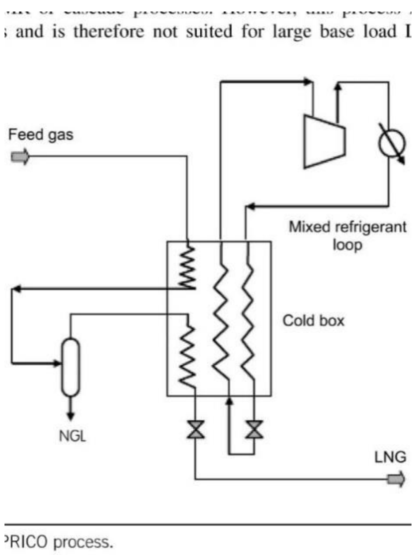

FIGURE 19: Black & Veatch Pritchard PRICO process [30].

B.2. Dual Mixed Refrigerant Cycle

The dual mixed refrigerant cycle (DMR) uses, as the word implements, two independent MR processes in order to liquefy the NG stream. A heavy MR is used to precool the NG in the first cycle and as follows in the second cycle, the cooled NG gets condensed by a lighter MR in a second heat exchanger. The design of the heat exchangers in DMR cycles differs from the size of the SMR. More specifically, since the cooling process is split into two cycles, the heat exchangers are half the size and height of an SMR exchanger. A lot of different versions of the dual-stage cooling cycles with mixed refrigerants have been developed. Worth mentioning are the Liquefin^{TM} process developed by IFP/Axens and the Shell DMR process.

B.2.1. IFP/Axens DMR Liquefin^{TM} Process

The Liquefin^{TM} process developed by IFP/Axens, is a dual mixed refrigerant process where each cycle uses a different mixture of refrigerants. In both refrigeration cycles, MR is compressed, condensed and expanded in one or several steps. The cooling duty generated by the expansions is transferred through Brazed Aluminum Heat Exchangers (BAHX) to the natural gas, eventually allowing its liquefaction.

The composition of each of the MR cycles can be optimized to improve heat exchange in the natural gas liquefaction process. It also allows to balance the refrigeration power from one cycle to the other. This gives the advantage of having identical compressor drivers for the two refrigeration cycles [31]. Apart from the identical compressors, the method has a lot of advantages when used for a large scale LNG plant; it can support capacities from 1.5 to 5+ MTPA. It is very efficient, it has a very compact design and delivers a cost competitive product. The process design is presented in Figure 20.

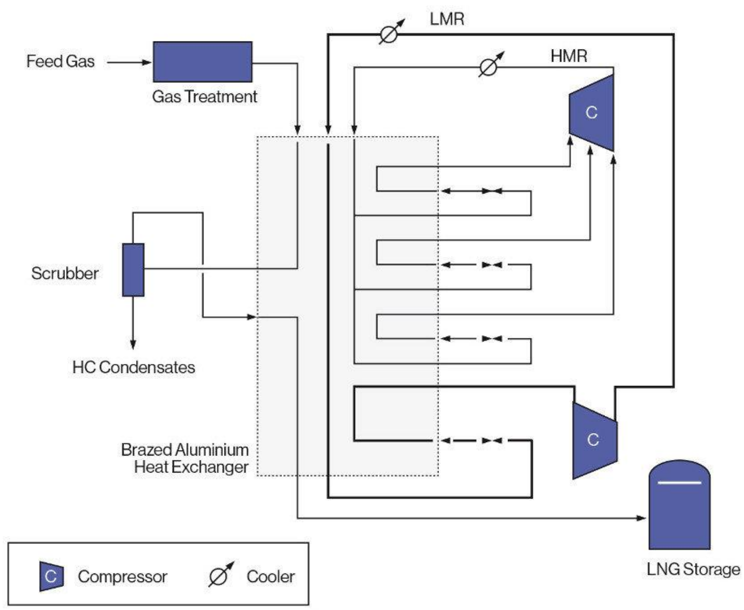

FIGURE 20: Liquefin™ Dual Mixed Refrigerant process [31].

As illustrated above, the feed gas enters the precooling section where an MR drops its temperature to about –50 ºC to –80 ºC in a brazed aluminum plate-fin heat exchangers (PFHEs). The cooled stream that exits the heat exchanger, is then flashed in order to get a chilled, high in methane stream. The natural gas is headed back to the heat exchanger and enters the cryogenic section where it is liquefied with a second MR.

The first MR is used at three different pressure levels to precool the feed gas and the second MR is used to liquefy and sub-cool the process gas. The MR entering the precooling section is completely condensed by the time it leaves the cryogenic section. After leaving the cryogenic section, the refrigerant is expanded and recycled back to the cryogenic section where both the process gas and the refrigerant are condensed.

B.2.2. Shell's DMR Process

The Dual Mixed Refrigerant (DMR) developed by ***Shell*** was based at the original design of APCI, C3MR. This process is suitable for colder climates where the mixed refrigerant can be condensed by heat exchangers that use atmospheric air or cooling water.

Pre-cooling of the gas occurs in spiral wound heat exchangers using a heavy mixed refrigerant. Final cooling and liquefaction of the gas occurs using a light mixed refrigerant. DMR has been applied for the Sakhalin LNG facilities and the Prelude FLNG project. The process is presented in Figure 21.

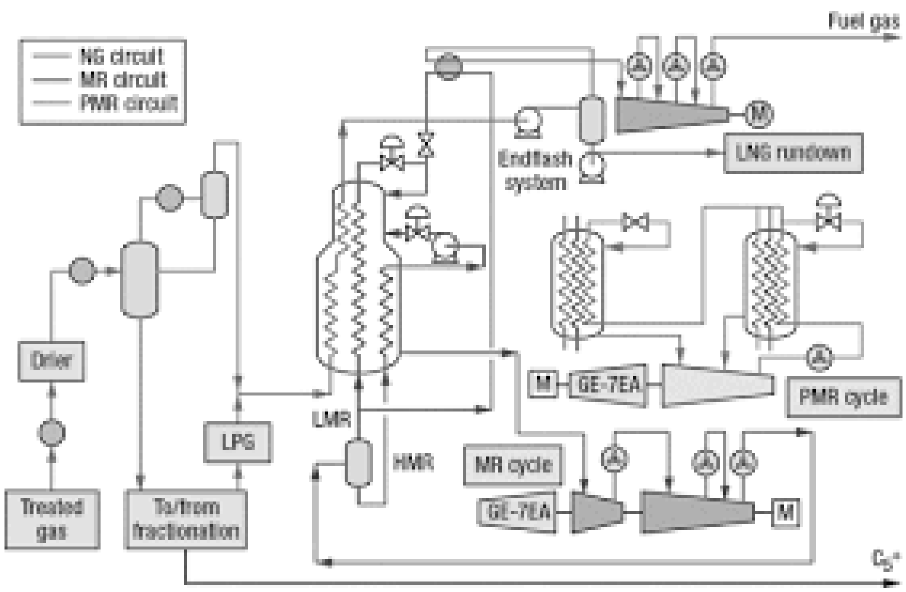

FIGURE 21: Shell's DMR process [29].

The process operates with two separate mixed refrigerant cooling cycles, one for precooling of the gas to approximately –50 ºC and one for the sub-cooling and liquefaction of the gas, so that it is possible to change the load on each cycle. Shell's DMR process was initially deployed at the Sakhalin project in Russia; a base load plant with two 4.8 MTPA trains, SWHEs and favorable cold climate conditions to enhance the air cooling. As mentioned before, the DMR process is suitable for cold climates since the mixed refrigerant can be precooled by exploiting the cold weather conditions and avoid a significant part of the propane cycle. SWHEs are preferred rather than the typical core-in-kettle or plate-fin heat exchangers. The cooling duty for liquefaction of the natural gas is provided by a second mixed refrigerant cycle. The refrigerant of this cycle consists of a mixture of nitrogen, methane, ethane, and propane. The vapor leaving the MCHE passes through an axial compressor and a two-stage centrifugal compressor.

The DMR process was further developed into an electrically driven Parallel Mixed Refrigerant PMR design [33], using a parallel line-up of electrically driven refrigerant compressors and SWHEs for the liquefaction. Even though the cost of electrically driven LNG trains is higher than those operating mechanically, the availability of powerful electrical motors breaks it even. In addition, the electric drivers offer flexibility in the size and speed of their design, a higher vendor base and maybe a chance of reducing the plant's CO_2 footprint by operating with highly efficient electric power generation plants [34].

C. Gas Expander Cycles

The last category of refrigeration cycles used for LNG production is the gas expander cycles. Before 1996 conventional liquefaction processes for natural gas operated with the high pressure condensation phase flow of liquefied natural gas expanding across a Joule-Thomson valve. Replacing the JT-valve with a cryogenic LNG liquid turbine substantially increased the thermodynamic efficiency of the refrigeration process resulting in a total LNG output increase of 3 to 5% [35].

The expander-based concept builds on different layouts of the reverse Brayton cycle, where the working fluid is compressed and expanded through a turbine to generate the refrigeration effect. One, two or multiple turbo-expanders can be used for the process, driven by electric motors or gas engines, offering high efficiency; typically, over 85% [36]. The refrigerant used is a light volatile component, methane or most commonly nitrogen, which always stays in the gaseous phase and is a better option for low temperature cooling. The heat curves for the gas expander cycles, present significant differentiations in the temperatures between the refrigerant and the cooling gas at the end of the natural gas cooling curve. This observation is the reason for working with the low temperature cooling range rather than the high temperature range in the beginning of the gas cooling, and it helps particularly when the feed gas contains significant amounts of C_{3+} components.

The refrigerants remain in the gaseous state throughout the expansion cycle. Being a single component, the process is simplified since there is no need to adjust the composition compared to hydrocarbon mixtures that will suffer phase change throughout the process. In addition, because the heat exchangers operate with relatively wide temperature approaches, they are less sensitive to changes in feed gas compositions. Hence, precise temperature control is not as critical as it is for mixed-refrigerant cycles. This cycle is considered to be more stable over a range of liquefaction conditions but it is the least efficient refrigeration method compared to the ones mentioned before [37]. When increasing the number of expanders, significant mechanical design challenges occur in order to handle the increasing volumetric flow and change in density whilst providing flexibility of operating range. The low efficiency of this cycle makes it more suitable for small-scale LNG plants (such as BOG liquefaction) rather than large baseload plants. Being a gas phase cycle, the system performance is not much impacted by ship motion and is more suitable for ship-based floating liquefaction plants (FLNG).

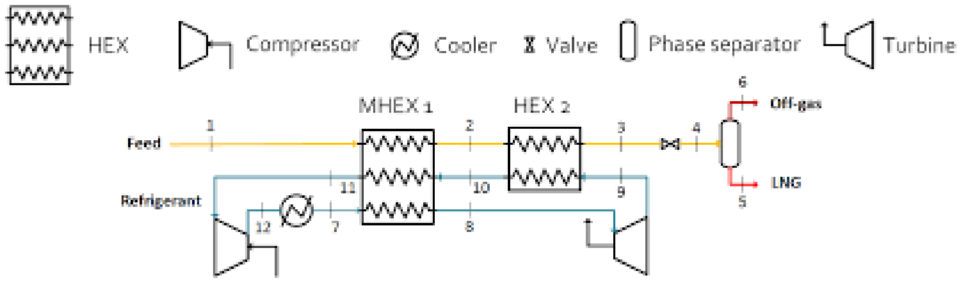

FIGURE 22: Single-expander configuration with one compression stage [38].

A simple configuration of the process is presented above, where nitrogen is initially compressed and cooled before entering the cold box. A compressor is used to liquefy and sub-cool the natural gas through expansion. Nitrogen at the turbine outlet is in super-heated or saturated conditions. A dual-expansion process entails a two-stage compression process, such that the intermediate-pressure and low-pressure nitrogen streams mix without major energy loss [38].

A.4.2. Liquefaction in arctic climates

In some arctic and subarctic regions there are significant amounts of exploitable natural gas and production has already started at some of them, like the Snøhvit LNG plant in Norway, the Sakhalin project in eastern Russia and the Kenai in Alaska. The natural gas that is produced in such regions is mainly liquefied and transported through sea to LNG consumers. LNG plants in arctic regions have to overcome design and operational difficulties, yet the low temperatures can actually be helpful in liquefaction steps by reducing the power consumption.

The step that is mostly affected by the environment’s temperature is the precooling stage where most of the heat is rejected to ambient air. Cooler or arctic climates can lower the refrigeration requirements due to the work of the precooling stage. Seasonal variations in temperature, especially in arctic climates, are extremely wide (from 26 ^{o}C in the summer to -18 ^{o}C in winter in the Sakhalin LNG plant), affecting the LNG production. During winter time, the air gets denser due to the low temperatures and the gas turbines have a higher power output one can take advantage of. The excess in LNG production during winter months could also cause a logistics issue since shipping could be delayed due to bad weather conditions. Hence, the increased LNG production during cold months is taken into consideration and liquefaction plants are usually not designed to utilize the full capacity of the drivers during winter.

The variations in temperature are dealt by adopting a DMR process and using MRs for precooling instead of a single pure refrigerant, varying their composition to match natural gas’s cooling curve according to the environment’s temperature and minimize energy consumption. Waste heat in all cases should be recovered and used to heat certain parts of the equipment that use water in order to keep it from freezing.

Other issues that occur in colder climates have to do with natural phenomena (i.e. snowfall, high winds) that make it difficult for the personnel to transport and the equipment to operate properly. Snow and ice that accumulates on building and structure designs, forces a significant load and can have a great impact on the integrity and operation of each unit. Constructing an LNG plant indoors in a heated and ventilated environment could protect the rotating equipment (pumps, generators, turbines and compressors) and prevent fluid freezing, wax and hydrate formations and liquid losses [39]. Replacing the propane refrigerant with a lower boiling point gas (e.g. ethane or ethylene) or a multicomponent mixed refrigerant could also reduce the precooling temperature which in turn will allow a better power balance and better machinery selection between precooling and liquefaction duty. Thus, in winter the lower ambient air can be used to condense a lighter refrigerant at a lower temperature, reducing the work forced by compressors [40].

At colder climates the cooling medium selection is also very important when designing an LNG plant. The choice to be made is between air and seawater and could significantly affect the cost when designing an LNG plant. The temperature of seawater stays relatively stable when compared to the air during seasonal changes. As mentioned before, the cool air could increase production by increasing the power output of the turbines, but due to wide variations in air temperature from winter to summer, the opposite effect occurs during summertime. On the other hand, seawater temperature seems to stay relatively constant throughout the year, but it doesn't lead to much increased production during winter and it can also have a negative effect in operations if frozen. Warm seawater discharge is also an issue since it could affect a fragile marine ecosystem. For these reasons, with financial gain being the major factor, air cooling is usually preferred over seawater cooling.

A.4.3. Refrigeration cycles comparison

In order to decide which refrigeration method should be used in an upcoming LNG plant, a number of factors should be taken into consideration, such as the size, the location, with the most important being the power consumption. In a liquefaction cycle, the largest cost corresponds to the compressors used for expansion and compression of the refrigerant, which due to their operation are very energy consuming. Hence, the power consumption of the compressors is a factor which needs to be examined in every case. The second most energy consuming unit in a liquefaction plant is the heat exchangers configuration. The cost of the heat exchangers can be estimated only by knowing the surface area needed for the heat transfer. Especially MCHEs, which are capable of achieving heat transfer between multiple streams, come with large heat exchanging surfaces, resulting in their high cost. Another factor that should be examined when comparing different refrigeration methods, is the cooling curves of the refrigerant and the natural gas. It is the most commonly used comparison in liquefaction technologies and as mentioned before, the more the refrigerant's cooling curve approaches the one of the natural gas, the less are the heat losses in the process. However, this comparison alone is not enough to conclude which refrigeration cycle is the best. Financial analysis and life-cycle costs in each case, are essential to make a comparison which corresponds closer to the reality.

A.5. Offshore natural gas liquefaction

The need to exploit remote, smaller gas fields and the recent developments in the LNG industry, have led in the evolution of FPSO (Floating Production Storage and Offloading) and FLNG (Floating Liquefied Natural Gas) units, which are offshore facilities that directly produce and store LNG [65].

An **FPSO** unit is a floating vessel used for the production and processing of oil or gas. Even though it has storage capacity, it can only store oil. FPSOs are preferred in offshore locations due to their easy installation.

FLNG facilities are moored directly above a natural gas field and route the gas from the field to the facility with drilling risers, conduits that unite a subsea well to a surface drilling facility. The gas is then processed following the same steps as it would in an onshore facility. The first stage of processing involves the production of natural gas, gas condensates and Liquefied Petroleum Gases (LPG-ethane, propane, butane). The natural gas is then exempt from impurities, sweetened and finally liquefied. The LNG that is produced, is stored in proper tanks. Offloading of LNG and LPG occurs from carriers that will deliver the products to the final consumers.

Liquefaction technologies for offshore units, do not differ from the ones used in onshore installations. However, a refrigeration method in an offshore unit, is not chosen by its efficiency, capacity and low cost of installation and operation. Instead, environment is the primary factor to be considered. An offshore facility tries to minimize its footprint, due to the direct contact with marine life and also minimize the number of operating units and their weight for easier installation. An offshore installation requires a compact module and needs to provide easy access in the equipment for maintenance.

B. Case Study:

Simulation and technoeconomic analysis of C3MR liquefaction process.

A. Liquefaction process selection

In this diploma thesis, the **C3MR** liquefaction process (***APCI's*** technology) was modeled in HYSYS environment, using large **Wound Coil Heat Exchangers** (LNG exchangers), for a baseload LNG plant of **3.1 MTPA** capacity.

A wound coil heat exchanger (WCHE), as designed by ***Linde***, consists of several layers of tubes that are helically wound around a large center tube, known as a mandrel. This tube bundle can be designed in such a way that it can contain more than one tube fraction to accommodate different process media. Both ends of the tubes are welded to tube sheets or ring pipes. A pressure vessel, which is manufactured in parallel to the tube bundle, encloses the entire bundle. This type of exchangers is ideal for natural gas liquefaction plants as it can operate in temperatures ranging between –269°C to +650°C [46].

A large scale LNG project, of more than 1 MTPA, can take place close to a natural gas reservoir of significant amount of reserves to minimize the cost of intermediate transport infrastructure and gas shrinkage due to transport. More than 75% of LNG plants, are using the propane pre-cooled Mixed Refrigerant cycle for natural gas liquefaction [18,41].

Constructing a large scale LNG project is of high financial risk, hence contracts and concessions for extended periods are essential before the investment decision. The APCI (Air Products and Chemicals, Inc.) developed technology of C3MR is one of the most mature liquefaction technologies (more than four decades of operating experience) and is considered a cost-effective and reliable method for large baseload LNG plants.

B. Site selection

Greece is considered to have significant amount of oil and gas reserves. According to Professor Antonis Foskolos and previous research on the site [42,64], the basin of the eastern Mediterranean is estimated to have 48 tcf of gas reserves, 27 of which are attributed to Greece, excluding the gas reserves that exist between Crete and Cyprus and the natural gas reserves that exist in the Ionian Sea, the western Greece and Thermaikos.

In the past few years, Crete has been in the center of interest for extraction of natural gas, since it is estimated that south of Sfakia and west of Selino, natural gas reserves are of high economic interest to be extracted and transported. The possibility of natural gas reserves south of Crete has led French and American companies (Total and ExxonMobil) to invest on the site and a possible production is estimated to start as early as 2022-2023.

In this study, the region between south of Sfakia and west of Selino, was selected as the site where the liquefaction simulation will take place. This site is in contact with the Cretan Sea that makes it a desirable spot for transportation of LNG offshore through LNG vessels and could also be a desirable location for the erection of FLNG facilities. Another benefit this site offers, is the cold weather during winter, which could further favor the production of LNG.

C. Model development

The simulation was modeled in Aspen HYSYS software. HYSYS is a modelling software used for steady-state and dynamic operations and is widely used in the oil and gas industry. The software contains a broad database of thermodynamic and chemical properties of chemical compounds and pure components. After the selection of components used in the simulation, a fluid package needs to be determined (equation of state) in order for the software to make calculations of pressure, temperature, density, enthalpy and other properties.

The simulation is based on mass, material and energy balance to provide an outcome that does not defy the thermodynamic laws (e.g. degrees of freedom). Specification of flow rate, composition, operating conditions (pressure and temperature) of the inlet streams and operating conditions into the process, result in calculation of material and energy streams, calculation of all process conditions and sizing of the unit operations. The sizing of the equipment can further be used to calculate the cost of the whole process. In this study, HYSYS was used to simulate APCI's C3MR liquefaction process using Wound Coil Heat Exchangers. The results and conditions of the simulation are presented below [41,43].

C. Process Flow Diagram (PFD)

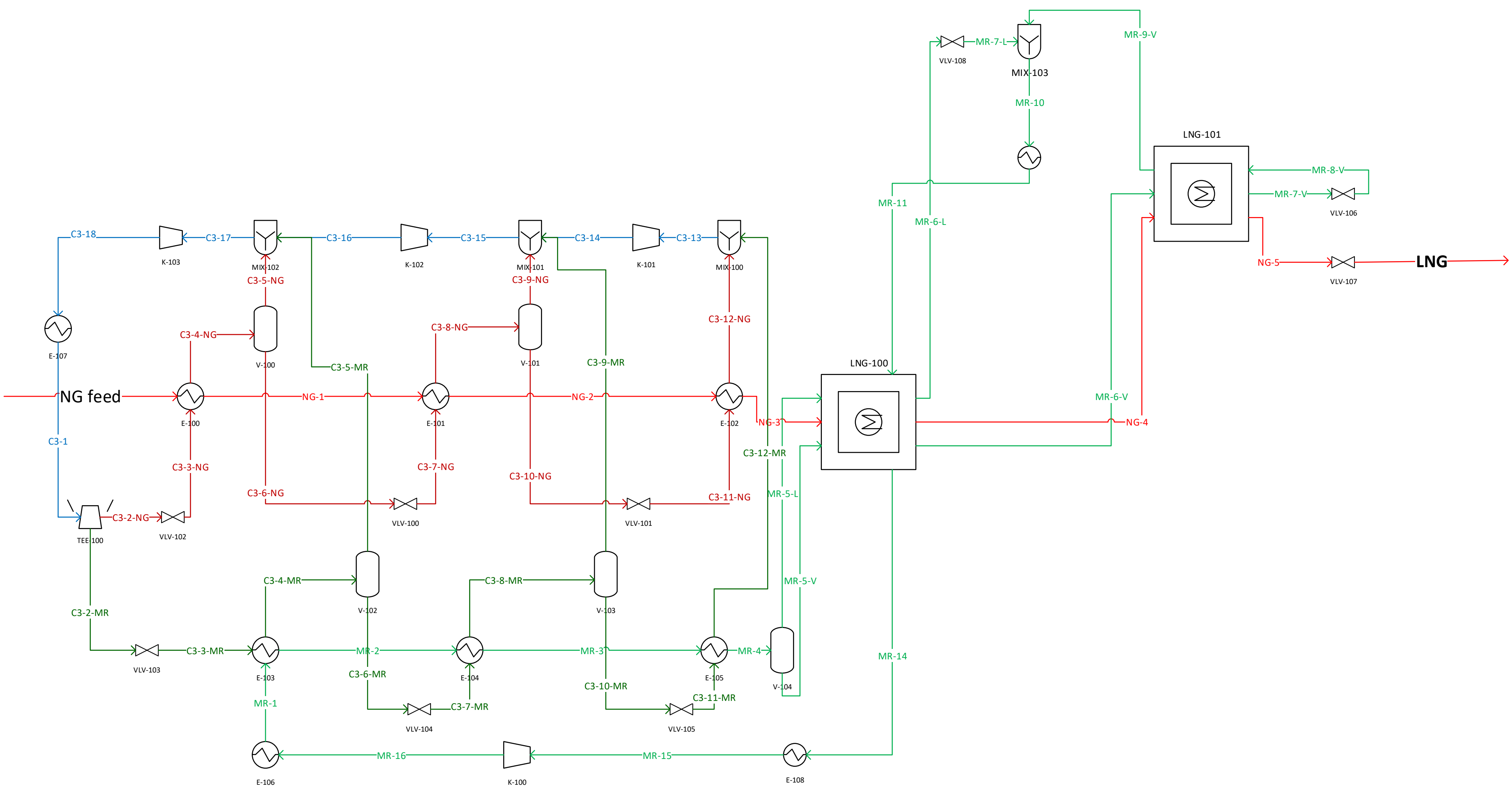

FIGURE 23: Process Flow Diagram of the C3MR process.

D. Results and Discussion

TABLE 6: Process Parameters.

Natural Gas	Value
NG inlet pressure	4000 kPa
NG inlet temperature	30 ºC
NG feed	3.102 MTPA
Mixed Refrigerant	
MR inlet pressure	4790 kPa
MR inlet temperature	30 ºC
MR feed	8.156 MTPA
Propane	
C3 inlet pressure	1081 kPa
C3 inlet temperature	0.6003 ºC
C3 feed	3.102 MTPA

TABLE 7: Natural gas stream composition after passing through the sweetening unit (elimination of hydrogen sulfide content) [44].

Composition	%mole concentration
Methane	89.7
Ethane	5.5
Propane	1.8
n-Butane	0.1
Nitrogen	2.9

TABLE 8: Mixed Refrigerant composition [44].

Composition	%mole concentration
Methane	45
Ethane	45
Propane	2
Nitrogen	8

The PFR in Figure 23 illustrates the C3MR liquefaction process as it was modelled in HYSYS. The propane stream (light blue color) enters a splitter unit (TEE-100) where it is divided into two streams; one to precool the natural gas (dark red color streams) and one to precool the MR (dark green color streams). The composition of the natural gas is assumed after passing through the sweetening unit and the composition of the MR are based on previous work of Dag-Erik Helgestad [44]. The inlet conditions of the three main streams, the natural gas, the MR and the propane are shown in Table 6.

Heat exchangers E-100, E-101 and E-102 use one of the split propane streams (C3-NG) to pre-cool the feed gas (bright red color stream) before entering the main cryogenic heat exchanger (MCHE), at a temperature of -32.91 ^{o}C. The stream that exits each heat exchanger is led to a 2-phase separator where the liquid outcome is transferred to a mixer with other liquid components to be recycled in the process, while the gaseous phase passes through a valve and is used in the following heat exchanger unit. Valves are used throughout the whole process to regulate the pressure of streams entering units that operate within a certain range of pressure.

The other half of the split propane stream is used to pre-cool the MR (light green color stream) to a final temperature of -32.09 ^{o}C before entering the MCHE. This stream follows the same path as the previous, entering successively heat exchangers (E-103, E-104 and E-105) while in between, vapor gets separated from the liquid with the liquid led to be recycled and vapor being used for heat exchange.

Between the mixers MIX-100 to MIX-102, compressors (K-101, K-102 and K-103) are used to decompress the propane stream and a cooler (E-107) to bring it to its inlet conditions (stream C3-1, temperature: 0.6 ^{o}C, pressure: 1081 kPa).

The MCHE unit consists of two Wound Coil Heat Exchangers (LNG-100, LNG-101), where the feed gas is chilled to meet final product's specifications. Each LNG exchanger is capable of utilizing more than two streams for heat transfer. LNG-100 has three inlet streams; a vapor and a liquid MR stream and the pre-cooled natural gas stream. The gaseous MR stream and the feed gas exiting LNG-100 enter LNG-101. The liquid MR stream passes through a valve to become gas again by pressure reduction and then enters a mixer (MIX-103), where it mixes with the outlet vapor MR stream from LNG-101. The outlet stream then is heated passing through a heater (E-109) and reenters the LNG-100. After exiting the LNG exchanger, the stream passes through two coolers (E-108 and E-106) and a compressor (K-100) to recompress and cool to its inlet conditions (MR-1 stream, temperature: 30 ^{o}C, pressure: 4790 kPa). At the second LNG exchanger, the vapor MR stream that exits the unit is depressurized passing through a valve and led back to the heat exchanger as an inlet (recycling of the streams). The final step is to bring the liquified natural gas at atmospheric pressure which occurs after passing the final NG stream through a valve. The temperature of the final LNG is -162 ^{o}C.

D.1. Liquefaction unit cost

Valves:

TABLE 9: Operating conditions and equipment cost of pressure regulator valves for gas and liquid stream [45].

	Material	Diameter [m]	Pressure Drop [kPa]	Cost of Equipment [USD]
VLV-100	SS	0.05	216.4	1,458
VLV-101	SS	0.05	120.5	1,458
VLV-102	SS	0.05	597.8	1,458
VLV-103	SS	0.05	597.8	1,458
VLV-104	SS	0.05	216.4	1,458
VLV-105	SS	0.05	120.5	1,458
VLV-106	SS	0.05	3,100	1,458
VLV-107	SS	0.05	2,719	1,458
VLV-108	SS	0.05	3,650	1,458
				13,122

<u>Mixers:</u>

TABLE 10: Operating conditions and equipment cost of double-arm sigma mixers (jacketed nonvacuum steel units with class I, group D motors) [45].

	Material	Product Temperature [ºC]	Product Pressure [kPa]	Capacity [m^3]	Cost of Equipment [USD]
MIX-100	SS304	-19.96	116.3	1.9296	13,560
MIX-101	SS304	-5.41	246.8	4.4706	21,493
MIX-102	SS304	17.42	473.2	6.1566	25,615
MIX-103	SS304	-130.9	490.2	6.4998	26,388
TEE-100	SS304	0.6	1,081	6.1566	25,615
					112,671

Separators:

TABLE 11: Operating conditions and equipment cost of 2-phase separators [45].

Separator Unit	**V-100**	**V-101**	**V-102**	**V-103**	**V-104**	
Temperature [ºC]	-0.021	-19.7	-0.019	-19.7	-32.09	
Pressure [kPa]	473.2	246.8	473.2	246.8	464.0	
Orientation	Vertical	Vertical	Vertical	Vertical	Vertical	
Height [m]	6.4008	5.0292	11.1252	5.6388	11.7348	
Diameter [m]	2.1336	1.6764	3.6576	3.9624	3.9624	
Internal separation	Liquid-Vapor	Liquid-Vapor	Liquid-Vapor	Liquid-Vapor	Liquid-Vapor	
Material of Construction	CS	CS	CS	CS	CS	
Cost of Equipment [USD]	38,406	33,818	57,134	45,342	60,625	**235,325**

Compressors:

TABLE 12: Operating conditions and equipment cost of centrifugal, motor-driven compressors [45].

Compressor Unit	K-100	K-101	K-102	K-103	
Molar Flowrate [kgmole/h]	3,900	6,268	14,520	20,000	
Driver Power [MW]	34.067	3.708	7.751	13.871	
Performance	0.75	0.75	0.75	0.75	
Material	SS316	SS316	SS316	SS316	
ΔT [ºC]	74.3	32.64	29.26	39.94	
ΔP [kPa]	2,844	130.5	226.4	617.8	
Cost of Equipment [USD]	45,704,510	4,975,345	10,131,270	18,609,315	**79,420,440**

Heat Exchangers and Coolers:

TABLE 13: Operating conditions and equipment cost of the heat exchangers and coolers [45].

	Material	Operation	Duty [MW]	Heat Transfer Area [m^2]	Cost of Equipment [USD]
E-100	SS	Heat Exchanger	6.669	25,903	589,680
E-101	SS	Heat Exchanger	4.591	17,831	456,845
E-102	SS	Heat Exchanger	3.747	14,553	397,611
E-103	SS	Heat Exchanger	18.222	70,773	1,476,409
E-104	SS	Heat Exchanger	27.438	106,570	2,223,121
E-105	SS	Heat Exchanger	27.383	106,354	2,218,664
E-106	SS	Cooler	52.027	6,238.76	1,972,200
E-107	SS	Cooler	113.166	4,535.75	847,200
E-108	SS	Cooler	-32.472	83.2	26,000
E-109	SS	Heater	-6.691	17.14	5,358
					10,213,088

Wound Coil Heat Exchangers:

A Coil Wound LNG Exchanger as designed by *Linde* has typical dimensions of:

Length: 60 m

Diameter: 5 m

Active area: 35,000 m²

Design temperature: –269°C to +650°C

Design pressure: up to 300 barg shell side and up to 1,400 barg tube side [46].

TABLE 14: Equipment cost of the heat exchangers and coolers [47].

	Heat Transfer Area [m^2]	UA [$\times 10^7$ kJ/^{o}C h]	Cold Duty Exchanger [MW]	Cost of Equipment [USD]
LNG-100	35,000	4.672	150.638	4,113,500
LNG-101	35,000	5.779	21.880	4,113,500
				8,227,000

Utilities:

TABLE 15: Cost of utilities used for the units' operation.

	Fluid	Rate	Rate Units	Cost [USD/y]
Electricity		524.321	MW	355,964,604
Cooling Water	Water	35.309	MTPA	1,119,326
Refrigerant-Freon 12	Refrigerant	21.655	MTPA	4,058,087
Steam @100psi	Steam	495.435	MTPA	8,890,874
Total				**370,032,891**

The **Total Cost of Equipment** based on the previous calculations, excluding the utilities, is estimated at: **98.222×10^6 USD.**

The Total Cost that occurs after the installation of the equipment (**Cost of Installed Equipment**), according to the assumption that the Installation accounts for 43% of the Equipment cost, is estimated at: **140.457×10^6 USD [48].**

Figures 24 and 25 that follow, present a brief comparison between the operating units and the proportion of the total equipment cost each unit occupies.

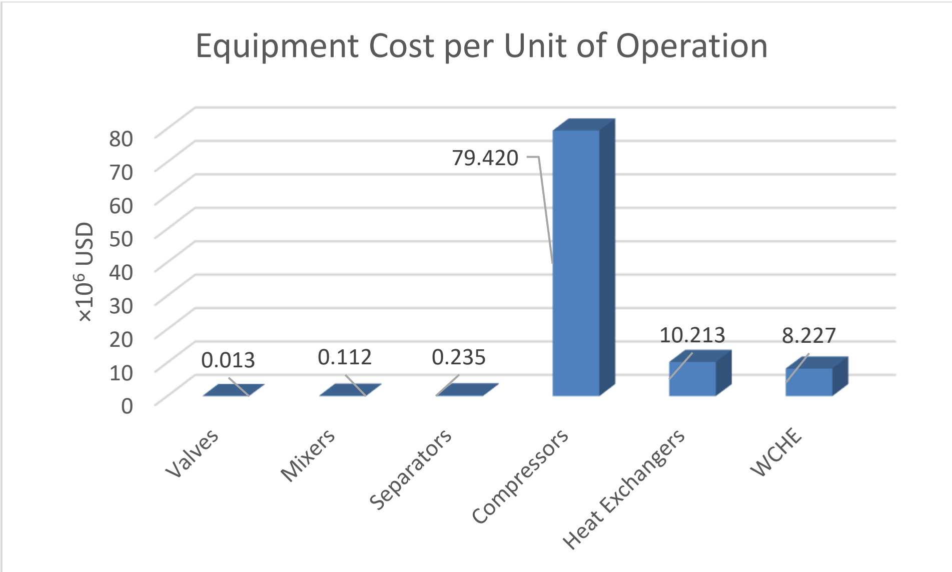

FIGURE 24: Estimated Cost of Installed Equipment for each operating unit.

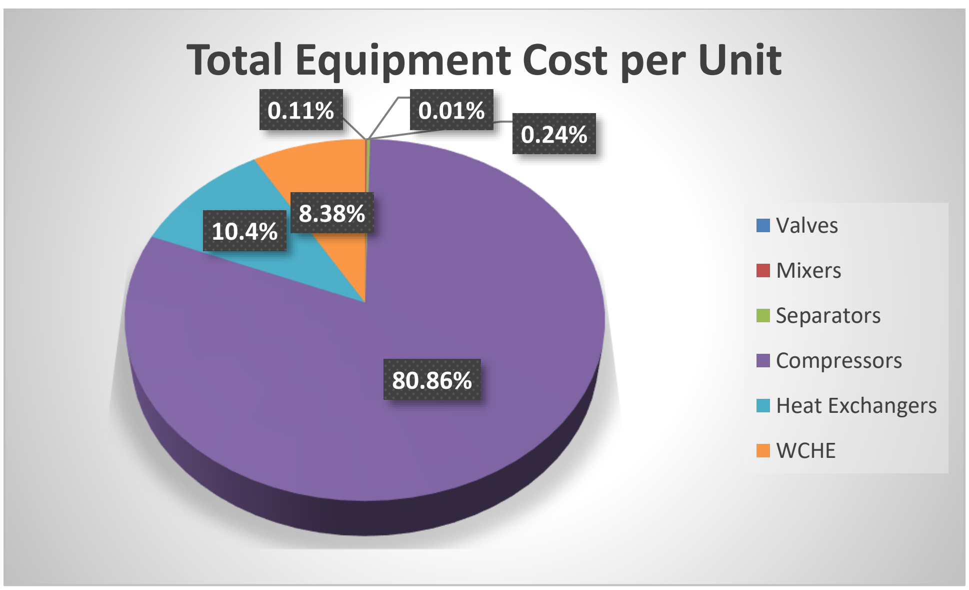

FIGURE 25: Analysis of the equipment contribution to the Total Liquefaction Unit Cost.

It is clear that the majority of the equipment cost corresponds to the compressors' operation (80.86%) with the total of heat exchangers (10.4%) and WCHE (8.38%) to follow, while the rest of the equipment is of minor importance when compared to the total cost of equipment. The operation and cost analysis of the compressors are presented below.

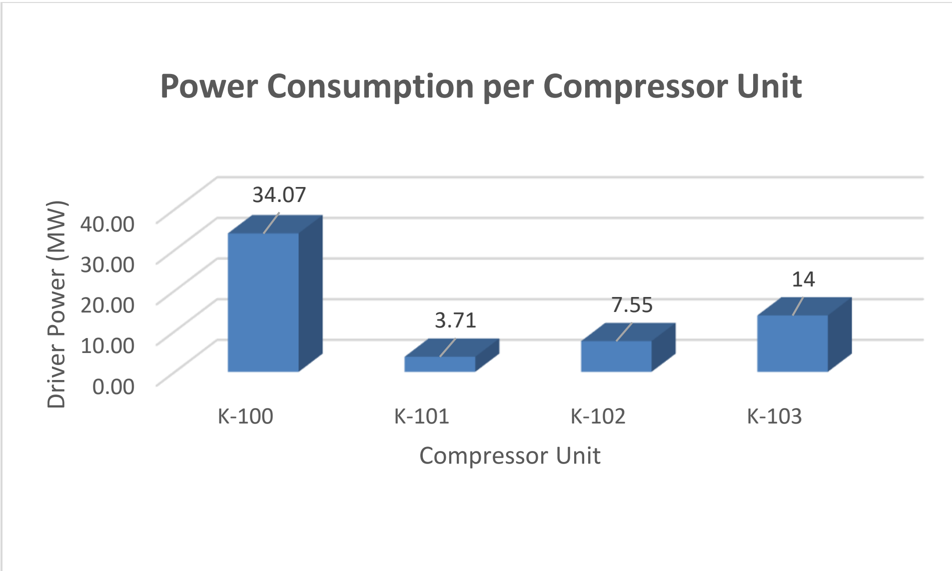

FIGURE 26: Power Consumption of each Compressor Unit.

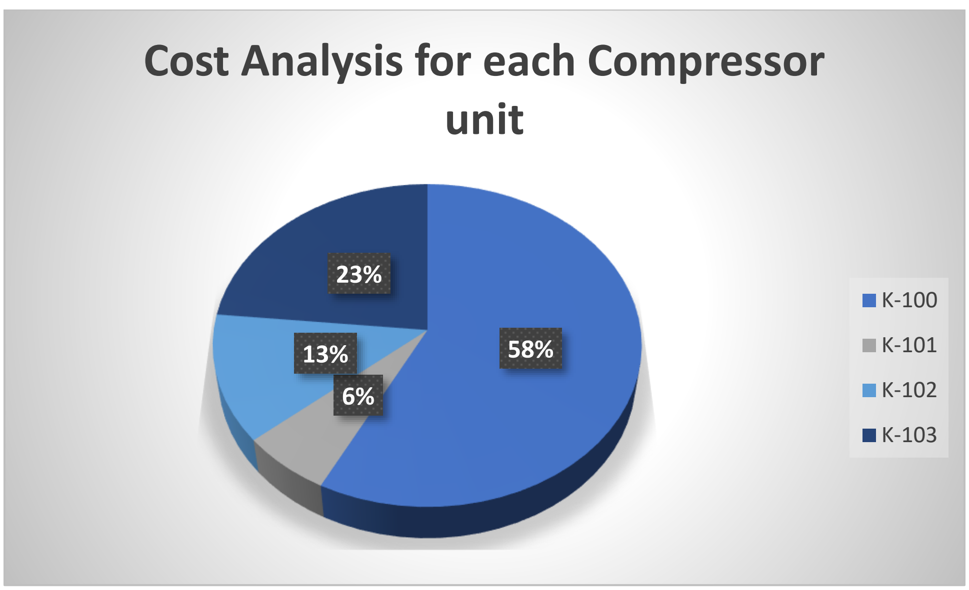

FIGURE 27: Quantitative cost analysis of each compressor unit to the total compressor cost.

From Figures 26 and 27, it occurs that compressor K-100 has a significant power consumption and operates to achieve a very high pressure difference (2,844 kPa), which results in its high cost. Unit K-100 contributes the most at the total equipment cost and at the total cost of compressors (58%). The high power required for the operation of this unit, is due to the recompression of the MR stream (stream MR-18) that exits the LNG exchanger (LNG-100) which is recycled in the process. In order to recompress this stream, multiple compressors could be used in order to reduce the workload of unit K-100, with the addition of coolers in between to further minimize their workload. However, since HYSYS allows an operation of this magnitude and the same process in the industry utilizes 85 MW compressors, in this diploma thesis it was decided to continue with one compressor. For future reference, it would be advised to spread the workload of recompression to more than one compressors.

Unit K-103 constitutes the 23% of the total cost of compressors and units K-102 and K-101 follow with 13% and 6% respectively. The work of these three compressors is to recompress the pure propane stream (streams C3-13 to C3-18), which will be recycled in the process and used to precool the MR and NG streams.

Heat Exchangers also influence the total cost equipment at an 18.8% as presented in Figure 29. A cost analysis and the heat transfer area of each heat exchanging unit, are presented below.

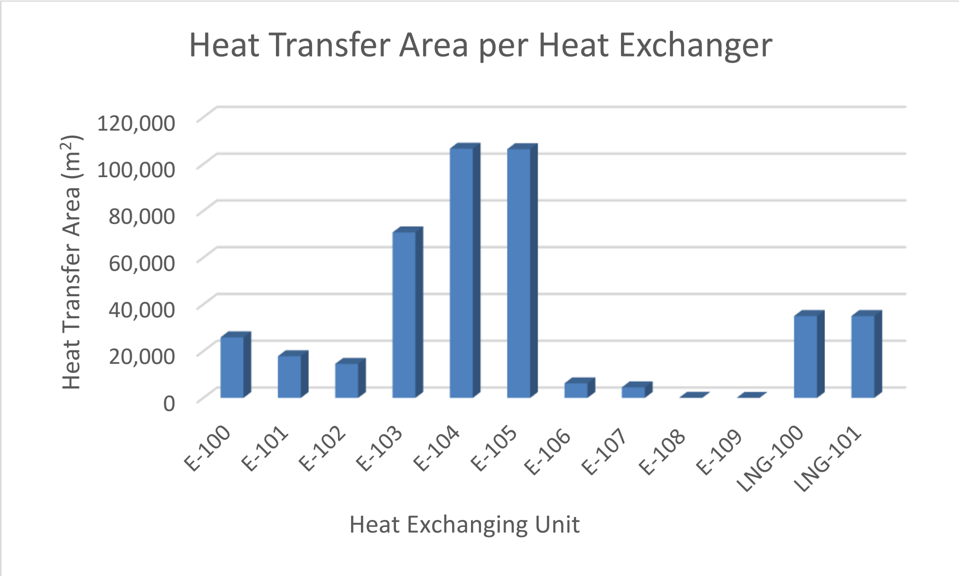

FIGURE 28: Heat transfer area per Heat Exchanging Unit.

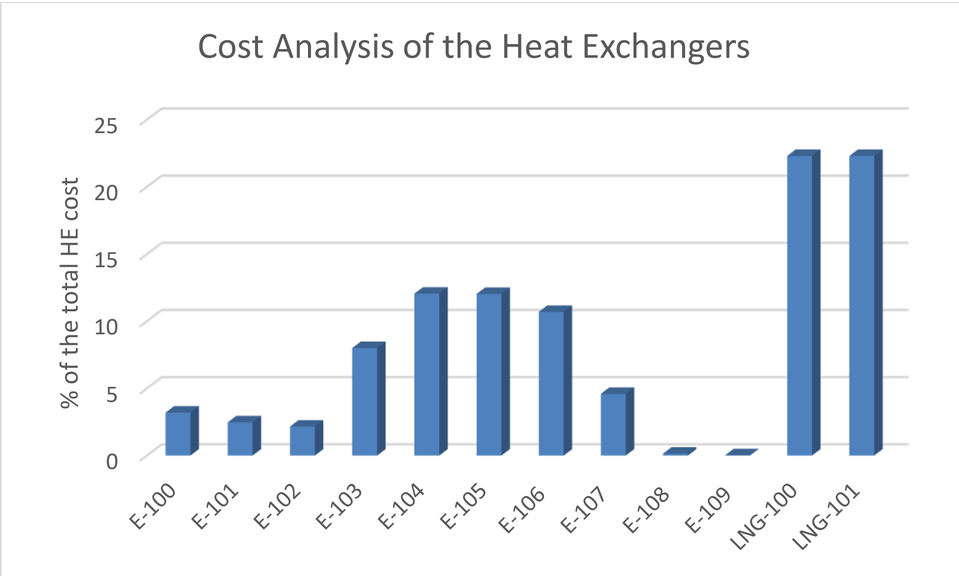

FIGURE 29: Cost Analysis of the heat exchanging units to the total Heat Exchangers' cost.

As seen from Figures 28 and 29, units LNG-100 and LNG-101 represent almost the 50% of the total heat exchanging cost. This conclusion is consistent since these LNG exchangers are part of the MCHE and achieve heat transfer between multiple streams. Hence, even though they have a lower heat transfer area than other units, their cost remains high because they manage temperatures and pressures from multiple streams. Heat exchangers E-100 to E-102 are used to precool the NG stream before it enters the MCHE and E-103 to E-105, are used to precool the MR stream. It is worth mentioning that units E-103 to E-105, which precool the MR stream, present the highest surface areas offered for heat transferring. Unit's E-103 surface area is almost double the size of the LNG exchangers (70,773 m^2), while units E-104 and E-105 have a three times larger surface area than WCHEs (~106,000 m^2). These extreme deviations concern only the MR stream and are logical when considering that the MR works to bring natural gas at cryogenic conditions, producing the final product; MR does most of the heat transferring.

D.2. Final Cost Analysis for the Liquefaction Unit

Revenues:

The revenues occur from exporting LNG. It is estimated that 1 ton of LNG is equal to about 52 MMBtu [49].

The LNG production from the liquefaction plant is 3.102 MTPA, which according to the previous correlation, corresponds to 161.304×10^6 MMBtu.

As of October 2018, the average landed LNG price for Europe (United Kingdom) was 8.8 USD/MMBtu [50].

Based on these data, the annual revenues from exporting LNG would be **1.420×10^9 USD**.

Expenses:

TABLE 16: Total Expenditures of the Liquefaction Unit per annum. *

	Annual Expenses	**$\times10^6$ USD/y**
Cost of the Raw Materials*	Mixed Refrigerant	0.567
	Propane	123.659
Cost of Utilities		370.033
Cost of Labor*		2.210
Total Expenditures		**496.469**

*Estimations are analyzed in the Appendix.

Fixed Capital Cost:

The Fixed Capital Cost or Fixed Capital Investment (FCI) is the capital cost that occurs at the start of the project regardless of the size or architecture of the power system [51]. The fixed capital cost, for the liquefaction unit, is given by the following relation [52]:

FCI=5-6.5×Cp_{total}, where Cp_{total} is the total installed equipment cost.

The Fixed Capital Cost is estimated to be: **912.970×10^6 USD**.

TABLE 17: Final Cost Analysis – Cumulative Cost Results.

Raw Materials Cost- C_{RM} ($\times 10^6$ USD/y)	124.226
Utilities Cost-C_{UT} ($\times 10^6$ USD/y)	370.033
Operating Labor Cost-C_{OL} ($\times 10^6$ USD/y)	2.210
Total Production Cost-TPC ($\times 10^6$ USD/y)	**496.469**

Estimated cost of the LNG plant:

The cost breakdown by LNG plant unit is given in the following chart:

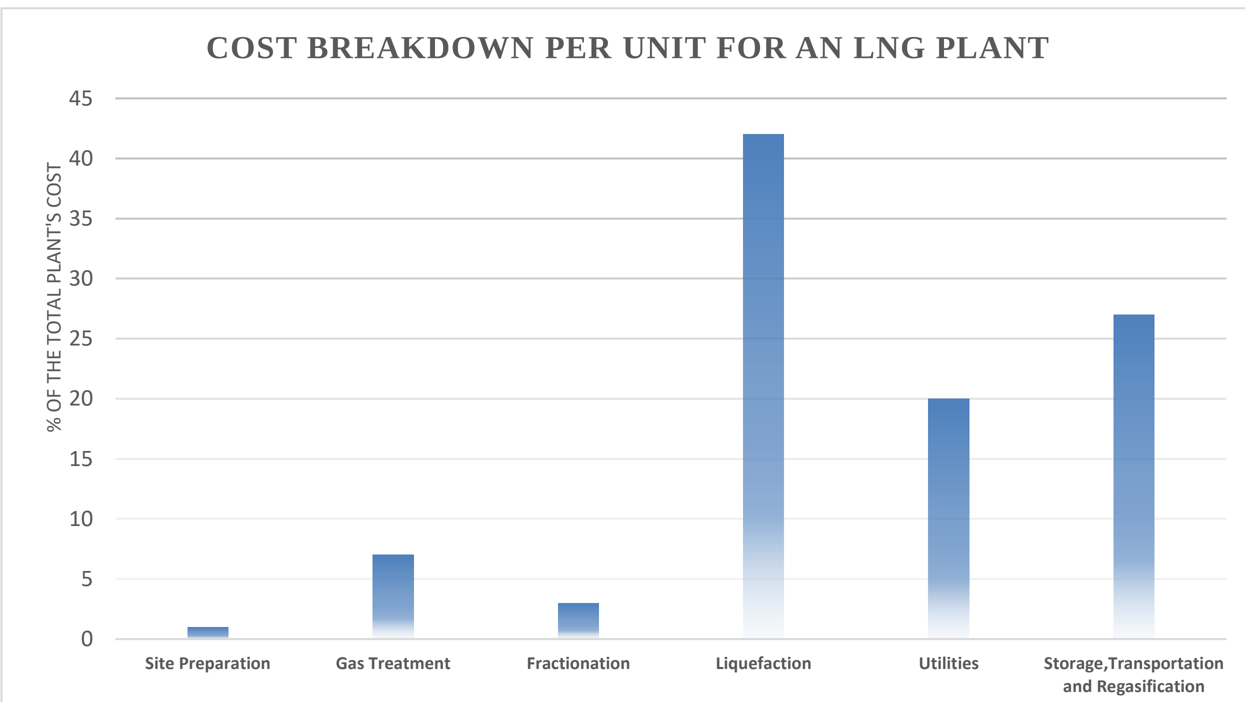

FIGURE 30: Cost Breakdown of an LNG plant per unit operation [53].

According to the data presented in Figure 30, the liquefaction process accounts for almost half of the LNG plant's total cost. If the Liquefaction unit accounts for the 42% of the total plant's cost, then a rough estimation of the whole LNG plant's cost should be: **2.605 $\times 10^9$ USD**.

Business plan for the Liquefaction Unit:

A financial analysis is essential when creating a new plant, in order to estimate and conclude if the construction and operation of the plant will be profitable. A profitable project will be the one presenting a Net Present Value (NPV) greater than 0. The Net present value corresponds to the difference between the present value of cash inflows and the present value of cash outflows over a period of time.

The case where NPV is negative then the operation of the plant will not be profitable and the project should not continue. The case where NPV=0, means that the operation will neither be harmful nor profitable and more data need to be examined in order for the project to continue. A basic financial analysis for the Liquefaction Unit of an LNG plant is presented below.

TABLE 18: Business plan of the Liquefaction unit operation.

Equity Financing (50% of the FCI)	456.485×10^6 USD
Loan Financing (50% of the FCI)	456.485×10^6 USD
LNG production	3.102 MTPA
Time of Construction	2-3 years
Life of asset	30 years
Fixed Capital	912.970×10^6 USD
Working Capital (15% of the FCI)	136.945×10^6 USD
Depreciation (20% of the FCI)	182.594×10^6 USD
Capital Cost or MARR	15%
Discount Rate	10%
Income Tax	29% [54]

- 50% of the FCI will be covered by the owner's capitals and the rest 50% will be covered by a loan which will be over a 30-year period, with an annual interest rate of 10% and equal annual payments.
- The working capital has been estimated to be 15% of the FCI, **136.945×10^6 USD**.
- The total expenditures for the utilities, the raw materials and labor have been estimated at **180.994 ×10^6 USD**.
- The depreciation of the equipment will occur in 30 years (when the loan expires) and the annual depreciation will be stable and equal to 20% of the FCI, **182.594×10^6 USD**.
- The tax rate of income will be equal to 29%.
- The discount rate will be 10%.
- The Minimum Acceptable Rate of Return (MARR) the company uses is stable and equal to 15%.

This analysis is based on a few assumptions:

- LNG prices will remain relatively stable in 30 years period.
- The loan interest will be annual instead of monthly after a special agreement with the lender.
- The products that exit the NGL fractionation unit are recycled to form the Mixed Refrigerant for the Liquefaction Unit.

By adding all these information into an Excel spreadsheet, the Net Present Value (**NPV**) is estimated to be **20.368×10^9 USD** and the Payback Period (**PBP**) equal to **0.880 years** (~9 months). The NPV is extremely high and the PBP is less than a year, meaning that is an approved project since the Liquefaction Unit will be profitable almost immediately after its operation. It is worth mentioning that these values correspond only for the liquefaction unit and not for the whole LNG plant.

E. Conclusions and Recommendations for Future Work

Conclusions that occur from the present work are:

- The cost of compressors as it was expected, was the highest cost when constructing the liquefaction unit and also had the highest cost of operation since their workload corresponded to the highest power consumption units. The total cost of compressors accounts for almost 81% of the total equipment cost, with 58% attributed to unit K-100, which required the highest power for operation, due to the recompression of the MR stream exiting the LNG exchanger and recycled in the process.

- The refrigeration cycle was based not in one, but two WCHEs that consisted the MCHE in order to achieve enough heat transfer between the streams and finally bring natural gas to -162 ºC at atmospheric pressure (LNG).

- The heat exchangers used for cooling the MR, had an extremely high surface area to achieve the required heat removal from the stream, before that entered the MCHE unit. Units E-103 to E-105 that were used for this process, presented two and three times larger heat transfer area compared to the WCHEs.

- The liquefaction unit turned out to be very profitable, since the PBP was estimated to be almost nine months of operation, while the NPV was estimated at 20.368×10^9 USD. These numbers are only indicative and would definitely alter if a technoeconomic analysis was contacted for the whole liquefaction plant instead of just the liquefaction unit.

Recommendations for future work include:

- The first recommendation involves energy integration in order to take advantage of the hot and cold duty of streams exiting the heat exchangers and as a result, minimize the cost of the utilities.

- Another suggestion for future work would be to test different MR compositions and compare the results to find the one who has the closest cooling curves to the ones of natural gas, thus eliminating the heat losses and helping the refrigeration process.

- A way to minimize compressors' work would also be considered for a future work, since compressors have the highest cost of operation in a liquefaction unit.

- Last but not least, a simulation and an economic evaluation of the same process is suggested for an offshore application this time, in an FLNG facility.

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Appendix

Costing of Vertical Pressure Vessels-Separators

By knowing the height and diameter of the vessels, the cost of equipment and the cost of the installed equipment is calculated based on the following equations.

$$\mathbf{Cp_o}\left(\$@1968\right) = 935,6 \cdot \mathrm{H}^{0,81} \cdot D^{1,05}$$

where Cp_o is the cost of equipment on USD of money value from 1968, H is the height and D is the diameter of the vessel.

$$\mathbf{Cp_o}\left(\$@2017\right) = \frac{CEPCI_{2017}}{CEPCI_{1979}} \mathbf{Cp_o}\left(\$@1968\right)$$

where Cp_o is the equipment cost corrected to an up to date cost of equipment index ($CEPCI_{2017}$) and $CEPCI_{2017}$ = 562.1 and $CEPCI_{1979}$ = 238.7 [55].

$$\mathbf{C}_{BM}\left(\$@2017\right) = [(\mathrm{F}_{BM}\text{-}1)+(\mathrm{F}_m\mathrm{F}_p)] \cdot \mathbf{Cp_o}\left(\$@2017\right)$$

where C_{BM} is the cost of the installed equipment, F_{BM}= 4.23 is the cost correction factor for the installed equipment, Fm=2.25 is the cost correction factor for the material and Fp=1.05 is the cost correction factor for the pressure [56].

For the shell:

$$\mathbf{Cp_o}\left(\$@1979\right) = \exp(8.6 - 0.21651\ln(W) + 0.04576\ln(W)^2)$$

where W is the weight of the vessel and is given by the following correlation:

$W = 155.6D(H + 0.812D)$ = 625,9788 kg.

For the platforms and stairs:

$$\mathbf{Cp_o}\left(\$@1979\right) = 1017 D^{0.7396} H^{0.70684}$$

Costing of the remaining units

The costing of the remaining units was estimated by using the online cost calculator of Mc Graw Hill Education [45,57]. The Utilities Cost was estimated by HYSYS software. The Wound Coil Heat Exchangers cost was estimated by the online cost calculator of Matches [47].

Raw Material Cost Estimation

Ethane and propane are produced from the NGL fractionation unit and methane can be obtained from the sweetened natural gas feed in order to compose the Mixed Refrigerant. The only component that needs to be imported is Nitrogen.

For the pure propane cycle, the propane that is produced from the NGL fractionation unit is not sufficient, hence most of it needs to be imported.

Cost of Nitrogen for the MR [58]:

Nitrogen Flowrate: 0.652 MTPA = 27,074,499 US gal/y = 3,619,500 cf/y.

12,065 cylinders are needed per year, each of 300 cf costing 47 USD.

Final Nitrogen Cost: **567,055 USD/y**.

Cost of propane as a coolant [59]:

Propane Flowrate: 3.102 MTPA=128,811,497 US gal/y.

0.96 USD/gal as of October 2018.

Final Propane Cost: **123,659,037 USD/y**.

Labor Cost Estimation

To make some calculations for the salary cost, it is estimated that 3.5 workers are needed per shift, hence a total of 11 workers per day (3 shifts/day) [60,61]. Table 19 shows how many employees are needed per unit of operation. With the assumption that the same workers work for 40 hours a week and with Ulrich's assumption that 4.5 employees are needed per position to cover positions on holidays, days-off and maintenance, the total labor cost is estimated at 1.93×10^6 USD/y for a minimum annual gross salary being 39,000 USD as of 2015 [62]. By further assuming working on the liquefaction unit, three engineers (for technical safety), a supervisor and a production manager the labor cost rises at a total of **2.21×10^6 USD/y**.

TABLE 19: Estimation of the number of employees needed for the liquefaction unit [60].

Equipment	**Employees/shift**	**Total Number of employees/shift**
Mixers (5)	0.2	1
Compressors (4)	0.1	0.4
Heat Exchangers (12)	0.05	0.6
Separators (5)	0.1-0.3	1.5
Valves (9)	0	0
		3.5

Net Present Value Estimation

For the calculation of the **NPV** it is necessary to know the net annual cash flow **(CF_k)**. The annual cash flow is the sum of: the net profits (**NI**), the depreciation (**D**), the loaned capitals (**LC**), the payments (**PP**), the invested capitals (**IC**) and the working capitals (**WC**).

Each category is estimated as follows:

Fixed Capital Cost (FCI) has been calculated at 912,970,000 USD.

Taxable Income (TI): TI=(R-E) **-** (D+I), where R is the revenues, E are the expenses, D is the depreciation which is estimated as the 20 % of the FCI and I is the annual interest rate.

Taxes (T): T=29%TI, where the taxes equal the 29% of the Taxable Income.

Net income (NI): NI=TI-T.

Invested Capital (IC): The negative value of the FCI.

Working Capital (WC): equal tov15% of the FCI.

Payments (PP): It is estimated by the PMT function of Excel for interest rate 10%, payback period 30 years and initial investment equal to the 50% of the FCI.

Loaned Capital (LC): 50% of the FCI.

Net Present Value (NPV) Calculation:

$NPV = \sum_{k=0}^{n} \frac{CF_k}{(1+MARR)^k}$, where CF_k is the annual cash flow and MARR (Minimum Acceptable Rate of Return) which could be achieved by the existing investments. A 15% MARR was chosen since the LNG plant is new [63].