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Energy Solution for Floating LNG Production System

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MASTER THESIS

for

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Spring 2012

Energy solutions for floating LNG production systems

Energiløsninger for flytende LNG-produksjonsanlegg

Background

Over the last decade there have been significant developments in concept solutions for offshore floating LNG production, where a complete LNG plant is placed on a FPSO (Floating Production Storage and Offloading System). Several field developments are now being planned based on floating LNG (FLNG), and the first investment decision has already been made by Shell for the Prelude field offshore Australia.

Any FLNG unit needs large refrigeration compressor drivers and also has several other rotating machinery units for compression of fuel gas, boil off gas, flash gas, amine pumping, thrusters, etc. In addition there are often large heat demands for amine regeneration, drier regeneration gas heating, fractionation column reboilers, etc. The energy (power and heat) supply system thus becomes an important part of the conceptual solution for FLNG, including the sourcing of fuel for power/heat generation. Fuel sources may include flash gas, end flash gas, regeneration gas, and boil off gas, but also liquids like LPG or condensate if these are not exported.

Drivers for refrigerant compressors may be gas turbines (e.g. aeroderivatives), steam turbines, electric motors, or combinations of these. Combined cycle system may have too large space requirement for a FPSO, but for large production capacity this may offer a highly efficient alternative. A power generation system will in any case be needed to drive various pumps, compressors, utility systems etc. Process heat may be distributed by a hot oil system or a steam system. Currently, the main compressor driver alternatives are gas turbines or steam turbines. The various system configurations differ in terms of efficiency, fuel utilization, production availability and environmental impact. Gas turbines have the disadvantage of given power output, while steam turbines can be manufactured for any size and capacity. There are also differences between the alternatives in terms of cooling water needs, deck space requirements, and limitations for fuel specifications (calorific value, pressure).

Objective

The objectives will be to analyze and compare various power and process heat solutions for floating LNG, with respect to thermal efficiency, fuel utilization, production availability, size/weight and deck space requirements, and fuel parameters.

The following items should be considered in the master thesis work:

1. Overview of solutions for FLNG energy systems, focusing on compressor driver, electric power and process heat supply, based on available literature and publications
2. Establishment of a few cases that can provide a basis for system analyses, including production capacity, feed gas specifications, process configuration, product requirements, and climatic conditions.
3. Establishment of process models for the main processing elements of the FLNG system, including inlet separation, gas treatment, liquefaction, heavy hydrocarbon extraction, power generation, process heat generation and fuel system.
4. Establishment of a framework for analysis and comparison, based on reasonable alternative system configurations for the cases defined above.
5. Analysis and comparison of energy system alternatives for FLNG
6. Conclusion of study with main findings, key features of the analyzed systems, and recommendations for further work.

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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.

The candidate is requested to initiate and keep close contact with his/her academic supervisor(s) throughout the working period. The candidate must follow the rules and regulations of NTNU as well as passive directions given by the Department of Energy and Process Engineering.

Risk assessment of the candidate's work shall be carried out according to the department's procedures. The risk assessment must be documented and included as part of the final report. Events related to the candidate's work adversely affecting the health, safety or security, must be documented and included as part of the final report.

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 (http://daim.idi.ntnu.no/). 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. The final reports, with summery and all other material and documents have to be given to the supervisor in digital format on a CD.

Department of Energy and Process Engineering, 16. January 2012

Olav Bolland
Department Manager

Jostein Pettersen
Academic Supervisor

Preface

This report, Energy Solution for Floating LNG Production System, is written as a master thesis at the Norwegian University of Science and Technology (NTNU). This thesis comprises 30 out 30 credits in the 10th semester for the 5-year Master of Science (McS) program, and the work was carried out at the Department of Energy and Process Engineering at the faculty of Engineering Science and Technology, with Prof II Jostein Pettersen as supervisor.

Trondheim, June 11, 2012

Magnus Nordahl Andersen

Abstract

This report considers different energy solutions for a floating LNG production vessel. The two alternatives analyzed are gas turbine and steam turbine. In addition to being stand-alone alternatives they are the basis for other alternatives, such as electric drive and combined cycle. Light side studies have been performed on the two latter ones.

A LNG process model has been built in Aspen HYSYS, and from this several cases has been run. There are 3 levels of different parameters that have been run: 1) Energy system, 2) Feed Gas Composition, and 3) Liquefaction process. As mentioned the two energy systems are GT and ST. Three different feed gas compositions have been analyzed: 1) Low content of both CO_2 and N_2 (0.5 % and 1 % respectively), 2) High content of CO_2 (9.5 %), and 3) High content of both CO_2 and N_2 (9.5 % and 3 %). The liquefaction processes analyzed are two of the most promising for a floating LNG application: Dual mixed refrigerant and dual N_2 expander.

The feed gas compositions was chosen to give a wide area of applications for the results, and to give illustration on how the two different energy solution would respond to changing feed gas composition. The DMR liquefaction process was chosen mainly because this is the one being implemented in Shell Prelude FLNG. Being the most proposed solution for offshore application the dual N_2 expander was a natural alternative to the DMR.

The analysis show a clear advantage for gas turbine and DMR process, when exclusively looking at efficiency. However; as the objective of the study states, important factors such as safety, vessel motion sensitivity, reliability, availability is also to be considered. The results show 245 kWh/ton LNG energy consumption with the DMR liquefaction process, whereas the dual N_2 expander requires 424 kWh/ton LNG; over 70 % increase. However; the side- cases run in this report show advantages to the N_2 dual expander in safety, weight/space requirements and ease of start-up and shut down.

The ST/N_2 has fuel gas consumption 4% higher than the GT/N_2. On the basis of the results in this report and other studies performed on FLNG a selection of the ST/N_2 setup will be favorable as long as there is a high CO_2 content in the feed. With low CO_2 content, hence heat demand, the advantage of the ST is smaller thanks to lower heat recovery demand.

Sammendrag

Denne rapporten vurderer ulike energiløsninger for et flytende LNG-produksjonsskip. De to alternativene som er analysert er gassturbin og dampturbin. I tillegg til å være frittstående alternativer er de grunnlaget for andre alternativer, for eksempel elektrisk drift og combined cycle. Kun enkle analyser har blitt utført på de to sistnevnte seg.

En LNG prosess modell har blitt bygget i Aspen HYSYS, og fra denne casen har flere andre caser har blitt kjørt. Det er 3 nivåer av forskjellige parametere som har blitt kjørt: 1) energisystemet, 2) fødegass komposisjon, og 3) flytendegjøringsprosessen. Som nevnt er de to energisystemer GT og ST. Tre forskjellige fødegasskomposisjoner har blitt analysert: 1) lavt innhold av både CO_2 og N_2 (0,5 % og 1 % henholdsvis), 2) Høyt innhold av CO_2 (9,5 %), og 3) Høyt innhold av både CO_2 og N_2 (9,5 % og 3 %). Flytendegjøring prosessene analysert er to av de mest lovende for en FLNG: Dual mixed refrigerant (DMR) og dual N_2 ekspander.

Fødegass komposisjonene ble valgt for å gi et bredt område for bruk av resultatene, og å gi en illustrasjon på hvordan de to ulike energiløsningene ville reagere på skiftende fødegass komposisjon. DMR LNG-prosessen ble valgt hovedsakelig fordi dette er den som blir implementert i Shell Prelude FLNG. Ved å være den mest foreslåtte løsningen for offshore anvendelse var dual N_2 ekspander et naturlig alternativ til DMR.

Analysen viser en klar fordel for gassturbin og DMR prosess, da utelukkende ser på effektivitet. Men, som målet med denne studien stater, er viktige faktorer som sikkerhet, fartøybevegelsene følsomhet, pålitelighet, tilgjengelighet også vurderes. Resultatene viser 245 kWh / tonn LNG energiforbruk med DMR LNG-prosessen, mens dual N_2 ekspander krever 424 kWh / tonn LNG; over 70 % økning. Side-casene som kjøres i denne rapporten, viser derimot fordeler ved N_2 dual ekspander både når det gjelder sikkerhet, vekt / plassbehov og enkel oppstart og stans.

ST/N_2 har brenngass forbruk 4 % høyere enn GT/N_2. På bakgrunn av resultatene i denne rapporten og andre studier utført på FLNG vil konklusjonen være at ST/N_2 vil være gunstig så lenge det er et høyt CO_2-innhold i fødestrømmen. Med lavt CO_2-innhold, fordelen til ST mindre takket være lavere varmegjenvinningsbehov.

Acknowledgements

I would like to express my gratitude to my supervisor, Jostein Pettersen, for his help and guidance through this semester. He has always been available for me, and I especially appreciate that he has taken his time for biweekly meetings. Through the meetings I have acquired new knowledge in several engineering areas, and gotten an insight in a scientist's way of working.

Thanks to the students I have shared office with for all the discussions, coffee breaks and the memorable tradition; cake Friday.

Nomenclature

Latin Symbols

T	temperature (K; ^{0}C)
Q	heat duty (W)
W	power (W)
h	enthalpy (kJ/kg)
s	entropy (kJ/kg K)
T_0	ambient temperature (K; ^{0}C)
C_p	heat capacity (kJ/kg K)
$\dot{m}$	mass flow (kg/s)
Δh_{fg}	heat of evaporation (kJ/kg)
ρ	density (kg/m^3)
V	velocity (m/s)
A	area (m^2)

Abbreviations

BOG	Boil off Gas
LNG	Liquefied Natural Gas
FLNG	Floating Liquefied Natural Gas
FPSO	Floating Production Storage and Offloading
FG	Fuel Gas
MTPA	Million tons per annum
PPM	Parts per Million
LHV	Lower Heating Value
HHV	Higher Heating Value
GT	Gas Turbine
ST	Steam Turbine
HR	Heat recovery
M-GT	Mechanical Gas Turbine
E-GT	Electricity Generating Gas Turbine

Table of Contents

1 Introduction

1.1 Motivation

The last decades several land-based LNG plants have been built and set into operation, utilizing offshore gas fields. The location of the plant is typically the closest piece of land to the offshore reservoir, hence minimizing the need for costly offshore piping. For such a plant to be economically profitable the gas field has to be above a certain size. As the world's energy demand increase, utilization of small and remote gas fields are getting more and more interesting.

One solution to utilization of smaller offshore gas fields is floating LNG. The idea of putting the whole LNG plant on a vessel is not new, but first from the mid-1990s substantial experimental testing on FLNG was performed. [1] Eliminating the need for long pipelines and having a mobile plant able to operate on several fields during its lifespan, are two of the main benefits with and floating LNG (FLNG). The FLNG makes it profitable to monetize smaller or remote fields of non-associated gas. However; as with every new technology several challenges arises, some of which just recently has been overcome. This led to the first investment decision for an FLNG solution (3.5 MTPA) in 2011. Shell's FLNG will be operating on the Prelude field offshore Australia, and move the technology from the drawing board to reality.

A floating LNG has a different set of requirements for the energy solution than an equivalent land-based plant. As the FLNG will typically operate at remote locations reliability and availability is a key factors. In addition; a FLNG in more exposed to harsh weather, and therefore the ease of start-up and shut down of the plant is a more important parameter than with an onshore plant. Last but not least the FLNG demands for a more versatile plant able to operate with several different feed gas compositions.

This set of requirements, and new key factors with an FLNG plant ask for a different energy solution analysis than an onshore plant.

1.2 Objective

In the work with this report the focus is the energy system for a floating LNG production vessel, with respect to thermal efficiency, fuel utilization, production availability, and size/weight and deck space requirements.

The objective of this study is divided into three parts:

1) Get an overview of solutions for FLNG, focusing on compressor driver, electric power and process heat supply.
2) Establish simulation cases for different driver configuration, feed gas specification and liquefaction processes.

3) Run the simulations, analyze the results and present a conclusion of the main findings, key features of the analyzed systems, and recommendations for further work.

1.3 Outline

Chapter 2 contains background and overview over different solutions for the energy system for floating LNG. Especially chapter 2.3 should be read.

In Chapter 3 the design basis containing all the input data and assumptions made are presented. All the input data in this chapter is common for all the different cases in this report, hence this chapter is essential reading for proper understanding of the results presented later in the report. You will find a block diagram representative for all the cases in Figure 3.1.

Figure 4.1 and Figure 4.2 in chapter 4 contains an overview of the different cases and the system of which the cases is organized. Further this chapter present case specific assumptions and input values.

In chapter 4.2.4 the results and a thorough discussion is presented. The first part presents the table of results from the base cases and a following elaboration of the table. The two other feed gas specifications are discussed relative to the base cases; hence the elaboration of the base case numbers is essential for proper understanding of the results in this report.

Most of the tables and graphs from the simulation models are put in the appendix.

2 Theory and Background of FLNG

2.1 Historical Progress

The base-load LNG industry now has over 40 years of history starting with permanent operations. A floating LNG facility however; does not have any current permanent operations. A lot of concept studies have been conducted through the last two decades, with the main motivation being that FLNG offer a route to unlocking natural gas resources stranded because of their remote location, complex piping or environmental issues.

2.1.1 Challenges

Here are some of the challenges of FLNG, which is all included to show some of the current challenges. They are all beyond the scope of this report, but is included to form a wider background.

- Cyclone survival is an important factor I the development of FLNG. The vessel has to withstand met oceans conditions, especially considering the turret mooring. How the mooring the LNG carrier to the FLNG vessel is to be performed is also a challenge. Side by side or Stern to bow is the two different principals for the offloading. [1]
- The industry has already developed flexible risers for the inlet fluid transport from the subsea surface to the floating facility. Recently, the transfer of LNG through flex hoses has been tested and additional development is ongoing in this area for further optimization and cost reduction. [2]
- "Recently, the transfer of LNG through flex hoses has been tested and additional development is ongoing in this area for further optimization and cost reduction. A specific design for floating storage tanks associated with LNG liquefaction facilities has recently been developed. The LNG storage tank is designed to help reduce the overall cost in this area, as well as addressing some of the safety issues associated with liquid sloshing in the marine atmosphere with partially filled LNG tanks." [1]
- "Achieving higher capacities in the floating liquefaction plants reaching up to 3.0 MTPA are dependent on the deck space driven by the hull design, which is affected by the hydrodynamics of the offshore location." [2]

2.1.2 Shell Prelude

In 2011 Shell took the final investment decision on their Prelude FLNG project in Australia, after spending 1.6 million working hours on Front End Engineering and Design process. Shell claim the Prelude FLNG will produce 3.6 MTPA of LNG once operational in 2017. Once built the floating facility will be 468 m long, 74m wide and displacing 600 000 tons of water. The liquefaction unit uses a single train Shell Dual Mixed Refrigerant (DMR) and the facility has LNG storage capacity of 220 000 m^3.

Shell has weighted robustness end reliability more than efficiency, when choosing a steam turbine over gas turbine. [3]

Shell reports the overall efficiency is getting better thanks to no long piping. In other word; this is not unique for this system, hence equally relevant for a gas turbine setup and N_2 dual expander liquefaction.

The Shell analysis clearly is relevant for the plant being simulated in this report. However; the high CO_2 and condensate argument is not to be added too much importance in a general study considering several feed composition. If the reboiler duty is considerably lowered, in other words less CO_2 and/or condensate, the steam turbine will have a lot of potential heat recovery not needed. In that case a gas turbine setup may be more feasible.

2.2 Benefits and Requirements

2.2.1 Benefits

- Avoid flaring or reinjection of associated gas
- Avoid Pipelines
- Monetize smaller or remote fields of non-associated gas.
- Reduce exposure to public and increase security of facilities.

2.2.2 Requirements

- Ease of shutdown/startup because of bad weather and harsh environment
- Flexibility to different gas compositions from different fields
- Vessel motion. Stresses in partly loaded tanks relative motion carrier to production facility
- High degree of safety given the location on a vessel: i.e. large inventories of hydrocarbon refrigerants

2.3 Energy Systems

2.3.1 Drivers for Rotating Equipment and Electric Power Generation

There are two categories under power production. One is the mechanical power driving the liquefaction compressors, which in this paper is called M-GT, and the other is electrical power consumers driven by a GT generating electricity, called E-GT in this paper. In Table 2.1 the different consumers are listed. With a liquefaction plant utilizing gas turbines and producing 3.3 MTPA a typical setup ratio between the two categories will be 4:2. In other words; four compressors driving the whole liquefaction process through Shaft Power and 1 generating electric power and one in backup

Table 2.1 - Drivers for rotating equipment and Power Generation

Drivers for Rotating Equipment	Electric Power Generation
Precooling compressors	BOG compressor
Liquefaction compressors	Flash compressor
Sub cooling compressors	Utility consumers

2.3.1.1 Gas Turbine

Industrial and aero derivative are the two main categories of gas turbines. For offshore use the industrial turbine is regarded as less suitable than the aero derivative given the following arguments:

- None of the small to medium industrial GT has been qualified for offshore application, and a qualification is not in sight in near future. [4]
- Easier maintenance on aero derivative GT than industrial GT
- Lower weight/space requirements than industrial GT
- Qualified for offshore use.
- No need for large starter motors and associated electrical equipment.
- Several models can be operated at variable speeds (e.g. LM6000)
- Replacement of modules of the GT package is possible onboard and repair works can be carried out onshore

Industrial GT is not considered in this study given the arguments listed above, hence the aero derivative is the GT used in this report. To point out the two key parameters weighted in this selection: The design is optimized with respect to large power/weight ratio and with a multi shaft setup the aeroderivative only requires a small starting motor, which again ease start up and shut down.

The LM6000 is chosen to be the GT referred to in this report. There has not been a detailed analysis performed in accordance to this choice; however the LM 6000 is well proven (i.e. at Statoil's Melkøya plant) and has suitable specifications; an ISO rated power of 43MW and an efficiency of 45 % and low power/weight ratio.

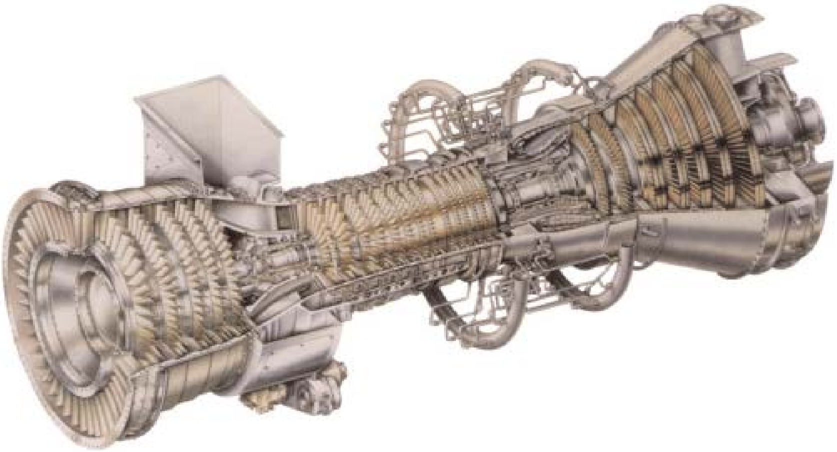

Figure 2.1 – LM6000 Gas Turbine

As reported by de producer General Electrics; The status of the LM6000 program, as of February 2000, includes [5]:

- 300 units produced since introduction in 1991
- 208 units in commercial operation
- 12 month rolling average engine availability = 96.8%

2.3.1.2 Steam Turbine

Whereas a LM6000 gas turbine have a power production of 43 MW ISO, a steam turbine setup can be tailor made for the given power demand. To cover the power demand a GT setup requires several trains and strings, which reduce the availability. A steam turbine will require lower fuel gas pressure levels (in this report set to 10 bar vs. 50 bar in the GT), and be more suited for easy integrating with a steam waste heat distribution system. This is because part of the steam can be extracted from the turbine at the desirable temperature, providing heat to the heat consumers; hence the heat recovery potential is quite large. This will favor a heat demanding natural gas processing, which in essence mean high concentration of CO_2 and high condensate content. If the large heat demand for CO_2 removal is not present, there will be a large heat surplus from the power generation, hence low overall energy utilization.

Some disadvantages will also be present with a steam turbine setup. The main disadvantages being bulky and space demanding layout, and lower efficiency (~25%). than the GT. The ST needs fresh water for boilers, hence a ST setup must be able to generate their own fresh water onboard the vessel. Fresh Water Generators are installed to convert the seawater from the sea to freshwater. This adds to the already large space and weight requirements.

However in the paper presented by Shell at the International Technology Conference held in Bangkok 7-9 February this is the considerations made treating driver choice for the Prelude FLNG:

- "High CO_2 and condensate of feed gas requires a reboiler duty of about 200 MW. The medium heat transfer fluid temperature not being higher than making steam or warm water an excellent fluid option yields a lower size and weight of the heat exchangers and topside piping choosing a steam turbine."
- "The FLNG facility is located directly above the well, which implies that any disturbance in the upstream are felt immediately on the processing facility, and disturbances on the topside can lead to flaring. This "close coupling" represents an important difference with onshore plants where the trunk lines acts as a buffer (varying pressure), decoupling the upstream facilities from the LNG plant. The flow assurance method requires depressurizing the upstream flow lines and a length re-start period, upon a disturbance of the FLNG. The reliability in single train steam turbine topside secures a high availability."
- "Being offshore the maintenance and service works are more expensive as compared to onshore. This fact favors units with good track record to make the systems simple and robust. Additionally our experience shows that the occurrence of damage and leak in cryogenic equipment is often connected to the number of plant trips."[3]

In summary; after considering steam turbines, gas turbines and electrical drivers, steam turbine, on the basis of the three items mentioned above, steam turbine was found to be the best option. With the number of trips ratio between steam turbine and gas turbine being 1:5-10, a steam turbine is considered a much more reliable system.

2.3.2 Liquefaction for offshore use

As stated in Barcley et als paper: Two processes that have been previously identified as offering potential for offshore liquefaction are nitrogen expander cycles and dual mixed refrigerant cycles. [6] This is supported by the requirements of a FLNG with its key factors that impose a simple and compact liquefaction setup with key factors being; flexible, adaptable to natural gas of different components; fast start-up/stop; safe, reliable and insensitive to the motion of FLNG.[7]

In this report Dual Mixed Refrigerant and N_2 dual expanders are compared in the simulation, because both hold the key factors mention in the above paragraph. All this factors are also supported by the conclusion in Total's liquefaction selection report. [8] In addition Total concluded that the availability, size and weight will be more or less the same with a small favor of a N_2 setup.

2.3.2.1 Dual Mixed Refrigerant

The development of the dual mixed refrigerant (DMR) is driven by the need for a precooling stage that covers a wider range of temperature than pure propane is able to. Shells DMR has been applied in the Sakhalin LNG Plant in Russia. While the

C3MR consist of a pure propane precooling, which limits the temperature out of the precooling to about -35^{0}C, the DMR precooling consists of a mix of ethane and propane.[9] Hence the DMR are able to cool further down in the precooling, given the NBP of ethane being -88^{0}C, which again gives a more equal load between the precooling and sub cooling. Two mixed refrigerant cycles also has the advantage of adding more flexibility, because the mixing can be optimized for different operating conditions, which will be of interest for an FLNG. A DMR cycle has an exergy efficiency of about 45%, which gives it a clear advantage over the N_2 dual expander.

2.3.2.2 N_2 Dual Expander

The main advantages with this kind of liquefaction are that the equipment count for a single train is low, the configuration is simple and there is no phase change of the refrigerant.

Cost-effective and efficient liquefaction plant designs have been based on generating refrigeration by gas compression and subsequent work expansion in turbo-expanders. This provides:

- Inherent safety by avoiding the need for any hazardous liquid hydrocarbon refrigerants
- Insensitivity to vessel movement, as the refrigerant is always in the gaseous phase
- Simplicity of operation and flexibility to feed gas changes
- Ease of start-up and shutdown
- A small number of equipment items, small area and low weight
- Ease of modularization and fabrication
- Use of conventional well-proven equipment that maximizes the opportunity for competition among suppliers and means lower cost equipment.

The advantage of the nitrogen refrigeration cycle is its process simplification and ease of operation. [1] [2]

However; a considerable loss in exergy efficiency, hence increased power demand, for a given production rate is a clear disadvantage when compared to DMR.

In their paper Wood et al [1], conclude with almost identical argument for the N_2 dual expander as mentioned above, and in addition point out that the limiting factor in a 3 MTPA plant is power requirements.

The Kollsnes II, utilizing dual N_2 expander process, has an energy demand reported to 510 kWh/ton LNG.[10]

2.3.2.3 N_2 vs. DMR comparison

Safety:

The N_2 dual expander has a clear advantage when it comes to safety; the reason being no flammable inventories or flare requirements.

Compactness:

With the N_2 dual expander has no need for refrigerant storage, because it is gaseous. The DMR cycle require large HC storage, although not as large as the C3MR. Also, a larger safety distance is needed in the use of the DMR, and two phase flow equipment will be heavier and more complex.

Exergy Efficiency:

The gliding temperature profile in a DMR process results in a better match to the liquefaction of natural gas. The different boiling point of the components in the fluid results in this gliding temperature profile. The composition of the mixed refrigerant has to be adjusted to match the composition of the natural gas to be liquefied.

The N_2 dual expander operating without evaporation has almost constant specific heat; hence the way to cover the temperature profile of natural gas is to vary the N_2 flow rate.

Operation:

Dual N_2 expanders will require less complex operation due to no inert refrigeration composition and low equipment count. In addition; a quick start-up time is an advantage of N_2.

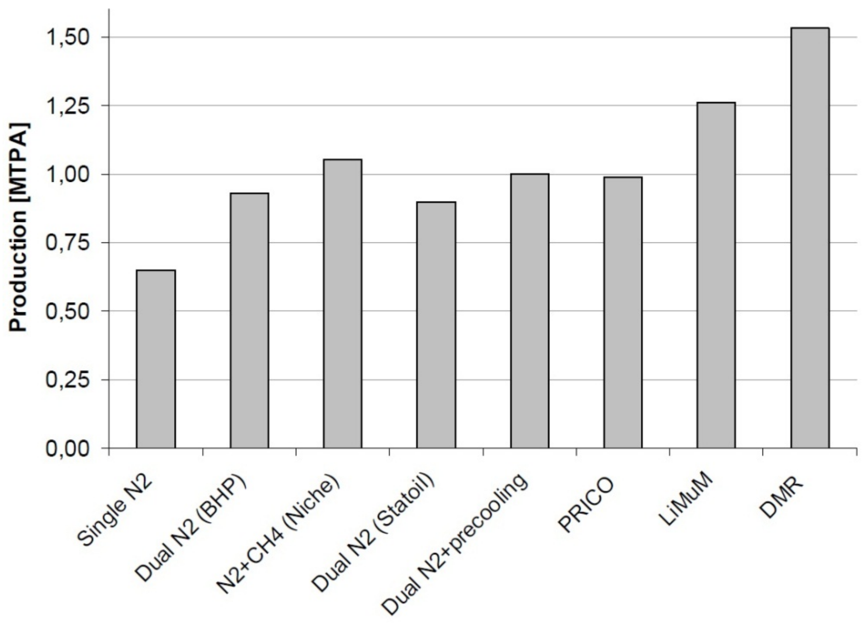

Figure 2.2 – Comparison Liquefaction Principles train capacity [10]

In Figure 2.2 the train capacity of different liquefaction technologies is compared. As can be seen a DMR has a train capacity about 50% higher than a dual N_2 (BHP). This has consequences for the production capacity chosen in this this report for the two different technologies. See chapter 4.2.4 for more information

In summary; the efficiency of the DMR is superior, but on every other factor the dual N_2 expander has an advantage over the DMR.

3 Design Basis

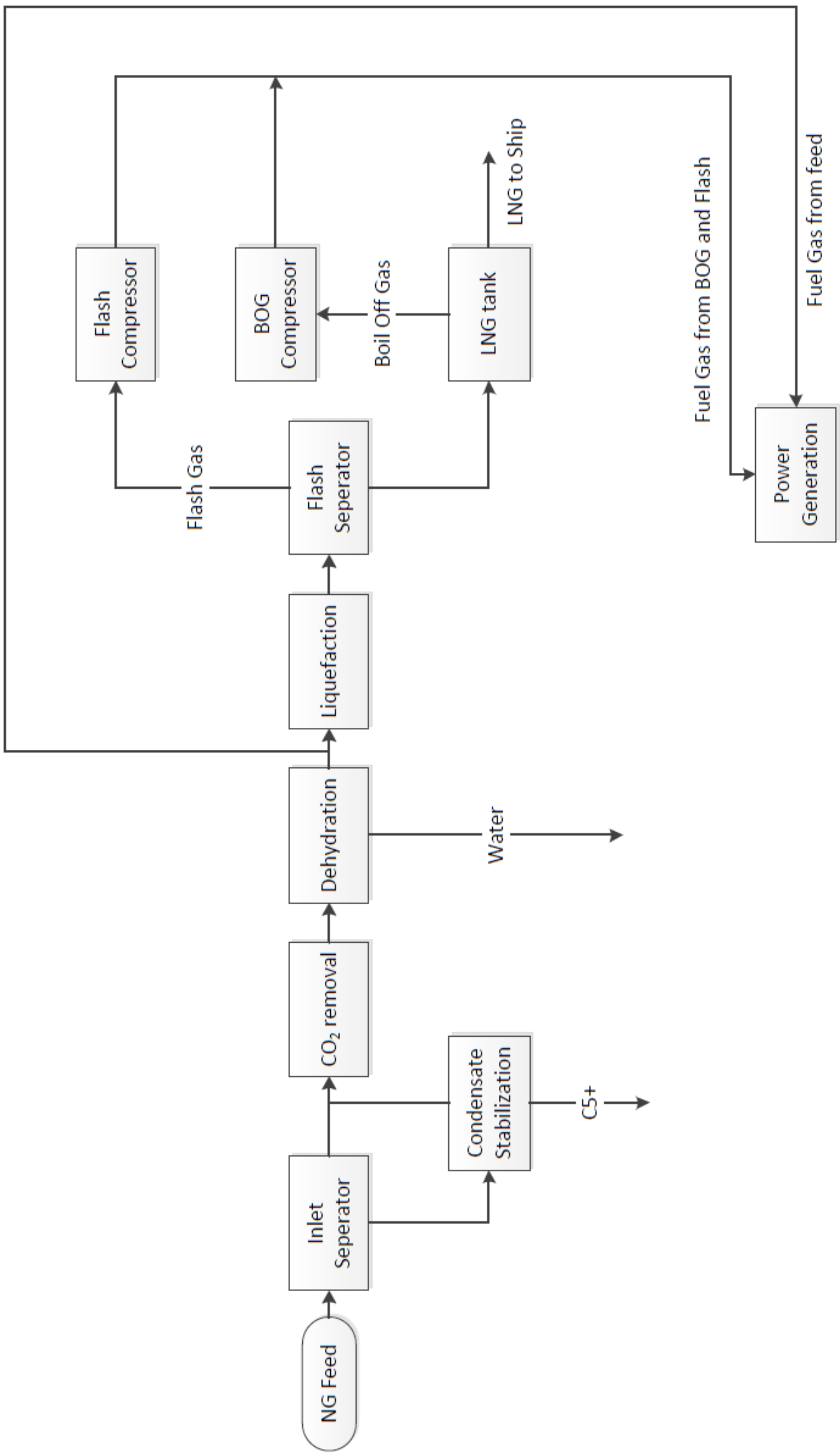

Figure 3.1 – Block Diagram

3.1 Input Data and Assumptions

Figure 3.1 show a block diagram of the simulation model. There are several parameters changed in between the different cases analyzed in this report. However; the setup of the block diagram is the same for all cases. In this chapter the setup and all the assumptions that is common for all the cases are described, hence all that is included in this chapter applies as a basis for every case. In chapter 4 the different cases and their specific assumptions are described. Note that all described in this chapter still applies.

3.1.1 Sea Water Temperature

Sea water temperature 15^{0}C (288K), pressure 1.03 bar

3.1.2 Pressure losses

Over heat exchangers and splitters there has been assumed 0.5 bar pressure loss. The dehydration splitter is set with a 3 bar pressure drop, whereas the total pressure loss over the liquefaction part is set to 8 bars.

3.1.3 Feed gas

The natural gas entering the plant is assumed to have a pressure of 70 bars and a temperature of 353 K. These conditions are based on a relatively short pipeline travel from the reservoir to the FLNG intake; hence a low temperature- and pressure drop in respect to general well parameters (~ 90 bars and ~373K). [9] The 3 different feed gas compositions used in this report are listed in chapter 0.

3.1.4 Feed Valve

This valve is put into the flow diagram to be able to adjust the intake pressure independent of the reservoir pressure. The pressure is adjusted to assure the required degree of HHC removal in the inlet separator. The maximum limit of 0,1 mole% of C5+ into the liquefaction plant is the specification that has to be met. Given that the FLNG will be connected to different reservoirs, this is a necessity.

3.1.5 Inlet separator end Condensate Stabilization

The inlet separator splits the liquid from the vapor and by adjusting the inlet pressure one can achieve the desired process of splitting HHC from the lighter one, because of different boiling point. The condensate from the inlet separator is sent into condensate stabilization, where a fraction of the lighter HC is sent to the sweetening, whereas the C5+ product is not processed any further in the simulation. This separator works in exactly the same way as the inlet separator, by splitting liquid and vapor. The lack of any further C5+ simulation is chosen given the scope of this paper. The C5+ content is the same in all the different cases simulated in this paper, hence there is set a fixed heat demand for this process in the simulation models.

In the simulation both of the processes are modeled with a component splitter to ease the separation. The mole% fractions that the splitter is set to are listed in Table 3.1. A liquid/vapor separator unit has been inserted to the model; and by using the phase splits from the separator and switching the separator with component splitter the mole fractions has been set constant. The scope of this paper is the energy solutions and not detailing the different separators for the FLNG, hence the model has been simplified with component splitters instead of liquid/vapor separator units. The heat required in relation to the four component splitters has been set based on numbers from other comparative simulations. [11]

The only specification that the C5+ stream has to meet is a vapor pressure (Reid 37,8C) below 10 psi. This is to assure no gauge pressure, and is one of the product requirements for the condensate.

An energy stream has been attached to the Condensate stabilization to simulate the heat demand of this process. This has no practical use in the model but describe the process in a better way. To compensate for the temperature increase of the stream a sea water cooler has been attached. The pressure loss is set to zero, because this cooler would not be incorporated in an actual plant.

Table 3.1 - Inlet and Condensate Split Fractions.

Numbers describing the streams going to the CO_2 removal in the block diagram. In mole %

	Inlet Separator	Condensate Stabilization
C1	95	99,9
CO2	95	100
N2	100	100
H20	95	100
C2	95	98,8
C3	95	99
iC4	95	98
nC4	95	98
iC5	12,5	1
nC5	12,5	1
C6	10	0,5
C7	10	1
C8	0,1	1
C9	0,5	1
C10	0,1	1
C11	0,2	1
C12	0,2	1

3.1.6 CO2 removal and dehydration

Once again the processes are both simulated with a component splitter. The reason for the simplification is the same as for the inlet separator and condensate stabilization; simplifying process beyond the scope of this report. The energy demand for the CO_2 removal was given for the 9.5 mole% feed composition, and set to 100 MW. See Table F.11. The basis for this number is a comparable (3,3MTPA and same CO_2 content in feed), but more complex simulation performed for Statoil. However; the scope of that simulation is not energy solution.

The content of water is set constant in all the cases, hence the heat demand for the dehydration will be constant.

As for the condensate stabilization; an energy stream is attached to both of the splitters to simulate the heat demand in the process, and again a sea water cooler is attached to compensate for the heat added in the simulation. There is set no pressure loss over the seawater cooler. This seawater cooler is not needed in an actual FLNG. Given that the requirements for liquefaction is in the range of ppm for both CO_2 and H_2O there has been set a 100% removal of both.

3.1.7 Fuel gas split

A stream splitter that splits off a part of the dry natural gas to fuel gas (FG) is located downstream of the dehydration. The split fraction is manually adjusted to cover the power demand and it is adjusted in the range from 0-10 % of the energy content in the stream.

3.1.8 Liquefaction

The dry natural gas enters the liquefaction at the conditions given in Table 3.2. These values are set based on common practice in today's LNG plant, but will of course differ from plant to plant. In this report and simulations all the values are fixed for the ease of simulation. The whole liquefaction is modeled as one single heat exchanger. The reason for such a crude simplification is the fact that even this model gives all the numbers needed for the exergy analysis performed later in this report. There is no need to build a complex liquefaction model, and even more; a more complex model would have required time for modeling both a dual mixed refrigerant (DMR) and a dual expander N_2. In this simple model the enthalpy, entropy and a given exergy efficiency is all needed to perform the energy analysis over the liquefaction part. H_2O

Table 3.2 - Dry Gas Conditions

Parameter	Value
Pressure	65,5 bar
Temperature	45 C
C5+ content	0,1 mole %
H_2O content	0 mole%
CO_2 content	0 mole%

3.1.9 Flash- Valve and Separator

The flash valve is contributing to the final refrigeration by the Joule Thompson effect, whereas the flash separator is flashing out the N_2 content to give the LNG product a mole fraction of N_2 below 1 %. The separator is a vapor/liquid separator, hence utilizing different boiling point of the stream content. In the feed compositions with 1 % N_2 fraction a sub cooling in the liquefaction could be favorable to avoid flashing out unnecessary amount of N_2 and also hydrocarbons. This will however require large amount of refrigeration power. Numbers acquired from the simulation model is in the range of 5 MW/K for both MR and N_2 refrigeration. In other words; there will be a tradeoff between the flash gas and liquefaction power.

An adjuster is used to control the temperature out of the liquefaction, to match this to the desirable temperature into the flash separator.

3.1.10 LNG tank

With production capacity of 3.3 MTPA and 330 operational days, the production per day equals 10000 tons. The tank is designed to have a capacity of 8 days of production. This equals a tank volume of about 180 000 m^3.

For the ease of simulation the boil off gas rate has been set constant at 0.15% of half the tank volume per day, 90000 m^3. This equals a BOG rate of 133m^3/day. Calculating with a LNG density of 450 kg/m^3 equals 60 tons per day or 2500 kg/h of BOG. This value is set fixed in the simulation and an energy stream is attached to the tank to simulate heat loss. To adjust the BOG rate an adjuster is set du adjust the heat added by the energy stream to match a BOG rate of 2500 kg/h. [12]

3.1.11 Flash and Boil off Gas compression

The flash gas and boil off gas has to be compressed before entering the fuel gas system.The first source of fuel gas used in the simulation will be the flash and boil off gas. Flash gas and the boil off gas will be recompressed and used as fuel gas. To cover the rest of the power demand a split, upstream of the liquefaction part of the plant, is used to transport natural gas to the power generation. See chapter 3.1.7 for more details.

There will be a great difference in the fuel gas pressure required in a gas turbine versus a steam turbine. In this report a fuel gas pressure of 10 bars for the steam turbine and 50 bars for the gas turbine is applied. This will of course affect the compressor work, but with the liquefaction part as the major power consumer (number from the simulation is ~95%), the flash gas and the BOG compressors only counts for about 5% of the power consumption. In other word; even a 100% change in the compressor power will not affect the overall demand more than 5%.

Table 3.3 - Fuel Gas Pressure

	Fuel Gas Pressure
Gas Turbine	50 bar
Steam Turbine	10 bar

The default value in Aspen Hysys with an adiabatic efficiency of 75% is used on all the compressors in the simulation.

4 Cases

4.1 Overview

In all the simulations the Design Basis, which is described in chapter 3, is the common basis. In this chapter the unique parameters and assumption is described.

The different cases main category is what type of drivers for rotating equipment and power generation. In other words; the cases are divided into gas turbine (GT) and steam turbine (ST). In Figure 4.1 all the GT cases are organized and Figure 4.2 all the ST.

The two main cases are divided into 3 subcases that are the same for both GT and ST. These cases are organized by their feed composition. Then the liquefaction process divides each of the subcases based on whether there is used a DMR or N_2 liquefaction process.

In this report the main focus is the comparison between GT and ST, i.e. GT/Low CO_2/MR is interesting to compare with ST/Low CO_2/MR.

During the work with the report there was found to be most practical to divide the report by the feed gas composition, hence under all the headings in the report marked “base cases” you will find both the GT and ST cases with low CO_2 content in feed. Equivalent system is used on both of the two other feed gas compositions analyzed.

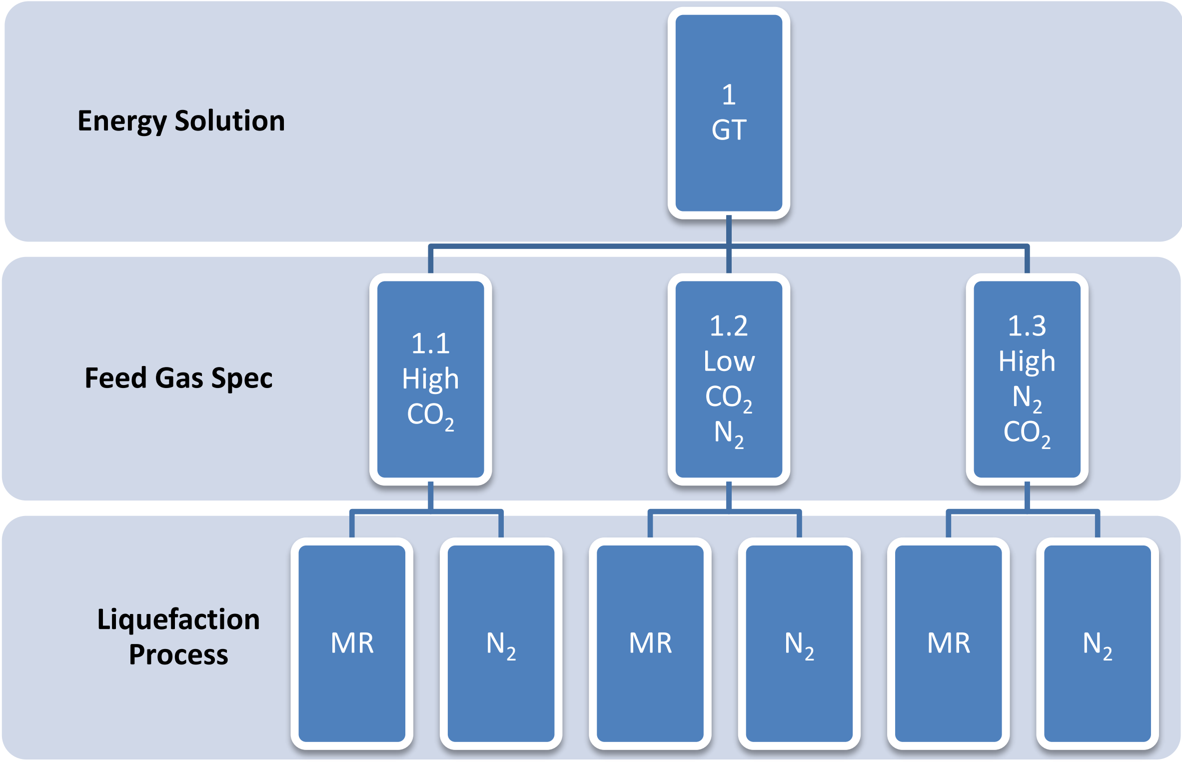

Figure 4.1 - Gas Turbine Cases and Parameters

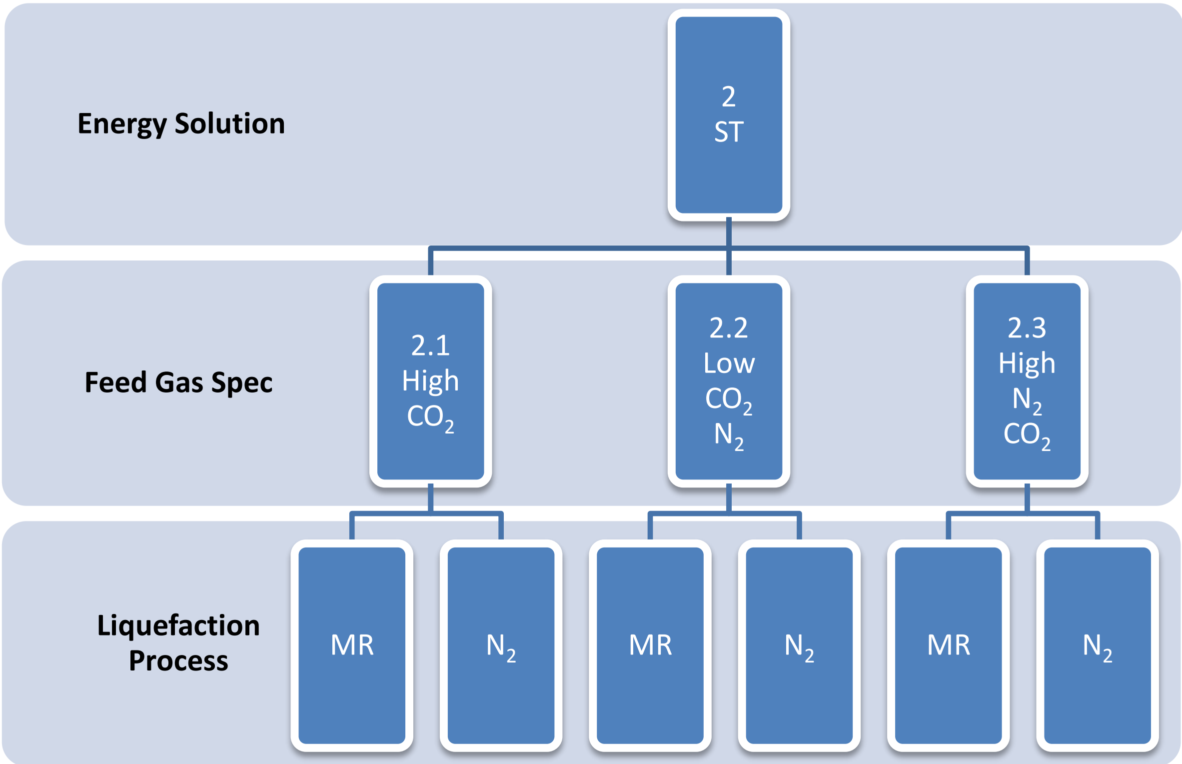

Figure 4.2 - Steam Turbine Cases and Paramaters

Base Cases

The gas turbine setup and steam turbine setup with low CO_2 and N_2 content in feed is set as the base cases for the simulation. The reason for this is the obvious benefit of having the feed specification ease adjustable to the two other feed compositions (increasing only CO_2 content on one hand and increasing both CO_2 and N_2 on the other hand) The two other feed case spec and the associated liquefaction process is compared to the base cases. The base cases are placed straight down from the top of the hierarchy in Figure 4.1 and Figure 4.2 with the two other feed gas specifications on each side

High CO_2 Cases

These cases are placed to the left in Figure 4.1 and Figure 4.2. The high CO_2 content increase the heat demand, hence may favor the ST which have a higher possibility for heat recovery.

High CO_2 & N_2 Cases

These cases are placed to the right in Figure 4.1 and Figure 4.2. Higher flash mass flow may severely increase the flash compressor power need, which may lead to great difference between the 10 and 50 bars FG system in ST and GT respectively.

4.1.1 Gas Turbines

As there is a limited deck space available there has been made a decision that the power setup will be 4+2 LM 6000 gas turbines. The power production per unit is set to 31 MW, which is realistic. The ISO power production is 43 MW, which takes into account: Total pressure 1 atm, total temperature 15 C and relative humidity 60%. This is a more theoretical value and is rarely achieved with normal operation. [11] Another way to look at the power production of 31 MW per unit is 72% load of the ISO power.

14 units are currently in operation on floating production and storage vessels according to GE. [5] To avoid condensation in the LM 6000 gas turbines there is a set a temperature requirement of 28K over the dew point. Heat efficiency is set to 40% and the recovery potential is set to 35% of the LHV in the fuel gas. See Table F.11.

4.1.2 Steam Turbine

For steam turbines the setup is two driver strings for a single process train with power output tailored for the power demand.

The efficiency is set to 25% and the heat recovery is set to 45% of the LHV in the fuel gas. See Table F.11.

4.2 Case Specific Assumptions

4.2.1 Feed Gas Composition

Table 4.1 shows the feed gas composition in the three different feed gas specification shown in Figure 4.1 and Figure 4.2. The first three highlighted rows are the only ones changes in between the Cases. The numbers are mole% of the feed gas.

Table 4.1 - Feed gas composition in mole %

	High CO_2	BASE - Low CO_2	High CO_2 and N_2
C1	80	89	78
CO_2	9,5	0,5	9,5
N_2	1	1	3
H_2O	2	2	2
C2	4	4	4
C3	1,5	1,5	1,5
iC4	0,3	0,3	0,3
nC4	0,4	0,4	0,4
iC5	0,2	0,2	0,2
nC5	0,2	0,2	0,2
C6	0,2	0,2	0,2
C7	0,15	0,15	0,15
C8	0,15	0,15	0,15
C9	0,08	0,08	0,08
C10	0,3	0,3	0,3
C11	0,01	0,01	0,01
C12	0,01	0,01	0,01

4.2.2 Heat Demand in the Gas Processing

The CO_2 removal heat demand is the only heat demand varying between the different cases. Simply because, as described in chapter 0, the C5+ and H_2O content in the feed is exactly the same in all feed compositions. The 100 MW heat demand in the high CO_2 content feed with production capacity of 3.3 MTPA is the reference value. This number is obtained from a similar study and more detailed described in chapter 3.1.6.

For the 0.5 mole% there has been assumed a linear coherence between CO_2 concentration and power demand. In other words; the mole% is lowered (9.5/0.5) 19 times and so are the power demand for the CO2 removal; to the value 5.3 MW.

During the work with this report, there was found that a production capacity of 3.3 MTPA with the available 4+2 GTs and the dual N_2 expander were not possible, hence the production capacity was decrease; to 2.3 MTPA. For these cases the heat demand was also decreased (1/3.3=>) 30%. Hence: i.e. the CO_2 heat demand in the high CO_2 cases will be 75 MW and so on.

Table 4.2 - Splitter Heat Demand in the 3.3 MTPA Cases

	High CO_2	Low CO_2	High N_2 and CO_2
C5+	4	4	4
CO_2	100	5.3	100
Dehydration	4	4	4

4.2.3 Liquefaction

The liquefaction requires large amount of power, which is provided by units (ST or GT) driving the process through shaft power. The liquefaction demand displayed in Table 5.1 is calculated by an exergy flow analysis using equation 4.1, which is derived from exergy rate balance for a control volume. Kinetic and potential energy are neglected. The T_0 ambient temperature is set to 288 K or 15 C, based on common practice.

$$e_{DRY\ NG} - e_{LNG} = h_{LNG} - h_{DRY\ NG} - T_0(s_{LNG} - s_{DRY\ NG}) \qquad 4.1$$

Equation 4.1 is at specific energy basis, hence giving a value in kJ/kg, which is then multiplied with the mass flow in kg/s through the liquefaction heat exchanger to give the power demand. The last thing is to multiply this power demand with the exergy efficiency of the given liquefaction process (45% for MR and 27% for N_2). One important note: the production capacity of LNG is set fixed (respectively 2.3 MTPA for N_2 and 3.3 MTPA for MR); hence will the mass flow through the liquefaction vary, depending largely on the N_2 content. This will be discussed further in the case with large N_2 content in chapter 5.3.

4.2.4 Production capacity

After running some of the simulations and setting up the 4+2 driver and electricity compressors it was, as expected, clear that the dual N_2 expander would have no chance keeping a 3.3 MTPA production rate. The production rate had to be lowered to 2.3 MTPA, when the power production available was set constant.

The limited power available with a GT setup will require a lower production capacity in the dual N_2 expander, because of the lower efficiency compared to the dual mixed refrigerant liquefaction process, hence there are set two different production capacities. For the DMR liquefaction simulation there is set a production capacity of 3.3 MTPA, whereas for the dual N_2 expanders there is set a fixed 2.3 MTPA. For the calculations in Table 4.3 there are assumed 330 operational days and 24 hour operation/day. The daily and hourly numbers are approximate. The stream set to the production capacities described is the stream entering the LNG tank. In other words; the BOG is a loss from the production capacity, because it is used as FG and does not contribute to the LNG sold.

Table 4.3 - Production Capacities

Liquefaction principal	Annual Production	Daily Production	Hourly production
Dual Mixed Refrigerant (DMR)	3.3 MTPA	10000 tons	417 tons
Dual N_2 Expanders (N_2)	2.3 MTPA	7000 tons	290 tons

5 Results and Discussion

5.1 Base Cases

Table 5.1 - Base Cases Low CO_2 and N_2

	Unit	1.2 Gas Turbine		2.2 Steam Turbine	
		MR	N_2	MR	N_2
Key Numbers					
Energy consumption liquefaction	kWh/ton LNG	254	424	254	424
Pressure FG	bar	50		10	
Feed mass flow	tons/h	500,8	359,8	520,0	383,0
LHV Feed	kJ/kg	46986,0		46986,0	
Production capacity of LNG	tons/h	417,0	290,0	417,0	290,0
LHV LNG	kJ/kg	49077,0		49077,0	
Power Demand:					
Utilities	MW	50	50	50	50
Liquefaction	MW	106	123	106	123
BOG Compressor	MW	0,3	0,3	0,1	0,1
Flash Compressor	MW	1,4	1	0,6	0,4
Total Power demand	MW	158	174	156,7	173,5
Power Production:					
Power Potential from Bog and Flash	MW	58,2	43,9	36,2	27,3
Power Production from Feed	MW	109	131	126	150,8
Total Power production	MW	167,2	174,9	162,2	178,1
# of LM6000 (31MW)	#	4+2	4+2		
Power consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	2,2	2,3	2,1	2,2
"FG from feed"/ Feed	% Energy	4,2	7,0	7,43	12,08
Total Fuel gas consumption	% Energy	6,4	9,3	9,53	14,28
"LNG"/ "Feed"	% Energy	87,0	84,2	83,8	79,1
"Condensate"/"Feed"	% Energy	6,6	6,5	6,7	6,6
Heat recovery @ given Power Production					
Heat demand	MW	13	9	13	9
HR from BOG and flash	MW	51	38	65	49
HR from feed	MW	96	115	226	271
Total Heat recovery	MW	147	153	291	321

5.1.1 Elaboration and Comments to the Base Case Table

This chapter contains elaborations of the numbers presented in Table 5.1, which reappear in most of the other table of results in this report. This chapter is essential for proper understanding of the report.

Table 5.1 will be explained from top to bottom divided into the chapters based on the headings in the table. All streams can be viewed in the block diagram on page 11.

5.1.1.1 Key Numbers

The “energy consumption liquefaction” is a specific value (per unit mass) describing the energy consumption of the whole liquefaction process. This number can easily be compared in between the different cases, and is mainly influenced by the exergy efficiency of the liquefaction process (see chapter 2.3.1 for more details). The energy consumption is calculated using equation 5.1. In the equation the power given in kW is the total power demand for the exergy change across the liquefaction heat exchanger as described in chapter 5.1.1.2. The mass flow in the denominator is exported from the simulation model, from the stream entering the LNG tank.

$$\frac{kWh}{ton\ LNG} = \frac{kW * 24\frac{h}{d}}{\frac{ton\ LNG}{day}} \qquad 5.1$$

The Pressure FG in Table 5.1 is the pressure required for the fuel gas. This represents a large difference between GT and ST, and the numbers 50 and 10 bars are user defined values. See Table F.11.

The next four numbers: Feed mass flow, feed lower heating value (LHV), production capacity of LNG and LHV LNG is numbers exported directly from the Hysys simulation model. From the user defined NG Feed stream, and from the LNG stream entering the LNG tank respectively. The LNG stream is at ambient pressure; 1,013 bar and the production capacities are set fixed. All these four numbers are used later in the table under the heading Power Consumption; hence a more detailed elaboration is found in chapter 5.1.1.4.

5.1.1.2 Power Demand

In the Hysys model the liquefaction power demand, and the BOG and flash recompression power demand are simulated. The rest of the power consuming utilities are assumed to have a total power consumption of 50 MW and are listed in Table 5.2. Even though these values will vary depending on i.e. seawater volume flow required for cooling, this is set as constant in all the different cases. This is for the ease of simulation, and to focus the work of the report on the on the main power consumer; the units driving the liquefaction. See side case in chapter 5.4.5 for a more detailed analysis on the seawater pump power demand.

Table 5.2 - Units included in utility power

Amine pumps
Loading pumps
Ballast pumps
Seawater pumps
Freshwater circulation pumps
Vessel stabilization thrusters

For the boil off gas (BOG) and flash compressors the number in Table 5.1 is exported directly from the value of the energy stream attached to the compressor in Hysys. For both of the values the pressure out of the compressor has been adjusted to meet the fuel gas pressure requirement; set as 10 bars for ST and 50 bars for GT. As can be calculated from the values in Table 5.1, the liquefaction shaft power stands for about 65% of the total power demand. There is important to note the difference between 1) the units delivering direct shaft power driving the liquefaction (M-GT) and 2) the drivers delivering electrical power to the other power consumers (E-GT). This is elaborated in the next chapter.

5.1.1.3 Power Production

This heading include, as mentioned, both shaft power production and electrical power generation. In other words; power production to cover the demand from the refrigeration, liquefaction and sub cooling process, and electrical power to provide all the other consumers.

“Power potential from BOG and flash” is numbers for the power possible to utilize in either a GT or ST. The numbers presented in Table 5.1 is calculated using equation 5.2.

$$(LHV * \dot{m})_{Relevant\ stream} * \eta_{effeciency\ energy\ system} \qquad 5.2$$

The working progress to find the total power production has been as follows:

- Setting up a spreadsheet calculating the power potential from the BOG and flash using equation 5.2. $\eta_{\text{efficiency energy system}}$ being the efficiency of the GT or ST, which is set to 40% and 25% respectively.
- Calculating the power deficit between the potential power production from the BOG and flash to the “Total Power Demand”
- Adjusting the “Fuel Gas from feed” mass fraction in the split upstream of the liquefaction to match the power deficit. See Figure 3.1 for block diagram. The mass fraction has been adjusted by one decimal, i.e. 4.2 %. By adjusting the

mass fraction with just one decimal some of the cases in this paper end up with a power production surplus of up to 7%. This could have been avoided by using mass fraction splits with more than one decimal. However; for the scope of this report this is not evaluated to be that important.

The assumption that each LM 6000 gas turbine are able to produce 31 MW is a conservative and realistic number assuming ambient temperature at 27 C, and taking into account the loss in efficiency over the years of operation. [11]

5.1.1.4 Power Consumption

The specific lower heating value is used as a calculation basis for the potential power generation in the GT or ST. Using the LHV instead of the HHV gives a more conservative energy analysis. The LHV is multiplied with the actual mass flow for each case. This gives the theoretical power for the stream.

In Table 5.1 there is a heading called "Power consumption" in energy basis. Equation 5.3 shows how this percentage is calculated. The percentage is included to give a relative perspective on where the energy entering from the feed ends up. A typical fuel gas consumption (energy basis) in an LNG plant utilizing mixed refrigerant is between 5-10%.[9]

$$\frac{(LHV * \dot{m})_{Relevant\ stream}}{(LHV * \dot{m})_{Feed}} \tag{5.3}$$

In total the "LNG product", the "Fuel Gas Consumption", and the "Condensate" does add up to a 100 %. This is the reason for using equation 5.3 instead of the simpler mass balance equation. In a mass balance some percent would be "lost" in the CO_2 removal and dehydration. In the work with this report the energy accounting was found to be a better than a mass balance, because it is more relevant to account for energy than mass in the plant.

The Condensate fraction varies according to meet the pre-liquefaction requirement set to be of less than 0.1 mole % C5+.

5.1.1.5 Heat Recovery

This heading in Table 5.1 displays the heat demand, which is a user specified value depending on the CO_2 removal, condensate stabilization process, and dehydration heat demand. The two latter have the same heat demand (4 MW each) in all the cases discussed in this report, while the CO_2 removal heat demand varies proportional to the CO_2 content in the feed. The heat demand for the CO_2 removal with 9.5% CO_2 in the feed and production capacity of 3.3 MTPA is set to 100 MW. [11]

The heat recovery is divided into heat recovery from BOG and flash, and from feed. This is simply because this is two separate streams in the simulation model, hence practical to look at the heat recovery possibilities in each stream. The heat recovery efficiency is set to 35% in the gas turbine setup and 45% in the steam turbine.

5.1.2 Discussion Base Case

Whereas the previous chapter contains elaboration and comments to the numbers in Table 5.1, this chapter contains discussion and a conclusion for the base case based on the numbers in the same table. Once again the base case will be discussed thoroughly, and the other cases will be discussed relative to the base case.

5.1.2.1 Key Numbers

The energy consumption per unit mass clearly show the difference between the dual N_2 expander and dual mixed refrigerant liquefaction. The higher efficiency of the DMR results in 40% less energy consumption per unit mass. To even consider the N_2 expander liquefaction it has to hold other major advantages. Below is a list of advantages to consider:

- Lower hydrocarbon inventory leads to increased safety, which is an important factor on a vessel. With liquid hydrocarbon refrigerants, the liquefaction part has to be mounted further from the barracks.
- A gaseous refrigerant will be much less influenced by vessel motion than a vaporizing and condensing cycle with HC refrigerant.
- No need for a refrigerant tank as is the case for mixed refrigerant.
- Ease of start-up and shut down, which is extremely relevant in harsh environment with respect to storms and bad weather demanding shut down.

This is the three advantages considered most important, and in addition some others are mentioned in chapter 2.3.2.2. However; the discussion whether these arguments are strong enough or not is basically a trade-off between efficiency and robustness. As a general note the efficiency of the liquefaction may be less important than the robustness and reliability of the plant when operated far from a maintenance port. However; when the production capacity differs 40% one can also argue that a lot of time can be spent one down-time with DMR, still being able to produce equal amounts of LNG as the dual N_2 expanders.

The numbers marked as red in Table 5.1 is important differences that have to be noted, because they dictated different conditions for the separate cases, meaning that one have to take care when comparing cases.

First the different fuel gas pressure required for GT (50 bar) and ST (10 bar), which directly influence the flash and boil off gas compressor work.

Second the two different production capacities in the MR and N_2 case. As can be seen the production capacity are the same for the equivalent case of ST and GT.

5.1.2.2 Power Demand

The liquefaction process is the major influence on the power demand. In other words; there is no difference in power demand whether a GT or ST setup is chosen. To confirm this see Table A.2 in appendix A; where the GT and ST within the base case are compared.

Even though the fact that the dual N_2 expander liquefaction is known to have a lower efficiency than DMR, there is interesting to see the results on how large the difference really is. The N_2 is set to have an exergy efficiency of 27%, compared to the DMR, which is set to have an exergy efficiency of 45%. Even with the production rate lowered 30% (see Table 4.3) the N_2 liquefaction still needs more power than the DMR. Here it is important to note that the liquefaction process is the only parameter changed in between the two cases, hence all the production "loss" is due to lower exergy efficiency!

One thing that can be noticed as a little strange in Table 5.1 is that the BOG compressor power is the same as for both DMR and N_2 liquefaction process cases even with production capacities far apart. This is simply because the BOG rate is set fixed related to the 3.3 MTPA. This is for the ease of simulation and chosen to be set fixed because the BOG compressor work accounts for less than 1% of the total power demand in all the cases, hence having a small impact on the total power demand.

In contrast to the BOG compressor the flash compressor power is higher in the MR than in the N_2 case; the reason being that the mass flow is higher. This explanation applies for both the ST and GT case.

The difference between BOG and flash compressor power is due to the pressure requirements of GT and ST in fuel gas.

5.1.2.3 Power Production

In both the GT/MR and GT/N_2 case 4 LM6000 will be sufficient for driving the precooling, liquefaction and sub cooling of the natural gas. Respectively 106 and 123 MW required. 2 LM 6000 have to be attached to a generator producing electric power for the rest of the power consumers such as utility consumers, BOG- and flash compressors

When looking at the power potential from BOG and flash it is steady falling from the peak value in case GT/MR to the bottom in ST/N_2. The reason is first (from GT/MR to GT/N_2) the lower mass flow, hence energy flow in the feed. In other words; the energy production potential is only decreasing because the starting point (feed energy) is decreasing. Over from the GT cases to the ST cases the reason for decreasing potential is the lower efficiency set in the ST than GT. The decreasing potential in the BOG and flash has to be compensated for by directing a larger mass flow of the pre-liquefaction feed directly to fuel gas.

When there has now been stated that a ST setup will require more power input to be able to drive all the power consumers, there is interesting to see how the availability influences the two different energy solutions.

From an operational point of view the availability is a very important factor in an FLNG. Remote location can require several days before spare parts or qualified personnel arrives, and as pointed out in several references, replacing key equipment items at site may not always be feasible. [1, 3, 13]

A possible and reasonable setup for the M-GT, given the power demand from the simulation, will be 2 train and 4 strings; each train with two GT. In addition the two E-GT generating electricity. For the ST the setup chosen is a single train with two strings. The following calculations are presented to emphasize the availability factor as a factor that should be regarded when making the choice between ST and GT. The exact number should be viewed as more of an example than taken literary. As showed in Table F.11 the availability factors are assumed to be 99% for a single ST and 98% for single GT.

$$GT\ availability = 0{,}98^{\,4} = 92{,}2\ \% \tag{5.4}$$

$$ST\ availability = 0{,}99^{\,2} = 98\ \% \tag{5.5}$$

The results show a rather large difference between the two driver setups. The difference of almost 6% will in operation represent a difference over time of 6% in production capacities, which is a *major argument for steam turbine*. This applies to DMR as well as to N_2 liquefaction.

5.1.2.4 Power consumption

This heading in Table 5.1 contains several interesting numbers. At first sight the obvious trend is, as expected, that the fuel gas consumption increase as the energy solution efficiency and liquefaction exergy efficiency decrease. However; when comparing gas turbine MR and steam turbine MR the difference is lower than one may assume. With the feed energy flow increased by 4 % and energy % used as fuel by about 3%, the absolute value increase of energy consumption is about 7.1% (1.03*1.04=1.0712) in LNG production between the GT and ST. In other word; running for the same amount of days with equivalent production rate the GT will require 7.1% less energy than the ST. This number is lower than expected.

When the availability factor described in chapter 5.1.2.3 is taken into account the steam turbine seems like a rather good choice with the result of 6% higher availability. Even though the GT, when running, uses less energy to produce a certain amount of LNG, the ST will have a longer mean time to failure, hence producing more. The value of being able of producing more LNG over a year is a *major argument for the ST*. See chapter 5.4.2 for a side-case on this argument.

The energy% needed as fuel gas in the simulation model is proportional to the efficiency of the ST and GT; hence the GT will require less of the feed as fuel gas than the ST. In the base case the effect of this results is a need for a higher feed mass flow to assure the given production rate. However; it is important to note that the availability factor is not taken into account in the numbers presented in Table 5.1.

One interesting comparison to make from Table 5.1 is between the power consumption in the GT/N_2 case with the ST/MR case. First, remember that the production rate is different with the N_2 1 MTPA below the MR. However; since the numbers presented under this heading is relative (%) one can compare the numbers. When running a side case increasing the production capacity of GT/N_2 to 3.3 MTPA Hysys calculated a decrease of 1% in the “FG from feed” value; down from 7% in the base case. This side case was run to double check that the result would not differ much when the MTPA was increase. However; increasing the production capacity with dual N_2 expanders has several issues:

- The liquefaction process requires the additional power of two LM6000 (2*31MW)
- This will require more deck space and add much weight to the vessel, which directly increase the capital cost of the vessel.

This is more of a theoretical solution, and may not be possible. Lowering of the production capacity of the ST/MR is more realistic. In other words; if considering the building of a 2.3 MTPA FLNG vessel the GT/N_2 and ST/MR are two alternatives, which in this report score more or less the same on power consumption!

Wood et al. give a crude estimate of the fuel gas consumption in a dual N_2 expander plant of about 12%. This is in the same range as the two N_2 cases in Table 5.4. [1]

5.1.2.5 Heat recovery

In this case there are small CO_2 amount in the feed needed to be removed, hence there is a small heat demand for running the MDEA regenerating process described in chapter 3.1.6. As can be seen under the heat recovery heading in Table 5.1, there is a large heat surplus. Building a heat recovery unit only attached the E-GT units will be more than enough to cover the small heat demand from the condensate stabilization and dehydration (4 MW each). In the ST cases only a HR attached to one of the strings is needed to cover the heat demand. The HR is far from a bottle neck is these cases.

For the heat recovery there is assumed that all the gas turbines running have a heat recovery system attached. A more common practice is to only have heat recovery on the 2 E-GT units. For the base case with its low CO_2 content a HR only on the E-GT units would be sufficient to cover the total of 13 MW. However; in the cases with high CO_2 content, a side case is run to illustrate how the lower heat recovery might require additional heat sources. See chapter 5.4.6 for this side case.

5.1.2.6 Heat distribution

There are two alternatives for distributing heat; hot oil or steam.

Hot oil vs. Steam

A hot oil system connected to the GT may be required, because of the high temperatures. In addition a hot oil system adds complexity, because of a higher pressure

$$\dot{Q} = C_p \dot{m} \Delta T \tag{5.6}$$

A steam heat distribution will be easy to integrate into a ST, because the steam is already there. The thing to do is to extract steam at the required temperature and send this to the heat consumer.

$$\dot{Q} = \dot{m} \Delta h_{fg} \tag{5.7}$$

In Figure 5.1 a comparison of hot oil and steam as heating distribution medium is performed. The basis is equation 5.6 and 5.7, and the graph is an extremely simple illustration. However: it clearly shows that a steam distribution require larger diameter to transport the same heat. In the calculation all numbers are picked from the "The Fundamentals of Thermodynamics" textbook by Moran et al. [14]

$$\dot{m} = \rho V A \tag{5.8}$$

Table 5.3 - Numbers Heat Distribution Systems

$\boldsymbol{C_p\ oil}$	1.9 kJ/kg K
$\boldsymbol{\Delta h_{fg}\ steam\ at\ 4\ bar}$	2133 kJ/kg
$\boldsymbol{\Delta T}$	20K
$\boldsymbol{\rho\ oil}$ at 300 K	884 kg/m^3
$\boldsymbol{\rho\ steam\ at\ 4\ bar}$	2,2 kg/m^3
V for hot oil and steam	2 m/s

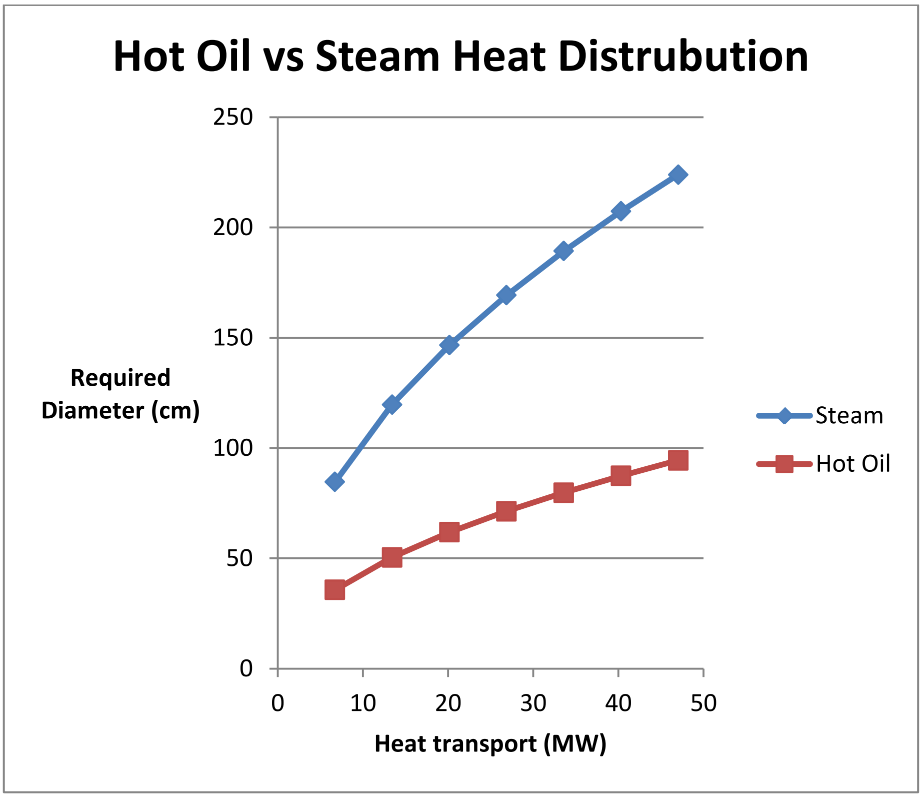

Figure 5.1 - Hot Oil vs Steam Pipe diameter

Figure 5.1 display one of the benefits with hot oil instead of steam as a heat transport medium. The required piping diameter is going to be about the double in a steam setup. By increasing the pressure level of the steam the sizing of the pipe can be reduced. [15] In this crude analysis a single pressure level of 4 bars is assumed. However; it is important to remember that a hot oil system will require larger pressure levels, hence thicker piping, which again will add weight. This is displayed in the analysis performed here:

> "The equipment list for the Trondheim study specifies sizes and weights for equipment in the hot oil system. These have been compared with equipment sizes for a steam system as calculated with GTPRO. Both systems are intended used in connection with 8 GE LM6000 gas turbines. The delivered heat duty is roughly similar, the main difference being that the steam system also is designed to supply a MEG regeneration/reclamation unit." [15]

Case	Steam	Hot Oil
HRSG's, Weight [tons]	568.0	1560.0
Pumps, Weight [tons]	5.8	62.0
Other equipment, Weight [tons]	17.0	70.0
Total Weight [tons]	590.8	1692.0

Figure 5.2 – Steam vs hot oil weight [15]

As can be seen a hot oil setup will be significantly heavier than equivalent steam system. However: as stated in the study the steam system include a MEG regeneration unit, and the sizing philosophy and methodology is not consistent between the two data sets. This introduces a large uncertainty in the comparison. In other word; take the number as an indication rather than exact numbers.

In the Trondheim Study mentioned earlier the waste heat recovery a steam system was used due to the following reasons:

- Extremely large hot oil mass flow would be required for process heating with hot oil
- Hot oil increases danger of fire on board
- Possibility to optimize overall thermal plant efficiency by utilizing a steam turbine for power generation [4]

This heat distribution is included to give an insight in this part of building a FLNG plant. The simplicity and ease of operation with a ST and steam distribution is a clear advantage. However; an analysis on how much of the ST work that is lost due to steam extraction should be performed in HYSYS. This has not been done in this study, hence is a recommendation for further work.

5.1.2.7 Conclusion

Basing the choice of energy production solution solely on the numbers presented in Table 5.1 will clearly make the GT/MR the best choice. Utilizing the fuel gas more efficiently by higher exergy efficiency in the GT makes a higher percent of the feed gas as LNG product, which is the main goal of a LNG plant. As can be seen in Table A.2 the total fuel gas consumption in the GT cases is about 2/3 of the ST cases.

Availability, robustness, and reliability, which in sense all are related, favors the ST. The almost equivalent fuel gas consumption in the GT/MR and ST/N_2 will favor the ST with its higher availability. Even though the N_2 has several benefits over the MR (see chapter 5.1.2.1) the conclusion is that a higher availability is better than the benefits of the N_2 expander. However; going for the ST/N_2 setup is clearly the most reliable solution based on the results and discussion presented above. The prize to pay is lower power efficiency and exergy efficiency in the liquefaction process. Through the work with this case the conclusion is that both of these efficiencies is less valuable than the robustness and availability of the ST/N_2 setup.

5.2 High CO_2 Cases

The GT and ST case will be compared within the case (see Table B.4), and relative to the base cases (see Table 5.4).

Table 5.4 - Relative Values "High CO_2" to Base Case

Numbers in percent, but the units are kept to specify what the numbers are relative to

	Unit	1.1 Gas Turbine		2.1 Steam Turbine	
		MR	N_2	MR	N_2
High CO2 Feed					
Energy consumption liquefaction	kWh/ton LNG	99	99	99	99
Pressure FG	bar	100		100	
Feed mass flow	tons/h	124	124	124	124
LHV Feed	kJ/kg	81		81	
Production capacity of LNG	tons/h	100	100	100	100
LHV LNG	kJ/kg	100		100	
Power Demand:					
Utilities	MW	100	100	100	100
Liquefaction	MW	99	99	99	99
BOG Compressor	MW	100	100	100	100
Flash Compressor	MW	107	100	117	125
Total Power demand	MW	99	99	99	99
Power Production:					
Power Potential from Bog and Flash	MW	101	100	102	101
Power Production from Feed	MW	100	100	99	99
Total Power production	MW	100	100	100	100
Power Consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	100	100	105	100
"FG from feed"/ "Feed"	% Energy	99	99	99	99
Total Fuel gas consumption	% Energy	100	99	101	99
"LNG"/ "Feed"	% Energy	99	99	99	99
"Condensate"/"Feed"	% Energy	110	110	108	109
Heat recovery @ given Power Production					
Heat demand	MW	812	833	812	833
HR from BOG and flash	MW	101	102	102	101
HR from feed	MW	100	100	100	100
Total Heat recovery	MW	100	100	100	100

Comparing these cases to the base case (Table 5.4) makes one interesting result. By increasing the CO_2 and lowering the C1 mole % in the feed, the LHV decreases. Since the production capacity of LNG is set fixed this result in a need for a higher feed mass flow than the base case. As Table B.5 show the LHV (mass basis) is about 20% lower in this cases than base cases, hence to have the same production capacity the feed mass flow is compensating by increasing (about 24%). An increase in mass flow through the plant will in reality increase the utility power. This is a factor that does not come into view in the simulation model since the utility power is set fixed. It will also have a harder toil on the equipment, meaning decrease in lifespan.

For the heat recovery part, there is still more than enough HR potential in all the different setups considered in this case. The heat demand, which in Table A.2 is displayed to increase 800%, is 108 MW and 75 MW. This is solely because of the increased CO_2 mole % in the feed. The heat demand of the CO_2 removal is a input value. However; as in the base case the HR numbers presented is based on heat recovery from all the power producing units. A more common practice is to only have HR on the electric power generation units, and not the units driving the liquefaction process. This will reduce the HR with 2/3, hence a deficit of about (150*1/3= 50 MW, 108-50→) 58 MW in both of the GT cases (see the side case of this in chapter 5.4.6). This can be covered by burning natural gas or condensate. The condensate represents such a small amount of the feed that it may be favorable to utilize the condensate as an energy source on the vessel, instead of storing and shipping it. However; a more detailed analysis of a condensate combustion solution is not included in this report.

When looking at the comparison between GT and ST in Table B.4 the numbers are much alike the equivalent comparison in the base case (Table A.2). Hence the argumentation is more or less the same as in the base case. The most interesting number is the difference in fuel gas consumption, with the GT about 2/3 lower than ST. this is because of the higher efficiency of the GT.

The side cases will illustrate different arguments in the choice of energy solution. The conclusion after running this case does not differ from the conclusion made after the discussion of the base case (see chapter 5.1.2.6)

5.3 High CO_2 & N_2 Cases

This is the cases with both high CO_2 and N_2 in the feed. In Table C.7 the GT/ST relative numbers is presented, and in Table 5.5 the comparison with the base case is presented.

Table 5.5 - Relative Values "High CO_2 & N2" to Base case

Numbers in percent, but the units are kept to specify what the numbers are relative to

	Unit	1.1 Gas Turbine		2.1 Steam Turbine	
		MR	N_2	MR	N_2
High CO2 Feed					
Energy consumption liquefaction	kWh/ton LNG	95	94	95	94
Pressure FG	bar	100		100	
Feed mass flow	tons/h	129	128	127	127
LHV Feed	kJ/kg	78		78	
Production capacity of LNG	tons/h	100	100	100	100
LHV LNG	kJ/kg	99		99	
Power Demand:					
Utilities	MW	100	100	100	100
Liquefaction	MW	95	94	95	94
BOG Compressor	MW	100	100	100	100
Flash Compressor	MW	443	480	467	550
Total Power demand	MW	99	99	99	99
Power Production:					
Power Potential from Bog and Flash	MW	343	362	344	365
Power Production from Feed	MW	0	11	25	46
Total Power production	MW	100	100	100	100
Power Consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	341	370	352	364
"FG from feed"/ "Feed"	% Energy	0	11	25	46
Total Fuel gas consumption	% Energy	100	99	101	99
"LNG"/ "Feed"	% Energy	99	99	99	99
"Condensate"/"Feed"	% Energy	110	110	108	109
Heat recovery @ given Power Production					
Heat demand	MW	812	833	812	833
HR from BOG and flash	MW	343	363	344	364
HR from feed	MW	0	10	25	46
Total Heat recovery	MW	100	100	100	100

As can be seen in Table C.7 the total power consumption in GT/ MR is only 20% lower than the equivalent ST/MR setup. When comparing this value to the same calculations in the other cases, one will find that the difference between GT and ST has decreased with about 12%. In other word; something in the case (GT/MR vs. ST/MR) suits a steam turbine setup better than all the other cases. When looking into this more carefully one find that the mass flow of flash gas is rather large and that the energy content in this stream is more than enough to cover the required power production for all the power consumers. When comparing power demand to potential power production, a power surplus of 40 MW is revealed. Ending up with a power surplus of 20% is not a good design, hence returning to the simulation model, and trying possible solution to this problem. The analysis revealed several solutions.

- Recycle some of the BOG and flash gas upstream of the liquefaction. The reason for mixing it in here is simply because all the pretreatments have been performed to the recycled stream.
- Installing an expander after the liquefaction and before the expansion valve. In other word; ending up with a larger amount of liquid in the stream entering the flash separator, because of the isentropic (expander) in addition to an all isenthalpic (valve) expansion.
- Decoupling the BOG gas from the FG, and putting up a small re-liquefaction process for this stream reinjection it into the tank instead. This leads to a loss of "energy" input to the power generating gas turbines.

After testing the different solutions, there was found that a combination of recycling 10% of the fuel gas, and de-coupling the BOG stream from the FG system made the best improvement (remember: in this case all the FG comes from the flash gas). With this setup the energy surplus decreased to 6%, and with power demand for a re-liquefaction setup for the BOG in mind (not simulated) it is *assumed* that the energy surplus not will be a problem anymore. In many of the other cases the difference between power demand and production is about 5% (see 5.1.1.3); hence solution to the problem gives results in the same range as for the other cases. Absolute values for the side case run to solve the energy surplus problem is found in Table D.9.

5.4 Side-cases

Through the work with the results, several interesting scenarios came up. To better understand the sensitivity of the different parameters some side-cases have been run. The side-cases are included to illustrate general trends. However; several of them reveal very interesting aspects concerning a FLNG vessel. All the side-cases have been run from the base case simulation model with both low CO_2 and N_2.

1. Sub cooling.
2. Availability. Comparing realistic days of operation with ST and GT.
3. Electrical drive.
4. Combined cycle.
5. Number on seawater and cooling pump related to actual flow rates
6. HR from only the power generating turbine (E-GT).

5.4.1 Sub cooling

In the GT/MR base case the mole fraction of N_2 in the LNG is 0.63%. This point is the marked in Figure 5.3 and Figure 5.4. From the base case there have been run several cases to plot a trend on the mole fraction of N_2 in the LNG. One can see from the figures that further sub cooling requires more energy per ton LNG, which is as expected. In this GT/MR the lowest possible temperature is slightly less than 110 K. If cooled further the sales gas requirement of mole fraction N_2 < 0.1 is not fulfilled.

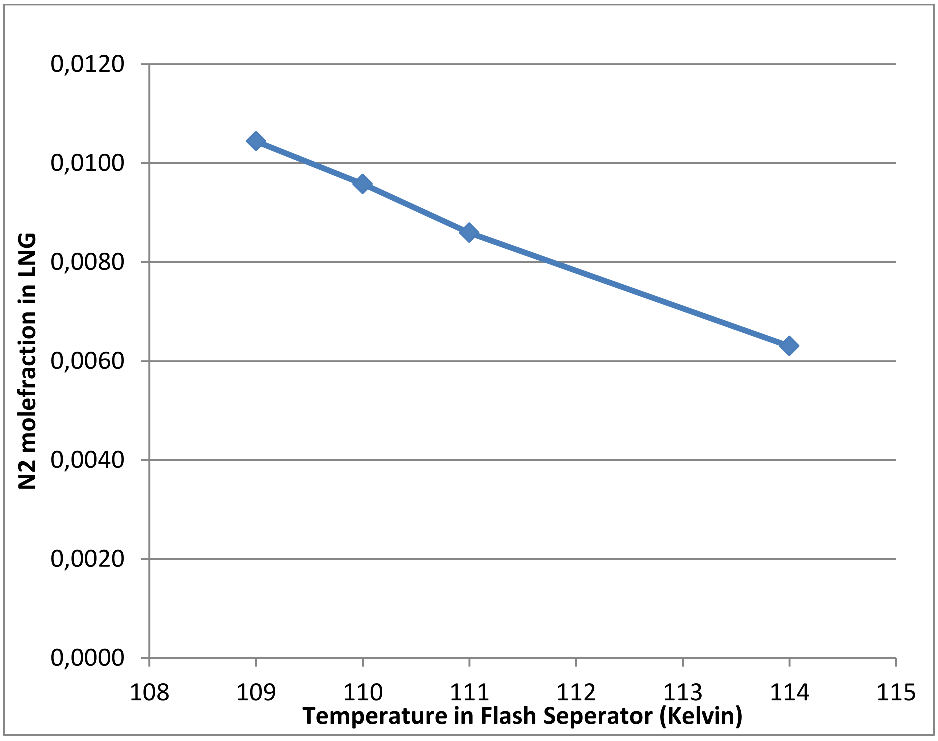

Figure 5.3 – Temperature vs N_2 molefraction

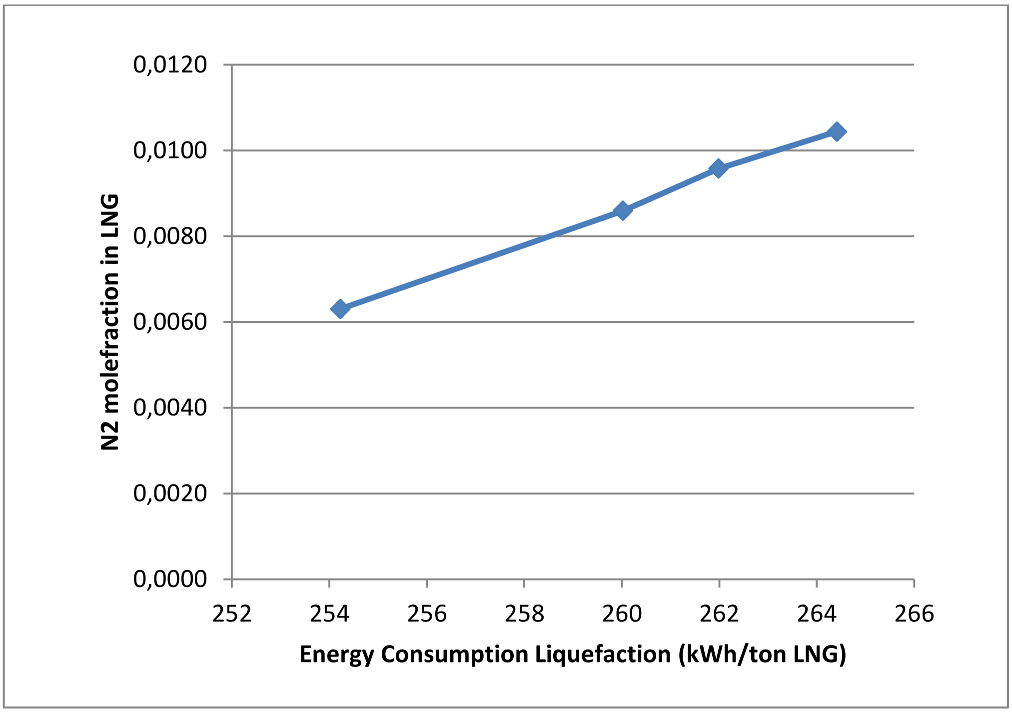

Figure 5.4 – Energy Consumption vs N_2 molefraction

The degree of sub cooling after the liquefaction highly influences the driver power required in the liquefaction, but it also influences the mass flow and mole fraction of the flash stream. In the simulation the parameter adjusting the temperature out of the liquefaction has been ending up with a N_2 mole % between 0.5% and 1%. A N_2 mole % under 1 % is one of the sales gas requirements. With a lower degree of sub cooling a higher fraction of the feed gas energy content enters the flash gas.

The reason for ending the plot at 0.63 % N_2 is because further decrease (higher flash separator temperature) leads to more and more of the hydrocarbons in vapor state, hence a larger percent ending up in the flash gas. This is not desirable, because it will lower the amount of LNG production. In addition; as can be seen in Figure 5.4 the energy consumption increase by almost 4 % by trying to increase the N_2 content in the LNG.

5.4.2 Availability Factor

In this case the daily production rate is still fixed at 10 000 ton/day; however the number of operational days are varied. A simple analysis based on the availability factors discussed in chapter 5.1.2.3 is performed: Lowering the instant LNG production in the ST. with 365 days a year as the starting point and the GT availability 0.93 and ST 0.98. Adding 3 % extra margin to get a more conservative estimate led to availability of 0.90 for GT and 0.95 for ST. Resulting in 325 operational days for the gas turbine and a annual production in the GT/MR base case decrease (from 3.3 MTPA) with 0.05 MTPA, whereas the steam turbine with 345 operational days increase the annual production with 0.15 MTPA. In other words; there is a difference of 0.2 MTPA between GT and ST just because of different availability.

Now considering an even lower availability for the GT caused by the fact that the GT is an open system and may need even more maintenance. Lowering the GT availability from 0.98 to 0.97 leads (0.97^4=0.88; 88% - 3% margin = 85%) 310 days of operation. This leads to 3.1 MTPA, which means that the difference between the GT and ST with this calculation is 0.35 MTPA.

The simple availability calculation shows that the ST has an advantage over the GT when it comes to days of operation. A difference of 0.35 MTPA is a difference of about 10 % of the base case production capacity of 3.3 MTPA. In other words; this should be evaluated as a major advantage for the ST.

5.4.3 Electrical drive

Shell report that after several studies electrical drivers was vote down mainly because the system being overly complex, costly and call for a lot of plot space. [13] The main advantage being that the power generation is decoupled from the compressor units. Power generation efficiency has to be added for accurate calculation. This means a loss overall efficiency. A simple availability shows that an electric drive setup will increase availability when compared to splitting gas turbines into M-GT and E-GT. A favorable setup here could be running all 6 GTs on 80% load, and in the incident of one GT down turning the 5 remaining GT up to 100%. Part load of gas turbine is beyond the scope of this report. The probability for 3 of the E-GTs to be down at the same time is the probability of one being down times 3. 2%^3= 0,0008%. In other word; a almost 100% availability of four of the E-GT to be running at the same time. This is the major argument for electrical drive.

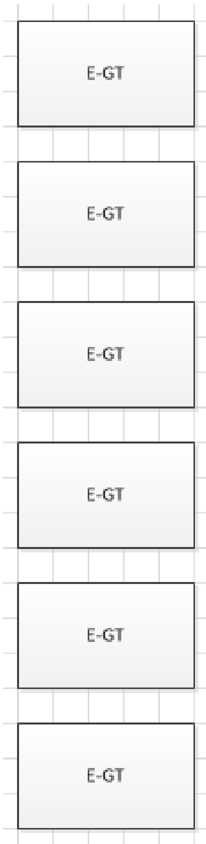

Figure 5.5 - E-GT

All GT in parallell

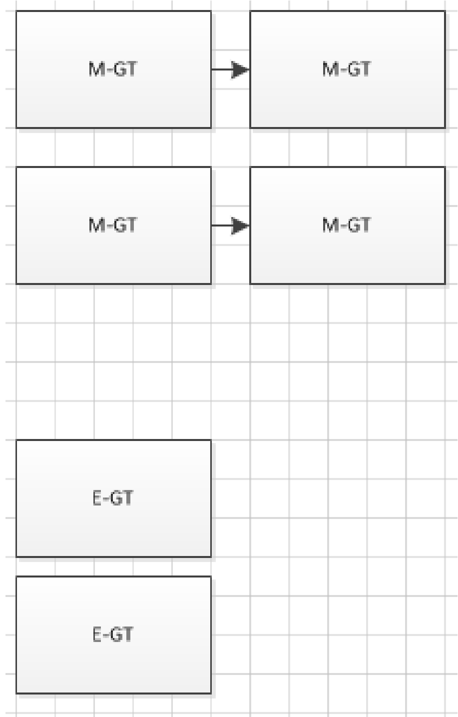

Figure 5.6 - M-GT 2 and 2 in serie

Figure 5.5 and Figure 5.6 illustrate the difference between having an all-electric drive vs. mechanical drive+ electricity production. With all E-GT solution one will have all the GTs in parallel, hence much higher redundancy, whereas M-GT setup will require 2+2 GTs in series driving the liquefaction. If one of the M-GT is down 50% of the production capacity is lost.

5.4.4 Combined Cycle

The main benefit in a combined cycle is an increase in overall plant efficiency. This means smaller fuel gas consumption, hence larger amount of natural gas for sale. An easy integration of steam heat distribution is also a benefit compared to a simple GT setup. The waste heat is dissipated almost entirely in the exhaust in a GT. This results in a high temperature exhaust stream that is very usable for boiling water in a combined cycle, or for cogeneration. There has not been performed a analysis of a combined cycle, so this is a recommendations for further work.

5.4.5 Cooling water

Cooling system runs on 50 000 m^3/h and utilize the lower temperature deep under the surface by having the intake a 150-200 m below sea level. Seawater is pumped up through a free hanging bundled riser. The reason for not guiding the risers through the turret and connecting the riser to the seabed is simply a question of available space in the turret. With 9 risers of 1 m in diameter each there no way the required cooling system could go through the turret says Shell.

A gas turbine will need a total of about 16 000 m3/h sea water cooling, which is much less than the identical steam turbine with sea water demand of 50000 m3/h. The main consumer of sea water cooling is cooling of the refrigerant in the liquefaction process. As a simplified analysis the theoretical pump power is calculated in the GT and ST cases.

$$Pump\ power = v * \Delta P \qquad 5.9$$

With equation 5.9 and the assumed values for the sea water volume flow there is found that the ST will require more than 3 times as much pump power as the GT. Putting in a pressure difference of 1 bar, give a required pump power of about 450 KW for GT, and 1.3 MW for ST. Even though this does not count for much in big picture when looking at power demand, one also has to note that a mass flow 3 times larger will require larger/more pumps, hence more weight/space.

5.4.6 Heat Recovery only from E-GT

In all the result tables there are assumed, as stated in the design basis, heat recovery from all the GTs is assumed. This gives a heat recovery far above the heat demand in all the cases analyzed. A more common way of integrating heat recovery from GT is to only have heat recovery from the electricity generating GT (E-GT), hence reducing the HR by 4/6. [1] This results in a heat recovery lower than the heat demand in all the GT cases with high CO_2 concentration. These cases have a heat demand set fixed at 108 MW and 75 MW (see Table 4.2), and with only 1/3 of the HR, light analysis based on the base case results show the HR form the E-GT will cover 50-60 % of the total heat demand. To cover the rest the small condensate amount or more of the natural gas could be burnt in a furnace solely for heat purpose.

According to Lieberman et al one can achieve 98 % efficiency in a modern industrial furnace. [16]. An analysis of the extra natural gas needed to cover the rest of the heat demand has been performed on the GT/High CO_2/MR case in Hysys. The LHV of the FG stream pre-liquefaction is 48600 kJ/kg. By setting the furnace efficiency to 93%, in other words; conservative in respect to the reported value by Lieberman, the analysis show a required mass flow of 1.03 kg/s to cover the ~50MW of heat required in addition the HR.

$$48600 \frac{kJ}{kg} * 1.03 \frac{kg}{s} = 50\ MW \qquad 5.10$$

$$\frac{50MW}{0.93} = 54MW \qquad 5.11$$

$$\frac{LHV_{furnace\ stream} * \dot{m}}{LHV_{feed} * \dot{m}} = \frac{54MW}{6571\ MW} = 0.8\% \qquad 5.12$$

The three equations above show the calculation performed to get an idea of the extra fuel gas consumption because of no HR from the M-GT and, hence an addition of a furnace. This gives almost a 1% increase of the total fuel gas consumption. What is interesting to compare now is the fuel gas consumption of GT/High CO_2/MR with the ST/High CO_2/MR. In Figure 5.7 the blue part represent the extra fuel gas needed in the furnace to cover the heat demand. In the GT/N_2 case it is assumed an equal increase of 0.8%, hence the blue part of the column. As was discussed in 5.1.2.4, the GT/N_2 and ST/DMR are almost equivalent when comparing fuel gas consumption. However; with the extra fuel gas consumption as a consequence of lower heat recovery potential from the GT, the FG consumption in the GT/MR case is now higher than the ST/N_2.

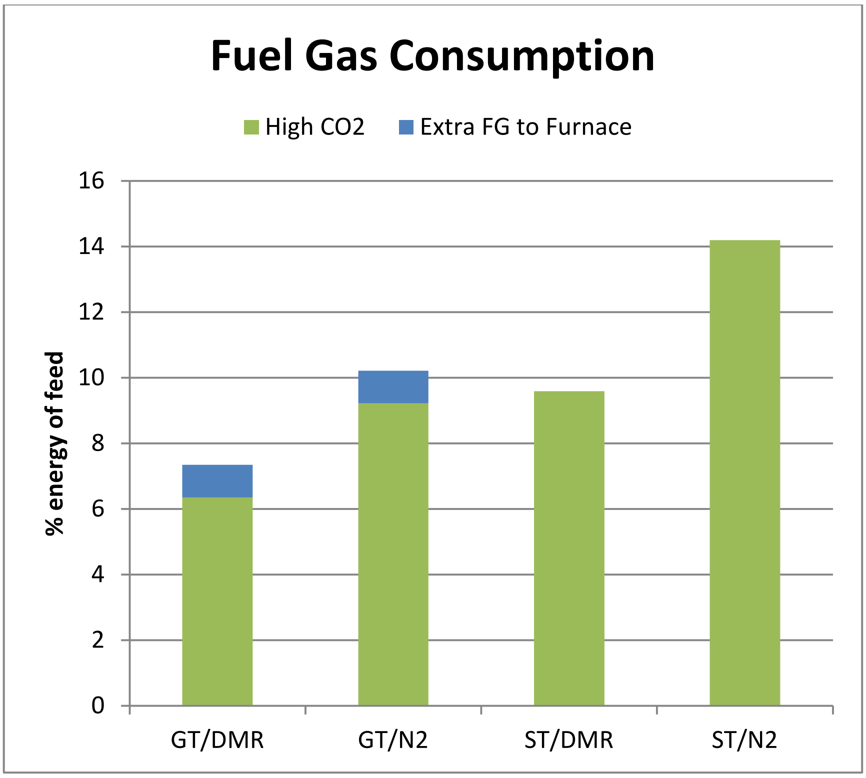

Figure 5.7 – Fuel Gas Consumption with reduced HR

In the ST cases this is not a problem. Even with HR on only one of the strings would cover the heat demand in all the cases. Hence; this factor will favor the ST.

6 Conclusion

Through this study several aspects of the selection of energy system for floating LNG has been revealed. In chapter 2 references and other studies is presented to get an overview of the different energy solution. The Shell Prelude project is the study encountered most when searching for FLNG studies, and is the only FLNG project close to reality; investment decision made in 2011.

The main part of this study was to run simulation models on a FLNG plant. After building and verifying a base case model, different cases for liquefaction process and feed gas consumption was analyzed. Figure 6.1 summarize the fuel gas consumption in all the different cases. With respect to fuel gas consumption and efficiency the selection of energy solution and liquefaction process will clearly be GT/MR. However; the total reliability look like the inverse, with ST/N_2 as the most reliable solution.

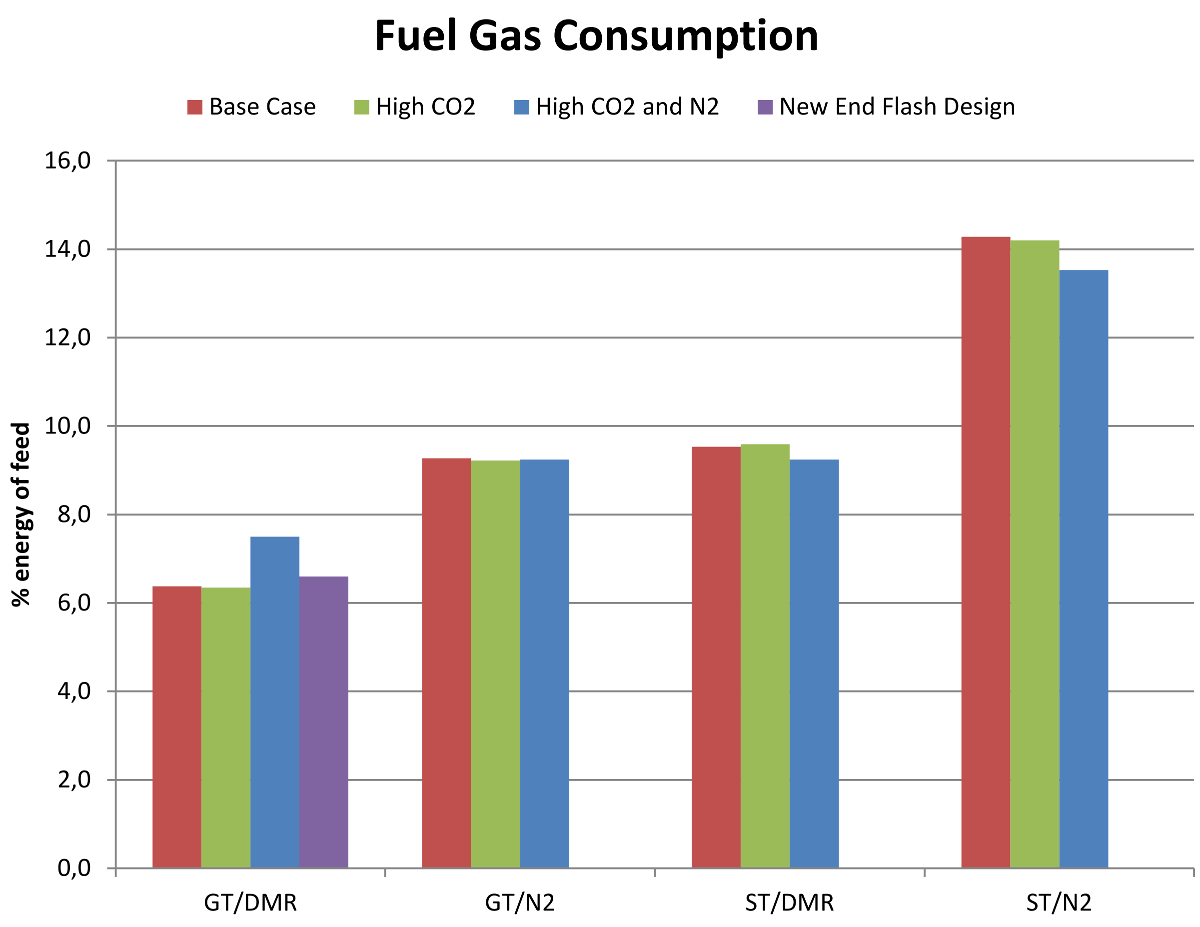

Figure 6.1 – Fuel Gas Consumption Chart

The selection of energy system is practical to divide into two categories. With a gas turbine one will have a high efficiency, whereas factors as safety, availability and robustness will favor steam turbine rather than gas turbine.

As described in chapter 2.3.2.3 the dual N_2 expander has an advantage over the DMR on all of the following factors: Start-up and shutdown time, compactness, weight, sensitivity to vessel motion and safety. However as Figure 6.1 show, the DMR has an edge when solely looking at fuel gas consumption. The fuel gas consumption is mainly influenced by the exergy efficiency of the liquefaction process.

A conclusion on whether to choose a GT/N_2 or a ST/DMR is hard. They both have one reliable unit and one “unreliable” unit. However; remember that the production capacity is 1 MTPA for the ST/DMR, and installing 2 extra GTs to increase the GT/N_2 to the same production capacity will certainly increase the weight and decrease the availability given that one will need more than 4 M-GT for the liquefaction process. If comparing a production capacity of 2.3 MTPA it all comes down to which factors one will weigh the most. Through the work with this report a selection of GT/N_2 will be the conclusion. The reason being all the mentioned advantages of the dual N_2 expander process with special emphasize on vessel motion insensitivity and safety.

Given emphasize on reliability, robustness and safety the 14% fuel gas consumption in the ST/N_2 setup could be favorable. In respect to the more realistic fuel gas consumption of the GT with lower HR capacity in Figure 5.7, the ST/N_2 have a fuel gas consumption 4% higher than the GT/N_2. On the basis of the results in this report and other studies performed on FLNG a selection of the ST/N_2 setup will be favorable as long as there is high CO_2 removal heat demand. With low CO_2 content, hence heat demand, the advantage of the ST is smaller thanks to lower heat recovery demand.

An estimate of the capital cost could be made on a weight basis; hence a lightweight setup with small numbers of equipment parts should be prioritized. The size and weight of the dual N_2 expander would be lower than DMR, because of no need for storage for refrigerant, smaller safety area range, and fewer equipment units overall. The ST is more bulky and heavier than an aeroderivative GT, hence requiring more space and adding weight. However there has not been performed any detailed analysis on the weight and space difference of the GT and ST.

7 Further Work

There are several possible further studies with great interest. An experimental study of the vessel sensitivity of the two different liquefaction processes; dual N_2 expander and dual mixed refrigerant will be very interesting. This however; requires more time and an accessible laboratory.

In addition some other interesting further work:

- The condensate represents such a small amount of the feed that it may be favorable to utilize the condensate as an energy source on the vessel, instead of storing and shipping it. Performing an analysis of condensate utilization could reveal new energy solutions
- Further analysis of combined cycle and electric drive. Both of these energy solutions has not been prioritized in this report; hence a more detailed analysis is left for future work.
- Based on side case for sub cooling there is interesting to see further into the end flash solution. Other master thesis has been performed on end flash solution at EPT this semester. Implementing more of this into an energy solution analysis would be interesting. See the side case sub cooling in chapter 5.4.
- An analysis on how much of the ST work that is lost due to steam extraction should be performed in HYSYS.
- Increase the efficiency of the steam turbine to 30%. This is the efficiency reported by Shell on their Prelude project. This 5% increase from the efficiency used in this report, will make the efficiency difference between GT and ST smaller.
- More detailed analysis on space and weight cost for a FLNG vessel.

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Appendices

A Base Case Simulation Results in Tabular Form

Table A.1 - Absolute Values Base Case

	Unit	**1.2 Gas Turbine**		**2.2 Steam Turbine**	
		MR	N_2	MR	N_2
Key Numbers					
Energy consumption liquefaction	kWh/ton LNG	254	424	254	424
Pressure FG	bar	50		10	
Feed mass flow	tons/h	500,8	359,8	520,0	383,0
LHV Feed	kJ/kg	46986,0		46986,0	
Production capacity of LNG	tons/h	417,0	290,0	417,0	290,0
LHV LNG	kJ/kg	49077,0		49077,0	
Power Demand:					
Utilities	MW	50	50	50	50
Liquefaction	MW	106	123	106	123
BOG Compressor	MW	0,3	0,3	0,1	0,1
Flash Compressor	MW	1,4	1	0,6	0,4
Total Power demand	MW	158	174	156,7	173,5
Power Production:					
Power Potential from Bog and Flash	MW	58,2	43,9	36,2	27,3
Power Production from Feed	MW	109	131	126	150,8
Total Power production	MW	167,2	174,9	162,2	178,1
# of LM6000 (31MW)	#	4+2	4+2		
Power consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	2,2	2,3	2,1	2,2
"FG from feed"/ Feed	% Energy	4,2	7,0	7,43	12,08
Total Fuel gas consumption	% Energy	6,4	9,3	9,53	14,28
"LNG"/ "Feed"	% Energy	87,0	84,2	83,8	79,1
"Condensate"/"Feed"	% Energy	6,6	6,5	6,7	6,6
Heat recovery @ given Power Production					
Heat demand	MW	13	9	13	9
HR from BOG and flash	MW	51	38	65	49
HR from feed	MW	96	115	226	271
Total Heat recovery	MW	147	153	291	321

Table A.2 - Relative Values GT/ST

In %

	Comparison	
	GT MR/ ST MR	GT N2/ ST N2
Key Numbers		
Feed mass flow	96,3	93,9
Power Demand:		
Utilities	100	100
Liquefaction	100	100
BOG Compressor	300	300
Flash Compressor	233	250
Total Power demand	101	100
Power Production:		
Power Potential from Bog and Flash	161	161
Power Production from Feed	87	87
Total Power production	103	98
Power consumption		
"FG from Bog and Flash"/ "Feed"	105	105
"FG from feed"/ Feed	56	58
Total Fuel gas consumption	67	65
"LNG"/ "Feed"	104	106
"Condensate"/"Feed"	99	98
Heat recovery @ given Power Production		
Heat demand	100	100
HR from BOG and flash	78	78
HR from feed	42	42
Total Heat recovery	50	48

B High CO_2 Simulation Results in Tabular Form

Table B.3 - Absolute Values "High CO_2"

	Unit	**1.1 Gas Turbine**		**2.1 Steam Turbine**	
		MR	N_2	MR	N_2
High CO2 Feed					
Energy consumption liquefaction	kWh/ton LNG	252	420	252	420
Pressure FG	bar	50		10	
Feed mass flow	tons/h	622,0	447,0	645,7	475,3
LHV Feed	kJ/kg	38030,0		38030,0	
Production capacity of LNG	tons/h	417	290	417	290
LHV LNG	kJ/kg	49000,0		49000,0	
Power Demand:					
Utilities	MW	50	50	50	50
Liquefaction	MW	105	122	105	122
BOG Compressor	MW	0,3	0,3	0,1	0,1
Flash Compressor	MW	1,5	1	0,7	0,5
Total Power demand	MW	157	173	155,8	172,6
Power Production:					
Power Potential from Bog and Flash	MW	59	44	36,9	27,7
Power Production from Feed	MW	109	131	125	150
Total Power production	MW	168	175	161,9	177,7
# of LM6000 (31MW)	#	4+2	4+2		
Power Consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	2,2	2,3	2,2	2,2
"FG from feed"/ "Feed"	% Energy	4,15	6,92	7,39	12
Total Fuel gas consumption	% Energy	6,35	9,22	9,59	14,2
"LNG"/ "Feed"	% Energy	86,4	83,6	83,2	78,6
"Condensate"/"Feed"	% Energy	7,3	7,2	7,2	7,2
Heat recovery @ given Power Production					
Heat demand	MW	108	75	108	75
HR from BOG and flash	MW	52	39	66,4	49,9
HR from feed	MW	96	114	226	271,1
Total Heat recovery	MW	147	153	292,4	321

Table B.4 - Relative Values “High CO_2”

Numbers in %

	Comparison	
	GT MR/ ST MR	GT N2/ ST N2
Key Numbers		
Feed mass flow	96,3	94,0
Power Demand:		
Utilities	100	100
Liquefaction	100	100
BOG Compressor	300	300
Flash Compressor	214	200
Total Power demand	101	100
Power Production:		
Power Potential from Bog and Flash	160	159
Power Production from Feed	87	87
Total Power production	104	98
Power consumption		
"FG from Bog and Flash"/ "Feed"	100	105
"FG from feed"/ Feed	56	58
Total Fuel gas consumption	66	65
"LNG"/ "Feed"	104	106
"Condensate"/"Feed"	101	100
Heat recovery @ given Power Production		
Heat demand	100	100
HR from BOG and flash	78	78
HR from feed	42	42
Total Heat recovery	50	48

Table B.5 - Relative Values "High CO_2" to Base Case

Numbers in % of equivalent base case. Identical to Table 5.4

	Unit	1.1 Gas Turbine		2.1 Steam Turbine	
		MR	N_2	MR	N_2
High CO2 Feed					
Energy consumption liquefaction	kWh/ton LNG	99	99	99	99
Pressure FG	bar	100		100	
Feed mass flow	tons/h	124	124	124	124
LHV Feed	kJ/kg	81		81	
Production capacity of LNG	tons/h	100	100	100	100
LHV LNG	kJ/kg	100		100	
Power Demand:					
Utilities	MW	100	100	100	100
Liquefaction	MW	99	99	99	99
BOG Compressor	MW	100	100	100	100
Flash Compressor	MW	107	100	117	125
Total Power demand	MW	99	99	99	99
Power Production:					
Power Potential from Bog and Flash	MW	101	100	102	101
Power Production from Feed	MW	100	100	99	99
Total Power production	MW	100	100	100	100
Power Consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	100	100	105	100
"FG from feed"/ "Feed"	% Energy	99	99	99	99
Total Fuel gas consumption	% Energy	100	99	101	99
"LNG"/ "Feed"	% Energy	99	99	99	99
"Condensate"/"Feed"	% Energy	110	110	108	109
Heat recovery @ given Power Production					
Heat demand	MW	812	833	812	833
HR from BOG and flash	MW	101	102	102	101
HR from feed	MW	100	100	100	100
Total Heat recovery	MW	100	100	100	100

C High CO_2 & N_2 Simulation Results in Tabular Form

Table C.6 - Absolute Values “High CO_2 and N_2”

	Unit	1.3 Gas Turbine		2.3 Steam Turbine	
		MR	N_2	MR	N_2
Key Numbers					
Energy consumption liquefaction	kWh/ton LNG	242	399	242	399
Pressure FG	bar	50		10	
Feed mass flow	tons/h	648,0	460,0	661,6	485,0
LHV Feed	kJ/kg	36860,0		36860,0	
Production capacity of LNG	tons/h	417,0	290,0	417,0	290,0
LHV LNG	kJ/kg	48757,0		48757,0	
Power Demand:					
Utilities	MW	50	50	50	50
Liquefaction	MW	101	116	101	116
BOG Compressor	MW	0,3	0,3	0,1	0,1
Flash Compressor	MW	6,2	4,8	2,8	2,2
Total Power demand	MW	158	171	153,9	168,3
Power Production:					
Power Potential from Bog and Flash	MW	199,4	159	124,6	99,6
Power Production from Feed	MW	0	13,9	31,2	69
Total Power production	MW	199,4	172,9	155,8	168,6
# of LM6000 (31MW)	#	4+2	4+2		
Power consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	7,5	8,5	7,4	8
"FG from feed"/ Feed	% Energy	0,0	0,7	1,84	5,53
Total Fuel gas consumption	% Energy	7,5	9,2	9,24	13,53
"LNG"/ "Feed"	% Energy	85,1	83,4	83,4	79,1
"Condensate"/"Feed"	% Energy	7,4	7,4	7,4	7,4
Heat recovery @ given Power Production					
Heat demand	MW	108	75	108	75
HR from BOG and flash	MW	175	139	224	179
HR from feed	MW	0	12	56	124
Total Heat recovery	MW	175	151	280	303

Table C.7 - Relative Values “High CO_2 and N_2”

Numbers in %

	Comparison	
	GT MR/ ST MR	GT N2/ ST N2
Key Numbers		
Feed mass flow	97,9	94,8
Power Demand:		
Utilities	100	100
Liquefaction	100	100
BOG Compressor	300	300
Flash Compressor	221	218
Total Power demand	102	102
Power Production:		
Power Potential from Bog and Flash	160	160
Power Production from Feed	0	20
Total Power production	128	103
Power consumption		
"FG from Bog and Flash"/ "Feed"	101	106
"FG from feed"/ Feed	0	13
Total Fuel gas consumption	81	68
"LNG"/ "Feed"	102	105
"Condensate"/"Feed"	101	100
Heat recovery @ given Power Production		
Heat demand	100	100
HR from BOG and flash	78	78
HR from feed	0	10
Total Heat recovery	62	50

Table C.8 - Relative Values "High CO_2 and N_2" to Base Case

Numbers in % of equivalent base case. Identical to Table 5.5

Energy Solution Liquefaction Process	Unit	1.1 Gas Turbine MR	1.1 Gas Turbine N_2	2.1 Steam Turbine MR	2.1 Steam Turbine N_2
High CO2 Feed					
Power consumption liquefaction	kWh/ton LNG	95	94	95	94
Pressure FG	bar	100		100	
Feed mass flow	tons/h	129	128	127	127
LHV Feed	kJ/kg	78		78	
Production capacity of LNG	tons/h	100	100	100	100
LHV LNG	kJ/kg	99		99	
Power Demand:					
Utilities	MW	100	100	100	100
Liquefaction	MW	95	94	95	94
BOG Compressor	MW	100	100	100	100
Flash Compressor	MW	443	480	467	550
Total Power demand	MW	99	99	99	99
Power Production:					
Power Potential from Bog and Flash	MW	343	362	344	365
Power Production from Feed	MW	0	11	25	46
Total Power production	MW	100	100	100	100
Power Consumption					
"FG from Bog and Flash"/ "Feed"	% Energy	341	370	352	364
"FG from feed"/ "Feed"	% Energy	0	11	25	46
Total Fuel gas consumption	% Energy	100	99	101	99
"LNG"/ "Feed"	% Energy	99	99	99	99
"Condensate"/"Feed"	% Energy	110	110	108	109
Heat recovery potential @ given Power Production					
Heat demand	MW	812	812	812	812
HR from BOG and flash	MW	343	363	344	364
HR from feed	MW	0	10	25	46
Total Heat recovery potential	MW	100	108	100	100

D Simulation Results remake case GT/High CO_2 and N_2/MR

Table D.9 - Recycle Case "High CO_2 and N_2"

Better design of demand/production.	Unit	**Gas Turbine** MR
Key Numbers		
Power consumption liquefaction	kWh/ton LNG	251
Pressure FG	bar	50
Feed mass flow	tons/h	643,0
LHV Feed	kJ/kg	36860,0
Production capacity of LNG	tons/h	417,0
LHV LNG	kJ/kg	48716,0
Power Demand:		
Utilities	MW	50
Liquefaction	MW	104
BOG Compressor	MW	0,3
Flash Compressor	MW	6,1
Total Power demand	MW	160
Power Production:		
Power Potential from Bog and Flash	MW	172,9
Power Production from Feed	MW	0
Total Power production	MW	172,9
# of LM6000 (31MW)	#	4+2
Power consumption		
"FG from Bog and Flash"/ "Feed"	% Energy	6,6
"FG from feed"/ Feed	% Energy	0,0
Total Fuel gas consumption	% Energy	6,6
"LNG"/ "Feed"	% Energy	85,7
"Condensate"/"Feed"	% Energy	7,7

E Results from side cases in Tabular and Graphical form

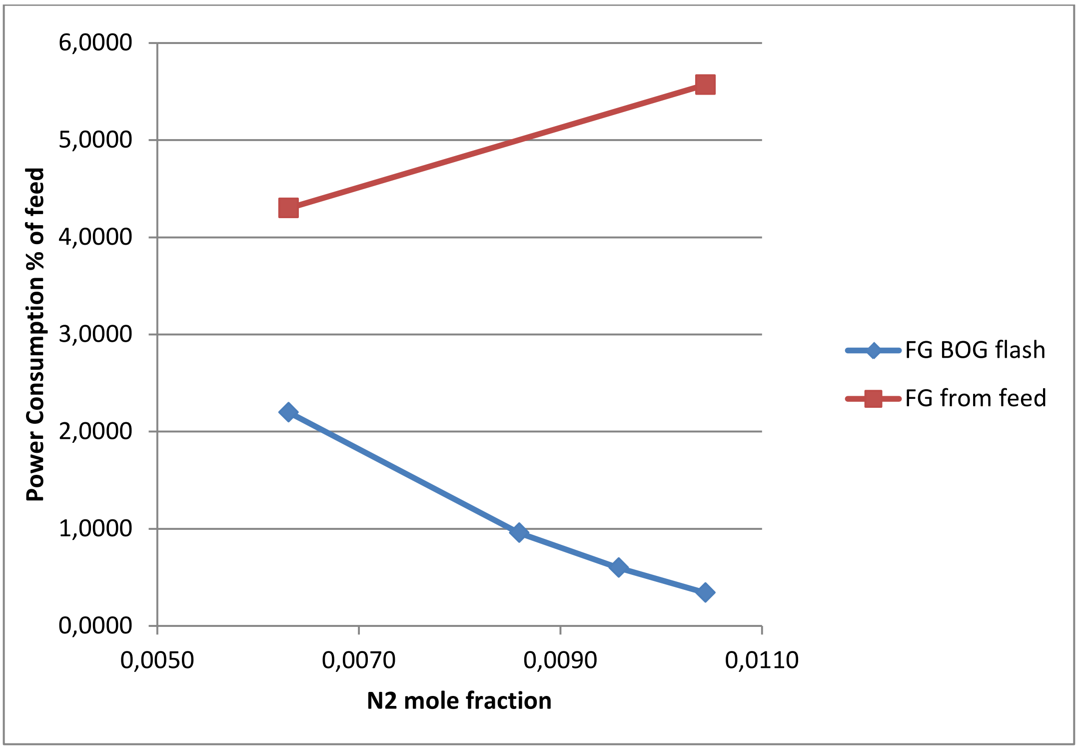

Figure E.1 - Effects on FG from feed under high subcooling.

The lower the subcooling temperature (increased mole fraction N_2) the more of the feed has to go directly to FG to compensate for the loss of FG from flash.

Table E.10 - Absolute and relative values further sub-cooling

110 K, 0.96% N2	Unit	**1.2 Gas Turbine** MR	**2.2 Steam Turbine** MR		**GT relative to base**	ST relative to base case
Key Numbers						
Power consumption liquefaction	kWh/ton LNG	262	262		103	103
Pressure FG	bar	50	10		100	100
Feed mass flow	tons/h	497,2	516,9		99	99
LHV Feed	kJ/kg	46986,0	46986,0		100	100
Production capacity of LNG	tons/h	417,0	417,0		100	100
LHV LNG	kJ/kg	48826,7	48825,5		99	99
Power Demand:						
Utilities	MW	50	50		100	100
Liquefaction	MW	109	109		103	103
BOG Compressor	MW	0,3	0,1		100	100
Flash Compressor	MW	0,2	0,1		14	17
Total Power demand	MW	160	159		101	102
Power Production:						
Power Potential from Bog and Flash	MW	15,4	9,6		26	27
Power Production from Feed	MW	144,7	151		133	120
Total Power production	MW	160,1	160,6		96	99
# of LM6000 (31MW)	#	4+2				
Power consumption						
"FG from Bog and Flash"/ "Feed"	% Energy	0,6	0,6		27	29
"FG from feed"/ Feed	% Energy	5,6	8,92		133	120
Total Fuel gas consumption	% Energy	6,2	9,5		97	100
"LNG"/ "Feed"	% Energy	87,2	83,8		100	100
"Condensate"/"Feed"	% Energy	6,6	6,7		100	100

F Numbers provided by my supervisor

Table F.11 - Numbers provided by supervisor

Description	Value
Starting Production capacity	3,3 MTPA
Heat demand CO2 removal **(with 9,5 mole % in feed @ 3.3 MTPA)**	100 MW
Heat demand Condensate Stabilization	4 MW
Heat demand dehydration	4 MW
Pressure into liquefaction	65,5 bar
Temperature into liquefaction	45 C
C5+ content liquefaction	0,1 mole %
Exergy Efficiency DMR	45%
Exergy Efficiency N_2	27%
Maximum N_2 **content LNG**	1 mole %
Availability factor steam turbine	99%
Efficiency steam turbine (W/Q)	25%
Heat recovery steam turbine	45%
Availability factor gas turbine	98%
Efficiency gas turbine (W/Q)	40%
Heat recovery gas turbine	35%