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Near Shore FLNG Concept Evaluations

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MASTER THESIS

for

Students

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Near Shore FLNG Concept Evaluations

Konseptvurderinger for FLNG-anlegg plassert ved land

Background and objective

Near-shore or "at-shore" installation of a jetty-moored floating LNG (FLNG) unit may give advantages such as reduced cost, shorter development time, reduced land use and reduced project risk compared to traditional onshore LNG plant construction. Compared to offshore FLNG the floater can be simplified and more standardized, thus saving cost.

In the fall 2014 specialization project, various cases and possible configurations for near shore FLNG were analysed, using process/equipment models and reference data. Three different combinations of climatic areas and feed gas composition were addressed, with recommendations for system design adapted to the individual cases. Even though the process and system design was varied, a potential for standardization of the floater and topside design was indicated. Several more detailed aspects of process and system design were not addressed, however, including influence of driver capacity and power generation schemes, liquefaction process selection, NGL extraction and fractionation system selection and configuration, heat recovery and heat generation concepts, layout and standardization of floater, and onshore cooling system details. A combination of warm climate and rich feed gas with high CO_2 content may also be addressed, to provide a more complete picture for the case studies.

The objective of this work will therefore be to further investigate possible conceptual solutions for near-shore FLNG systems, to find favourable design concepts for the identified locations while still keeping focus on standardization. The specialization project focused on comparing data and results for different cases/locations, while the present work should focus on the influence of design choices for a limited number of selected locations.

The following tasks are to be considered:

1. Review of earlier work and any updated input and literature of relevance.
2. Establish updated design basis and location definitions for further modelling and analysis, focusing on climatic data, local requirements and restrictions, and expected feed gas data.
3. Review of facilities and process "building blocks", and identification of main conceptual configurations for a complete installation, including floater and onshore plant.
4. Analysis of potential system configurations for each of the given locations/cases, to clarify pros and cons, and to identify and quantify consequences of different concept selections.
5. Summary of results and findings, with recommendations for system selection and further work needed.

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Within 14 days of receiving the written text on the master thesis, the candidate shall submit a research plan for his project to the department.

When the thesis is evaluated, emphasis is put on processing of the results, and that they are presented in tabular and/or graphic form in a clear manner, and that they are analyzed carefully.

The thesis should be formulated as a research report with summary both in English and Norwegian, conclusion, literature references, table of contents etc. During the preparation of the text, the candidate should make an effort to produce a well-structured and easily readable report. In order to ease the evaluation of the thesis, it is important that the cross-references are correct. In the making of the report, strong emphasis should be placed on both a thorough discussion of the results and an orderly presentation.

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Pursuant to "Regulations concerning the supplementary provisions to the technology study program/Master of Science" at NTNU §20, the Department reserves the permission to utilize all the results and data for teaching and research purposes as well as in future publications.

The final report is to be submitted digitally in DAIM. An executive summary of the thesis including title, student's name, supervisor's name, year, department name, and NTNU's logo and name, shall be submitted to the department as a separate pdf file. Based on an agreement with the supervisor, the final report and other material and documents may be given to the supervisor in digital format.

☐ Work to be done in lab (Water power lab, Fluids engineering lab, Thermal engineering lab)
☐ Field work

Department of Energy and Process Engineering, 14. January 2015

Olav Bolland
Department Head

Jostein Pettersen
Academic Supervisor

Preface

This report is written as a master thesis at the Department of Energy and Process Engineering, Norwegian University of Science and Technology. The master thesis comprises 30 out of 30 credits in the 10^{th} semester and is a compulsory part of the MSc in Mechanical Engineering.

We would like to thank our supervisor Jostein Pettersen for dedicated support and excellent guidance throughout the master thesis. We especially appreciate that he has taken his time for biweekly meetings.

We would also like to thank Eirik Rødstøl, which has provided guidance for the optimization of the different process models in HYSYS.

Finally, we would like to thank Lars Petter Revheim at Höegh LNG for inputs on several challenging issues that could not be quantified by simulation in HYSYS.

Trondheim, 10^{th} June, 2015

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Sammendrag

Denne rapporten vurderer ulike konseptløsninger for LNG anlegg plassert delvis på land og delvis på en flyter – såkalt at-shore FLNG. Det er mange fordeler med en slik løsning, der redusert kostnad, kortere oppstartstid og muligheter for standardisering trekkes frem som de viktigste.

For å belyse utfordringene rundt FLNG-anlegg plassert ved land har to hovedscenario blitt identifisert, hvor begge knyttet opp mot en lokasjon som er aktuell for LNG produksjon. Lokasjonene som er valgt er Mexicogolfen og Nord-Norge. For disse er det etablert en basiskonfigurasjon for flytendegjøringsprosess, kompressordriver, NGL-ekstraksjon, varmegenerering og kjøling basert på fødegasskomposisjon, klima, statlige restriksjoner og lokale forhold. Resultatene fra basiskonfigurasjonen er kalkulert ved hjelp av HYSYS og referansedata. Videre er ett system byttet ut om gangen og målt opp mot resultatet for basiskonfigurasjonen. Dette identifiserer og kvantifiserer konsekvensene for hver alternative konfigurasjon. Den ønskede produksjonsraten er i utgangspunktet ca. 4 MTPA, men varierer sterkt for de ulike konfigurasjonene. I tillegg er alle simuleringene utført med to temperaturer, slik at konsekvensene for effektivitet og produksjonskapasitet ved gjennomsnittlig og høy omgivelsestemperatur blir belyst.

Deretter er det etablert tre underscenario for å undersøke resultatet av å kombinere flere alternativer på en gang. Også disse er knyttet opp mot en aktuell lokasjon for LNG produksjon. Systemkonfigurasjonen som er valgt for underscenariene er ansett som mest sannsynlig for den enkelte lokasjonen. De tre lokasjonene som er valgt er vestkysten av Canada, nordvestkysten av Russland og nordvestkysten av Australia. I likhet med hovedscenariene er også underscenariene simulert med gjennomsnittlig og høy omgivelsestemperatur.

Komplette prosessmodeller av de ulike scenariene og underscenariene er laget i HYSYS, og resultatet fra simuleringene danner hovedgrunnlaget for sammenlikningen av de ulike konfigurasjonsalternativene. Ved bruk av optimaliseringsfunksjonen Hyprotech SQP i HYSYS er flytendegjøringsprosessene PRICO og Niche optimalisert. Målet med optimaliseringen er å oppnå et best mulig sammenligningsgrunnlag for de ulike prosesskonfigurasjonene og hjelpesystemene. Optimaliseringen er basert på å gjøre flytegjøringsprosessene mest mulig effektiv på bakgrunn av tilgjengelig kompressorkraft i de ulike scenariene, altså å oppnå et lavest mulig spesifikt kraftbehov. Basert på resultatene for kraft- og varmebehov er drivstofforbruk og CO_2-utslipp regnet ut for de ulike scenariene.

Resultatene fra simuleringene og optimaliseringen av flytendegjøringsprosessene viser at PRICO har høyere produksjonsrate og lavere spesifikt kraftbehov enn Niche. Videre viser resultatene tydelig overlegenheten til sjøvannskjøling kombinert med elektrisk driver i forhold til gassturbiner og luftkjøling, målt ut fra spesifikt kraftbehov, produksjonsstabilitet og CO_2 utslipp. Kombinasjonen av varmt klima, luftkjøling og kompressorer som drives direkte av gassturbiner, viser seg å være den minst effektive konfigurasjonen. Resultatene viser tydelig at spesifikt kraftbehov øker i takt med økende kjølevannstemperatur, og dermed synker produksjonsraten. Ved bruk av gassturbiner

vil tilgjengelig kraft fra disse synke ved økende lufttemperatur, noe som reduserer produksjonen ytterligere.

I tillegg til simuleringer, har forhold som ikke kan kvantifiseres i HYSYS blitt undersøkt. Dette gjelder hovedsakelig kompleksiteten av anlegget og påliteligheten til de ulike løsningene. Det er blitt lagt mest vekt på påliteligheten til flytendegjøringstogene og NGL-ekstraksjon da disse er ansett som mest utslagsgivende for den totale påliteligheten. Som ventet viser beregningene at elektrisk driver gir flere produksjonsdager ved 100% kapasitet enn gassturbindriver. Derimot opererer alle togene uavhengig av hverandre, slik at dersom en av driverne svikter, kan full produksjon opprettholdes i de andre togene. Dette viser at den høyere påliteligheten til en elektrisk driver gir et mindre utslag på flere PRICO eller Niche tog enn den ville for et enkelt DMR tog, hvor driverne er i serie.

Videre har det vært stort fokus på standardisering av systemene som er plassert på flyteren, uavhengig av lokasjon og lokale forhold. Denne studien viser at flytendegjøringsprosessen, lagringstanker, lastesystem og fakkelsystem til en viss grad kan standardiseres og plasseres på flyteren. Derimot avhenger produksjonskapasitet og effektivitet av type kompressordriver og tilgjengelig kjølevannstemperatur. Der en elektrisk driver kan ha konstant ytelse uavhengig av lufttemperatur, faller ytelsen til en gassturbin drastisk når temperaturen stiger. Dette medfører store variasjoner i ytelsen til en gitt gassturbin ved de undersøkte lokasjonene, noe som igjen fører til store variasjoner i LNG produksjonen. Disse variasjonene indikerer også at det kan være vanskelig å operere anlegget og vanskelig å ta ut anleggets fulle potensial og optimale produksjon.

Som et resultat av dette bør deler av anlegget konfigureres etter at lokasjon og fødegasskomposisjon er kjent, slik at effektivitets- og produksjonspotensialet kan utnyttes fullt ut. Dette gjelder spesielt driver, kjølesystem og systemene for gassprosessering oppstrøms for flytendegjøringsprosessen. En generell løsning er foreslått, der dekket på flyteren kan ha en standardisert seksjon og en seksjon som kan tilpasses den aktuelle lokasjonen og gasskomposisjonen.

Videre viser studien at omplassering av flyteren er fullt mulig, men ikke gunstig dersom det er store forskjeller på de lokale forholdene eller fødegasskomposisjon. Dersom gassturbiner er brukt som kompressordrivere vil disse være enten underdimensjonert eller overdimensjonert, avhengig av om flyteren flyttes fra kaldt til varmt klima eller omvendt. Ut fra dette er elektrisk driver kombinert med sjøkjøling den eneste konfigurasjonen som egner seg for omplassering uten tap i enten driverutnyttelse eller produksjon. Denne konfigurasjonen er også sett på som det mest gunstige alternativet for standardisering før lokasjon og fødegasskomposisjon er kjent. Dette blir likevel det dyreste alternativet, spesielt hvis kraft ikke kan forsynes utenfra, selv om denne studien ikke har fokusert på kostnadene for de ulike alternativene.

Abstract

This study evaluates different solutions for a LNG facility, partially placed on shore and partially placed on a floater, hereby referred to as at-shore FLNG. There are several advantages with this solution where reduced cost, shorter development time and potential for standardization is highlighted as the greatest.

To illustrate the challenges for an at-shore FLNG project, two main scenarios linked to a potential location for LNG production have been identified. The chosen locations are the Gulf of Mexico and Northern Norway. An initial configuration for liquefaction, refrigerant compressor driver, NGL extraction, heat generation and cooling has been established based on weather data, governmental restrictions and local conditions at the locations. The result of this configuration has been calculated using HYSYS and reference data. Next, the process or utility systems have been swapped with other configuration alternatives. This is done one alternative at the time, and the result has been measured against the initial result to identify and quantify the consequence of other process or utility systems. The desired production rate is approximately 4 MTPA, but this varies at the different configuration alternatives. Additionally, all configurations are simulated with average and high temperature to identify and quantify the consequences this have for the plant efficiency and capacity.

Next, three subcases, each linked to other potential locations for LNG production, has been identified to evaluate the consequences of combining more than one alternative at the time. The alternative system combination is considered the most likely combination at the given location. The potential locations for the subcases are the west coast of Canada, the Northwest coast of Russia and the Northwest coast of Australia. As for the scenarios, the subcases are simulated with average and high temperature.

Complete process models of the different scenarios and subcases have been made in HYSYS and the simulation results forms the main basis for comparison for the different configuration alternatives. The PRICO and Niche liquefaction process have been optimized with the optimization function Hyprotech SQP in HYSYS. The reason for using the optimizer is to achieve a good basis of comparison between the different process configurations and utility systems. The optimizer is configured to obtain a liquefaction process as efficient as possible based on the available compressor power different scenarios and subcases, which in this case means a specific power as low as possible. Based on the simulation results for power demand and heating duties, fuel gas consumption and CO_2 emissions are calculated for the different scenarios.

The simulation and optimization results for the liquefaction processes show that PRICO has both higher production rate and lower specific power than Niche. Next, the simulation results clearly underline the superiority of seawater cooling combined with electrical drive compared to gas turbines and any air based cooling system in terms of specific power, production stability and CO_2-emmissions. The combination of a warm climate, air cooling and gas turbine driven compressors proves to be the least efficient combination. The results imply that specific power increase with increasing cooling water temperature, thus the production rate decrease. If gas turbine compressor drivers

are used, the power output drops with increasing air temperature, thus reducing the production even further.

In addition to the simulations, issues that cannot be quantified by HYSYS have been addressed. This mainly regards the complexity each system entails and the reliability of the alternatives. The reliability evaluation focuses mostly on the liquefaction trains and NGL extraction as these are regarded to have the greatest impact on the overall plant reliability. As expected, electrical drive results in more operating days with 100% production when compared to gas turbine drive. However, each train operates independently of each other, meaning that if one driver fails, the rest of the liquefaction trains can still maintain their production. This indicates that the increased reliability of electrical drive has a smaller impact on the multiple liquefaction trains in this study than it would in a single DMR train where the drivers are configured in series.

It has also been a great focus on the potential for standardization of systems placed on the FLSO, regardless of feed gas composition, local conditions and climate. The study shows that the liquefaction process, storage tanks, offloading and flare system can be standardized to a certain point and placed on the FLSO. However, the production capacity and efficiency of the facility largely depends on the type of driver and available cooling water temperature. Whereas electrical motor has a constant power output despite temperature, gas turbine performance drops rapidly when the ambient temperature increase. This results in large variation in the gas turbine output at the evaluated locations, which further results in large variation in LNG production. These variations also indicate that the plant will be more challenging in operation and may be hard to operate at the optimal specifications.

The results imply that to fully exploit the potential for high and efficient production, part of the process systems should be selected after the location and feed gas composition is known. This mainly regards driver, cooling and gas processing systems required upstream of the liquefaction process. A general solution is proposed, where the deck of the FLSO may have a standardized section and a field specific section that can be fitted to the given location and feed gas composition.

Next, the results show that relocation of the floater is possible but not favourable, especially if the variations in ambient air temperature and feed gas composition are large. If gas turbine compressor drivers are used, they will be either be undersized or oversized depending on whether the floater is moved from cold to hot climate or opposite. Based on this, electrical compressor drive combined with a seawater based cooling system is regarded to be the only favourable option for relocation without a major loss in production or efficiency. This configuration is also the only one regarded to be favourable for standardization before location and feed gas composition are known. However, this is also the most expensive configuration, especially if power must be generated locally, but a detailed cost analysis for the different alternatives has not been performed in this study.

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Nomenclature

BAT	Best Available Technology
BC	British Columbia
BOG	Boil-off Gas
BZ	Benzene
C3MR	Propane-precooled Mixed Refrigerant
CAPEX	Capital Expenditure
CC	Combined Cycle
CCS	Carbon Capture and Storage
CT	Cooling Tower
DMR	Dual Mixed Refrigerant
DWT	Deadweight Tonnage
FLNG	Floating Liquefied Natural Gas
FLSO	Floating Liquefaction, Storage and Offloading
GoM	Gulf of Mexico
GE	General Electric
GT	Gas Turbine
GTG	Gas Turbine Generator
HHC	Heavy Hydrocarbon
HHV	Higher Heating Value
HP	High Pressure
HX	Heat Exchanger
J-T	Joule-Thomson
LHV	Lower Heating Value
LNG	Liquefied Natural Gas
LP	Low Pressure
LPG	Liquefied Petroleum Gas
MEA	Monoethanolamine
MDEA	Methyldiethanolamine
MFC	Mixed Fluid Cascade
MP	Medium Pressure
MR	Mixed Refrigerant
MTPA	Million Tonne per Annum
NBP	Normal Boiling Point
NG	Natural Gas
NGL	Natural Gas Liquids
NPV	Net Present Value
QRA	Quantitative Risk Assessment
RH	Relative Humidity
RVP	Reid Vapour Pressure
SMR	Single Mixed Refrigerant
ST	Steam Turbine
SW	Sea Water
WB	Wet Bulb

1 Introduction

1.1 Background

The world's energy need grows rapidly. As the focus on the environmental changes grows as well, there is a need for a cleaner energy source. To be able to replace coal and oil, the use of natural gas is expected to increase significantly. Meeting this huge demand requires efficient production and transportation. Liquefying the gas into LNG is an efficient method for transportation, but requires an extensive and costly process. Traditionally, LNG plants have been located at the shore and LNG carriers have transported it to the market. In the recent years, the contractors have focused more on cost reducing alternatives to be able to compete globally.

One of the alternatives is to locate the liquefaction unit, storage tanks and all other gas processing facilities on a vessel. This is known as FLNG (Floating Liquefied Natural Gas) and can be located offshore, near shore or at shore. If the floater is located offshore, hereby referred to as offshore FLNG, all the process facilities needed from riser to storage tanks, including power generation are required on board. With a near shore or at-shore FLNG configuration, some parts of the gas processing facilities may be located onshore and some on the floater, hereby referred to as Floating Liquefaction Storage and Offloading (FLSO). This gives more opportunities for alternative utility systems and standardization of the floater. For the same reason, it can also open up for a smaller FLSO, but with higher capacity than offshore FLNG. In this study, only the at-shore configuration has been considered. At-shore FLNG describes the whole LNG plant including FLSO and the onshore facilities.

Traditionally, gas turbine direct drive has been the solution for compressor drivers at LNG plants. However, large variation in daily and seasonal temperatures results in frequent variations in production rates, which makes optimal production and plant efficiency difficult to achieve. Additionally, the use of gas turbines directly leads to increased CO_2 emissions from the LNG plant. This thesis considers the possibilities for electrical driven compressors, which is more environmental friendly if combined with renewable energy sources. Seawater cooling is also emphasized, since it is a more stable heat sink than ambient air, where the latter has been the traditional choice. A stable heat sink also improves the possibilities for an optimal production and easier operation.

Furthermore, the potential for relocation of the FLSO has been greatly emphasized in this study. Moving the unit to a new location often includes changes in climate conditions, gas compositions and governmental restrictions. This leads to different process and equipment requirements in order to achieve a profitable production. Thereby, standardization or partly standardization of the unit becomes a key factor for relocation the FLSO, and this is one of the main focus areas in this report.

1.2 Scope of Work

Point 4 in the assignment text can be divided into identifiable and quantifiable consequences. The quantifiable criteria chosen to consider in this paper are listed below.

- Efficiency
- Power Demand
- Production Rate
- Feed Gas Consumption
- Capacity
- Cooling Demand
- Heat Demand
- Emissions

However, other consequences such as complexity and reliability are hard to quantify and these will be evaluated based on literature, site-specific conditions and regulations and feed gas composition.

To limit the amount of cases, some systems have already been selected to proceed with based on findings in the specialization project (Corneliussen and Samnøy, 2014). Firstly, the cooling system will be an indirect system to obtain a standardized cooling circuit on the FLSO, meaning that a direct seawater system will not be evaluated. This implies that titanium heat exchangers are not required, thus saving cost. Additionally, fouling will not be a problem in the indirect freshwater system.

Secondly, the paper focuses on two out of the three relevant liquefaction technologies for a FLSO. The dual mixed refrigerant (DMR) technology has not been studied in this paper due to the complexity it entails although it is has a great potential for a single train configuration.

1.3 Outline of the Report

Chapter 2 contains background information and motivation for FLNG, ongoing offshore and at-shore FLNG projects, advantages and challenges with floating LNG production. Furthermore, relevant liquefaction technologies, driver solution and other utilities are discussed.

Chapter 3 presents the different cases evaluated in this report. Two main scenarios with potential locations are presented along with given climate data and system configurations. Next, three subcases are identified and linked to other potential locations to get a broader range of the analysis.

Chapter 4 describes the HYSYS models used in the simulations. Next, the HYSYS simulation results are presented and discussed in Chapter 5.

Chapter 6 contains a reliability analysis of the liquefaction modules and NGL extraction alternatives. Next, an analysis of the possible layouts for the vessel, including size and system location is presented. Finally, a system configuration for a lean and rich gas scenario is suggested.

Chapter 7 contains the conclusion, and recommendations for further work is given in Chapter 8.

2 Systems and Components

2.1 FLNG

Although LNG has been produced for more than 40 years, there are currently no floating LNG production facilities in the world. Compared to at-shore FLNG, offshore FLNG projects has to meet stricter requirements regarding robustness and reliability compared to the at-shore configuration in this study. For the same reason, the vessel must have all gas pre-treatment, power generation and utility systems on board. With a near shore or at-shore configuration, some parts of the gas processing facilities may be located onshore. This gives more opportunities for alternative utility systems and standardization of the FLSO. For the same reason, it can also open up for a smaller FLSO, but with potential for a higher production capacity.

This makes at-shore and offshore FLNG very different from each other. While offshore FLNG is a new concept located above the offshore reservoir with all needed facilities on board, at-shore FLNG is basically a new way to build a base load LNG plant, with production from pipeline gas or offshore gas that is brought to shore.

2.1.1 Challenges with FLNG

Although base load LNG plants and traditional FPSO's have been in production for many years, combining them into a FLSO turns out to be very challenging. The main challenges are listed below.

- Must be suited for a marine environment
- Process system must be light and compact
- Should utilize field proven technologies to ensure reliability and robustness
- Economically attractive
- Should utilize a minimum of flammable inventory
- Safety

For a base load LNG plant, thermodynamic efficiency and train capacity is of great importance during the process system selection. For a FLSO on the other hand, safety, weight, footprint, ease of operation and maintenance play a key role along with the two criteria for the base load plant.

Due to the limited space on a vessel deck, a low equipment count, low weight and small footprint of the process systems becomes important. The limited space also makes it harder to separate the systems having flammable inventory with the ones occupied by personnel. Therefore it is desirable to minimize the flammable inventory on the FLSO. For a base load plant, this is usually not that much of an issue since safety distances is not a problem.

2.1.2 Motivation

Pre-assembled modules for LNG base load plants has proven to be cost and time saving in projects such as the Woodside Train V expansion and Snøhvit. The cost of a modular design is typically 10-15% higher than on site due to the extra steel to make the module able to withstand shipping (Habibullah et al. 2009). However, this is overcome by reduced construction time due to parallel manufacturing of the modules and reduced onsite construction costs, especially for remote areas with poor infrastructure.

Figure 1 illustrates the metric cost (specific cost) of LNG plants in US$/tonne per annum which is calculated by the formula below.

$$Metric\ cost = \frac{Cost\ of\ the\ plant\ in\ millon\ US\$}{capacity\ in\ millon\ tonnes\ per\ annum}$$

Note that the recent high cost plants encircled in the figure are all located on remote locations in Australia and Papua New Guinea with the exception of Angola. Additionally, the cost for a construction worker in Australia is twice as much as for one in Singapore, US or Quatar (Songhurst, 2014). This is a result of high competition of in-country resources as the availability of personnel resources is limited. This illustrates the key role location and labour costs plays for a LNG plant. Note that the Lavaca Bay Barge, which is an at-shore FLNG project referred to several times later in this report, has a moderate metric cost of approximately $700/tpa, which further underpins the economic potential for an at-shore FLNG solution. The metric cost for the offshore Prelude FLNG project, also referred to later in this report, is estimated to approximately $3500/tpa (BBC, 2013). If the production of LPG and condensate is included in the metric cost calculation, the number drops to approximately $2380/tpa. In other words, the condensate and LPG production has a major impact on the economy for this project. However, the huge difference in metric cost for Lavaca Bay and Prelude reflects the difference in complexity and requirements for an offshore FLNG compared to at-shore FLNG.

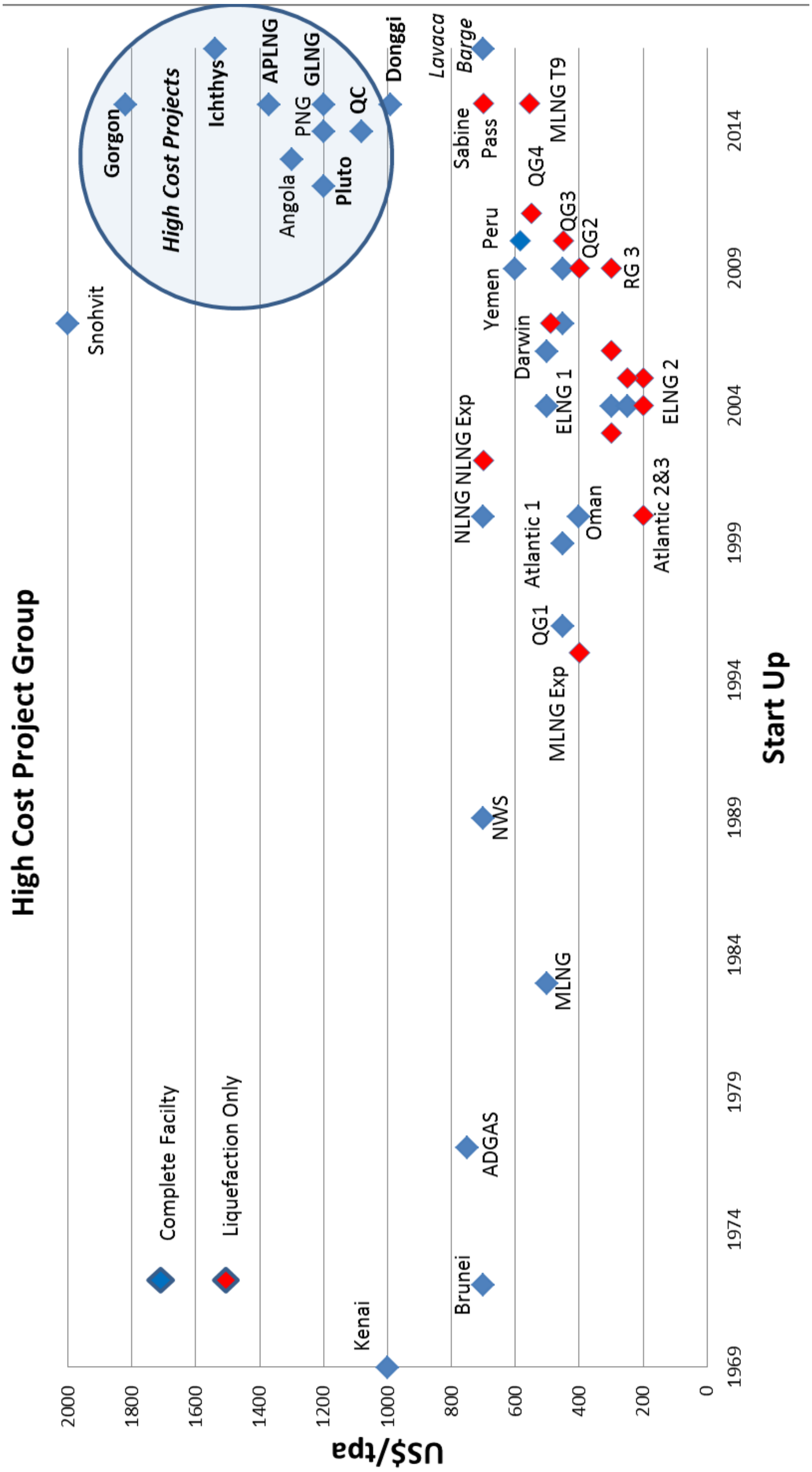

Figure 1: Metric cost for recent and ongoing LNG projects (Songhurst, 2015)

An at-shore FLNG takes the modularization a bit further. Each module can be built simultaneously at a factory where all the required construction equipment is present, thus saving CAPEX and construction time. By building the hull in a low cost country the capital cost may be significantly reduced. If the vessel is of moderate size, the number of shipyards able to construct it is quite many thus increasing the competition, which may reduce the cost. However, if the vessel is very large, more process equipment can be placed on deck but only a few shipyards can handle the size. The cost can become high since there is less competition. Unlike the modular projects mentioned, a vessel can be relocated to other potential fields. This opens up for a more economic feasible gas production from smaller gas field, which would normally not be profitable.
For this to be possible, the vessel must be standardized to be able to handle a wide range of gas compositions and climates. If so, the cost of building several vessels may be significantly reduced since the shipyard already have the building experience and less engineering hours are needed. The development of a typical base load LNG plant takes 10 years from concept to production including 4 years of site construction (Songhurst, 2014). The total schedule for a barge solution is expected to be significantly lower, especially if several standardized vessels are built.

The potential for standardization is underpinned by experiences from Höegh LNG, who has built two similar LNG carriers. Their estimate is that the cost savings for vessel number two lies in the range of 10-25% of CAPEX, mainly due to less engineering hours needed. The savings depends on how many changes in design that are made on vessel number two. From a life cycle perspective, it might be profitable to spend more to improve the design of the second vessel. Note that these numbers should be used as an indication, rather than exact numbers as a LNG carrier is quite standard, requiring significantly less engineering and construction hours compared to a complex FLSO.

Figure 2 shows the cost breakdown for a base load LNG plant producing from an offshore field. As shown in the figure, onshore cost represents roughly 60% of the total capital cost. For a lean gas scenario producing from already existing pipelines, the onshore site development cost will represent even more of the total capital cost. Next, site civil works such as dredging, jetty and harbour development is expected to be lower for an at-shore FLNG solution, since less process systems are required onshore. Other advantages with a barge solution are easier onshore de-commissioning when the project ends and less environmental footprint.

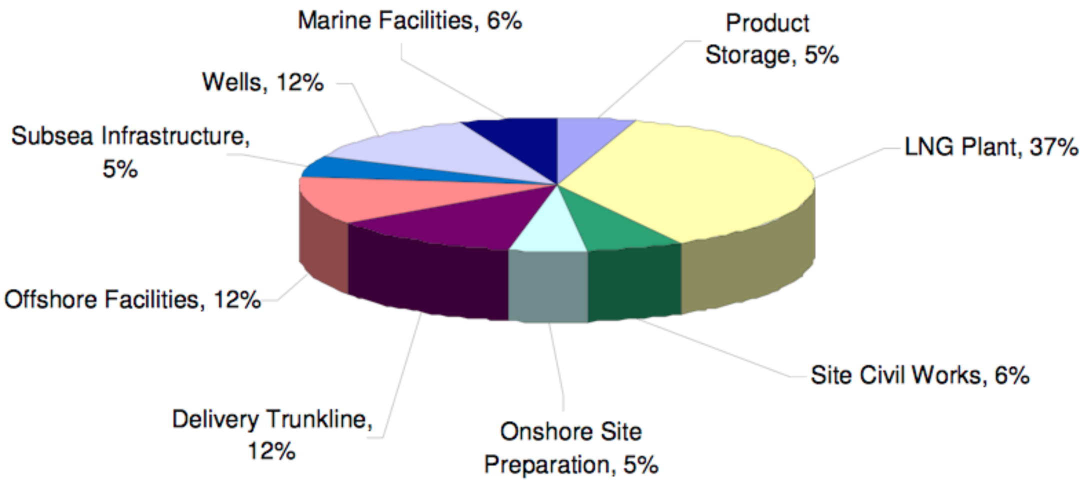

Figure 2: Typical cost breakdown for a LNG project (Habibullah et al. 2009)

Figure 3 shows the cost breakdown for the onshore development of a base load LNG plant. As shown here, the liquefaction, storage and refrigeration systems represent the largest part of the total cost. From a capital cost of view, these are most important to locate on the vessel. At a shipyard, all the necessary construction equipment is already there which makes the construction easier and faster. Transportation of the required topside process systems is also expected to be easier and faster as well. This means that the more remote the production location is, the more money can be saved by using an at-shore FLNG solution. This opens up for LNG production in countries with poor infrastructure such as the East Coast of Africa.

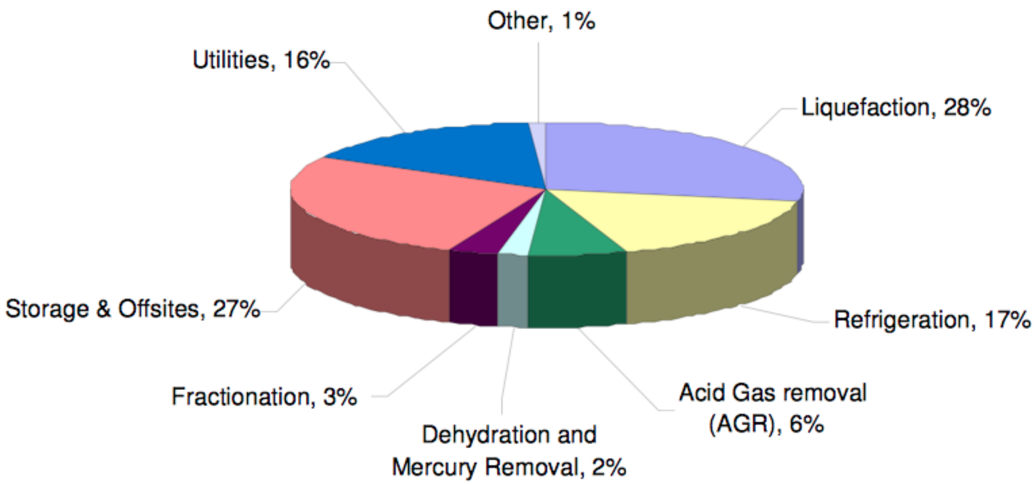

Figure 3: Typical breakdown for the onshore LNG plant development (Habibullah et al. 2009)

The advantages with at-shore FLNG discussed in this subchapter are summarized below.

- Lower CAPEX compared to an onshore development
- Shorter schedule
- Minimise the onshore requirements in terms of infrastructure and construction workers.
- Possible to standardize resulting in further reduction in CAPEX and lead time.
- Reusable for other gas fields.
- Less environmental footprint
- Easy de-commissioning

2.1.3 Similar Projects and Contractors

FLNG projects and contractors similar to the study in this thesis can be interesting to look into in terms of process solutions and concept configurations. Similar offshore and near shore projects are presented in this subchapter, with a short description and some key numbers. At the end, the projects and contactors are summarized in two tables.

Petronas Offshore FLNG

Petronas FLNG is an offshore FLNG project, developed for the Kanowit gas field 180 km off the coast of Sarawak in Malaysia. The FLNG unit is scheduled to be completed in the fourth quarter of 2015. The FEED and construction contractor for the project is Technip-Daewoo Consortium, which is a joint-venture between Technip and Daewoo Shipbuilding & Marine Engineering. The FLNG unit has an estimated production rate of 1.2 MTPA, and the liquefaction technology used is the dual N_2 expander process (Hashim et al. 2014) (Petronas 2014).

Shell Prelude Offshore FLNG

Another offshore FLNG project is the Shell Prelude FLNG, developed for the Prelude field located 475 km off the north western coast of Australia. FEED and construction contractors are Technip and Samsung Heavy industries. The FLNG vessel will be 488m long, 74m wide and weighing more than 600,000 tonnes fully ballasted, making it the largest floating vessel ever made. Shell's Dual Mixed Refrigerant (DMR) process with steam turbine compressor drivers will be used to liquefy the gas, and the estimated production rate is 3.6 MTPA LNG, 1.3 MTPA condensate and 0.4 MTPA LPG. Scheduled production start-up for the Prelude FLNG is around 2017 (Shell, 2014).

Pacific Rubiales Near Shore FLNG

The Pacific Rubiales FLNG facility will be located 3 km off the coast of Tolu, Columbia, and is defined as a near shore FLNG project. FEED and construction contractor for the project is Exmar NV. As shown in Figure 4, the floater will be moored at an offshore jetty, and gas will be supplied from an onshore field by a pipeline. Planned liquefaction technology is the PRICO SMR process, delivered by Black & Veatch, and estimated production rate of LNG is 0.5 MTPA. Latest news, however, is that Pacific Rubiales has decided to postpone the start-up of the FLNG facility due to unfavourable market conditions (Platts, 2015).

Figure 4: The Pacific Rubiales/Exmar FLNG project (Exmar, 2015)

Lavaca Bay At-shore FLNG

A more relevant project for this study is the Lavaca Bay FLNG by Excelerate Energy. This is an at-shore configuration consisting of two FLSO units that combined will produce up to 10 MTPA LNG from pipeline gas. Each FLSO uses four Black and Veatch PRICO trains to liquefy the gas. The site constructions was originally planned to begin in 2016, but due to the recent change in global market conditions, the project has been put on hold (LNG World News, 2014). Figure 5 shows an overview of the planned Lavaca Bay FLNG project (Excelerate Energy, 2013).

Figure 5: Plant overview of the Lavaca Bay project (Excelerate energy, 2013)

Flex LNG –Offshore FLNG

FLEX LNG is a FLNG contractor and has delivered FEED studies for offshore FLNG projects. FLEX LNG has suggested a modular topside system that is partly standardized and partly field specific. The partly standardized topside system includes acid gas removal, dehydration, mercury removal and dual N_2 expander liquefaction system with a LNG rundown capacity of 1.7-2.0 MTPA. Figure 6 shows a proposed topside layout of the offshore FLSO unit (FLEX LNG, 2015).

Figure 6: Principal sketch of the FLEX LNG FLSO unit (FLEX LNG, 2015)

Höegh LNG – Near Shore/At-shore FLNG

Höegh LNG is FLNG contractor that delivers Pre-FEED, full generic FEED and field specific studies for near shore/at shore FLNG projects. The FLNG solution has a LNG production capacity of 0.5-3.0 MTPA, depending of the liquefaction process and field size. The barge includes gas pre-treatment facilities as well as utility systems, and the liquefaction processes offered are SMR and DMR. Power generation will be on board or alternatively onshore, and LNG storage can be done in the hull of the FLSO or in an external FSO (Floating, Storage and Offloading) with buffer storage in the FLSO hull, as shown in Figure 7 (Höegh LNG, 2015).

Höegh LNG's near shore FLNG solution is based on liquefaction of pipeline gas quality for greenfield development. Since Höegh LNG's projects and FEED studies faces many of the same challenges as in this study, they have been an important source of information and experience for this thesis.

Figure 7: Principal sketch of the Höegh's near shore FLNG concept (Höegh LNG, 2015)

Summary

A summary of the similar offshore and near shore FLNG projects and contractors are presented in Table 1 and Table 2 on the next page.

Table 1: Summary of similar FLNG projects and contractors (Cott Oil and Gas, 2015) (LNG World News, 2012) (Robinson, 2012) (Air Products, 2014) (Ahmad, 2015)

Project/ contractor	Location	Rundown LNG Capacity [MTPA]	Feed Gas	Liquefaction Process	Number of Trains	LNG Storage Capacity [m^3]	Power Gen/ Driver	Project Status/ Production start
Petronas FLNG/ Technip-Daewoo Consortium	Offshore Malaysia	1.2	Lean	N_2 Expander	1	177,000	GT power gen/ direct drive	Under construction/ Q4 2015
Prelude FLNG/Shell	Offshore Australia	3.6	Rich	DMR	1	220,000	ST power gen/ direct drive	Under construction/ 2017

Table 2: Summary of similar near shore FLNG projects and contractors (Pacific Rubiales, 2010) (Pacific Rubiales, 2015) (LNG World News, 2014) (FLEX LNG, 2015) (Höegh LNG, 2015)

Project/ contractor	Location	Rundown LNG Capacity [MTPA]	Feed Gas	Liquefaction Process	Number of Trains	LNG Storage Capacity [m^3]	Power Gen/ Driver	Project Status/ Production start
Pacific Rubiales/ Exmar	Near shore Colombia	0.5	Lean	PRICO	1	16,500	GT power gen/el drive	On hold
Lavaca Bay/ Excelerate	At-shore Texas	8-10 (2 FLSO's)	Pipeline	PRICO	8	500,000 (2 units)	GT direct drive	On hold /2019
FLEX LNG	-	1.7-2.0	-	N_2 Expander	2	170,000	Field specific	-
Höegh LNG	-	3.0	-	SMR/DMR	1-2	Field specific	Field specific	-

2.1.4 Vessel Design

The size of the vessel depends largely on the required storage capacity. A standard LNG carrier has a capacity of 150 000 m^3. With a plant production rate of roughly 4 MTPA, this corresponds to roughly 5.5 days of full production. Some flexibility must also be added in case of bad weather, which may delay the loading or if the carrier is late. Based on this, a storage capacity of 250 000 m^3 is reasonable, which makes the facility able to maintain full production in case any of the situations above should occur.

Since this is an at-shore FLNG, which will be moored to a jetty, there is no need for a propulsion system, which would only result in unnecessary use of space, cost and maintenance. For the same reason, control room and living quarters can be located onshore so that the deck space can be fully exploited for process systems.

The hull efficiency also becomes less important since the vessel is expected to be towed minimally throughout the project life. This opens up for construction of a broader vessel than a regular shipping vessel, which can be crucial to exploit the deck space. However, for the shipyards to be able to construct the hull fast and affordable it will be very beneficial to have a hull similar to one they already have experience with.

To get an estimate of the ship dimension, light calculations have been performed on one of the cases studied in this report, presented in Chapter 6.4.

2.1.5 FLNG Safety Issues

Placing a LNG plant on a floating vessel entails a significant challenge from a safety point of view. The main focus is to minimize the risk for an incident and to prevent any escalation if this occurs. As there are no FLNG units in production it is difficult to identify all potential threats. The main issues identified are listed below.

- Fire and explosion
- Amount of flammable inventory
- Safety distances
- Cryogenic spill
- LNG offloading
- Personnel evacuation
- Blowdown and flare system
- High pressure systems
- Hot oil leakage

The greatest threat is possibly fire and explosion. Compared to a traditional onshore LNG plant, the process systems are much more compact. For instance, the Prelude FLNG project has a footprint of approximately one quarter of a typical onshore plant (Shell, 2014). One of the most common methods to avoid explosion and escalation is to have sufficient safety gaps between the process systems treating or containing hydrocarbons. This increases the ventilation and limits the size of the gas cloud if any hydrocarbon leakage occurs. A potential explosion flame front will also decelerate in the gaps (Haitsma, 2014).

However, large safety gaps imply that less deck space can be exploited and it reduces the cost saving potential. Another proposed solution to prevent escalation is to use blast walls between the modules. These will not require large gaps between modules, thus a large deck area can be fitted with process equipment. However, the walls prevent ventilation and if an explosion occurs, a high explosion pressure will be generated inside the wall. This requires all systems to be designed to handle the explosion pressure (Revheim, 2015).

A safe design and layout requires extensive analysis. A Quantitative Risk Assessment (QRA) should be performed and detailed simulations of potential explosions are necessary. This requires a detailed layout of the facility, including all piping and system configurations. As this is outside the scope of this study, it will not be studied any further. However, to account for the required plot area and to get a more accurate layout of the vessel deck, a safety gap of 15 meters has been assumed between the explosion zones.

2.2 Liquefaction Technologies for FLNG

There are several liquefaction technologies available to liquefy natural gas, and most of them are based on the mixed refrigerant cycle (MRC). Shell's Prelude FLNG project in Australia uses a single train DMR (dual mixed refrigerant) process with a capacity of 3.6 MTPA LNG. In addition to the DMR process, other mixed refrigerant processes such as the mixed fluid cascade (MFC) and propane-precooled mixed-refrigerant (C3MR) have certain advantages in terms of efficiency and capacity. This is illustrated in Figure 8 and Figure 9. However, due to drawbacks in term of complexity, space requirements and HSE issues, these processes have not been considered in this report.

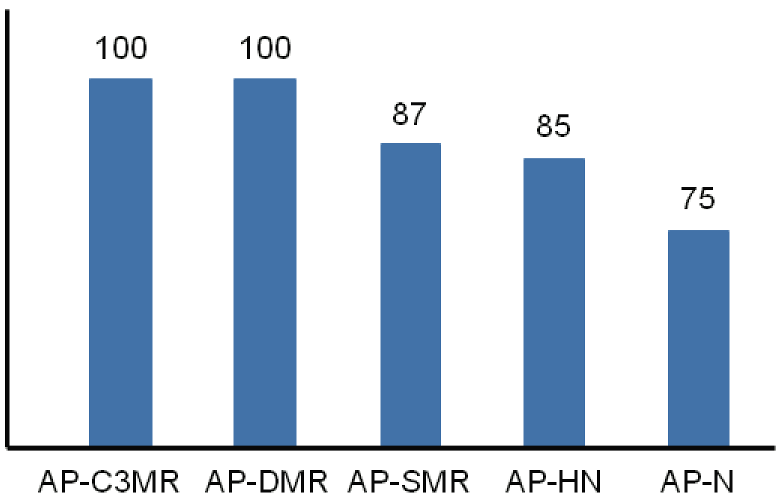

Figure 8: Relative process efficiency based on the C3MR process (Bukowsk & Boccella 2013)

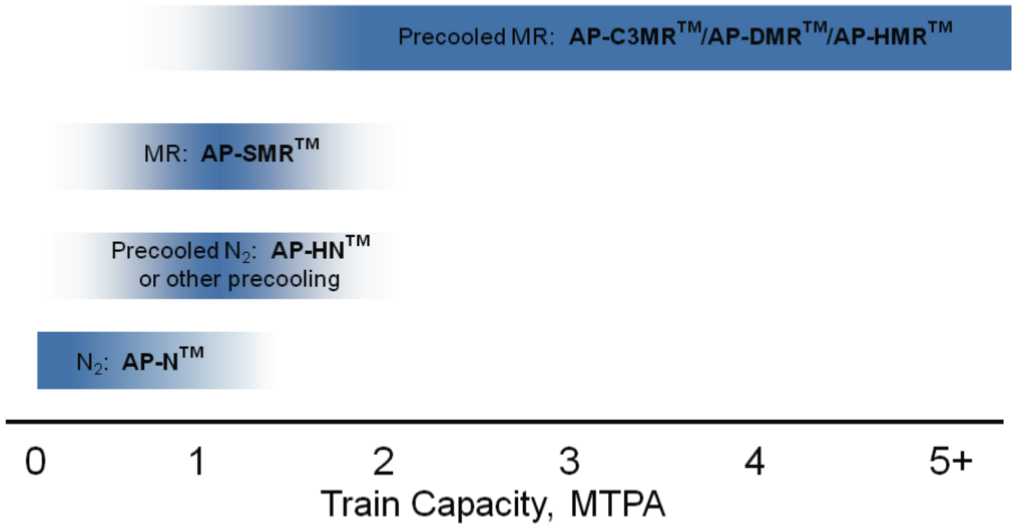

Figure 9: Single train capacity for a selection of liquefaction processes (Bukowsk & Boccella 2013)

2.2.1 The PRICO Process

The PRICO process is the simplest form of mixed refrigerant liquefaction cycle, as it consists of only one refrigeration circuit. The natural gas is precooled, liquefied and subcooled in the same heat exchanger, which also allows for integrated NGL-extraction, as shown in Figure 10. Due to low complexity and low amount of equipment required, the PRICO process is well suited for a FLNG.

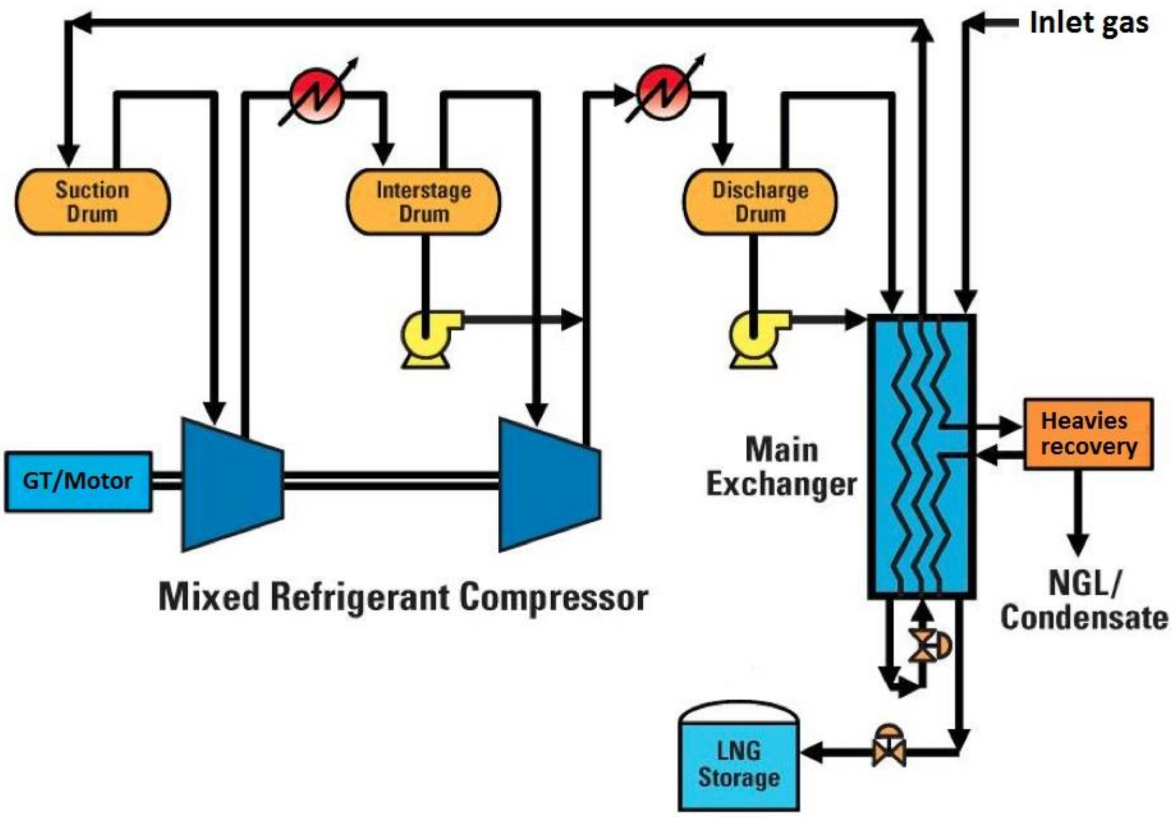

Figure 10: Modified principal sketch of the PRICO process (Talib et al. 2011)

2.2.2 The Niche Process

The Niche LNG process consists of two independent circuits that utilize both natural gas and nitrogen as refrigerants. A principal sketch of the Niche process is shown in Figure 11. Pretreated inlet natural gas is mixed with the natural gas refrigerant circuit, making it an open circuit, while the nitrogen circulates is a closed circuit. The mass flow of natural gas circuit regulates the cooling capacity and the production rate of LNG (Foglietta et al. 2013).

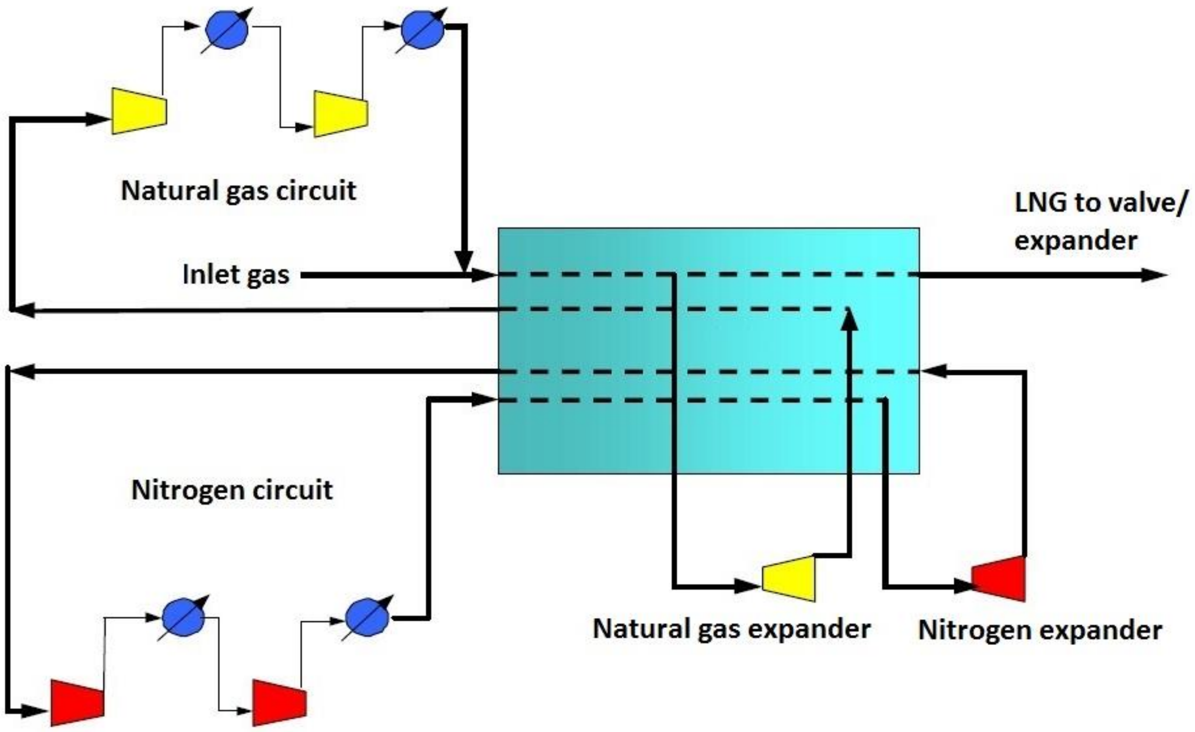

Figure 11: Modified principal sketch of the Niche process (Wijngaarden & Jos 2006)

The Niche process consists of a motor/gas turbine driven compressor and an expansion turbine driven compressor (hereby referred to as compander) on both refrigerant circuits. The efficiency of the liquefaction process refers to the specific power, meaning the LNG production rate divided by the added compressor work. Thereby, the compander work is not included when specific power is calculated. However, the specific power of Niche is somewhat higher than for PRICO. This is partly because the refrigerants are in gas phase in the low pressure side of the LNG heat exchanger.

The Niche process offers certain benefits in terms of operation. The process operates at high pressure, which makes the refrigerant gas highly compressed, resulting in smaller pipes and valves. The risk of a major leakage is reduced, and since the refrigerant is always in gaseous phase, no refrigerant storage, drums and separators are necessary. (Kuru & Iyagba, 2013).

2.3 Compressor Drive

The refrigerant compressors require a large amount of power and can either be driven directly by a gas turbine, an electric motor or a steam turbine. The choice of driver can affect the productions rate to a great extent under varying ambient conditions and is often the production bottleneck. While the power output for an electric motor remains constant despite changes in ambient temperature, gas turbine power output varies significantly. The choice of driver in this project depends largely on cost, complexity, reliability and production stability.

2.3.1 Gas Turbine Direct Drive

There are several gas turbines available as mechanical drive. For an onshore base load LNG plant, industrial gas turbines are the most common. For a FLSO, an aeroderivative gas turbine is regarded to be the best choice due to the compactness, low weight and high efficiency this type provides when compared to the industrial type. In this study, a configuration using the Rolls Royce Trent 60 DLE has been studied, but several other aeroderivative models, such as the GE LM6000, may be an alternative. The rated power for the Trent 60 is 53 MW at ISO conditions (15°C, 60% RH) and 42.4% efficiency (Centrax, 2015).

Note that the power output provided by the manufacturer is rated with no losses. To get a realistic number for the available power, derating factors must be included. The numbers used in this study are given in Table 3 and include pressure losses at the inlet and outlet. For gas turbines operating in cold climates, typically colder than 3°C, a de-icing unit must be installed, leading to a larger pressure drop and higher air inlet temperature, resulting in a lower power output (Pettersen, 2015).

Table 3: Derating factors for aeroderivative gas turbines (Pettersen, 2015)

Type of Derating	Derating Factor
Ageing	0.96
Fouling	0.98
API 617 margin	0.96
Engineering margin	0.96
De-icing	0.985
Total derating factor warm	**0.867**
Total derating factor cold (below 3°C)	**0.854**

The total derating factors results in a power output of 46 MW for Trent 60 at 15°C air temperature. For locations that experience temperatures below 3°C, the output drops to 45.3 MW at ISO conditions. The derating factors do not include the varying performance associated with changes in ambient temperatures. A typical estimate is that the output will fall 1.2%/°C (Schmidt et al. 2010).

The performance of the Trent 60 is given in Figure 12 and illustrates the declining power output when temperatures increase. As shown, the Trent 60 is quite sensitive to changes in temperature. Taking into account that temperature changes during the day can easily exceed 15°C indicates that operation with this gas turbine drivers may be challenging.

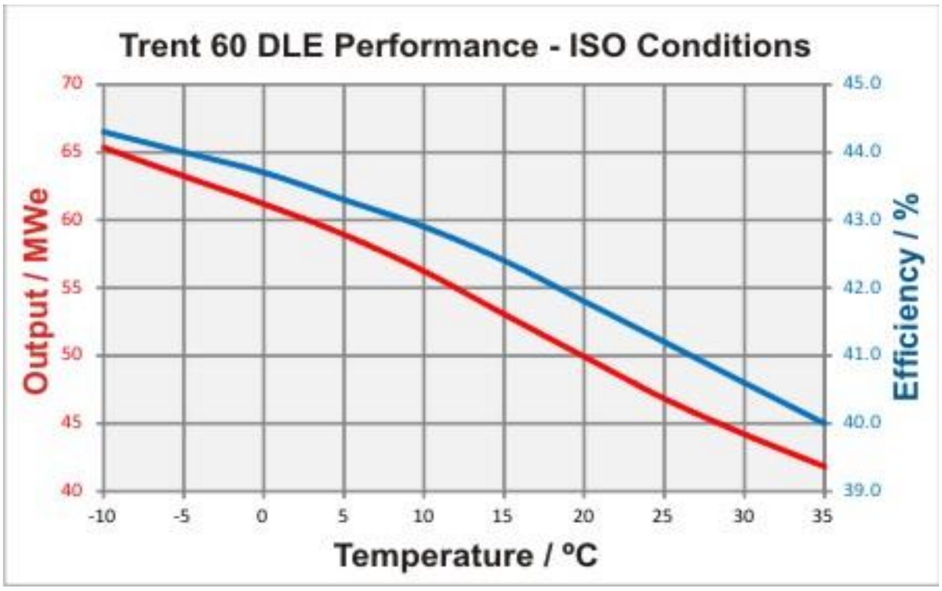

Figure 12: RR Trent 60 DLE performance with varying temperatures (Centrax Gas Turbines, 2015)

2.3.2 Electrical Drive

The second alternative is to use electric motors as compressor drive, supplied by power from the grid or from a local power plant. The Melkøya LNG plant, located in the northern part of Norway is currently the only plant in operation utilizing electric motors as compressor drive. Electric motors can be built in any size, where the 65 MW models at Melkøya are the largest to date. However, ongoing projects, such as the Freeport LNG in the US are planning to use six GE 75 MW electric motors to drive the refrigerant compressors for their two 4.4 MTPA trains. These motors will either use existing electric power from the grid or power generated locally in a combined cycle power plant, also delivered by GE (Business Wire, 2014).

The use of an electric motor offers several important advantages over a gas turbine. Unlike the gas turbine, a motor offers a constant power output despite high ambient temperature. Next, a motor requires less maintenance than a gas turbine, which results in an increased availability of approximately 2%. This translates into a significant amount of income (Schmidt et al. 2010).

On the downside, the use of electric motors will have higher CAPEX due to the cost associated with the power plant needed to generate electric power. However, the increased availability and reliability, shorter delivery schedule and stable power output can result in a net increase in NPV (Habibullah et al. 2009).

2.3.3 Steam Turbine

Although steam turbines as compressor drive and power generation was common earlier and is still used in large single train plants today such as the Prelude FLNG project, it has not been considered any further in this study. The reason for this is the high weight and large space requirements on the vessel deck. On the Prelude project, this disadvantage is overcome by the more robust and reliable system that steam turbine technology offers compared to gas turbines. Prelude utilizes a single train DMR process and the use of many aeroderivative gas turbines would reduce the plant availability, as the failure of one of the gas turbine might result in a shutdown of the whole plant. In this thesis, several independent and small capacity trains have been studied instead of a single large DMR train. If one of the gas turbines unexpectedly fails, the plant is still able to maintain 50% or more of the production, depending on the liquefaction process and number of trains. Therefore the advantage of several steam turbines does not outweigh the high weight and footprint.

2.4 Gas Processing Requirements and Product Specifications

The specific pre-liquefaction requirements are presented in Table 4 along with requirements for the LNG, LPG and condensate product.

Table 4: Requirements for pre-liquefaction, LNG, LPG and condensate (Pettersen, 2015)

Component	Pre-liquefaction requirement	LNG product	LPG product	Condensate product
C1		> 85 mol%	-	-
C2	-	-	< 1 mol%	
C4		< 2 mol%		
C5+	<0.1 mol%	<0.1 mol%	< 2 mol%	-
BZ	1 ppm	1 ppm	-	
CO_2	< 50 ppmv	< 50 ppmv		
H_2O	<0.1 ppmv	<0.1 ppmv	<0.1 ppmv	-
H_2S	<4 ppmv	<4 ppmv	-	-
N_2	-	<1.0 mol%	-	-
Maximum Gross Calorific Value	-	42 MJ/Sm^3	-	-
Reid Vapour Pressure at 37.8°C	-	-	-	11.5 psi

Figure 13 shows the solubility of different components in liquefied methane, based on the Non-Equilibrium Simulator developed at NTNU. Any concentrations to the left of the points in the diagram results in freeze out. An additional margin to these points must also be taken into account, resulting in very strict constraints for the concentration of some components as seen in Table 4.

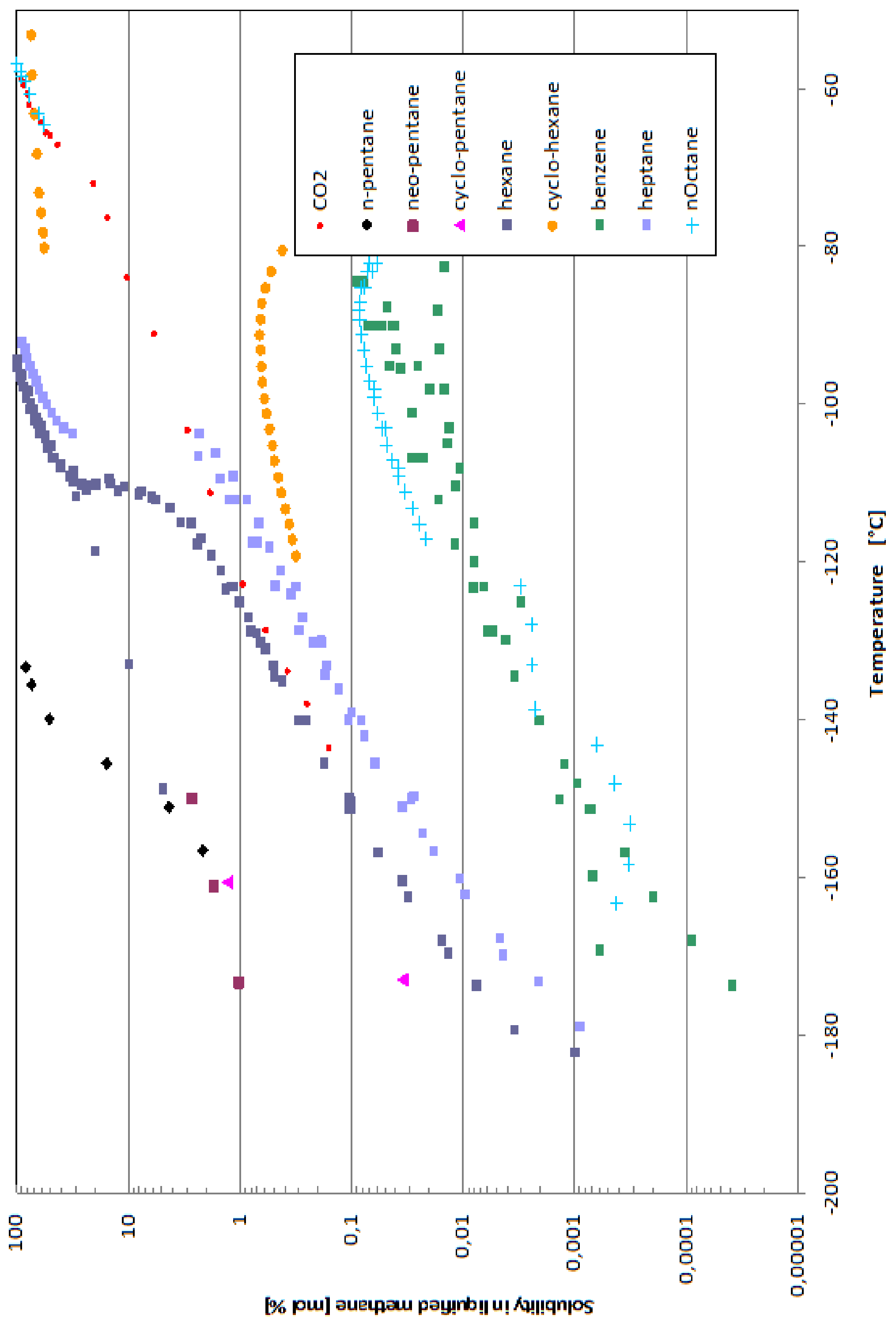

Figure 13: Solubility diagram (freeze out) for selected components in liquefied methane. (Pettersen, 2015)

2.5 NGL Extraction Options

The motivation for NGL extraction can be summarized in four points.

- To avoid freeze out of heavy hydrocarbons during liquefaction.
- To adjust the heating value to meet the market requirements.
- To produce a sellable LPG product.
- To supply makeup refrigerant to the liquefaction process.

As a result, most of the NGL needs to be extracted before or during the liquefaction process. This can either be done upstream or integrated after precooling in the liquefaction heat exchanger. However, removal of BZ and HHC is often challenging. The system must be able to extract enough, able to handle variations in feed gas composition and operate efficiently and reliably. The most common NGL extraction processes are presented and discussed in this subchapter.

If the NGL extraction is located upstream, the process is often referred to as frontend turboexpander NGL extraction and may become quite complex due to a number of rotating equipment (Chen and Ott, 2013). The use of a compander and booster compressor in series with the liquefaction process may reduce the availability of the plant and will in most cases lead to an increased cost. In lean gas cases with a low level of NGL extraction, the capital cost might not be justified. However, it offers an efficient removal of HHC, as the lower operating pressure makes separation easier.

The integrated NGL extraction does not require a compander or booster compressor like frontend does, making it easier to operate, cheaper and more reliable. However, to achieve the required level of extraction, the HHC column must be operated at a pressure with sufficient margin to the critical point (Chen and Ott, 2013). This will reduce the liquefaction efficiency and therefore increase the total power consumption. For a leaner feed gas, the pressure and temperature must be reduced even more. This will reduce the production capacity compared to a solution with a frontend system, as the latter can be operated at a pressure closer to the critical point. Additionally, if the gas is very lean there may not be a large enough amount of heavy components to achieve sufficient reflux to the HHC column.

Another option for integrated NGL extraction is partial condensation, which utilizes a separator instead of a scrub column. Black and Veatch have this option integrated in their PRICO modules. This is less expensive and simpler in operation but may not remove the required amount of C8, C9 and BZ, as removal of these components requires lower temperatures or a more extensive extraction process.

An option for the removal of specific components such as aromatics and heavy paraffins is frontend adsorption. This process does not operate at reduced pressure as it is not based on vapour-liquid equilibrium, and will therefore not affect the liquefaction efficiency. However, using adsorption alone is considered to be neither economical nor practical, as removal of a high amount of NGL will result in very large adsorption beds and a large amount of regeneration gas.

All of the options presented are challenging to operate with varying feed gas composition, which might be necessary, especially for a lean gas scenario fed by several pipelines from different reservoirs. Figure 14 shows the varying feed gas composition for a plant producing from pipeline gas. The methane content is nearly steady before it changes rapidly to another nearly steady value, varying between 92 mol% and 98 mol%. The change in the total amount of heavy hydrocarbons is also rapid, varying between 20 and 200 ppm (Chen and Ott, 2013).

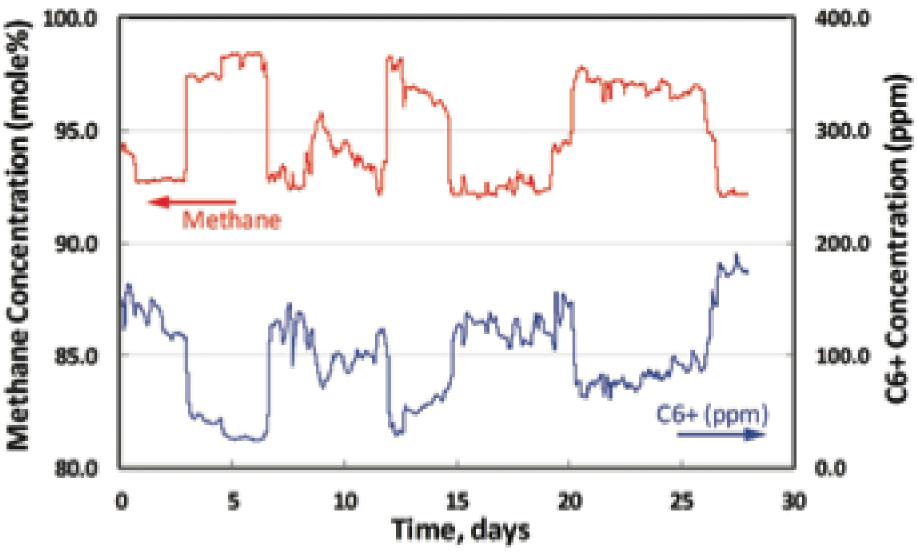

Figure 14: Change in feed gas composition for pipeline lean gas (Chen and Ott, 2013)

A process that handles this scenario is the APCI adsorption/partial condensation hybrid, shown in Figure 15. In this system, the adsorption process removes a portion of the HHC, and a portion is removed in the partial condensation drum. Leaving the feed gas that enters the liquefaction unit partly condensed, the less liquid is required in the drum and the process can thereby operate at higher pressure. According to Air Products, combining these two systems results in sufficient removal of HHC at varying gas compositions, keeps the equipment count low and improves the liquefaction efficiency.

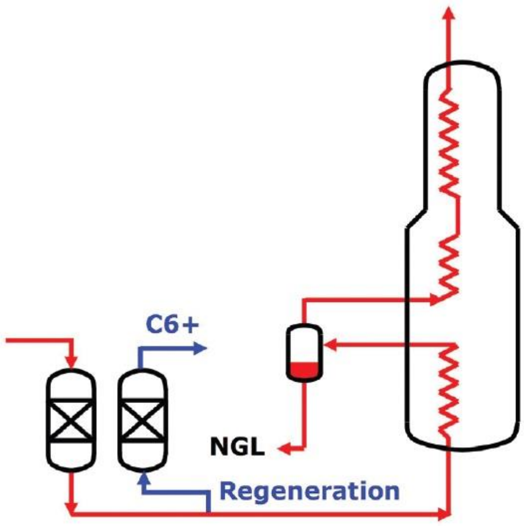

Figure 15: APCI adsorption/partial condensation hybrid (Chen and Ott, 2013)

The required NGL systems and the amount of LPG extracted from the gas will vary significantly with the feed gas composition. For a lean gas scenario, all the LPG can possibly be reinjected into the lean gas after the heavier components have been fractionated. For a richer gas, the amount of LPG extracted exceeds the maximum concentration of C4 in the LNG product and must be fractionated and stored separately. This variation in feed gas makes any standardization challenging.

2.6 Process Cooling

The process cooling system is an essential part of a LNG facility, as heat needs to be removed from different processes and components. Process cooling can be done either directly or indirectly, and the heat sink may be air, seawater or freshwater. Direct process cooling refers to configurations where the heat is rejected directly from the refrigerant to the heat sink. Although it often offers more effective heat rejection, it increases the need for a specific design for each location and thereby reduces the possibility for standardization of the FLSO.

An indirect cooling system consists of a recirculating circuit that removes heat from the processes and components and rejects it against a heat sink. The medium used for the indirect cooling circuit is often ionized water. Figure 16 illustrates the most common cooling alternatives for a near shore FLNG facility.

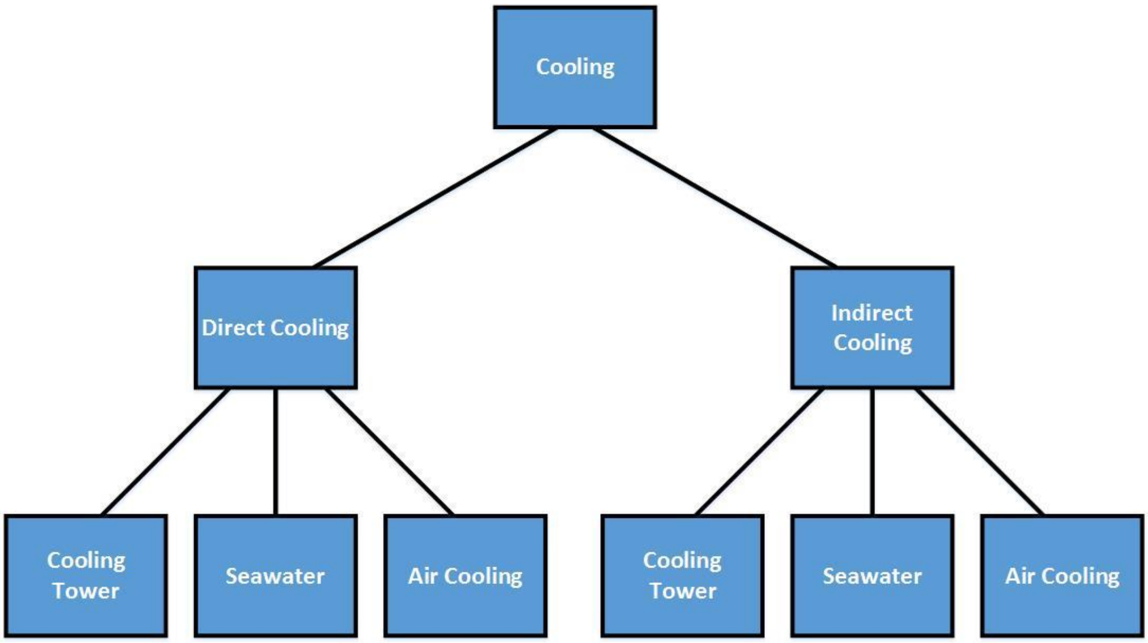

Figure 16: Process cooling alternatives for a near shore FLNG facility

As indicated in the specialization project, indirect cooling is the most suitable alternative considering standardization opportunities for the FLSO unit (Corneliussen and Samnøy, 2014). Possible heat sinks will be the same as described in the figure, meaning cooling towers, seawater and air coolers, and these will be further discussed in the subchapters below.

2.6.1 Cooling Towers

Cooling towers provide an alternative for discharging energy to the surroundings where sufficient cooling water cannot be obtained from natural sources, or at places where the temperature at which cooling water can be returned to the environment is regulated by concerns for the environment. However, if the ambient temperature is below the freezing point of water, cooling towers cannot be used. Cooling towers utilize the principal of evaporation, as a small portion of the incoming water evaporates and attracts heat from the rest. As a result, a large amount of makeup water is required. The cold water exiting the cooling tower can approach temperatures 2-4 °C higher than the wet-bulb air temperature, depending on the design of the cooling tower (Kroger, 2004).

For this study, a forced convection direct/open cooling tower will be considered. An extra heat exchanger between the cooling tower circuit and the indirect cooling circuit is suggested. A possible configuration of a cooling tower as heat sink for an indirect cooling circuit is shown in Figure 17.

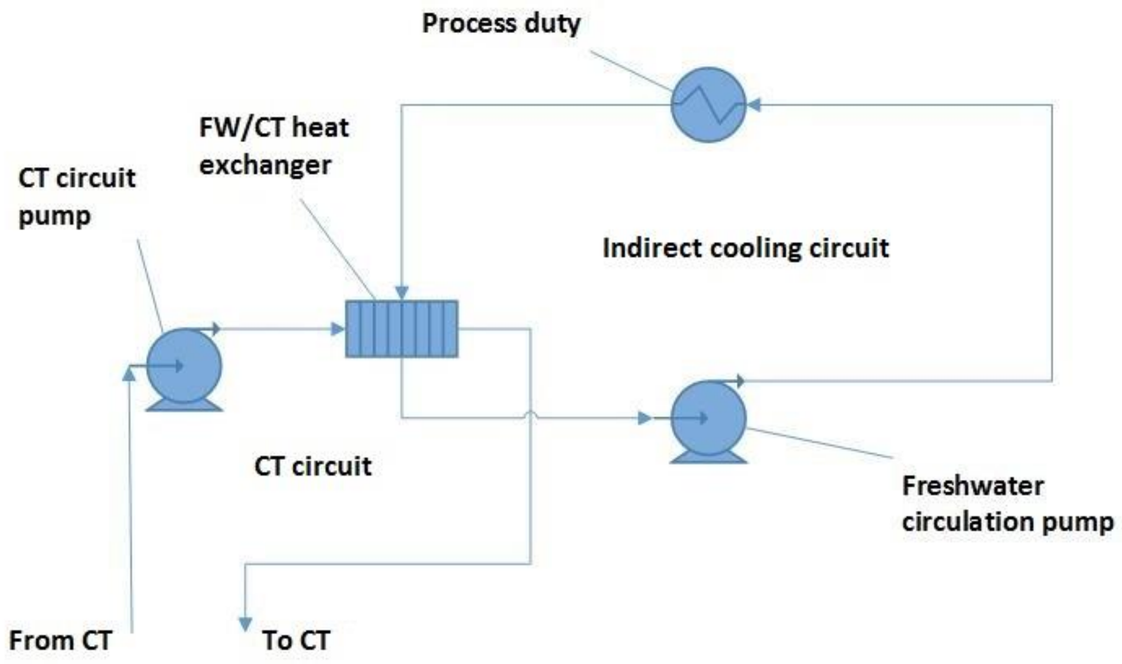

Figure 17: Possible configuration of a cooling tower as heat sink for an indirect cooing circuit

2.6.2 Seawater

Seawater represents a compact, reliable, energy- and cost efficient system for heat rejection for an indirect cooling system. Compared to cooling towers and air coolers, seawater has a more stable and predictable temperature range and the system configuration is less complex. The seawater intake is usually located far below the surface, for instance 150 meters for the Prelude FLNG (Shell, 2014). This decreases the effect of seasonal temperature variations and reduces the biological activity in the water. Due to governmental restrictions in some countries, however, seawater is not always available as heat rejection source. Additionally, many LNG facilities are located in a closed harbour, which makes the use of seawater an unfavourable option, as hot discharged water might recirculate back to the intake.

Figure 18 shows a possible configuration of seawater as heat sink for an indirect cooling circuit. Additionally, a weir box should be installed prior to the seawater outlet to prevent the water from flowing out and create slugging during restart, or in worst case vacuum in the system. The weir box head should be 5 meters above the highest seawater utility, resulting in a backpressure of 0.5 bar (Pettersen, 2015).

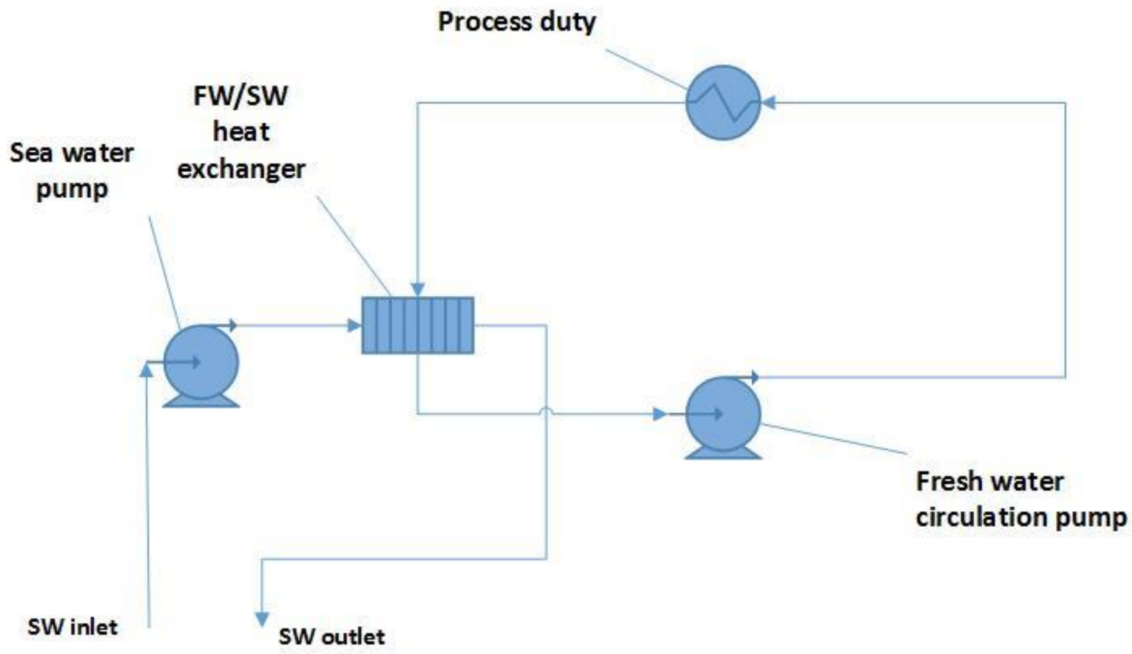

Figure 18: Possible configuration of seawater as heat sink for an indirect cooing circuit

2.6.3 Air Coolers

Air coolers are often economically advantageous at dry or semi-dry areas where the access of water is limited or the available water requires extensive treatment to reduce fouling. Air coolers are the second most used heat exchangers in the oil and gas industry, next to shell-and-tube heat exchangers. Due to the low heat transfer coefficient, air coolers will occupy a large area at the LNG plant. A possible configuration of air coolers as heat sink for an indirect cooing circuit is shown in Figure 19.

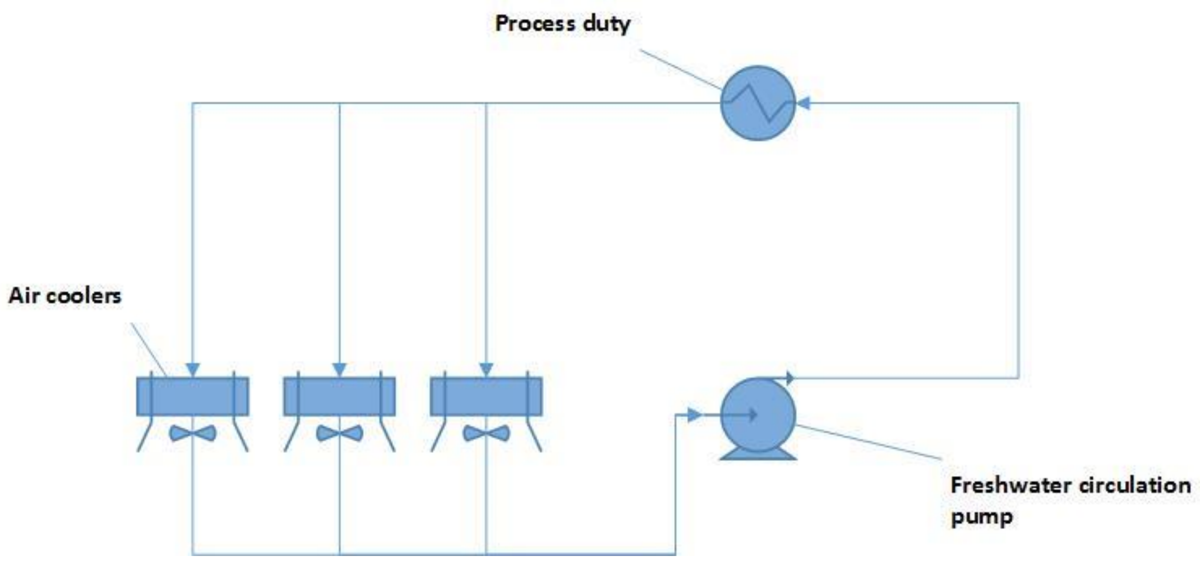

Figure 19: Possible configuration of air coolers as heat sink for an indirect cooing circuit

2.7 Power Generation

Regardless of the choice of refrigerant compressor driver, smaller duties such as pumps and other compressors are usually electrically driven. Power generation is usually done in a local power plant consisting of gas turbines, often with waste heat recovery. The efficiency depends largely on choice of driver and the amount of heat needed in the processes. If electric compressor drive is combined with a centralized gas turbine power plant Worley Parsons claims that the overall efficiency can be increased by 5-10% (Habibullah et al. 2009).

Another alternative is to use a combined cycle, which will offer a higher efficiency but also an added capital cost. Finally, the last alternative is to utilize power from the grid, generated externally.

The choice of power generation for projects in the development phase varies. For instance will LNG Canada in BC utilize renewable power for every system except the compressor drivers, delivered by BC Hydro through the grid (LNG Canada, 2014). Freeport LNG has suggested a GE combined cycle power plant to supply the electric compressor drivers (Business Wire, 2014).

2.8 Heat Generation and Transport Medium

The required heat needed in the different processes varies largely with the feed gas composition. The biggest consumers are usually the reboiler in the CO_2 removal process and regeneration in the dehydration process. For a rich feed gas, the fractionation train can also require a significant amount of heat as well. The heat transport medium can either be pressurized water, hot oil or steam and is usually generated in a heater or a gas turbine waste heat recovery system.

The use of steam offers advantages in terms of safety, as it is not flammable. However, as steam utilizes latent heat rejected during condensation, several pressure levels are required to deliver heat at different temperatures. High pressurized steam implies a certain risk in itself and the different piping and pressures can result in a complex system, depending on the required amount of heat and temperatures. Additionally, a large cooling demand, a more complex heat generation system and access to a large amount of makeup water is also required. A hot oil based system is simpler but utilizes sensible heat, which is much lower per unit mass when compared to steam. An indication of the difference in required mass flow and pipe dimensions is illustrated with a hypothetical heat demand of 50 MW and is presented in Table 5. The calculations are straightforward thermodynamics and the formulas are not included.

Table 5: Flow comparison for hot oil and steam (MatWeb, 2014) (Moran et al, 2012) (Gudmundsson, 2010)

Property	Abbreviation	Unit	Hot oil	Steam (at 15 bar)
Heat capacity	$C_{p,204°C}$	$kJ/kg\ °C$	2.59	-
Latent heat of vaporization	$h_{fg,198.3°C}$	kJ/kg	-	1947.3
Assumed temperature difference	ΔT	$°C$	25	-
Mass flow	$\dot{m}$	kg/s	772.2	25.7
Density	ρ	kg/m^3	854	7.59
Volume flow	$\dot{q}$	m^3/s	0.90	3.39
Maximum flow speed	u_{max}	m/s	6	60
Required cross sectional area	**A**	m^2	0.15	0.0565
Required diameter of pipe	**D**	m	0.437	0.268

Note that the recommended maximum flow speed for fluids is 6 m/s according to NORSOK P-100. For vapour, the recommended maximum flow speed is 60 m/s when the gas density is below approximately 12 kg/ m³ (Gudmundsson, 2010). This shows that the required pipe diameter for hot oil is larger than for steam, indicating that a hot oil system will require more space and material. Note that the wall thickness required due to the different pressure levels for steam has not been investigated further.

Next, the choice of heat generation depends largely on the choice of driver and power generation. If gas turbines are used as driver, a waste heat recovery system can become quite complex. If electric motors are used together with a centralized gas turbine power plant, integrating a WHR system will be easier. If a large amount of heat is required in the process, efficiency up to 70% can be achieved (Statoil, 2002). If a smaller amount of heat is needed, a steam based heating system can be implemented in a combined cycle power plant and a high efficiency can be maintained. For a configuration utilizing electric drive and power from the grid, the only option to generate heat is a furnace.

To keep the configuration as simple as possible, it is decided to use hot oil as heat transport medium in this study, which results in a less complex configuration. However, further studies on the potential for steam is recommended, especially if a large local centralized gas turbine power plant or a combined cycle plant is used.

2.9 Connections to Shore

Although the FLSO is moored to the shore, some movement must be accounted for. Especially high and low tide will result in some movement, depending on the location. This makes the flexible connections between the shore and the FLSO an important issue as several of these transports high pressure hydrocarbons and represents a high risk if any leakage should occur. Therefore, the number of connections will be of great importance when the location of different systems is evaluated.

The largest marine loading arm for CNG transfer on the market is currently 16 inch and can be operated at pressures up to 150 bar (Emco Wheaton, 2008) This is delivered by Emco Wheaton, which also delivers LNG marine loading arms up to 16 inches and 150 bar. The flow rate capacities and dimension of the available arms are presented in Table 6.

Table 6: Dimensions and flow rate of LNG loading arms

Arm Size	Flow rate [m^3/h]
4”	300
6”	600
8”	1100
10”	1700
12”	2500
16”	4000

2.10 Summary and Comparison of the Systems

Table 7: Summary and comparison of systems

System	Alternatives	Discussion
Liquefaction Process	PRICO Niche	PRICO operates more efficient and reliable. Niche is the one with lowest amount of flammable inventory.
Compressor Driver	GT Electric motor	GT is lighter and cheaper than electric motor. Electric motor is more reliable and less influenced by ambient temperature as long as sufficient power is available. An electric motor also requires less maintenance, which translates into higher annual production.
NGL extraction	Integrated Frontend	Integrated column must operate at reduced pressure, which decreases the liquefaction efficiency. May be critical for a lean gas, which requires a further pressure reduction. Frontend turboexpander requires more rotating equipment in series with the liquefaction, which may reduce the plant availability and reliability.
Power Generation	External power supply Local power plant	A local power plant with WHR can be beneficial in terms of efficiency, especially for a rich gas case. Power from an external source will reduce the capital cost and reduce the economic risk. It also opens up for significantly reduced emissions compared to a local power plant.
Cooling	Seawater Cooling Tower Air Coolers	Seawater is not considered to be an option in GoM and Russia. Cooling tower is not considered to be an option at locations with subzero temperatures. Seawater is a much more stable heat sink and offers a better temperature approach than the other alternatives. Cooling tower may give colder cooling water temperature than air coolers since it approach the wet bulb temperature.
Heat	Hot Oil Steam	Hot oil results in a simpler system than steam, which seems favorable for an at-shore FLNG. Steam might be combined with a combined cycle power plant, resulting in high efficiency of the power generation, especially for a lean gas.

3 Design Basis

In this thesis, two main scenarios with given climates have been studied. A potential location for each of these has been identified and the respective climate data has been used in the simulations. An initial scenario configuration of the alternatives below has been chosen based on ambient air temperature and feed gas composition. Each initial configuration is discussed further for the respective scenario in the subsequent sections. This discussion is based on background information given in Chapter 2 and findings in the specialization project (Corneliussen and Samnøy, 2014).

- Liquefaction process
- Driver
- NGL extraction
- Power generation
- Cooling method
- Heat generation
- Heat Transport Medium

Next, the configuration has been varied, one system at the time, and the result has been measured against the result from the initial scenario configuration. Especially production rate with 330 days of operation per year has been emphasized to identify any potential bottleneck. Ambient air temperature and local restrictions are taken into account resulting in that some configurations are not feasible and have therefore not been evaluated any further.

To get realistic results that can be applied on other potential locations, several variables must be changed at the time. Therefore, simulations have been performed on three selected subcases that are based on gas composition, site-specific conditions and regulations.

3.1 Temperature and Production Definitions

All simulations have been performed with two air temperatures, design and high. The results based on the design temperature illustrates the normal operating conditions with average temperature and the results from the high temperature describes the peak conditions, which the plant must be able to handle occasionally without any major loss in production. The high temperatures are obtained with the following procedure.

- Maximum ambient air temperature measured at the proposed location is obtained.
- 5°C is subtracted to obtain the high temperature used in the simulations.

The reason for the subtraction of 5°C is that the maximum temperatures occur rarely and the cost of constructing the facility for these is considered to exceed the lost revenue from a slightly reduced production rate. For the location using seawater cooling the following procedure has been used to calculate the temperatures.

- Average seawater temperature is obtained (design temperature) at the proposed location.
- 3°C is added to obtain the high temperature.

The reason for the low variation between average and high seawater temperature is that it varies little through the year, especially if the intake is locates far below the surface. As a result of this use of two temperatures, consequences of each of the potential system configurations for the locations are identified and quantified with varying temperatures and becomes valid for a large part of the year. Note that the plant is assumed to be designed for the design temperature, meaning that the results obtained from high temperature are with the same constraints, such as pressure ratios in the compressors, as for design temperature. This does not provide an optimal result at high temperature. For a detailed discussion regarding this topic, see Appendix E.

The production rate of LNG, LPG and condensate is measured in million tonnes per annum (MTPA), which is the standard production unit in LNG plants. At high temperature the production decrease and when this is given in production per year, it might give an inaccurate picture. It is important for the reader to know that the production at high temperature only represents a small part of the year even though it is measured in MTPA.

3.2 Gas Composition

The lean gas composition showed in Table 8 is provided by the supervisor for this study and will be used in all lean gas simulations (Pettersen, 2015).

Table 8: Lean gas composition (Pettersen, 2015)

Component	Mole fraction
Methane	0.9654
Ethane	0.0123
Propane	0.0044
i-Butane	0.0009
n-Butane	0.0011
Neopentane	4.E-5
i-Pentane	0.0004
n-Pentane	0.0003
Hexane	0.0003
Heptane	0.0013
Octane	0.0001
Nonane	0.0003
C10+	0.0012
Benzene	4.E-5
E-Benzene	0.0006
Toluene	5.E-5
Xylene	6.E-5
Nitrogen	0.0072
Carbon Dioxide	0.0002
Hydrogen Sulphide	2.E-5
Water	0.0041

The rich gas composition showed in Table 9 is based on the gas composition for the Snøhvit field. Note that the nitrogen content of 2.53 mol% might need an advanced end flash system to achieve the required product specification of below 1 mol%.

Table 9: Rich gas composition (Christiansen, 2012)

Component	Mole fraction
Methane	0.8102
Ethane	0.0503
Propane	0.0253
i-Butane	0.0040
n-Butane	0.0083
i-Pentane	0.0021
n-Pentane	0.0031
Hexane	0.0035
Heptane	0.0039
Octane	0.0032
Nonane	0.0014
C10+	0.0031
Benzene	0.0008
Toluene	0.0009
p-Xylene	0.0006
Nitrogen	0.0253
Carbon Dioxide	0.0526
Hydrogen Sulphide	5.e-6
Phenol	2.e-6
Helium	0.0002

3.3 Scenario 1 – Warm Climate

The first scenario is for a LNG facility located in a warm climate. Climate data has been collected from Houston, Texas, corresponding to the Gulf of Mexico and is presented in Table 10.

Table 10: Climate data from Houston, Texas (WeathersSpark, 2015) (NRK/NMI, 2015) (Intellicast, 2015) (NOAA, 2015)

Data	Value	
Design Air Temperature	20°C	
Maximum Air Temperature	41°C	
High Air Temperature	36°C	(maximum - 5°C)
Design Relative humidity	70 %	
High Relative humidity	90 %	
Design WB temperature	17°C	
High WB temperature	34°C	
Sea water design temperature	23°C	

Note that 5°C has been subtracted from the maximum temperature to define a high temperature for the simulations. A thorough review of the weather for 2014 indicates that Houston experienced 81 days with temperatures above 36°C (WeathersSpark, 2015). This corresponds to 22.2% of the year in total. However, number of hours above 36°C was only 346, which corresponds to 3.95%. In other words, the temperatures used in the simulations are valid for roughly 96% of the year.

An integrated NGL extraction column has not been simulated for the scenarios or subcases with lean feed gas. This decision is based on previous work on NGL extraction, with a lean feed gas composition similar to the one used in this study (Kusmaya, 2012). To achieve a sufficient vapour/liquid separation, the report concludes that the column pressure must be approximately 38 bar. This results in an increased specific power of 70% (kWh/tonne) compared to the base case simulated with a feed gas slightly leaner than the rich gas for this study. Meanwhile, the frontend turboexpander configuration only experienced a specific power increase of 6% compared to the base case, as the gas was liquefied at a pressure closer to the critical point (Kusmaya, 2012). This indicates that an integrated column is not favourable for a lean feed gas scenario and will therefore not be studied any further in this report.

The initial configuration will be very similar to the Lavaca Bay project, with the exception of NGL extraction method and is considered to be the most favourable configuration for this scenario. The similarity also offers a good basis of comparison for some of the results obtained in this study. A brief discussion and argumentation of the selected systems for the initial scenario configuration is given in Table 11.

Table 11: Initial configuration and alternative configuration for Scenario 1

System	Initial Configuration	Alternative Configuration	Discussion
Liquefaction process	PRICO	Niche	PRICO is more efficient than Niche.
Driver	Gas turbine	Electric motor	GT is cheaper and lighter.
Gas comp	Lean	Rich	-
NGL	Frontend		Integrated is not considered to be favourable for lean gas. GT driven booster compressor since the refrigerant compressors are GT driven.
Power generation	Local	External	Local power plant with Trent 60 Gas Turbine Generator (GTG). Same driver for compressor and power generation makes maintenance simpler.
Cooling	Cooling tower	Air cooler	CT can supply a lower cooling water temperature as long as the air is dry.
Heat	Waste heat recovery	Heater	WHR matches the initial choice of power generation best.

3.4 Scenario 2 – Cold Climate

Scenario 2 is for a LNG facility located in a cold climate. Climate data have been collected from Hammerfest, Norway, which is a potential location for an at-shore LNG plant producing from offshore fields. The climate data are presented in Table 12. Note that the design temperature is based on the average yearly air temperature for the last 15 years.

The seawater design temperature is based on numbers from Hammerfest and provided by the supervisor of this study. A temperature increase of 3°C is assumed to obtain the high temperature. This small variation during the year is a fair approximation when the fact that the seawater inlet will be located far below the surface is taken into account.

Table 12: Climate data from Hammerfest, Norway (NRK/NMI, 2015)(Pettersen, 2015)

Data	Value
Design Air temperature	1°C
Maximum Air Temperature	28°C
High Air Temperature	23°C
Relative humidity	Not relevant
Seawater design temperature	6°C
Seawater high temperature	9°C

The initial scenario configuration presented in Table 13 is based on Melkøya, which is the only large LNG plant currently producing in Norway, with rich feed gas, electric drive and seawater cooling.

The difference in this case compared to the Melkøya case is the power generation. Melkøya utilizes five LM6000 with WHR but due to recent environmental restrictions, the Norwegian government does not allow any use of gas turbines without a Carbon Capture and Storage (CCS) system. The CCS system is regarded to be too expensive and has not been studied any further. Therefore external power is the initial scenario configuration. As a result, heat needed in the process will be supplied by a local furnace, fed by end flash, BOG, return gas etc. The environmental focus of the Norwegian government also indicates that the system with the smallest environmental impact should be chosen for the initial scenario configuration.

However, to avoid linking too much towards a fixed location, gas turbine driver and power generation will still be simulated, but as an alternative configuration. A GTG power plant illustrates the combination of a local power and heat generation with electrical drive.

Table 13: Initial and alternative configuration for Scenario 2

System	Initial Configuration	Alternative Configuration	Discussion and Comments
Liquefaction process	PRICO	Niche	PRICO is more efficient than Niche
Driver	Electric motor	Gas turbine	Governmental restrictions.
Gas comp	Rich	Lean pipeline	-
NGL	Frontend	Integrated	Electrical driven due to governmental restrictions
Power generation	External	Local GTG with WHR	Governmental restrictions.
Cooling	Sea water	Air coolers	SW is more efficient, smaller and gives a lower cooling water temperature than AC. SW also varies less, making the production more stable and easier to operate.
Heat	Burner	Waste heat recovery if GT is used	Result of governmental restrictions on driver and power generation.

3.5 Potential Locations and Subcases

Some other possible locations for scenario 1 and 2 including considerations regarding climate, infrastructure, governmental restrictions and gas composition are presented in Table 14. Based on these considerations and the feasibility of a project at the location, three subcases are chosen to proceed with. By selecting these subcases, some of the system configurations regarded as the most relevant will be simulated. Note that the lean and rich feed gas compositions used in the scenario simulations will also be the basis for the subcases. This is a source of error as the actual feed gas composition for the subcase locations differs from the ones used in the scenario. However, it gives a better basis for comparison of the different configurations.

Table 14: Possible locations for scenario 1 and 2

Location	Climate	Infrastructure	Governmental Restrictions	Gas Composition
British Columbia, Canada	Cold	Good/moderate	Environmental focus	Lean Pipeline
The Barents Sea, Norway	Cold	Moderate	GT's can not be used	Rich
The Barents Sea, Russia	Cold	Poor	SW can not be used	Lean
East Coast of Africa	Warm	Poor		Lean
Gulf of Mexico	Warm	Good	SW may not be feasible due to closed harbour	Lean Pipeline
North Western Coast of Australia	Warm	Moderate		Rich

3.5.1 Subcase A – Prince Rupert, British Columbia (Cold Climate)

A highly relevant location for an at-shore FLSO is Prince Rupert in British Columbia. The weather data for this location is presented in Table 15. Several high capacity projects have been proposed recently and indicate a large focus on the environment. For instance, LNG Canada, if constructed, will utilize GE LMS100 gas turbine as compressor drive (LNG Canada, 2014), which has an efficiency of 53% (GE Power and Water, 2015). The required electricity is supplied by renewable energy sources through the grid. Although this is an onshore project it still gives an indication of the environmental focus in this region. For this subcase it is assumed that only the best available technology in terms of efficiency and emissions will be considered for driver and power generation.

With the environmental focus in mind, a configuration for this subcase has been chosen and is presented in Table 16. This is the configuration that is regarded to be most favourable for this area. PRICO liquefaction modules, electric drive combined with external power generation potentially opens up for a very environmental friendly plant.

The grid is assumed to be able to deliver all the required electric power. A consequence of this choice is that any process heat must be generated in a heater.

Table 15: Weather data for Prince Rupert, BC Canada (ClimaTemps, 2015) (Weather2, 2015) (WWC, 2015)

Temperature	Value
Design Air Temperature	6.9°C
Maximum Air Temperature	33°C
High Air Temperature	28°C

Table 16: Subcase A configuration

Configuration Variables	Subcase A Configuration
Liquefaction process	PRICO
Compressor driver	Electric motor
Gas composition	Lean Pipeline
NGL extraction	Frontend, electrical driven booster compressor
Power generation	External
Cooling	Air Coolers
Heat generation	Burner

3.5.2 Subcase B – Northwest Russia (Cold Climate)

Subcase B is for a LNG plant potentially located in Northwest Russia. Climate data has been collected from Murmansk, which is close to a potential location for LNG production from the offshore Shtokman gas field. This is presented in Table 17. An at-shore FLNG production requires the gas to be brought 500 km to shore through a multiphase pipeline. This distance is far beyond the longest multiphase pipeline today and is infeasible with current technology. However, the at-shore case offers an attractive combination of system configuration and has therefore been chosen to look further into.

Due to governmental restrictions, seawater cooling is not an option, making air coolers the only cooling solution (Pettersen, 2015). Furthermore, the gas is lean, thus frontend is the only option evaluated in this study. Due to the low average air temperatures for this location, gas turbines will perform effectively and have therefore been selected for both compressor driver and local power generation. Heat required in the process is limited and will be supplied by waste heat recovery from the local power generation. The complete configuration is presented in Table 18.

Table 17: Weather data for Murmansk, Northwest Russia (NRK/MMI, 2015) (Weather and Climate, 2015)

Temperature	Value
Design Air temperature	-1°C
Maximum Air Temperature	33°C
High Air Temperature	28°C
Seawater Design Temperature	Not relevant

Table 18: Subcase B configuration

Configuration Alternative	Subcase B Configuration
Liquefaction process	PRICO
Compressor driver	GT
Gas composition	Lean pipeline
NGL extraction	Frontend, GT driven booster compressor
Power generation	Local GTG
Cooling	Air coolers
Heat generation	WHR

3.5.3 Subcase C – Northwest Australia (Warm Climate)

Subcase C is located in Northwest Australia, which experience very high ambient air temperatures, as shown in Table 19. Due to the high temperatures, gas turbine compressor drive will not be efficient compared to the colder cases. Linear interpolation of Figure 12 shows that the power output from a RR Trent 60 will be approximately 34.3 MW at high temperature with the derating factors included. The high air temperature also results in higher cooling water temperature, reducing the plant efficiency further.

From an efficiency point of view, electrical drive with a constant power output would be better. However, the construction cost in Australia is already extremely high and GT is the driver option with the lowest CAPEX. Additionally, the gas composition contains probably more CO_2 and is richer than the rich gas used in the simulations, hence the required process heat will be high. With gas turbines, waste heat can be used to cover some of the heat demand. Based on this, GT driver and power generation seems to be the most favourable alternative and will therefore be used for Subcase C. The complete configuration is presented in Table 20.

Table 19: Climate data for Port Headland, Northwest Australia (AGBM, 2015) (WeatherSpark, 2015)

Temperature/humidity	Value
Design Air temperature	26.5°C
Maximum Air Temperature	49°C
High Air Temperature	44°C
Design Relative humidity	41%
High Relative humidity	85%
Design WB temperature	19°C
High WB temperature	40°C

Table 20: Subcase C configuration

Configuration Alternative	Subcase C Configuration
Liquefaction process	PRICO
Compressor driver	GT
Gas composition	Rich gas
NGL extraction	Frontend
Power generation	Local
Cooling	Cooling tower
Heat generation	WHR

3.6 Available Compressor Power at Design and High Temperatures

Based on the RR Trent 60 performance chart in Figure 12, derating factors and obtained temperatures for each scenario and subcase, the available compressor power for each gas turbine is estimated and presented in Table 21. As mentioned in chapter 2, the electrical motors have a constant power output despite changes in temperature. To obtain comparable results for each scenario and subcase, the designed power output will be equal to the gas turbine output at design temperatures for each location.

However, the different power output from the electric motors makes the results from different locations less comparable.

Note that for Scenario 2, Subcase A and Subcase B, a de-icing system must be installed, since the gas turbines may experience temperatures below 3°C.

Table 21: RR Trent 60 power output at design and high temperatures for all scenario and subcases

	Trent 60 available power at design temperature [MW]	Trent 60 available power at high temperature [MW]	Electric motor assumed power [MW]
Scenario 1	43.4	36.0	43.4
Scenario 2	51.7	41.0	51.7
Subcase A (British Columbia)	49.5	38.4	49.5
Subcase B (Northern Russia)	51.5	38.4	Not relevant
Subcase C (Northwest Australia)	40.3	34.3	Not relevant

4. Simulation Model

In this chapter, the HYSYS models of the different system configurations are presented. The main focus has been the liquefaction systems and NGL extraction and these are described in detail in this chapter. The rest of the process systems are described briefly in Section 4.2. The process stages for the initial configuration in Scenario 1 and Scenario 2 are shown in Figure 20 and Figure 21.

The PRICO and Niche liquefaction processes have been optimized in HYSYS by using the Hyprotech SQP optimizer. The objective is to establish a good comparison basis between the different configurations. Optimizing the simulation models manually would have been very time consuming, and the result would probably have been poorer than with the optimization tool. This especially applies for the PRICO process, as the optimal MR composition changes for each scenario configuration and subcase.

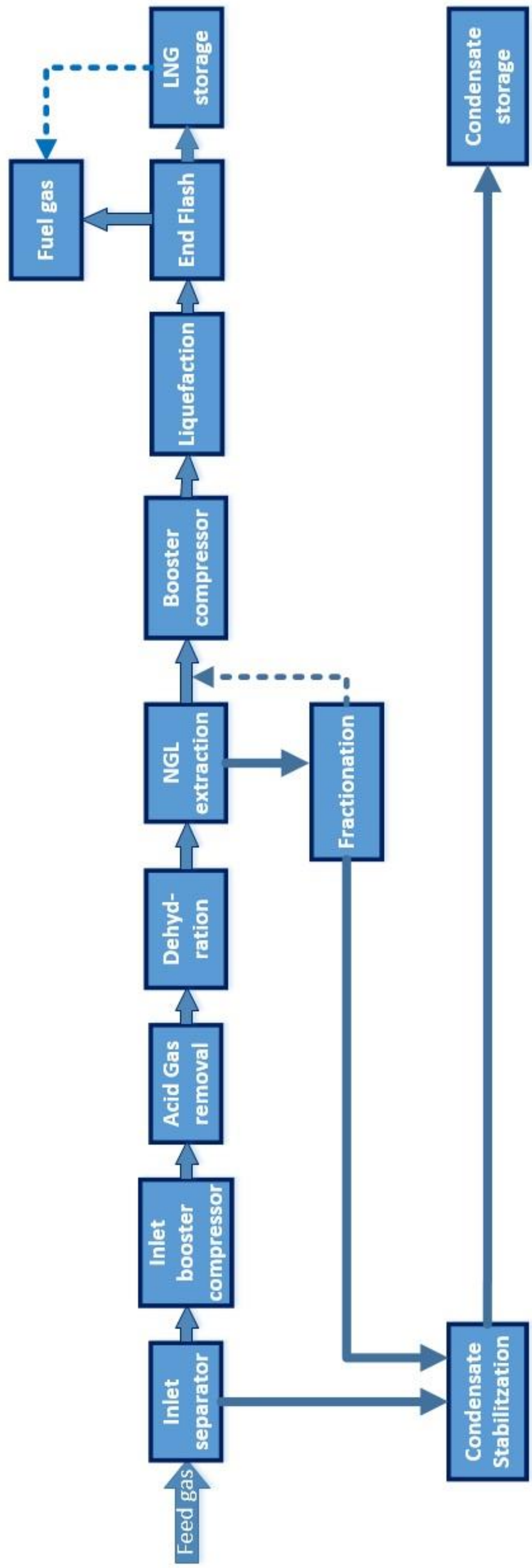

Figure 20: Process stages for the initial configuration in Scenario 1

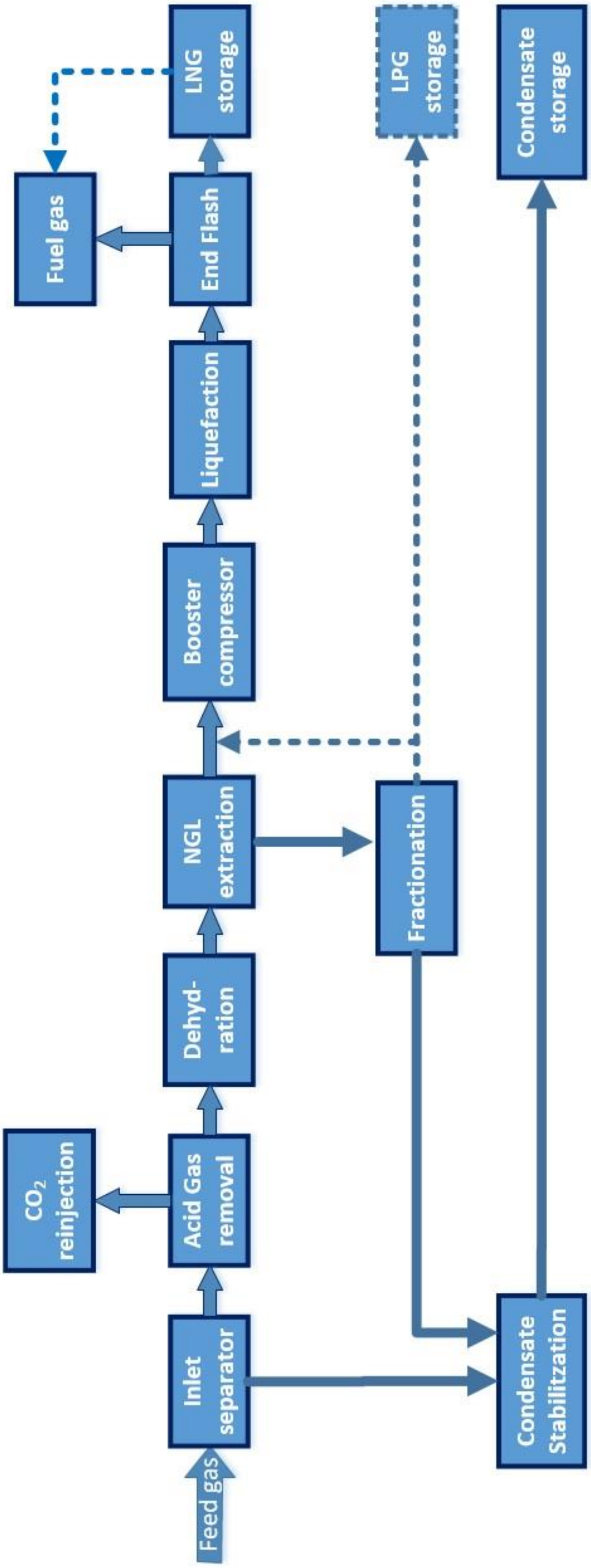

Figure 21: Process stages for the initial configuration in Scenario 2

4.1 Assumptions

General assumptions made for the HYSYS model are based on numbers provided by the supervisor of this thesis, and are listed in Table 22 and Table 23.

Table 22: Temperature approach for the heat exchangers (HX) in the HYSYS models

Unit	Temperature Approach
Water to water HX	5 °C
Water to air HX (Air coolers)	10°C
Cooling tower temperature approach	3°C (to WB temperature)
Water to HP gas HX	5°C
Water to MP gas HX	7°C
LNG HX	3°C

Table 23: Other assumptions used in the HYSYS models

Unit	Assumption
Process side pressure loss for all HX except air cooled HX	0.5 bar
Process side pressure loss for air cooled HX	1.0 bar
Pump adiabatic efficiency	75%
Compressor adiabatic efficiency	75%
Frontend NGL compander adiabatic efficiency	75%
Niche compander polytrophic efficiency	80%

The assumption of 75% adiabatic efficiency in the compressors and frontend NGL companders can be regarded as somewhat conservative. This often gives a polytrophic efficiency lower than 80%, which is a more usual assumption for compressors (Pettersen, 2015). By experimental work with the process models in HYSYS, it was found that the polytrophic efficiency of the compressors and frontend NGL extraction companders was ranging from 73-78% when the adiabatic efficiency was sat to 75%. This did not affect the simulation results particularly, meaning that the results were considered as reasonable and within range of what could be expected at the given conditions.

In the Niche liquefaction process, however, the polytrophic efficiency dropped below 70% in the turbine part of the companders, which led to very low specific power. To compensate for this, the polytrophic efficiency of the Niche companders were sat to 80%, which resulted in a more reasonable specific power of the liquefaction process. This is further described in Appendix E.

4.2 Process Model Overview

The complete model of the Scenario 2 initial configuration is shown in Appendix D. This model includes all the processes described in this chapter, and is therefore used as reference model. Note that some power, heating and cooling duties have been scaled linearly from other plants and references to provide an accurate result. This is specified when used.

4.2.1 Inlet Separator and Condensate Stabilization

For all the scenarios and subcases, light components are splitted from heavy components in an inlet separator (slug catcher) as it enters the facility. The inlet conditions for all the scenarios and subcases are shown in Table 24. The inlet temperature for Scenario 2 (Northern Norway), Subcase B (Northwest Russia) and Subcase C (Northwest Australia) with gas from subsea reservoir, corresponds to the average seawater temperature at these locations. The pipeline gas in Scenario 1 and Subcase A can have varying inlet pressure. Therefore, the feed gas pressure is set to a "worst case" of 40 bar, and an inlet booster compressor is installed after the inlet separator.

Table 24: Feed gas conditions for the scenarios and subcases at design temperature

	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
Inlet pressure	40 bar	70 bar	40 bar	70 bar	70 bar
Inlet temperature	26 °C	6 °C	26 °C	6 °C	10 °C

A condensate stabilization process is modelled downstream of the inlet separator, as shown in Figure 22. The heavy component stream is reduced in pressure before it enters a 3-phase separator where the gas, water and MEG (if any) are separated from the HHC liquids. The HHC liquids are further reduced in pressure and enter the condensate stabilization column as a two-phase stream. To meet the sales requirements, specifications for the column are Reid vapour pressure of 11.5 psi for the condensate exiting at the bottom. The reboiler temperature is then approximately 160°C, and the column operates at a pressure of 10-10.2 bar. Cooling and storage of the condensate product is also included in the model.

By compression and temperature reduction of the light components exiting at the top, approx. 4.4 % of the molar flow is sent back to the column as reflux. The remaining gas flow is mixed with the gas from the 3-phase separator, recompressed and reinjected into the main gas stream.

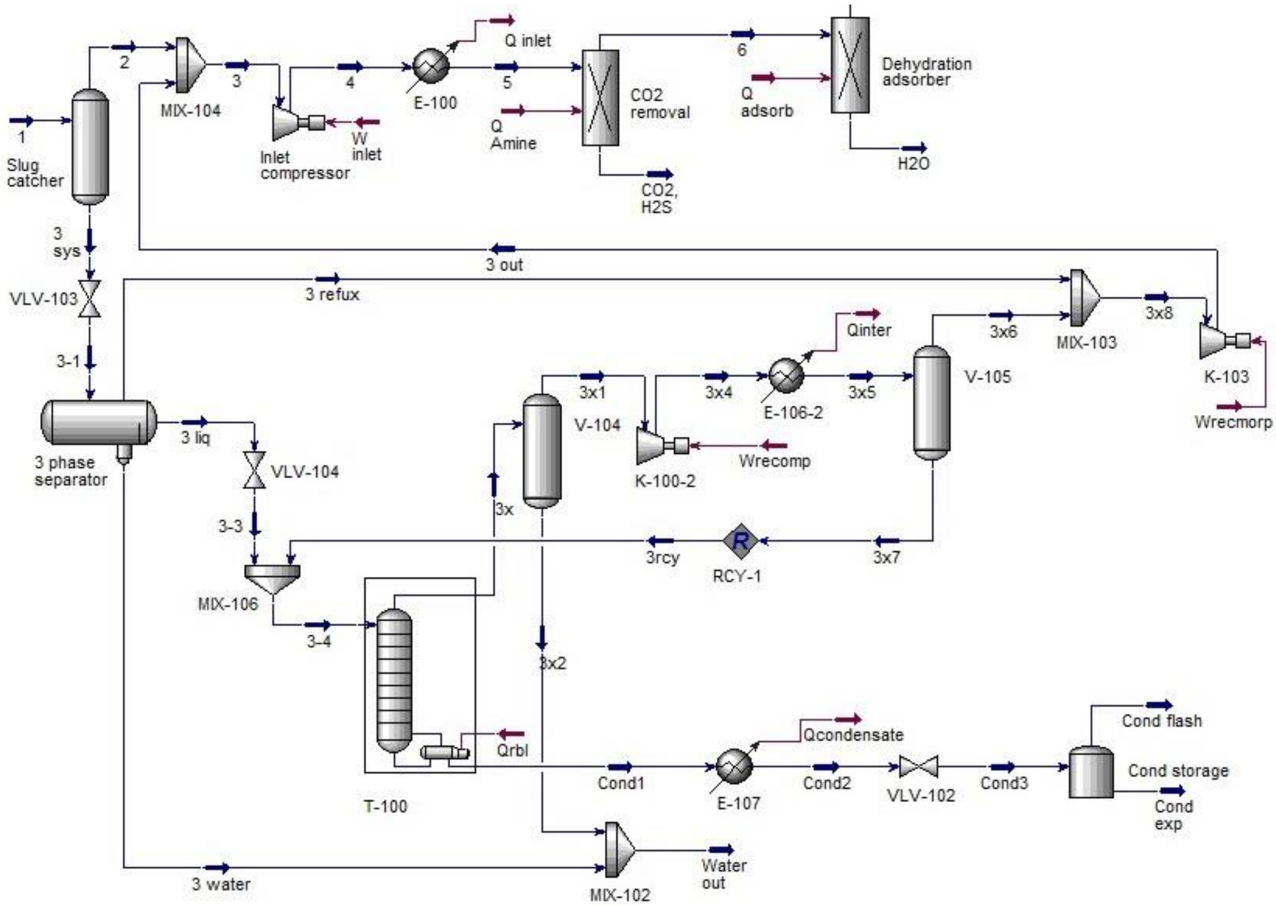

Figure 22: Inlet separator, condensate stabilization and storage, inlet gas compressor, acid gas removal and dehydration in HYSYS

The gas inlet and condensate stabilization system for the subsea gas reservoir scenarios and subcases are based on presence of water and MEG in the inlet gas stream, even though the compounds are not present in the composition. Therefore, no inlet gas preheater is added before the inlet gas separator.

4.2.2 Acid Gas Removal and Dehydration

Acid gas treatment and dehydration are both simulated with component splitters. The acid gas removal process has proven to be difficult to simulate with realistic numbers in HYSYS. This was also experienced in the master project thesis in the fall of 2014 (Corneliussen and Samnøy, 2014). The dehydration process cannot be simulated in HYSYS, as it is not a static process.

For the rich gas scenarios and subcases, linearly scaled numbers from Melkøya based on production are used to get an estimate of the required demand of heating, cooling and power for these processes. For the lean gas scenarios and cases, the numbers are scaled linearly based on production from a lean gas process model for FLNG provided by supervisor.

4.3 NGL Extraction

4.3.1 Frontend NGL extraction

The frontend turboexpander NGL extraction process is shown in Figure 23. The simulation model is based on the lean gas processing and liquefaction model provided by supervisor (Pettersen, 2015). However, some temperatures and pressure levels are sat by trial and error to achieve a converging process with reasonable results, due to different gas composition and inlet conditions than the reference model. Note that the downstream booster system is shown in Figure 25.

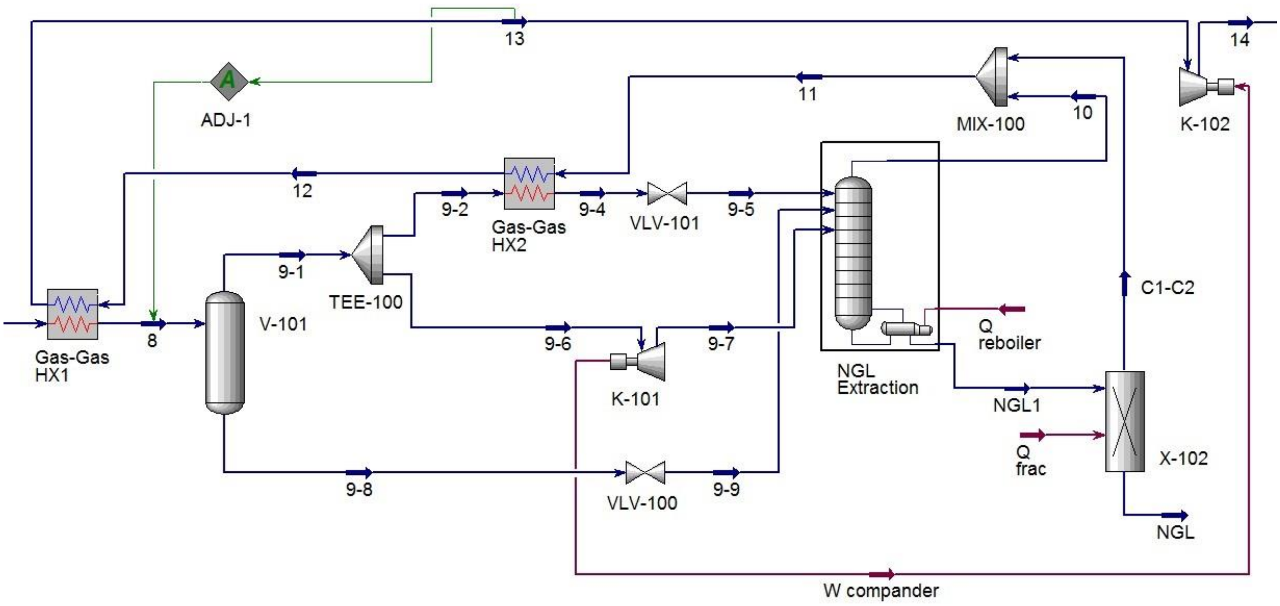

Figure 23: Frontend NGL extraction process in HYSYS

After acid gas removal and dehydration, the natural gas enters Gas-Gas HX1 at a pressure of 62 bar. The inlet temperature corresponds to the available cooling water temperature at the different scenarios and subcases. The gas is then cooled by the precooled and partly condensed top product of the NGL extraction column mixed with light component reinjection from fractionation. Liquid is then separated from the gas in a two-phase separator, and enters the column at mid stage after expansion to the column pressure.

The saturated vapour leaving the top is splitted in two equal streams; one entering the top of the column, and the other at the bottom. The one entering at the top is precooled and partly/completely liquefied in Gas-Gas HX2, before the pressure is reduced to the top stage pressure (30.5 bar). The temperature after the expansion valve is adjusted for the different scenarios and cases. For Scenario 1 and Scenario 2 with initial process configuration, it is sat to -80°C and -95°C respectively. This is based on trial and error to achieve sufficient NGL extraction. The other gas stream is expanded in the turboexpander part of the compander, and enters the column at the bottom stage pressure (31.5 bar).

The NGL extraction column is modelled as a reboiled absorber in HYSYS, specified with a reboiler temperature of 20°C. Due to model simplifications, the fractionation system has been modelled as a simplified component splitter with split factors of 0 and 1. Light hydrocarbons (C1 and C2 for rich gas, and C1-C4 for lean gas) are reinjected into the top product of the column. Further NGL processing is not included in this study.

Next, cold lean gas leaving the top of the column is heated by Gas-Gas HX2 and Gas-Gas HX1. An adjust function sets the Gas-Gas HX1 outlet (stream 13) to 3°C lower than stream 8, and thereby maintaining a temperature approach of 3°C in the heat exchanger. The lean gas is then compressed by the compressor part of the compander (K-102), intercooled, and then compressed by a booster compressor (K-100) to a pressure of 60 bar. This system is shown in Figure 25. A recycle function is installed between the intercooler and the booster compressor. This function is modelled for recycling the temperature, pressure and composition of the gas, as the mass flow is one of the variables to be optimized in the liquefaction process optimization. This will be further described in Chapter 4.4.

4.3.2 Integrated NGL extraction

The integrated NGL extraction process is only modelled for the PRICO liquefaction process in Scenario 2. The HYSYS model of the process is shown in Figure 24. Due to optimization complexity in HYSYS, the NGL extraction column is modelled as a two phase separator instead of a reboiled absorber. The consequences of this are further discussed in Chapter 5.2.3.

After LNG HX1, the partially liquefied natural gas stream is expanded in an expansion valve to a pressure of 50 bar and a temperature of -67°C (stream 20). The natural gas liquids are then separated in the two phase separator. As for frontend NGL extraction, the fractionation system has been modelled as a simplified component splitter with split factors of 0 and 1, and further NGL processing has not been included. C1 and C2 is reinjected into the top product of the separator, and the lean gas then enters LNG HX2 at a pressure of 50 bar and a temperature of -70°C.

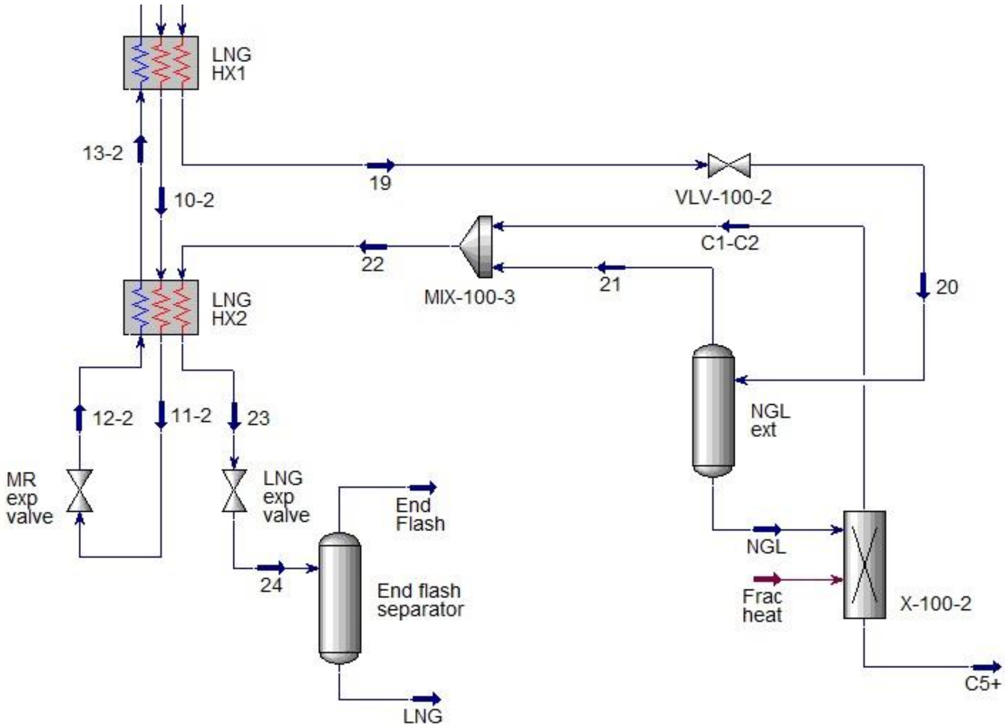

Figure 24: Integrated NGL extraction process in HYSYS

4.4 Liquefaction Systems

The simulation models of the liquefaction processes are based on pervious work and a reference model provided by supervisor (Corneliussen and Samnøy, 2014) (Pettersen, 2015). As for the NGL extraction models, some temperatures and pressure levels are sat by trial and error to achieve a converging process with reasonable results, due to different gas composition and inlet conditions than the reference model.

4.4.1 PRICO Liquefaction Process

The HYSYS model of the PRICO liquefaction process is shown in Figure 25. After the booster compressor and precooling, the pretreated natural gas enters the LNG heat exchanger with a pressure of 59.5 bar (stream 18). The inlet temperature corresponds to the available cooling water temperature at the different scenarios and subcases. The HX outlet is specified with a temperature of -155°C. Next, the liquid is expanded to a pressure of 1.05 bar in a J-T valve, which leads to sufficient end flash of nitrogen. The remaining liquid is sent to LNG storage.

The MR-circuit is modelled with two compressor stages and intercooling. Between the compressor stages, any present liquid will be separated from the MR stream. The pressure of this liquid is increased in a MR pump, as the gas enters the second stage of the MR compressor. A set function is used to maintain the same pressure level after the second stage compressor and the MR-pump. After the second MR cooler, the MR flow enters the LNG heat exchanger at the high pressure side and is fully condensed at the outlet (stream 11-2), where the temperature is set to -155°C. The cold, liquid MR flow is expanded in the MR expansion valve, and enters the cold side of the LNG heat exchanger with low pressure and some vapour present.

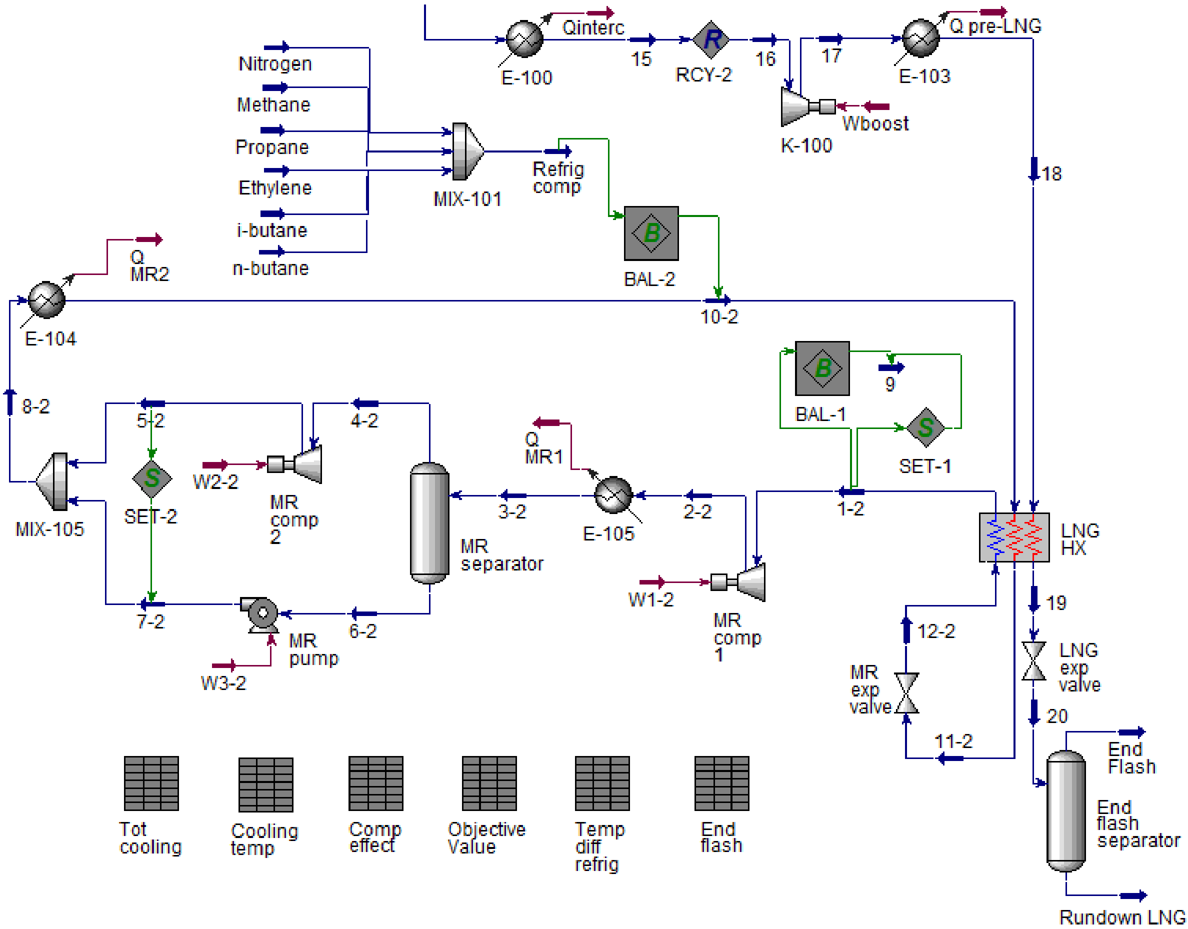

Figure 25: HYSYS simulation model of the PRICO liquefaction process

The low pressure, middle pressure and high pressure of the MR circuit are adjusted to achieve the optimal process at given ambient conditions. The same goes for the MR composition, which is defined in the stream "Refrig comp", and transferred to stream 10-2 by a balance function. These adjustments are done by the optimizer function Hyprotech SQP in HYSYS, which is further described in Appendix E.

4.4.2 Niche Liquefaction Process

As described in Chapter 2.2.2, the Niche process utilizes the refrigerant capacity of an open natural gas circuit and a closed nitrogen circuit. The HYSYS model of Niche is shown in Figure 26. Natural gas enters in the stream "NG inlet" with the same inlet conditions as for PRICO (59.5 bar and scenario dependent temperature). Based on limited LNG production capacity of Niche and the available compressor power in Scenario 1, the Niche process is modelled as one of three parallel trains, which means that the pre-treated NG inlet stream is splitted in three before entering the Niche process (not included in Figure 26), and the inlet natural gas flow is much lower than for PRICO.

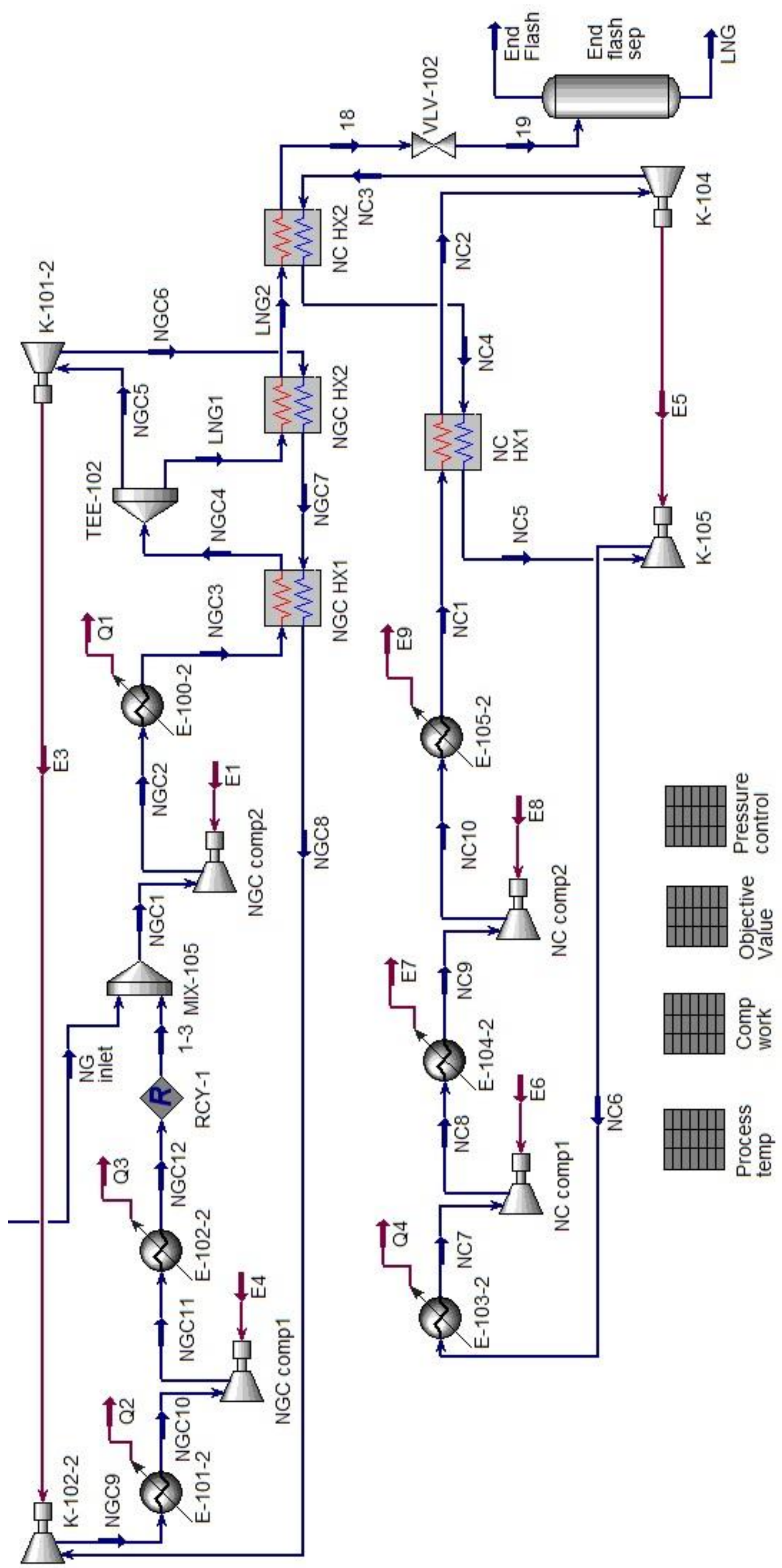

Figure 26: HYSYS model of the Niche process

Next, the inlet stream is mixed with the open natural gas refrigerant circuit and enters the second stage compressor. The gas is then precooled by the process cooling circuit and the LP side of the NG refrigerant circuit. TEE-102 then splits the natural gas flow by the relation of approximately 0.25/0.75, where 25% is sent further to liquefaction (stream LNG1). The natural gas is liquefied in NGC HX2 and subcooled in NC HX2. After expansion in VLV-102, the pressure and temperature is set to 1.05 bar and -163.2°C, as for PRICO. Methane and nitrogen in vapour phase are flashed off in the end flash separator, and the LNG product is sent to storage.

The two cooling circuits are both based on the principle of three compressor stages with intercooling. The first compression stage is the compressor side of a compander, driven by turbine expansion from the HP to LP side of the process. For Scenario 1, the second and third stage compression is driven by gas turbines, meaning one gas turbine is needed per circuit with available power given by the ambient temperature. For Scenario 2 on the other hand, the second and third stage compressors are driven by electrical motors, which can be adjusted to achieve an optimal process. The compressor drive and the pressure levels of the two circuits are the main basis for optimization of the Niche process, which is further described Appendix E.

4.5 End Flash System, Storage and BOG

The flow rate and composition of the end flash and BOG is necessary for calculating the available energy for GT drive, power generation and heating. Thereby, the end flash system and storage tank with correct BOG rate needs to be simulated. The HYSYS system is shown in Figure 27.

The storage tank is modelled with 0.15% BOG rate, which is a normal assumption for hull storage of LNG (Pettersen, 2015). The 0.15% BOG rate refers to 0.15% of the storage tank volume, and thereby 375 m^3/day with a storage capacity of 250 000m^3 as described in Chapter 2.1.4. The 375 m^3/day BOG rate need to be defined as an actual flow rate in liquid phase. A cooler (E-102) is therefore added after the stream "BOG" to liquefy the boil off gas, and the BOG rate of 375 m^3/day in the stream "BOG liq" is obtained by adjusting the energy stream "Heat leak" of the storage tank. A heater (E-101) is then added to re-heat the stream to the original BOG temperature, before it is mixed with the end flash and recompressed. Note that E-102 and E-101 is modelled without pressure losses to avoid affecting the process downstream.

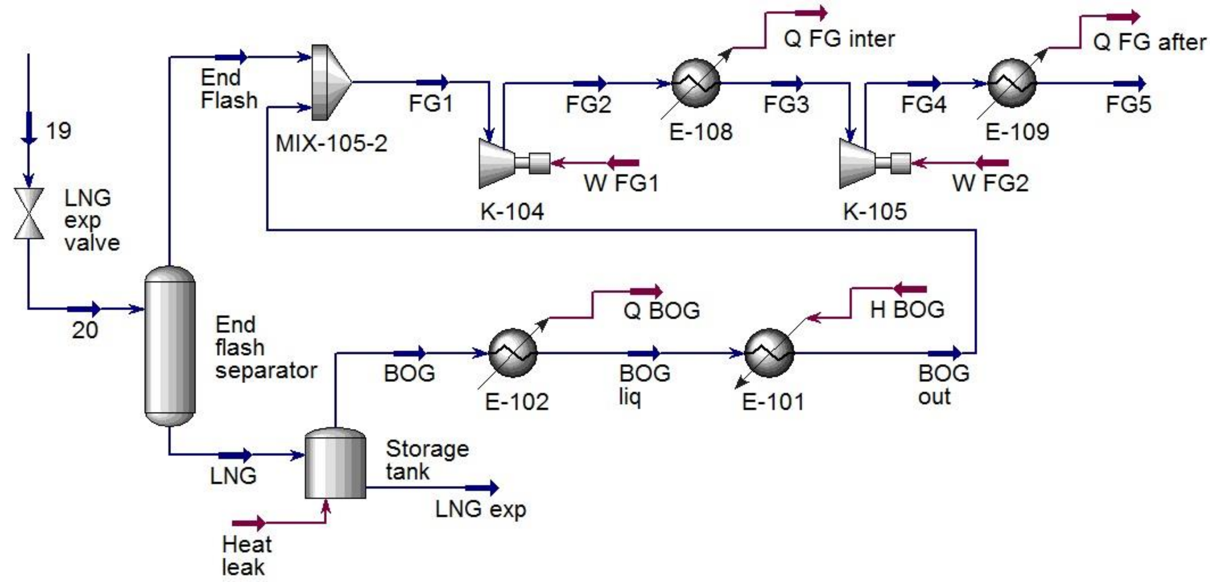

Figure 27: End flash, storage and BOG system with recompression in HYSYS

The LNG expansion has been simulated with a Joule-Thompson valve ("LNG exp valve"). This results in an isenthalpic expansion, which means that the LNG enters the two phase region somewhat further to the right than what it would with a liquid expander, resulting in a higher vapour fraction and thereby a higher end flash flow rate. Additionally, no power is generated in a J-T valve. However, J-T valves is common for FLNG process configurations, due to less complexity and space requirements than a liquid expander (Pettersen, 2015). Thereby, a liquid expander is not modelled in HYSYS.

4.6 Cooling System

As can be seen in the process model overview in Appendix D, all cooling utilities have been modelled with a simple process cooler. The outlet temperature of the processes is defined by the assumed temperature approaches in Table 22. Since all cooling utilities not are included in the process model, the total cooling demand has been summarized and modelled in a separate cooling circuit, as shown in Figure 28. This is to determine the required pumping power and circulation rate for the indirect circuit as well as the seawater/cooling tower circuit. The figure shows the simulation model with cooling tower as heat sink. The energy stream "Q tot" represents the summarized cooling demand, and is calculated by and exported from the spreadsheet "Tot cooling". It is assumed that the temperature of the indirect cooling circuit is increased by 10°C in heat exchanger "Total cooling". The same goes for the cooling tower water in "CT HX".

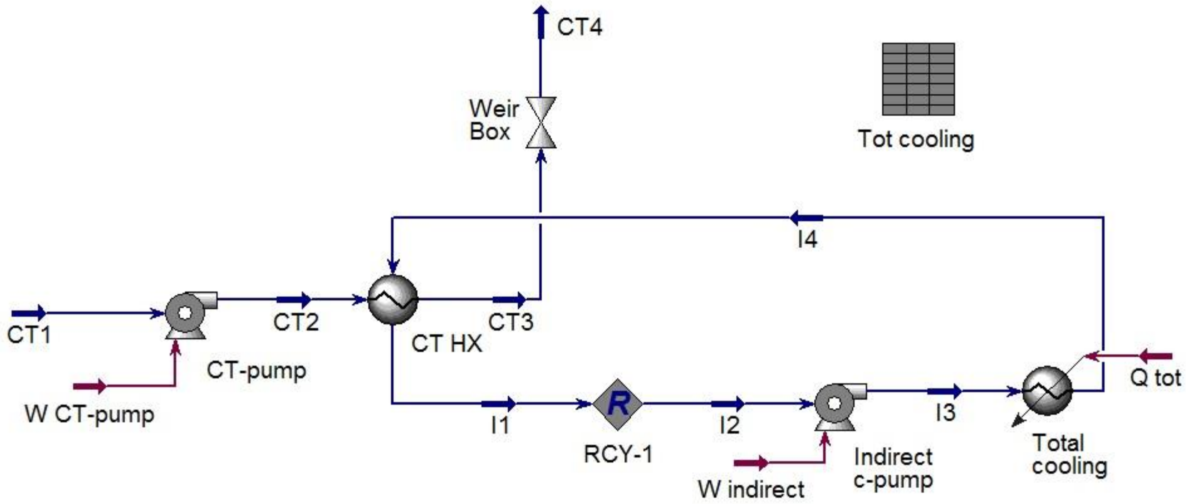

Figure 28: Simplified simulation model of the indirect cooling system

It is assumed that the process coolers in all the indirect cooling circuits are configured in parallel, meaning that the inlet temperature is the same for all the water cooled heat exchangers. This is a source of error, since several heat exchangers might be configured in series to avoid a very complex system, but has been simplified in the simulation performed in this study.

The simulation model is identical for seawater cooling. The air cooler configuration is modelled with a cooler directly connected to the indirect cooling circuit, as shown in Figure 29.

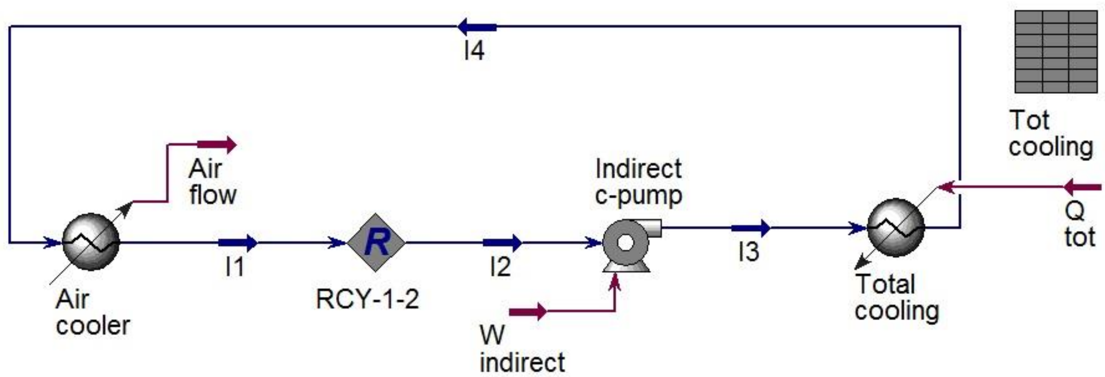

Figure 29: Air cooler system in HYSYS

5 Results and Discussion

In this chapter, the simulations results are presented and the different configuration alternatives are compared and discussed. Then, required power demand, fuel consumption and CO_2-emissions are calculated based on the simulation results.

When comparing the different configuration alternatives for the scenarios, the production rate and specific power for the initial configuration has been set to 100% and the alternative configuration has been measured against this. Furthermore, the result at design temperature for the configuration alternatives is measured against the result for the initial configuration at design temperature. Similarly, the result at high temperature for the configuration alternatives is measured against the result at high temperature for the initial configuration.

Note that for configurations with frontend NGL extraction, the booster compressor duty has been included in the specific power calculation. For each configuration, the available refrigerant compressor power is fixed at the given temperature, while the utility systems capacity such as cooling and heating is assumed to be unlimited, but with a fixed temperature approach.

5.1 Simulation Results for Scenario 1 (Gulf of Mexico)

Table 25 shows the simulation results for the initial configuration of Scenario 1 at design and high temperature. For more detailed simulation results, see Appendix A. Note that the heat demand is scaled linearly with respect to production rate from a lean gas processing and liquefaction model for FLNG provided by the supervisor.

Table 25: Simulation results for the initial configuration in Scenario 1

Property	Unit	Result at Design Temperature	Result at High Temperature
Ambient air temperature	°C	20	36
Total available refrigerant compressor power	MW	173.6	144.0
Annual LNG production rate	MTPA	3.77	2.64
Specific power (efficiency)	kWh/tonne	404.3	473.3
Booster compressor duty	MW	18.43	16.36
Total power demand	MW	228.58	188.12
Total cooling duty	MW	350.37	268.9
Indirect CW flowrate	m^3/h	28540	22230
Total heating duty (scaled)	MW	24.71	17.30
Annual LPG production rate	MTPA	0	0
End flash	mass%	6.50	6.50
LNG Higher Heating Value	MJ/Sm^3	38.49	38.49
LNG Wobbe Index	MJ/m^3	50.98	50.98
Annual condensate production rate	MTPA	0.08	0.06

The LNG production rate drops significantly at high temperature, and corresponds to 70% of design temperature production. Due to the fact that both driver and liquefaction are very sensitive to high ambient air temperatures makes this combination quite ineffective.

As stated earlier in this report, Black and Veatch claims to be able to produce 4.0 MTPA at 30°C with the Trent 60 compressor drive. This is done with integrated NGL extraction, which implies a lower liquefaction pressure, hence a lower efficiency and production. This indicates that the simulated result is quite low. However, the cooling method and number of production days Black and Veatch operates with, is uncertain. Next, Black and Veatch claims that the output from the Trent 60 is 43.6 MW at 30°C, indicating that no derating factors have been included. For the simulation model used in this study with 330 days of operation, 4.0 MTPA can be achieved with 350 production days per year, though at 20°C ambient temperature.

5.1.1 Results for the Alternative Configurations

Table 26 shows the production for the alternative configurations. The percentage value is obtained by comparing the alternative configuration result with the initial configuration result in Table 25. As described earlier, the results at high temperature are compared to the initial configuration result at high temperature.

Table 26: Production rate for the selected alternative configurations in Scenario 1

System	*Initial Configuration*	Alternative Configuration	Production at Design Temperature MTPA (%)	Production at High Temperature MTPA (%)
Liquefaction Process	*PRICO*	Niche	4.03 (106.9%)	2.75 (104.0%)
Driver	*Gas Turbine*	Electric Motor	3.77 (100%)	3.22 (122.0%)
Gas composition	*Lean*	Rich	3.82 (101.4%)	2.56 (97.0%)
Cooling	*Cooling Tower*	Air Coolers	3.63 (96.3%)	2.50 (94.9%)

The variation in production rate for the alternative configurations in Table 26 is further illustrated in Figure 30. The negative effect of gas turbine direct drive becomes clear when compared to electrical drive at high temperature. With a constant power output as for the electric drive, the variation is approximately half compared to the initial. Note that three Niche trains have been assumed in the simulation, resulting in six Trent 60 gas turbines in total. If four gas turbines were used in two Niche trains, the production would be 2.69 MTPA, corresponding to 71.4% of PRICO at design temperature.

With the exception of Niche and electric drive, the production for the other alternatives lies in the range of approximately 5% above or below the initial configuration. The electric drive reduces the large drop in production at high temperature, resulting in a 22% higher production than the initial configuration with GT drive.

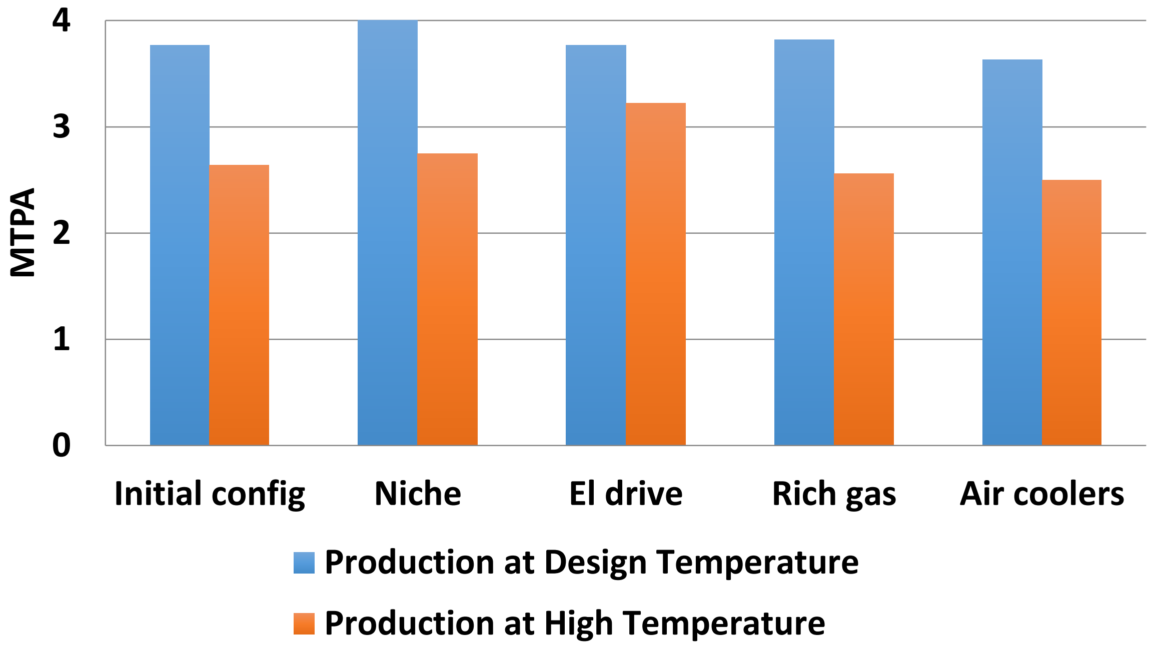

Figure 30: Production rate for the alternative configurations compared to initial configuration for Scenario 1

Table 27 shows the specific power for the alternative configuration. This is compared by percentage of the initial configuration result, which are 404.3 kWh/tonne and 473.3 kWh/tonne for design and high temperature, respectively. The results are further illustrated in Figure 31.

The results show that the type of liquefaction process and cooling method mainly affects the specific power. The configuration with cooling tower perform 6% better than air coolers for the given location at design temperature due to the approach on WB temperature.

Table 27: Specific power for the selected alternative configurations in Scenario 1

System	***Initial Configuration***	**Alternative Configuration**	**Specific power at design temperature [kWh/tonne]**	**Specific power at high temperature [kWh/tonne]**
Liquefaction Process	*PRICO*	Niche	539.4 (133.4%)	625.1 (132.1%)
Driver	*Gas Turbine*	Electric Motor	404.3 (100%)	471.9 (99.7%)
Gas composition	*Lean*	Rich	400.2 (99.0%)	483.5 (102%)
Cooling	*Cooling Tower*	Air Coolers	427.9 (106%)	497.7 (105%)

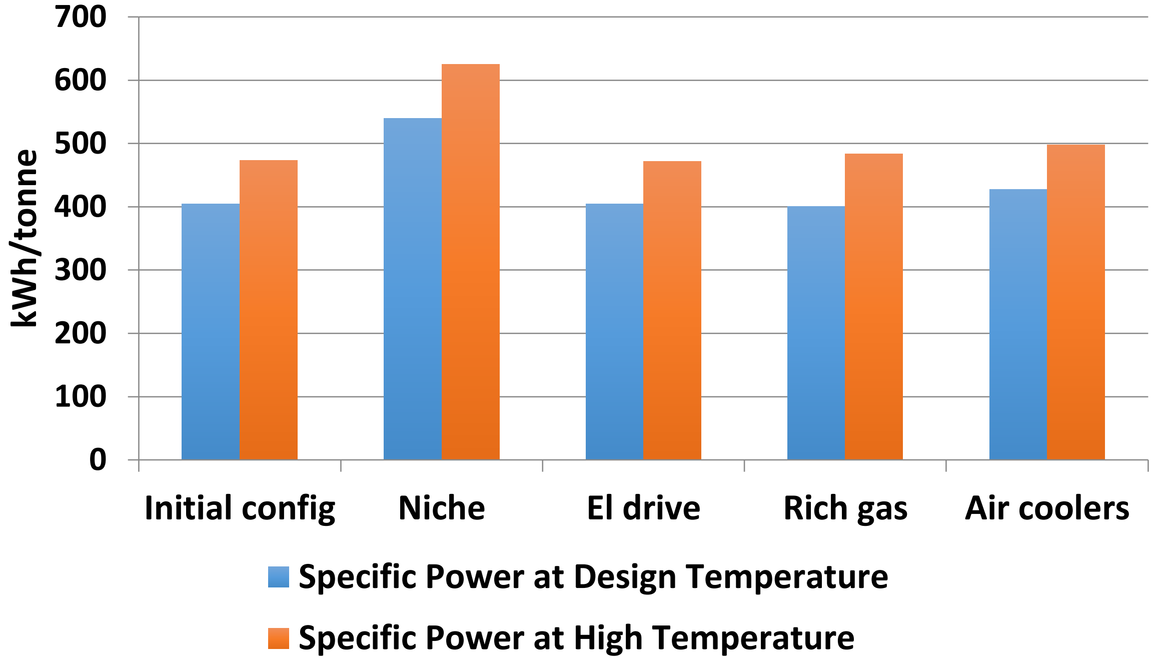

Figure 31: Specific power for the alternative configurations compared to initial configuration in Scenario 1

5.1.2 Consequences of Rich Gas in Scenario 1

As can be seen in Table 26 and Table 27, the simulation results for production rate and specific power are almost identical when the feed gas composition is varied from lean to rich. This is because the rich gas that enters the liquefaction system is extensively pretreated, making the composition very similar to the lean gas. Especially CO_2 removal, NGL extraction and fractionation demands more power, cooling and heat, which are not reflected in these results. A more thorough analysis of the gas processing systems for the rich gas configuration is presented in Figure 32 and Figure 33. The detailed results are given in Appendix I.

For detailed results for power, cooling and heat demand for initial configuration with lean feed gas, see Appendix F, G and H. Note that the cooling and heating duty for several of the pre-treatment consumers are provided by the supervisor and is based on numbers from Melkøya LNG plant and scaled linearly to match the production rate in Scenario 1.

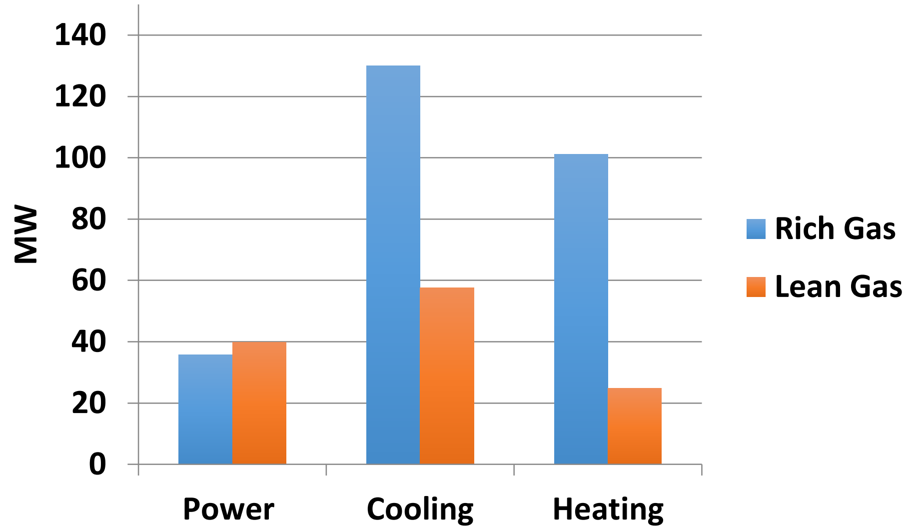

Figure 32: Gas processing power, cooling and heating duties for lean and rich gas in Scenario 1

As shown in Figure 32 the heating and cooling duties for the gas processing systems upstream of the liquefaction are higher for rich gas than for lean, mainly caused by the CO_2 removal system. The power demand is slightly lower for the rich gas, mainly due to the assumption that the rich gas is exploited from an offshore reservoir and no inlet compressor is needed. Figure 33 shows the difference in LPG and condensate production rate and indicates that a significantly more advanced gas processing system is required if the feed gas is rich.

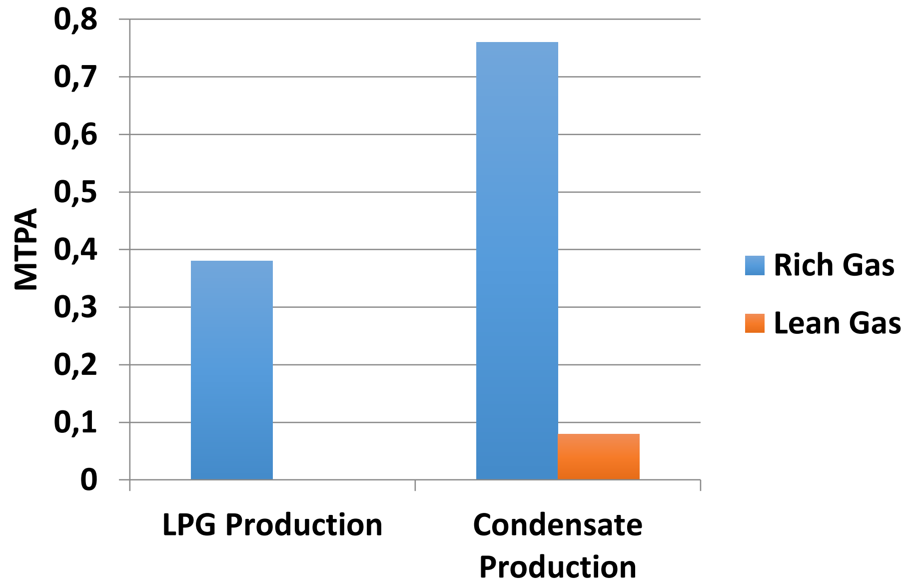

Figure 33: LPG and condensate production for lean and rich feed gas in Scenario 1

5.1.3 NGL Extraction Simulation Results

Table 28 shows for the NGL extraction results for the most important components. The complete composition before and after NGL extraction is shown in Appendix J. As shown, all critical components are within critical range. All C2, C3 and C4 are reinjected after fractionation to meet the market requirements for the LNG product.

Table 28: NGL extraction simulation results for Scenario 1

Component	Unit	Pre-liquefaction requirement	Upstream	Downstream
C1	mol%	-	97.196	97.401
C2	mol%	-	1.240	1.242
C3	mol%	-	0.440	0.441
C4	mol%	< 2	0.190	0.191
C5+	mol%	<0.1	0.182	0.000
BZ	ppm	<1	31.89	0.01

5.2 Simulation Results Scenario 2 (Northern Norway)

Table 29 shows the simulation results for the initial configuration of Scenario 2 at design and high temperature. For more detailed simulation results, see Appendix A.

Note that to obtain total cooling and heating duty, several numbers are scaled linearly from Melkøya with respect to production rate and added to give a more realistic and comparable result. This include processes such as CO_2 removal, MEG treatment, fractionation and condensate stabilization that have not been simulated in HYSYS. A CO_2 reinjection system has also been included since this is regarded to be the only option for CO_2 handling at this location. Detailed calculations of power, cooling and heating duties are given in Appendix F, G and H, respectively.

Table 29: Simulation results for the initial configuration in Scenario 2

Property	Unit	Result at design temperature	Result at high temperature
Ambient air temperature	°C	1	23
Seawater temperature	°C	6	9
Total available refrigerant compressor power	MW	206.0	206.0
Annual LNG production rate	MTPA	5.51	5.41
Specific power (efficiency)	kWh/tonne	335.2	342.7
Booster compressor duty	MW	23.17	22.98
Total power demand	MW	282.51	282.37
Total cooling duty	MW	583.96	576.8
Indirect CW flowrate	m^3/h	47350	46950
Total heating duty (scaled)	MW	164.4	161.4
End flash	mass%	9.03	9.03
Nitrogen Content in LNG	mol%	0.97	0.97
LNG Higher Heating Value	MJ/Sm^3	39.57	39.60
LNG Wobbe Index	MJ/m^3	51.27	51.29
Annual LPG production rate	MTPA	0.57	0.57
Annual condensate production rate	MTPA	0.98	0.96

As can be seen from the results in Table 29, the LNG production rate is high when the system operates at design temperature. Note that this is only based on available power and cooling, and does not reflect the increased size needed for the process equipment. Further comments on this aspect are given in Section 6.2.

At high temperature, the plant is still able to maintain a production of 98.2% of the design production due to the constant power output for the compressor drive and small increase in seawater temperature. The results for the initial configuration illustrates the potential of electric drive combined with seawater cooling with respect to production rate and efficiency, provided that sufficient electrical power is available. The low variation will also ease the operation of the plant.

The high production rate at design temperature indicates that 4.13 MTPA can be reached with only three PRICO trains with one 51.7 MW electric motor driving each train as long as the rest of the liquefaction module is designed for the increased flow. This might free vessel deck space where other process systems can be placed. At high temperature, the production rate will then be slightly reduced to 4.06 MTPA due to a higher cooling water temperature.

The rich feed gas has a nitrogen content of 2.53 mol% and the end flash flow is quite high (8.95 mass% of the total flow). Part of this flow is supplied to the heat generation, which is calculated in Section 5.4.3. The rest should have been reinjected into the natural gas flow after N_2 removal, which would result in a higher LNG production, but this has not been the mains focus of this study. Additionally, the nitrogen content in the LNG product is 0.97 mol%, just below the maximum concentration of 1.0 mol%, indicating that a more advanced end flash system might be required for the rich gas.

The higher heating value of 39.57 MJ/Sm^3 is obtained when all C2 is reinjected into the natural gas before liquefaction. None of the C3 and C4 extracted in the NGL process is reinjected after fractionation, and it is chosen to produce LPG from these hydrocarbons instead. This gives the rich gas scenario some flexibility in terms of HHV. If all C3 and C4 are reinjected into the natural gas, the higher heating value will increase to 41.40 MJ/Sm^3, but no LPG will then be produced.

5.2.1 Results for the Alternative Configurations

Table 30 shows the production for the alternative configurations. The consequence of a local GTG power generation is discussed in Section 5.4.2.

Table 30: Production rate for the selected alternative configurations in Scenario 2

System	***Initial Configuration***	**Alternative Configuration**	**Production at Design Temperature MTPA (%)**	**Production at High Temperature MTPA (%)**
Liquefaction Process	*PRICO*	Niche	3.15 (57.2%)	3.15 (58.2)
Driver	*Electric motor*	Gas turbine	5.51 (100%)	4.21 (77.8%)
Gas Composition	*Rich*	Lean	5.32 (96.6%)	5.29 (97.7%)
NGL Extraction	*Frontend*	Integrated	5.14 (93.3%)	5.02 (92.8%)
Cooling	*Seawater*	Air Coolers	5.51 (100%)	4.31 (79.7%)

As shown in Table 30, the production rate with gas turbine drops to 77.8% at high temperature. If a 3-train configuration is used, a production rate of 4.13 MTPA can be achieved at design temperature, but the production drops to 3.16 MTPA at high temperature. The variation in production rate for the alternative configurations in Table 30 is further illustrated in Figure 34.

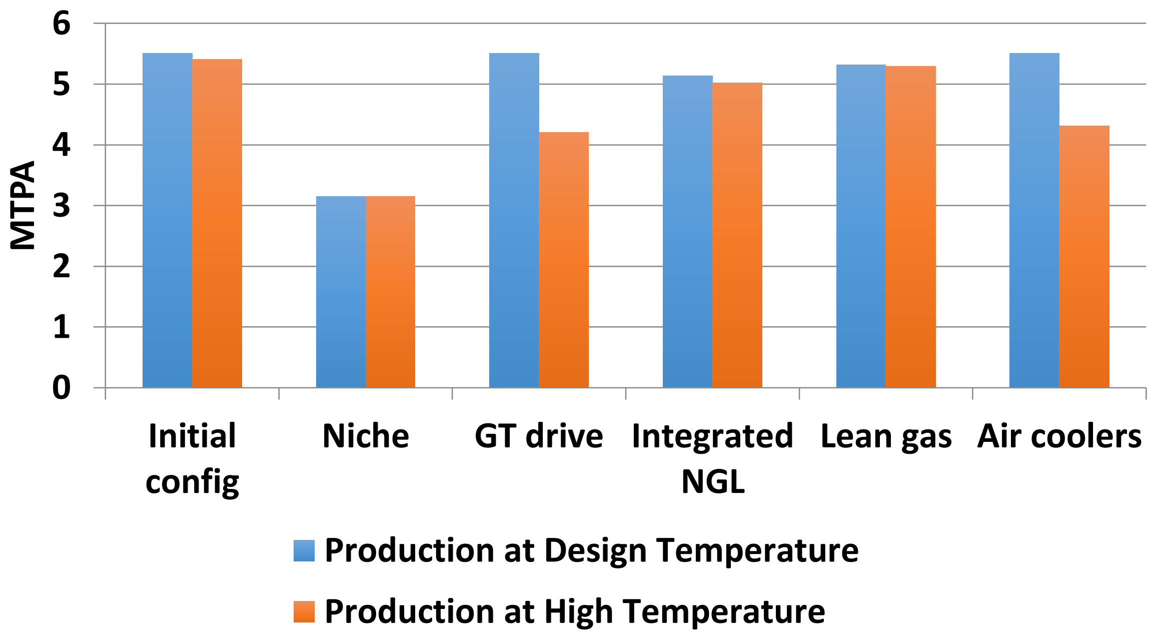

Figure 34: Production rate for initial and alternative configurations in Scenario 2

For the alternative with Niche liquefaction process, the power output from the electrical motors has not been fixed, but been allowed to vary in the simulations. This is done to obtain the optimal efficiency for the liquefaction process under these conditions. A maximum expander output of 15 MW combined with an electrical motor with the same high output as a Trent 60 gas turbine at the given ambient temperature would simply be too much available power, resulting in a very inefficient process. The two required electric motor outputs for the Niche process is presented in Table 31. Note that one motor drives the NG refrigeration circuit and one motor drives the N_2 refrigeration circuit. If gas turbine drivers were to be used on Niche, the results indicates that GE LM6000 PF would be better suited than Trent 60 due to the lower output of LM6000.

Table 31: Required electrical power for Niche liquefaction process in Scenario 2

Driver	Design Temperature	High Temperature
El motor, NG circuit [MW]	22.90	23.47
El motor, N_2 circuit [MW]	32.06	34.71

The specific power for different alternatives is listed in Table 32 and further illustrated in Figure 35. This is compared with the initial result of 335.2 kWh/tonne and 342.7 kWh/tonne for average and design temperature, respectively. As can be seen, the Niche process performs poorly even though the drivers have been allowed to vary. This is clearly illustrated in Table 32, where specific power is 33% larger than for PRICO. However, the specific power of Niche process is expected to be 30-40% above PRICO (Pettersen, 2015), and the simulations of Niche in this study is within this range.

The results in Table 32 show that the integrated NGL extraction is the most efficient, requiring 4.3% less power to liquefy one tonne LNG. The reason for this is that the frontend NGL extraction has a 24.58 MW booster compressor, which is included in the calculation of specific power. However, the liquefaction process in itself is less efficient for integrated due to a reduced liquefaction pressure (50 bar compared to 59.5 bar for frontend). This is reflected in the production rate, which is 6.0% lower for integrated NGL extraction compared to frontend. Which of the options that is the best solution is discussed further in Chapter 6 when complexity and reliability are taken into account.

As mentioned, the liquefaction pressure for the integrated NGL extraction alternative is reduced to 50 bar to achieve a sufficient NGL extraction. This is because heavy components such as benzene is hard to remove sufficiently at higher pressures with only a two phase separator. This is further discussed in Chapter 5.2.3.

Table 32: Specific power for the selected configuration alternatives in Scenario 2

System	*Initial Configuration*	Alternative Configuration	Specific power at Design Temperature [kWh/tonne]	Specific Power at High Temperature [kWh/tonne]
Liquefaction Process	*PRICO*	Niche	443.9 (133%)	468.1 (136.6%)
Driver	*Electric motor*	Gas turbine	335.2 (100%)	347.8 (101.5%)
Gas composition	*Rich*	Lean	346.6 (103.4%)	353.0 (103.0%)
NGL extraction	*Frontend*	Integrated	320.9 (95.7%)	329.5 (96.2%)
Cooling	*Seawater*	Air Coolers	335.2 (100%)	420.0 (122.6%)

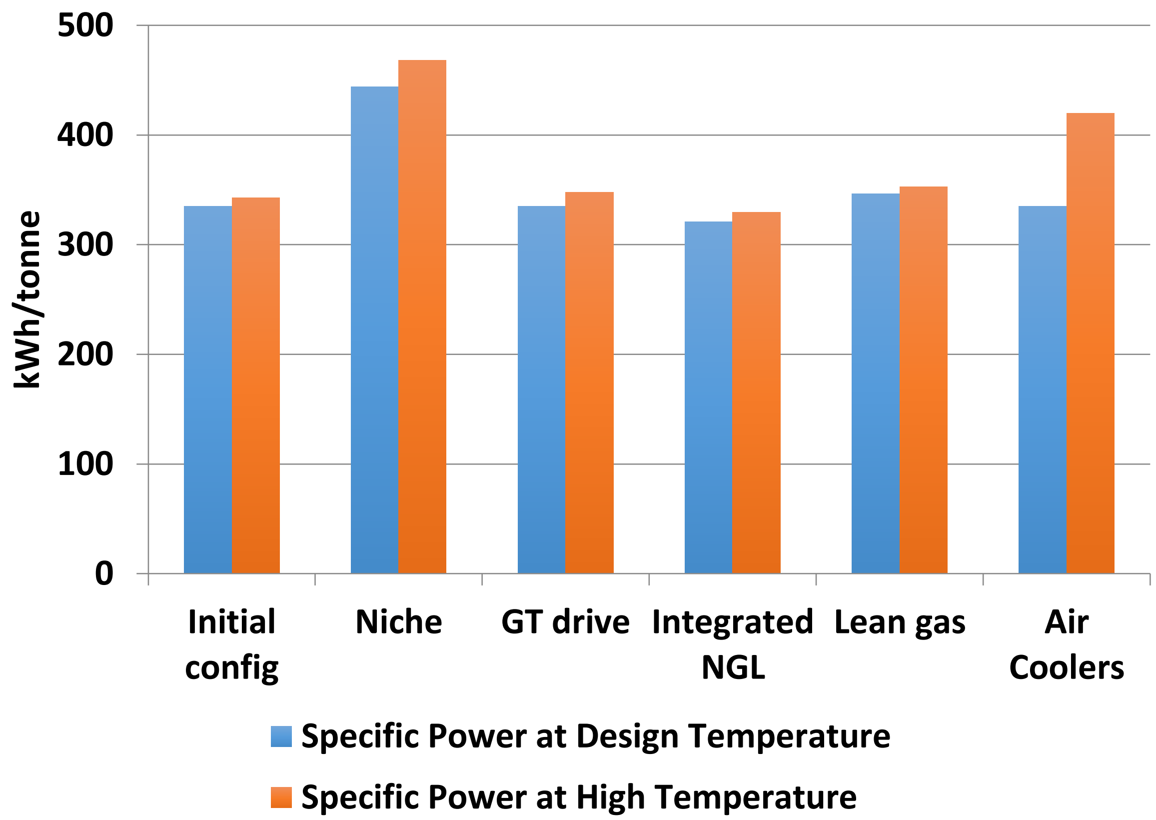

Figure 35: Specific power for alternative and initial configurations in Scenario 2

5.2.2 Consequences of Lean Gas in Scenario 2.

As for Scenario 1, the results for production rate and specific power are almost identical for lean and rich gas composition and do not reflect the variation upstream of the liquefaction process.

Figure 36 illustrates the difference in required power, cooling and heat demand in Scenario 2 with lean and rich gas. Figure 37 illustrates the difference in LPG and condensate production rate. For detailed results of power, cooling and heat demand for the initial configuration with rich feed gas, see Appendix F, G and H. The detailed results for lean gas in Scenario 2 are given in Appendix I. Note that the heating duty for all consumers are scaled linearly based on production at Melkøya and at a lean gas process model for FLNG provided by supervisor.

As expected, the required power, cooling and heating duties are lower if the feed gas is lean. Next, the condensate production will be significantly reduced and there will be no LPG production, as all C3 and C4 is reinjected into the natural gas. This indicates that the gas processing systems are less advanced and extensive for a lean gas scenario.

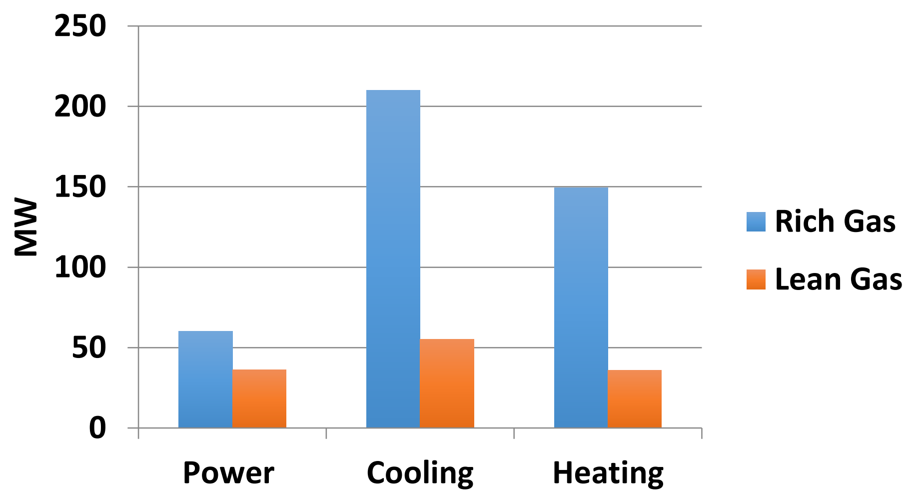

Figure 36: Gas processing power, cooling and heating duties for lean and rich feed gas in Scenario 2

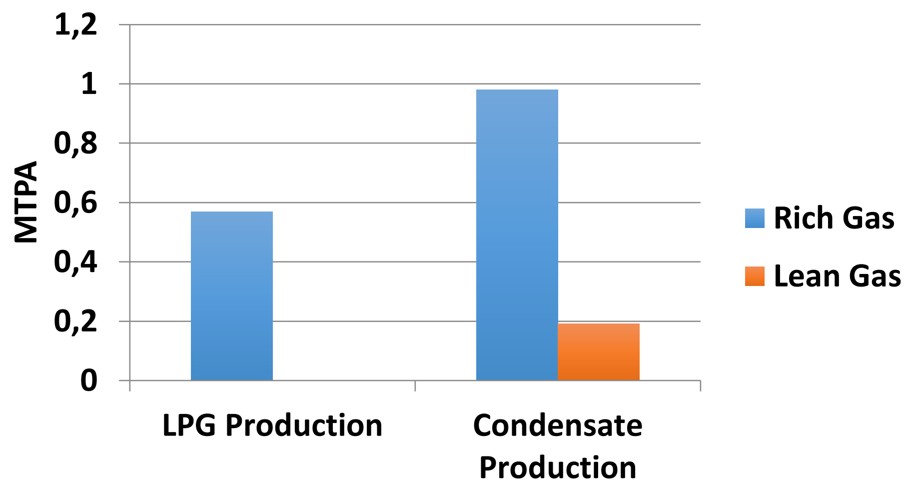

Figure 37: Annual LPG and condensate production in Scenario 2 with lean and rich feed gas

5.2.3 NGL Extraction Results

Table 33 shows the most important results for the integrated and frontend NGL extraction. The complete composition before and after NGL extraction is shown in Appendix J. As shown, all components are within critical range, and frontend offers a higher level of extraction for the HHC and BZ.

As described in Chapter 4.3, the integrated NGL extraction was modelled with a two phase separator instead of a reboiled absorber due to optimization complexity in HYSYS. However, if a reboiled absorber was used instead, the liquefaction process might become more efficient and the extraction results might also have been improved due to better separation. The use of a separator leads to an unrealistically high flow rate of the extracted stream. This flow mainly contains methane, which needs to be reinjected into the natural gas, resulting in an unrealistically high flow back and forth between fractionation and liquefaction. However, this has not studied any further in this thesis and is considered as further work.

Table 33: NGL extraction simulation results for Scenario 2

Component	Unit	Pre-liquefaction Requirement	Upstream	Downstream (frontend)	Downstream (integrated)
C1	mol%	-	87.743	90.890	91.183
C2	mol%	-	5.447	5.643	5.661
C3	mol%	-	2.736	0.590	0.048
C4	mol%	< 2	0.943	0.038	0.063
C5+	mol%	<0.1	0.362	0.002	0.005
BZ	ppm	<1	143.8	0.13	0.98

5.3 Simulation Results for the Subcases

This subchapter contains the most important results obtained from the simulation of the subcases at design and high temperatures. Compared to the scenario simulations, several variables have been changed to fit local conditions and constraints. The configuration of each subcase is presented in Section 3.5.

5.3.1 Simulation Results for Subcase A (British Columbia)

Table 34 shows the simulated results for Subcase A in British Columbia. Note that the heating and cooling duties for fractionation, acid gas removal and condensate stabilization have not been included in the simulations. These duties have been scaled linearly with respect to LNG production from a lean gas process model for FLNG provided by supervisor. As for the scenarios, the scaled results are added to the simulated to obtain the total cooling and heating duty. For detailed results, see Appendix G and H. Detailed simulation results are given in Appendix A.

For Subcase A and B, which uses air cooling, HYSYS has not been able to calculate the required number of fans and fan power. Instead, a sizing calculator from GEA Heat Exchangers has been used (GEA Heat Exchangers, 2010). The fan size is 5.5 meter, each requires 33.2 kW electrical power and the temperature approach between ambient air and cooling water is assumed to be 10°C.

Table 34: Simulation results for Subcase A

Property	Unit	Result at design temperature	Result at high temperature
Ambient air temperature	°C	6.9	28
Total available refrigerant compressor power	MW	198.0	198.0
Annual LNG production rate	MTPA	4.78	3.82
Specific power (efficiency)	kWh/tonne	372.0	452.9
Booster compressor duty	MW	24.22	21.12
Total power demand	MW	277.13	266.25
Heating duty	MW	31.33	25.04
Cooling duty	MW	424.01	376.9
Indirect CW flowrate	m^3/h	34530	30720
Number of fans required	-	441	392
End flash	mass %	6.50	6.50
LNG Higher Heating Value	MJ/Sm^3	38.49	38.49
LNG Wobbe Index	MJ/m^3	50.98	50.98
Annual condensate production rate	MTPA	0.10	0.08

5.3.2 Simulation Results for Subcase B (Northwest Russia)

Table 35 shows the simulated results for Subcase B in Northwest Russia. As for Subcase A, the small cooling duties have not been simulated, but scaled linearly with respect to LNG production from a lean gas process model for FLNG provided by supervisor and added to the simulated duty. See Appendix G for detailed cooling consumers and duties. To calculate the number of fans needed for the cooling system, GEA sizing calculator has been used (GEA Heat Exchangers, 2010).

Table 35: Simulation results for Subcase B

Property	Unit	Result at Design Temperature	Result at High Temperature
Ambient air temperature	°C	-1	28
Total available refrigerant compressor power	MW	206.0	153.6
Annual LNG production rate	MTPA	5.59	2.94
Specific power (efficiency)	kWh/tonne	332.0	457.7
Booster compressor duty	MW	27.09	16.20
Total power demand	MW	268.46	200.27
Heating duty (scaled)	MW	49.24	19.27
Cooling duty	MW	437.51	280.8
Indirect CW flowrate	m^3/h	35590	22890
Number of fans required	-	455	292
End flash	mass %	6.50	6.50
LNG Higher Heating Value	MJ/Sm^3	38.49	38.49
LNG Wobbe Index	MJ/m^3	50.98	50.98
Annual condensate production rate	MTPA	0.14	0.08

Note that the cooling duty is only 3.2% higher for Subcase B than for Subcase A, even though the production is 17% higher. Next, the power required for Subcase A is 3.2% higher than in Subcase B. This is because Subcase A has a large 27 MW cooler after the required inlet compressor. Since Subcase B produce from a reservoir, an inlet compressor and aftercooler is not needed.

5.3.3 Simulation Results for Subcase C (Northwest Australia)

Table 36 shows the simulated results for Subcase C in Northwest Australia. As for the other subcases, the small cooling duties have not been simulated, but scaled linearly with respect to LNG production from a lean gas process model for FLNG provided by supervisor. As expected, the LNG production rate is low, especially at high ambient temperature, due the combination of gas turbine direct drive and air cooling.

Table 36: Simulation results for Subcase C

Property	Unit	Result at Design Temperature	Result at High Temperature
Ambient air temperature	°C	26.5	44
Total available refrigerant compressor power	MW	161.2	137.2
Annual LNG production rate	MTPA	3.55	2.29
Specific power (efficiency)	kWh/tonne	396.7	513.8
Booster compressor duty	MW	16.44	11.37
Total power demand	MW	219.32	187.08
Heating duty (scaled)	MW	94.10	60.70
Cooling duty	MW	406.97	300.1
Indirect CW flowrate	m^3/h	33160	24440
End flash	mass %	8.87	8.80
LNG Higher Heating Value	MJ/Sm^3	40.22	40.51
LNG Wobbe Index	MJ/m^3	51.65	51.81
Annual LPG production rate	MTPA	0.27	0.18
Annual condensate production rate	MTPA	0.60	0.45

5.3.4 Comments on Air Cooled Heat Exchangers in Subcase A and B

The number of fans in Subcase A and B depends largely on the assumed temperature approach of 10°C. Figure 38 and 39 shows the required number of fans with varying temperature approach for Subcase A and B with the respective required cooling duty at design and high temperature. The calculation is performed by GEA sizing calculator (GEA Heat Exchangers, 2010). As expected, the required number of fans grows exponentially when the temperature approach goes towards zero. The graphs show that even a 10°C approach might be a bit radical, but this has not been studied any further in this thesis.

The low cooling duty at high temperature leads to a lower number of fans required. This number drops from 441 to 392 in Subcase A and from 455 to 292 in Subcase B. However, if all fans were used at high temperature, this would result in a better temperature approach. For Subcase A, the temperature approach would be reduced to 8.33°C, meaning that all processes in the plant could be cooled an additional 1.67°C. For Subcase B the gain is even greater as the temperature approach is reduced to 4.95°C. However, this would result in lower specific power, increase the production and demand more cooling, which would increase the required number of fans. Iteration on the potential production with the given number of air coolers have not been performed but the result in this subchapter indicates a potential for higher production than simulated at high temperature. Increasing the number of fans is also a relative inexpensive way to increase the production at high temperature.

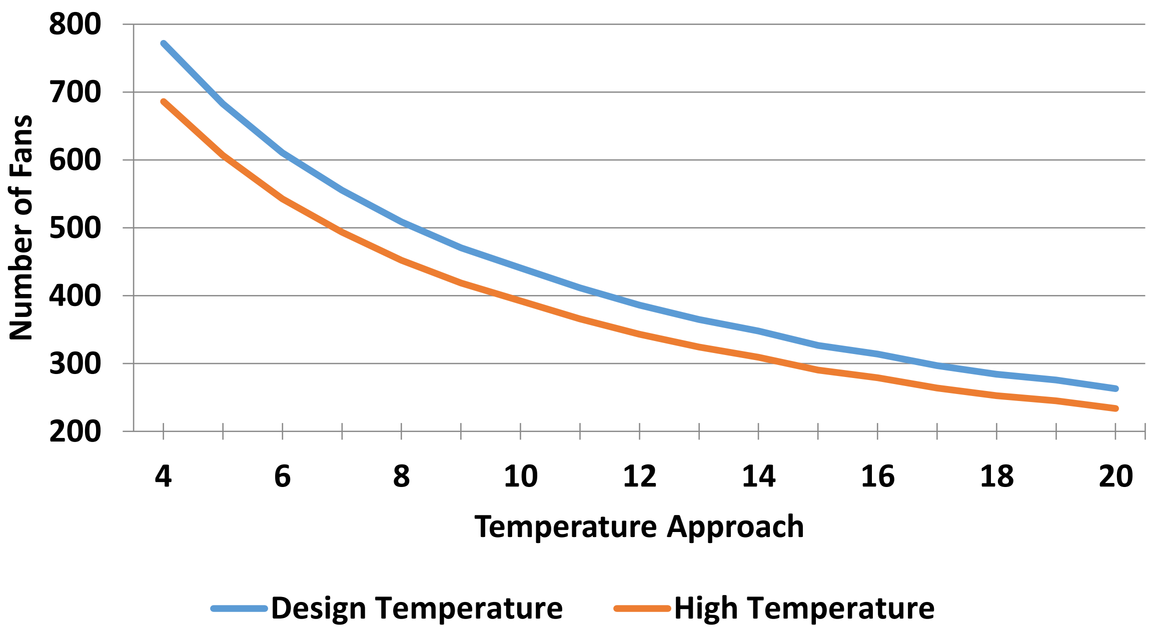

Figure 38: Required number of fans for Subcase A with varying temperature approach

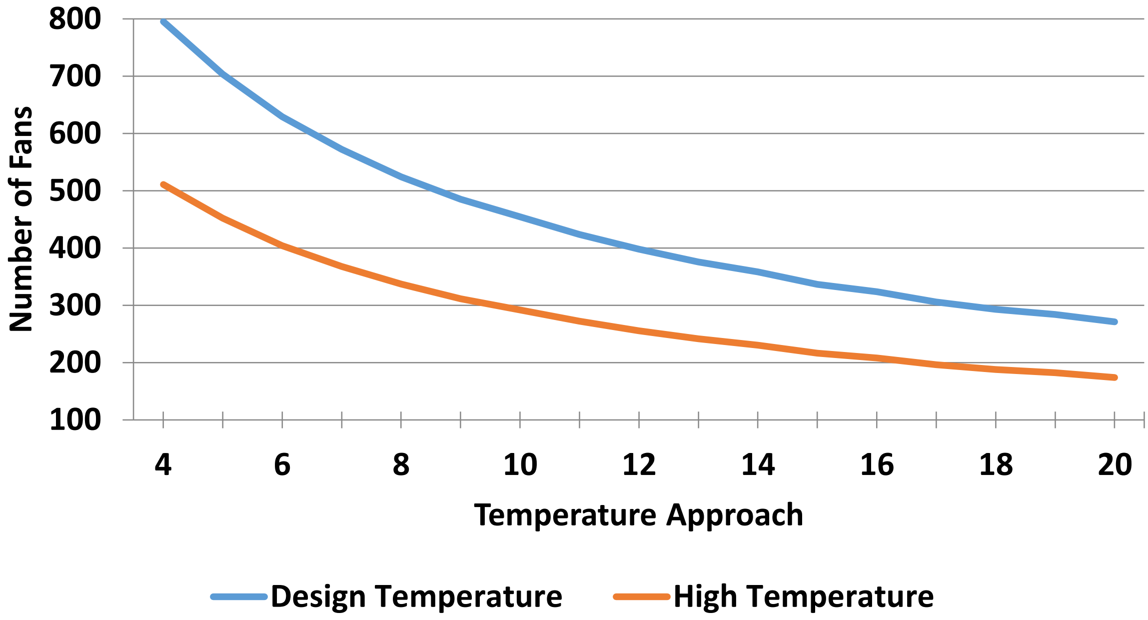

Figure 39: Required number of fans for Subcase B with varying temperature approach

5.3.5 Summary of Subcases

Figure 40 shows the production at design and high temperature for all subcases. The specific power is illustrated in Figure 41. As shown, Subcase B in Northwest Russia experience the greatest variation in production, dropping 47% at high temperature, which indicates the poor performance for GT driver, combined with air coolers. Subcase A in British Columbia experiences the lowest variation since the compressors are electrical driven, where the production is reduced 20% at high temperature. If GT drivers were used instead, the production at high temperature would be the same as for Subcase B as these two subcases have the same high temperature of 28°C and air coolers. GT drivers would then result in a production decrease of 38.5% for Subcase A. This clearly shows the advantage of electrical drive in terms of stable and high production.

Subcase C production is reduced 35.5% at high temperature and experiences the lowest production for all scenarios and subcases. Although gas turbine drivers and an air dependent cooling system are used, the reduction is lower than for Subcase B. This is mainly due to the use of cooling tower and a smaller temperature difference between design and high (17.5°C for Subcase C and 29°C for Subcase B). The air in Subcase C is very dry, making the cooling tower more efficient compared to air coolers. This is further discussed in Section 5.6.1.

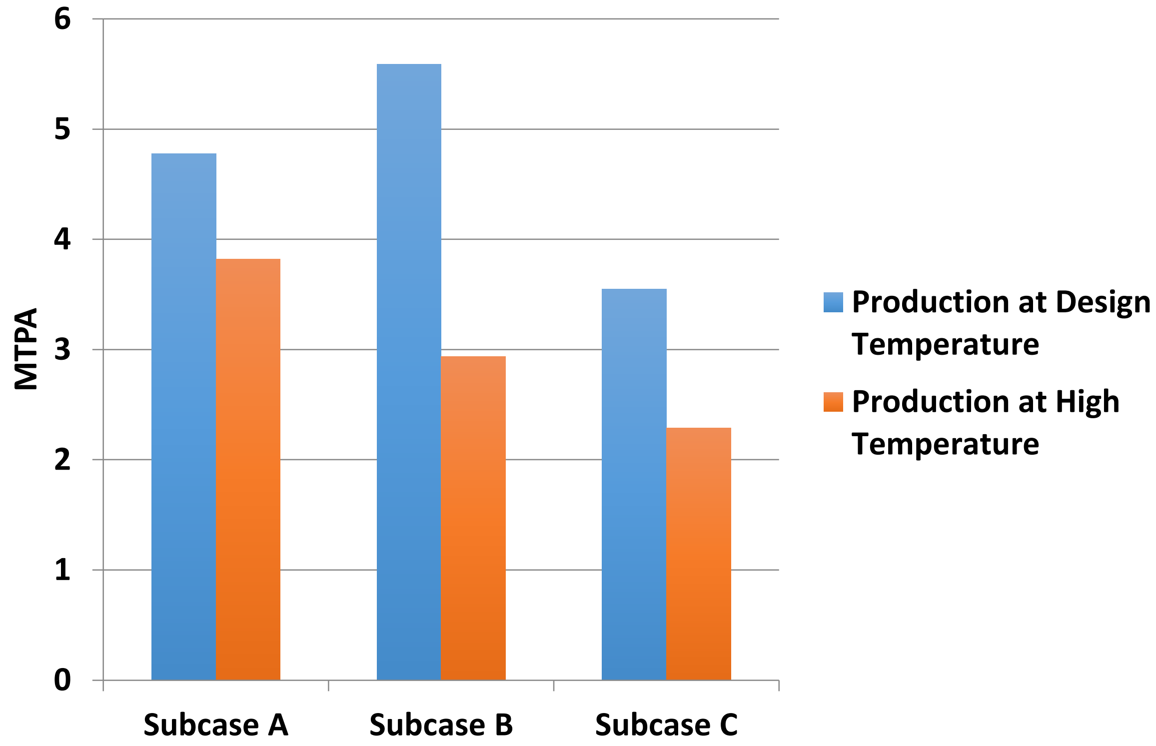

Figure 40: Production for the subcases at design and high temperature

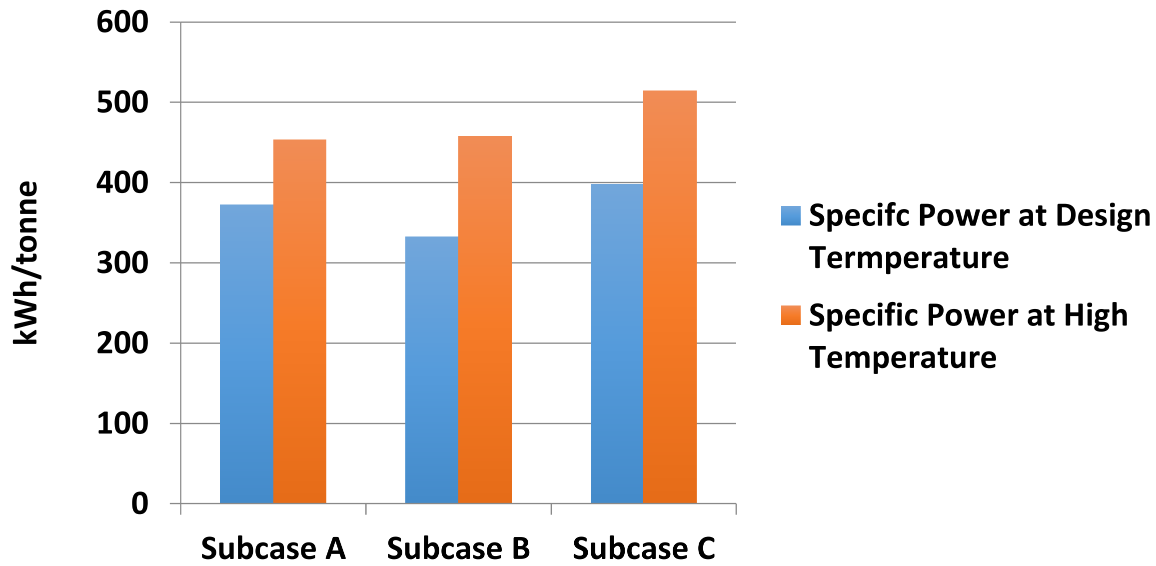

Figure 41: Specific power for all subcases at design and high temperature

5.4 Total Power Demand and Fuel Gas Consumption

In this subchapter, the required electrical power, required number of gas turbines and fuel gas consumption are calculated for each scenario and subcase. Finally, the fuel gas consumption is measured against the feed gas flow.

5.4.1 Required Power

Table 37 shows the total electrical power required for the initial configurations at design temperature. A more detailed table is given in Appendix F. For the lean cases, 10 MW has been added to include the consumers that have been excluded in the simulation. This covers loading pumps, fractionation, lighting etc. For the rich gas cases, 15 MW has been added since the fractionation, LPG and condensate system requires more. Note that the rich gas locations in Scenario 2 and Subcase C have a CO_2 reinjection system, where CO_2 removed from the process is reinjected into a reservoir. The power demand for this system is based on numbers from the Melkøya plant (Bjørge, 2014). The power demand is scaled linearly with respect to production rate.

Note that Scenario 1 and Subcase A, which produce LNG from pipelines, have an inlet booster compressor since the inlet pressure is assumed to be 40 bar. The booster compressor is assumed to be electrical driven for both Scenario 1 and Subcase A. The booster compressor after the frontend NGL extraction, however, is assumed to be electrical driven only if electrical motors are used for the liquefaction drivers. Finally, Subcase A and B have air cooled heat exchangers, where the fans requires a total of 14.64 MW and 14.92 MW, respectively.

Table 37: Total required electrical power for the scenarios and subcases at design temperature

Location	Required electric power [MW]
Scenario 1	36.57
Scenario 2	282.51
Subcase A	277.13
Subcase B	35.23
Subcase C	41.59

5.4.2 Number of Gas Turbines in the Power Plant

The number of gas turbines required for the power generation is presented in Table 38. This is based on the required electrical power in Table 37 and the performance of a GTG at the given location. Note that a N+1 configuration is assumed with one GTG in backup at design temperatures. This is done to improve the reliability and to prevent the GTG power plant from becoming the bottleneck at days with high air temperature.

The claimed power output and efficiency for the Trent 60 at the respective locations have been used to calculate the number of gas turbines required. Next, the derating factors defined in Section 2.3.1 have been included to calculate the actual power output. For calculation of the actual electrical power output, a generator efficiency of 98% has been assumed.

Table 38: Number of GTG for power generation in Scenario 1, Subcase B and C

Property	Scenario 1	Subcase B	Subcase C
GT efficiency at design temperature	41.8% (20°C)	43.5% (2.5°C)	41.0% (26.5°C)
GT claimed power output [MW]	50.0	60.0	46.5
GT actual power output [MW]	43.4	51.5	40.3
Generator efficiency	0.98	0.98	0.98
GTG electrical power output [MW]	42.5	50.5	39.5
Number of GTG in operation	1 (86% load)	1 (70% load)	2 (100% +5% load)
Total number of GTG in the power plant	2	2	3

As shown in Table 38, Scenario 1 and Subcase B can operate satisfactory with one GTG in operation at design temperature, resulting in two GTG in the power plant when redundancy and high temperature operation are taken into account. Subcase C needs slightly more power than one GTG can supply, resulting in 3 in total. Note that the numbers for required electrical power should be used as an indication rather than exact numbers and the actual requirements might need one GTG in operation instead of 1.05.

The alternative with a GTG power plant in Scenario 2 has been included to illustrate some issues when electrical drive is combined with local power generation. The electric motors require constant power input despite temperature, while the gas turbine output declines rapidly at increasing temperature. This is illustrated in Figure 42 with a GTG power plant with 6 and 7 gas turbines. Derating factors and anti-icing are included and a generator efficiency of 0.98 is assumed.

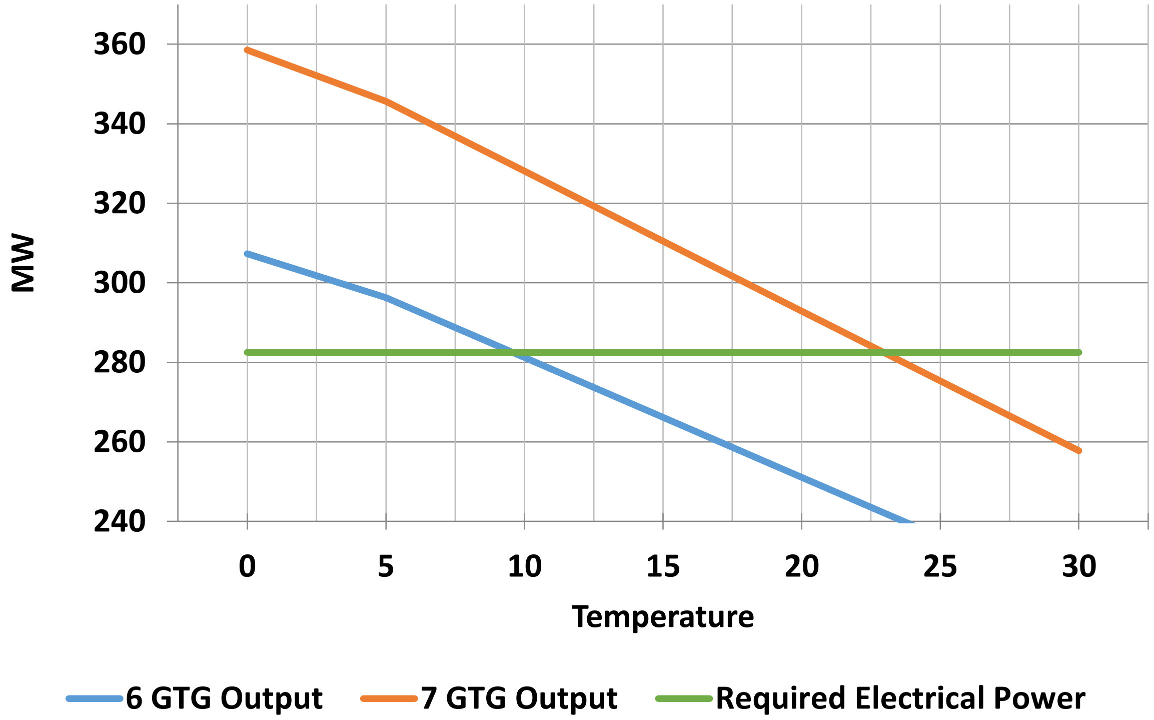

Figure 42: 6 and 7 GTG power output with increasing temperature compared with required electrical power

At design temperature at 1°C, 5.56 gas turbines are required, for instance five turbines on 100% load and one on 56% load or more realistically, 6 gas turbines, each running on roughly 93% load. As showed in Figure 42, a 6 GTG power plant is able to deliver sufficient power up to 9.5°C. Studies of the maximum air temperature in Hammerfest from June 2014 to May 2015 shows that this location will experience 132 days with maximum temperatures above 9.5°C (NRK/NMI, 2015), thus reducing the production rate. A 7 GTG power plant can supply sufficient power up to 23°C and numbers from the same period shows that only 5 days experience temperatures above this.

For this scenario alternative, a N+1 configuration with a total number of 7 GTG seems favourable. The plant can then operate with redundancy 64% of the year. If scheduled maintenance is done during the warm part of the year, the redundancy number will be higher.

5.4.3 Fuel Gas Consumption

To get an indication of the fuel gas consumption for all scenarios and subcases, the gas turbine drivers and power generation has been evaluated up against the required electrical power. The GT compressor drivers have been included where it is used. The claimed power output is divided by the efficiency to calculate the required fuel energy flow per GT or GTG. The calculation results are shown in Table 39.

Scenario 1, Subcase B and C uses a GT driven booster compressor in the NGL extraction. For the simplicity of this calculation, it is assumed that this has the same derating factor and efficiency as the Trent 60 at the given location.

Finally, the available energy stream from end flash and BOG are obtained to identify how much of the required fuel energy they cover. The BOG rate is assumed to be 0.15% of the storage tank volume per day and the end flash flow rate and higher heating value is obtained from the simulations. For detailed calculations, see Appendix B.

Table 39: Calculation of fuel gas consumption for Scenario 1, Subcase B and C

Property	Unit	Scenario 1	Subcase B	Subcase C
GT efficiency at design temperature	-	41.8% (20°C)	43.5% (2.5°C)	41.0% (26.5°C)
GT claimed power output	MW	50.0	60.0	46.5
Required fuel energy per GT	MJ/s	119.6	137.9	113.4
GT's for compressor drive	-	4	4	4
GT's for power generation (in operation)	-	1 (86% load)	1 (70% load)	2 (100% + 5% load)
GT driven booster (in NGL extraction) required fuel energy	MJ/s	50.85	72.92	46.25
Total required fuel energy	MJ/s	**631.59**	**721.05**	**618.92**
End flash energy flow	MJ/s	406.62	619.16	396.02
BOG energy flow	MJ/s	92.67	92.67	81.82
Total available energy flow from flash and BOG	MJ/s	**499.29**	**711.83**	**477.84**
Shortage energy flow	**MJ/s**	**132.3**	**9.22**	**141.08**

Scenario 2 and Subcase A must generate the required heat in a burner. The fuel gas consumption is shown in Table 40. The efficiency of the heater is assumed to be 90%.

Table 40: Calculation of fuel gas consumption for Scenario 2 and Subcase A

Property	unit	Scenario 2	Subcase A
Required Heat	MW	164.40	31.33
Burner efficiency	-	0.90	0.90
Required fuel energy	MJ/s	182.7	34.8
End flash energy flow	MJ/s	620.01	514.92
BOG energy flow	MJ/s	81.79	92.67
Total available energy	MJ/s	701.80	607.59
Excess energy flow	MJ/s	519.1	572.79

The required and available fuel energy from end flash and BOG are illustrated in Figure 43 for all scenarios and subcases. As shown, Scenario 2 in Norway and Subcase A in BC have a major surplus of available energy since fuel is only required in the heat generation. The surplus of BOG and end flash will need to be reinjected before the liquefaction process. In this study, only recompression of BOG and end flash is included. Scenario 1 in GoM and Subcase C in Australia needs more energy than the BOG and end flash can supply, which means that fuel gas needs to be taken from elsewhere in the process. Subcase B in Northern Russia needs slightly more energy than available, which is assumed to be achieved with minor adjustments in the flash system.

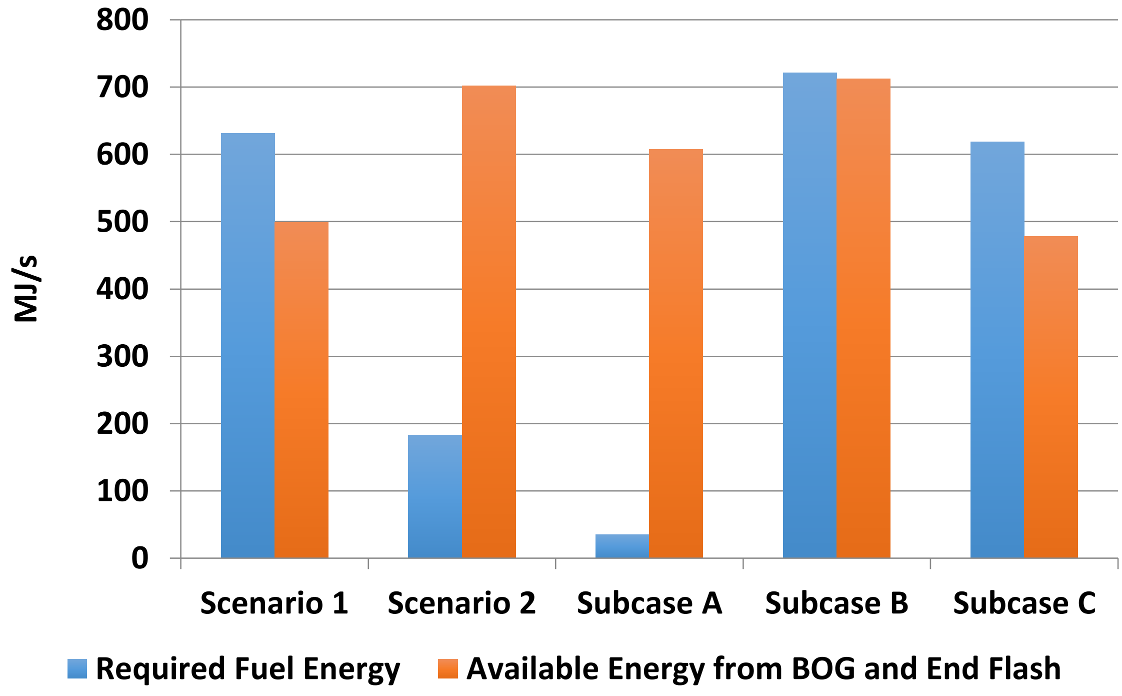

Figure 43: Required and available fuel energy for all scenario and subcases

For further comparison, the fuel gas consumption has been measured against the total feed gas flow for all scenario and subcases at design temperature. The shortage of energy in Scenario 1, Subcase B and C is assumed to be covered by gas with LNG product specifications. The results are given in Table 41 and detailed calculations are presented in Appendix B.

Table 41: Gas consumed for all scenario and subcases

Property	Unit	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
BOG and end flash flow rate	kg/s	11.23	21.28	13.68	15.66	14.41
Fraction of flash and BOG for fuel	-	100%	29.47%	5.73%	100%	100%
Required additional gas flow for fuel	kg/s	2.665	-	-	0.186	2.888
Total fuel gas flow	kg/s	13.895	6.27	0.78	15.846	17.298
Total feed gas flow	kg/s	145.8	295.3	185.8	216.4	187.2
Required energy flow	MJ/s	631.59	182.70	34.80	721.05	618.92
Feed gas energy flow	MJ/s	7131	12382	9087	10584	7849
Total gas consumed on mass basis	-	9.53%	2.12%	0.004%	7.32%	9.24%
Total gas consumed on energy basis	-	8.86%	1.48%	0.004%	6.81%	7.89%

5.5 CO_2 Emissions

To estimate the CO_2 emissions from scenarios and subcases, a simple combustion calculation has been performed on the gas turbines and burner. The calculation is done by equation 5.1, where α, b, c and d are constants and complete combustion is assumed. The fuel properties are based on the properties of methane where the LHV and combustion is only affected by methane since nitrogen is inert. Note that these calculations are based on kmol instead of kg and the emitted CO_2 per year is based on 330 days of production.

$$CH_4 + \alpha(O_2 + 3.76N_2) \rightarrow aH_2O + bCO_2 + cN_2 \qquad 5.1$$

For Scenario 1, Subcase A and Subcase B, the CO_2 removed from the feed gas is assumed to be vented out into the atmosphere. For Scenario 2 and Subcase C, the removed CO_2 is assumed to be reinjected into a reservoir.

For Scenario 2 and Subcase A, CO_2 is emitted when the gas fired burner is used to provide heat for the facility. The simple calculation for this is the same as for the gas turbines, but with slightly different fuel gas properties. A heat loss will occur in the pipes and processes, but is assumed to be zero for the calculations. For Scenario 1 and Subcase B and C, heat is assumed to be supplied by a waste heat recovery system on the gas turbines.

Figure 44 shows the total CO_2 emissions for all scenarios and subcases. For detailed calculations, see Appendix K. As expected, Scenario 1, Subcase B and C emits significantly more CO_2 than Scenario 2 and Subcase A since they have GT drivers and power generation. If the external power generation in Scenario 2 and Subcase A is generated from renewable energy, the total emitted CO_2 locally will be as shown. Note that these numbers does not include CO_2 equivalents for the renewables. However, if the energy is generated from fossil fuels, the total emitted CO_2 will be significantly more and might exceed Scenario 1 and Subcase B and C.

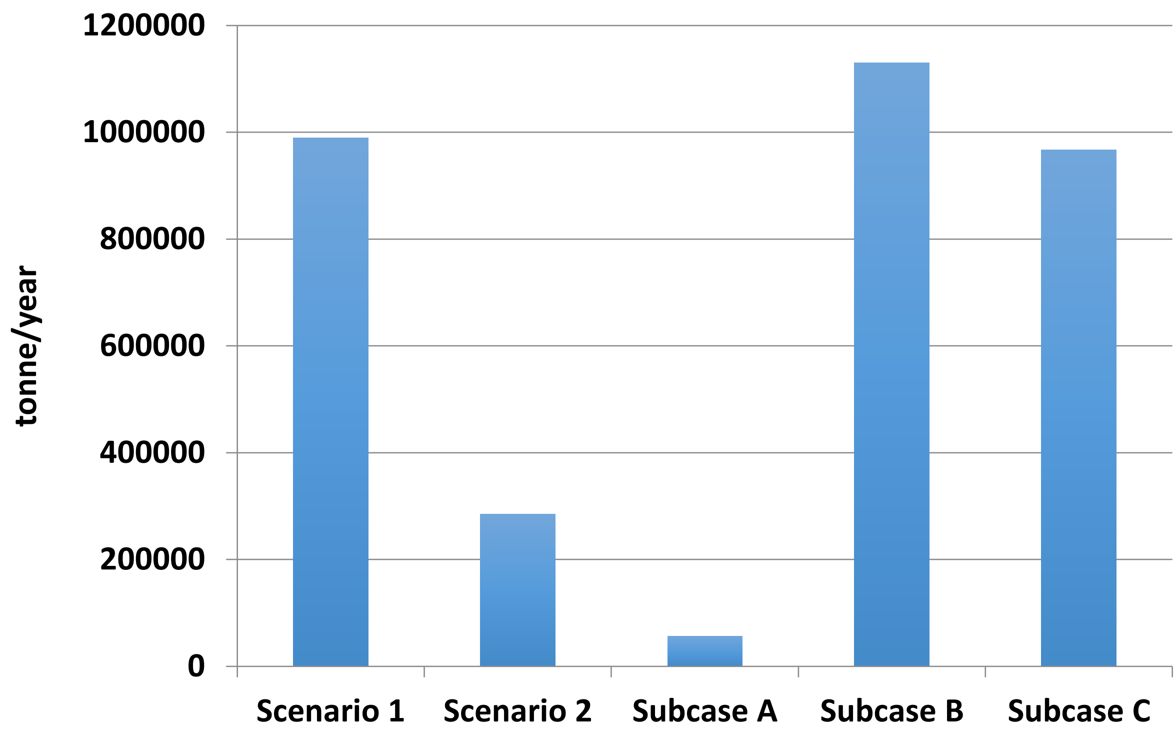

Figure 44: CO_2 emitted from initial configuration scenarios and subcases at design temperature

Since the production rate and fuel gas consumption for the scenarios and subcases are quite different, Figure 44 does not provide an accurate picture when total emissions are compared to the production. The emitted amount of CO_2 per tonne LNG produced is shown in Figure 45 for all scenarios and subcases. Although Subcase B emits most in total, it actually emits least of the options with gas turbines when LNG production rate is taken into account.

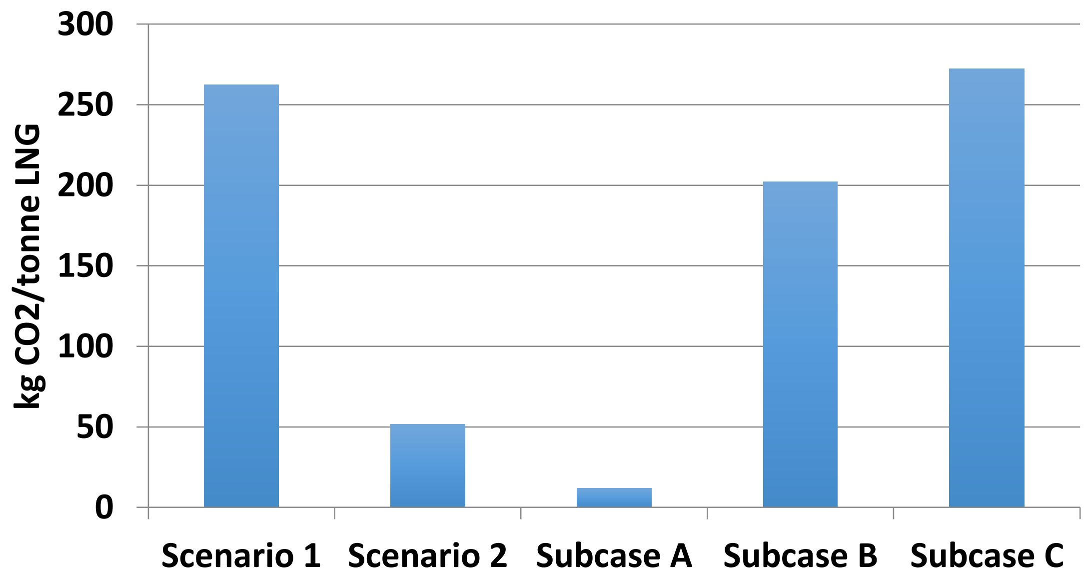

Figure 45: kg CO_2 emitted per tonne LNG produced

5.6 Discussion of the Simulation Results

The simulation results in this chapter underlines how the production varies with ambient temperature. Simulations performed in cold climates indicate a great potential for high and efficient production. The reason for this is the combination of a high available power output from the drivers and the high ability to cool the feed gas and refrigerant. However, the simulation is only based on production rate and specific power and does not reflect the need for increased size of pipes, heat exchangers etc., which in reality would restrict the flow and limit the production rate. This is further discussed in Section 6.2.1.

The great variation in production from location to location does not provide an accurate comparison when electric motors are used for compressor drive. The reason for this is that the output is matched against the gas turbine performance at the given location, thus leading to a different power output at each location. However, this does not apply when gas turbines are used as driver, as this is the same type for all scenario and subcases, thus giving a realistic result with respect to GT driver, despite location.

Another and possibly better method to compare the performance is to study the specific power. Figure 46 shows a summary of specific power for all scenario and subcases at design and high temperature. This excludes the varying production and driver output and illustrates the importance of a cold heat sink. The small variation in specific power for Scenario 2 underpins the advantage a seawater based cooling system has over an air based cooling system.

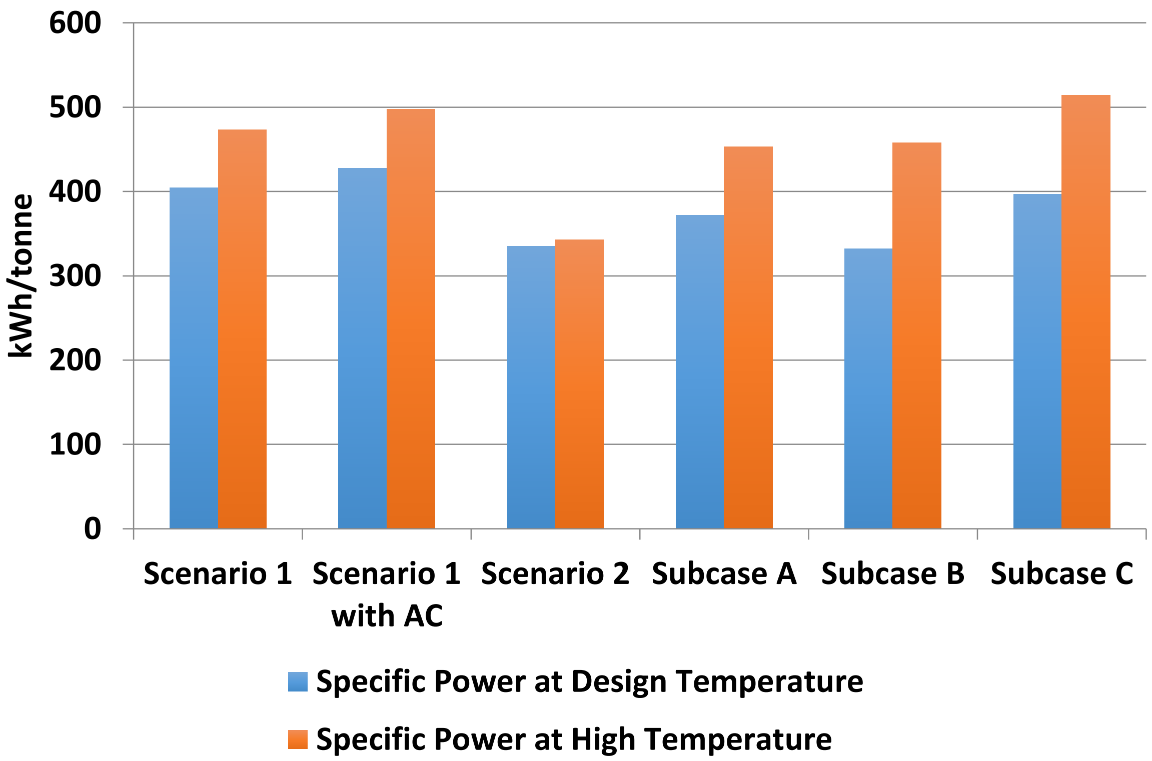

Figure 46: Specific power at design and high temperature

Scenario 1 with air coolers has been included in Figure 46 to illustrate the effect of cooling tower and air coolers. The results show that the specific power increase roughly 6% if air coolers are used instead of cooling towers. The effect depends largely on the humidity at the given location. For instance, Subcase C in Australia has an average temperature of 26.5°C but the dry air results in a wet bulb temperature of 19°C. This leads to a specific power of 396.7 kWh/tonne, which is slightly lower than for Scenario 1 with cooling tower (1.9%) even though the average temperature in Subcase C is 6.5°C higher than in Scenario 1. In other words, if an air based cooling system is to be used, the simulations results favours cooling tower, especially if the air at the given location is dry.

5.7 Summary of Chapter 5

The simulation results clearly shows that electrical drive combined with seawater cooling provides the highest, most efficient and stable production for all the alternatives evaluated in this chapter. Especially if several standardized units are the main goal, electrical drive should be chosen for compressor drive due to the constant power output despite air temperature, thus despite location. Seawater is colder, more stable and efficient than ambient air, which contributes significantly to a stable and efficient production. The advantage of this combination is clearly showed in the simulation results for Scenario 2 where the production only drops 1.8% at high temperature. If possible at a given location, seawater cooling should be chosen before air cooling.

6 Process Systems, Complexity and Reliability

This chapter evaluates possible layouts, complexity and reliability for the process systems. First, a simple analysis of the reliability and complexity has been performed on selected systems. Next, weight and dimensions of the FLSO are roughly estimated to get an indication of the physical size of the vessel and available deck space. Then, the location of different systems is discussed. Finally, a layout of the optimal process configuration is proposed, based on the results in Chapter 5, available deck area and the evaluation in the subsequent sections.

6.1 Process System Complexity and Reliability Analysis

To get a broader basis of comparison, some important factors that cannot be quantified in HYSYS have been studied. One of the most important is the reliability of the process systems. The reliability excludes the downtime due to scheduled maintenance, but this has already been taken into account in the number of production days per year.

6.1.1 PRICO Reliability Comparison

As mentioned earlier in this study, the choice of driver will affect the reliability of the plant. Based on data from previous studies, the compressor reliability with electric drive and gas turbine drive is 0.988 and 0.973, respectively (Miranda and Meira, 2008). The driver reliability comparison provides fair numbers for the whole PRICO liquefaction modules as long as the heat exchanger, connections, pumps etc. has an assumed reliability of 1. The configuration of PRICO is illustrated in a simple reliability block diagram in Figure 47.

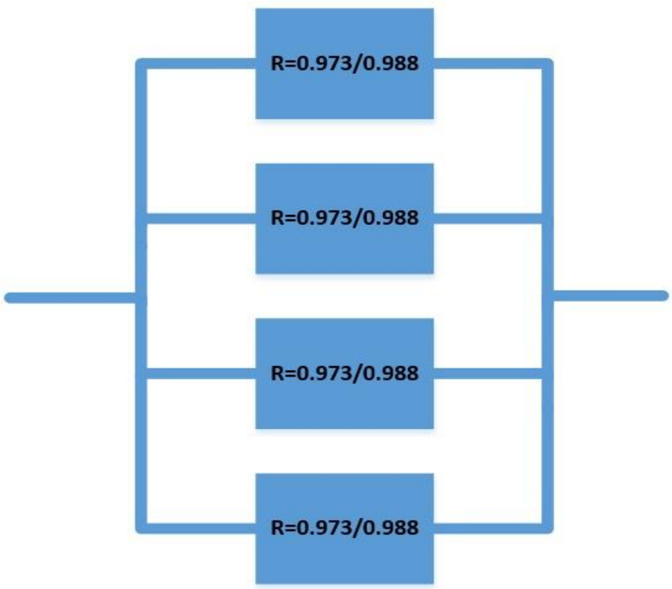

Figure 47: Reliability block diagram for 4x1.0 MTPA PRICO

Table 42 shows the reliability for the PRICO liquefaction process and is calculated with the binomial probability distribution equation C.1 in Appendix C. A production of 75% means that at least 3 out of 4 PRICO trains are in operation, etc.

Table 42: Reliability comparison for PRICO with GT and electrical compressor drive

	Production 100%	**Production 75% or more**	**Production 50% or more**	**Production 25% or more**
PRICO Electric drive	0.9529	0.9991	0.9999	≈1
PRICO GT drive	0.8963	0.9958	0.9999	≈1

For the PRICO trains, the electrical drive has a much better reliability when these are compared for a 100% production rate, resulting in 18.7 more days in operation with electrical drive. Note that the reliability of the power plant, frequency converters and other electrical components has not been included, which most likely would decrease the difference. This depends largely on the reliability and redundancy of the power plant or grid stability. Additionally, the fuel gas compressors for gas turbines has not been included, which would reduce the reliability of these alternatives even further.

However, the reliability difference is nearly equalized when the configuration results for a 75% production rate is calculated, since the liquefaction trains are independent of each other. In number of production days, the plant is able to produce at 75% or more approximately one day more per year with electrical drive. This analysis alone is probably not enough to justify the added cost and complexity with electrical drive. However, it clearly indicates that the increased reliability of electric drive has a much smaller impact on a multiple train PRICO process than it would for a single train DMR process where the drivers are configured in series.

6.1.2 Niche Reliability Comparison

While PRICO is configured in parallel, Niche is configured in serial parallel, leading to a lower reliability. A simple reliability block diagram of the plant with 3 Niche trains is shown in Figure 48. As for the PRICO calculation, the reliability of heat exchangers, connections etc. has been set to 1, but to get a more realistic result for the Niche process, it is chosen to include the reliability of the two companders in series with the drivers. The companders have a reliability of 0.99 each (Pettersen, 2015). Note that these are not included in Figure 48. The large number of rotating equipment results in a total reliability of 0.9279 for a GT driven Niche train and 0.9567 for an electrical driven Niche train. Table 43 shows the reliability for the complete plant with GT and electrical driven liquefaction compressors.

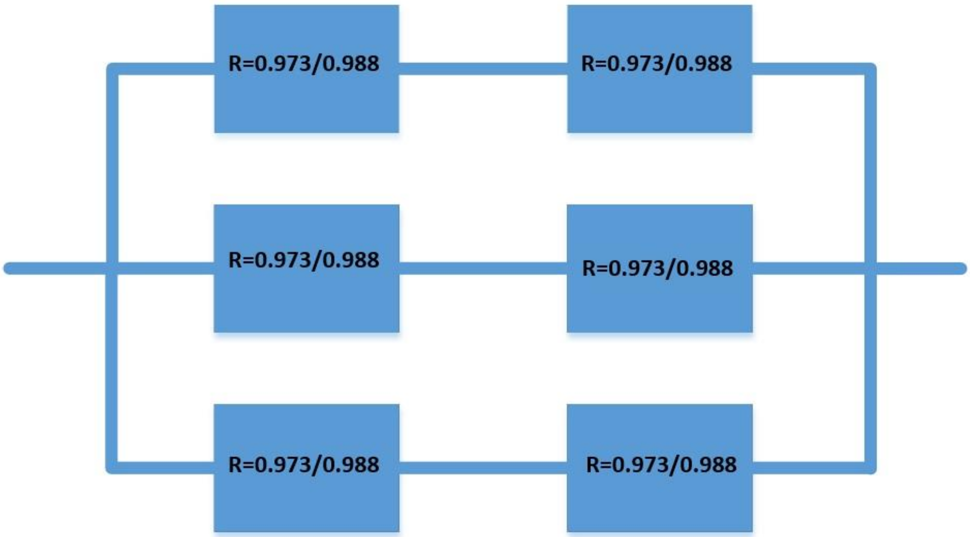

Figure 48: Reliability block diagram for Niche with three trains

Table 43: Reliability comparison for Niche with GT and electrical compressor drive

	100% Production	Production >67%	Production >33%
Niche GT drive	0.7989	0.9852	0.9996
Niche electrical drive	0.8757	0.9945	0.9999

As expected, the reliability of three trains in operation (100% production) for Niche is lower than for four PRICO trains in operation. The reliability in Scenario 1 with gas turbine driven liquefaction compressors is quite low. Scenario 2 in Northern Norway with electrical drivers is better, but still lower than PRICO with gas turbine drive. Comparing Niche GT and PRICO GT shows that the difference translates into 32 more days of 100% production for PRICO. If the electrical alternatives are compared, PRICO can maintain 100% production 25 days more than Niche.

6.1.3 NGL Extraction Comparison

To further identify the consequences of the two NGL extraction options evaluated in this study, a reliability comparison of the most important equipment has been done. Although an integrated distillation column has not been used in the simulations, it is regarded to be the most likely choice for integrated NGL extraction. The number of key equipment as well as the reliability is presented in Table 44.

It is assumed that the frontend compander has the same reliability as the ones in the Niche trains. Next, the pumps are configured with one in operation and one on standby, leading to a high reliability. Note that the two reliabilities for the booster compressor are for gas turbine driven and electric driven, respectively. For Scenario 1 and Subcase B and C, it is assumed that the booster compressor is driven by a gas turbine since the GT is used for refrigerant compressor drive. For Scenario 2 and Subcase A, the booster compressor is electrically driven.

Table 44: Equipment count and reliability comparison for integrated and frontend NGL extraction (Miranda and Meira, 2008), (Vicente, 2005)

Equipment	Integrated	Frontend	Reliability
Compander	0	1	0.990
Booster Compressor	0	1	0.973/0.988
Separator	1	1	1
Distillation Column	1	1	1
Heat Exchanger	0	1	1
Pumps (N+1)	2	0	0.9975
Equipment count	4	6	-
Total reliability	0.9975	0.9637/0.9781	-

The significant difference of roughly 3.3% between integrated and frontend with gas turbine drive translates into 11 days of production with 330 operational days. A possible solution is to have an electrical driven dual train frontend configuration. To make the frontend system able to compete with the integrated solution in terms of reliability, the dual train must be configured in a N+1 configuration, leading to a high reliability of 0.9995, if electrical driven. However, the added cost and complexity for this configuration will make this a very unfavourable solution. A more feasible solution will be to utilize a dual train running on partial load during normal operation. If one of the units unexpectedly fails, the other will run on full load and the plant can still be able to maintain a production in excess of 50% with 0.9995 reliability. The final and likely option is to have a J-T valve in standby that can bypass the compander if this should fail, thus maintaining some production. However, a detailed analysis of this has not been performed in this study.

6.1.4 Total Reliability

The overall reliability of the whole plant has been estimated to illustrate the impact each configuration alternative. In Table 45, a worst case and best case scenario has been outlined along with the results for the initial scenario configurations and subcases. The worst case consists of Niche liquefaction process, gas turbine driven compressors, frontend NGL extraction and local power generation with no redundancy. The best case for rich utilizes PRICO, electric drive and integrated NGL extraction. The best case for lean gas has frontend NGL extraction with an electrical driven booster compressor and with one unit in redundancy. Both of the best cases have redundancy in the power generation.

Note that only the most important components have been included in the analysis and only key numbers are presented in Table 45, meaning that utilities such as the cooling system and hot oil system is assumed to have a reliability of 1. For detailed description and calculation, see Appendix C.

Scenario 1, Subcase B and Subcase C uses a local power plant for the power generation. This is configured in a N+1 configuration, meaning that one gas turbine is on standby if one of the other should fail.

Scenario 2 in Norway and Subcase A in British Columbia use electrical compressor drive and electrical power is supplied through the grid. Collecting reliability numbers for a grid with a capacity in the range of 250-300 MW for the relevant locations have not been successful in this study. Therefore, the reliability of the grid is assumed to be the same as if the power was supplied from a local GTG power plant. This reliability is also assumed to account for frequency interference, voltage dips etc. that leads to production loss.

Table 45: Total reliability for selected systems

Case	Liquefaction Process	Type of power generation	Power generation	NGL Extraction	Reliability 100% production
Worst Case	0.7989	Local GTG N=1	0.973	0.9637	0.7491
Best case rich gas	0.9529	Grid	0.9860	0.9975	0.9372
Best case lean gas	0.9529	Grid/Local (N+1=7)	0.9860	0.9995	0.9391
Scenario 1	0.8963	Local GTG N+1=2	0.9993	0.9637	0.8632
Scenario 2	0.9529	Grid	0.9860	0.9781	0.9190
Subcase A	0.9529	Grid	0.9860	0.9781	0.9190
Subcase B	0.8963	Local GTG N+1=2	0.9993	0.9637	0.8632
Subcase C	0.8963	Local GTG N+1=3	0.9976	0.9637	0.8617

Note that the total reliability is for 100% production with every PRICO train in operation. Especially the reliability number for PRICO with GT drivers has a great impact on the total reliability. Since the trains are in parallel and independent of each other, the plant is still able to produce even though one driver fails. For instance, the reliability for 75% production in Scenario 1 will be 0.9590 compared to 0.8632 for 100% production.

The number of days with 100% production, adjusted with the reliability numbers, is shown in Table 46 for both scenarios and all subcases. The results indicates that Scenario 2 and Subcase A, where both operates with electrical drive, is close to the best case. The difference corresponds to roughly 6 days of 100% production for both Scenario 2 and Subcase A. If integrated NGL extraction is used in Scenario 2, the reliability will be equal to the best case. Note that the numbers in Table 46 are based on 330 days of operation per year.

Table 46: Number of days with 100% production for all scenario and subcases

Case	Number of days with 100% Production	Lost production days compared to best case
Worst Case	247.2	62.7
Best case rich	309.3	-
Best case lean	309.9	-
Scenario 1	284.9	25.0
Scenario 2	303.3	6.0
Subcase A	303.3	6.6
Subcase B	284.9	25.0
Subcase C	284.4	24.9

6.2 Flow Margins and Standardization Potential

In this subchapter, flow margins in the PRICO liquefaction modules are discussed and linked to the simulated results to determine the potential for standardization and relocation of the FLSO.

6.2.1 Flow Margins in the Process Systems

A flow margin of 10% for the modules has been suggested by Höegh LNG, meaning that the potential for high production for standardized modules is limited (Revheim, 2015). For instance, 4 PRICO modules designed to produce 1.0 MTPA each can only be pushed to produce 4.4 MTPA in total when the air temperature is low, thus becoming the bottleneck of the production. The gas turbines, if used, may then be running on partial load, thus not exploiting their full potential. This is further described in Section 6.2.2.

The flow margins are determined by the suppliers to ensure that the design capacity is achieved. An alternative is to design the systems with a higher flow margin, making the modules able to produce more at low temperatures and limit the lost production. However, at higher temperatures, the designed production rate cannot be reached. Additionally, the added cost associated with increased size must be carefully evaluated to determine which temperature to size the equipment for.

The simulated flowrates for all scenarios and subcases at design temperature are shown in Table 47. Note that this is the total flowrate for four PRICO trains. As both the natural gas flow rate and refrigerant flow rate depends on the available compressor power, the results varies for each location. However, the highest refrigerant flow rate in Scenario 2 is only 13.8% higher than the lowest in Subcase C, even though the natural gas flow rate is 53.7% higher in Scenario 2. These results indicate that the refrigerant side of the liquefaction modules (except compressor and driver) is significantly easier to standardize.

Table 47: Simulated natural gas and refrigerant flow rate at design temperature

Location	Unit	Natural Gas Flow Rate	Refrigerant Flow Rate
Scenario 1 – Gulf of Mexico	kg/s	142	1052
Scenario 2 – Northern Norway	kg/s	210	1133
Subcase A – British Columbia	kg/s	179	1124
Subcase B – Northern Russia	kg/s	210	1066
Subcase C – Northern Australia	kg/s	137	995

6.2.2 Standardization Potential and Relocation Issues

The findings in this part of the study indicate that the selection of driver, cooling method and design capacity of the plant should be performed after the location and climate is known. This will have a negative effect on the potential for standardization and possible relocation of the vessel.

The result for Scenario 2 in Norway and Subcase B in Russia at design temperature indicate that the compressor drivers are too powerful to drive the standard 1.0 MTPA PRICO modules. A possible solution for production in cold climate is to upscale each train to 1.33 MTPA, thus achieving a production rate of roughly 4.1 MTPA at design conditions with 3 trains and Trent 60 or corresponding electrical drive. Another solution is to use the 1.0 MTPA train, driven by a gas turbine with lower power output than the RR Trent 60 or corresponding electric motor. The GE LM6000 is a good and thoroughly field proven alternative with a power output of 47 MW at ISO conditions compared to 53 MW for the Trent 60. However, if a smaller gas turbine is used, this will have a negative effect at higher temperatures, especially if air cooling is used. This is reflected in Subcase B where the production rate drops 47% at high temperature.

The use of electric drive or seawater cooling is two options to stabilize the production through the year and simplify the sizing of process equipment. The use of electric drive instead of gas turbine drive is well illustrated if the alternative configuration results are compared to the initial results for Scenario 1 in the Gulf of Mexico. At high temperature, the plant with electrical drive is able to produce 22% more than if GT drivers are used. Note that this does not take the daily temperature variations into account, where an electrical driven liquefaction process will be easier to operate.

As shown in the results for Scenario 1 and Subcase B, the combination of gas turbine driver and air as heat sink results in large variations in production. If the ambient air temperature varies much throughout the year, the sizing of process equipment becomes even more challenging, making this configuration even less favourable.

Figure 49 shows the performance for two LNG plants with gas turbine compressor drive and air coolers. The first plant, illustrated by the green solid line, is designed to produce 3.63 MTPA at 20°C ambient air temperature (Scenario 1 with air coolers). With the 10% flow margin included, the production can be pushed to 4 MTPA already at 16°C. The green dotted line shows the potential production with an unlimited flow margin where the production increases when the Trent 60 driver output increase and available cooling water temperature decrease (based on numbers from Subcase B). In other words, if a FLSO with GT drivers designed to operate at 20°C is relocated to a location with cold climate with an average temperature of 6°C, the potential for the FLSO will not be fully exploited. The production will then be limited to 4.0 MTPA, but the driver and cooling has a capacity to produce 23% more (4.93 MTPA).

The black line illustrates the production for a plant designed to produce 3.63 MTPA at 6°C with the GT drivers running on 100% load. The required drivers are smaller than a Trent 60, but the sensitivity to temperature is assumed to be the similar. If this FLSO is relocated to a warmer climate, for instance Gulf of Mexico with 20°C average temperature, the production will be reduced to 73.6%. For the high temperature of 36°C in Gulf of Mexico, the production will be roughly 50% of the capacity the plant was initially designed for.

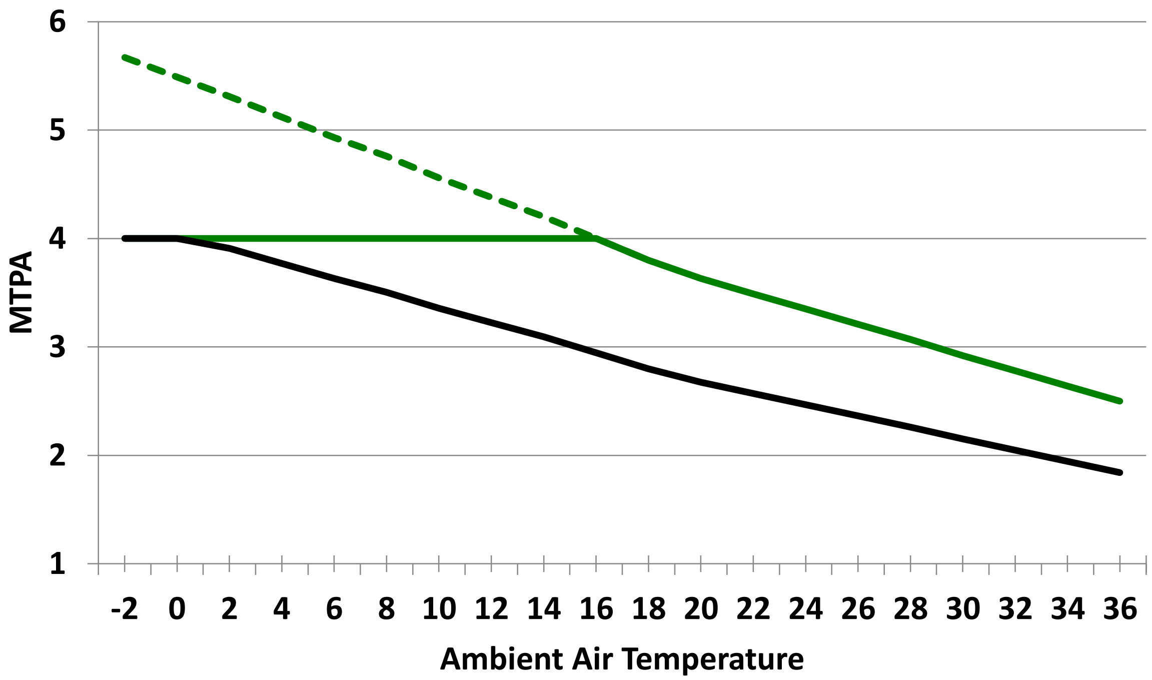

Figure 49: Production for two LNG plants designed to produce 3.63 MTPA at 20 ℃ (green) and 6 ℃ (black)

Figure 49 indicates that a standardized unit with potential for relocation is not favourable for a FLSO with GT drivers and air cooling if the design temperature at the two locations is very different. GT driver combined with cooling tower will behave similarly. A good alternative with respect to standardized liquefaction modules is to use

LM 6000 GT drive in cold climates and Trent 60 (which has a higher performance) in warm climate. As mentioned earlier, both are thoroughly field proven and the difference in power output will partially equalize the production rate. Optimal exploitation of the modules can then be achieved with lower flow margins.

6.3 Final Comparison of Process Systems

In this subsection, a final comparison of the process systems is performed. This combines the simulation results from Chapter 5 with the reliability results in Section 6.1.

6.3.1 Liquefaction Process

The specific power of Niche is roughly 33% higher than PRICO for both Scenario 1 and 2. Next, the higher reliability of PRICO results in 32 more days with 100% production when gas turbine driven PRICO and Niche are compared. If electrical drivers are used the difference is 25 days. In other words, PRICO liquefaction process is better than Niche in terms of efficiency, capacity and reliability. Although Niche is safer in operation, PRICO is regarded to be the preferred liquefaction alternative in all scenarios and subcases.

6.3.2 Compressor Driver

Electrical drive has proven to be more reliable and provides a stable power output as long as the power supply is sufficient. Next, comparing Subcase A and B, where both produce from a lean gas shows that the total emitted CO_2 can be reduced 95% if electrical drive is selected instead of GT. If electrical drive is combined with seawater cooling, the production rate and specific power is nearly constant, despite temperature. Without taking the increased cost into account, electrical drive should be chosen before gas turbine drive. However, a detailed comparison of the CAPEX of the two alternatives and the increased income with electrical drive should be performed, but this has been moved to further work in this study.

6.3.3 NGL Extraction

The simulation results from Scenario 2, presented and discussed in Section 5.2.1, shows that a rich gas configuration with frontend NGL extraction can produce roughly 7% (0.37 MTPA) more than a configuration with integrated NGL extraction. However, the specific power is 4.5% higher for the frontend configuration due to the increased power demand for a booster compressor. When the reliability in Section 6.1.3 is taken into account, a configuration using frontend is able to produce 5% (0.26 MTPA) more, but with the same specific power.

The added cost and complexity, reduced reliability and efficiency and more challenging operation for frontend NGL extraction does not justify the slightly increased production for a plant producing from rich gas. Frontend NGL extraction is therefore only considered to be favourable for plants producing from lean gas. For Scenario 2, this relieves the grid of approximately 23 MW (8% of total power duty) as the booster compressor is not required.

6.3.4 Electrical Drive vs Seawater Cooling

As mentioned, the cost of both electrical drive and seawater cooling is much higher than the less efficient alternatives. If only one of the most favourable alternatives for driver and cooling can be chosen, the section depends largely on the governmental restrictions and environmental focus at the location. From an efficiency point of view, seawater should be chosen before electrical drive.

From a production rate point of view, Scenario 2 shows that the production drops 23.6% at high temperature if gas turbines are used instead of electric motors. If air coolers are used instead of seawater, combined with electrical drive, the production drops 21.8%. This indicates that the gain of electric motor is greater than the gain of seawater. If the reliability is included, this leads to a higher production with electrical drive. Whereas the choice between seawater and any air cooled system has no effect on the reliability of the cooling system in this study, electrical drive makes the plant able to maintain 100% production 18.7 more days per year.

Where the use of seawater cooling might increase the risk of affecting the environment, electrical drive offers a huge potential for lower CO_2 emissions if the power is generated from renewable energy sources.

Next, electrical drive is regarded to be easier to implement than seawater cooling, especially if the plant is located in closed harbour. Additionally, governmental restrictions may not allow seawater cooling at all. The discussion in this subchapter shows that electrical drive should be the preferred choice if this stands between electrical drive and seawater cooling.

6.4 Vessel Design and Dimensions of Selected Process Systems

The first priority of the vessel is to design a hull with a sufficient storage capacity. A possible way to calculate this is to assume that one meter ship length is equivalent to 1000 m^3 storage capacity for a vessel with 60 meter beam and 32.5 meters height. This implies that 250 meters is needed just for the 250 000 m^3 storage capacity (Revheim, 2015). Note that this capacity is based on a LNG production of approximately 4.0 MTPA. If the production is similar to the simulated results in Scenario 2 or Subcase B, the capacity should be increased.

Compared to a traditional LNG carrier or offshore FLNG, there are certain issues that can be excluded for at-shore FLNG. For instance, there is no need for a large engine room since the vessel will be towed and moored to shore. Next, LNG tanks cannot be located under living quarter, but for the vessel options in this study, all living quarters will be onshore. Hence, LNG storage tanks can occupy a larger part of the hull. However, ballast tanks are needed in the stern and bow. In this study it is assumed that these will occupy 30 meter aft of the LNG storage and 20 meters in the bow, resulting in a total length of 300 meters (Revheim, 2015).

To further illustrate ship dimensions, simple calculations have been performed on the weight of different components, shown in Table 48. The configuration is based on Scenario 1 with the initial configuration. As there are no FLNG or FLSO in operation per today, the dead weight of the hull is based on the LNG carrier type Q-max. This carrier is 345 meters long, 53.8 meters wide and has a deadweight of 124 690 tonnes (Chubu, 2010). However, the hull must most likely be strengthened to be able to handle the increased topside weight when the process systems are installed. Additionally, the membrane type storage tanks must be lowered to get a flat deck, and the wheelhouse and engine room can be removed etc. In other words, take the deadweight tonnage (DWT) for the hull in this section as an indication rather than exact numbers.

Another large source of error is the weight of the topside equipment. The table only accounts for the main components of the process. These do not occupy the whole available deck area, meaning that other systems will be placed on the FLSO. The realistic weight also includes piping, utilities, offloading and flare, which implies that the topside weight will be significantly higher than the one outlined in Table 48.

Table 48: Weight estimation for the vessel (Talib et al. 2011) (Centrax, 2015)

System	Volume/number of units	Specific Weight	Total Weight	Footprint m x m
LNG Storage	250 000 m^3	450 kg/m^3	112 500 T	-
PRICO 1.0 MTPA Liquefaction Module (B&V)	4	3150 T	12 600 T	26 x 46 (each unit)
BOG Module (B&V)	1	3640 T	3640 T	-
Refrigerant Makeup Module (B&V)	1	840 T	840 T	-
Fractionation module (B&V)	1		1380 T	-
Rolls Royce Trent 60 Generator	2	300 T	600	29.4 x 8.5 (each unit)
Total Weight Topside	-	-	19060	-
Hull Deadweight	-	124690 T	124690 T	-
Total deadweight	-	-	143750	-
Full load displacement	-	-	256250	-

6.5 Location of Process Systems

In this subchapter, the location of the different process systems is discussed. As mentioned in section 2.1.1, placing some of the systems on the FLSO can result in a lower investment cost and shorter delivery time. However, some of the systems may differ significantly from case to case due to the different climate and feed gas, while some systems are identical.

This limits some of the standardization potential, as part of the systems will need to be suited for different configurations. However, part of the vessel can be standardized, as several of the systems can be almost identical from case to case. This opens up for a partly standardized vessel where one section can be fully standardized and the other can be fitted with field specific systems based on climate, feed gas composition and governmental restrictions. This will result in a vessel similar to the design of FLEX LNG.

If the feed gas composition requires a large amount of process equipment, one option is to reduce the number of trains, thus freeing some deck space to these. This depends on factors like remoteness of the facility, CAPEX, size of the reservoir etc. A life cycle cost analysis should be conducted to optimize the configuration but this has not been the main focus of this study and has therefore not been studied any further.

As discussed in Section 2.9, the flexible connections between the FLSO and the shore are an important issue. The number of connections will be of great importance when the location of different systems is evaluated. To avoid too many connections, the location of each system should follow Figure 20 and Figure 21 in Chapter 4. For instance, if the liquefaction and storage is on the FLSO, the end flash should be placed on the FLSO as well. This means that it is not regarded as an option to place the CO_2-removal system on board and the frontend NGL system (if used) onshore and so on.

6.5.1 Liquefaction, End Flash, Storage and Offloading

As shown in Figure 3 in Chapter 2, the liquefaction and storage is a large part of the investment cost and therefore has a great cost reducing potential if these are placed on the FLSO. The liquefaction can be nearly standardized to fit all cases despite different feed gas composition and climates and should therefore be placed on the standardized section of the vessel. Different compressor drivers are assumed to be relatively easy to implement. To avoid connections back and forth, the end flash and offloading system should also be placed on the vessel. When carriers come to load the LNG, the FLSO serves as a jetty.

6.5.2 Cooling and Heating System

As mentioned in Section 1.1, it is decided to use an indirect cooling system for all scenarios and subcases. The decision is based on evaluation in the specialization project where an indirect system fits the standardization criteria best due to the different heat rejection methods utilized in the different scenarios and subcases (Corneliussen and

Samnøy, 2014). For the same reason, the final heat rejection method should be placed onshore, whether it is a seawater system, air coolers or cooling towers.

If waste heat recovery is being used, the heat generation should obviously be placed where the gas turbines are. If the plant utilizes electric drive and power from the grid, the heater should be located onshore to minimize the ignition sources on the vessel.

6.5.3 Power Generation

The location of the power generation depends largely on the choice of compressor drive. If gas turbines are used as driver, less gas turbines are needed in the power plant. However, if electric motors are used with local power generation, a larger power plant will be required and this will most likely not fit the vessel deck.

Since the use of gas turbines does not apply for all cases, these should be placed on the field specific section.

6.5.4 Inlet Separation, Condensate Stabilization and Storage

As the feed gas composition varies from case to case, the condensate production rate and system size will also vary. The inlet separation and condensate stabilization system should therefore be located onshore. Another argument is that it is probably not enough space on the FLSO for the downstream processes (amine, dehydration and NGL system) and if the inlet separation were placed on the vessel, connections back and forth would be required.

Lighter components separated in the condensate stabilization will be recompressed and reinjected into the lean gas. For a rich gas scenario, where the production rate of condensate can become quite high, a large onshore storage tank is the best alternative. The condensate for rich gas scenarios should therefore be loaded of the side of the FLSO to condensate carriers, resulting in one connection to the shore. For a lean gas scenario, the production rate is expected to be quite low, which opens up for transportation by trucks.

6.5.5 NGL Extraction, Fractionation and LPG Storage

Frontend NGL extraction could be placed on the vessel, as it is the process system closest to the liquefaction trains. However, the heavy components separated out must be fractionated. For a lean feed gas scenarios, the fractionation system may be a module similar to the one that Black and Veatch delivers and can be placed on board in the field specific section. This module separates condensate and the NGL components are returned to the liquefaction process.

Black and Veatch also offer an integrated NGL separator for their PRICO modules, which is only relevant for a rich feed gas composition in this study. However, extensive studies should be done to determine if this could extract a sufficient amount of NGL. For a rich gas, the large fractionation train required should most likely be located onshore due to the large footprint it entails. However, if available deck space on the FLSO is not exploited, the fractionation system for rich gas could also be placed on the FLSO. Since a LPG storage tank is only needed for a rich feed gas, these tanks should be located onshore.

Table 49 shows the number of connections between the FLSO and shore for the possible NGL and fractionation configurations if placed on the vessel. Note that the natural gas is already assumed to be on the vessel for all alternatives in the evaluation. Next, the required connections for hot oil have not been included. Two connections for hot oil might be required if there is no other heat consuming processes onboard. Note that the connections between storage and offloading for LPG and condensate in a rich gas case have not been included. To determine the size of the connections, the respective flow rates from Scenario 2 have been used for the rich alternatives while Subcase B has been used for the lean gas alternatives. All connection arms are of the type B0300 developed by Emco Wheaton, presented in Section 2.9.

Table 49: Number of connections between FLSO and shore for different NGL extraction systems and locations

Possible Scenario	Number of Connections	Flow Description	Flowrate [m^3/h]	Connection Arm Size
Lean gas, frontend NGL extraction and fractionation on FLSO.	1	Condensate to storage	9.63	4”
Lean gas, frontend NGL extraction and fractionation onshore.	2	Feed gas to shore	2474	12”
		Reinjection	45.26	4”
Rich gas, integrated NGL and fractionation on FLSO.	2	Condensate to storage	122.4	4”
		LPG to storage	112.0	4”
Rich gas, integrated NGL and fractionation onshore	2	NGL to fractionation	625.0	8”
		Reinjection	490.6	6”
Rich gas, frontend NGL on FLSO and fractionation onshore	2	NGL to fractionation	196.8	4”
		Reinjection	48.9	4”

As shown in Table 49, the most favourable alternative for a lean gas scenario with frontend NGL extraction and fractionation on the FLSO only requires one 4” connection arm to transport condensate to storage.

All rich gas cases results in the same number of connections. Note that the flow rate numbers for integrated NGL extraction could be somewhat unrealistic since the NGL extraction is simulated with a separator. This leads to a very high methane fraction of the NGL extracted, which then again leads to a large connection arm for both NGL to fractionation and reinjection of methane and ethane. The required arm size would most likely be smaller if a distillation column was used in the simulations, but this has not been done due to optimization complexity in HYSYS, as described in Chapter 4.3.

6.5.6 CO_2 Removal and Dehydration

As suggested earlier in this subchapter, the inlet separator and condensate treatment system should be placed onshore due to large variations in the feed gas composition. For the same reason, the CO_2-removal and dehydration should be placed onshore. As the CO_2 and water content varies with feed gas compositions and cooling water temperature, the required size of these systems also varies. Additionally, if the CO_2 is to be reinjected into a reservoir (for rich gas cases), the pipeline exit should be at the feed gas intake, which is suggested onshore. This would lead to an extra connection between the FLSO and shore for the CO_2 if these systems were placed at the vessel. The same goes for the liquid water after dehydration, as further treatment most likely will be located onshore.

However, if available deck space on the FLSO is not utilized, the CO_2 removal and dehydration system could be placed at the field specific part of the vessel. This especially applies for a rich gas scenario with integrated NGL extraction, as this system is included in the liquefaction modules.

6.6 Standardized FLSO Section

Based on the evaluation is this chapter, some systems clearly points out to be placed on the FLSO while others should be placed onshore. The liquefaction modules, LNG storage, end flash, offloading system and flare system can be standardized to a certain point and should be placed on the FLSO for all scenarios and subcases. The rest of the vessel deck can be occupied by field specific equipment and can vary from location to location. The field specific section is discussed further in the next subchapter.

Figure 50 and Figure 51 illustrates two possible layouts for the standardized section. As shown in the first figure, the deck is simply split in two between the standardized and field specific section. Due to the required safety gaps of 15 meters between each fire zone, the deck is not fully utilized in width.

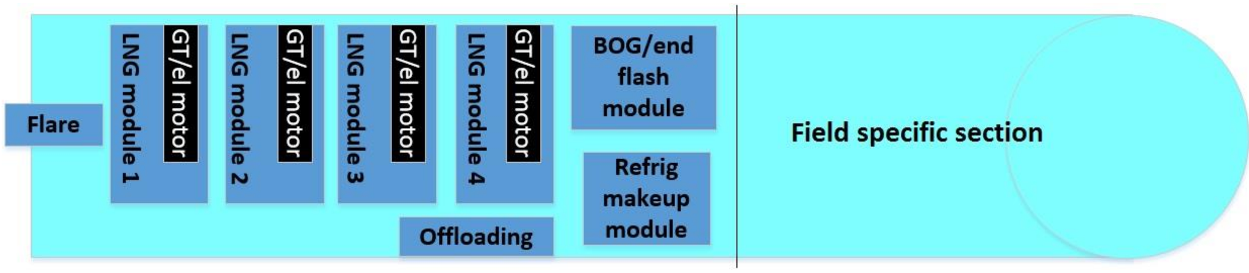

Figure 50: Proposed layout for the standardized section of the FLSO with LNG modules across the length direction

The next possible layout is to place the PRICO modules with the longest side of the module parallel to the ship length as shown in Figure 51. This is similar to the layout of the Lavaca Bay project.

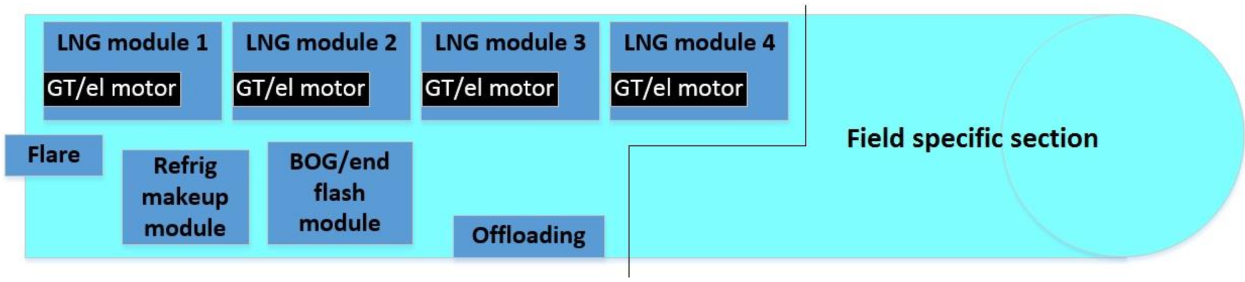

Figure 51: Proposed layout for the standardized section of the FLSO with LNG modules lengthwise

6.7 Field Specific FLSO Section

For any scenario or subcase producing from lean pipeline gas, the frontend NGL extraction and fractionation module can be placed on the FLSO. With respect to the number of connections between the FLSO and shore, this seems like an attractable alternative.

Earlier projects such as Melkøya as well as ongoing projects such as the Lavaca Bay indicate that the power generation should be placed on the FLSO if gas turbines are used, making this an attractable alternative as well.

Determining the optimal system to place in the field specific section requires detailed information regarding weight, footprint and potential cost savings. However, systems for the field specific section are proposed in the subsequent sections, based on the findings in this study.

6.8 Proposed Layout for the FLSO

In the following subchapters, a proposed field specific layout of the FLSO for a lean gas and a rich gas scenario is presented. Note that on the standardized section for the rich gas scenario, the liquefaction modules includes integrated NGL extraction, making them not completely standardized for both rich and lean gas. However, if completely standardized liquefaction modules (besides driver and compressor) are desired, frontend turboexpander NGL extraction must be installed for rich gas as well.

6.8.1 Proposed Layout for a Lean Gas Scenario

Figure 52 shows a proposed layout for a the initial configuration in Scenario 1 with PRICO liquefaction, frontend NGL extraction, fractionation, power generation, makeup refrigerant and BOG module on board. Note that the frontend NGL extraction footprint has not been obtained and is therefore assumed. Compared to the layout of Lavaca Bay, this project has a larger available deck area in the bow of the vessel since living quarters have been moved ashore.

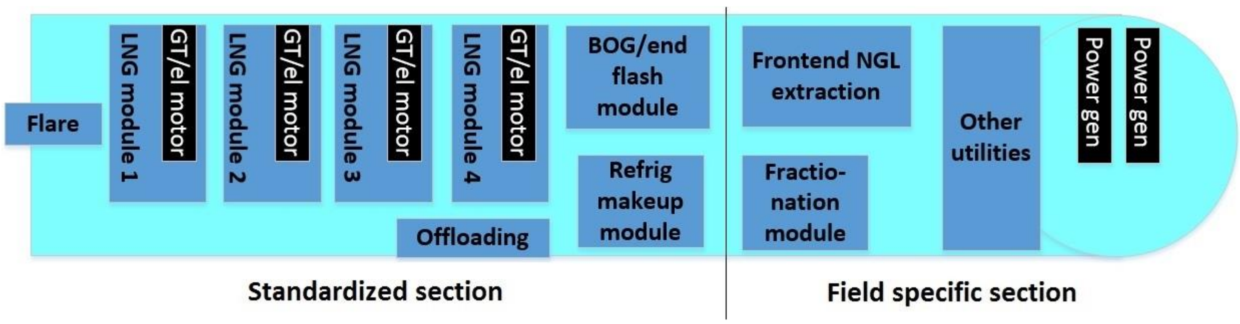

Figure 52: Proposed layout of the FLSO for a lean gas scenario

The number of connections between the FLSO and shore for the layout in Figure 52 is given in Table 50. The connections transporting hydrocarbons are assumed to be of the type B0300 by Emco Wheaton, which was introduced in Section 2.9. The arm size is based on the given arm capacity in Table 6 and simulated flow data. All heat is generated on board, making another two connections to shore necessary for heat transport. The connections for cooling water and hot oil do not have the same requirements as the natural gas and condensate. In total, this layout requires six connections to shore, but several more is needed for electrical power cables, refrigerant makeup etc., but this has not been included in this study.

Table 50: Connections to shore for a lean gas scenario

Flow	Arm Size	Number of connection arms required
Natural gas to NGL extraction	12”	1
Condensate to shore	4”	1
Cooling water	-	2
Hot oil	-	2
Total		6

6.8.2 Proposed Layout for a Rich Gas Scenario

Figure 53 shows a proposed layout for Scenario 2 with integrated NGL extraction. Note that integrated NGL extraction is included in the liquefaction modules in the standardized section, and that the dimensions for fractionation, CO_2 removal and dehydration have not been successfully obtained and is therefore assumed.

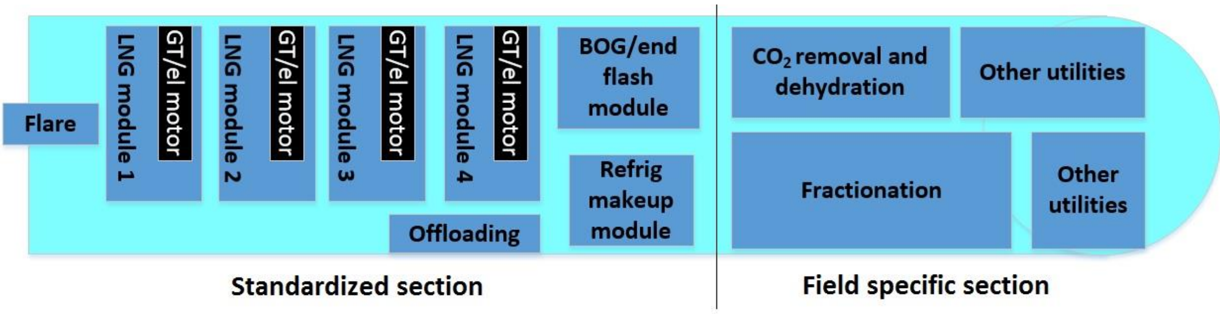

Figure 53: Proposed layout of the FLSO for a rich gas scenario

Number of connections between the FLSO and shore for the layout in Figure 53 is given in Table 51. In total, this layout requires eleven connections to shore, but several more is needed for electrical power cables, refrigerant makeup etc., but this has not been included in this study.

Table 51: Connections to shore for a rich gas scenario

Flow	Arm size	Number of connection arms required
Natural gas	16”	1
Condensate to shore	4”	1
Condensate storage to FLSO for offloading	-	1
LPG to shore	4”	1
LPG storage to FLSO for offloading	-	1
CO_2 removed in amine system	4”	1
Fuel Gas for heater	4”	1
Cooling water	-	2
Hot oil	-	2
Total		11

7 Conclusion and Recommendations

Through this report, various process and utility systems have been evaluated for at-shore FLNG based on five locations with potential for LNG production.

Regarding the liquefaction processes evaluated in this study, the simulation results clearly favours PRICO, as Niche operates with a specific power 33.4% and 33.0% higher than PRICO in Scenario 1 and Scenario 2, respectively. Next, PRICO requires less rotating equipment, which makes it easier in operation, easier to modularize and more reliable.

The common use of gas turbine driver and air cooling gives the poorest performance in terms of production capacity and stability. These alternatives are both directly dependent of the ambient air temperature, which results in large variations in production. The large temperature variation of 29°C from average to high in Subcase B in Northern Russia is the most extreme in this study and leads to a 47% drop in production rate at high temperature.

The simulation results clearly favours electrical drive combined with seawater cooling, which provides the highest, most efficient and stable production for all the alternatives evaluated. This is well illustrated in the initial configuration for Scenario 2, where the high production of 5.51 MTPA at design temperature decreases 1.8% at high temperature. The advantage of this combination is further underpinned by the higher reliability an electrical motor offers compared to a gas turbine if the reliability of the power generation or grid is sufficient. The difference in reliability for the electrical motor alone translates into approximately 19 days of 100% production.

However, both seawater cooling and electrical drive are major cost drivers and have not been the traditional choice due to the high CAPEX. On the other hand, the growing global environmental focus may change the view on driver combined with power from a renewable energy source. A simple comparison of the CO_2 emission from Subcase A and B, which both produce from lean gas and have approximately the same power demand, shows that the total CO_2 emitted locally can be reduced 95% if electrical drive is used instead of gas turbines. Next, the use of seawater makes the plant more efficient due to a lower temperature, but the large amount of heat rejected might disturb the biological life in the sea, making other alternatives more favourable at some locations. If the FLNG plant can be designed with only one of these alternatives, electrical compressor drive should be chosen before seawater cooling.

The two NGL extraction options evaluated in this study are frontend turboexpander and integrated, both common in LNG plants. When the reliability of the two options is taken into account, Scenario 2 in Northern Norway is able to produce 5% more with frontend, but with a 4% lower efficiency. This indicates that the added cost, footprint and complexity, reduced reliability and efficiency as well as more challenging operation for frontend NGL extraction does not justify the slightly increased production for a plant producing from rich gas. For a lean gas, however, a frontend NGL system has been regarded to be the only option that can sufficiently extract critical components, such as benzene, without a major decrease in liquefaction efficiency and production.

Due to the large variation in feed gas composition, air temperature and local conditions at the locations in this study, a fully standardized unit is difficult to achieve. Especially the refrigerant compressor driver, cooling system and upstream gas processing systems have turned out to be hard to standardize while obtaining an optimal production and efficiency at the same time. Comparing the results from Subcase B in Northwest Russia and Subcase C in Northwest Australia shows that the production capacity at design temperature (average ambient temperature) is roughly 57% higher for Subcase B with the same gas turbine driver. Next, the required process systems upstream of the liquefaction process will vary significantly in size and complexity when the feed gas composition varies. This indicates that these systems should be selected after the location and feed gas composition is known.

However, the liquefaction process, storage tanks, offloading, end flash and flare system can be standardized to a certain point and placed on the FLSO. To be able to standardize the liquefaction modules, a custom driver and compressor will be required. If gas turbine compressor drivers are to be used, the LM6000 is suggested for a cold climate and Trent 60 is suggested for a warm climate. If electrical motors are to be used, the output should correspond to the performance of the respective GT. Next, if standardized liquefaction modules (besides driver and compressor) are desired for production from both a lean gas and rich gas scenario, the only option for NGL extraction is the frontend turboexpander process. Finally, the complexity of the end flash system will vary with the nitrogen content in the gas. The simulation results for the end flash in a rich gas case show that the maximum limit of 1 mol% in the LNG product is fulfilled. However, if the feed gas contains more than 2.53 mol% nitrogen as in this study, a more advanced end flash system might be required and cannot be standardized.

This leads to a FLSO layout with standardized systems in one section and field specific equipment in the other. The field specific section may be fitted with gas processing systems, power generation or other utilities, depending on the local conditions and restrictions. Another solution is to have one standardized section layout for a lean gas scenario and another for a rich gas scenario, as suggested in Chapter 6.8.1 and 6.8.2, which makes integrated NGL extraction possible for a rich gas scenario.

However, even if the FLSO is partly outfitted with field specific equipment, relocation can lead to inefficient exploitation of the plant. This is due to production limitations caused by flow margins in the natural gas side of the liquefaction modules in the standardized FLSO section. Higher production can be achieved with higher flow margins, but this also leads to higher cost due to increased size of pipes, heat exchangers etc. Based on the simulations, the refrigerant side of the liquefaction modules is easier to standardize with the exception of driver and compressor. By using a driver with lower output than the Trent 60 in cold climate, the large difference in production can be decreased, thus decreasing the required flow margins to achieve optimal production.

8 Further Work

There are several interesting issues that should be studied further to get a better basis for comparison.

- A thorough cost analysis should be performed on electrical and gas turbine compressor drive to compare the CAPEX with the reliability and production stability. For an electrical motor, the higher reliability leads to more days in production and the better stability results in higher production and easier operation throughout the year. Whether or not the increase in production is high enough to justify the increased CAPEX needs to be determined.
- A thorough cost comparison of indirect seawater cooling and air cooling to determine if the increased cost of seawater cooling can be justified by the plant efficiency and increased production.
- Consider the potential for a direct cooling water system for the liquefaction cooling and an indirect cooling circuit for the rest. Some of the important criteria that need to be evaluated are cost, standardization options, efficiency compared to direct and indirect seawater, compactness and maintenance requirements.
- This study shows that the capacity of each train varies at the different locations although the compressor driver is the same. An optimization of train design capacity at a given location and with a given cooling method should be performed to exploit the potential of the driver.
- Evaluation of integrated NGL extraction shows a great potential with respect to efficiency, reliability and equipment count. However, the model in this study uses a two-phase separator instead of the traditional reboiled absorber (distillation column), thus leading to some unrealistically results. A more detailed and realistic evaluation of the NGL process should be performed to get a better basis for comparison for the two commonly used options.
- A detailed analysis of a centralized combined cycle power plant with steam extraction should be performed on a lean gas scenario. If the compressor drivers are electrical driven, the power plant will be large and the required steam flow rate for heat in the processes will be limited due to the lean gas. One should investigate how the variable steam extraction will affect the efficiency of the steam turbine. This configuration seems to have a good potential, especially when the heat demand is low compared to the exploitable waste heat from the gas turbines. The different temperature requirements for the heat processes should be defined to find out what pressure the steam needs to be extracted at. In addition, the layout should be roughly specified to identify the heat losses in pipes, connections etc.

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Appendix A – Detailed Simulation Results

Detailed HYSYS data for the initial configuration of Scenario 1

Property	Unit	Result at design temperature	Result at high temperature
LNG HX refrigerant HP inlet	bar	29.07	39.50
LNG HX refrigerant LP inlet	bar	5.45	7.14
Intermediate Pressure (outlet refrigerant compressor 1)	bar	16.94	22.74
Indirect CW circulation rate	m^3/h	28540	22230

Detailed HYSYS data for the initial configuration of Scenario 2

Property	Unit	Result at design temperature	Result at high temperature
LNG HX refrigerant HP inlet	bar	33.00	35.25
LNG HX refrigerant LP inlet	bar	4.79	5.07
Intermediate Pressure (outlet refrigerant compressor 1)	bar	17.90	19.06
Indirect CW circulation rate	m^3/h	47350	46950

Detailed HYSYS data for Subcase A

Property	Unit	Result at design temperature	Result at high temperature
LNG HX refrigerant HP inlet	bar	35.85	39.50
LNG HX refrigerant LP inlet	bar	6.20	6.75
Intermediate Pressure (outlet compressor 1)	bar	20.05	22.02
Indirect CW circulation rate	m^3/h	34530	30720

Detailed HYSYS data for Subcase B.

Property	Unit	Result at design temperature	Result at high temperature
LNG HX refrigerant HP inlet	bar	36.17	31.14
LNG HX refrigerant LP inlet	bar	5.49	4.83
Intermediate Pressure (outlet compressor 1)	bar	18.48	16.01
Indirect CW circulation rate	m^3/h	35590	22890

Detailed HYSYS data for Subcase C.

Property	Unit	Result at design temperature	Result at high temperature
LNG HX refrigerant HP inlet	bar	36.46	38.54
LNG HX refrigerant LP inlet	bar	7.54	7.93
Intermediate Pressure (outlet compressor 1)	bar	21.27	22.45
Indirect CW circulation rate	m^3/h	33160	24440

Refrigerant composition for Scenario 1.

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.2469	0.2333
Ethylene	0.2708	0.2810
Propane	0.1111	0.0500
i-Butane	0.0942	0.1201
n-butane	0.1534	0.1558
Nitrogen	0.1236	0.1548

Refrigerant composition for Scenario 2.

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.2274	0.2208
Ethylene	0.2982	0.3316
Propane	0.1443	0.0837
i-Butane	0.0856	0.1112
n-butane	0.1095	0.1084
Nitrogen	0.1351	0.1443

Refrigerant composition for Subcase A.

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.2470	0.2392
Ethylene	0.3064	0.2847
Propane	0.1118	0.0648
i-Butane	0.0961	0.1450
n-butane	0.0949	0.1169
Nitrogen	0.1438	0.1495

Refrigerant composition for Subcase B.

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.2203	0.2288
Ethylene	0.3326	0.2689
Propane	0.1296	0.1158
i-Butane	0.0588	0.1236
n-butane	0.1000	0.1497
Nitrogen	0.1587	0.1132

Refrigerant composition for Subcase C.

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.2479	0.1828
Ethylene	0.2839	0.2682
Propane	0.0889	0.0797
i-Butane	0.1389	0.1353
n-butane	0.0831	0.1428
Nitrogen	0.1574	0.1912

End flash composition for Scenario 1

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.9287	0.9287
Nitrogen	0.0713	0.0713

End flash composition for Scenario 2

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.7658	0.7658
Nitrogen	0.2342	0.2342

End flash composition for Subcase A and B

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.9287	0.9287
Nitrogen	0.0713	0.0713

End flash composition for Subcase C

Component	Mole fraction at design temperature	Mole fraction at high temperature
Methane	0.7651	0.7650
Nitrogen	0.2349	0.2350

Detailed NGL Extraction Simulations Results for the Subcases

Component	Subcase 1 [mole fraction]	Subcase 2 [mole fraction]	Subcase 3 [mole fraction]
Methane	0.9740	0.9741	0.8999
Ethane	0.0124	0.0124	0.0559
Propane	0.0044	0.0044	0.0144
i-Butane	0.0009	0.0008	0.0008
n-Butane	0.0010	0.0010	0.0009
Nitrogen	0.0072	0.0072	0.0281

Appendix B – Available Energy from End Flash and BOG and Fuel Gas Consumption

Available Energy for Scenario 1 and 2 at design temperature)

Property	Unit	Scenario 1	Scenario 2
End flash flowrate	Sm^3/s	12.872	23.810
End flash LHV	MJ/Sm^3	31.59	26.04
End flash energy flow	MJ/s	406.62	620.01
BOG liquid flowrate (0.15% per day)	m^3/d	375	375
BOG standard flowrate	Sm^3/s	2.877	2.970
BOG LHV	MJ/Sm^3	32.21	27.54
BOG energy flow	MJ/s	92.67	81.79
Total energy available from BOG and end flash	MJ/s	499.3	701.8

Available Energy for the Subcases at design temperature

Property	Unit	Subcase A	Subcase B	Subcase C
End flash flowrate	Sm^3/s	16.30	19.60	15.22
End flash LHV	MJ/Sm^3	31.59	31.59	26.02
End flash energy flow	MJ/s	514.92	619.16	396.02
BOG liquid flowrate (0.15% per day)	m^3/d	375	375	375
BOG standard flowrate	Sm^3/s	2.877	2.877	2.970
BOG LHV	MJ/Sm^3	32.21	32.21	27.55
BOG energy flow	MJ/s	92.67	92.67	81.82
Total energy available from BOG and end flash	MJ/s	607.59	711.83	477.84

Calculation of Fuel Gas Consumption Compared to Total Feed Gas Flow.

Property	Unit	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
BOG and end flash flow rate	kg/s	11.23	21.28	13.68	15.66	14.41
Fraction of flash and BOG for fuel	-	100%	29.47%	5.73%	100%	100%
LNG LHV	MJ/kg	49.65	-	-	49.65	48.85
Required additional fuel energy	MJ/s	132.3	-	-	9.22	141.08
Required additional gas flow for fuel	kg/s	2.665	-	-	0.186	2.888
Total fuel gas flow	kg/s	13.895	6.27	0.78	15.846	17.298
Total feed gas flow	kg/s	145.8	295.3	185.8	216.4	187.2
LHV feed gas	MJ/kg	48.91	41.93	48.91	48.91	41.93
Required energy flow	MJ/s	631.59	182.70	34.80	721.05	618.92
Feed gas energy flow	MJ/s	7131	12382	9087	10584	7849
Total gas consumed on mass basis	-	9.53%	2.12%	0.004%	7.32%	9.24%
Total gas consumed on energy basis	-	8.86%	1.48%	0.004%	6.81%	7.89%

Appendix C – Detailed Reliability Analysis

To calculate the reliability of several independent trains in section 5.4.1, the binomial probability distribution formula A.1 has been used. Here n is the number of trains, k is the number of trains in operation and p is the probability that each train is in operation.

$$P(N \geq k) = \sum_{k}^{n} \frac{n!}{k!(n-k)!} p^k (1-p)^{n-k} \qquad \text{C.1}$$

Formula B.2 has been used for systems configured in series, while formula B.3 has been used for parallel configurations with one or more units in redundancy.

$$R = R_x \times R_y$$

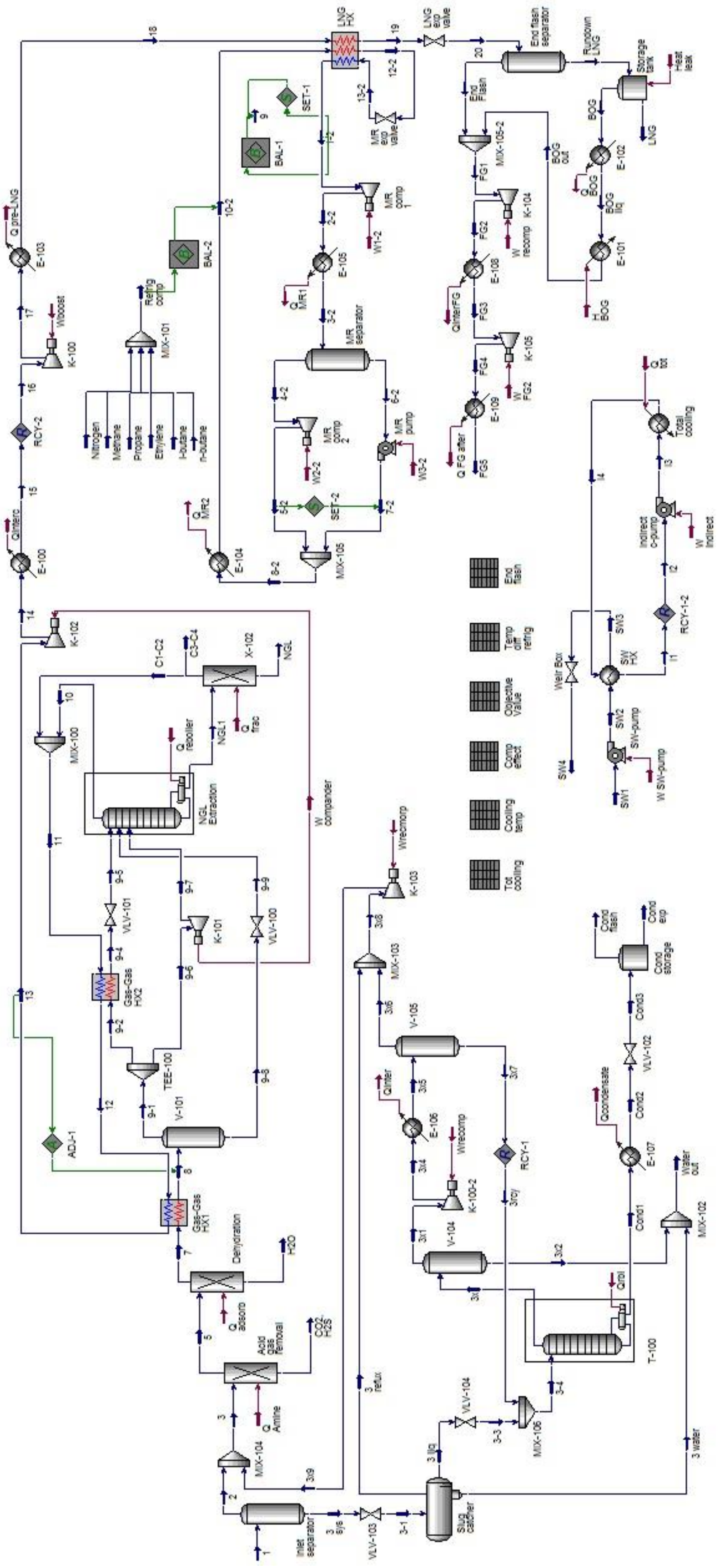
C.2

$$R = 1 - (1 - R)^2 \qquad \text{C.3}$$

Component	Reliability
Gas turbine driven compressor	0.9730
Electric driven compressor	0.9880
Frontend NGL, gas turbine	0.9637
Frontend NGL, electric	0.9781
Integrated NGL	0.9975
Compander	0.9900
PRICO train, GT drive	0.9730
PRICO train, electric driver	0.9880
Niche train, GT drive	0.9279
Niche train, electric drive	0.9567
Local GT Power Plant, N+1=2	0. 9993
Local GT Power Plant, N+1=3	0. 9976
Local GT Power Plant, N+1=6	0.9898
Local GT power plant, N+1=7	0.9860

Indirect cooling system (N+1)	1 (assumed)
Seawater pumps (N+1)	1 (assumed)
Hot oil system (N+1)	1 (assumed)

Appendix D – Complete HYSYS Model



Appendix E – Liquefaction Process Optimization

Based on previous work and statements made by Aspentech, the Hyprotech SQP optimizer is proved to be the most reliable optimizer in HYSYS (Rødstøl, 2014), and is therefore used in this study. Settings for the optimizer are based on the same study and are listed in the table below.

Optimizer setting	Value
Max Iterations	2000
Objective Scale Factor	0
Gradient Calculations	2-sided
Diagnostic Print Level	None
Objective Function	Minimize
Accuracy Tolerance	1.00E-8
Step Restriction	0.0001
Perturbation	0.001
Max Feasible Point	500

Briefly summarized, the optimization tools in HYSYS are based on an objective function, which is to be minimized or maximized based on process constraints and variables (Rødstøl, 2014). For the optimization in this study, the specific power of the liquefaction process is the objective function to be minimized, meaning the compressor power in the liquefaction process and booster compressor divided by the LNG production rate. The specific power is calculated in the spreadsheet "Objective value", and is calculated by the unit kWh/tonne LNG. Process variables and constraints are added in a derivative sheet in the model analysis tab in HYSYS.

Common process constraints in the simulation models are available compressor power in the liquefaction process and a minimum temperature pinch of 3°C in the LNG heat exchangers. To achieve valid evaluations of the simulation result, the process has been optimized with the same component configuration at high and design ambient temperature. Therefore, the pressure ratio in the liquefaction process compressors is added as a constraint at high ambient temperature simulations, where the constraint value is the pressure ratio of the compressors at design temperature. Other process constraints specific for the two liquefaction processes are described in the next subchapters.

The process variables in the simulation models are different in the two liquefaction processes, and is further described in the next subchapters. However, a common variable for PRICO and Niche (in Scenario 1) is the inlet flow of the natural gas. This is included to achieve variable LNG production rates at different configurations and ambient temperatures. As for ordinary process simulation, some values are sat as constants to achieve a functioning/iterating process, and thereby not included in the optimization. These values are also different in the two liquefaction processes, and are listed in the next subchapters.

PRICO Optimization

The optimization of the PRICO process is configured to achieve the maximum production rate with given available power in the MR compressors. The following optimization variables are used in the simulation:

- MR composition
- Low, medium and high pressure of the MR circuit
- Natural gas molar flow

The MR pressure levels are adjusted in stream 13-2 (low pressure), stream 3-2 (middle pressure) and stream 5-2 (high pressure) in Figure 25. The natural gas molar flow is adjusted in stream 16. The recycle function “RCY-2” transfers temperature, pressure and composition from the upstream process. This function is installed because the molar flow cannot be optimized upstream the liquefaction process. However, the molar flow stream 15 is iterated to match the optimized molar flow in stream 16 by adjusting the model feed stream manually (stream 1 in Figure 22).

An important optimization variable is the MR composition. To achieve the optimal composition, all the MR components are modelled as individual streams that enter the mixer “MIX-101” in Figure 25. The optimal composition is then defined in the stream “Refrig comp”, and a balance function transfers the composition to stream 10-2.

The following optimization constraints are used in the simulation

- Minimum 0.5°C superheating before the first stage compressor
- Temperature pinch of 3°C in the LNG heat exchanger
- Total available MR compressor power defined by the ambient temperature
- Pressure ratio in the MR compressors at high ambient temperature simulations

The total available compressor power is calculated in the spreadsheet “Comp effect” in Figure 25. At stream 1-2 before the first compressor stage, a balance function together with a virtual stream (stream 9) and a set function, is used to keep the stream 1-2 in vapour phase at all time. The molar flow and pressure of stream 1-2 is transferred to Stream 9 by the balance function and the set function. Stream 9 is also defined as pure vapour phase . The temperature difference between stream 1-2 and stream 9 is calculated in the spreadsheet “Temp diff refrig”, and used as a constraint in the optimization with a minimum of 0.5°C. This way, the temperature of stream 1-2 will always be superheated minimum 0.5°C, meaning that the gas entering the first stage compressor will always pure gas phase.

The following process values are constant in the optimization process:

- 60 bar pressure after booster compressor (stream 16)
- -155°C LNG outlet temperature of LNG HX (stream 18)
- LNG product specification of -163.2°C and 1.05 bar (stream 20)

Niche Optimization

Unlike the optimization of PRICO, different process variables and constraints are used when optimizing Niche in Scenario 1 and Scenario 2. In Scenario 1, the second and third stage compressors are powered by a direct drive Trent 60 gas turbine, and the production capacity depends mainly on the power output of this gas turbine. Thereby, the compressor power is defined as a constraint in the Scenario 1 simulations. In the Scenario 2 simulations, however, where compressors are driven by electrical motors, the compressor power can be adjusted in order to achieve the optimal production.

As a consequence of constrained compressor power in Scenario 1, the inlet molar flow is added as a variable in the scenario 1 simulations. All process variables and constraints for Scenario 1 and Scenario 2 are listed in the tables below. The stream names (in parentheses) corresponds to Figure 26.

Scenario 1	Scenario 2
Molar flow, NG inlet (NG inlet)	Molar flow, NG circuit (1-3)
Molar flow, NG circuit (1-3)	Low pressure, NG circuit (NGC6)
Low pressure, NG circuit (NGC6)	Temperature before NG split (NGC4)
Temperature before NG split (NGC4)	LNG temp. before subcooling (LNG2)
LNG temp. before subcooling (LNG2)	Low pressure, nitrogen circuit (NC3)
Low pressure, nitrogen circuit (NC3)	Temperature, nitrogen circuit (NC4)
Temperature, nitrogen circuit (NC4)	Middle pressure, nitrogen circuit (NG8)
Middle pressure, nitrogen circuit (NG8)	Middle pressure, NG circuit (NGC11)
Middle pressure, NG circuit (NGC11)	Flow split, NG circuit (TEE-100-2)
Flow split, NG circuit (TEE-100-2)	

Scenario 1	Scenario 2
Temperature pinch of 3°C in the LNG heat exchangers	Temperature pinch of 3°C in the LNG heat exchangers
Available compressor power defined by the ambient temperature	Pressure ratio in the compressors at high ambient temperature
Pressure ratio in the compressors at high ambient temperature	Max compander duty of 15 MW
Max compander duty of 15 MW	

The following process values are constant in the optimization process:

- 55.5 bar pressure after booster compressor (stream 16)
- 8000 kmol/h inlet molar flow (NG inlet) for Scenario 2 simulations
- 55.5 bar as second stage pressure in NG circuit (NGC11)
- 80 bar as high pressure in NG and N_2 circuit (NGC2 and NC10)
- -155°C LNG outlet temperature of LNG HX (stream 18)
- LNG product specification of -163.2°C and 1.05 bar (stream 19)
- Polytrophic efficiency of 80% in the companders

Note that the inlet molar flow is sat to 8000 kmol/h in the Scenario 2 simulations. This is due to 8000 kmol/h is an optimal inlet flow for Niche process when the compressor power can be easily adjusted (Pettersen, 2015), as with el motor drive in scenario 2. By experimental work in HYSYS, it was found that the inlet molar flow became very low (below 6000 kmol/h) if it were allowed to vary in the Scenario 2 simulations.

As described in Chapter 4.4.2, the Niche process is modelled as one of three parallel trains. As a result of this, the required booster compressor power is divided by 3 when calculating the objective function. The available compressor power of each process is calculated in the spreadsheet "Comp Work" in Figure 26.

By experimental work in with the optimization model in HYSYS, it was discovered that the polytrophic efficiency became very low in the turbine part of the companders with constant adiabatic efficiency of 75%. Subsequently, the constraint of 3°C in the subcooling heat exchanger (NC HX2) was not adhered. This lead to very high objective values in the optimisation, meaning that the specific power of the Niche process became too high in the simulation models (50-60% above PRICO), and the comparison basis for Niche became poor.

To solve this issue, the polytrophic efficiency was sat to 80% in the companders, which is a fair assumption for these components (Pettersen, 2015). The results of this was specific power of 30-40% above PRICO, which is a normal ratio between PRICO and Niche (Pettersen, 2015). Note that the polytrophic efficiency was sat only for the compander, meaning that the compressors still had an adiabatic efficiency of 75%. This was done to maintain as good comparison basis with PRICO as possible, since the MR compressors in PRICO also has an adiabatic efficiency of 75%.

Appendix F – Total Power Demand for Initial Configuration

Simulated Results at Design Temperature

Consumer	Scenario 1 [MW]	Scenario 2 [MW]	Subcase A [MW]	Subcase B [MW]	Subcase C [MW]
MR compressor 1	122.69	151.47	141.68	143.85	111.20
MR compressor 2	50.89	55.29	57.62	62.28	50.09
MR pump	0.49	1.17	0.84	0.94	0
Booster compressor	18.43	23.17	24.22	27.09	16.44
Indirect CW circulation pump	1.05	1.72	1.91	1.96	1.26
Seawater pump	-	3.40	-	-	-
Cooling tower pump	4.18	-	-	-	4.90
Inlet booster compressor for pipeline gas	15.46	-	19.70	-	-
CH_4/N_2 from BOG and end flash	5.39	9.11	6.52	7.41	6.27
Other Compressors	-	1.68	-	-	0.96
Total	218.58	247.01	252.49	243.53	191.12

Simulated Results at High Temperature

Consumer	Scenario 1 [MW]	Scenario 2 [MW]	Subcase A [MW]	Subcase B [MW]	Subcase C [MW]
MR compressor 1	101.80	152.60	138.60	103.50	93.54
MR compressor 2	42.24	54.49	58.83	49.94	43.68
MR pump	0.07	1.08	0.13	0	0
Booster compressor	13.87	22.98	21.12	16.20	11.44
Indirect CW circulation pump	0.83	1.72	1.71	1.27	1.49
Seawater pump	-	3.39	-	-	-

Cooling tower pump	3.25	-	-	-	3.65
Inlet booster compressor for pipeline gas	11.97	-	15.74	-	-
CH_4/N_2 from BOG and end flash	4.09	8.96	5.48	4.44	4.40
Other Compressors	-	1.65	-	-	0.68
Total	178.12	246.87	241.61	175.35	158.88

Rich gas cases scaled linearly from production rate from Melkøya numbers provided in the course TEP08 (Bjørge, 2014).

Consumer	Melkøya [MW]	Scenario 2 [MW]	Subcase C
Production [MTPA]	4.3	5.51	3.55
Amine pump	3.3	*4.2*	*2.7*
CO_2 reinjection compressor	12	*15.4*	*9.9*
CO_2 reinjection pump	0.7	*0.9*	*0.6*
Total	16.0	*20.5*	*13.2*

Total Power Demand at Design Temperature

	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
Total Simulated	218.58	247.01	252.49	243.54	191.12
Total Scaled	-	*20.5*	-	-	*13.2*
Air cooler fans	-	-	14.64	14.92	-
Added for hot oil pumps, fractionation, lighting etc.	*10*	*15*	*10*	*10*	*15*

Total power demand	228.58	282.51	277.13	268.46	219.32
Total electrical power demand	36.57	282.51	277.13	35.23	41.59

Total Power Demand at High Temperature

	Scenario 1	**Scenario 2**	**Subcase A**	**Subcase B**	**Subcase C**
Total Simulated	178.12	246.87	241.61	175.35	158.88
Total Scaled	-	*20.5*	-	-	*13.2*
Air cooler fans	-	-	14.64	14.92	-
Added for hot oil pumps, fractionation, lighting etc.	*10*	*15*	*10*	*10*	*15*
Total power demand	188.12	282.37	266.25	200.27	187.08
Total electrical power demand					

Appendix G – Total Cooling Demand for Initial Configuration

Simulation Results at Design Temperature

	Scenario 1 [MW]	Scenario 2 [MW]	Subcase A [MW]	Subcase B [MW]	Subcase C [MW]
MR cooler 1	155.80	232.80	205.40	225.80	145.50
MR cooler 2	135.40	137.80	139.20	145.60	126.20
Inlet Compressor Aftercooler	18.31	-	27.44	-	-
Condensate Reflux Cooler	0.02	1.05	0.02	0.03	0.63
Condensate Cooler	1.05	10.97	1.39	1.97	6.27
Compander Aftercooler	2.41	4.51	1.37	7.48	2.41
Booster Aftercooler	22.43	29.70	29.86	34.09	20.36
$CH4/N_2$ cooler	1.56	3.23	2.12	2.68	1.80
Total	**336.98**	**420.06**	**406.80**	**417.65**	**303.17**

Simulation Results at High Temperature

	Scenario 1 [MW]	Scenario 2 [MW]	Subcase A [MW]	Subcase B [MW]	Subcase C [MW]
MR cooler 1	117.90	238.30	165.30	114.60	103.80
MR cooler 2	112.70	129.20	155.90	133.70	108.80
Inlet Compressor Aftercooler	9.07	-	13.48	-	-
Condensate Reflux Cooler	0.01	1.03	0.02	0.01	0.33
Condensate Cooler	0.74	10.57	1.00	0.89	3.76
Compander Aftercooler	1.87	4.49	1.15	0.89	1.74
Booster Aftercooler	16.36	29.22	25.00	19.17	13.63

CH_4/N_2 cooler	0.85	3.06	1.25	1.01	1.05
Total	**259.5**	**415.9**	**363.1**	**270.3**	**233.1**

Lean Gas Cases at Design Temperature Scaled Linearly from Reference Model Provided by Supervisor

	Reference model	Scenario 1	Scenario 2 with lean gas	Subcase A	Subcase B
Production at design temperature [MTPA]	3.28	3.77	5.32	4.78	5.59
Acid gas removal [MW]	7.82	*8.99*	*12.68*	*11.40*	*13.33*
Fractionation [MW]	3.83	*4.40*	*6.21*	*5.81*	*6.53*
Total at design temperature [MW]	**-**	***13.39***	***18.89***	***17.21***	***19.86***
Total at high temperature [MW]	-	*9.38*	*18.78*	*13.75*	*10.45*

Rich Gas Cases Scaled Linearly from Melkøya Numbers Provided by Supervisor

Consumer	Unit	Melkøya	Scenario 1 with rich gas	Scenario 2	Subcase C
Production	MTPA	4.3	3.82	5.51	3.55
Waste water cooler	MW	0.1	-	*0.13*	-
MEG Clarification	MW	1.8	-	*2.31*	-
Lean MEG cooler	MW	0.28	-	*0.36*	-
CO_2 stripper condenser	MW	17.8	*15.81*	*22.81*	*14.7*
MDEA cooler	MW	36	*31.98*	*46.13*	*29.72*
CO_2 compressor 1 intercooler	MW	4.1	-	*5.25*	*3.38*
CO_2 compressor 2 intercooler	MW	3.4	-	*4.36*	*2.81*
CO_2 condenser	MW	6.8	-	*8.71*	*5.61*
CO_2 subcooler	MW	0.68	-	*0.87*	*0.56*
Treated Gas water precooler	MW	6.30	*5.60*	*8.07*	*5.20*
Light condensate cooler	MW	1.10	*0.98*	*1.41*	*0.91*
LPG cooler	MW	1.00	*0.89*	*1.28*	*0.83*
Depropaniser condenser	MW	7.70	*6.84*	*9.87*	*6.36*

Benzene removal condenser	MW	4.30	*3.8*	*5.51*	*3.55*
C4/C5 recycle cooler	MW	1.05	*0.93*	*1.35*	*0.87*
Tempered water cooler	MW	35.50	*31.53*	*45.49*	*29.31*
Total at design temperature	MW	127.9	***98.4***	***163.9***	***103.8***
Total at high temperature	MW	-		***160.93***	***66.96***

Total Simulated and Scaled Cooling Demand at Design Temperature.

	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
Simulated cooling duty	336.98	420.06	406.80	417.65	303.17
Scaled cooling duty	13.39	163.9	17.21	19.86	103.8
Total	350.37	583.96	424.01	437.51	406.97

Total Simulated and Scaled Cooling Demand at High Temperature.

	Scenario 1	Scenario 2	Subcase A	Subcase B	Subcase C
Simulated cooling duty	259.5	415.9	363.1	270.3	233.1
Scaled cooling duty	9.4	160.9	13.8	10.5	67.0
Total	268.9	576.8	376.9	280.8	300.1

Appendix H – Total Heating Duty at Design Conditions

Rich Gas Cases Scaled from Melkøya numbers provided by supervisor

Consumer	Melkøya	Scenario 1 with Rich gas	Scenario 2	Subcase C
Production rate [MTPA]	4.30	3.82	5.51	3.55
CO_2 removal reboiler duty [MW]	62.5	55.50	80.10	51.60
CO_2 regeneration heater	0.6	0.53	0.77	0.50
Dehydration regeneration heater	8.90	7.91	11.4	7.35
Rich MEG reclaimer package [MW]	5.75	0	7.35	0
MEG regeneration package [MW]	6.10	0	7.80	0
Rich MEG preheater [MW]	0.35	0	0.45	0
Inlet HC condensate preheater [MW]	3.40	3.00	4.35	2.81
HHC removal reboiler [MW]	7.00	6.20	8.95	5.78
Inlet gas preheater	2.20	-	2.82	-
Condensate stabilizer reboiler [MW]	9.00	8.0	11.5	7.43
Demethanizer reboiler [MW]	6.15	5.45	7.90	5.08
Deethanizer reboiler [MW]	7.50	6.65	9.60	6.19
Depropanizer reboiler [MW]	8.90	7.90	11.4	7.35
LM6000 anti-acing heater, per GT [MW]	2.1	-	-	-
Total [MW]	**-**	**101.1**	**164.4**	**94.1**

Lean Gas (Reference Provided by Supervisor)

	Reference	Scenario 1	Scenario 2 with lean gas	Subcase A	Subcase B
Production rate [MTPA]	3.14	3.77	5.46	4.78	5.59
Required heating duty [MW]	20.58	24.71	35.79	31.33	36.64
LM6000 anti-acing heater, per GT [MW]	2.1	-	-	-	12.6
Total	-	24.71	35.79	31.33	49.24

Appendix I – Detailed Pre-treatment Result for Gas Composition Alternative in Scenario 1 and 2

System	Scenario 1 rich gas [MW]	Scenario 2 lean gas [MW]
Power		
Booster Compressor	16.90	26.29
Added for loading pumps, hot oil pumps, fractionation, lighting etc.	15	10
Amine pump	2.93	-
Other compressors	1.03	-
Total power demand	**35.86**	**36.29**
Cooling		
Compander aftercooler	3.32	1.48
Booster aftercooler	20.95	32.88
Condensate cooling	7.44	1.89
Other fractionation and condensate	13.44	6.21
CO_2 removal	47.79	12.68
Other	37.13	-
Total cooling duty	**130.1**	**55.14**
Heating		
Fractionation and condensate	45.64	-
CO_2 removal	55.5	-
Total heating duty	**101.14**	**35.79**

Appendix J – Detailed NGL Extraction Results for Scenario 1 and Scenario 2

Scenario 1 NGL extraction results. Note that values smaller than 0.005 ppm are adjusted to 0.

Component	Upstream mole fraction	Downstream mole fraction
Methane	0.97196	0.97401
Ethane	0.01240	0.01242
Propane	0.00440	0.00441
i-Butane	0.00086	0.00086
n-Butane	0.00104	0.00104
Neopentane	0.00004	0.25 ppm
i-Pentane	0.00037	0.99 ppm
n-Pentane	0.00024	0.47 ppm
Hexane	0.00026	0.08 ppm
Heptane	0.00078	0.03 ppm
Octane	0.00003	0
Nonane	0.00005	0
C10+	0.00009	0
Benzene	32 ppm	0.01 ppm
E-Benzene	178 ppm	0
Toluene	28 ppm	0
Xylene	17 ppm	0
Nitrogen	0.00723	0.00724
Carbon Dioxide	0	0
Hydrogen Sulphide	0	0
Water	0	0

Scenario 2 NGL extraction results. Note that values smaller than 0.005 ppm are adjusted to 0.

	Upstream mole fraction	Downstream mole fraction (frontend)	Downstream mole fraction (integrated)
Methane	0.87743	0.90890	0.91183
Ethane	0.05447	0.05643	0.05661
Propane	0.02736	0.00590	0.00482
i-Butane	0.00337	0.00019	0.00028
n-Butane	0.00606	0.00019	0.00035
i-Pentane	0.00102	0.00001	0.00002
n-Pentane	0.00133	0.00001	0.00002
Hexane	0.00073	0.54 ppm	4.17 ppm
Heptane	0.00036	0.03 ppm	0.67 ppm
Octane	0.00013	0	0.08 ppm
Nonane	0.00002	0	0
C10+	0.00002	0	0
Benzene	144 ppm	0.13 ppm	0.98 ppm
Toluene	65 ppm	0.01 ppm	0.16 ppm
p-Xylene	16 ppm	0	0.01 ppm
Nitrogen	0.02740	0.02838	0.02606
Carbon Dioxide	0	0	0
Hydrogen Sulphide	0	0	0

Appendix K – CO_2 Emissions at Design Temperature

CO_2 emissions for Scenario 1, Subcase B and C with gas turbines as compressor driver and for power generation.

	Unit	Scenario 1	Subcase B	Subcase C
LHV methane	*MJ/kmol*	802.7	802.7	802.7
Total required fuel energy [MJ/s]	*MJ/s*	631.59	721.05	618.92
Required CH4 flow rate	*kmol/s*	0.7868	0.8983	0.7710
CO_2 emission rate from GT	*kmol/s*	0.7868	0.8983	0.7710
CO_2 vented	*kmol/h*	5.580	8.280	-
CO_2 emitted per year	*tonne/year*	**989047**	**1129804**	**967301**

CO_2 emissions for Scenario 2 and Subcase A with electrical drive, external power generation and gas fired burner.

Property	Unit	Scenario 2	Subcase A
LHV methane	*MJ/kmol*	802.7	802.7
Required energy from fuel	*MW*	182.7	34.8
Required methane flow	*kmol/s*	0.2276	0.0434
CO_2 emission rate from burner	*kmol/s*	0.2276	0.0434
CO_2 vented	*Kmol/h*	-	7.200
CO_2 emitted per year	*tonne/year*	**285539**	**56898**