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Mass balance model for Hammerfest LNG plant Snøhvit

Anders Lauvdal

Master of Science in Energy and Environment
Submission date: December 2007
Supervisor: Jostein Pettersen, EPT

Norwegian University of Science and Technology
Department of Energy and Process Engineering

Problem Description

The goal of the task is to develop and test a mass balance model for the entire Snøhvit facility which makes it possible to balance all streams inn and out. The model should be implemented in an excel sheet. Also production rate calculations for LNG, LPG and condensate should be made based on updated well and field data.

1.Go through and sum up the process steps and streams at the Snøhvit onshore facilities and establish the basis and principles for the mass balance model. Special precautions to consider are the maximal heating value of the LNG product, other limits in product specifications and a realistic modeling of fuel gas consumption.
2.Develop the mass balance model and implement it in an excel sheet while documenting the model and excel sheet in a good way. Test and tune/modify the model based on existing process simulation data.
3.Carry out model calculations based on new reservoir data and field production data.
4.Evaluate further possibilities of developing the model including a better representation of the extraction of heavier hydrocarbons and implementation of capacity limits for the process and equipment.
5.Make suggestions for further tuning of the model based on real production data.

Assignment given: 06. August 2007
Supervisor: Jostein Pettersen, EPT

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ADDRESS:
NTNU
DEPARTMENT OF ENERGY AND PROCESS Engineering
Kolbjørn Hejes vei 1A
N-7491 Trondheim - NTNU

TELEPHONE
Switchboard NTNU: 73 59 40 00
Department office: 73 59 27 00
Hydropower section: 73 59 38 57

TELEFAX
Department office: 73 59 83 90
Hydropower section: 73 59 38 54

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Abstract
The objective of this work has been to develop a relatively simple Excel model that can be used for production forecasting purposes for Snøhvit LNG plant. Effort has been put into making the model intuitive and user-friendly. The model has further been used to forecast the production from updated production profiles (December 2007).

The development of the model has been based on detailed descriptions of processes at Melkøya which are affecting the composition and flow rates of the six defined products; LNG, LPG, Condensate, Fuel Gas, CO2 and Nitrogen. Six base cases made by Linde have been used to represent the system behaviour for all the years of production.

The necessary input data to perform a production forecasting with the model, are the composition of the Feed Gas on mol basis, the total Feed Gas flow rate in kmol/hr, the thermal power consumption in MW, the distribution coefficients (K-values) for the Feed Gas and the annual number stream days.

The calculated composition and flow rate of LNG, Fuel Gas and Condensate are believed to be quite accurate (<0.5 mol% error in composition and < 1% error in flow rate on mass basis compared to Linde base cases), while some uncertainties are connected to the LPG product (<2.1 % error in composition and < 5% error in flow rate on mass basis compared to Linde base cases).

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Group 2	Mass balance model	Massebalanse-model
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Institutt for energi- og prosessteknikk

EPT-H-2007-29

MASTEROPPGAVE

for

Stud.techn. Anders Lauvdal

Høsten 2007

Massebalanse-modell for Hammerfest LNG-anlegg

Mass balance model for Hammerfest LNG plant

Bakgrunn

Snøhvit-anlegget tar inn ubehandlet brønnstrøm fra gass/kondensatfelter ute i havet, skiller ut vann/glykol, CO_2, nitrogen, og andre uønskede komponenter, og behandler hydrokarbon-strømmen fram til de tre produktene fra anlegget: LNG, LPG, og kondensat. Samtidig forbrukes det hydrokarboner til drift av anlegget, inkludert kraft/varmeproduksjon og fakling.

I forbindelse med oppdateringer av produksjonsprofil fra gass/kondensatfeltene er det ønskelig å kunne gjøre enkle vurderinger av endringer i produktmengder og sammensetning av produktene, samtidig som en tar hensyn til prosessmessige behov og begrensninger. Hovedoppgaven skal utvikle og utprøve en enkel modell som skal benyttes i slike vurderinger, med tanke på planlegging av drift av anlegget, og prognosering av produksjonstall. I første omgang kan modellen baseres på bruk av konstante/variable "ekstraksjonsfaktorer" for de enkelte komponentene i brønnstrømmen, mens en på sikt vil tilstrebe mer fysisk korrekte modeller.

Oppgaven vil utføres i nært samarbeid med Snøhvit Drift.

Mål

Målet med oppgaven er å utvikle og utprøve en massebalanse-modell for hele Snøhvit-landanlegget som muliggjør balansering av alle strømmer inn/ut, og implementering av denne modellen i et regneark. Det skal også gjennomføres beregninger av produksjonsrater for LNG, LPG og kondensat basert på oppdaterte brønn- og feltproduksjonstall.

Oppgaven bearbeides ut fra følgende punkter:

1. Gjennomgang og oppsummering av prosesstrinn og strømmer i Snøhvit-landanlegget, og etablering av grunnlag og prinsipper for en massebalanse-modell. Spesielle hensyn som må tas her er krav til maksimal brennverdi for LNG-produktet, eventuelle andre begrensninger i produktspesifikasjoner, og realistisk modellering av fyrgassforbruk.

2. Utvikling av en massebalanse-modell og implementering av denne i et regneark, samtidig som modellen og regnearket dokumenteres på en god måte. Utprøving og tuning/modifikasjon av modellen basert på eksisterende prosess-simuleringsdata.

3. Gjennomføring av beregninger basert på nye reservoardata og feltproduksjonstall.

4. Vurdering av videre utviklingsmuligheter for modellen, inkludert bedre representasjon av utskilling av tyngre hydrokarboner, og implementering av kapasitetsgrenser for prosess og utstyr.

5. Forslag til videreutvikling og tuning av modellen basert på faktiske produksjonsdata.

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Senest 14 dager etter utlevering av oppgaven skal kandidaten levere/sende instituttet en detaljert fremdrift- og evt. forsøksplan for oppgaven til evaluering og evt. diskusjon med faglig ansvarlig/ veiledere. Detaljer ved evt. utførelse av dataprogrammer skal avtales nærmere i samråd med faglig ansvarlig.

Besvarelsen redigeres mest mulig som en forskningsrapport med et sammendrag både på norsk og engelsk, konklusjon, litteraturliste, innholdsfortegnelse etc. Ved utarbeidelsen av teksten skal kandidaten legge vekt på å gjøre teksten oversiktlig og velskrevet. Med henblikk på lesning av besvarelsen er det viktig at de nødvendige henvisninger for korresponderende steder i tekst, tabeller og figurer anføres på begge steder. Ved bedømmelsen legges det stor vekt på at resultatene er grundig bearbeidet, at de oppstilles tabellarisk og/eller grafisk på en oversiktlig måte, og at de er diskutert utførlig.

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Institutt for energi og prosessteknikk, 30. juli 2007

Johan Hustad
Instituttleder

Jostein Pettersen
Faglig ansvarlig/veileder

Kontaktperson(er)/medveileder(e):

Sivert Vist, Statoil - Snøhvit Drift

Preface

This Master theses has been performed at the Norwegian University of Science and Technology (NTNU) during the autumn 2007. Most of all, I would like to thank my co-supervisor Jostein Pettersen at StatoilHydro for his invaluable assistance. I would also like to thank Sivert Vist at Snøhvit Drift for valuable help in the beginning of the working process.

Anders Lauvdal
22.12.2007

Abstract

The objective of this work has been to develop a relatively simple Excel model that can be used for production forecasting purposes for Snøhvit LNG plant. Three mass balance models have been developed and tested, but only one of them has the sufficient accuracy to perform valid calculations. Effort has been put into making the model intuitive and user-friendly. The model has further been used to forecast the production from updated production profiles (December 2007).

The development of the model has been based on detailed descriptions of processes at Melkøya which are affecting the composition and flow rates of the six defined products; LNG, LPG, Condensate, Fuel Gas, CO_2 and Nitrogen. Six base cases made by Linde have been used to represent the system behaviour for all the years of production. Linde has produced this data by performing extensive process simulations on six different design feedstocks. Additional stream data and process flow diagrams have been used to obtain more specific information about the internal streams and units.

The necessary input data to perform a production forecasting with the model, are the composition of the Feed Gas on mol basis, the total Feed Gas flow rate in kmol/hr, the thermal power consumption in MW, the distribution coefficients (K-values) for the Feed Gas and the annual number stream days.

The calculated composition and flow rate of LNG, Fuel Gas and Condensate are believed to be quite accurate (<0.5 mol% discrepancy in composition and <1% discrepancy in flow rate on mass basis compared to Linde base cases), while some uncertainties are connected to the LPG product (<2.1 mol% discrepancy in composition and <5% discrepancy in flow rate on mass basis compared to Linde base cases).

Sammendrag

Målsetningen med denne oppgaven har vært å utvikle en enkel Excel modell som kan benyttes til å foreta produksjonsprognoser for Snøhvit LNG-anlegg. Tre massebalanse-modeller har blitt utviklet og testet, men bare en av dem har den nødvendige nøyaktighet til å foreta gyldige beregninger. Det har blitt lagt vekt på å gjøre modellen intuitiv og brukervennlig. Modellen har videre blitt benyttet til prognosere produksjonen basert på oppdaterte produksjonsprofiler (desember 2007).

Utviklingen av modellen er basert på detaljerte beskrivelser av prosesser ved Melkøya som påvirker sammensetningen og strømningsraten til de seks definerte produktene; LNG, LPG, Kondensat, Brenngass, CO_2 og Nitrogen. Seks basiscase laget av Linde har vært brukt for å representere oppførselen til systemet for hele anleggets levetid. Linde har produsert data gjennom omfattende prosessimuleringer for seks forskjellige fødegassammensetninger. I tillegg er det blitt brukt strømdata og prosessflytskjemaer for å få mer spesifikk informasjon angående interne strømmer og enheter.

De nødvendige inputdata for å gjøre produksjonsprognoser med modellen, er sammensetningen til fødegassen på molbasis, strømningsraten til fødegassen i kmol/time, termisk effektbehov, distribusjons koeffisienter (K-verdier) for fødegassen og antall produksjonsdager per år.

Den beregnede sammensetningen og strømningsraten til LNG, Brenngass og Kondensat er trolig temmelig nøyaktige (<0.5 mol% avvik i sammensetning og < 1% avvik i strømningsrate sammenlignet med Linde sine basiscase), mens noe usikkerhet er tilknyttet LPG-produktet (<2.1 mol% avvik i sammensetning og <5% avvik i strømningsrate sammenlignet med Linde sine basiscase).

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Abbreviations

LNG	Liquefied Natural Gas
LPG	Liquefied Petroleum Gas
HHC	Heavy Hydrocarbons
GHV	Gross Heating Value
C4(-)	Butane and lighter hydrocarbons
C5+	Pentane and heavier hydrocarbons
sd/yr	Stream day per year
wt ppm	Parts per million on mass basis
vppm	Parts per million on volume basis

1. Introduction

The LNG plant at Melkøya receives pipeline gas from three offshore gas fields. It is necessary to extract water, MEG, CO_2 and nitrogen from the Feed Gas in order to produce the three defined products LNG, LPG and Condensate. Some of the hydrocarbons from the pipeline stream are used as fuel gas to generate power and heat for the LNG plant. The production rates of the different products are depending on the Feed Gas composition due to the different production profiles for the three gas fields.

The production forecasting at Melkøya is at the moment based on extensive process simulations made by Linde. Due to updated production profiles, it is necessary to make new estimations for the production of LNG, LPG and Condensate. Regarding updated production profiles, it is desirable to make quick and simple estimations of the production without using advanced and time demanding methods. A relatively simple Excel model will be developed in this work to represent the plant behaviour and satisfy the criteria connected to product specifications and production limitations at Melkøya.

Scope of work

Excel Mass Balance Model
Effort will be put into making the mass balance model as independent as possible to be able to do calculations without the necessity of other programs. It is therefore not desirable to link/import data from for instance HYSYS. It will be important to test the accuracy of the mass balance model in order to evaluate the validity of the production forecasting. From the results it will hopefully be possible to determine if the Excel model is representative for the LNG process and can be used for production forecasting.

Methods for representing the LNG process in Excel
This work will mainly consider those processes at Melkøya where the composition and flow rates of the plant products are determined. Process simulation data will form the basis for the behaviour of the LNG process in the excel models. Additional simulations in HYSYS may be performed to investigate the behaviour of specific sections in the process.

Refrigerant cycles and refrigerant make up
The refrigerant make-up system will not be included in the mass balance models because the fraction of the Feed Gas used for this purpose is considered insignificant. The refrigerant cycles are not covered in detail because they are not affecting the product compositions directly; they are only providing the cooling energy necessary to produce the different products within their specifications. Only the power consumption connected to the refrigerant cycles are of interest.

Steady State
The models are to operate on a monthly/yearly steady state basis. The flow rates will be calculated for the plant at normal operation (100%). Storage and loading issues are not covered in detail because the plant is acting in a more transient way during this period.

This thesis is written under the assumption that the reader is familiar with the LNG process at Melkøya and with thermodynamics in general.

Structure

Chapter 2 is containing the process description of the LNG plant at Melkøya to establish the physical basis for the mass balance model. Then Chapter 3 follows with a description of the background data used in the model design. Chapter 4 is containing key figures for the production and the product specifications that the model have to achieve. The theory used in the development of the models is presented in Chapter 5. The descriptions of the model development is described in Chapter 6. Production forecasting is treated in Chapter 7. The developed models are use to estimate the production both for the outdated and updated Feed Gas composition. The results from the development of the models and the production forecasting can be found in Chapter 8 followed by a discussion about the results in Chapter 9. This is followed by Chapter 10 which presents the conclusions from this work. Finally, Chapter 11 contains recommendations for further work.

2. The LNG Process at Melkøya

2.1. General Description

Gas Treatment

The pipeline gas from the offshore fields needs extensive treatment in order to produce LNG. The feedstock from the production fields arrives in the *Slug Catcher* at the onshore LNG plant. MEG, water and condensate is separated from the natural gas in this unit. It is then necessary to remove CO_2 and remaining water from the gas phase which would otherwise freeze and cause plugging in the downstream process. Mercury must also be removed because it will cause destruction of the aluminium plate fin heat exchangers in combination with moist air.

LNG

The natural gas is pre-cooled, liquefied and sub-cooled in a series of heat exchangers (see Figure 3). A fraction of heavier hydrocarbons are extracted in the pre-cooling section. Nitrogen is reduced to a value below 1 mol% after the three stage liquefaction process and the LNG product is then routed to the LNG storage tanks. LNG will periodically be loaded onto LNG ships and the flash, displacement and boil-off gas due to loading will be sent back to the onshore LNG plant where it is compressed together with the LNG tank boil-off gas.

Refrigerant Make-Up and LPG

The heavier hydrocarbons extracted from the LNG liquefaction process are fractionated and used to produce the necessary Refrigerant Make-Up. LPG (mixture of propane and butane) is also produced in this fractionation process, stored in tanks and periodically loaded onto LPG carriers.

Condensate

The heavy ends of the fractionation (C5+) are joining the condensate stream from the inlet treatment facilities, stored in tanks and periodically loaded onto condensate shuttle tankers.[1]

2.2. Detailed Process Description

This section will form the basis for the Excel models developed in this work. Each section in this chapter is a describing main system at Melkøya which is of importance to the composition and flow rate of the different products. The block diagram attached in Appendix A is of valuable help to easier understand how the systems are connected together and where the main streamlines go.

2.2.1. System 11 – Slug Catcher and Pig Trap

The purpose of the *Slug Catcher* is to receive the three-phase feed stream from the pipeline, to buffer liquid slugs, to separate the gas phase and the two liquid phases, and to provide continuous gaseous and liquid feed to the LNG plant.

The *Slug Catcher* is designed for a three-phase flow of 876 838 Sm^3/hr from the pipeline. For normal operation of Snøhvit A the feed from the pipeline is divided into the following streams.

Natural gas	820 431	Sm^3/hr
Condensate	56 406	Sm^3/hr
MEG/Water	10 164	Sm^3/hr

The inlet pressure varies between 70 – 90 Bara. The inlet pressure may go down to 35 bar after several years of production. The inlet temperature may vary between -5 °C and +4 °C.

After several years of operation, the Slug Catcher inlet pressure will drop below 70 Bara. A Feed Gas compressor will then be installed downstream of the Slug Catcher, which increases the pressure from a minimum of 35 Bara to 70 Bara. An inlet pressure of 70 Bara is required for the liquefaction of the natural gas. [1]

2.2.2. System 12 – Inlet Facilities

The gas stream and liquid stream from the *Slug Catcher* are further treated in System 12. The gas is heated to a temperature that prevents hydrates from forming in the subsequent gas expansion. Liquid droplets and solid particles are removed in the *Inlet Filter Separator* (12-CB-101 A/B) to satisfy requirements in System 22 – CO_2 Removal.

Treatment of Gas Phase

The gaseous feed steam from the *Slug Catcher* enters System 12 at a pressure between 70 – 90 Bara and temperatures between -1 to +4 °C. The temperature will be 2 – 5 °C above the temperature for which hydrates may form. To prevent hydrates from forming during expansion in the control valve (12-FV-1140), the gas is heated upstream of the expansion process in the *Inlet Gas Preheater* (12-HA-102). The pressure downstream this valve is more or less constant because of the required

pressure in the liquefaction process, but the pressure upstream may vary significantly (70 – 90 Bara, maximum 115 Bara). Therefore the temperature drop due to expansion will vary. The temperature downstream the control valve shall be kept constant independent of the Feed Gas pressure. The control valve temperature is therefore used as signal to regulate the preheater duty.

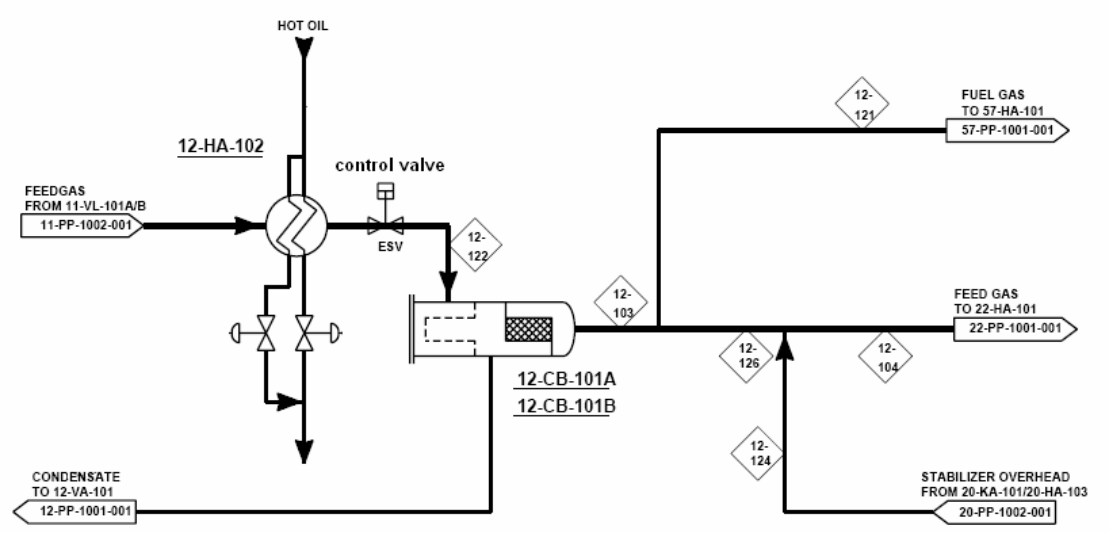

Figure 1: Main stream lines for the gas phase in System 12 (edited from PFD)

The pressure of the Feed Gas downstream the expansion valve is controlled by the *HHC Removal Reflux Drum* 25-VD-107 from Chapter 2.2.7. The pressure after the control valve is fixed at 68 Bara and independent of the Feed Gas pressure upstream of the valve.

After the expansion the liquid droplets and solid particles are removed in the *Inlet Filter Separator* (12-CB-101 A/B). Downstream the filter separator, a part of the Feed Gas is taken out and sent to the Fuel Gas system (normally 48 000 Sm^3/hr, respectively 40 t/hr). The remaining part of the Feed Gas is mixed with Stabilizer Overhead gas from System 20. The flow rate of the gas from System 20 depends mainly on the content of heaver hydrocarbons in the condensate stream from the *Slug Catcher*. The mixed gas is then routed to System 22 – CO_2 Removal. [1]

Condensate Treatment

The condensate stream from the *Slug Catcher* is first heated in the *Inlet HC Condensate Preheater* (12-HA-101) to prevent hydrates from forming in the subsequent expansion. The condensate temperature before entering the expansion valve is depending on the pressure upstream the preheater. The condensate is then expanded to a pressure of 20 Bara.

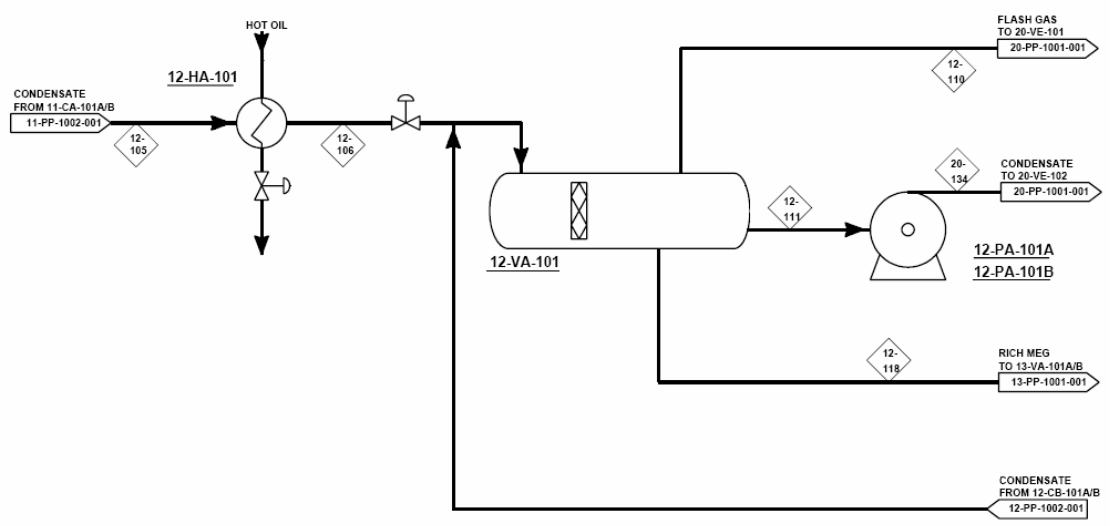

Figure 2: Main stream lines for the liquid phase in System 12 (edited from PFD)

After the expansion, the condensate is sent to the Condensate Separator (12-VA-101). This is a three phase separator where the flash gas (top product) and condensate with low water content is routed to System 20 – Condensate Treatment. The heavy liquid phase consisting of rich MEG is sent to System 13 – MEG Recovery (not covered in this work). [1]

2.2.3. System 20 – Condensate Treatment

The intention of this system is to remove MEG and lighter hydrocarbons from the condensate stream coming from System 12. Hydrocarbons lighter than C5 are removed in the Stabilizer Column (20-VE-101) in order to produce a stabilized condensate product. The overhead gas from this column is then compressed and recycled back to System 12 where it is mixed into the Feed Gas stream. The C4(-) content in the bottom of the stabilizer shall be lower than 2.5 wt%. The Condensate product is then routed to System 43 – Condensate Storage.

The MEG Removal Column (20-VE-102) in System 20 removes the traces of MEG in the condensate to a value of 2 wt ppm (max value < 10 wt ppm) to satisfy Condensate specifications.

There are hydrocarbon streams formed in the plant at a lower pressure than the Feed Gas pressure, which are to be recycled back to the Feed Gas stream. Flash Gas from 12-VA-101 and light hydrocarbon fractions from System 26 (stream 26-105 in block diagram) are mixed to the top section of the stabilizer. These two streams are mostly containing light hydrocarbons and will therefore leave the top of the stabilizer and be recycled back to System 12.[1]

2.2.4. System 21 - Mercury Removal

The purpose of the mercury removal unit is to remove mercury to a satisfying level to protect the downstream units from mercury attack. This unit is not included in the mass balance model because the amount of mercury compared to the amount of hydrocarbons is insignificant. [1]

2.2.5. System 22 – CO2 Removal

The CO_2 content in the LNG can only be 50 vppm (0.005 vol%) for any Feed Gas case to avoid CO_2 to freeze out in the liquefaction process. The CO_2 is separated from the gas using amines in an absorption process. The amine used is called MDEA (methyldiethanolamine). The system is designed for a Feed Gas stream of 823 285 Sm^3/h with a CO_2 content of 5.3 vol% (Snøhvit A). The extracted CO_2 is routed to System 24 – Drying and Compression form where it is transported with pipeline for re-injection.

For material balance purposes in the mass balance models, the CO_2 content in the LNG product is fixed to be 0.005 vol% because this is an absolute criterion that has to be fulfilled under any circumstance. [1]

2.2.6. System 23 – Dehydration

The Feed Gas coming from System 22 – CO_2 Removal is saturated with water and needs to be dried to prevent water from freezing out in the subsequent cryogenic process. Feed Gas from System 22 contains about 700 mol ppm of water and this amount is reduced to 0.1 mol ppm in a molecular sieve drier.

System 23 is not included in the mass balance model because it is only a technical criterion that has to be fulfilled to make the process run. The amount of water removed is insignificant. [1]

2.2.7. System 25 – Natural gas liquefaction

Main functions:
- Receive Feed Gas from System 21 and route it to the pre-cooling process
- Extract heavier hydrocarbons and lead these to System 26
- Lead the light natural gas through the liquefaction and sub-cooling heat exchangers
- Extract nitrogen from the natural gas and lead the LNG to System 42

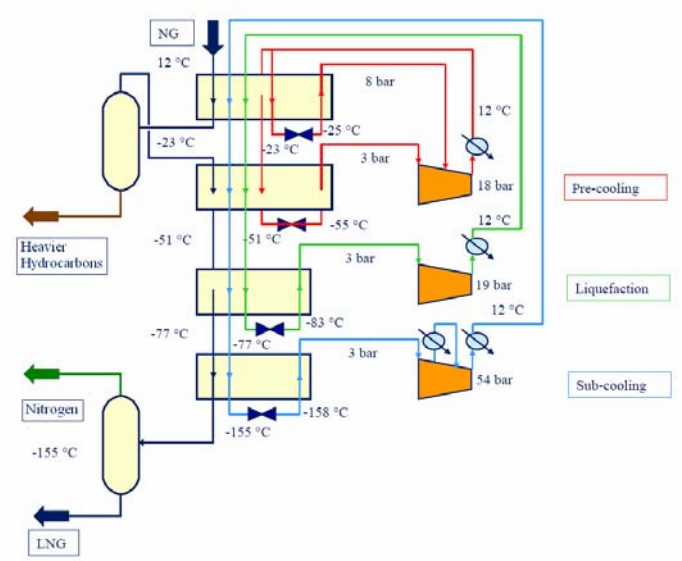

Figure 3: Flow diagram of System 25

Heavy Hydrocarbon (HHC) Removal Column 25-VE-101

The Feed Gas from System 21 enters System 25 at 61 Bara and 27°C and is cooled in the *Treated Gas Water Precooler* (25-HA-101). The gas is cooled against sea water down to a temperature of 13 °C before it goes through the warm section of the *Treated Gas Precooler* (25-HG-101). The natural gas outlet temperature is important for the GHV of the gas leaving the top of the *HHC Removal Column* (25-VE-101) and the set point for this temperature is depending whether the Feed Gas composition is light or heavy. [2]

Reflux

The lighter hydrocarbons exiting through the top of the column are directed back to the cold part of the *Treated Gas Pre-cooler* where it is further cooled down to -53 °C. Some of the natural gas is condensed and taken out as liquid in the *HHC Removal Reflux Drum* (25-VD-107) and sent back as reflux to the lower part of the upper section of the *HHC Removal Column*. The temperature where the reflux enters is just -30 °C and to avoid instabilities, a section for intense mixing is installed in the column to give a smooth temperature profile. The reflux provides removal of the heavy hydrocarbons. Except of too high nitrogen content, the vapour leaving the reflux drum (stream 12-144) is almost at LNG specification. [2]

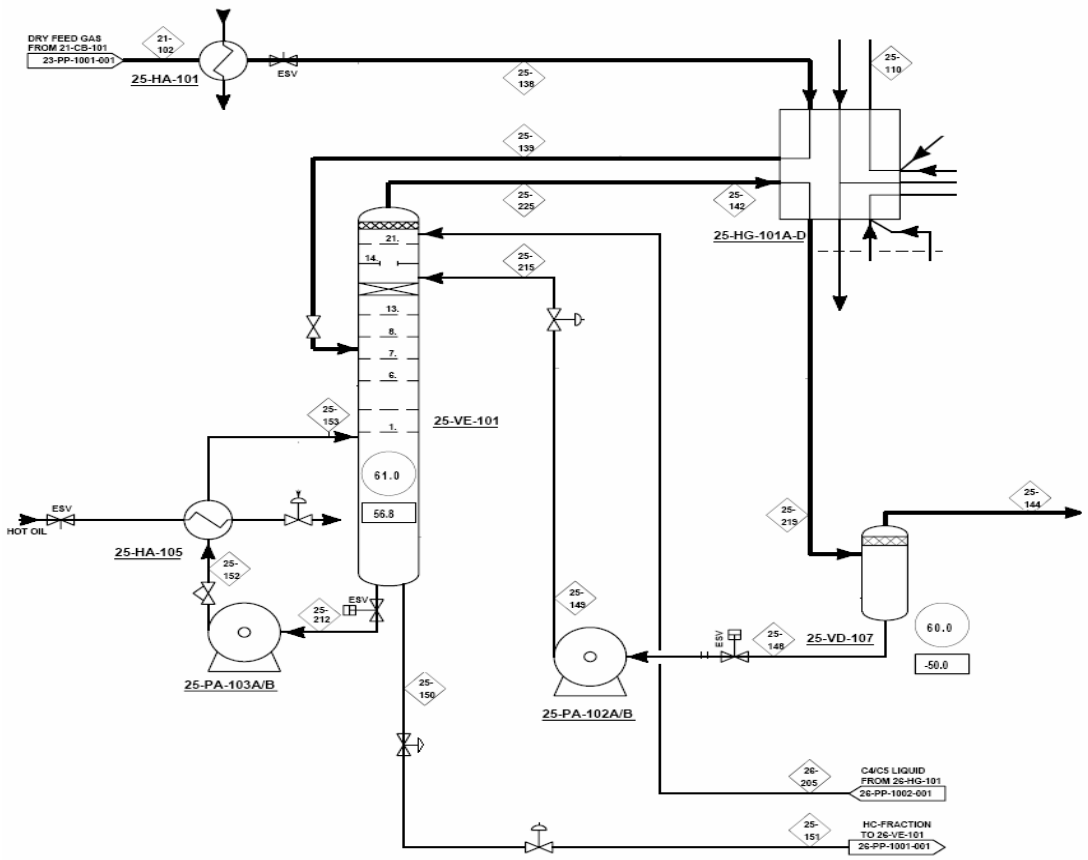

Figure 4: HHC Removal Column (edited from PFD)

GHV Control

The natural gas leaving the *HHC Removal Reflux Drum* for liquefaction need to have a GHV of 39.0 MJ/Sm^3, which result in a GHV of 40.2 MJ/Sm^3 at the loading terminal. It is necessary to remove propane from the top product in order to satisfy the GHV requirements. To adjust the GHV and control the amount of propane, a reflux from System 26 is directed into the top of the *HHC Removal Column* with a lower temperature than the gas leaving the column. The reflux consists mainly of butane and pentane and the resulting temperature profile contributes to condense residues of propane. Propane has a large solubility in the butane/pentane liquid and therefore this reflux contributes to remove propane from the Feed Gas. Insignificant amounts of butane and pentane evaporates at this stage and most of it runs down the column.

The function of the lower part of the column (below the feed inlet tray) is to remove as much methane from the bottom product as possible to reduce the recycled amount of methane in the system (see C1/C2 stream in Chapter 2.2.8). The design methane content for the bottom product is set to maximum 21 mol%. The bottom product is routed to System 26 for LPG and condensate fractionation. The reboiler duty is depending on whether the feed is a light or a heavy. [2]

Liquefaction and Sub-Cooling

The liquefaction and sub-cooling process will not be further treated in this section of reasons stated in Chapter 1.

N2 Removal Column 25-VE-102

The intention of the Nitrogen Removal Column, 25-VE-102, is to remove nitrogen and adjust the nitrogen level in the LNG-product to 1 mol%. The column works as a stripping column where the feed enters in the top and the boiler provides the necessary gas flow in the column.

The bottom product is a mixture of the following streams going into the column:
- LNG-liquid from *N2/CH4 Heat Exchanger* (25-HX-103) (about 3 wt%)
- LNG-liquid from System 27 - N2 Removal (about 5 wt%)
- LNG-liquid from *N2 Removal Reboiler* (25-HG-103) (about 92 wt%)

The LNG coming from the heat exchanger for sub-cooling is routed through the reboiler (25-HG-103) and the expansion turbine (25-CT-102) where the pressure is reduced to 0.2 Barg. Some of the LNG evaporates in the expansion turbine and this amount of gas contributes to 50 % of the total amount of gas going out of the top of the *N2 Removal Column*.

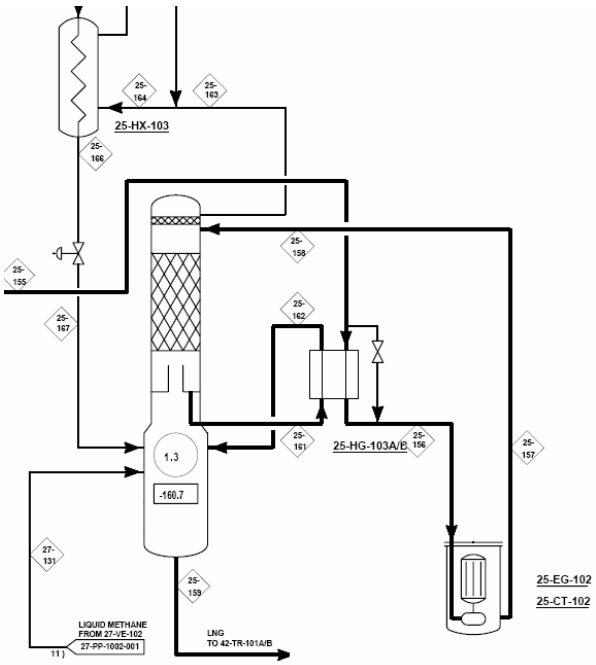

Figure 5: N2 Removal Column (edited from PFD)

To generate the necessary internal vapour flow the N2 column is equipped with a reboiler. LNG is withdrawn from a column tray located above the column bottom section. It is then heated to produce a vapour fraction of 3 wt% and routed back to the bottom section of the column below the tray it was withdrawn from. Heat from the LNG (25-155) is used as reboiler energy. This provides a further sub-cooling of the LNG before it enters the expansion turbine.

The LNG going into the top of the column has a higher content of nitrogen than the LNG in the trays below. By adding heat to the LNG in the lower section, the evaporated LNG will get a lower level of nitrogen than the LNG fed into the top section. This makes the driving potential of the N2 stripping process. As the LNG flowing downwards makes contact with LNG vapour flowing upwards, the vapour gradually becomes more enriched with nitrogen because of the stripping of nitrogen from the LNG liquid.

The LNG (-160.9 °C and 4.5 Bara) exiting through the bottom of the column is routed to System 42 - LNG Storage (Stream 25-160 in block diagram)

The gas going out through the top section is mixture of nitrogen and methane. This stream is routed *to* System 27 – N2 Removal. [2]

2.2.8. System 26 – Fractionation and Refrigerant Make Up

The main design intention of this system is to fractionate the bottom products from the *HHC Removal Column* in System 25 into a LPG product, a condensate product, a C1/C2 recycle stream and a C4/C5 rich stream which is used as top reflux in the *HHC Removal Column* described in Chapter 2.2.7. A second design intention is to produce high purity methane, ethane and propane which are used for Refrigerant Make-Up.

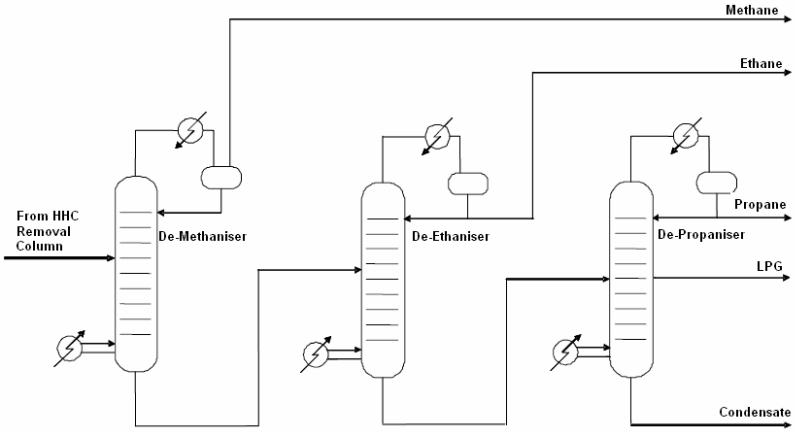

Figure 6: Product streams in System 26

De-Methaniser (26-VE-101)

The *De-Methaniser* receives the bottom stream from the *HHC Removal Column* and will produce a nearly methane free bottom product and a methane top product with a

minimum ethane content (20 mol%) to be used as methane refrigerant make up. The surplus gas (C1/C2) which is not used for refrigerant make up is routed to *System 20* (recognized as stream line 26-105 in the block diagram). [1]

De-Ethaniser

The purpose of this column is to produce a nearly ethane free bottom product (maximum 0.3 mol% ethane) and a high purity ethane top product to be used for ethane refrigerant make up. [1]

De-Propaniser

The purpose of the *De-Propaniser* is to produce a light condensate bottom product (maximum 2.5 wt% C4-), withdraw a side stream to produce the LPG product and a high purity vapour and liquid top product which is used as propane Refrigerant Make-Up. The column is operated differently depending on whether the Feed Gas is light of heavy.

Light Condensate

The bottom product from the *De-Ethaniser* enters the *De-Propaniser* in the middle section. As the liquid drops down to the bottom of the column, butane is stripped of until the design content of 2.1 wt% C4(-) is achieved. This bottom product is recognized as stream 26-137 in the block diagram.

Column Overhead Stream

The overhead stream from the column is almost pure propane (2.9 mol% ethane, 96.5 mol% propane and 0.6 mol% butanes for Snøhvit A). One portion of the overhead stream is used for Propane Refrigerant Make-Up, one portion is used as reflux in the *De-Propaniser* to achieve the desired purity of the column overhead stream and the surplus is mixed together with the LPG product.

LPG

The LPG product is withdrawn as a side stream from the upper section of the column and the flow rate will vary for different to Feed Gas compositions. The LPG is then mixed with the surplus from the column overhead stream, sub-cooled to a temperature of -34 °C and routed to the LPG storage tank. The LPG stream going to the storage tank is recognized as stream 26-124 in the block diagram and LPG specifications that must be satisfied are listed in Table 4. [1]

Benzene Removal Column

A vapour flow is withdrawn from the lower section of the *De-Propaniser* and routed to the *Benzene Removal Column*. The intention of this column is to produce a C4/C5 enrich overhead stream with a low benzene concentration which is used as top reflux in the *HHC Removal Column* to reduce the GHV of the LNG. The composition of the C4/C5 overhead stream is approximately 0.5 mol% propane, 60 mol% butanes and 39,5 mol% pentanes during normal operation for Snøhvit A feedstock. The overhead stream is sub-cooled down to -34 °C before it can serve as reflux in the *HHC Removal Column*. The bottom product from the *Benzene Removal Column* is sent back to the *De-Propaniser* to the stage from which it was withdrawn. [1]

Refrigerant Make-Up

The pre-cooling, liquefaction and sub-cooling cycles are closed loops and under normal conditions the Refrigerant Make-Up streams are insignificant. The Refrigerant Make Up is generated from the overhead streams of the De-Methaniser, De-Ethaniser and the *De-Propaniser* where more or less pure methane, ethane and propane respectively are extracted. Refrigerant Make Up is necessary to compensate for refrigerant losses, for adjusting the optimal refrigerant composition and for initial filling or refilling after major refrigerant losses [1].These minor flow rates are neglected in this work and not considered in the development of the mass balance models.

2.2.9. System 27 - Nitrogen Removal

The intention of this system is to recover methane from the Nitrogen Removal Column overhead stream in System 25. This is done in the Nitrogen Removal Unit (NRU). The nitrogen is purified and released to the atmosphere while the recovered methane is sent back to System 25. Nitrogen released to atmosphere is recognised as the Nitrogen product in the data from Linde and is represented in the block diagram with the stream number 27-127. The NRU is also handling the LNG boil-off gas, but as stated in Chapter 1 this effect has not been considered. [1]

3. Background Data

Feedstock Data

The mass balance models developed in this thesis are based on six different base cases developed by Linde. These cases state the material balance for the expected products according to different production profiles. The data is given in the Excel document "*Material balance component.xls*" and is described in the report "*Snøhvit LNG Project Calculated LNG, LPG and Condensate yearly production"* [3]. The results from Linde are based on an extensive process simulation for the whole system at Melkøya. The six different feedstock compositions used as basis for these process simulations are illustrated in Figure 7 with the names 100 % Snøhvit, #1, #2, #3, #4 and #5. More detailed data of composition and flow rate of these feedstocks are attached in Appendix B. Other names have been used in this work to describe the same feedstocks mentioned above. These are according to the names in the attachment which are Case 01, 2007, 2013, 2015, 2020 and 2022 referring to the different years.

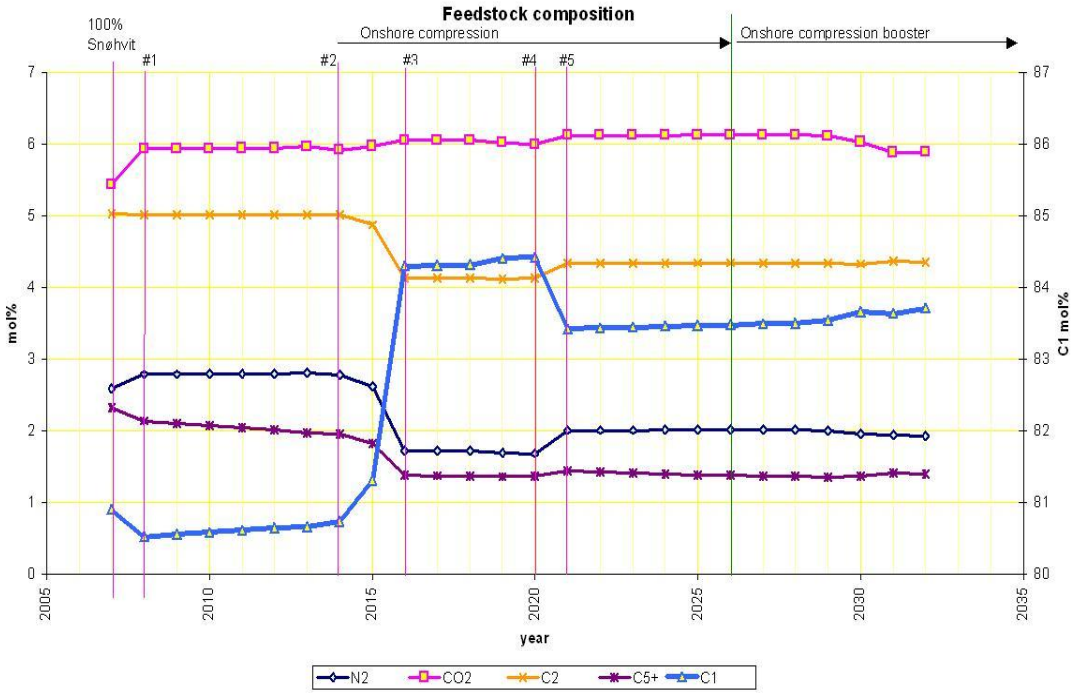

Figure 7: Composition of the Feed Gas from 2007 to 2035 [3]

System descriptions

PFDs (Process Flow Diagrams) of Melkøya LNG plant has been used to get better insight in the internal process lines within the main systems in the block diagram. These PFDs are internal documents in Statoil and are therefore not attached in appendix or fully illustrated. The segments from the PFDs shown in some of the

sections in this document are edited, simplified and does only show the main stream lines.

Stream data describing the internal streams in *System 11, 12, 25* and *26* in the block diagram has been applied to make valid assumptions for the system behaviour. The data is collected from PIM (Process Information Management) which is a data base with data for all the Statoil projects. The cases used are Case 01, Case 07, Case 11 and Case 13. These names correspond to Snøhvit A, Snøhvit B, Snøhvit C and year 2013 respectively [4]. The cases are describing the production at design capacity with no loading operations . Snøhvit B and Snøhvit C do not correspond to any of the Feedstocks presented in Figure 7. The data is internal and is therefore not attached.

4. Design Basis

The contents in this chapter is based on a document called "*Snøhvit LNG Overall Technical Design Basis*" [5].

4.1. Feed Gas

The pipeline Feed Gas is produced from the three offshore fields Snøhvit, Albatross and Askeladd. The initial reservoir composition can be seen in Table 1.

Table 1: Initial reservoir composition [mol%]

	Snøhvit	Albatross	Askeladd
N_2	2,526	3,554	0,88
CO_2	5,262	7,932	5,699
C_1	81,027	78,844	86,88
C_2	5,028	4,975	3,59
C_3	2,535	2,196	1,25
IC_4	0,4	0,32	0,19
NC_4	0,829	0,67	0,35
C5+	2,393	1,509	1,161

The gas in the Askeladd field is a light feedstock because of the high fraction of methane (light component) and the low fraction of C5+; also referred to as heavier hydrocarbons in this work. The gas in the Snøhvit and Albatross fields are heavier feedstocks as can bee seen from the fractions of methane and C5+.

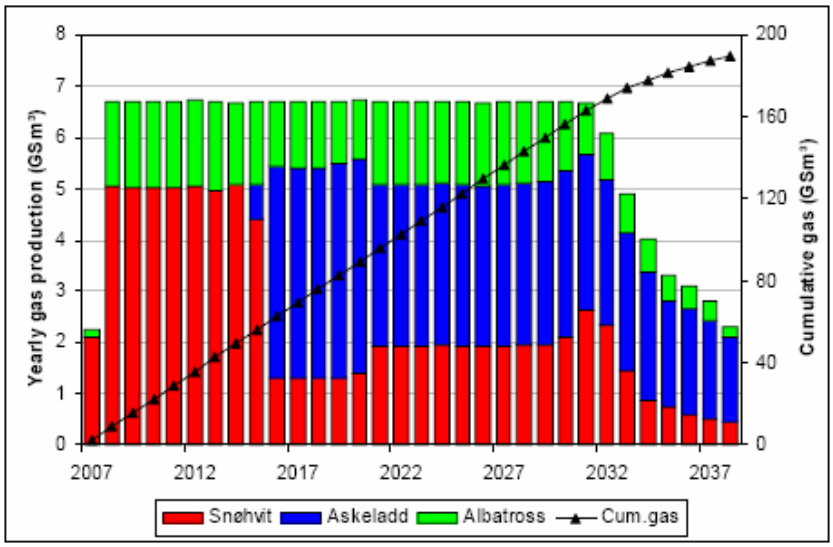

Figure 8: Scheduled production from Snøhvit, Albatross and Askeladd [5]

The feedstocks from the three reservoirs are mixed and led by pipeline to Melkøya. The Feed Gas composition entering the plant (Figure 7) is depending on how the reservoirs are produced (Figure 8). The well stream compositions will also change over time as a result of decreasing pressure in the reservoirs.

4.2. Key Figures

Design Feed Gas flow (MEG/water free basis):	20.8	MSm^3/sd
Number of yearly days of operation:	331	sd/yr

Three sales products are produced from the Feed Gas at Melkøya:

- LNG	4.20	MT/yr
- LPG	0.20	MT/yr
- Condensate	0.45	MT/yr

Three other products will also be produced from the Feed Gas
- Fuel Gas
- CO_2
- Nitrogen

The Fuel Gas is used in the gas turbines to power the LNG plant, the CO_2 is re-injected into subsea reservoirs and the Nitrogen is released to the atmosphere.

4.3. Product Specifications

The product specifications for LNG are strict in order to obtain the right properties for the sales product and to protect the equipment used to produce it. There are also restrictions for the LPG and Condensate but they are not as strict as for LNG. The specifications in the mass balance models will be, as far as possible, according to the specifications from the technical design basis document [5].

There are also other specifications concerning components which may damage equipment if the content is too high. These are not included because they are just trace components and will not influence product compositions

4.3.1. LNG Specifications

The most important specifications for LNG composition that the model need to satisfy, are the ones listed in Table 2. There are also restrictions for trace components, but they are not affecting the model calculations.

Table 2: LNG specifications

In mol%	Min.	Max.
Nitrogen	0	1,00
Methane	84,55	100,00
Ethane	0	9,20
Propane	0	3,25
I-Butane	0	0,60
N-Butane	0	0,75
C5+	0	0,15
CO_2	0	0,01

Table 3: LNG specifications at loading arm at Melkøya

At loading arm outlets	Ideal Gas calculation [MJ/Sm^3]	
	Min.	Max.
GHV [25°C;V(15°C;101,325 kPa)]	38.1*	40.2

*LNG plant will not produce LNG with GHV less than 39 MJ/Sm^3

4.3.2. LPG Specifications

The LPG product specification is a mix of propane and butane with preferably maximum 50 % of butanes total and 17 % I-Butane content.

Table 4: LPG product specification

Component	**Specification**
N_2	trace
CO_2	max 0.1 mol%
Methane	max 0.05 mol%
Ethane	max 1.0 mol%
C5+	max 2.0 mol%

4.3.3. Condensate Specifications

The only specification relevant for the mass balance models is the content of butanes and lighter material. C4(-) has a maximum value of 2.5 wt%

5. Theory

5.1. General Theory

Gas calculations are based on ideal gas behaviour and the ideal gas law has been applied. It is assumed that the reader is familiar with thermodynamics in general (refer to *Moran Shapiro – Fundamentals of Engineering Thermodynamics* for further reading).

Thermal Power

$$Q = \dot{m}_f \cdot GHV_f \tag{6.1}$$

where

Q is the thermal power consumptions in gas turbines [J/s]
$\dot{m}_f$ is the fuel flow rate on mass basis [kg]
GHV_f is the Gross Heating Value of the fuel on mass basis [J/kg]

5.2. Split Factors

The split factors are defined as the ratio of the mol flow of component *i* in the product to the material flow of component *i* in the Feed Gas:

$$s_{i,j} = \frac{\dot{n}_{i,j}}{\dot{n}_{i,feed}} \tag{6.2}$$

Where

$s_{i,j}$ is the split factor for component i in product j
$\dot{n}_{i,j}$ is the molar flow of component i in product j
$\dot{n}_{i,feed}$ is the molar flow of component i in the Feed Gas

Split Balance (conservation of mass)

$$\sum_{j=1}^{N} s_i = 1 \tag{6.3}$$

where

N is the total number of products

5.3. Gross Heating Value (GHV)

The Gross Heating Value (or *Higher Heating Value*) is calculated according to ISO 6976 (second edition 1995-12-01, table 3) with a reference temperature (combustion) of 25 °C [2]. The GHV is calculated at standard condition which means t_2 = 15 °C and p_2 = 1 atm = 101325 N/m^2 [6]. The formulas for calculating the GHV from ISO 6976 [7] are attached in Appendix C.

5.4. Vapour-Liquid Equilibrium Fundamentals

Chemical components can be separated in distillation columns when there are differences in the concentration of these components in the liquid and vapour phases. Vapour-liquid equilibrium data, also called VLE data, are vital for distillation design.

Binary systems consisting of two chemical components form a set of VLE data expressing the dew points and bubble points as function of composition, pressure and temperature at equilibrium. Liquid compositions are usually expressed as the mol fraction of the light component where x is used as symbol. Vapour composition is expressed as the mol fraction of the light component where the symbol y is used. The VLE data are usually presented in phase diagrams. The most relevant diagram is the T-xy diagram because the pressure is more or less constant in all the cases of separation treated in this work.

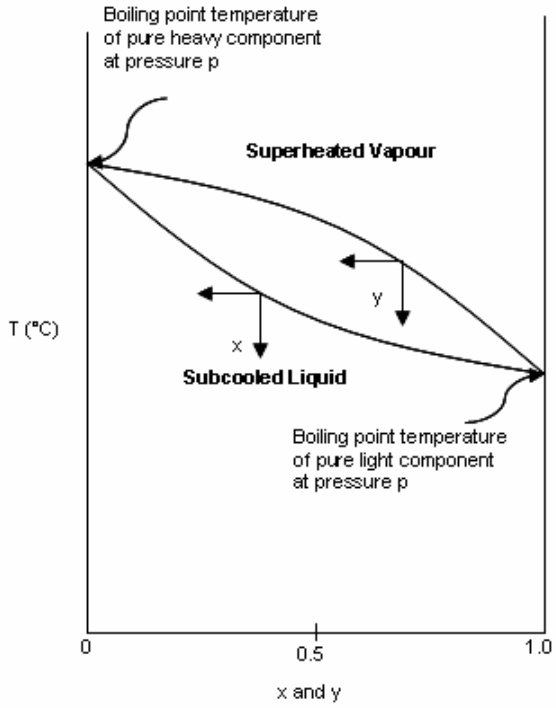

Figure 9: Temperature vs. composition of binary mixture at constant pressure

To determine the composition of liquid and vapour at a given pressure one can draw a horizontal line at the given temperature and read the corresponding x and y values. [8]

5.5. Distillation Fundamentals

Figure 10 illustrates an ordinary distillation column with one feed (**F**) introduced in the middle section of the column and a distillate product (**D**) in the top and a bottom product (**B**).

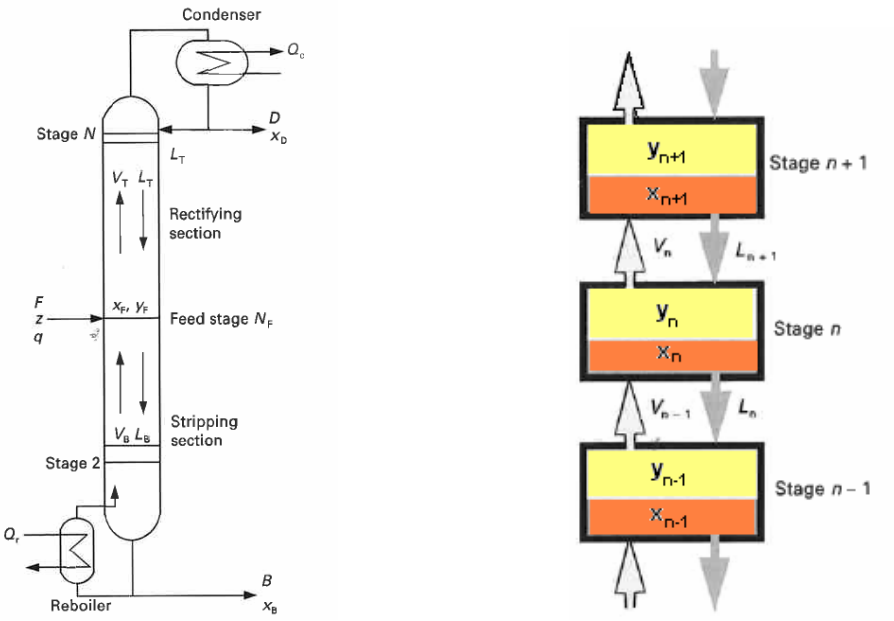

Figure 10: Two product distillation column (left) with column sections (right)

The column is divided into a certain number of stages which promotes the mass transfer of light components into the vapour flowing up the column and of heavy components into the liquid flowing down the column. The reboiler in the bottom section generates the vapour flowing upwards while the condenser located in the top of the column provides a liquid reflux stream by condensing the top product. The reflux is returned to the top stage and provides the necessary liquid flow downwards the column. The temperature in the column has a decreasing profile from bottom to top. Vapour-liquid equilibrium determines the vapour and liquid composition on each stage. According to vapour-liquid equilibrium theory from Chapter 5.4 the heavier components in the gas flowing upwards will condense as a result of reduction in temperature on each stag. The product exiting through the top will only consist of the lightest components while the heavier components are flowing downwards with the liquid flow to the bottom product. The purity of the top product is depending on the reflux ratio. Higher reflux ratio will increase product purity but will also increase the power consumption in the condenser. [8]

5.6. Distribution Coefficients

Distribution coefficients (K_j) can be applied to calculate the composition of the vapour and the liquid in a vapour-liquid system at equilibrium. For each component, the K_j value is defined as the ratio of vapour composition (y_j) to liquid composition (x_j). [8]

$$K_j = \frac{y_j}{x_j} = \frac{P_j \gamma_j}{P_T} \tag{6.4}$$

where

P_j is the vapour pressure of *j*th component in any pressure unit

P_T is the total system pressure in any pressure unit

γ_j is the activity coefficient of the *j*th component in the liquid phase at the conditions of temperature and composition of the liquid. γ_j = 1 for "ideal" systems.

5.7. Isothermal Flash Calculations

Combining vapour-liquid equilibrium relationships with total mass and component balances leads to the equations for calculating the products from an isothermal flash process. A material flow of known composition z_j is fed into a drum at a given rate of F kmol/hr. The drum operates at constant temperature and pressure. The flow rate and composition of the vapour and liquid fraction are unknown. [8]

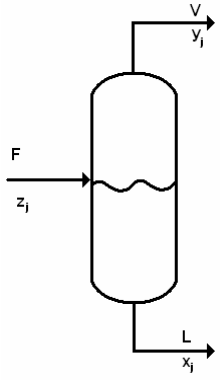

Figure 11: Isothermal flash

The equations describing the system are:

$$F = L + V \tag{6.5}$$

$$z_j F = x_j L + y_j V \tag{6.6}$$

$$y_j = x_j \frac{P_j}{P_T} \tag{6.7}$$

6. Model Development

6.1. Microsoft Excel

The mass balance models are to be made in Microsoft Excel. This program is suitable for making relatively simple calculation models which are intuitive and easy to operate if it's done in the right way. Larger and more complex models are more difficult to represent in an Excel sheet because the operations/formulas are hidden behind the interface and it can be difficult to get a clear overview. In some cases it may be necessary to use programming because Excel has its limitations when it comes to more advanced problem solving. For this purpose Visual Basic (VB) can be used in combination with Excel, but applying VB will make the program less intuitive.

In consultation with Snøhvit Drift it has been found most favourable to make an intuitive model with sufficient accuracy to perform valid calculations for product compositions and flow rates. Visual Basic has therefore been used to a limited extent.

The model will be made for steady state conditions as stated in Chapter 1, but it is necessary to determine whether the steady state condition should be on a monthly or a yearly basis. Transient behaviour due to LNG storage and loading has been left out of this work, but variations in Fuel Gas consumption as a result of seasonal variations may be necessary to include and is therefore discussed in the following chapter.

6.2. Yearly Variation in Fuel Gas Consumption

The Fuel Gas is mainly used to fuel the five LM6000PD gas turbines covering the power demand at Melkøya. As described in Chapter 2.2.2, a part of the Feed Gas in System 12 is separated out and used as Fuel Gas. Vaporized LNG can alternatively be used in situations when Feed Gas is not available. Feed Gas will be used as fuel during normal operation [1]. Since the Fuel Gas is produced from the Feed Gas stream the production of the other products will be affected as a result of variations in the Fuel Gas consumption.

6.2.1. Parameters Influencing Fuel Gas Consumption

This chapter is based on the report *Reference Data for Power and Heat Balance* [9]. There are several parameters treated in the report influencing the Fuel Gas consumption. The ones that are most important are described below:

Ambient Air Temperature

Ambient air temperature is directly connected to the power output from the gas turbines (see figure in Appendix B). A reduction in power output makes it necessary to increase the Fuel Gas consumption in order to maintain the power production at the same level.

Sea Water Temperature

The sea water temperature is affecting the power consumption of the compressors in the pre-cooling, liquefaction and sub-cooling cycle. These cycles are seawater cooled which can be seen from for Figure 12. An increase in sea water temperature will make it necessary for the compressors to increase the duty to get a higher temperature lift in order to transfer the heat from the refrigerant cycles to the sea water. The power consumption in these compressors sums up to about 70 % of the plant overall power consumption. The change in plant overall efficiency as function of the sea water temperature is found in Appendix B. It can be seen from the curve that the plant loses about 1% efficiency when the sea water temperature goes up one degree.

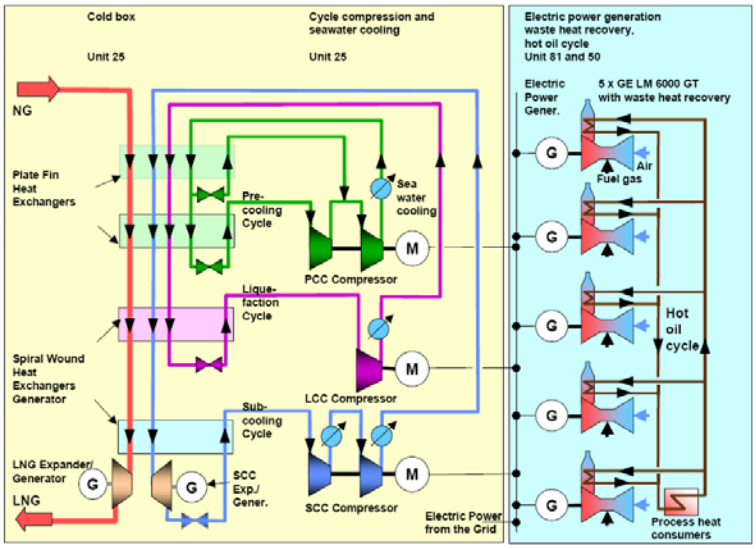

Figure 12: Power supply and demand at Melkøya

Ageing

The plant power demand will increase due to degrading. This is called the aging factor in the figure in Appendix B. Every third year the plant will have a hot section repair where the power losses due to aging will be reduced (see Appendix B).

Loading Factor

The efficiency of the system is not linear with the loading factor. At reduced load, pressure losses in the process lines are reduced by the reduction in flow rate and consequently the necessary delivery pressure from the compressors is lower. The loading factor is not considered in this work because of reasons stated in Chapter 1.

Variation in Fuel Gas Consumption

The data for the Fuel Gas consumption (in MW) for the first twelve years of production has been provided from the power and heat balance report. The shape of the curve in Figure 13 is a result of factors described in the previous sections (except loading factor). The first six years covers production without Feed Gas compression and the next six years are with Feed Gas compression. The growing tendency in Fuel Gas consumption is a result of the plant ageing factor and the influence from seasonal variations in air and seawater temperature can be seen from the curve oscillation. The increasing tendency of the curve goes a bit down after the 36th month as a result of the hot section repair. Two red lines are also drawn to show the average fuel consumption for the two periods. This graph has further been used in this work to evaluate the significance of seasonal variations in Fuel Gas consumption and to decide the time resolution for the model calculations.

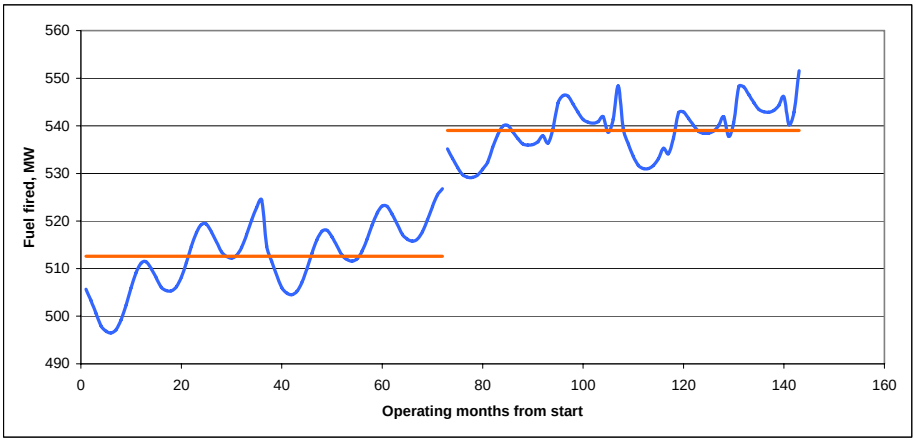

Figure 13: Resulting fuel consumption (month 1 is October)

6.2.2. Significance of Seasonal Variations

Variations in the Fuel Gas consumption will directly affect the LNG production because the two products are taken from the same feedstock. An increase in Fuel Gas consumption leads to roughly the same decrease in LNG production since the two products have almost the same composition. Equation 6.1 states that for a constant GHV the Fuel Gas flow rate will be proportional to the thermal power consumption. A rough calculation applying this relationship can provide a reasonable estimation of how the Fuel Gas and *LNG* flow rates changes with varying thermal power demand. Equation 6.1 on mass basis can therefore be rewritten into:

$$c \cdot \dot{m}_f \cdot GHV_f = Q \cdot c \qquad (7.1)$$

c is the percentage change in fuel fired compared to year average value

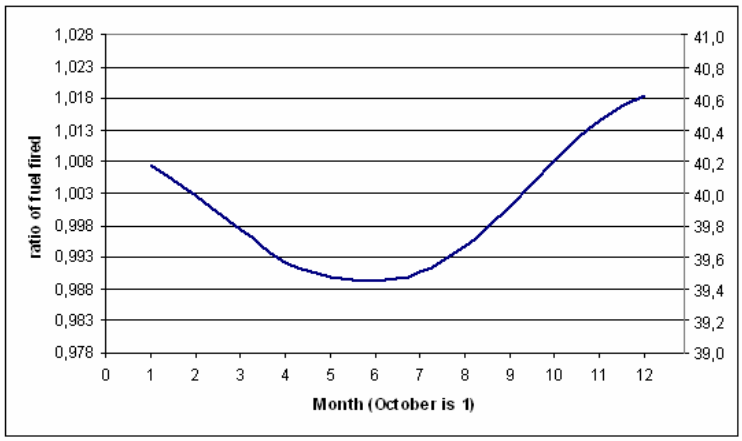

Figure 14: Variations in Fuel Gas consumption for the first year of production

Figure 14 illustrates the variation of the c factor (left y-axis). The c values are based on a year average power consumption of 502.17 MW during the first year in Figure 13. The GHV is considered constant and the yearly variation in Fuel Gas flow rate (right axis) is obtained simply by multiplying the c values for each month with the Fuel Gas flow rate found in Linde Case 01 (year average value) which is 39.9 T/hr. The highest Fuel Gas consumption is found in the 12^{th} month and it is 0.7 T/hr (1.8 %) higher than the year average value. Since the flow rate of LNG in Case 01 is about 13 times higher (529 t/hr) than the Fuel Gas flow rate, the percentage change in LNG production due to increased Fuel Gas consumption is just 0.1%. This effect has therefore been ignored and it has been decided to make the model on a yearly basis.

6.3. Model 1: Constant Split Factors

Model 1 is based on the production forecasting model that is used at the Kårstø gas processing facility where natural gas is treated and exported by pipeline. The excel model used at Kårstø calculates the flow rates and composition of the products by using constant split factors which means that constant fractions of each component in the Feed Gas are distributed to the different products [10]. The same principle has been applied in Model 1 but some adjustments have been made to make the model more representative to the changes in feedstock composition.

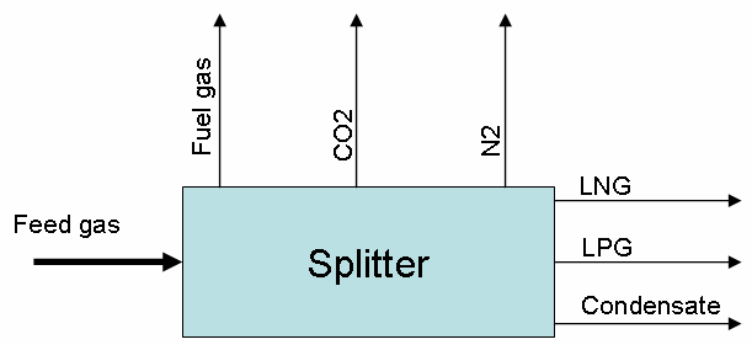

Figure 15: Principle sketch for Model 1

Figure 7 has been used to provide the basis for the constant split factors used in Model 1. It has been found sensible to divide the period from 2007 to 2032 into three main periods where the feedstock composition is approximately constant in each period. The partition made in the model has been based on when the different fields are phased in. The first period goes from 2007 to 2015, the next goes from 2016 to 2020 and the last goes from 2021 until 2032.

It may be assumed that the LNG plant operates under approximately constant conditions in each period. Three sets of split factors have therefore been calculated from the Linde base cases to represent the behaviour of the LNG process in the three main periods. The average split factors from 2007 and 2013 have been used to represent the first period. The average value of the split factors in 2015 and 2020 represents the second period. No average values have been available to represent the period between 2022 and 2032 and therefore the split factors for 2022 have been used. No mechanisms have been included in the model to control the GHV of the LNG.

6.4. Model 2: Mathematical Approach

The intention of Model 2 was to make a model that represents all the three main periods of production by adjusting the split factors as functions on the composition of the Feed Gas. The model was developed before detailed data for more than System 25 and 26 data were available therefore the approach was more mathematical than physical. Some of the assumptions made in the development of this model have later been proven to be questionable or incorrect and are discussed in Chapter 9.

6.4.1. Method for Estimating Split Factors

In Model 2 it has been assumed that some of the split factors can be found on component basis from correlation between the mole fraction in the Feed Gas and the split factors for the products. The logical reasoning behind this assumption has been that if a Feed Gas with for instance higher methane fraction is introduced, the methane fraction in LNG and Fuel Gas will go up. Process data from the six Linde base cases with the corresponding split factors found for Model 1 has formed the basis for the model behaviour. Correlations proving to be more or less linear have been linearized and used to make the split factors into linear functions of the mole fraction in the Feed Gas. The method used to find the linearized functions are the tool in Excel called "add trendline" which finds the best linear representation of the plotted data. The components included in the correlation analysis are Nitrogen, CO_2 and C1-C5. Split factors for heavier hydrocarbons (C6+) have been assumed to be constant because most of these components are going to the Condensate product. Iteration has been used to adjust the content of heaver hydrocarbons in the LNG to satisfy the GHV criteria. A parameter for the change in power consumption was not included in the model because it didn't seem relevant at the time the model was made.

Figure 16 shows a section of the split matrix which has been used to calculate the product compositions in Model 2. The split factors in the white cells have been modified while the ones in the green cells have been kept constant because the values are close to zero and they don't vary much. C6+ have also been given constant split factors because these components are basically all found in the Condensate product, hence the condensate split factors for C6+ are very close to one. The constant split factors have been based on the split factors found for Case 01. More information about how the split matrix works in Excel can be found in Appendix E.

Split Matrix: Function of feed gas composition

	LNG	LPG	CONDENSATE	FUEL GAS	CO2	N2
N2	0,2685	0,0000	0,0000	0,0687	0,0000	0,6627
CO2	0,0006	0,0000	0,0000	0,0673	0,9321	0,0000
METHANE	0,9339	0,0000	0,0000	0,0661	0,0000	0,0000
ETHANE	0,9273	0,0025	0,0000	0,0703	0,0000	0,0000
PROPANE	0,5292	0,4047	0,0008	0,0654	0,0000	0,0000
I BUTANE	0,2848	0,6272	0,0275	0,0604	0,0000	0,0000
N BUTANE	0,2644	0,5967	0,0820	0,0569	0,0000	0,0000
I PENTANE	0,1349	0,0730	0,7515	0,0405	0,0000	0,0000
N PENTANE	0,0504	0,0218	0,8922	0,0355	0,0000	0,0000
C6	0,0001	0,0001	0,9898	0,0100	0,0000	0,0000

Figure 16: Segment from the split matrix

6.4.2. Nitrogen and CO_2

LNG

The split factor for nitrogen in LNG was based on the linear approximation of the relationship between mol% nitrogen in the Feed Gas and the corresponding split factors found in the base cases. This assumption seems to have some validity when looking at Figure 17 but is not representative for all the cases. No adjustment has been made to this linear function because all the LNG product compositions from the Linde base cases have nitrogen content well below the maximum value in Table 2. The same scale has been used on the y-axis for most of the linearized split functions presented in the following sections in order to compare the linearity for the different components. The functions written on the graphs are the functions used in the split matrix to calculate the split factors for the respective components.

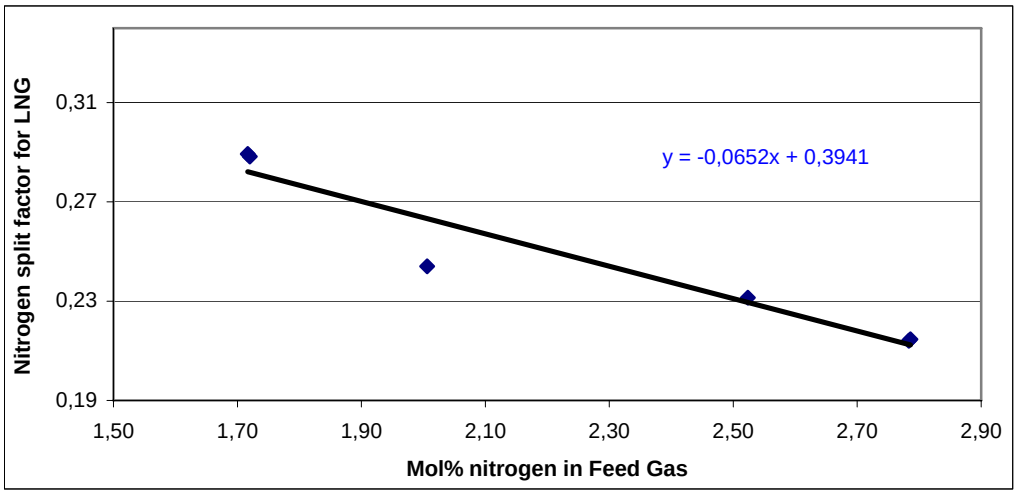

Figure 17: Split function for nitrogen in LNG

The amount CO_2 going to LNG is only 0.08% on molar basis of the total amount in the Feed Gas. This amount is insignificant and has no effect on the composition of the other products. Figure 18 illustrates how the split factor of CO_2 decrease for an increasing fraction of CO_2 in the Feed Gas.

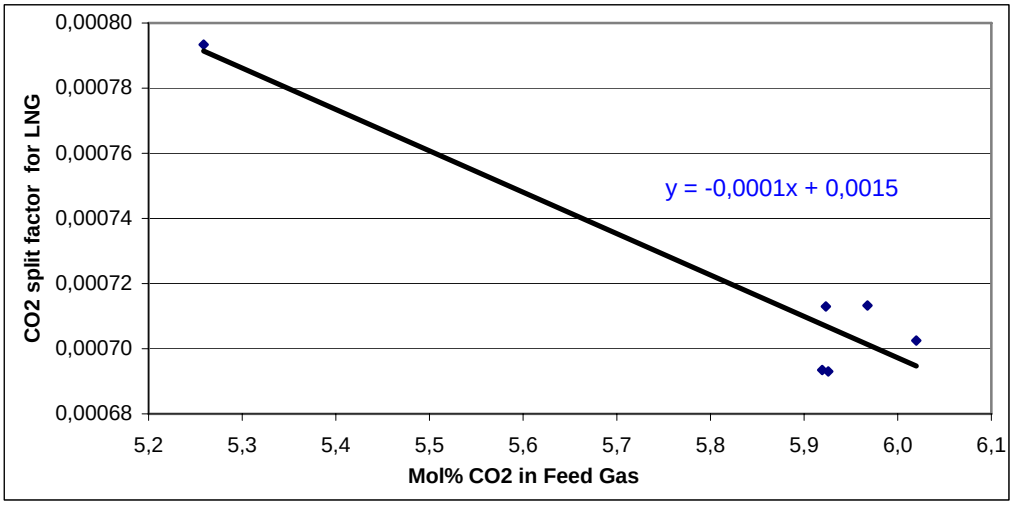

Figure 18: Split function for CO_2 for LNG

The plotted data in Figure 18 doesn't necessarily give an obvious linearization and the scale on the y-axis is not comparable to the other graphs in this chapter because the split factor for CO_2 for LNG is very small. Still, the most important task for this function is to keep the CO_2 fraction in the LNG below the critical mol% stated in Table 2. Therefore the line must be adjusted so that all the plots are above the line. This was done by making the curve steeper with $y = -0.00015x + 0.0015$ which satisfies the maximum CO_2 content in all the six base cases.

Fuel Gas

The linear approximation of the split factors for nitrogen and CO_2 for the Fuel Gas are shown in Figure 19 and Figure 20. The linear approximation seems to be valid for both components without major discrepancies.

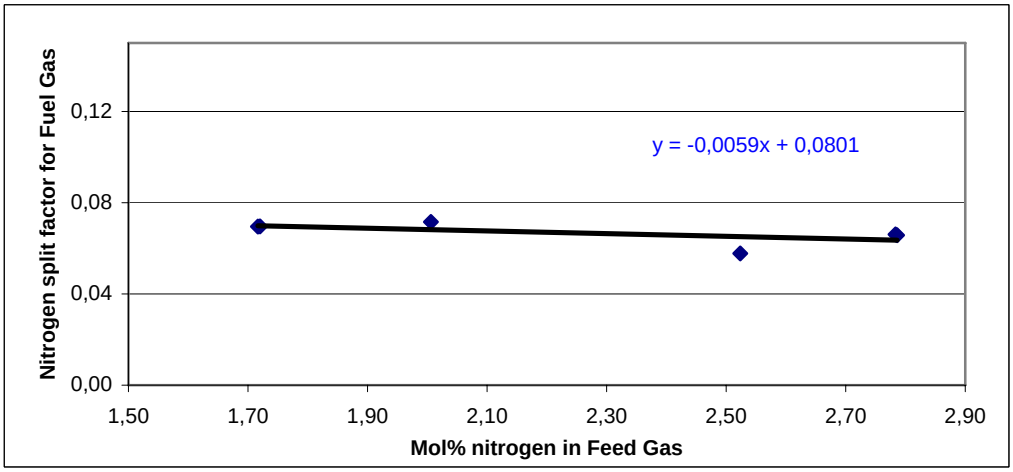

Figure 19: Split function for nitrogen in Fuel Gas

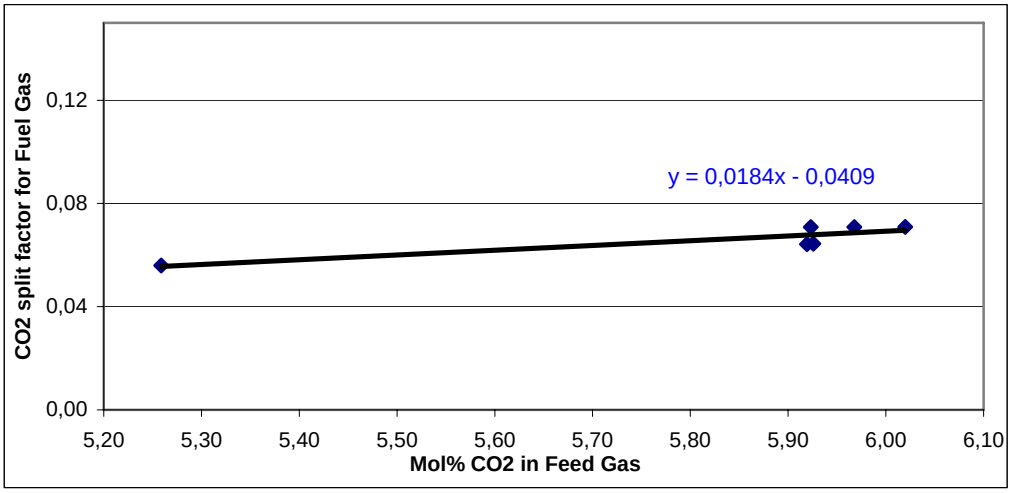

Figure 20: Split function for CO_2 in Fuel Gas

After the split factors for CO_2 and nitrogen was decided for LNG and Fuel Gas, the surplus of CO_2 going to the CO_2 product and the surplus nitrogen going to the Nitrogen product was determined by split balance because none of these components goes to the LPG or Condensate. Equation 6.3 then gives:

$$s_{N2,Nitrogen} = 1 - s_{N2,LNG} - s_{N2,FuelGas} \quad (7.2)$$

$$s_{CO2,CO2} = 1 - s_{CO2,LNG} - s_{CO2,FuelGas} \quad (7.3)$$

6.4.3. Methane and Ethane

Methane and ethane are mainly distributed to the LNG and the Fuel Gas products. It was therefore only necessary to make a linear approximation for one of the products because the split factors for the other product would then be given by split balance. The linear approximations for the split factors for methane and ethane in the LNG product are shown in Figure 21 and Figure 22.

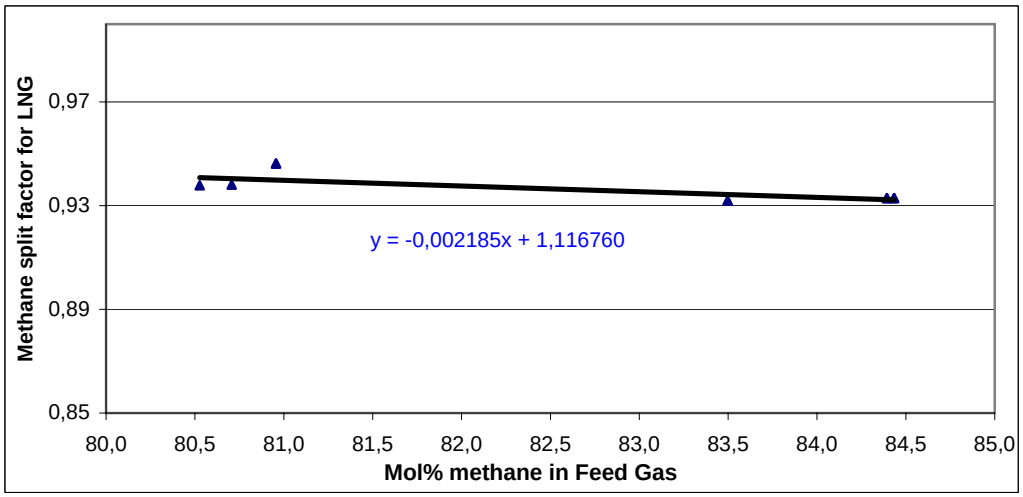

Figure 21: Split function for methane in LNG

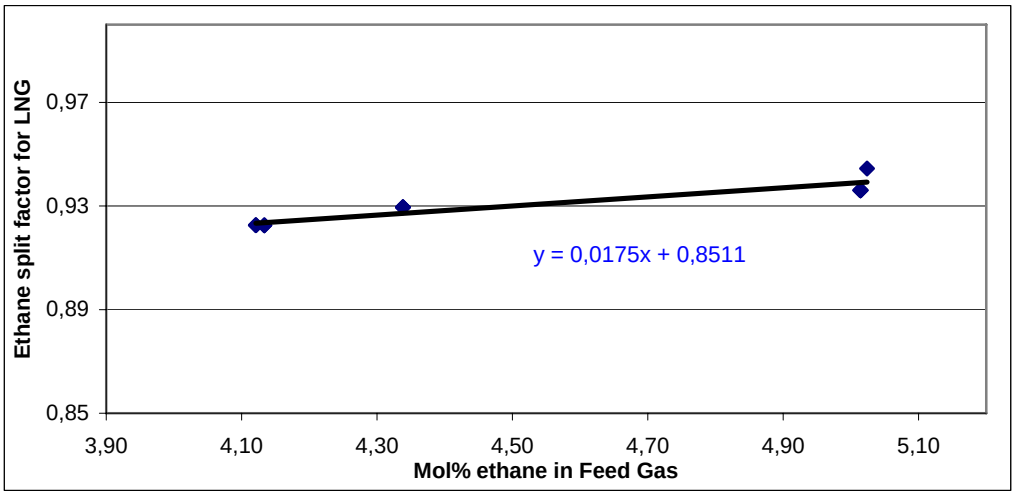

Figure 22: Split function for ethane in LNG

The split factor for methane and ethane in the Fuel Gas is given by split balance:

$$s_{C1,FuelGas} = 1 - s_{C1,LNG} - s_{C1,CO2} \quad (7.4)$$

$$s_{C2,FuelGas} = 1 - s_{C2,LNG} - s_{C2,CO2} \quad (7.5)$$

6.4.4. Propane

Propane is mainly distributed to the LNG, the Fuel Gas and the LPG product as can be seen in Figure 16. It was therefore necessary to determine two of the three split factors in the split matrix. From Chapter 2.2.7, the propane content in the *HHC Removal Column* overhead stream is controlled to keep the GHV in the LNG product below the maximum value from Table 2. The split factor for propane in the LNG has therefore been determined by iteration to satisfy the GHV criteria.

The propane split for the Fuel Gas was assumed to be determined as a function of the mol% propane in the Feed Gas. The propane going to LPG is, as described in Chapter 2.2.7, strongly depending on the fractionation in the *HHC Removal Column* and was therefore evaluated by split balance as a surplus product.

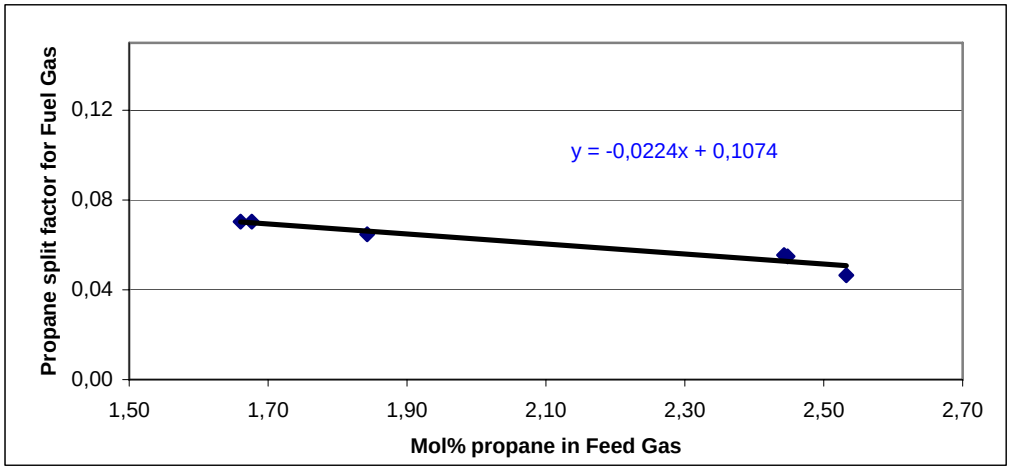

Figure 23: Split function for propane in Fuel Gas

The propane split for LPG is given by split balance:

$$s_{C3,LPG} = 1 - s_{C3,LNG} - s_{C3,FuelGas} \quad (7.6)$$

6.4.5. Butanes and Pentanes

LNG

The GHV of the LNG is not only depending on the content of propane but also the content of butanes and pentanes and these split factors have to be adjusted as well. By investigating the Linde data it seem as if the fractions of these components in the LNG product are in more or less constant ratio to each other and to the propane fraction. The mathematical approach to set these split factors has therefore been to make them dependent on the iterated propane split factor for the LNG. The relationships between the split factors for propane and the butanes and pentanes are shown in Figure 24. The plotted data was obtained by dividing the split factors found for butanes and pentanes in the base cases by the split factor of propane in the same cases.

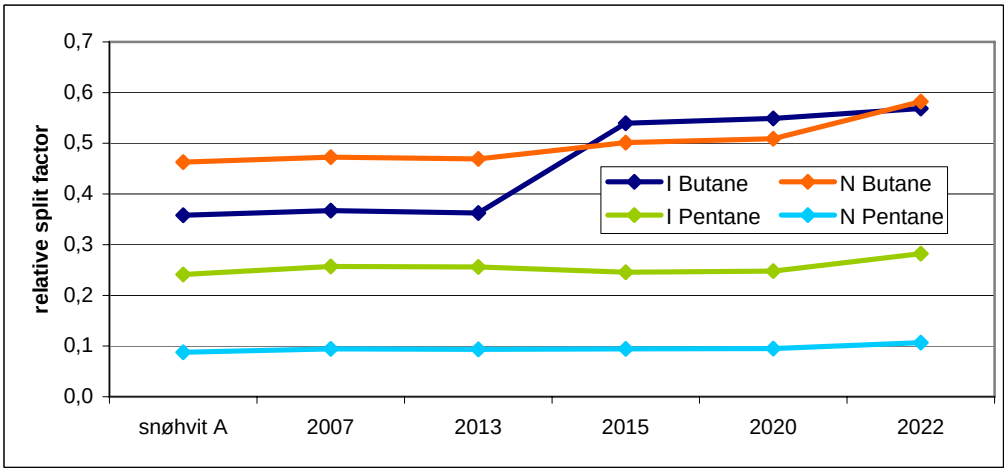

Figure 24: Split factors for butanes and pentanes relative to propane split factor in LNG

The relative split factors for N-Butane, I-Pentane and N-Pentane are more or less in a constant relationship to the propane. The average values of the relative split factors for these components were further used to represent this relationship in Model 2. It has not been considered valid to use the average value for I-Butane because it is clearly not in a constant relationship to the propane split. The split factor for this component varies significantly (between 0.13 and 0.34) in the different base cases. This will not affect the LNG composition significantly because the I-Butane fraction is less than 0.1 mol% in most of the base cases. On the other hand the LPG composition and flow rate will be affected quite significantly because the I-Butane content is 24-28 mol% in the base cases. A better approximation which is more representative for the I-Butane split was found to be a linear function depending on the mol% I-Butane in the Feed Gas. The linear function is illustrated in Figure 25.

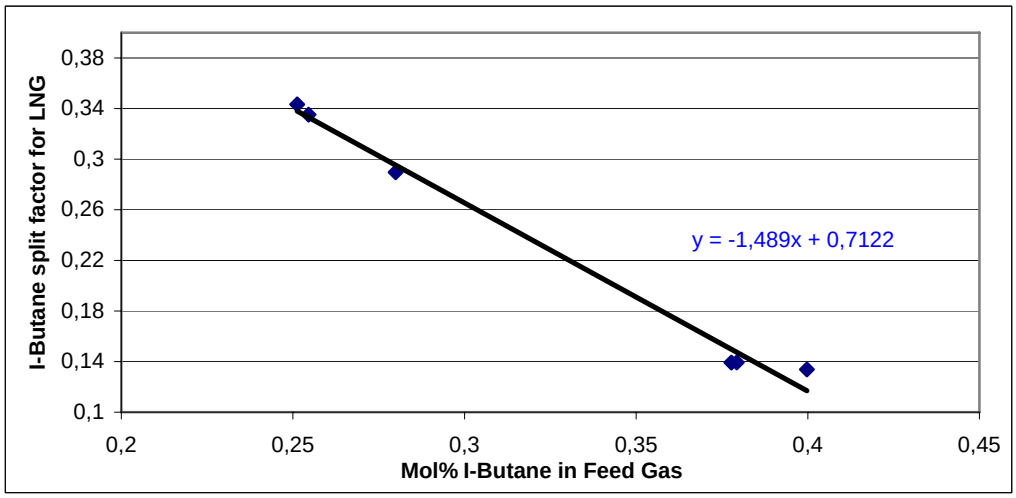

Figure 25: Split function for I-Butane in LNG

Fuel Gas

There are no GHV restrictions for the Fuel Gas and therefore the split factors for C3 to C5 have been calculated as linear functions of the mol% of each component in the Feed Gas. The plotted data and the linear functions of the split factors for these components are found in Appendix D and assuming linearity seems to be a valid approximation.

LPG and Condensate

The two figures below illustrate how the split factors for butanes and pentanes vary for LPG and Condensate for the six Linde based cases.

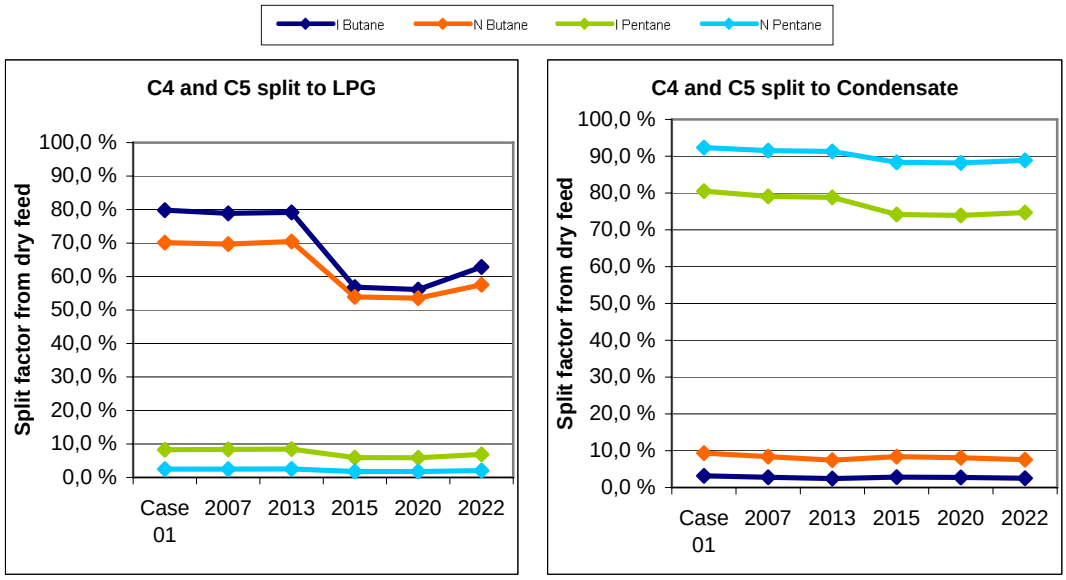

Figure 26: C4 and C5 split factors for LPG and Condensate

To decide the split factors for butanes and pentanes in the LPG and Condensate it was necessary to set one split factor for each component constant. By requiring split balance, the corresponding split factor for the other product will then be given.

In Figure 26 (left) the split factors for I-Pentane and N-Pentane proves to be more or less constant (2.55% variation) for LPG while the split factors for the butanes are varying significantly (23.7% variation) and may not be represented with constant split. Therefore the I-Pentane and N-Pentane were chosen to be constant for LPG based on the average value from all the cases. The split factor for the pentanes in the Condensate is then given by split balance:

$$s_{I-Pent,Condensate} = 1 - s_{I-Pent,LNG} - s_{I-Pent,FuelGas} - s_{I-Pent,LPG} - s_{I-Pent,Nitorgen} - s_{I-Pent,CO2} \quad (7.7)$$

$$s_{N-Pent,Condensate} = 1 - s_{N-Pent,LNG} - s_{N-Pent,FuelGas} - s_{N-Pent,LPG} - s_{N-Pent,Nitrogen} - s_{N-Pent,CO2} \quad (7.8)$$

In Figure 26 the solution is not as obvious as for the pentanes in the LPG. Still the split factors for the butanes are more constant than for the pentanes. Therefore these were given the constant split value equal to the mean values in the six cases. The split factor for the butanes in the LPG is then given by split balance:

$$s_{I-But,LPG} = 1 - s_{I-But,LNG} - s_{I-But,FuelGas} - s_{I-But,Condensate} - s_{I-But,Nitrogen} - s_{I-But,CO2} \quad (7.9)$$

$$s_{N-But,LPG} = 1 - s_{N-But,LNG} - s_{N-But,FuelGas} - s_{N-But,Condensate} - s_{N-But,Nitrogen} - s_{N-But,CO2} \quad (7.10)$$

6.4.6. GHV Requirements

Figure 27 shows a plot for the GHV of the LNG product as a function of the methane content in the Feed Gas. It was necessary to find such mathematical relationship between the Feed Gas composition and the GHV value of the LNG product to be able

to make the iteration of the propane split factor in Model 2. The reason for this relationship was not found in Model 2, but linear approximation seemed to be representative for the GHV behaviour.

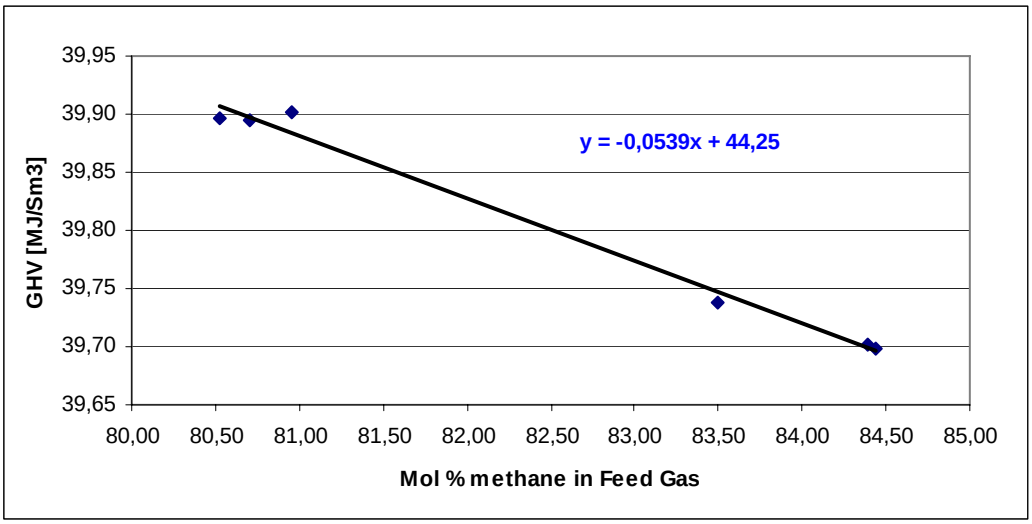

Figure 27: GHV as function of mol% methane in Feed Gas

6.5. Model 3: Physical Approach

Model 3 has a more physical approach than Model 1 and Model 2 and therefore more detailed data of the LNG process at Melkøya was necessary to develop it. Stream data and process flow diagrams (PFDs) for System 11, 12, 25 and 26 has been used to investigate certain units in the respective systems. The process description presented in Chapter 2 has formed the basis for the physical behaviour of the LNG process. Stream data from PIM has also been used to get information about internal system streams.

The findings from Model 2 (see Chapter 9) have formed the goals for the development of Model 3 which are as follows:

- Improved estimation of methane split for LNG and Fuel Gas
- Investigate factors determining nitrogen content in LNG
- Evaluate/develop better methods to represent the fractionation process in the *HHC Removal Column*
- Improved accuracy of Fuel Gas and LPG composition and flow rate
- Identify factors determining the GHV

6.5.1. Improved Fuel Gas estimation

The estimated composition of the Fuel Gas in Model 2 needs to be improved; especially for the components CO_2 and methane (see Figure 34 and Figure 35). Improving the methane split will automatically improve the estimation of the LNG product since the surplus of methane after the Fuel Gas is taken out goes to the LNG product. Improved accuracy of the LNG composition will also improve the composition of LPG and Condensate because they are depending on the LNG split factors and modelled as surplus products evaluated by split balance in Model 2.

The Fuel Gas composition in the Linde base cases are taken from the split downstream the *Inlet Filter Separato*r described in Chapter 2.2.2. Based on the system descriptions for the *Slug Catcher* and the *Inlet Facilities*, the Fuel Gas composition is more or less determined by the flash process in the *Slug Catcher* and the filtering process in System 12. The amount of heavy hydrocarbons extracted in the filter separator is neglectable and the Fuel Gas composition will therefore be entirely determined by the Feed Gas separation in the *Slug Catcher*.

The separation in the *Slug Catcher* is more or less similar to the isothermal flash separation described in Chapter 5.7 but the *Slug Catcher* in System 11 has three products instead of two. The Feed Gas does not contain MEG and water and the separation will therefore just have to products. To do the flash separation for the Feed Gas without concerning the MEG and water stream from the bottom is considered valid after a closer look at the stream data for the condensate stream (12-105) and the MEG/water stream (12-112). In Case 01 the MEG/water stream contains 68.6 mol% water and 30.5 mol% MEG and only 0.9 mol% hydrocarbons. The bottom product is

therefore more or less a pure MEG/water stream. The total amount of MEG/water in the condensate stream sums up to 1 mol% and will therefore not influence the flash process between the natural gas and the condensate significantly. To model the *Slug Catcher* as an isothermal flash separation with two products is therefore considered valid to estimate the composition of the gas entering System 12. As stated in Chapter 5.4, the vapour composition in the *Slug Catcher* is depending on the temperature, pressure and the composition of the Feed Gas from the pipe line. The temperature is more or less constant but the pressure can vary between 90 Bara to 35 Bara which is believed to influence the vapour composition and flow rate. A simple flash tank has been modelled in the process simulation program Aspen HYSYS to investigate the significance of the variation in inlet pressure.

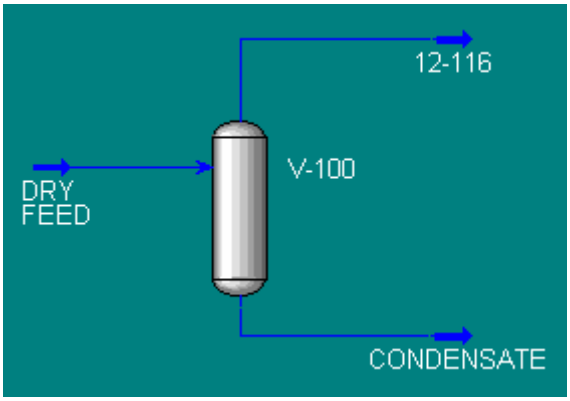

Figure 28: Flash tank in HYSYS

Feed Gas from Case 01 has been used in the isothermal flash tank simulation in HYSYS (Peng-Robinson fluid package). The temperature in the flash tank is set constant to 1 °C for all the flash simulations performed in this section and the flash tank is simulated for inlet pressures of 35, 70 and 90 Bara. For an inlet pressure of 35 Bara, the natural gas will be compressed before entering System 12 as described in Chapter 2.2.1. From the phase envelopes (see Appendix D) generated for the vapour stream (12-116) it can be seen that compression of the gas phase from 35 Bara to 70 Bara will not change the composition because both temperature and pressure are increasing in the compression and the gas will remain in the gas phase area to the right of the dew point line. This does not apply for a Feed Gas pressure of 90 Bara. As described in Chapter 2.2.2 the pressure of stream 12-116 will be reduced to 70 Bara before the gas enters the *Inlet Filter Separator*. The gas is at the dew point when it enters System 12 and some of the gas may condense due to the pressure drop and the composition may also change. A heat exchanger and a valve were therefore added to the simulation model (see Appendix D) to see whether a condensed amount had any significance for the Fuel Gas composition. The heat exchanger was set to compensate the temperature due to throttling and keep it to 1 °C and the pressure drop in the valve was set to 20 Bara according to system descriptions in Chapter 2.2.2. The condensed amount was found to be insignificant and the composition remained more or less constant.

The composition of stream 12-116 for the cases at 35 Bara and 90 Bara does not deviate significantly for the same stream at 70 Bara (less than 0.0014% for the component with highest discrepancy) and the pressure is not so important for the vapour composition. The separation in the *Slug Catcher* is therefore considered to be mainly depending on the Feed Gas composition.

6.5.2. Calculation of Fuel Gas composition in Excel

The necessary equations to calculate the Fuel Gas composition can be derived from the equations listed in Chapter 5.6.

Rearranging Equation 6.4 gives the following expression

$$y_j = K_j \cdot x_j \tag{7.11}$$

Substituting y_j with $K \cdot x_j$ in Equation 6.6 makes it possible to calculate x_j as a function of K_j.

$$x_j = \frac{F \cdot z_j}{(V \cdot K_j) + L} \tag{7.12}$$

The same principle can be applied for y_j which gives the following relationship:

$$y_j = \frac{F \cdot z_j}{V + \frac{L}{K_j}} \tag{7.13}$$

It is necessary to estimate the K-values for each component to be able to calculate y_j and x_j in Equation 7.12 and Equation 7.13. HYSYS has been found the most suitable tool for this purpose because it calculates the K values directly for any composition at a given temperature and pressure.

The flow rates V and L can be calculated by iteration in Excel. In addition to Equation 6.5, two other equations are needed to satisfy the conservation of mass at equilibrium:

$$\sum_{j=1}^{N} x_j = 1 \tag{7.14}$$

$$\sum_{j=1}^{N} y_j = 1 \tag{7.15}$$

The iteration routine in Excel is further described in Appendix E.

6.5.3. Calculation of Fuel Gas flow rate in Excel

Equation 6.1 readily show that the material flow of Fuel Gas is depending on the heating value (GHV) of the gas and the thermal power demand for the whole LNG process. The power demand will vary for the different phases of production but it will always have a known value. Therefore it's favourable to include the power demand expressed as Fuel Gas fired in MW as one of the parameters in the Fuel Gas calculation. The GHV is calculated based on ISO 6976 (see Chapter 5.3) using the Fuel Gas composition given from the iterative calculation from the chapter above.

As stated in Chapter 6.2.2 the change in Fuel Gas consumption is not expected to have significant influence on the production of LNG on a yearly basis. When considering the whole lifetime of the plant, the increase in power consumption due to Feed Gas compression must be added to provide a valid estimate of the Fuel Gas consumption.

The Fuel Gas consumption is found by rewriting Equation 6.1:

$$\dot{n}_{fuel} = \frac{Q_{GT}}{GHV_{fuel}} \tag{7.16}$$

When the Fuel Gas composition and flow rate is determined, the improved split factors for the Fuel Gas can be calculated by Equation 6.2.

6.5.4. Improved Nitrogen Split Factors for LNG

The nitrogen content in the LNG leaving the Nitrogen Removal Column described in Chapter 2.2.7 has a nitrogen content of 1 mol%. From the Linde base cases it can be found that the nitrogen content in the LNG product is between 0.59 and 0.73 mol%. This means that the nitrogen content in the LNG is determined in the LNG storage tank. Due to LNG tank return gas (mostly nitrogen and methane), the nitrogen fraction in the LNG will go down [1]. The composition of the LNG in the storage tank will determine the content of nitrogen in the LNG products according to vapour liquid equilibrium theory. Storage and loading will also effect the composition in the tank, but this factor is outside the scope of work. It is not possible to utilize the same principle as used for determining the Fuel Gas composition. This is because the LNG composition is not known. No method has been found for estimating the split factor and the best alternative is probably to use the mean split values found for Model 1. If-statements in excel based on the methane content in the Feed Gas has been used to assign the correct constant split factor corresponding to light and heavy feedstocks.

6.5.5. GHV Estimation for Model 3

Figure 27 shows that the GHV goes down for a lighter Feed Gas composition. The GHV control in the *HHC Removal Column* is set to keep the GHV of stream 25-144 (see Figure 4) below 39.0 MJ/Sm^3 to satisfy LNG product specifications. Stream data from Case 01 and Case 07 representing light and heavy feedstocks has been used estimate the GHV of stream 25-144. The estimated values using ISO 6976 reveals that the GHV for Case 07 is higher than the value estimated for Case 01 (39.3 MJ/Sm^3 and 39.0 MJ/Sm^3 respectively). The process in the *HHC Removal Column* is therefore not the reason for the linear profile in Figure 27. The reason is not found in the Nitrogen Removal Column either because this column is just set to bring the fraction of nitrogen in the LNG down to a value of 1 mol% without changing the content of the other components. The reason for the reduction in GHV for the LNG product is therefore found in the LNG tank. The concentration of nitrogen in the boil-off gas is given by the composition of the LNG according to vapour-liquid theory. A better estimation has not been found because the LNG composition is unknown. No improvement has therefore been made to the linear GHV function developed for Model 2 which has also been used in Model 3.

6.5.6. Split factors for C3 to C5

LNG

A HYSYS simulation of the *HHC Removal Column* might be helpful to get a better insight in how the split factors for propane, butanes and pentanes for the LNG product are influenced by the different Feed Gas compositions. This was originally the planned method for evaluating these split factors for Model 3, but due to recycled streams in the system, this method failed to deliver any results. This is further described in the following sections.

To be able to make the simulation of the *HHC Removal Column*, it was necessary to isolate this unit from the rest of the plant to avoid simulating the whole plant. To isolate parts of a larger system is usually possible when the different systems operate independently. This is unfortunately not the case for the *HHC Removal Column* because two of the streams entering the column are depending on several of the units and systems described in Chapter 2.2. The stream 12-124 which is recycled back to System 11 from System 20 and mixed into the main Feed Gas line is not constant but shows a rather large variation in composition and flow rate (from stream data Case 01, 07, 11 and 13). The Feed Gas entering the *HHC Removal Column* will therefore not be just depending on the Feed Gas but also stream 12-124 which has a significant influence on the composition and flow rate. The C4/C5 reflux stream described in Chapter 2.2.7 is depending on the fractionation in system 26, but this stream proves to have a more or less constant flow rate and composition and could have been used for simulation purpose. A better explanation for the variation in the relationship between propane and I-Butane has therefore not been found (Figure 24). To represent the split factor for I-Butane as a function of the mol% I-Butane in the Feed Gas in Model 2 is questionable of two reasons. The first reason is the effects from the recycled stream

12-124 described above. The other reason is that the fractionation process in the *HHC Removal Column* is manipulated by the C4/C5 reflux stream to control the GHV. The LNG composition is therefore a function of the C4/C5 reflux stream. These effects are not suited for implementation in Excel and far more advanced methods are probably necessary to perform the estimation. The best representation of the I-Butane split has therefore been to represent the *HHC Removal Column* for the three main periods with average constant split factors found in Model 1. An if-statement in excel based on the methane content in the Feed Gas has been used to assign the correct constant split factor corresponding to the different main periods of production.

LPG and Condensate

The results in Table 6 shows that the C4(-) fraction in the Condensate product is too high in some of the cases for Model 2 according to specifications in Chapter 4.3.3. The improved calculation of LPG and Condensate has been developed based on the separation process in the *De-Propaniser* column. It has been found from *Model 2* that the average split factors assigned to the butanes and pentanes may not be representative to the configuration of the *De-Propaniser*. Manipulation of the LPG and Condensate products has been found necessary to satisfy the column behaviour for different feedstocks.

The LPG composition is determined in the *De-Propaniser* as described in Chapter 2.2.8. The column has a different configuration depending on whether the Feedstock is light or heavy. The maximum level of C5+ in the LPG product is stated in Table 4, but this is not the level that the column operates at in the base cases. Figure 29 shows a plot of the C5+ level on mass basis for LPG in the six Linde base cases. It has been assumed that the *De-Propaniser* operates according to the light and heavy Feedstock set points illustrated by the orange lines in the figure. This assumption was necessary in order to define the LPG product in the manipulation.

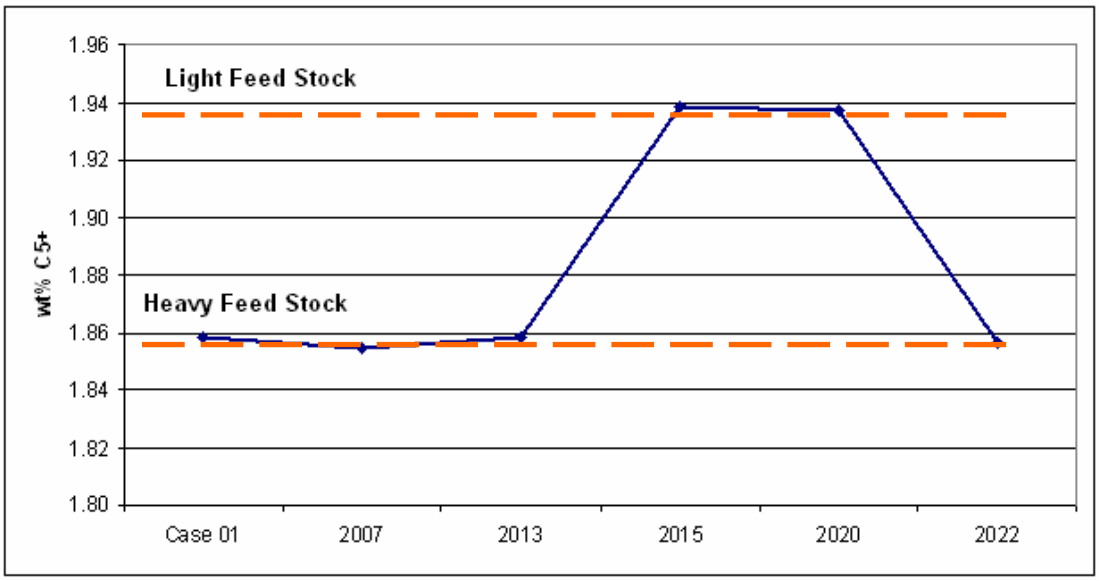

Figure 29: Light and heavy feedstock set points for LPG in De-Propaniser

From the six Linde base cases it has been found that the C4(-) content in the Condensate product has a value of value of 2.1 wt% in all cases. This is the same value as the *De-Propaniser* in Chapter 2.2.8 is set to achieve. This value has been assumed constant in order to define the Condensate product in the manipulation.

The Fuel Gas and the LNG has already been defined for Model 3, which implies that the surplus (except CO_2 and Nitrogen) from the Feed Gas is going to the LPG and Condensate products. The same split factors from Model 2 has been used for these products to satisfy the split balance. After the temporarily composition of LPG and Condensate has been determined, the two products are manipulated in order to obtain the correct content of C5+ in the LPG and C4(-) in the Condensate. In Figure 26 it can be seen that the butanes in the Condensate are in a more or less constant ratio to each other and the same relationship can be found for the pentanes in the LPG. One iteration factor multiplied with the butanes in the Condensate has been used to take out or add butanes in the Condensate and the butanes in the LPG have been adjusted correspondingly to satisfy mass balance. Another iteration factor multiplied with the pentanes in the LPG has been used to take out or add pentanes in the LPG and the pentanes in the Condensate have been adjusted correspondingly. The right level of C5+ in the LPG and C4(-) in the Condensate has then been achieved by iteration. A further description can be found in Appendix E.

7. Production Forecasting

7.1. Forecasting compared to Reference Data

Model 1 and 3 has been found valid (see Chapter 9.1) to forecast the production of LNG, LPG and Condensate in the period 2007 to 2032. The results from these models have been compared to more accurate reference data to se how precise the forecasted results are.

7.1.1. Reference data

A more detailed production forecast report called "*Snøhvit LNG Project Calculated LNG, LPG and Condensate yearly production*" [3] has been used as reference data for the estimated product forecasting. The process program used for material balance calculations in this report is the one developed by Linde for design of the plant. The production rates and product composition are calculated for normal operation without product loading.

The six representative feedstock compositions have been selected in the above mentioned report to represent the different phases of production. These can be seen in Figure 7. Plant start-up is set to 1st of June 2007.

For the calculation of the yearly production, the following effects are included

- Decrease in Feed Gas flow rate during LNG ship loading
- Gas displacement from ship
- Flaring
- Overhauls every third year

Assumptions made in the report
Increase in Fuel Gas consumption due to onshore and offshore compression is considered. The Feed Gas flow rate is reduced 0,6 % for feedstock #3 and #4 and 0,4 % for feedstock #5 compared to the maximum flow rate of 20,8 Sm^3/sd. This is due to design limitations in some of the plant equipment.

The plant ramp up during the planned start up phase in 2007 is defined as follows:

- June to July: 25%
- August to September: 50%
- October to November: 75%
- December: 100%

The reference forecast data is based on Feed Gas compositions for all the years from 2007 to 2032. The number of stream days and annual Feed Gas flow rates can be found in Appendix B.

7.1.2. Comments to Reference Data

The material balance calculations made in the reference report is believed to be quite accurate. Model 1 and Model 3, which will be used to perform the production forecasting, do not include all the effects listed above directly. Therefore some assumptions have been made in the paragraphs below related to these effects.

The planned start-up date has been delayed and at present time (December 2007) it is still not known when the plant will be in normal operation (100%) due to problems in the start-up phase. The estimated yearly production in the following sections is therefore also based on 1st of June 2007 as the start-up date. The same production schedule (Feed Gas flow rate and number of stream days) is used for the following estimations of the production. Overhauls every third year will therefore be included as a result of the reduced number of stream days for the respective years. The ramp-up period is not included because it is not known how this is included in the reference data. It is therefore assumed that this effect is reflected in the number of "effective stream days" with 100% production.

The yearly production of LNG given in the reference data is given in [MT/yr] with only one decimal. A comparison between the calculated data and the reference data is considered to be too inaccurate since the yearly production is about 4.0 – 4.3 MT. An increase in LNG production from 4.0 to 4.1 corresponds to 2.5 % which is a significant amount. Nevertheless, the reference data is included in the graphs for LNG to give an illustrative impression of the accuracy.

7.1.3. Modification in Excel Models

It has been absolute necessary to implement macro scripts, especially in Model 3, in order to make the models calculate the production year by year and to avoid calculating each year by hand. Documentation for how the scripts work can be found in the macro scripts in the Excel models (from CD).

Power consumption
Accurate data of the thermal power consumption at Melkøya for the whole production period is not available. The power consumption has therefore been estimated based on the calculated heating values and flow rates in the six Linde base cases using Equation 6.1.

K-values
The K-values are estimated according to dry feed pressure given in the Linde base case data. In the periods 2007 to 2014 and 2022 to 2032 the K-values are calculated in HYSYS at a dry feed pressure of 70 Bara and a temperature of 1°C. For the period from 2015 to 2021 the K-values are calculated at a pressure of 35 Bara and a temperature of 1 °C.

7.2. Production Forecast for Updated Feed Gas Data

New and updated Feed Gas data have become available [11] and have been used for the updated production forecasting. The new Feed Gas from the pipeline is given for the period 2007 to 2039 and the profile can be seen in Appendix F. The composition as function of time is quite different for some of the components compared to the data in Figure 7 and this is favourable regarding validation of the two models.

The number of stream days, Feed Gas flow rates and the thermal power consumption are the same as the data used in Chapter 7.1.3. The extended last period (2032 to 2039) is following the same pattern as the period 2022 to 2032.

The product specifications in Chapter 4 are the only available specifications to validate the calculations. There are no reference production data available because this data has recently been updated (December 2007).

In the results in Chapter 8.2.3, the new production figures for LNG, LPG and Condensate have been compared to the reference data used in Chapter 7.1.

8. Results

8.1. Model Accuracy

The graphs in this section are comparing the estimated compositions and flow rates for all the three models to the Linde base cases.

8.1.1. Model 1 Compared to Linde Base Cases

The results shown in the figures below are obtained by subtracting the Linde product compositions from the calculated product compositions.

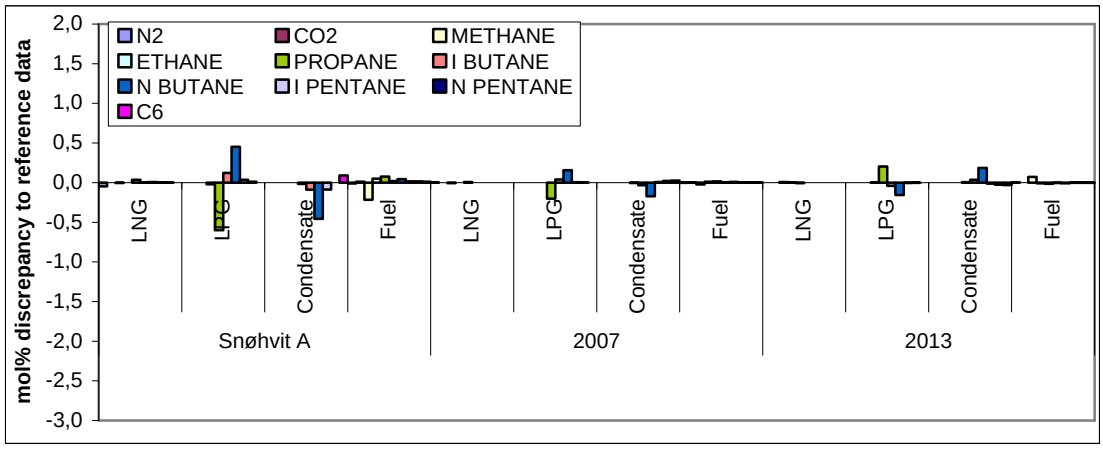

Figure 30: Composition discrepancy for Model 1 for Case 01, 2007 and 2013

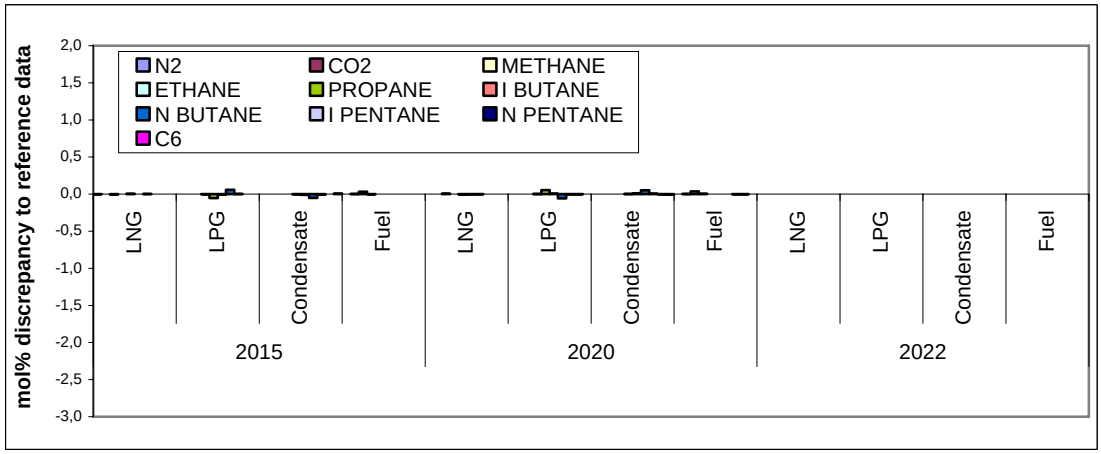

Figure 31: Composition discrepancy for Model 1 for 2015, 2020 and 2022

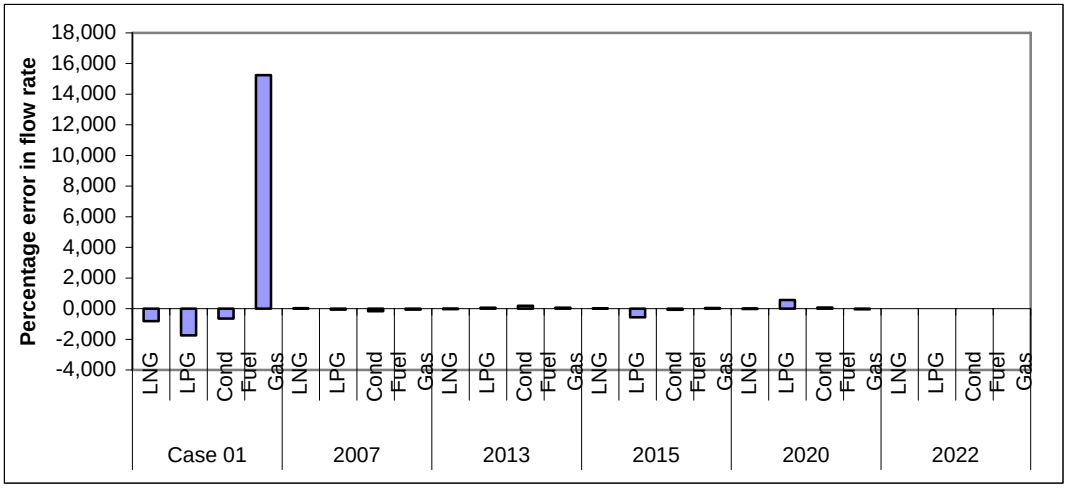

Figure 32: Percentage error in product flow rate on mass basis for Model 1

Table 5: Product specifications Model 1

		Case 01	2007	2013	2015	2020	2022
LNG							
Nitrogen	mol%	0.66	0.73	0.73	0.59	0.60	0.59
Methane	mol%	92.16	92.12	92.14	93.31	93.34	93.11
Ethane	mol%	5.71	5.72	5.71	4.52	4.51	4.83
Propane	mol%	1.17	1.14	1.14	1.24	1.23	1.12
I-Butane	mol%	0.07	0.06	0.06	0.10	0.10	0.10
N-Butane	mol%	0.18	0.17	0.17	0.19	0.18	0.20
C5+	mol%	0.05	0.04	0.04	0.04	0.04	0.04
CO_2	mol%	0.005	0.005	0.005	0.005	0.005	0.005
LPG							
CO_2	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Methane	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	mol%	0.52	0.52	0.52	0.52	0.52	0.52
C5+	mol%	1.32	1.28	1.27	1.36	1.35	1.28
Condensate							
C4-	wt%	1.78	1.98	2.24	2.06	2.14	2.10

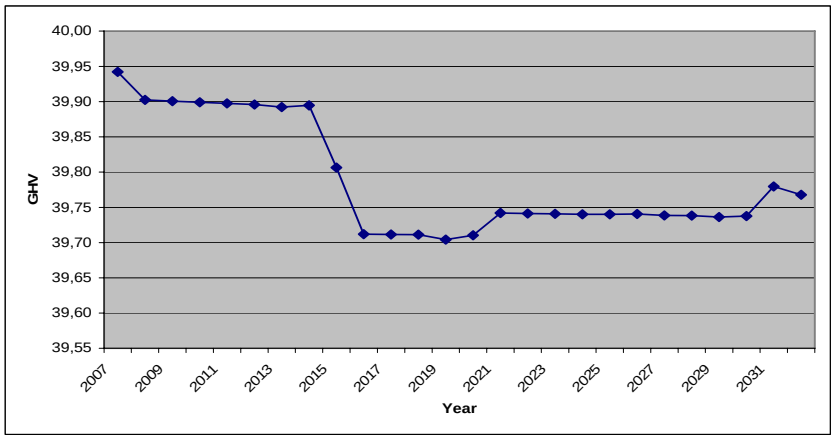

Figure 33: GHV for LNG from Model 1 in the period 2007 – 2032

8.1.2. Model 2 Compared to Linde Base Cases

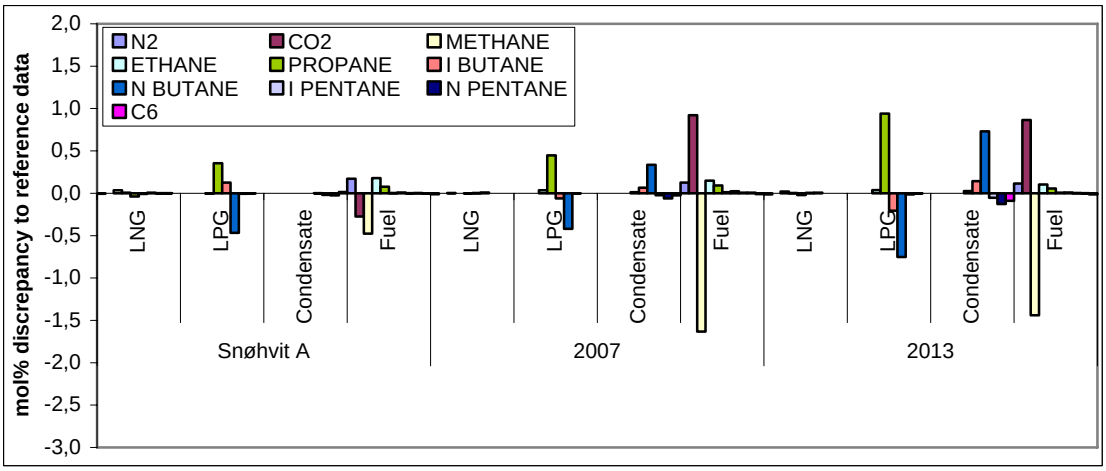

Figure 34: Composition discrepancy for Model 2 for Case 01, 2007 and 2013

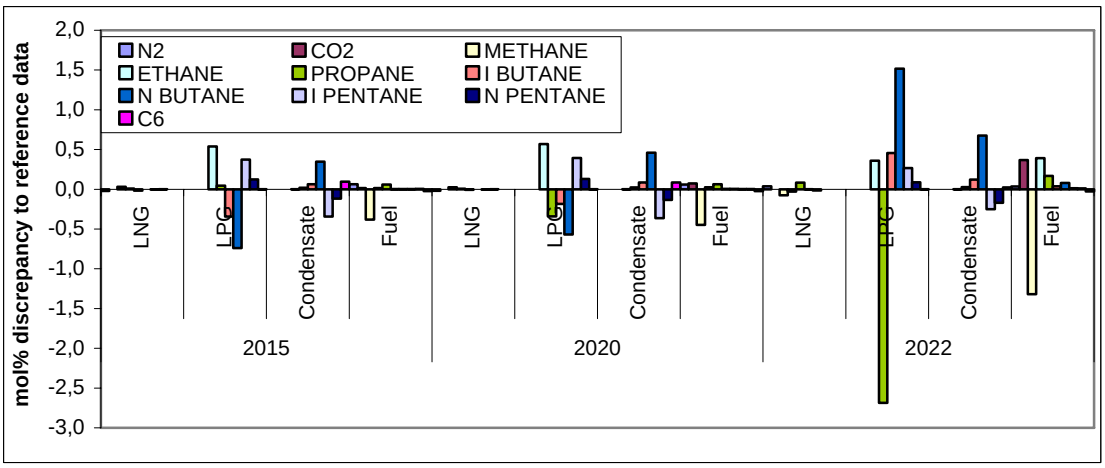

Figure 35: Composition discrepancy for Model 2 for 2015, 2020 and 2022

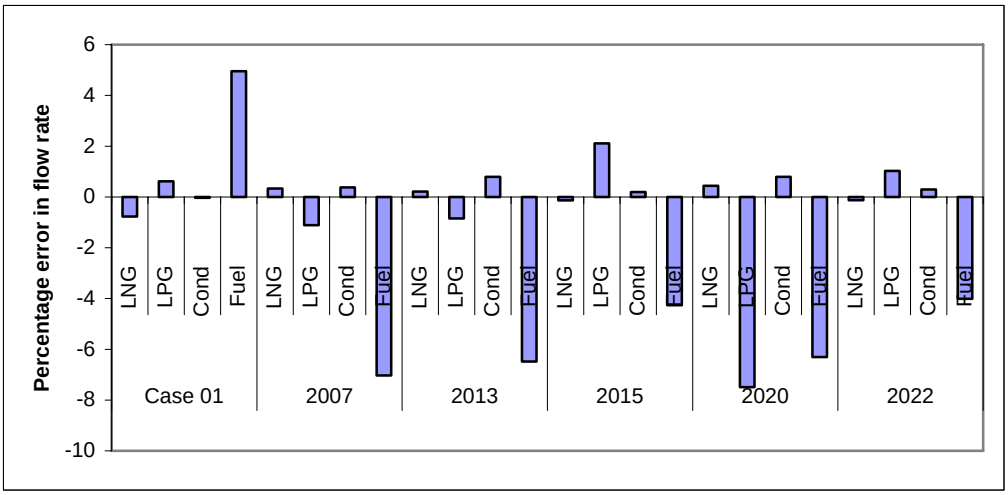

Figure 36: Percentage error in product flow rate on mass basis for Model 2

Table 6: Product specification Model 2

		Case 01	2007	2013	2015	2020	2022
LNG							
Nitrogen	mol%	0.70	0.72	0.72	0.57	0.58	0.63
Methane	mol%	92.20	92.13	92.15	93.35	93.36	93.04
Ethane	mol%	5.72	5.72	5.72	4.53	4.51	4.80
Propane	mol%	1.10	1.13	1.12	1.22	1.22	1.21
I-Butane	mol%	0.06	0.07	0.07	0.10	0.10	0.10
N-Butane	mol%	0.18	0.18	0.18	0.18	0.18	0.18
C5+	mol%	0.04	0.04	0.04	0.04	0.04	0.04
CO_2	mol%	0.005	0.004	0.004	0.004	0.004	0.004
LPG							
CO_2	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Methane	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	mol%	0.05	0.05	0.05	0.05	0.05	0.05
C5+	mol%	1.27	1.27	1.26	1.86	1.88	1.63
Condensate							
C4-	wt%	2.10	2.34	2.64	2.35	2.43	2.59

*Values outside product specifications are written in red

8.1.3. Model 3 Compared to Linde Base Cases

The K-values used to evaluate the accuracy of Model 3 are based on a *Slug Catcher* temperature of 1 °C and a pressure of 70 Bara in Case 01, 2007,2013 and 2022. In case 2015 and 2020 the pressure is set to 35 Bara. This is done according to the pressure and temperature for the *Dry Feed* in the Linde base cases. The Fuel Gas flow rate is adjusted to the same value as the ones in the Linde cases because the power consumption for the different cases is not known.

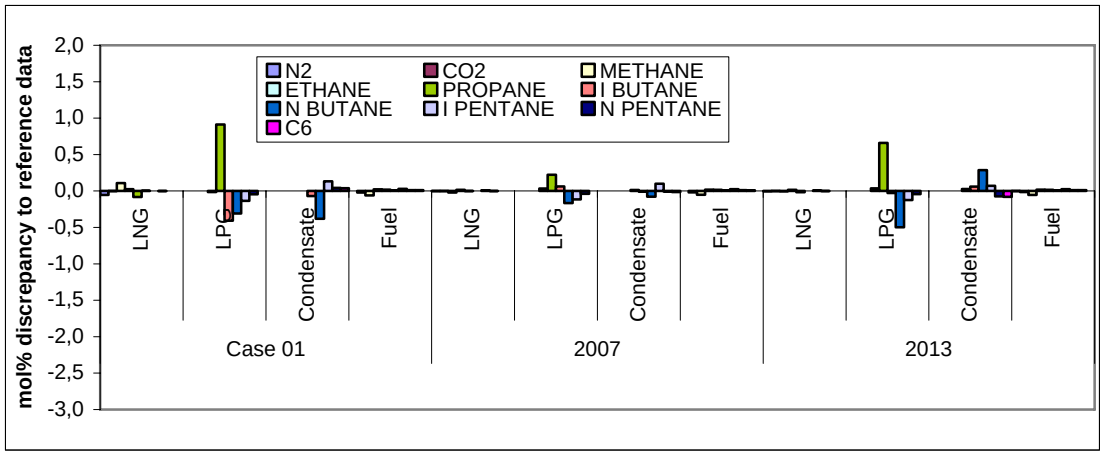

Figure 37: Composition discrepancy for Model 3 for Case 01, 2007 and 2013

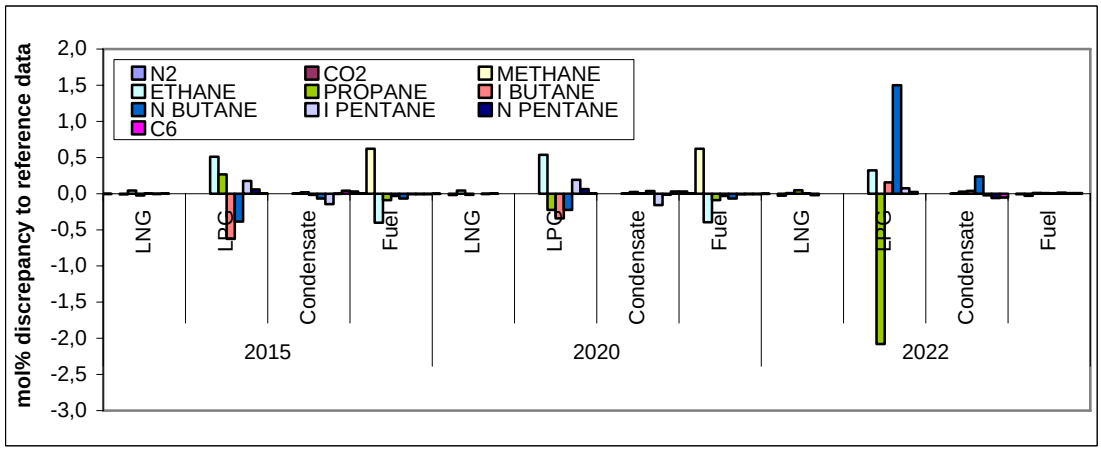

Figure 38: Composition discrepancy for Model 3 for 2015, 2020 and 2022

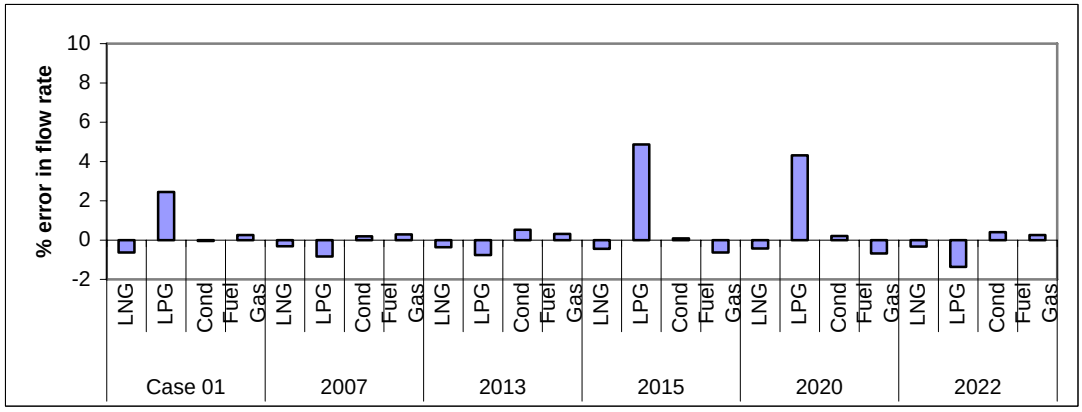

Figure 39: Percentage error in product flow rate on mass basis for Model 3

The Fuel Gas composition and flow rate calculated by Model 2 for the year 2007 has been chosen to verify the improvement for the Fuel Gas calculation in Model 3. The results from both models have been compared to the reference data from Linde case 2007 and are expressed as the calculated composition minus reference composition.

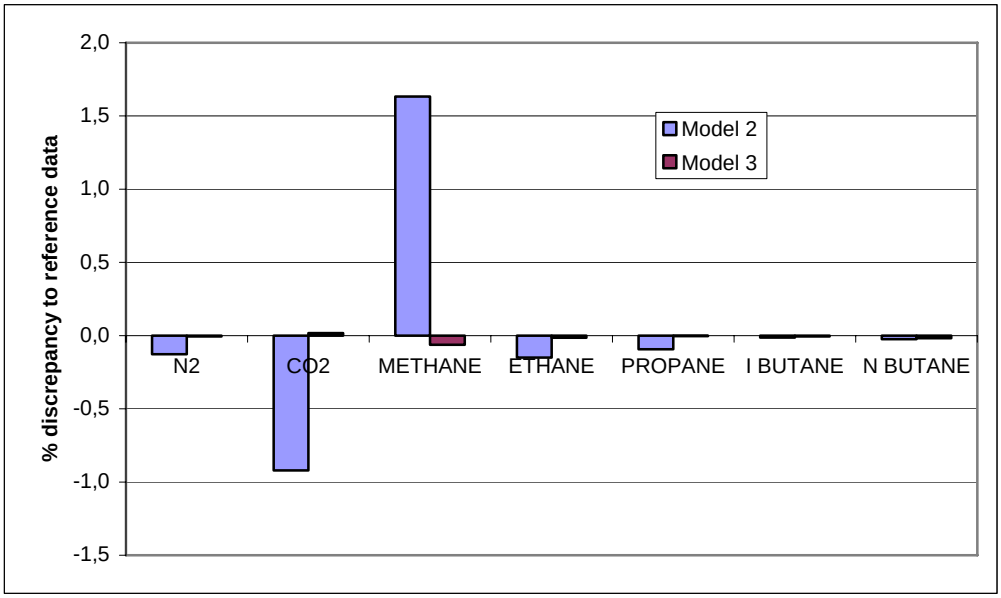

Figure 40: Calculated Fuel Gas composition from Model 2 and Model 3 compared to Linde 2007

Table 7: Product specification Model 3

		Case 01	2007	2013	2015	2020	2022
LNG							
Nitrogen	mol%	0.65	0.73	0.73	0.59	0.59	0.59
Methane	mol%	92.27	92.10	92.12	93.31	93.31	93.08
Ethane	mol%	5.73	5.74	5.73	4.56	4.55	4.84
Propane	mol%	1.05	1.13	1.13	1.21	1.21	1.17
I-Butane	mol%	0.07	0.06	0.06	0.10	0.10	0.19
N-Butane	mol%	0.18	0.18	0.18	0.18	0.18	0.17
C5+	mol%	0.04	0.04	0.03	0.04	0.03	0.04
CO_2	mol%	0.005	0.004	0.004	0.004	0.004	0.004
LPG							
CO_2	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Methane	mol%	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	mol%	0.05	0.05	0.05	0.05	0.05	0.05
C5+	mol%	1.26	1.27	1.28	1.35	1.35	1.28
Condensate							
C4-	wt%	2.10	2.10	2.10	2.10	2.10	2.10

8.2. Results from forecasted production

8.2.1. Model 1 Production Forecast

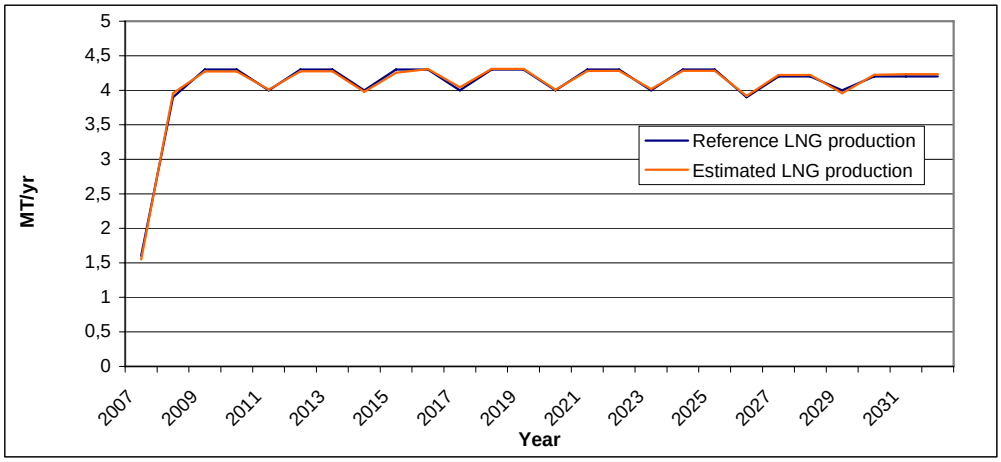

Figure 41: Forecasted yearly LNG production from Model 1 compared to the reference yearly LNG production

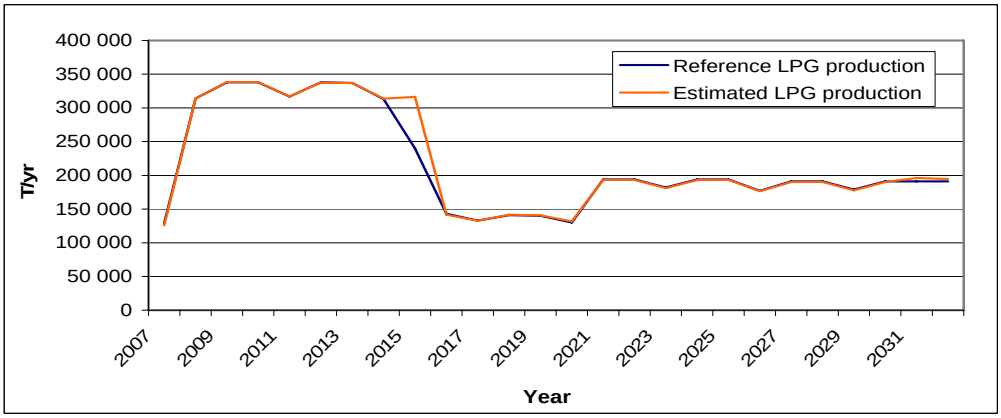

Figure 42: Forecasted yearly LPG production from Model 1 compared to the reference yearly LPG production

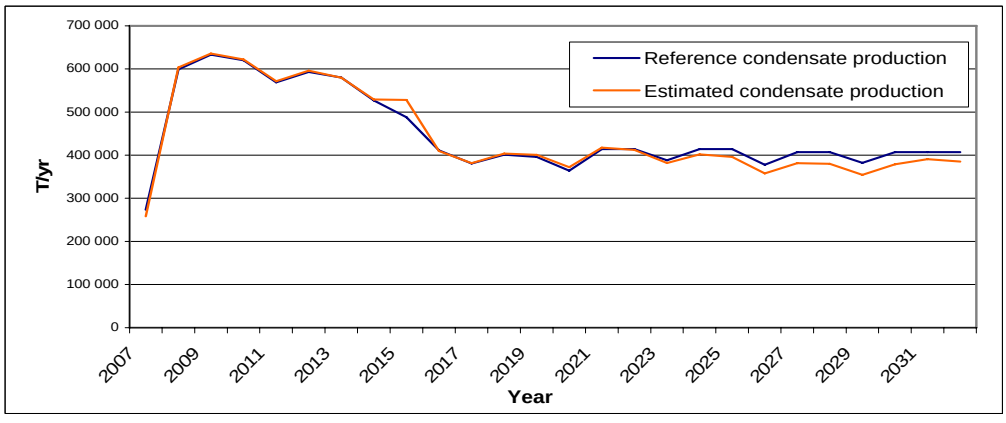

Figure 43: Forecasted yearly Condensate production from Model 1 compared to the reference yearly Condensate production

8.2.2. Model 3 Production Forecast

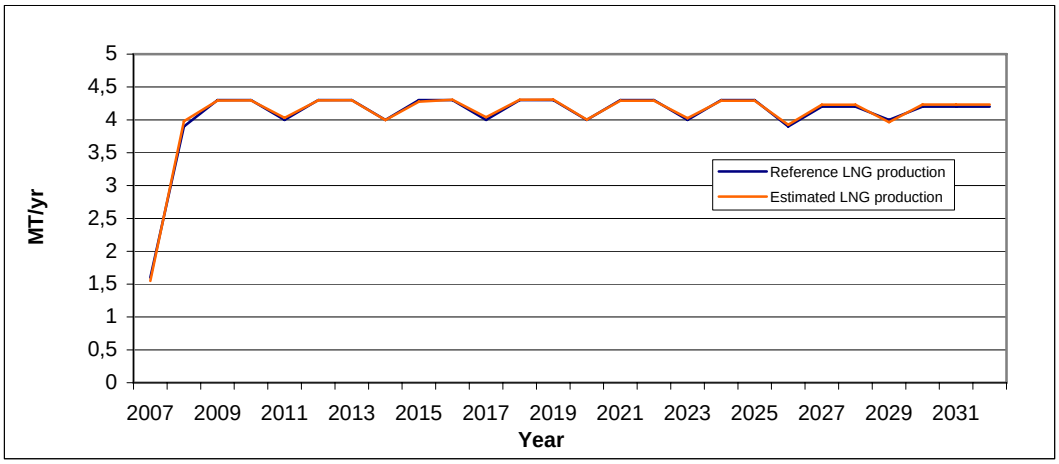

Figure 44: Forecasted yearly LNG production from Model 3 compared to the reference yearly LNG production

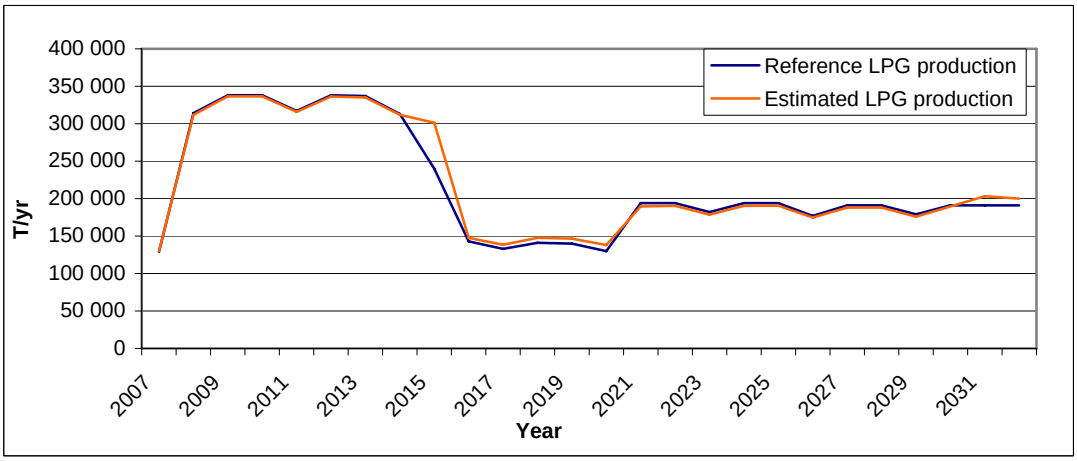

Figure 45: Forecasted yearly LPG production from Model 3 compared to the reference yearly LPG production

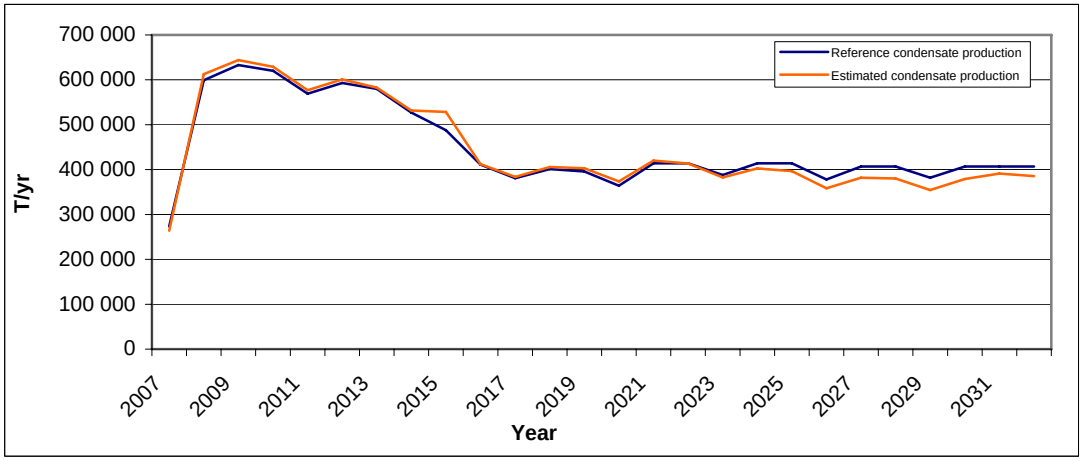

Figure 46: Forecasted yearly Condensate production from Model 1 compared to the reference yearly Condensate production

8.2.3. Production Forecast for Updated Feed Gas Data

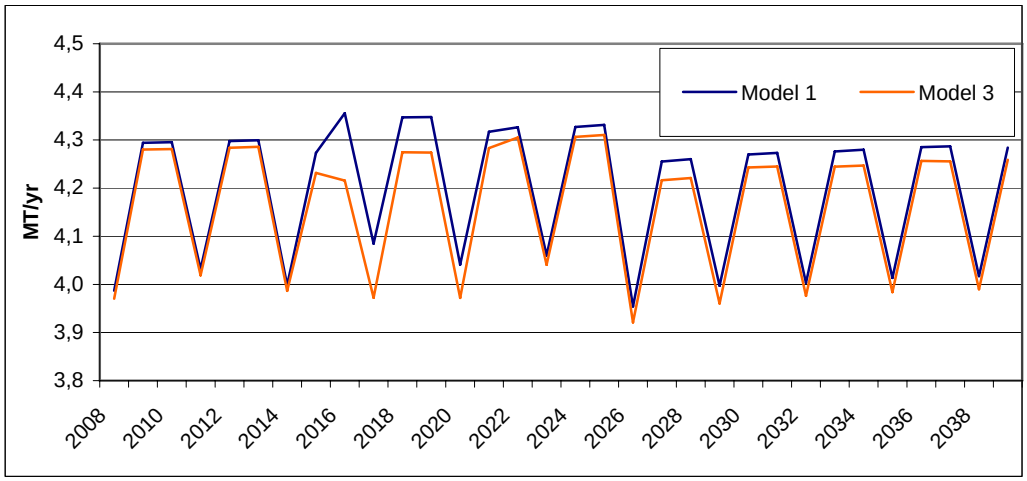

Figure 47: LNG production from Model 1 and 3 based on new Feed Gas data (2007 not included)

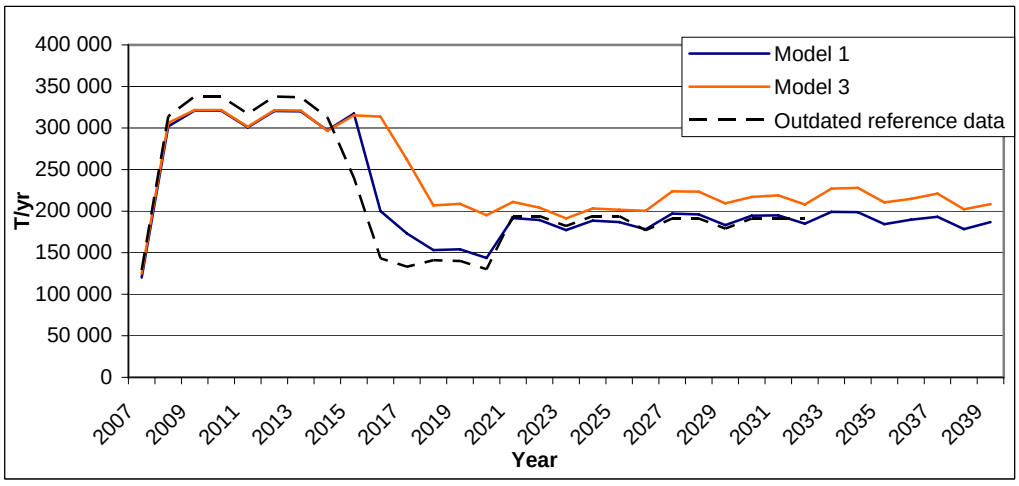

Figure 48: LPG production from based on new Feed Gas data

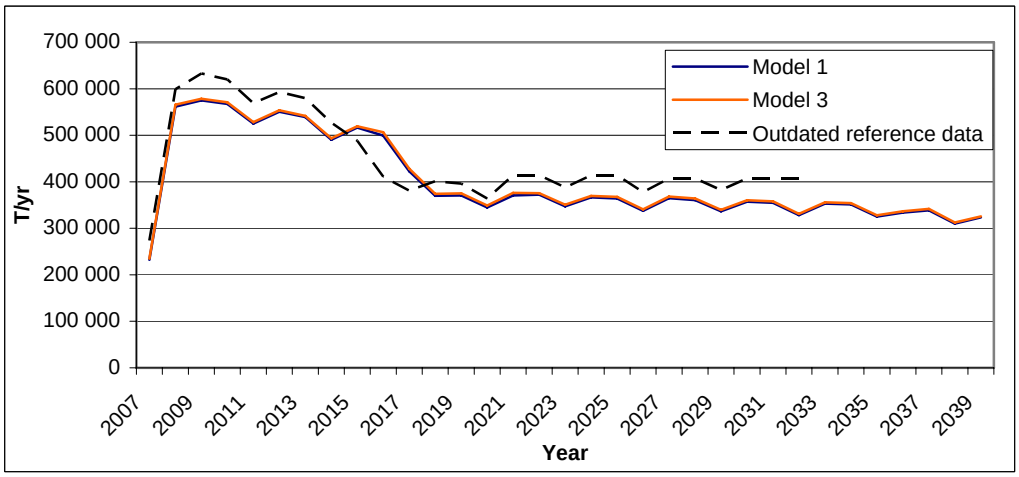

Figure 49: Condensate production based on new Feed Gas data

Table 8: Product specification for updated Feed Gas for Model 1 and Model 3

		Model 1		Model 3	
LNG		min	max	min	max
Nitrogen	mol%	0.43	0.87	0.52	0.66
Methane	mol%	91.34	93.18	92.33	93.26
Ethane	mol%	4.97	5.66	4.98	5.68
Propane	mol%	1.09	1.77	0.95	1.08
I-Butane	mol%	0.06	0.14	0.06	0.12
N-Butane	mol%	0.16	0.27	0.14	0.16
C5+	mol%	0.04	0.06	0.03	0.04
CO_2	mol%	0.004	0.005	0.004	0.005
GHV	MJ/Sm^3	39.8	40.31	39.68	39.86
LPG					
CO_2	mol%	0.00	0.00	0.00	0.00
Methane	mol%	0.00	0.00	0.00	0.00
Ethane	mol%	0.44	0.56	0.53	0.82
C5+	mol%	1.21	1.36	1.27	1.34
Condensate					
C4-	wt%	1.93	2.51	2.1	2.1

*Values outside specifications are written in red

9. Discussion

9.1. Model Accuracy

Model 1

Model 1 performs a high level of accuracy for calculating the composition of the four products as can be seen in Figure 30 and Figure 31. High accuracy was expected because the Feed Gas does not vary much in the three main periods of production. This may indicate that a small variation in the Feed Gas composition in these periods is of minor importance for the composition and flow rate of the products. The operation of the different systems (temperature, pressure and duty) will be depending on whether the feedstock is light or heavy as described in Chapter 2.2. Assigning constant split factors will therefore probably represent the unit configurations in a good way for a given feedstock. If the system configuration is the determining factor for the production, then minor changes in the Feed Gas composition may not be that important. The excel model used for product calculation at Kårstø has only one set of split factors which may indicate sufficient accuracy, but it must also be said that the natural gas specifications at Kårstø are not as strict as for LNG.

Even though the error is very close to zero, this model may not be accurate for a Feed Gas compositions deviating significantly from the characteristic compositions in the three main periods of production. This assumption can not be confirmed because simulation data for alternative Feed Gasses with corresponding product compositions and flow rates are not available.

The weakness of this model is that Fuel Gas consumption is not included as a parameter. Figure 32 clearly illustrates the discrepancy for the Fuel Gas consumption compared to Linde results in Case 01. The split factors in the first main period are based on the Fuel Gas consumption for *2007* and *2013* which is higher than in Case 01. In order to handle an increase in Fuel Consumption, a new set of split factors should be calculated when the power consumption is changed.

Based on the graph in Figure 33, the constant split model will provide an estimation of the LNG product which satisfies the GHV criteria of the LNG product. This is at least valid if the gas composition is as expected in the different time periods in Figure 7. In cases where the gas composition in one period deviates significantly from the expected composition, this model may not provide a satisfying GHV. There are no mechanisms in the model to control the propane content in the LNG like the mechanism in the HHC column described in Chapter 2.2.7.

Model 2

Comments to model validity
The mathematical approach in Model 2 is not a fully valid way of modelling the LNG process. To represent the split factors as a linear function of the Feed Gas may only be applied to the Fuel Gas because the composition of this product is entirely depending on the separation in the *Slug Catcher* which is depending on the Feed Gas composition (see Chapter 6.5.1). Linear approximation may also be applied to the split factors for methane and ethane in the LNG because these two components are only found in the Fuel Gas and the LNG. Since they add up to one by split balance and linear approximation is valid for the Fuel Gas, then linear approximation will also be valid for these components in the LNG. The compositions of LNG, LPG and Condensate are, in addition to the composition of the Feed Gas, also depending on the different product specifications which will be adjusted by the configuration for the different systems/units. Adjustments made to the product compositions makes the LNG process complex because several hydrocarbon streams are taken out and recycled back to System 12 into the main feed stream. The composition of the LNG, LPG and Condensate can therefore not entirely be represented as linear functions of the composition of the Feed Gas because they are also depending on the recycled streams. Other alternatives for the split factors for nitrogen and I-Butane in Model 2 must therefore be considered in Model 3.

The accuracy of the calculated compositions from Model 2 compared to Linde base cases in Figure 34 and Figure 35 reveals that the component errors are quite significant. Particularly the Fuel Gas and LPG composition have significant discrepancies (up to 2 mol% for some components). This is mainly a result of the linear approximation of the split factors for methane and to some extent the ethane in the LNG (Figure 21 and Figure 22) which is too inaccurate. The flow rate of LNG is about 13 times higher than the flow rate of Fuel Gas, and the LNG consists mainly of methane. A small deviation from the correct split factor for methane or ethane in of LNG will not affect the composition significantly for LNG and therefore this composition is more or less correct in all the six compared cases. The Fuel Gas on the other hand will experience major errors in composition as a result of the small deviations in the split factors for methane and ethane combined with the large flow rate of LNG. When the content of methane and ethane in the LNG is wrong, the iteration of the GHV will provide the wrong split factors for C_3 to C_5 for LNG. The split balance, which determines the composition of LPG and Condensate, will therefore transfer the error to the LPG and Condensate which can be seen from Figure 34 and Figure 35.

Model 2 does not calculate the flow rate of LPG and Fuel Gas with sufficient accuracy (Figure 36). The error is almost 8% for these products in some of the cases. The calculation of LNG and Condensate proves to have better accuracy but with the major flow rate of LNG; one percent error is still considerable. The main reasons for the poor estimation of the flow rates are partly the inaccurate estimation of the Fuel Gas composition but also that the Fuel Gas consumption is not included as a parameter in the model. Variations in Fuel Gas consumption were neglected in the

linearization of the split factors which has led to the considerable errors for the Fuel Gas flow rate. The error in LPG flow rate is mainly a result the wrong LNG composition from the GHV iteration.

Model 2 will not be representative for the LNG process at Melkøya and the specification for Condensate in Table 6 are not within the maximum values stated in Chapter 4.3.3 in some of the cases. It is necessary to improve the composition of the LNG and Fuel Gas in order to obtain results with better accuracy. It is also necessary to include the power consumption as a parameter to decrease the flow rate discrepancies. These issues have formed the basis for the improvements in Model 3. Model 2 has not been used for product forecasting because improvements are necessary.

Model 3

The product composition accuracy in Model 3 (Figure 37 and Figure 38) has generally been improved for all products in all cases when compared to the results from Model *2*, but it is still not as accurate as Model 1. The discrepancies for the Fuel Gas composition from Model 2 have been reduced considerably to almost zero due to the flash calculation described in Chapter 6.5.2. This can clearly be seen in the compared results in Figure 40. Detailed information about the temperature and pressure in the *Slug Catcher* for all the six cases has not been available and this is assumed to be the reason for minor composition discrepancies in the years 2015 and 2020 (Figure 38).

The constant split factors assigned to the nitrogen depending on whether the Feed Gas is a heavy or a light feedstock is believed to give a sufficient accuracy of the nitrogen fraction in the LNG. The error due to small variations in the nitrogen fraction in the LNG is believed to be of minor significance because the fraction is very low (less than 1 mol%).

The LPG composition has been improved to some extent compared to the results from Model 2, but it still differs a bit from the target values; especially for the year 2022. The reduction in the percentage error is mainly a result of the flash calculation which provides a more accurate estimation of the methane and ethane content in the LNG product. This makes a better basis for the GHV iteration determining the split factors for propane and the heavier components in the LNG product. The percentage error for C3 to C5 in the LPG is therefore a result of the assumption that the split factors for the butanes and the pentanes (except I-Butane) are in a constant ratio to the split factor of propane. A better estimation of these split factors for LPG and Condensate will probably be far more difficult to obtain in an Excel model. More accurate split factors for C_3 to C_5 will then be necessary and must be obtained by a better representation of the fractionation process in the *HHC Removal Column*. Such solution will be far too complex to represent in an Excel model and it is questionable if it is worth the effort. The accuracy of the Linde data is not known and the base cases may not be absolute solutions. It is probably advisable to compare the model to real production data before such an extensive analysis and model development should be carried out.

The flow rate discrepancies in Figure 39 are considered insignificant for LNG, Fuel Gas and Condensate and are better than for Model 2. The LPG flow rate is still containing some error in some of the cases but the improvement from Model 2 is quite clear. Improvements have been achieved by including the Fuel Gas consumption in the model. The increased power demand due to Feed Gas compression represents an important factor for the accuracy of the model. An accurate Fuel Gas composition and flow rate will give the right amount of methane and ethane in the LNG. This will improve the split factors for C_3 to C_5 from the GHV iteration and consequently improve the LPG flow rate.

The development of Model 3 has made it necessary to include HYSYS to obtain K-values data in order to perform the flash calculation. The intention of this work was to make a fully independent excel model without the requirement of additional programs to perform valid estimations. Based on the results in Model 3, it has been found favourable to include the flash calculation in order to make a model that is valid for different Feed Gas compositions. Even though Model 1 is more accurate for the given Feed Gas compositions presented in the Linde base cases, it is not necessarily the most correct model; especially not if the Feed Gas has a different composition than in Figure 7. The composition of the Feed Gas is decided by the mixing of the gas from the different production fields because the gas composition at Snøhvit, Albatross and Askeladd are not the same. The production rate from these fields will vary over time and may vary from the forecasted schedule of production; hence the Feed Gas composition may be different. At some point it is believed that Model 3 will be more accurate than Model 1, but with the reference data available it is not possible to validate at which point this will be true. When the LNG plant is in full operation and production data becomes available, this matter should be investigated.

9.2. Forecasted Production Accuracy

Model 1

LNG forecast
The forecasted LNG production is illustrated in Figure 41. The estimation proves to be quite accurate and the deviations are small. The fluctuating shape of the curve is caused by the overhauls every third year. The constant split factor model appears to provide a good estimation of the LNG production when the set of split factors is based on constant Feed Gas composition. It is not yet known how the model is affected by greater variations in Feed Gas composition. The reference data is only provided with a 0.1 MT/yr resolution, which makes the comparison too inaccurate.

LPG forecast
The consequences of a Feed Gas composition that deviates significantly from compositions used as basis for the calculated split factors is clearly seen from the forecasted LPG production in Figure 42. Year 2015 is included in the first production period and by inspections in Figure 7 the composition of the Feed Gas this particular year deviates significantly from the Feed Gas composition for the period 2008 – 2014. The considerable discrepancy between the reference value and the estimated value for

2015 is 76 350 tonnes. A part from year 2015, the model appears to provide accurate results for the yearly LPG production.

Condensate forecast
As for the forecasted LPG production, the forecasted Condensate production also shows discrepancy for the year 2015. The production this particular year is about 40 000 tonnes higher than the reference value. A part from year 2015, the constant split is representative for the production in the period 2007 – 2024. Between 2024 and 2032, the model is more inaccurate compared to the reference data. The split factors for this period are only based on the year 2022 and from Figure 7, it is possible to see a slight declining profile for C5+ the following years and it should be expected that the Condensate production goes down. This is further commented in the paragraph below regarding the reference data.

Model 3

LNG
The LNG production seem to be more or less correct judged from the forecasted result and will not be further commented due to the low resolution of the reference data.

LPG
A comparison between the results from Model 3 and Model 1 reveals that Model 3 has a bit poorer accuracy. The forecasted production of LPG seems to be quite accurate from 2007 to 2014 and from 2022 to 2032. The period in the middle shows more deviation and this was also expected based on the results in Figure 39 where the production of LPG is about 6 % too high on mass basis. It is difficult to obtain a more accurate split factor for the propane in the LPG because this split factor is very sensitive to the GHV of the LNG. The year 2015 has the largest discrepancy but compared to Model 1, where the discrepancy is 76 350 T/yr, the discrepancy is reduced to 62 683 T/yr.

Condensate
The Condensate product has more or less the same profile as the Condensate product calculated by Model 1. Significant discrepancy for the year 2015 can also be seen from the results from Model 3.

Further Comments to Reference Data

Although the reference data is supposed to be the most accurate, there are some logical correlations that seem a bit wrong for the reference values in 2015 and the last main period of production. More detailed information about the calculation of the reference data has not been available so the content of this paragraph is just speculations.

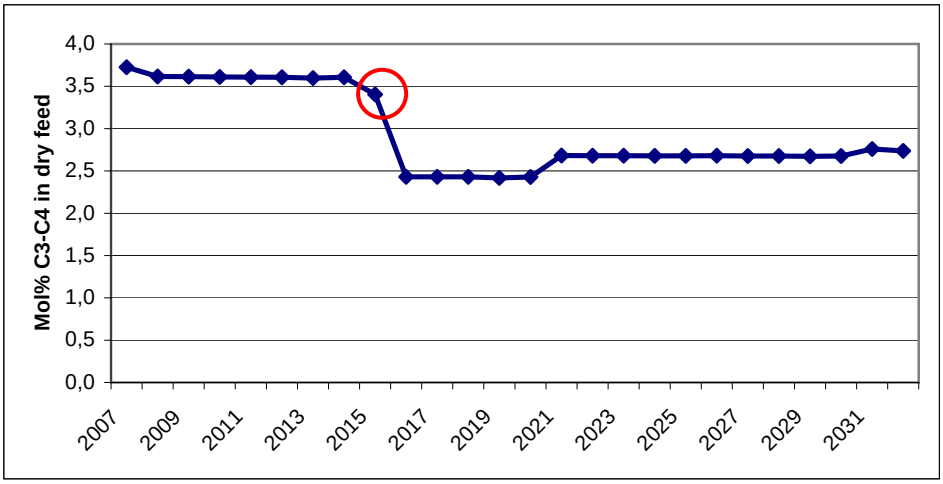

Figure 50: Total mol fraction C3 and C4 in Feed Gas

Figure 50 shows the sum of the mol% propane and butanes in the Feed Gas. A small drop in the mol fraction occurs for these components in 2015. These are the main components of the LPG product and this small drop should lead to a small drop in the LPG production this year followed by a more significant drop in 2016. The component fractions illustrated for the Feed Gas in Figure 7 also implies that the change in the production rate for the different products between 2014 and 2015 should be smaller than the change between 2015 and 2016 where the Feed Gas composition is changed far more considerably. When inspecting the reference production rate for LPG in Figure 45 the drop in LPG production in 2015 seems to be higher than the change in Feed Gas composition implies. In this particular year, the estimated data seems to give a better representation of this effect. The same relationship can bee seen for the Condensate the same year. The declining profile of the C5+ in the period 2022 to 2032 should correspond to a reduced production rate of Condensate in this period, which is not the case for the reference data. The calculations from the reference data may not be as accurate as hoped and therefore it is difficult to give absolute answers for the accuracy of the forecasted production. Real production data will probably be the best source to validate the accuracy of Model 1 and 3.

9.3. Updated Feed Gas Data

The composition of the Feed Gas in the updated data is significantly different from the original composition used to develop the mass balance models (see appendix F). Even though there are no updated reference data to compare the results to, there are clear indications that Model 3 is most correct. First of all Model 3 fulfils all the product specifications (in Table 8) which Model 1 doesn't (see GHV criteria for LNG and C4(-) criteria for Condensate). Another indication is the high diversification of the max and min values in Table 8 for Model 1 compared to the ones for Model 3. This was also expected since Model 3 is a more physical representation of the real plant behaviour with mechanisms included to control/manipulate the composition and specifications of the different products.

The results from forecasted production of LNG for the years 2008 to 2039 (2007 not included) is found in Figure 47. As stated in Chapter 7.1.1, the flow rate will go down in the second and third main periods. This means that the production of LNG will go down if the Feed Gas composition remains the same. The amount of produced LNG will also be mostly depending on the fraction of methane in the Feed Gas since this is the main component. Especially between 2015 and 2022, the updated Feed Gas has a lower fraction of methane than the Feed Gas in the "outdated" reference date. Both these factors should therefore indicate that the LNG production goes down. From the figure it can be seen that Model 3 is following the expected production behaviour while Model 1 doesn't.

Both models calculate the same Condensate production which is corresponding well to the profile of the C5+ fraction in the Feed Gas. Good accuracy was more or less expected since most of the C5 and approximately all the C6+ components are found in the Condensate. This may also strengthen the theory that the forecasted condensate production from the "outdated" reference data is wrong.

The calculated LPG production from the two models is very different as can be seen from Figure 48 and Figure 49. It is difficult to state the accuracy these calculations because the LPG product is not clearly defined by the product specifications. Based on Figure 45 it is believed that the forecasted LPG production from Model 3 will be quite accurate. The calculation of LPG in Model 1 is questionable. As a result of the inaccurate LNG calculation for this model, it is assumed that the LPG calculation is also wrong because the LPG product is very much depending on the LNG composition.

It is believed that Model 3 has sufficient accuracy to perform qualified calculations of the production at Melkøya. For further validation of Model 3, comparison to real production data from the LNG plant at Melkøya will probably be the best solution.

10. Conclusion

Three mass balance models have been developed in this work based on simulation data from six base cases developed by Linde. The base cases are representing the different phases of production according to six design feedstocks which have been used in the design of the LNG plant at Melkøya. The models are made in Microsoft Excel and developed with the intention to make quick estimations of the production of LNG, LPG, Condensate, Fuel Gas, CO_2 and Nitrogen at Melkøya LNG plant.

All the models are based on the use of split factors that distributes the components in the Feed Gas to the different products. Model 1 is based three sets of average constant split factors used in the three main periods of the production. Model 2 is a mathematical approach that calculates some of the split factors as a linear function of the Feed Gas composition to make the model able to handle a wider range of Feed Gas compositions. Model 3 is based on a more careful study of the physical process at Melkøya. The mechanisms implemented in this model are based on real system behaviour in order to fulfil product specifications. These mechanisms have made the model a bit more advanced and HYSYS has been necessary in order to obtain the necessary input data.

All the three models have been tested for composition and flow rate accuracy against six Linde base cases. Model 2 proved to be too inaccurate and therefore only Model 1 and Model 3 have been used for production forecasting.

Feed Gas and production data from a production forecasting report made in January 2006 has been used as reference data to validate the accuracy of the forecasted production from Model 1 and Model 3. The forecasted results from the two models correspond well to the LNG and LPG production in the reference report, but both models have a considerable discrepancy to the reference data for the estimation of Condensate in the period 2022 to 2032. This discrepancy is assumed to be a result of inaccuracy in the calculation of Condensate in the reference data. The forecasted production from the two models is quite similar and therefore it has not been possible to state which model that is most suited for production forecasting based on reference data from Linde.

The production from recently updated Feed Gas data (December 2007) have been calculated by Model 1 and Model 3. Model 1 has been found invalid because it does not satisfy product specifications. The products calculated from Model 3 were all within the product specifications. The calculated composition and flow rate of LNG and Condensate are believed to be quite accurate (<0.5 mol% discrepancy in composition and <1% discrepancy in flow rate on mass basis compared to Linde base cases), while some uncertainties are connected to the LPG product (<2.1 mol% discrepancy in composition and <5% discrepancy in flow rate on mass basis compared to Linde base cases).

11. Recommendations for Further Work

Before further improvements should be made to Model 3, it is favourable to compare the results from the model to real production data from Melkøya. Comparison to real production data was originally a part of this assignment, but due to start-up problems at Melkøya, such data has not become available. It is not known when the plant will start up again (December 2007) and when it does, it will take time to ramp the production up to 100% (about 7 months from start up). The model, as it is now, is only based on "outdated" simulation data used in the design of the LNG plant. The updated Feed Gas data has a quite different composition over time and old data may therefore not be fully representative for the physical LNG process at Melkøya as it is today. When production data becomes available, it will probably be necessary to tune the model to better represent the plant configuration for light and heavy feedstocks.

The updated Feed Gas data arrived just a few days before the deadline for this work and therefore the results from the forecasted production from Model 3 have not been fully analysed. A more careful analysis of the results in relation to the Feed Gas composition may be necessary to validate the production according to limits in the system equipment. Due to limitation in the equipment, the Feed Gas flow rate is reduced in 2015 as a result of the change in Feed Gas composition. The maximum flow rate of Feed Gas for the production years may therefore have changed and this will affect the yearly production. Updated Feed Gas flow rates should be implemented in the model.

The implemented mechanisms in Model 3 are relatively simple because Excel is not suited to perform advanced calculations. For this purpose it is necessary to make more advanced macro functions in Visual Basic. If Model 3 turns out to be too inaccurate, even after further tuning based on production data, it may be necessary to include a better representation of the *HHC Removal Column* to better determine the fraction of C3 to C5 when the Feed Gas is changing. It will therefore also be necessary to include System 26 and System 20 in order to include the recycled hydrocarbon streams which are of significance for the fractionation process in the *HHC Removal Column*. This will be a more complex model and it will probably be necessary connect the Excel model to HYSYS to perform some of the calculations.

The Excel model is not connected to HYSYS to provide the K-values (for Fuel Gas calculation) automatically due to limited experience with Visual Basic. It is possible to make a script that does this operation and it will certainly make the model more user-friendly.

12. References

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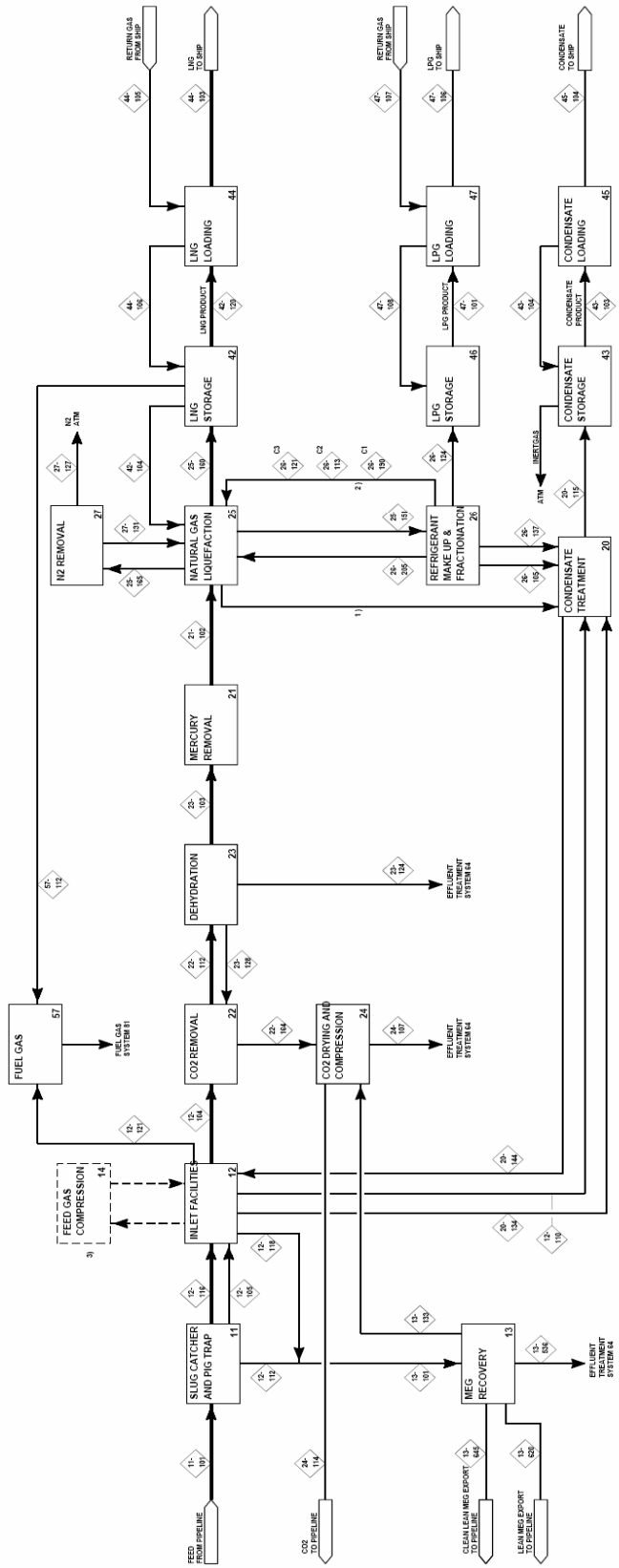
Appendix A



1. STREAM NUMBER	UNITS	26-124	26-205	27-127	42-120	43-103	44-103	44-105	45-104	47-101	47-106	47-107	47-108
2. PHASE		LIQ TOT.	LIQ TOT.	VAP	LIQ TOT.	LIQ TOT.	LIQ TOT.	VAP	LIQ TOT.	LIQ TOT.	LIQ TOT.	VAP	VAP
3. MOLAR FLOW RATE	KMOL/H	882.9	348.2	686.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
4. TOTAL FLOW RATE	10^3 KG/H	43.8	22.3	19.1	5677.8	3548.8	5677.8	35.6	3548.8	2034.7	2034.7	10.0	10.0
5. FLOWRATE AT 15° C	SM^3/H	20875.0	8232.9	16242.1	7727517	806975	7727517	46532	806975	969828	969827	5241.0	5241.0
6. FLOWRATE AT F.C.	M^3/H	73.0	33.7	1321.6.1	12584	4802.8	12597	14862	4806.8	3398.5	3402.3	2800.8	2762.8
7. MOLECULARWEIGHT	KG/KMOL	49.63	63.92	27.77	17.37	103.98	17.37	18.09	103.58	49.60	49.60	45.11	45.11
8. OPERATING PRESS.	BARA	8.7	60.8	1.3	5.7	6.4	3.2	1.8	2.5	6.2	2.7	1.8	1.7
9. OPERATING TEMP.	°C	-34.0	-33.8	17.3	-162.4	17.5	-162.2	-106.58	17.7	-33.3	-33.1	-8.1	-6.4
10. DENSITY F.C.	KG/M^3	659.83	659.69	1.44	451.21	738.91	450.72	2.39	738.29	598.71	598.04	3.84	3.61
11. VISCOSITY	10^3 NS/M^2	0.2325	0.3449	0.0174	0.1545	0.5495	0.1535	0.0073	0.5465	0.2303	0.2291	0.0074	0.0075
12. SPECIFIC HEAT	KJ/KG°C	2.208	2.107	1.045	3.188	2.027	3.191	1.846	2.027	2.211	2.214	1.616	1.617
13. ENERGY CONTENT	MW	-5.131	-3.011	-0.044	-1411.237	-381.936	-1410.907	-2.445	-381.968	-284.077	-284.050	-0.169	-0.161

1. STREAM NUMBER	UNITS	11-101	11-101	11-101	12-104	12-116	12-116	12-121	13-101	13-133	13-536	13-620	13-645	26-115	26-134	20-144	21-102	22-112	23-103	24-114	25-160
2. PHASE		VAP	LIQ TOT.	LIQ2	VAP	VAP	LIQ TOT.	VAP	LIQ TOT.	VAP	LIQ TOT.	LIQ TOT.	LIQ TOT.	LIQ TOT.	LIQ TOT.	VAP	VAP	VAP	VAP	LIQ TOT.	LIQ TOT.
3. MOLAR FLOW RATE	KMOL/H	34698.5	2365.6	429.3	34819.2	34695.1	21.0	2030.9	435.3	5.6	281.5	192.9	87.1	901.5	1240.9	2150.1	32946.8	36095.1	32946.8	1824.2	31137.9
4. TOTAL FLOW RATE	10^3 KG/H	681.7	134.8	13.6	703.5	681.3	1.2	39.9	13.8	0.2	5.1	9.3	4.3	93.2	103.9	61.8	622.1	681.5	622.1	80.2	541.6
5. FLOWRATE AT 15° C	SM^3/H	820431.4	56406.2	10163.8	823284.4	820351.6	495.8	48020.7	10293.6	132.2	6657.0	4561.9	2060.4	21316.2	29341.0	50837.2	779012.8	853452.4	779012.8	4312.2	736242.6
6. FLOWRATE AT F.C.	M^3/H	8721.3	195.4	12.7	10195.2	8833.0	1.7	551.7	12.9	139.9	5.1	8.3	3.9	126.4	150.2	1051.8	11329.2	11829.0	11100.7	92.5	1205.0
7. MOLECULARWEIGHT	KG/KMOL	19.65	56.50	31.59	20.20	19.64	56.57	19.64	31.61	31.15	18.02	48.23	49.89	103.34	83.69	28.74	18.88	18.88	18.88	43.98	17.39
8. OPERATING PRESS.	BARA	70.0	70.0	70.0	66.8	69.2	66.8	69.2	1.2	5.5	364.0	364.5	4.6	24.0	67.0	67.0	62.3	64.9	63.6	150.0	4.5
9. OPERATING TEMP.	°C	-1.0	-1.0	-1.0	14.9	-1.4	-1.4	1.5	-1.4	74.8	25.0	13.4	12.9	17.4	28.1	162.4	26.9	26.6	27.5	27.9	-160.8
10. DENSITY F.C.	KG/M^3	78.16	690.00	1073.20	69.00	77.14	690.42	72.31	1069.85	1.24	995.86	1119.61	1123.27	737.27	691.27	58.75	54.91	57.61	56.04	866.47	449.45
11. VISCOSITY	10^3 NS/M^2	0.0130	1.0179	11.7174	0.0130	0.0129	1.0281	0.0128	11.7224	0.0129	0.3063	17.9936	19.5416	0.5397	0.3212	0.0155	0.0126	0.0127	0.0126	0.0791	0.1482
12. SPECIFIC HEAT	KJ/KG°C	2.803	2.236	3.123	2.599	2.792	2.233	2.704	3.119	1.544	4.182	2.576	2.536	2.030	2.148	2.579	2.615	2.646	2.627	2.343	3.191
13. ENERGY CONTENT	MW	-27.024	-20.152	-6.435	-19.470	-26.951	-0.178	-1.453	-6.509	0.003	-3.442	-3.218	-1.474	-10.035	-10.346	3.907	-11.464	-13.274	-11.464	-5.676	-133.396

Appendix B

Linde base case data

Stream	Dry Feed															
Feedstock No.		Snohvit A		1		2		3		4		5				
Year		Case 01		2007		2013		2015		2020		2022				
Run ID		MH56.C1R4		MH56.Sens_F07_R1		MH56.Sens_F13_R1		MH56.Sens_F15_R1		MH56.Sens_F20_R1		MH56.Sens_F22_R1				
N2	mol / wt %	2,5234	3,2270	2,7832	3,5844	2,7859	3,6276	1,7165	2,3504	1,7200	2,3616	2,0060	2,7235			
CO2	mol / wt %	5,2588	10,5652	5,9258	11,9894	5,9193	12,1090	5,9233	12,7421	5,9678	12,8728	6,0198	12,8396			
METHANE	mol / wt %	80,9561	59,2897	80,5264	59,3912	80,7065	60,1841	84,3934	66,1795	84,4332	66,3908	83,4976	64,9199			
ETHANE	mol / wt %	5,0239	6,8964	5,0139	6,9312	5,0144	7,0087	4,1336	6,0757	4,1211	6,0737	4,3383	6,3223			
PROPANE	mol / wt %	2,5324	5,0979	2,4480	4,9627	2,4428	5,0071	1,6765	3,6137	1,6604	3,5886	1,8426	3,9378			
I BUTANE	mol / wt %	0,3998	1,0607	0,3793	1,0136	0,3777	1,0205	0,2545	0,7232	0,2512	0,7158	0,2799	0,7886			
N BUTANE	mol / wt %	0,8295	2,2010	0,7874	2,1041	0,7824	2,1140	0,5012	1,4239	0,4944	1,4085	0,5602	1,5780			
I PENTANE	mol / wt %	0,2808	0,9250	0,2606	0,8646	0,2574	0,8634	0,1655	0,5836	0,1627	0,5752	0,1835	0,6415			
N PENTANE	mol / wt %	0,3078	1,0138	0,2880	0,9554	0,2838	0,9519	0,1760	0,6206	0,1730	0,6117	0,1976	0,6908			
C6	mol / wt %	0,3518	1,3644	0,2497	0,9755	0,2415	0,9535	0,1548	0,6427	0,1509	0,6282	0,1695	0,6978			
C7	mol / wt %	0,3908	1,6625	0,3330	1,4269	0,3150	1,3648	0,1885	0,8586	0,1813	0,8280	0,2096	0,9465			
C8	mol / wt %	0,3168	1,5547	0,2982	1,4739	0,2744	1,3714	0,1765	0,9272	0,1688	0,8892	0,1864	0,9709			
C9	mol / wt %	0,1419	0,8143	0,1738	1,0041	0,1563	0,9130	0,1312	0,8059	0,1263	0,7780	0,1278	0,7784			
BENZENE	mol / wt %	0,0780	0,2780	0,0688	0,2470	0,0691	0,2508	0,0452	0,1725	0,0448	0,1715	0,0499	0,1889			
TOLUENE	mol / wt %	0,0899	0,3783	0,0805	0,3409	0,0809	0,3464	0,0679	0,3057	0,0678	0,3061	0,0701	0,3130			
M XYLENE	mol / wt %	0,0610	0,2955	0,0550	0,2684	0,0552	0,2724	0,0686	0,3559	0,0690	0,3590	0,0651	0,3349			
C10	mol / wt %	0,1409	0,8556	0,0796	0,4866	0,0688	0,4252	0,0698	0,4537	0,0672	0,4380	0,0646	0,4163			
C11	mol / wt %	0,0630	0,4225	0,0584	0,3946	0,0479	0,3272	0,0464	0,3333	0,0439	0,3162	0,0421	0,2999			
C12	mol / wt %	0,0620	0,4497	0,0435	0,3179	0,0331	0,2446	0,0311	0,2417	0,0284	0,2213	0,0269	0,2072			
C13	mol / wt %	0,0490	0,3867	0,0317	0,2521	0,0220	0,1769	0,0202	0,1708	0,0179	0,1517	0,0168	0,1408			
C14	mol / wt %	0,0330	0,2830	0,0234	0,2022	0,0148	0,1293	0,0134	0,1231	0,0115	0,1059	0,0105	0,0956			
C15	mol / wt %	0,0250	0,2304	0,0178	0,1653	0,0099	0,0929	0,0087	0,0859	0,0069	0,0683	0,0061	0,0597			
C16	mol / wt %	0,0150	0,1458	0,0112	0,1097	0,0052	0,0515	0,0047	0,0494	0,0034	0,0360	0,0028	0,0294			
C17	mol / wt %	0,0150	0,1574	0,0112	0,1184	0,0052	0,0556	0,0047	0,0534	0,0034	0,0389	0,0028	0,0318			
C18	mol / wt %	0,0100	0,1127	0,0060	0,0681	0,0021	0,0241	0,0017	0,0205	0,0010	0,0121	0,0008	0,0096			
C19	mol / wt %	0,0070	0,0824	0,0042	0,0498	0,0015	0,0180	0,0012	0,0151	0,0007	0,0088	0,0005	0,0063			
C20+	mol / wt %	0,0170	0,2443	0,0205	0,2968	0,0062	0,0908	0,0043	0,0662	0,0025	0,0386	0,0017	0,0259			
H2S	mol / wt %	0,0005	0,0008	0,0005	0,0008	0,0005	0,0008	0,0005	0,0008	0,0005	0,0008	0,0005	0,0008			
COS	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000			
PHENOL	mol / wt %	0,0002	0,0007	0,0002	0,0007	0,0002	0,0007	0,0002	0,0007	0,0002	0,0007	0,0002	0,0007			
METHANOL	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000			
MEG	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000			
HE	mol / wt %	0,0200	0,0037	0,0200	0,0037	0,0200	0,0037	0,0200	0,0039	0,0200	0,0039	0,0200	0,0039			
H2O	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000			
Temperature	°C	-1,0		-1,0		-1,0		-5,0		-5,0		-1,0				
Pressure	bar(a)	70,0		70,0		70,0		35,0		35,0		70,0				
Flow	Sm3/hr	866667		866667		866667		861467		861467		863200				
	MSm3/day	20,80		20,80		20,80		20,68		20,68		20,72				
	t/hr	802,93		797,31		788,56		745,38		743,36		753,29				

Stream	LNG												
Feedstock No.		Snohvit A		1		2		3		4		5	
Year		Case 01		2007		2013		2015		2020		2022	
N2	mol / wt %	0,7080	1,1417	0,7319	1,1799	0,7328	1,1814	0,5955	0,9676	0,5944	0,9659	0,5917	0,9605
CO2	mol / wt %	0,0050	0,0127	0,0050	0,0127	0,0050	0,0127	0,0050	0,0128	0,0050	0,0129	0,0051	0,0129
METHANE	mol / wt %	92,1608	85,1060	92,1238	85,0505	92,1293	85,0558	93,3169	86,8374	93,3341	86,8650	93,1140	86,5651
ETHANE	mol / wt %	5,7082	9,8801	5,7255	9,9075	5,7116	9,8836	4,5199	7,8836	4,5054	7,8593	4,8256	8,4088
PROPANE	mol / wt %	1,1383	2,8894	1,1334	2,8762	1,1422	2,8985	1,2343	3,1572	1,2303	3,1473	1,1232	2,8703
I BUTANE	mol / wt %	0,0644	0,2153	0,0645	0,2157	0,0640	0,2141	0,1011	0,3409	0,1022	0,3446	0,0970	0,3268
N BUTANE	mol / wt %	0,1726	0,5776	0,1723	0,5762	0,1717	0,5743	0,1850	0,6239	0,1864	0,6286	0,1988	0,6697
I PENTANE	mol / wt %	0,0305	0,1265	0,0310	0,1288	0,0308	0,1279	0,0299	0,1252	0,0299	0,1251	0,0316	0,1320
N PENTANE	mol / wt %	0,0121	0,0504	0,0126	0,0522	0,0124	0,0515	0,0122	0,0511	0,0122	0,0510	0,0128	0,0537
C6	mol / wt %	0,0000	0,0002	0,0000	0,0002	0,0000	0,0002	0,0000	0,0002	0,0000	0,0002	0,0000	0,0002
C7	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C8	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C9	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
BENZENE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
TOLUENE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
M XYLENE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C10	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C11	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C12	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C13	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C14	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C15	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C16	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C17	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C18	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C19	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C20+	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2S	mol / wt %	0,0000	0,0001	0,0000	0,0001	0,0000	0,0000	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001
COS	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
PHENOL	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANOL	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
MEG	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
HE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Temperature	°C	-162,6		-162,7		-162,7		-162,3		-162,3		-162,3	
Pressure	bar(a)	1,05		1,05		1,05		1,05		1,05		1,05	
Flow	kmol/hr	30468,0		30047,6		30122,1		30740,0		30747,5		30508,0	
	t/hr	529,32		522,14		523,44		529,96		530,02		526,47	
	t/day	12704		12531		12563		12719		12720		12635	
	m3/day	28168		27774		27842		28402		28408		28195	

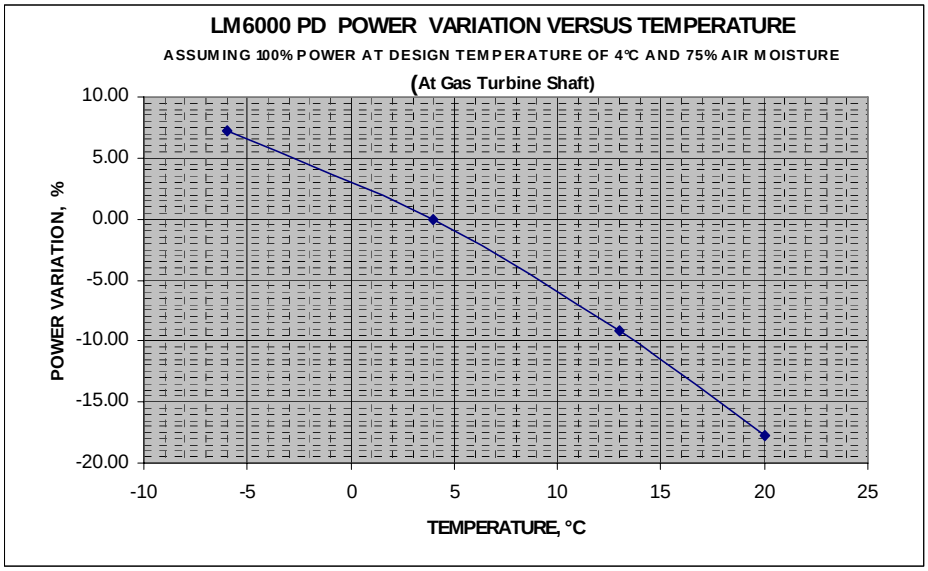
Stream	LPG												
Feedstock No.		Snohvit A		1		2		3		4		5	
Year		Case 01		2007		2013		2015		2020		2022	
N2	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
CO2	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
ETHANE	mol / wt %	0,5166	0,3131	0,5166	0,3132	0,5166	0,3129	0,5178	0,3082	0,5178	0,3082	0,5167	0,3125
PROPANE	mol / wt %	60,9645	54,1930	61,0043	54,2350	60,6854	53,9024	54,5300	47,5969	54,5159	47,5828	60,1598	53,3561
I BUTANE	mol / wt %	13,1914	15,4563	13,1240	15,3791	13,1947	15,4479	15,1990	17,4865	15,1610	17,4422	13,4367	15,7079
N BUTANE	mol / wt %	24,0499	28,1790	24,0805	28,2183	24,3249	28,4788	28,3963	32,6702	28,4489	32,7295	24,6077	28,7671
I PENTANE	mol / wt %	0,9600	1,3963	0,9569	1,3920	0,9592	1,3939	1,0308	1,4722	1,0300	1,4709	0,9662	1,4020
N PENTANE	mol / wt %	0,3160	0,4596	0,3165	0,4603	0,3178	0,4619	0,3247	0,4638	0,3250	0,4641	0,3116	0,4521
C6	mol / wt %	0,0013	0,0022	0,0010	0,0016	0,0010	0,0017	0,0010	0,0018	0,0011	0,0018	0,0010	0,0017
C7	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C8	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C9	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
BENZENE	mol / wt %	0,0002	0,0003	0,0002	0,0003	0,0002	0,0003	0,0002	0,0003	0,0002	0,0004	0,0002	0,0003
TOLUENE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
M XYLENE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C10	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C11	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C12	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C13	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C14	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C15	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C16	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C17	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C18	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C19	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C20+	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2S	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
COS	mol / wt %	0,0001	0,0001	0,0001	0,0001	0,0001	0,0001	0,0002	0,0002	0,0002	0,0002	0,0002	0,0002
PHENOL	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANOL	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
MEG	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
HE	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	mol / wt %	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Temperature	°C	-33,5		-33,5		-33,4		-31,7		-31,7		-33,3	
Pressure	bar(a)	1,05		1,05		1,05		1,05		1,05		1,05	
Flow	kmol/hr	886,8		835,7		830,6		346,9		339,0		478,2	
	t/hr	43,99		41,45		41,24		17,53		17,13		23,78	
	t/day	1056		995		990		421		411		571	
	m3/day	1765		1663		1654		701		685		953	

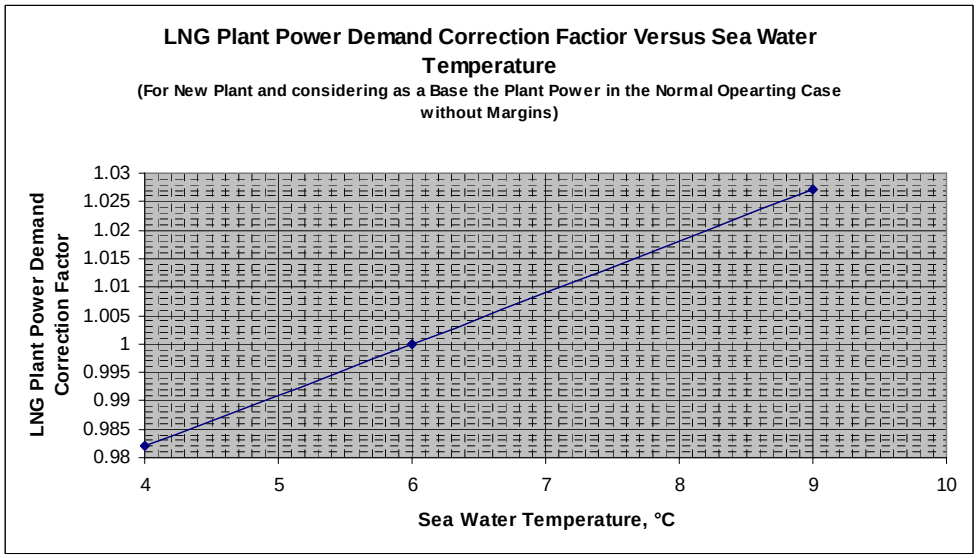
Stream	Condensate												
Feedstock No.		**Snohvit A**		**1**		**2**		**3**		**4**		**5**	
Year		**Case 01**		**2007**		**2013**		**2015**		**2020**		**2022**	
N2	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
CO2	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANE	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
ETHANE	*mol / wt %*	0,0012	0,0003	0,0012	0,0003	0,0011	0,0003	0,0014	0,0004	0,0014	0,0004	0,0013	0,0004
PROPANE	*mol / wt %*	0,0799	0,0341	0,0763	0,0328	0,0709	0,0319	0,0759	0,0329	0,0747	0,0327	0,0735	0,0328
I BUTANE	*mol / wt %*	0,5179	0,2913	0,5046	0,2859	0,4740	0,2808	0,5315	0,3032	0,5236	0,3020	0,5051	0,2968
N BUTANE	*mol / wt %*	3,1542	1,7742	3,1426	1,7809	3,0167	1,7870	3,0919	1,7635	3,0600	1,7649	3,0120	1,7700
I PENTANE	*mol / wt %*	9,1952	6,4203	9,7985	6,8929	10,5185	7,7346	9,0368	6,3982	9,1972	6,5848	9,7490	7,1116
N PENTANE	*mol / wt %*	11,5591	8,0709	12,5268	8,8121	13,4388	9,8820	11,4465	8,1043	11,6705	8,3556	12,4949	9,1147
C6	*mol / wt %*	14,1491	11,6331	11,7020	9,6933	12,3284	10,6748	11,1461	9,2926	11,2831	9,5123	11,7807	10,1193
C7	*mol / wt %*	15,8143	14,2632	15,7296	14,2933	16,2231	15,4095	13,7557	12,5805	13,7439	12,7107	14,7527	13,9012
C8	*mol / wt %*	12,8544	13,3725	14,1345	14,8145	14,1852	15,5411	12,9477	13,6585	12,8658	13,7242	13,1973	14,3436
C9	*mol / wt %*	5,7629	7,0101	8,2443	10,1038	8,0866	10,3595	9,6394	11,8901	9,6418	12,0264	9,0657	11,5212
BENZENE	*mol / wt %*	3,0722	2,3224	3,1547	2,4026	3,4462	2,7435	3,1880	2,4437	3,2794	2,5419	3,3804	2,6697
TOLUENE	*mol / wt %*	3,6141	3,2226	3,7733	3,3898	4,1311	3,8794	4,9277	4,4555	5,1105	4,6726	4,8906	4,5559
M XYLENE	*mol / wt %*	2,4696	2,5373	2,6024	2,6938	2,8481	3,0816	5,0276	5,2378	5,2540	5,5350	4,6023	4,9400
C10	*mol / wt %*	5,7264	7,3703	3,7794	4,9008	3,5632	4,8298	5,1341	6,7007	5,1362	6,7785	4,5900	6,1721
C11	*mol / wt %*	2,5592	3,6407	2,7737	3,9754	2,4817	3,7180	3,4144	4,9252	3,3568	4,8964	2,9931	4,4484
C12	*mol / wt %*	2,5189	3,8758	2,0664	3,2034	1,7153	2,7796	2,2889	3,5713	2,1720	3,4269	1,9130	3,0753
C13	*mol / wt %*	1,9909	3,3330	1,5060	2,5401	1,1402	2,0103	1,4868	2,5240	1,3691	2,3502	1,1949	2,0899
C14	*mol / wt %*	1,3408	2,4394	1,1117	2,0377	0,7671	1,4697	0,9863	1,8196	0,8796	1,6409	0,7468	1,4195
C15	*mol / wt %*	1,0158	1,9856	0,8457	1,6655	0,5131	1,0563	0,6404	1,2694	0,5278	1,0579	0,4339	0,8861
C16	*mol / wt %*	0,6095	1,2563	0,5321	1,1050	0,2695	0,5850	0,3496	0,7308	0,2639	0,5577	0,2027	0,4365
C17	*mol / wt %*	0,6095	1,3565	0,5321	1,1932	0,2695	0,6317	0,3496	0,7891	0,2639	0,6023	0,2027	0,4714
C18	*mol / wt %*	0,4063	0,9712	0,2851	0,6865	0,1088	0,2740	0,1251	0,3033	0,0765	0,1875	0,0569	0,1421
C19	*mol / wt %*	0,2844	0,7101	0,1995	0,5019	0,0777	0,2044	0,0883	0,2236	0,0535	0,1371	0,0356	0,0928
C20+	*mol / wt %*	0,6907	2,1056	0,9739	2,9911	0,3213	1,0316	0,3165	0,9783	0,1912	0,5977	0,1209	0,3851
H2S	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
COS	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
PHENOL	*mol / wt %*	0,0033	0,0030	0,0034	0,0031	0,0035	0,0034	0,0035	0,0032	0,0035	0,0033	0,0037	0,0035
METHANOL	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
MEG	*mol / wt %*	0,0003	0,0002	0,0002	0,0001	0,0002	0,0001	0,0002	0,0001	0,0002	0,0001	0,0001	0,0001
HE	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Temperature	*°C*	17,6		17,5		17,5		17,4		17,4		17,4	
Pressure	*bar(a)*	1,05		1,05		1,05		1,05		1,05		1,05	
Flow	*kmol/hr*	901,5		771,3		707,0		494,9		476,2		513,1	
	t/hr	93,16		79,11		69,37		50,43		47,99		50,75	
	t/day	2236		1899		1665		1210		1152		1218	
	m3/day	3035		2581		2286		1637		1561		1661	

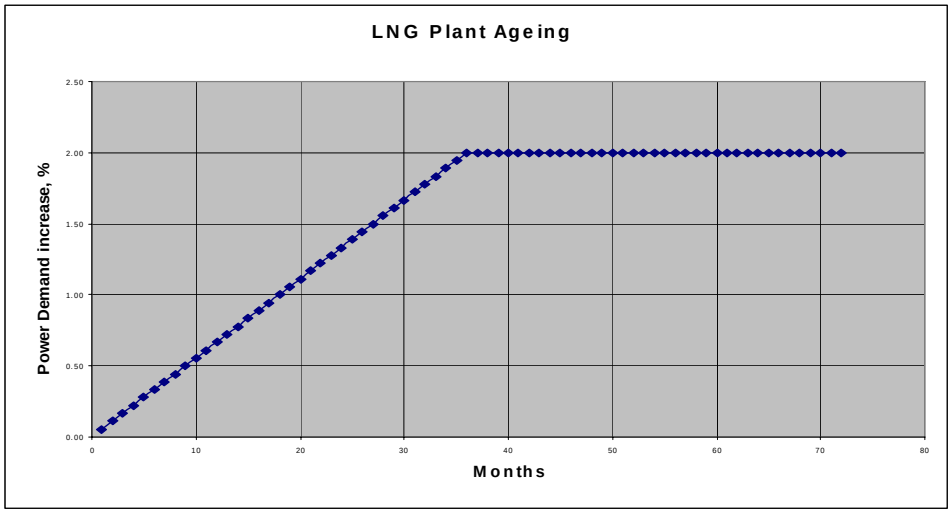
Stream	Fuel	(split downstream Inlet Filter Separator)											
Feedstock No.		Snohvit A		1		2		3		4		5	
Year		Case 01		2007		2013		2015		2020		2022	
N2	*mol / wt %*	2,6493	3,7785	2,9001	4,0919	2,8923	4,0766	1,7178	2,4671	1,7207	2,4711	2,0525	2,9457
CO2	*mol / wt %*	5,3110	11,9001	5,9746	13,2435	5,9624	13,2027	5,9677	13,4649	6,0103	13,5602	6,0450	13,6296
METHANE	*mol / wt %*	83,9463	68,5662	83,0334	67,0941	82,9865	66,9864	85,0570	69,9588	85,0575	69,9547	84,9523	69,8228
ETHANE	*mol / wt %*	4,8595	7,4395	4,8736	7,3812	4,8850	7,3908	4,5263	6,9780	4,5051	6,9448	4,2716	6,5806
PROPANE	*mol / wt %*	2,1241	4,7689	2,1047	4,6746	2,1254	4,7156	1,6777	3,7928	1,6609	3,7547	1,6892	3,8161
I BUTANE	*mol / wt %*	0,2662	0,7878	0,2662	0,7793	0,2722	0,7960	0,2481	0,7393	0,2451	0,7303	0,2256	0,6718
N BUTANE	*mol / wt %*	0,4835	1,4309	0,4904	1,4356	0,5044	1,4752	0,4656	1,3875	0,4607	1,3726	0,4166	1,2405
I PENTANE	*mol / wt %*	0,1081	0,3970	0,1108	0,4025	0,1154	0,4191	0,1074	0,3972	0,1071	0,3962	0,1032	0,3816
N PENTANE	*mol / wt %*	0,1013	0,3721	0,1057	0,3842	0,1107	0,4020	0,0988	0,3656	0,0989	0,3660	0,0990	0,3660
C6	*mol / wt %*	0,0636	0,2752	0,0517	0,2212	0,0540	0,2307	0,0455	0,1980	0,0455	0,1984	0,0524	0,2281
C7	*mol / wt %*	0,0287	0,1360	0,0285	0,1337	0,0296	0,1389	0,0205	0,0980	0,0204	0,0973	0,0284	0,1354
C8	*mol / wt %*	0,0089	0,0490	0,0099	0,0536	0,0101	0,0548	0,0067	0,0366	0,0066	0,0362	0,0100	0,0548
C9	*mol / wt %*	0,0021	0,0135	0,0031	0,0193	0,0031	0,0195	0,0025	0,0161	0,0025	0,0160	0,0036	0,0235
BENZENE	*mol / wt %*	0,0102	0,0405	0,0104	0,0408	0,0114	0,0446	0,0101	0,0405	0,0103	0,0412	0,0112	0,0449
TOLUENE	*mol / wt %*	0,0041	0,0193	0,0043	0,0200	0,0048	0,0224	0,0046	0,0219	0,0048	0,0226	0,0059	0,0277
M XYLENE	*mol / wt %*	0,0010	0,0056	0,0011	0,0059	0,0012	0,0066	0,0016	0,0088	0,0017	0,0092	0,0021	0,0111
C10	*mol / wt %*	0,0007	0,0047	0,0005	0,0031	0,0005	0,0031	0,0005	0,0032	0,0005	0,0032	0,0006	0,0042
C11	*mol / wt %*	0,0001	0,0010	0,0001	0,0011	0,0001	0,0010	0,0001	0,0010	0,0001	0,0010	0,0002	0,0013
C12	*mol / wt %*	0,0001	0,0004	0,0000	0,0003	0,0000	0,0003	0,0000	0,0002	0,0000	0,0002	0,0000	0,0003
C13	*mol / wt %*	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001
C14	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C15	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C16	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C17	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C18	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C19	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C20+	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2S	*mol / wt %*	0,0005	0,0008	0,0005	0,0008	0,0005	0,0008	0,0005	0,0009	0,0005	0,0009	0,0005	0,0008
COS	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
PHENOL	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANOL	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
MEG	*mol / wt %*	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001
HE	*mol / wt %*	0,0211	0,0043	0,0209	0,0042	0,0208	0,0042	0,0200	0,0041	0,0200	0,0041	0,0205	0,0042
H2O	*mol / wt %*	0,0096	0,0088	0,0096	0,0087	0,0096	0,0087	0,0209	0,0193	0,0208	0,0192	0,0096	0,0089
Temperature	°C	1,6		1,6		1,6		11,7		11,7		1,6	
Pressure	*bar(a)*	67,0		67,0		67,0		67,0		67,0		67,0	
Flow	*Sm3/h*	48021		55352		55266		60555		60619		60957	
	MSm3/day	1,15		1,33		1,33		1,45		1,45		1,46	
	t/h	39,9		46,5		46,5		50,0		50,0		50,3	

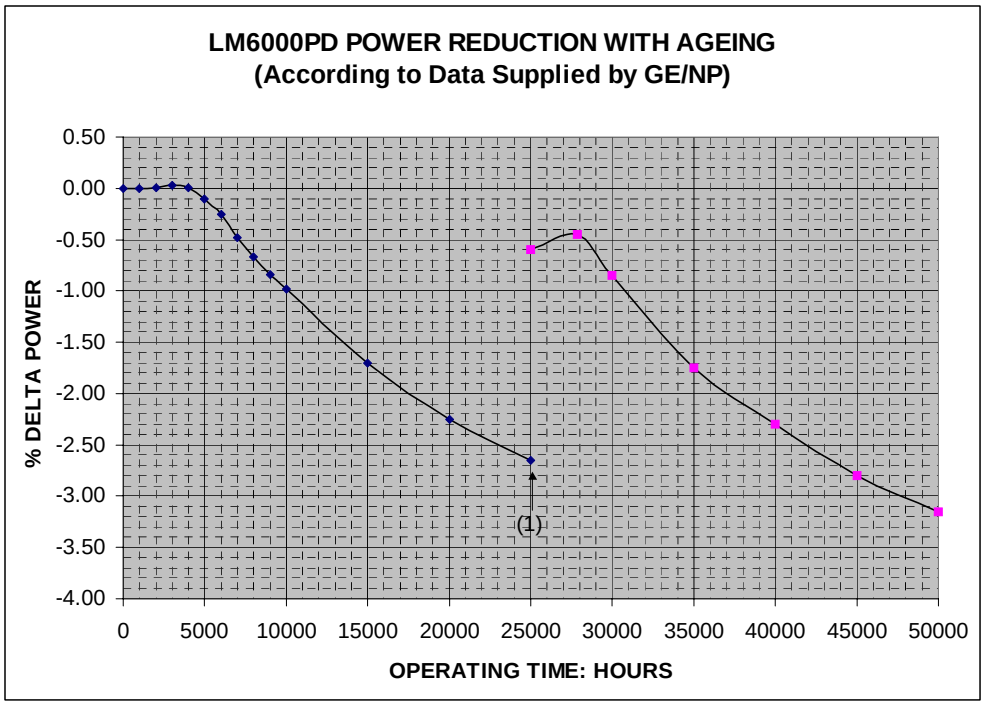
Stream	CO2 Export												
Feedstock No.		Snohvit A		1		2		3		4		5	
Year		Case 01		2007		2013		2015		2020		2022	
N2	*mol / wt %*	0,0017	0,0011	0,0018	0,0012	0,0018	0,0012	0,0007	0,0005	0,0007	0,0005	0,0013	0,0008
CO2	*mol / wt %*	99,6650	99,7259	99,7029	99,7578	99,6947	99,7445	99,7553	99,8058	99,7559	99,8058	99,7098	99,7621
METHANE	*mol / wt %*	0,1929	0,0704	0,1716	0,0626	0,1729	0,0631	0,1507	0,0546	0,1496	0,0546	0,1742	0,0636
ETHANE	*mol / wt %*	0,0318	0,0217	0,0285	0,0195	0,0287	0,0196	0,0193	0,0131	0,0191	0,0131	0,0236	0,0162
PROPANE	*mol / wt %*	0,0142	0,0142	0,0125	0,0125	0,0128	0,0128	0,0080	0,0079	0,0078	0,0079	0,0095	0,0095
I BUTANE	*mol / wt %*	0,0032	0,0042	0,0028	0,0037	0,0029	0,0038	0,0020	0,0027	0,0020	0,0027	0,0022	0,0029
N BUTANE	*mol / wt %*	0,0076	0,0101	0,0068	0,0090	0,0071	0,0093	0,0051	0,0066	0,0050	0,0066	0,0054	0,0071
I PENTANE	*mol / wt %*	0,0028	0,0046	0,0025	0,0041	0,0027	0,0044	0,0023	0,0037	0,0023	0,0037	0,0022	0,0037
N PENTANE	*mol / wt %*	0,0032	0,0052	0,0029	0,0048	0,0032	0,0052	0,0027	0,0044	0,0027	0,0044	0,0026	0,0043
C6	*mol / wt %*	0,0037	0,0072	0,0027	0,0052	0,0029	0,0056	0,0027	0,0053	0,0027	0,0053	0,0027	0,0052
C7	*mol / wt %*	0,0042	0,0089	0,0038	0,0080	0,0041	0,0086	0,0034	0,0073	0,0034	0,0073	0,0038	0,0080
C8	*mol / wt %*	0,0029	0,0072	0,0030	0,0074	0,0033	0,0080	0,0025	0,0061	0,0025	0,0061	0,0031	0,0076
C9	*mol / wt %*	0,0011	0,0030	0,0015	0,0042	0,0016	0,0046	0,0013	0,0038	0,0013	0,0038	0,0018	0,0051
BENZENE	*mol / wt %*	0,0363	0,0645	0,0310	0,0550	0,0337	0,0599	0,0211	0,0380	0,0214	0,0380	0,0280	0,0497
TOLUENE	*mol / wt %*	0,0166	0,0348	0,0145	0,0303	0,0160	0,0336	0,0114	0,0246	0,0117	0,0246	0,0167	0,0350
M XYLENE	*mol / wt %*	0,0033	0,0079	0,0028	0,0067	0,0031	0,0075	0,0033	0,0083	0,0035	0,0083	0,0046	0,0110
C10	*mol / wt %*	0,0006	0,0018	0,0004	0,0012	0,0005	0,0014	0,0003	0,0010	0,0003	0,0010	0,0005	0,0016
C11	*mol / wt %*	0,0001	0,0004	0,0002	0,0005	0,0002	0,0006	0,0001	0,0003	0,0001	0,0003	0,0002	0,0006
C12	*mol / wt %*	0,0001	0,0002	0,0001	0,0002	0,0001	0,0002	0,0000	0,0001	0,0000	0,0001	0,0001	0,0002
C13	*mol / wt %*	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000	0,0001
C14	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C15	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C16	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C17	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C18	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C19	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
C20+	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2S	*mol / wt %*	0,0088	0,0068	0,0078	0,0061	0,0078	0,0061	0,0077	0,0059	0,0077	0,0059	0,0077	0,0059
COS	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
PHENOL	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
METHANOL	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
MEG	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
HE	*mol / wt %*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	*mol / wt %*	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000	0,0001	0,0000
Temperature	*°C*	27,9		27,9		27,9		27,8		27,8		27,9	
Pressure	*bar(a)*	150,0		150,0		150,0		150,0		150,0		150,0	
Flow	*kmol/hr*	1824,2		2036,7		2035,0		2008,6		2023,6		2046,2	
	t/hr	80,23		89,59		89,52		88,35		89,01		90,01	
	t/day	1926		2150		2148		2120		2136		2160	
	m3/day	92,6		103,3		103,3		101,9		102,6		103,8	

Parameters affecting power consumption






LNG Plant Ageing
Power Demand increase, %
2.50
2.00
1.50
1.00
0.50
0.00
0
10
20
30
40
50
60
70
80
Months


LM6000PD POWER REDUCTION WITH AGEING
(According to Data Supplied by GE/NP)
% DELTA POWER
0.50
0.00
-0.50
-1.00
-1.50
-2.00
-2.50
-3.00
-3.50
-4.00
(1)
0
5000
10000
15000
20000
25000
30000
35000
40000
45000
50000
OPERATING TIME: HOURS

Reference data with number of stream days and dry feed flow rate

Year	Stream days	Dry feed flow rate MSm^3/sd	Dry feed flow rate GSm^3/yr
2007	123	20,78	2,55
2008	316	20,78	6,56
2009	341	20,78	7,08
2010	341	20,78	7,08
2011	320	20,78	6,64
2012	341	20,78	7,08
2013	341	20,78	7,08
2014	317	20,78	6,58
2015	340	20,70	7,05
2016	340	20,62	7,02
2017	319	20,62	6,59
2018	340	20,62	7,02
2019	340	20,62	7,02
2020	316	20,62	6,53
2021	340	20,66	7,03
2022	340	20,66	7,03
2023	319	20,66	6,60
2024	340	20,66	7,03
2025	340	20,66	7,03
2026	311	20,66	6,43
2027	335	20,66	6,93
2028	335	20,66	6,93
2029	314	20,66	6,49
2030	335	20,66	6,93
2031	335	20,66	6,93
2032	335	20,66	6,93
Average plateau	332	20,69	6,87

Appendix C

Formulas from ISO 6976 for calculating GHV

Calculation of GHV on volumetric basis for an ideal gas
The ideal gas calorific value on a molar basis as a function of t_1 for a mixture of a known composition metered at a temperature t_2 and pressure p_2 is calculated by the following equation.

$$\tilde{H}^{\circ}[t_1, V(t_2, p_2)] = \overline{H}^{\circ}(t_1) \times \frac{p_2}{R \cdot T_2} \qquad \text{(C.1)}$$

Where

$\tilde{H}^{\circ}[t_1, V(t_2, p_2)]$ is the ideal calorific value on a volumetric basis of the mixture (either superior or inferior) in $[kJ \cdot Sm^{-3}]$

$\overline{H}^{\circ}(t_1)$ is the ideal molar calorific value of the mixture (either superior or inferior)

R is the molar gas constant (= 8,314 510 $J \cdot mol^{-1} \cdot K^{-1}$)

T_2 (= t_2 + 273,15) is the absolute temperature in Kelvin

To calculate the GHV of the mixture of an ideal gas on molar basis, the following formula can be used.

$$\overline{H}^{\circ}(t_1) = \sum_{j=1}^{N} x_j \cdot \overline{H}_j^{\circ}(t_1) \qquad \text{(C.2)}$$

Where

$\overline{H}^{\circ}(t_1)$ is the ideal molar calorific value of the mixture (either superior or inferior) in $[kJ \cdot mol^{-1}]$

$\overline{H}_j^{\circ}(t_1)$ is the ideal molar calorific value of component j (either superior or inferior) in $[kJ \cdot mol^{-1}]$

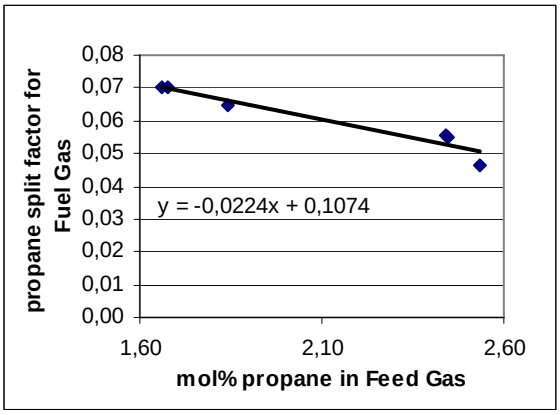
x_j is the mole fraction of component j

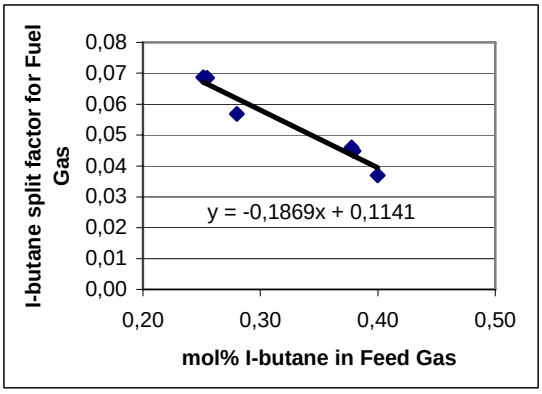
N is the number of components in the mixture

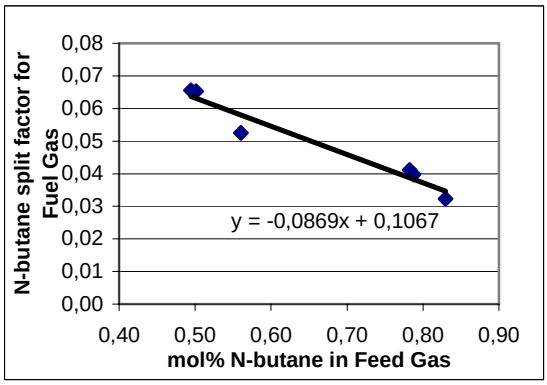
Numerical values of $\overline{H}_j^{\circ}$ for t_1= 25 °C are given in table 3 in ISO 6976.

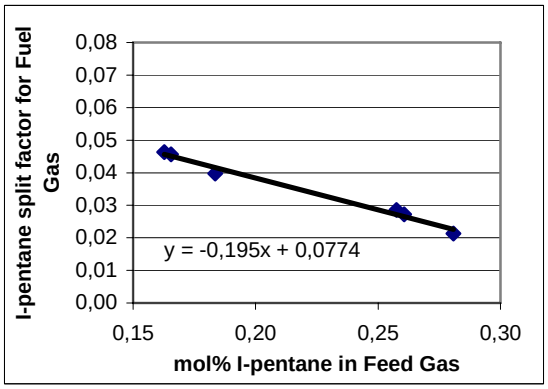
Appendix D

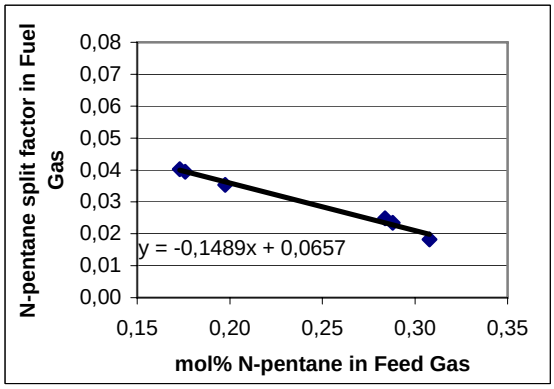
Linear split factors for Fuel Gas in Model 2



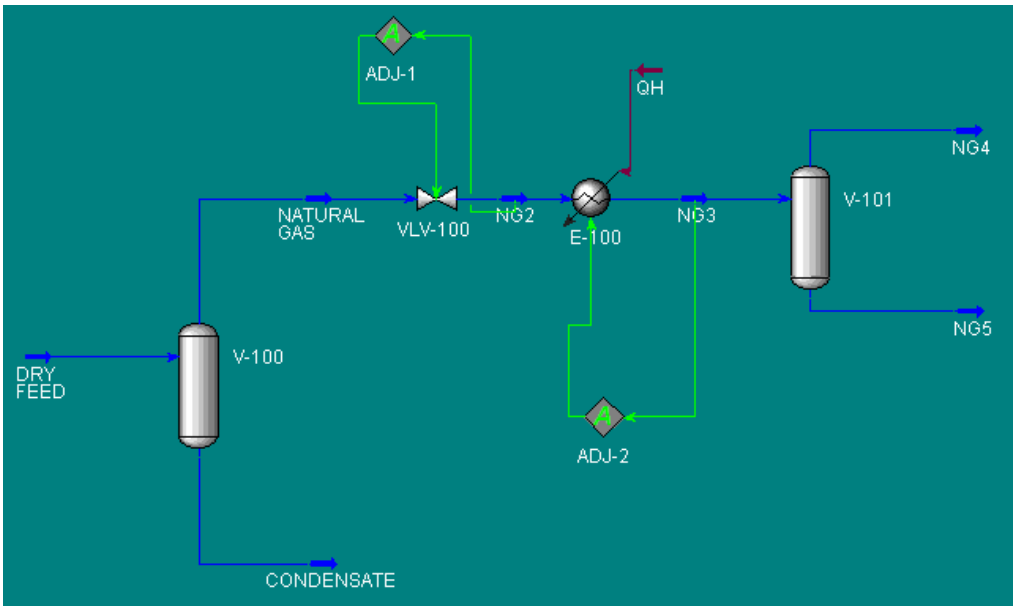




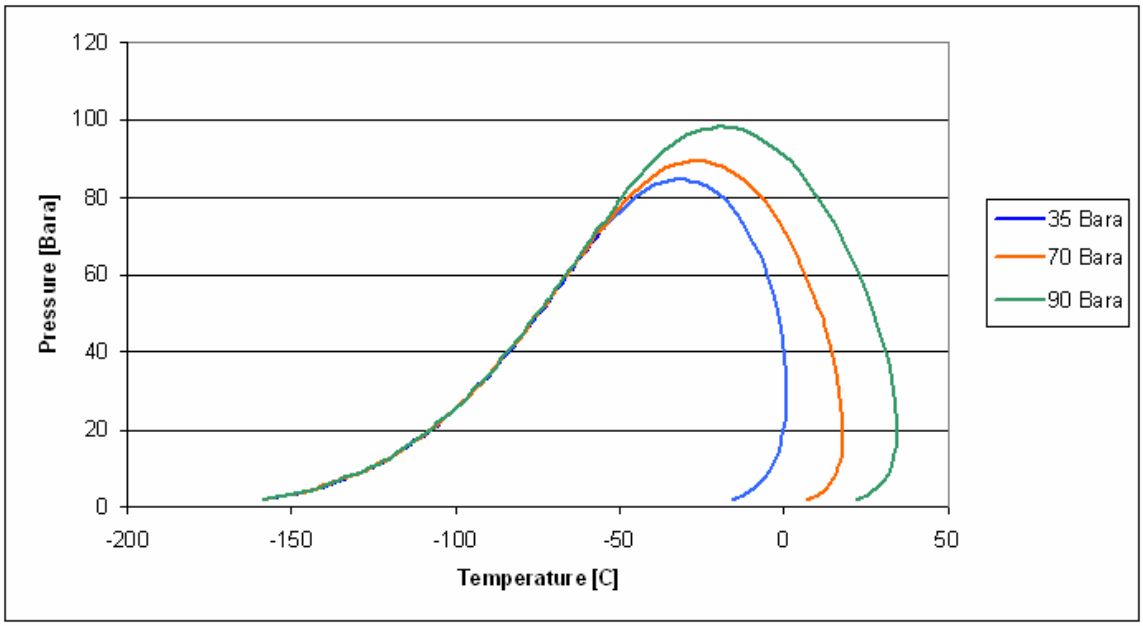




HYSYS model for Fuel Gas calculation at 90 Bara



Phase envelope for Feed Gas in Slug Catcher at 35, 70 and 90 Bara



Appendix E

User Manual for Mass Balance Model (Model 3)

How to use the model:

The model starts in the Excel sheet called “Input data”. To perform a production forecast it is necessary to get hold of the data described below and put them into the cells where they belong.

Necessary input data:

- Flow rate of Feed Gas (MEG/Water free)	kmol/hr
- Feed Gas composition on molar basis	mol %
- Thermal power consumption (Year average)	MW
- Distribution coefficients (K-values form HYSYS)	-
- Yearly number of stream days	sd/yr

The default values for the thermal power consumption are based on the Fuel Gas flow rate from the Linde Base cases and may be used if other data is not available.

The K-values for the Feed Gas in the Slug Catcher must be calculated by another program. HYSYS may be used for this purpose. It is only necessary to make a stream line and define the composition, flow rate, temperature and pressure for the Feed Gas based on the conditions in the Slug Catcher (1° C and 70 Bara can be used as default values). The K-values must be calculated for each year and put into the respective input data cells in the Excel model. If any of the components in the Feed Gas are equal to *zero*, it is necessary to assign a very small value to this component in order to obtain a K-value.

When the data is put into their respective cells, it is only necessary to press the “***Forecast***” button and a macro function will calculate the production for all the years.

How the model works

The following description shows how the model works sheet by sheet.

Macro function

The macro function is used to loop the calculation of the products each year in order to calculate the whole production period. Comments are added to the macro script in the Excel file in order to make it easier to understand.

Fuel calculation

The composition and flow rate for the Fuel Gas is calculated in this sheet. The macro function first collects data for the K-values, the Feed Gas composition and power consumption for the given year from *Input data.* The x and y value are calculated by Equation 7.12 and 7.13. The macro function then performs an iteration to satisfy the criteria in Equation 7.14 by changing the cell for vapour flow. This will determine the

composition of the Fuel Gas. The flow rate is then calculated from Equation 7.16. The GHV of the fuel gas is calculated from ISO 6976 from the formulas in Appendix C.

Calculate
This sheet contains the split matrix that distributes the components in the feed gas to the different products. All the cells coloured in green are constants based on split factors from Case 01. The other factors are described in Chapter 7.4 and Chapter 7.5. The if - statements used to determine split factors for nitrogen and I-Butane for LNG are based on the methane content in the three main periods of production (see Model 1) and the corresponding mean split factors for nitrogen have been used. The same principle has been used for the I-Butane. The GHV iteration to determine the content of heavier hydrocarbons in the LNG is done by the macro function. The right GHV is obtained by adjusting the propane content in the LNG.

Print out
In this sheet the temporary products are calculated by the split factors from the sheet *Calculate.*

De-propaniser
This sheet performs the correction of the fraction of C4 and C5 in the LPG and Condensate. The macro function takes the flow rates from *print out* and past them into this sheet. The operation of this column are depending on whether the feedstock is light of heavy. This is determined by the methane content in the Feed Gas (The if-statement in Cell F:31 which states the wt% C5+ in the LPG by the methane fraction in the Feed Gas and the typical methane content for the three main periods of production). The macro function is changing the iteration factors for butanes and pentanes to adjust the amount of these components in the LPG and the Condensate to satisfy the criteria.

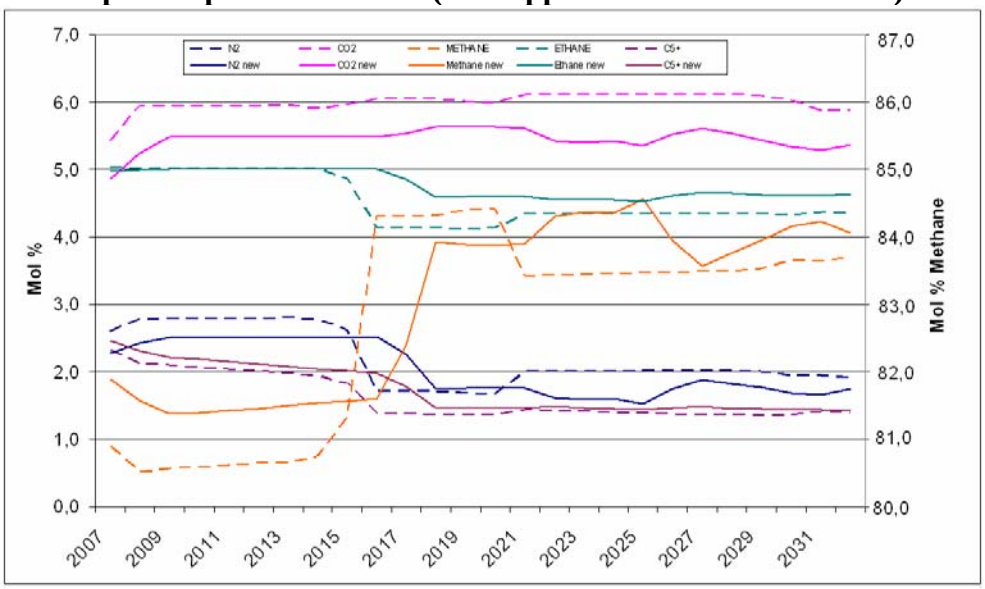
Forecast kmol/hr
For each year the macro function takes the flow rate of LNG, Fuel Gas, CO_2 and N_2 on kmol/hr basis from *Print out* and presents them in tables in *Forecast kmol/hr*. The LPG and Condensate are taken from *De-propaniser*. The yearly production in T/yr is also calculated in this sheet for each product.

Forecast composition
The corresponding composition for each product each year are calculated from *Forecast kmol_hr* and presented the similar way in tables in *Forecast composition.*

Graphs
In this sheet the forecasted production of LNG, LPG and Condensate on mass basis are presented in graphs and tables to see how the production of the different products are changing over time.

Appendix F

Updated production data (the stippled lines are the old data)



Forecasted production based on updated field production data for Model 3

Year	LNG [MT/yr]	LPG [T/yr]	Condensate [T/yr]
2007	1,6	124 937	234 562
2008	4,0	306 250	565 830
2009	4,3	321 560	578 686
2010	4,3	321 580	570 935
2011	4,0	301 634	528 265
2012	4,3	321 455	553 796
2013	4,3	321 041	541 646
2014	4,0	296 614	493 629
2015	4,2	315 168	519 491
2016	4,2	313 662	506 625
2017	4,0	262 103	428 217
2018	4,3	206 907	374 195
2019	4,3	208 790	375 032
2020	4,0	195 153	348 934
2021	4,3	211 138	376 020
2022	4,3	204 189	375 615
2023	4,0	191 177	350 552
2024	4,3	203 117	369 765
2025	4,3	201 662	367 565
2026	3,9	200 220	340 683
2027	4,2	223 709	368 146
2028	4,2	223 414	363 970
2029	4,0	209 423	339 813
2030	4,2	217 071	360 359
2031	4,2	218 980	357 857
2032	4,0	208 067	331 082
2033	4,2	227 018	355 867
2034	4,2	227 999	354 231
2035	4,0	210 438	327 830
2036	4,3	214 765	336 655
2037	4,3	221 178	341 688
2038	4,0	202 227	312 639
2039	4,3	208 351	325 647