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Regulering av brennverdi i LNG anlegg

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Preface

This thesis has been written during my last semester at the Norwegian University of Science and Technology (NTNU) and concludes five years of education in the study programme Energi og Miljø.

I would like to thank my supervisor Geir Owren and Knut Håvard Nordstad at Statoil for help and guidance through writing this thesis. I would also like to thank friends and family for their support and motivation while writing this thesis and during my years at NTNU.

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Abstract

With today's ever-increasing energy consumption and increasing demand for more environmentally friendly energy sources, natural gas has become one of the most important energy options for the future. There are large gas resources along the Norway Continental Shelf and further discoveries are expected. Norway will therefore continue to play a significant role in the future gas market. An important factor to remain a provider in the gas market, is maintaining efficient gas extraction, as well as getting a product that meets specifications according to end user.

Natural gas specifications such as GHV have several purposes. These specifications are related to gas sales contracts and prevails unproblematic transport and storage of gas. The European market can accept LNG with a GHV at 41 MJ/sm^3, while the US market GHV specifications cannot exceed 40 MJ/sm^3. But due to shale gas production, the US market is off less interest, and it is possible increasing GHV on LNG to 41 MJ/sm^3. This adjustment can make it possible to include associated gas to an already existing LNG treatment facility.

To get a view of GHV control, a description of methods used to control GHV is presented in this thesis. There are several methods which can adjust the GHV on LNG, and the most efficient way is to vary the amount of LPG in the gas flow. This can be done by integrating a turbo-expander upstream, but since this installation is volatile for flow changes and requires extra installation, the best way is to do modifications to the already existing scrub column prior to the fractionation system. Reboiler and condenser temperature, amount of trays, feed flow inlets and reflux flows in the scrub column are all modification which effect separation efficiency and consequently the heating value.

A presentation of GHV control is described by a simulation of the pretreatment facilities, where the process is designed with a maximum GHV 40 MJ/sm^3. This simulation represents existing facilities and extracts heavy hydrocarbons prior to the fractionation systems both by a condensate stabilizer and by a scrub column. Propane recovery on the bottom product is about 60% and combined with butane the total LPG recovery in the bottom product is about 70%. This gives an LNG product at about 17.1 million sm^3/day

When increasing max GHV specifications up to 41 MJ/sm^3, the process can be simplified. This is done by removing the recycled flow from the fractionation system, C_4/C_5 reflux flow respectively. A presentation of simulation results show how propane recovery in the bottom product is reduced to 40%. LPG follows top flows instead and consequently increases heating value on the LNG product. No other modifications are done, so heating value becomes 40.45 MJ/sm^3. This heating value is below new limit, but over original GHV specifications. This process simplification gives a LNG product 17.4 million sm^3/day, where all other gas specifications are reached.

To assess whether it is possible to include associated gas to an existing LNG treatment plant, five new cases are simulated. The LPG amount in feed flow is increased from added 20% to added 200% of original feed flow. In these simulations the temperature in the condenser is increased, so heating value on LNG product reaches 41 MJ/sm^3. These cases hardly change the LNG product amount, since most of added LPG follows the bottom product.

The scrub column performance decreases when LPG added exceed 80%, due to higher risk for flooding. Energy use also increase for added LPG, which limits possibilities for adding LPG. In addition, when increasing LPG, other scrub column adjustments must be performed to achieve full convergence. By increasing methane following the bottom product, the column handles higher amounts of LPG in feed flow. However, this can challenge the fractionation unit and there might be necessary to install a demethanizer to extract added methane.

When considering energy use and mechanical limitations in the scrub column, a feed flow flexibility assessment shows that added LPG should not exceed 60% of original LPG in feed flow. When exceeding this amount, gains decline. Additional LPG can increase up to 68 002 kg/h in feed flow, which results in an LNG product at 17.3 million sm^3/day.

The results are based on several assumptions, which make it difficult to get fully conclusive results. However, despite uncertainties, all results indicate that GHV adjustments provide an opportunity to include associated gas in the future.

Sammendrag

Med stadig økende energiforbruk og økende krav til mer miljøvennlige energikilder vil naturgass bli en av de viktigste energialternativene for fremtiden. Det er funnet store gassressurser langs kysten av Norge og det er forventet ytterligere funn. Norge vil fortsette å spille en betydelig rolle i det fremtidige gassmarkedet. En viktig faktor for å forbli en aktør på gassmarkedet baserer seg på å ha kontinuerlig effektiv gassutvinning, samt få et produkt som innfrir spesifikasjoner som settes av sluttbruker.

Det amerikanske gassmarkedet tillater et LNG produkt med en brennverdi (GHV) på 40 MJ/sm^3, mens det europeiske markedet kan tillate gass med høyere brennverdi. Siden USA produser skifergass, er det er mindre aktuelt med salg til det amerikanske markedet. Det har derfor blitt mer aktuelt å rette salg mot det asiatiske og europeiske markedet. Ved å ekskludere det amerikanske markedet kan gass produseres med høyere brennverdi. Dette åpner for muligheter til å inkludere assosiert gass til et allerede eksisterende LNG behandlingsanlegg.

For å få en oversikt over hvordan brennverdispesifikasjoner påvirker et LNG prosessanlegg, er det i denne masteroppgaven beskrevet et utvalg av metoder for å regulere brennverdien til naturgassen. Den mest effektive måten å variere brennverdien er ved å variere mengden av LPG i gasstrømmen. Ved å skille ut LPG både oppstrøms, under og nedstrøms for nedkjølingsprosessen, endres brennverdien. Dette kan gjøres ved å integrere en turboekspander oppstrøms, men siden denne installasjonen er sårbar for gasstrømendringer og krever ekstra installasjon, er den et uaktuelt alternativ. Den beste måten å regulere brennverdien er ved å gjøre endringer på skrubbkolonnen før fraksjoneringssystemet. Både koker- og kondensertemperatur, antall trinn, inngangstrinn og refluks i en skrubbekolonne påvirker separasjonseffektiviteten og påvirker brennverdien på sluttproduktet.

En presentasjon av GHV kontroll er beskrevet ved å designe en simuleringsmodell av et LNG forbehandlingsanlegg, der prosessen er utformet med en maksimal GHV 40 MJ/Sm^3. Denne simuleringen representerer eksisterende anlegg og viser at tunge hydrokarboner skilles ut før fraksjoneringssystemet både ved hjelp av en kondensat stabilisator og etter en skrubb kolonne. Propanutvinningen i bunnproduktet er omtrent 60% og i kombinasjon med butan er den totale LPG-utvinningen i bunnprodukt omtrent 70%. Dette gir et LNG- produkt på ca. 17,1 millioner sm^3/dag.

Ved å øke brennverdien opp til 41 MJ/Sm^3, kan prosessen forenkles. Dette gjøres ved å fjerne den resirkulerte strømmen fra fraksjoneringssystemet, henholdsvis en refluksstrøm bestående av hovedsakelig C_4 og C_5. Resultater fra simuleringen viser hvordan propanutvinningen i bunnproduktet blir redusert til 40%. LPG følger toppstrømmer i stedet, noe som øker

brennverdien på LNG produktet. Ingen andre endringer ble gjort i simuleringene, noe som resulterte i en brennverdi på 40.45 MJ/Sm^3 . Denne brennverdien er under ny grenseverdi, men over opprinnelige spesifikasjoner . Denne prosessforenklingen gir et LNG produkt på 17.4 millioner sm^3/dag hvor gasspesifikasjoner er nådd .

For å vurdere om det er mulig å inkludere assosiert gass til en eksisterende LNG behandlingsanlegget, er fem nye tilfeller simulert. LPG-mengde i inngangsstrøm er økt med en tilsats på 20 % til en tilsats på 200 % av den opprinnelige inngangsstrømmen. I disse simuleringene er temperatur i kondensatoren økt, slik at brennverdien på LNG- produktet når 41 MJ/Sm^3. Resultatene viser at en økning av LPG knapt endrer LNGproduktet, da mesteparten av økt LPG følger bunnproduktet.

Ytelsen til skrubbkolonnen reduseres når ekstra LPG i inngangsstrøm overgår 80 % på grunn av høyere risiko for oversvømmelse i kolonnen. I tillegg øker energibruken ved økt LPG, noe som begrenser mulighetene for å legge til LPG. Med økende LPG, må ytterligere justeringer gjøres på skrubbkolonnen. For å oppnå konvergens i skrubbkolonnen må andelen metan som følger bunnproduktet oppjusteres. Dette kan utfordre fraksjoneringssystemet ytterligere, og det kan være nødvendig å installere en demetaniserer for å utskille ekstra metan.

Basert på en vurdering av energibruk og mekaniske begrensninger i skrubbkolonnen er det antatt at en LPG økning ikke bør overstige 60% av opprinnelig LPG i inngangsstrømmen. Dette resulterer i en økt LPG-mengde på 68 002 kg/h og et LNG-produkt på 17.3 millioner sm^3/dag. Disse resultatene er basert på en del forutsetninger, som gjør at resultatene blir noe usikre. Men til tross usikkerhetsmomenter, indikerer samtlige resultater at ved å justere brennverdien på LNG-produktet, er det muligheter for å inkludere assosiert gass i fremtiden.

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Abbreviations

AGRU	Acid Gas Remover Unit
EPA	United States Environmental Protection Agency
GDP	Gross Domestic Product
GHV	Gross Heating Value
HHC	Heavy Hydro Carbons
LHV	Lower Heating Value
LMTD	Log Mean Temperature Difference
LNG	Liquefied Natural Gas
LPG	Liquefied Petrol Gas
MDEA	Methyl Diethanol Amin
MEG	Mono Ethylene Glycol
NGL	Natural Gas Liquids
NPD	Norwegian Petroleum Directorate
NTU	Number of Transfer Units
o.e	Oil Equivalents
PDO	Plan Development and Operation
PFD	Process Flow Diagram
S	Standard Conditions (p=100kpa, T=275,15K)
TVP	True Vapour Pressure
UHV	Upper Heating Value
VLE	Vapour and Liquid Equilibrium
WI	Wobbe Index

1. Introduction

1.1. Norwegian Petroleum Industry

The petroleum industry is currently the largest industry in Norway. In the last 40 years the petroleum production on the Norwegian Continental Shelf has added more than 900 billion NOK to the national gross domestic product (GPD). 23% of the value creation came from the petroleum sector in 2012. As shown on figure 1, the main production of petroleum comes from oil production, which represents the green columns on the figure. However, for the last ten years, gas production and export has become a much more noticeable player in the international energy marked. Norway is regarded as the world's third largest gas exporter, and produced in 2012 about 115.1 million Sm^3 of which 113.7 million Sm^3 were exported (Norwegian Petroleum Directorate, 2013).

Figure 1 also predicts that gas production will remain unchanged until 2016, with a gas production over 100 million sm^3 o.e/year. Natural gas accounts for more than 20% of the world`s total energy demand (NPD, 2013), and as a consequence natural gas production will remain important for the Norwegian Petroleum Industry.

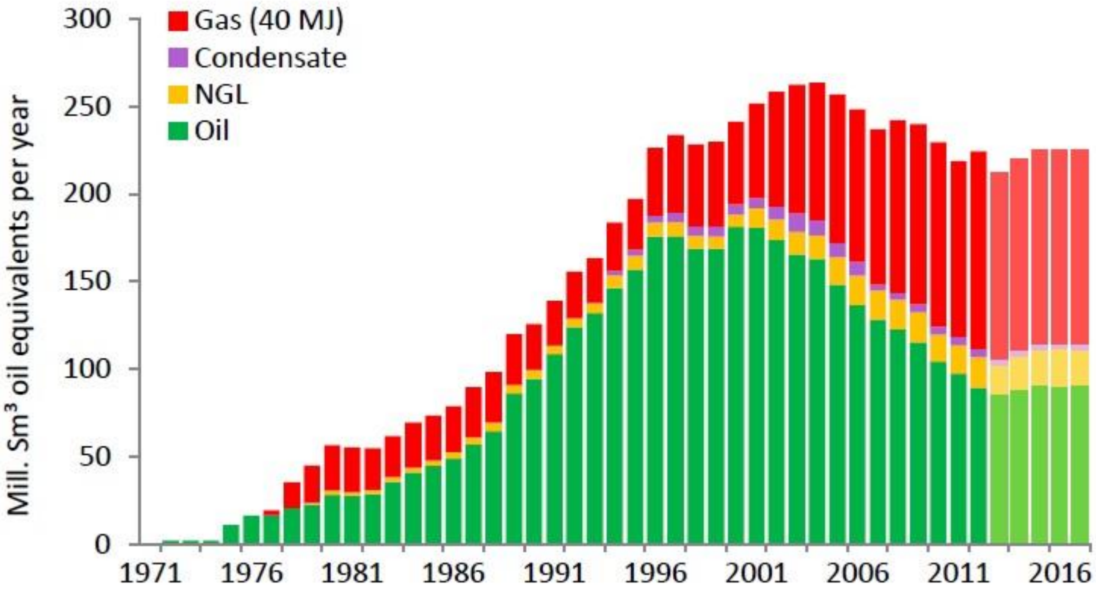

Figure 1 Historical and expected production of oil and gas in Norway (NPD, 2013)

Another important aspect showing natural gas forecast is illustrated in figure 2. This figure depicts how gas production will remain important in both near and distant future. The figure shows the Norwegian Petroleum Directorate (NPD) basic estimate of resources[1] and reserves[2] on the Norwegian Continental Shelf, respectively for gas (red area), liquid (green area) and combined (blue area). The figure illustrates remaining reserves and resources, where total resources estimated 13.6 (6.0+7.6) billion sm^3 oil equivalents (o.e) in 2012. Sold and delivered products counted 44% of this amount. The remaining amount, 7.6 billion sm^3 o.e, consists of both resources and reserves, which have not yet been technically exploitable or planned. This suggests that production will continue for many years to come.

A more detailed presentation of figure 2 with associated table can be found in attachment I.

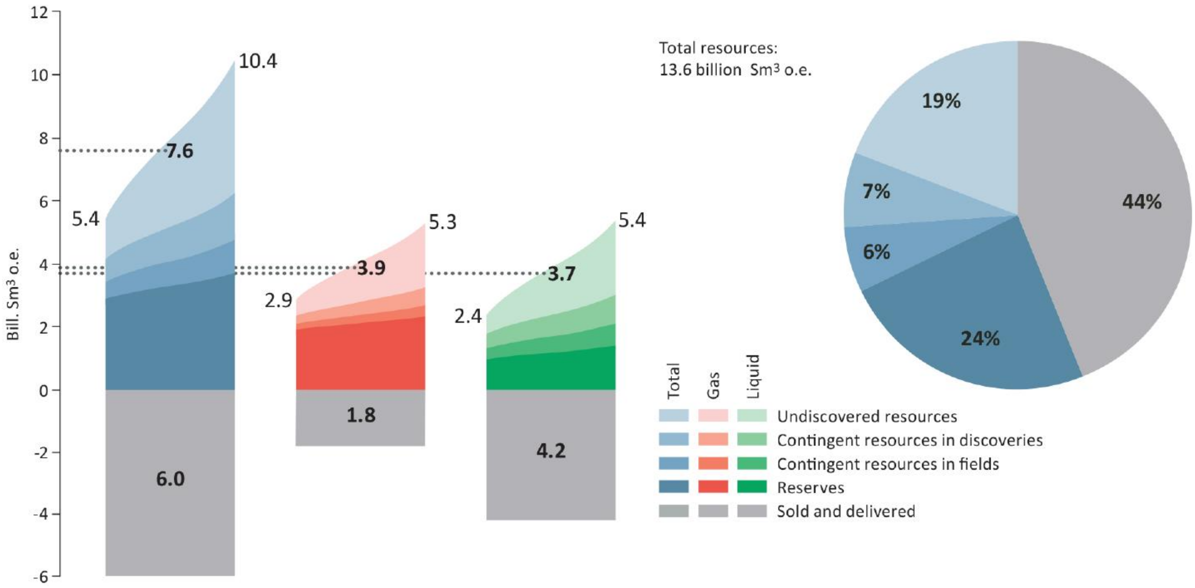

Figure 2 Norwegian Petrol Resources and Reserve Estimates in 2012 (NPD, 2013)

The Norwegian Petroleum Industry will in the future be important at a global and local level due to expected increased energy consumption, increased need for more environmentally friendly energy sources and large expected findings.

However, an important criteria for Norway to remain as a major player in the international gas market, is to have a continuously effective production, while meeting sale specifications. One of the key assignments is therefore to always consider improvements and new solutions for petroleum production.

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[1] Generic term for all quantities of petroleum which are technically exploitable

[2] Consists of remaining stored petroleum resources, government granted with a license to plan, develop and operate (PDO), and in production.

1.2. Objective

The objective of this thesis is to assess whether an increased max Calorific Heating Value (GHV) specification can be used to allow a more Liquefied Petroleum Gas (LPG) rich feed to an LNG (Liquefied Natural Gas) Plant.

1.3. Scope

A possible advancement in gas production is to recover gas not only from pure gas fields, but also include associated gas from tie-ins fields, where oil is produced together with gas. However, this intervention increases amount of heavy hydrocarbons in feed gas and consequently increase heating value.

Objectives of this master is to look at the gas production in an existing LNG plant and to assess whether an increased max GHV specification can allow more LPG to an LNG Plant without creating problems for the LPG and the fractionation systems in the plant.

The thesis includes the following considerations:

1. A description of how GHV is controlled in a typical LNG plant and the methods used to allow for a max GHV of 40 MJ/sm^3. The process shall be simulated with either PROII or HYSYS.
2. A description of how the process can be simplified if the GHV max specification is increased to 41 MJ/sm^3. The process shall be simulated with the increased GHV max specification.
3. An assessment of how the increased max GHV specification can be used to improve the feed flow flexibility.

In the first part of this thesis an overview of background of an LNG process and a description of methods used to control GHV is presented. Furthermore, simulation results are presented. Both results from a typical LNG plant and in a plant where simplifications are done, are presented. The last part presents an assessment on feed flow flexibility, by introducing various feed gases to the simplified LNG plant.

This will shed light on the possibility to achieve greater gains by including the associated gas in gas production in the future.

2. Background

Production of LNG has several advantages. By condensing natural gas, the volume decreases which enables transportation in tanks over long distances. However, this condensation process is highly energy demanding and requires several important facilities installments. To get an understanding of the LNG process, this chapter presents a background for LNG treatment, where the facilities are described from well to market.

2.1. LNG - From Well to Market

LNG processing is a demanding process and to achieve desired end product with the right specifications, the natural gas must go through several processing steps. This process can be seen on the block diagram in figure 3 and will be further presented in this chapter.

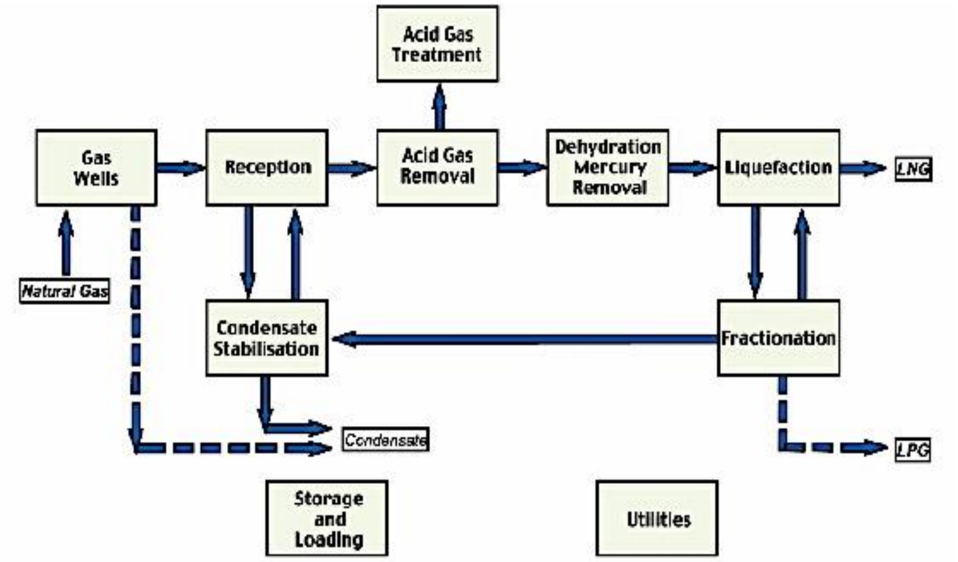

Figure 3 Principal block flow diagram for an LNG plant (Fredheim et al., 2012)

2.1.1. Reception

After natural gas leaves the gas wells, the gas as a multiphase flow, enters reception facilities. Unprocessed multiphase flow often leads to formation of slug flow, which needs to be removed. This is done by treating the well stream in a slug catcher. The slug catcher is a separation unit which minimizes the slug from the gas and separates the multiphase flow into gas, condensate and water/Mono Ethylene Glycol (MEG). The latter is recycled back to the slug catcher, while condensate and gas will be further processed.

2.1.2. Condensate Stabilization

Condensate from the slug catcher is sent to a stabilizing column, which distills and transmits light hydrocarbons back to the feed gas, while stable condensate is separated out. This stabilization is important to reduce gas pressure. In order to avoid phase change when the condensate is exported to atmospheric tanks the gas pressure is reduced.

2.1.3. Acid Gas Removal

After reception facilities the overhead gas contains sour gases such as CO_2 and H_2S, which needs to be removed. Acid gas is typically removed in an amine unit using the principle that alkaline solvent will remove acid gas components. (Acid Gas Removal, 2011) In LNG production this is often done by a wet absorption process, where amine solvents counter – currently contacts the source gas inside an absorber column. Chemical solvents such as Methyl Diethanol Amin (MDEA) will then absorb the acid gas substances, before they are boiled off in a stripper column. The amine solvent is recycled and the effluent gas in the absorber column, now sweetened, can continue to the next processing step.

2.1.4. Dehydration and Mercury removal

Since sweet gas after acid gas removal contains water, the gas needs to be dehydrated. This is done by cooling, where the sweet gas is cooled down and water is condensed out. Non-condensed gas goes through an adsorption process, with use of molecular sieves. The adsorption process is a regenerative process, where temperature and pressure can be used as parameters for the regenerator. First, gas enters a drying tower containing desiccant. A mass transfer between adsorbent and sweet gas takes place, and dry gas leaves the tower bottom. Meanwhile, adsorbent rich gases are heated and send to a regeneration tower. The gas is cooled down and thereby condenses out adsorbed material. A similar process is used for mercury removal. The gas can then, assuming it meets specifications, return to the gas stream. (Campbell, 2001)

2.1.5. HHC Extraction

To achieve a desired end product consisting of methane, heavier hydrocarbons (HHC) must be extracted from the gas stream. This process, NGL –extraction and fractionation, can be done upstream or integrated in the liquefaction, and separates the natural gas into various components, such as LPG and C_1, C_2 and C_3. It can done by various methods and will be described in the following chapters.

2.1.6. Liquefaction

Liquefaction takes place with pressurized feed gas, typically at a level between 40 and 80 bars and occurs at a gliding temperature, where the natural gas goes through stages of precooling, liquefaction and sub cooling. By using a refrigerant medium the threated gas is sub cooled to approximately -155°C. After reaching this temperature, the now pressurized liquid undergoes near-isentropic expansion slightly above boiling point. At this stage, the flow is in two-phase.

2.1.7. End Flash

The pressure reduction in the liquefaction gives flashing. This flash gas generally consists of a nitrogen/methane mixture. Since nitrogen is more volatile than methane, the nitrogen can be separated in an end flash separator. By doing this one achieves an end product consisting of pure methane in a liquid state. LNG is now ready for storage and transportation. The end flash on the other hand, can be used as fuel gas. (Fredheim et al., 2012)

The end plash is important in relation to composition of the LNG product and is illustrated in figure 4. The x- axis show pressure, the y- axis show enthalpy and red lines depicts temperature in the range from 50 °C to -200 °C. The curves on the figure represent dew point lines. Cooling the natural gas occurs when following the blue line horizontally through isobaric precooling, liquefaction and subcooling. LNG is ordinary cooled down to about -163 °C.

Higher temperature after sub cooling increases components in the end flash gas. This is illustrated by vertical blue line on the graph. By increasing sub cooling temperature, moving blue line further to the right, more nitrogen and methane end up in end flash gas instead of the LNG product. This increases heating value.

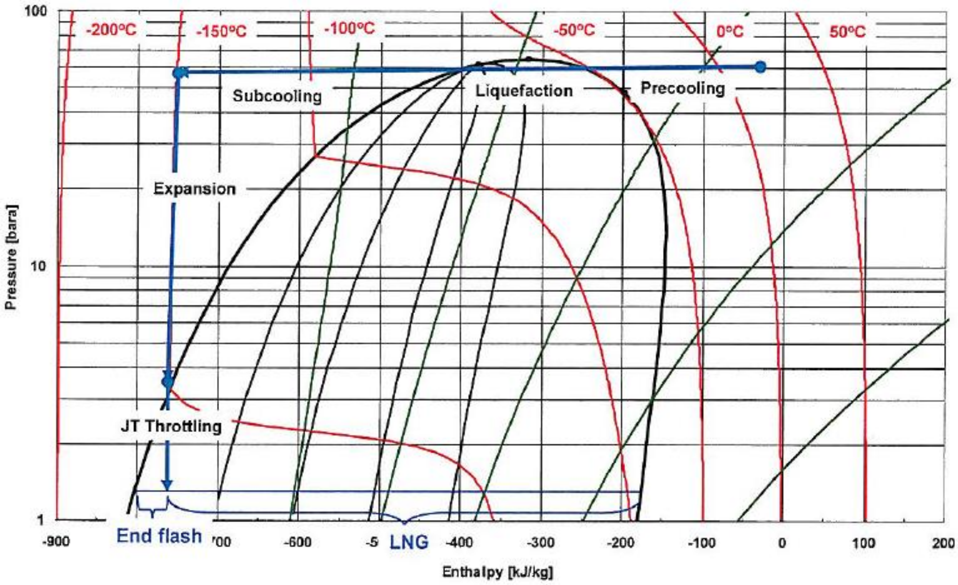

Figure 4 Pressure- enthalpy diagram for Natural Gas through precooling, liquefaction, sub cooling and end flash (ibid.)

3. GVH Control

As mentioned previously several processing steps are required to achieve wanted end product. To measure quality and to control the end product, the main parameters are heating value and Wobbe Index (WI).

The WI is a density-corrected heat value that reflects the behavior of the fuel during combustion. The value is indicates relationships between gas composition and usability and is defined as follows. (Fredheim et al., 2012)

$$\mathrm{WI} = \frac{\mathrm{GHV}}{\sqrt{\mathrm{spgr}}} = \frac{\mathrm{GHV}}{\sqrt{\frac{\mathrm{MW}}{28.964}}}$$

Where

GHV: Calorific Heating Value or Gross Heating Value [MJ/Sm^3]
Spgr = specific gravitation [-]
MW = Molar weight [kg/kmole]

The higher a gases' Wobbe number is, the greater the heating value of quantity of gas that will flow through a hole of given size in a given amount of time is. (*The Wobbe Index*, 2007)

The heating value or GHV is a measure of heating power and is dependent upon the composition of the gas. It refers to the amount of heat released when a known volume of fuel is completely combusted under specified conditions. It can be expressed as an upper value, Upper Heating Value (UHV or GHV), or a lower value, Lower Heating Value (LHV). In this thesis all values are given in upper value and expressed as both heating value and the abbreviation GHV.

As shown in table 1, the heating value varies for different hydrocarbon components. The heating value increases with the amount of heavy hydrocarbons in the natural gas. (Coyle et al., 2007) A more detailed view on heating value control will be described in the following chapter, by presenting motivation, methods and design.

Table 1 Heating Values of Components in Natural Gas (Fredheim et al.,2012)

Substance	**GHV [kJ/kg]**	**GHV [MJ/Sm^3]**	**LHV [kJ/kg]**	**LHV [MJ/Sm^3]**
Nitrogen	0	0	0	0
Methane	55 496	37.66	50 010	33.93
Ethane	51 875	65.97	47 484	60.39
Propane	50 345	93.90	46 353	86.45
Butane	49 500	121.69	45 714	112.38
Pentane	49 011	149.56	45 351	138.39

3.1. Motivation

Natural gas specifications such as GHV have several purposes. These specifications are related to gas sales contracts and prevails unproblematic transport and storage of gas.

One reason for GVH control is to avoid liquid dropout in pipelines. Propane and heavier hydrocarbons have a higher freezing point than LNG, and if the percentage is large enough and is combined with CO_2 and water, it may freeze out in the cryogenic heat exchangers before liquefaction. This can be prevented by removing heavier hydrocarbon components. Therefore natural gas pipeline companies limit the amount of butane, pentane and heavier components. (Fredheim et al., 2012)

Another advantage for this type of heavy hydrocarbon extraction is increased value of the natural gas. The possibility for increased economic incomes is higher when a variety of end products, such as LPG and condensate, are produced. Products such as propane and butane can in addition be used as refrigerant internally in the plant, which increase the power efficiency.

A third reason for considering the heating value in the end product is due to boil-off of LNG during transportation. When LNG at -163° C are transported in tanks, the fluid will start boiling (primarily of nitrogen), which will increase heating value. This must be encountered for, when setting restriction on maximum GHV of the LNG at the receiving terminal. (ibid.)

Other reasons for GHV control is to adapt to the various markets. Different countries demand different GHV on their product, and since GHV increases with heavy hydrocarbon concentration, the composition in the gas must adjust to end user.

Figure 5 shows representative heating value specifications for different countries, where the x-axis represents en extract of end consumers and the y-axis embodies heating values in the range from 34 MJ/sm^3 to 46 MJ/sm^3. As shown the European market, with the UK, France and Spain respectively, allows relatively high GHV specifications on the natural gas, with values in the range from 37 MJ/sm^3 to 45 MJ/sm^3. The US market, on the other hand limit the natural gas end product to consist of heating value around 40 MJ/sm^3. To reach both of these markets, the GHV should intersect between these ranges.

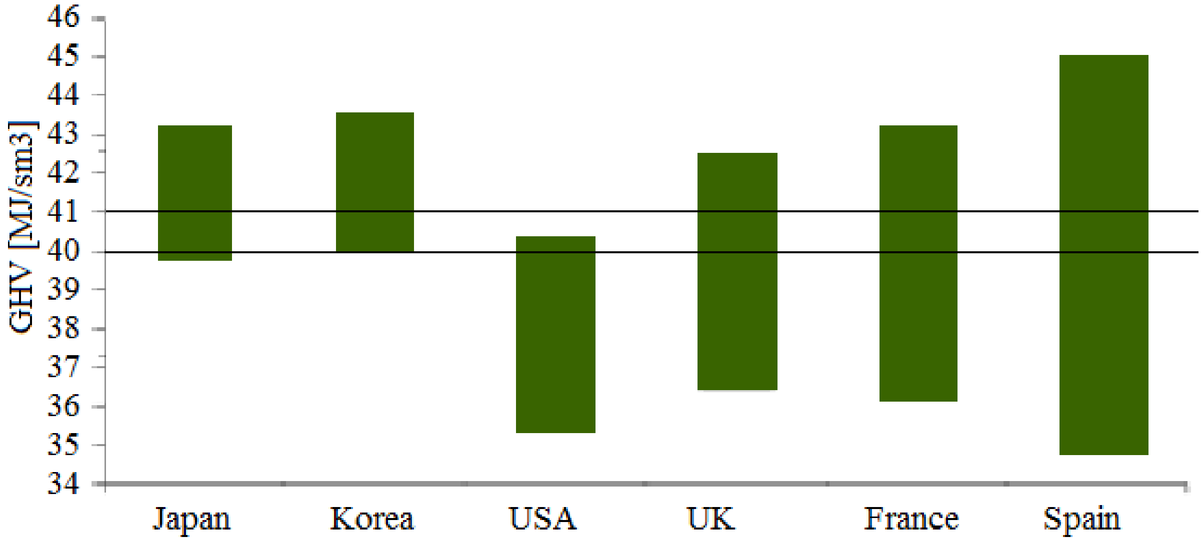

Figure 5 Worldwide Heating Value Specifications (Coyle et al., 2007)

To get a wide range of customers, the lower horizontal line on the figure includes all markets. This can be achieved by selling an end product with a heating value of 40 MJ/sm^3. This will include sales to the US market as well as to the asian market. However, due to shale gas production in USA, selling to the US market is of less interest. By excluding the US market, it might be possible to increase heating value specifications for the LNG plants. This is presented as the upper horizontal line on the figure, with a GHV on 41 MJ/sm^3 and outside of US market range.

There are several methods to control GHV. An extract of these methods will be presented as followed. This is firstly done by presenting use of nitrogen, and then followed by presenting GHV control by LPG Recovery.

3.2. Nitrogen Dilution

One way of reducing the heating value is to dilute the natural gas with nitrogen. The richer the gas, the more nitrogen is required to lower the heating value to an acceptable level. However, most gas markets have limitations on content of inert gas in the mixture. Typical US pipeline specifications allow for a maximum of 3 mole% nitrogen. (Carnell et al., 2009)

At many sites the nitrogen is already present in the natural gas and in some cases the gas composition contains over 5 mole% nitrogen. To keep the nitrogen with the LNG it is only necessary to chill the LNG even colder, which increases power demand (Coyle et al., 2007).

When considering using nitrogen to meet natural gas specifications, one must also consider consequences at the shipping and receiving end. Since boil-off gas from a 3 mole% LNG contains over 50% nitrogen, an increase of nitrogen in LNG will have large effect on nitrogen in the boil-off gas (ibid.).

Nitrogen ballasting is not without complications. The scale of usage means that the nitrogen must be available as liquid and this adds the requirement of access to liquefaction plant. (*The Wobbe Index*, 2007) Furthermore, nitrogen removal requires installation units such as adsorption units or membrane separation units. This additional installation would in most cases be uneconomical. (Kidnay et al., 2011) In addition, nitrogen dilution provides low feed gas flexibility, which makes it less desirable for GHV manipulation .

3.3. LPG Extraction Units

As mentioned the heating value are affected by the hydrocarbon composition. A way of modifying the heating value will usually mean adding or extracting heavy hydrocarbons such as ethane, propane and butane. There are several ways of extracting heavy hydrocarbons in the LNG value chain, both before liquefaction, integrated in the liquefaction and at the receiving end.

3.2.3. Turbo-expander Extraction

If it is necessary for a deep extraction it is cost efficient with a turbo-expander extraction process. (Coyle et al., 2007). For an extraction over 80% propane, installation of a turbo-expander should be considered. This installation can be done front end to the liquefaction process. By front end installation LPG extraction occurs independently of the fractionation unit.

A typical turbo-expander extraction process is shown in figure 6. There are many variations possible, but the key feature is dynamic expansion of the natural gas to produce liquids followed by re-compression. The turbo expander will have an isentropic expansion, which makes both high recoveries and high efficiencies. It is applied in processes with a high input pressure of the feed gas. Turbo expander will operate with gas-phase and one can separate cryogenic individual components from the gas, respectively, ethane and propane.

A common way to integrate turbo expander in the extraction process is to first cool down the gas through a gas-gas heat exchanger and then through a separator before expanding. (Bretz & Maddox, 1976) This can also be illustrated on figure 6. The top product, cold gas, enters the turbo expander with high pressure. The gas passes through the expander, an expansion wheel and through the compressor. The gas performs work, losing enthalpy and get a lower pressure and lower temperature. Through the cooling process, the hydrocarbon gas temperatures fall below their respective dew point temperatures and condense out. The two phase product enters a distillation column which separates condensate. Typically, a turbine expander produces 6-8% condensate of the feed gas. (Bloch & Soarez, 1998)

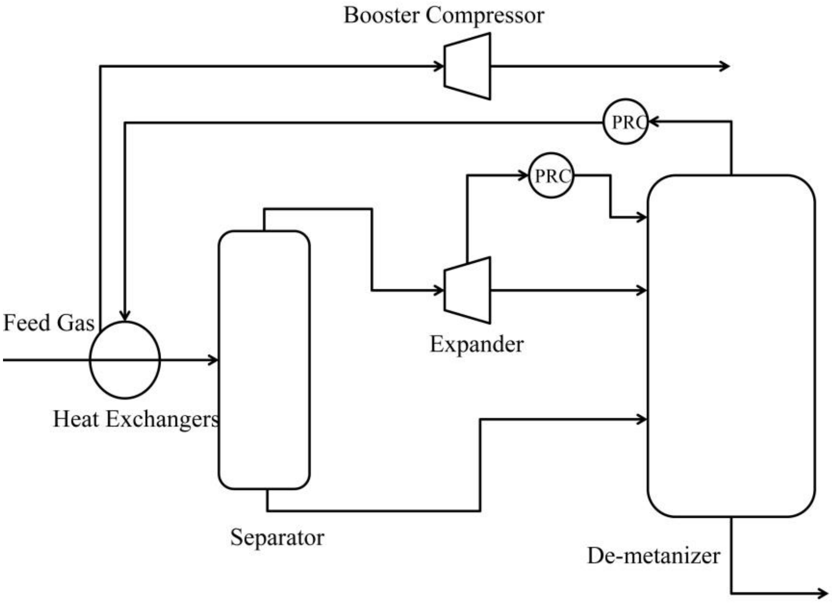

Figure 6 Turbo-expander Process (Simms,2009)

However, despite efficiency and practical installation, the expander shows weakness when it comes to system changes. The expander must respond to the plant's control functions of pressure and flow control. (Campbell, 2001). If there are system changes in upstream units and downstream facilities or different flow conditions, expander operation might be reduced or shut down completely. For treatment facilities where flow compositions and conditions might vary, installation of a turbo expander becomes inefficient.

3.2.4. LPG Fractionation and Recycle

Heating value is also adjusted by the fractionation system in LNG treatment facilities. The fractionation process is situated as an integrated unit after scrub column before liquefaction. A more detailed description of scrub column will be discussed in a later chapter, but the main target of the scrub column is to separate the feed gas into C_{3+} and C_2. Mainly methane and ethane will exit the column in a gaseous state and continue to liquefaction. Ethane and heavier hydrocarbons will exit the column as a liquid bottom product and continue to fractionation. A typical LPG recycle flow diagram is shown in figure 7.

The fractionation process takes the liquid product and purifies the different products. The first separation starts in a deethanizer, where ethane exits at the top and C_{3+} as a liquid bottom product which flows to the next fractionation unit. The following steps are similar, where a depropanizer splits propane and C_{4+} and the last unit, a debutanizer, separates C_4 and $C_{5+.}$ The bottom product is exported as natural gasoline. (Campbell, 2001) Extracting these components from the natural gas will reduce the heating value and is one of the most efficient ways of controlling the heating value.

Another regulation on the heating value is as shown in the figure, by a recycle of top streams back to the scrub column. By recycling the components, the efficiency of the scrub column improves since more heavy hydrocarbons follow the bottom stream. The top product is a leaner gas with a lower heating value. A more detailed view on this process will be further discussed in chapter 4.

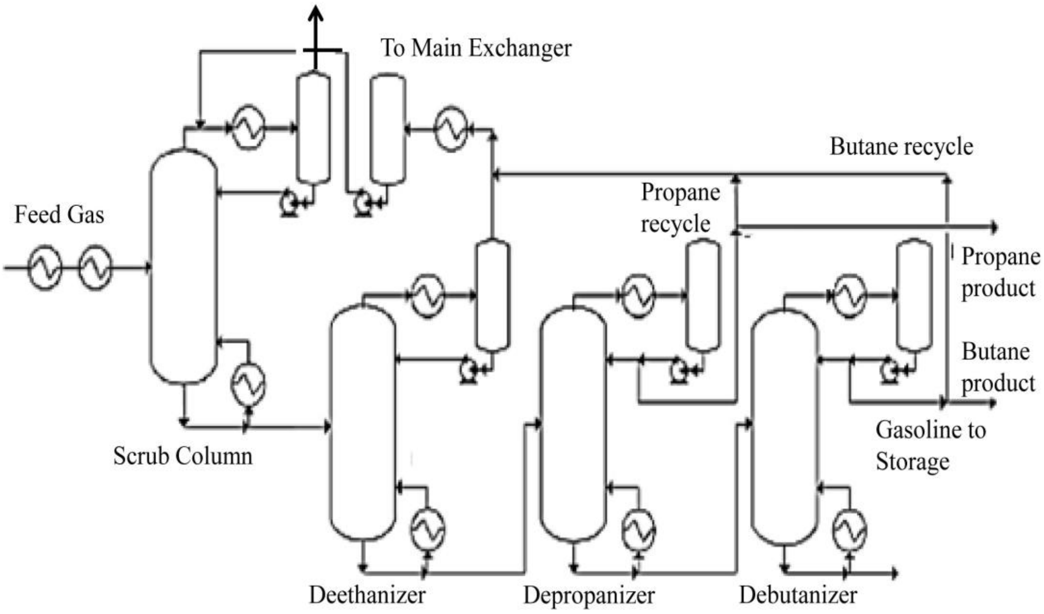

Figure 7 LPG Recycle Process (Coyle et al., 2007).

3.2.5. Scrub Column Modifications

Another method to increase or decrease LPG recovery, and hence affect the heating value is to vary the temperatures in the scrub column. (Madouri, 2004) By using a lower temperature refrigerant in a modified scrub condenser, reflux flow increases. This reduces heating value on end product. Opposite results occur by increasing the condenser temperature or removing this reflux stream completely. More heavy hydrocarbons follow the top stream to further processing, and the heating value increases.

Adjustments on the reboiler can also reduce the heating value. By decreasing temperature on hot oil used in reboiler, or eliminating the reboiler completely, heating value is reduced. However, by doing this, the column sends an increased amount of methane to fractionation. To adjust this increased methane amount in the bottom product, a solution could be to install a demethanizer prior to the fractionation system. This can be seen on figure 8.

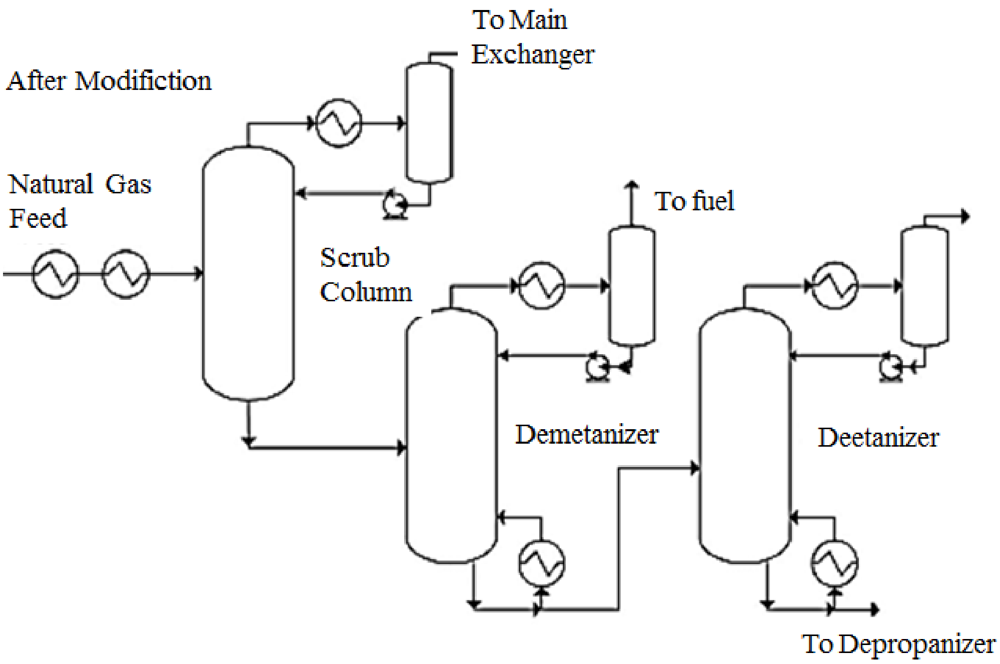

Figure 8 LPG Extraction by Scrub Column Modifications (Chrétien, 2006)

All three of the methods presented may be applied to existing plants, with various degrees of modifications required. The turbo- expander option may actually involve very little change to the existing equipment, since it can be installed front end and the LPG extraction acts independently of the liquefaction unit. The LPG recycle however, is integrated in the liquefaction unit and would require significant modifications or equipment replacements in the fractionation unit. In contrast to a separate installation an integrated installation can also use the same cooling circuits. The cooling for the extraction process can be extracted from the pre-cooling. Economy of scale yields reduced capital cost per unit of heat removed when the size of the liquefaction system is increased. The easiest way to control the GHV would be to do modifications on the scrub column. Since this does not require any new installations, it is a cheap alternative.

4. Design of GHV Control

To get a complete view of heating value control a mathematic fundament and equipment design of a distillation column is presented.

4.2. Distillation Column

A distillation column is normally used to achieve a fine split between components. The device splits the feed into two parts, where at least one of which will have a controlled composition or vapour pressure. In the LNG value chain distillation columns are for fractionation and separation of LPG.

The construction of such a column can be seen in figure 9. Feed enters a distillation column where liquid and gas flow counter and ideally reach equilibrium through several steps within the vessel. Gas will flow upwards, where more of the heavier components are stripped from the gas at each step and ending at the bottom of the column. Some of the heavier components at the boiling point circulate through a reboiler for supplying hot return. (Mokhatab & Poe, 2006). The part of the distillation column above the feed point is referred to as the "Enriching Section", while the part of the column below feed point is known as "Stripping Section".

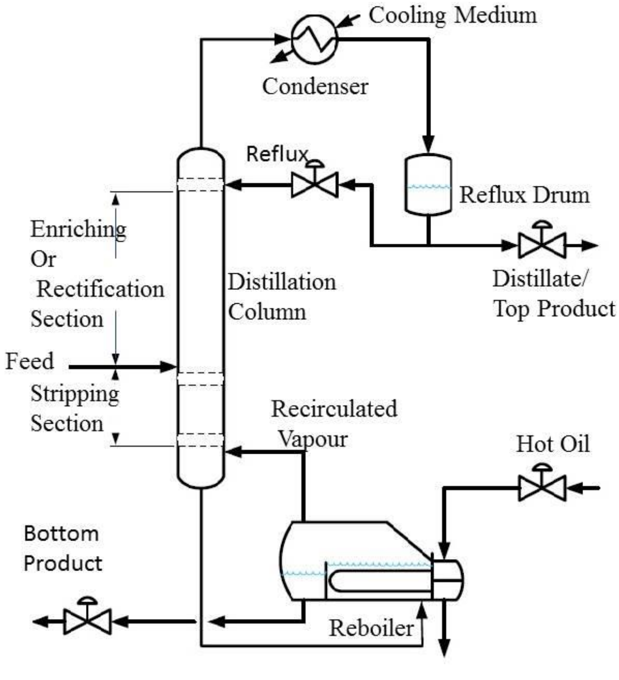

Figure 9 Distillation Column (Cheremisinoff, 2000)

The top product, distillate, enters a condenser and a stripper, and is thus refluxed. In the condenser heat is removed by using a refrigerant. Here the temperature is selected and designed based on the practical temperature difference between the refrigerant and top product. In the reflux drum further cooling is added. The product will leave the reflux accumulator either as a liquid by bubble point, or vapour at dew point. This depends on the use of a partial or total condenser. Using a total condenser, reflux and distillate will have the same composition, and the product is in gaseous form. When using a partial condenser on the other hand, the reflux at the boiling point of liquid is in equilibrium with the distillate and gives a product in liquid form. (Campbell, 2001)

4.2. Mathematic Fundament

The strategy to achieve separation of a mixed component fluid and column design is based on the following fundamental assumptions. (Halvorsen & Skogestad 2000)

- Equilibrium
- Volatility
- Molar Flow

These assumptions, as well as the following equations, are collected from Distillation Theory by Halvorsen and Skogestad (2000) and will be further derived.

A column consists of several steps, where an equilibrium between vapour and liquid is assumed. By temperature and pressure change, separation between these two phases occurs where the liquid will be sent to the lower levels, and vapour is sent to the upper stage. A general equation for the gas and liquid equilibrium is given as follows

$$[y_1, y_2, \dots, y_{Nc-1}, T] = f(P, x_1 x_2, \dots., x_{Nc-1})$$

$$[y, T] = f(P, x) \tag{1}$$

Where f is degrees of freedom, defined from Gibbs' phase rule. N_c is components which do not correspond to each other, T is temperature and $\sum_{i=1}^{n} x_i = 1$ and $\sum_{i=1}^{n} x_i = 1$ are defined as the mole fractions to vapour and fluid.

For ideal mixtures the vapour-liquid equilibrium can be derived from Raoults law, which states that the partial pressure p_i, consisting of i components, is proportional to the vapour pressure p_i of the pure component and the liquid mole fraction x_i.

$$p_i = x_i p_i^0(T) \tag{2}$$

Furthermore, since total pressure is given $P = P_1 + P_2 + \ldots \ldots P_{Nc} = \sum_i p_i = \sum_i x_i p_i^0(T)$, a derivation gives the following relation between liquid and vapour

$$y_i = x_i \frac{x_i p_i^0(T)}{\sum_i x_i p_i^0(T)} \tag{3}$$

The following empirical formula is used to calculate pressure for pure components where *a* to *e* is collected from component databases.

$$\ln p^0(T) = a + \frac{b}{c + T} + dln(T) + eT^f \tag{4}$$

Volatility is defined as the relationship between to components equilibrium constant, K-value, which is very temperature- and pressure dependent. This value reflects whether two components are easily separated.

$$a_{ij} = \frac{(\frac{y_i}{x_i})}{(\frac{y_j}{x_j})} = \frac{K_i}{K_j} \tag{5}$$

For ideal mixtures that satisfy Raoult`s law one gets

$$a_{ij} = \frac{(\frac{y_i}{x_i})}{(\frac{y_j}{x_j})} = \frac{K_i}{K_j} = \frac{p_i^0(T)}{p_j^0(T)} \tag{6}$$

Since $p_i^0(T)$ is temperature dependent the K-values will be constant in the end of the columns where the temperature is approximately constant. (Halvorsen & Skogestad 2000)

Furthermore, a common reference component *r*, is chosen. This component is often the least volatile/heavy component. This gives

$$a_i = a_{ir} = p_i^0(T)/p_r^0(T) \tag{7}$$

Then the equilibrium relation between vapour and liquid will be

$$y_i = \frac{a_i x_i}{\sum_i a_i x_i} \tag{8}$$

Without index, where easy components are expressed as $x=x_1$ and $x_2=1-x$ are expressed as heavy components, the relation will be

$$y = \frac{ax}{a + (a-1)x} \tag{9}$$

This relationship is also expressed in the equilibrium curve on figure 10 where the y- axe represents mole fraction to a light component in vapour phase, while the x- axe represents mole fraction to a light component in liquid phase. The difference y-x gives the amount of separation that can occur at a given level. Large relative volatility implies large differences in boiling point and great separation, while similar boiling points between phases suggest a low relative volatility close to one unit, *a=1,* making separation more difficult. (Halvorsen & Skogestad, 2000).

For hydrocarbon fractionation a relative volatility is normally below 3, and in some cases below 1.5. (Campbell, 2011)

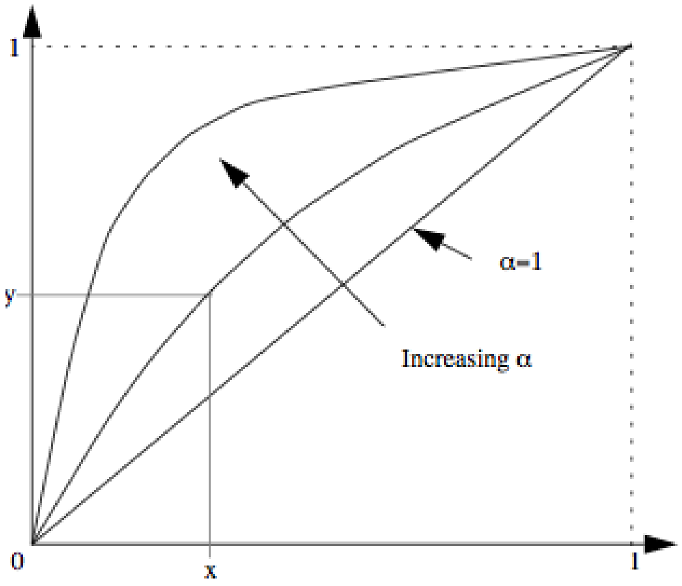

Figure 10 Equilibrium Curve for Ideal Binary Mixture (Halvorsen & Skogestad, 2000).

In addition to establishing equilibrium relations and volatility an important aspect is view of mass balance in the fractionation. A mass balance is done to decide top product, bottom product and composition exiting the column. (Campbell, 2001) To do so, there are certain assumptions regarding component distribution, which will be presented as followed.

A column section is shown on figure 11, where L_n and V_n is the total liquid- and vapour molar flow rates leaving stage *n*.

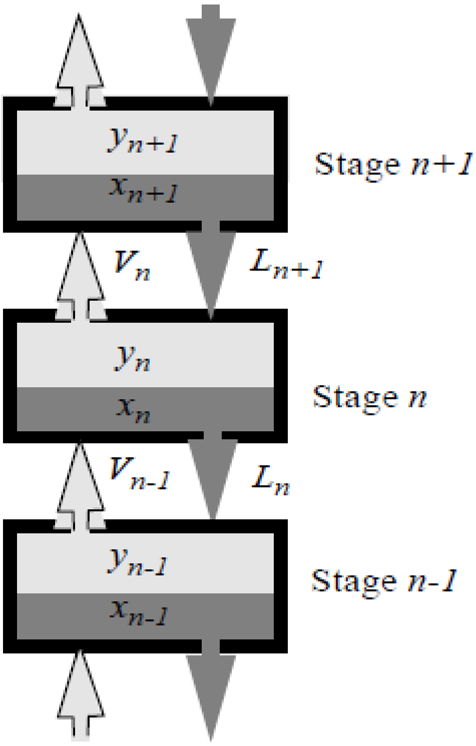

Figure 11 Connected Equilibrium Stages in a Distillation column (Halvorsen & Skogestad 2000)

The material balance for a component *i* by a given stage *n* is given

$$\frac{dN_{i,n}}{dt} = \left(L_{n+1}x_{i,n+1} - V_n y_{i,n}\right) - \left(L_n x_{i,n} - V_{n-1} y_{i,n-1}\right) \tag{10}$$

Where $N_{i,n}$ is amount of mole components consisting of *i* components and *n* stages. It is convenient to define the molar flow w_i by *i* components going upwards from tray *n* to *n+1*. Furthermore, by steady state $\frac{dN_{i,n}}{dt} = 0$, the molar flow is given

(11)

$$w_{i,n} = V_n y_{i,n} - L_{n+1} x_{i,n+1}$$

or

$$y_{i,n} = \frac{L_{n+1}}{V_n} x_{i,n+1} + \frac{1}{V_n} w_i$$

Combined with equilibrium relations this gives the opportunity to compute mole fractions in a section of equilibrium stages. This is also illustrated in figure 12. The resulting curve is known as the operating line.

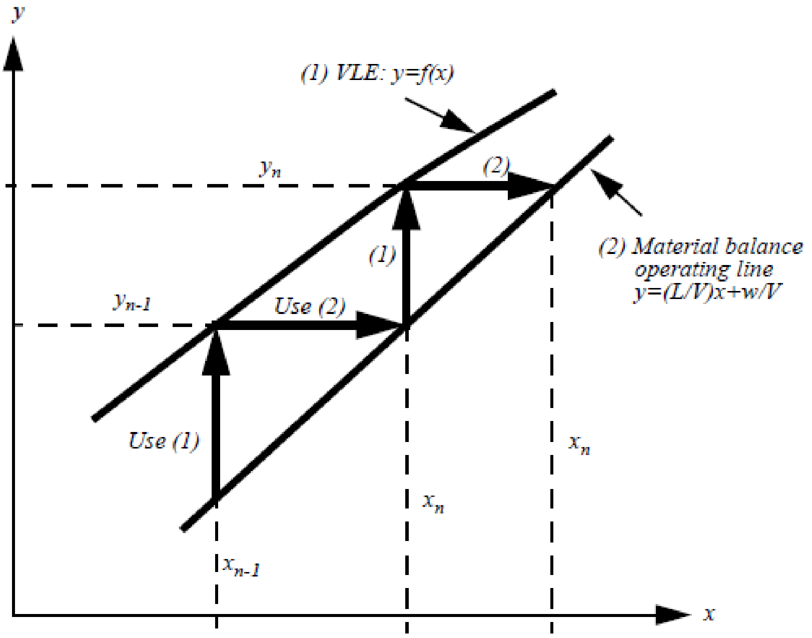

Figure 12 Computation of mole fractions in a section of equilibrium stages (Halvorsen & Skogestad 2000)

Furthermore, it is practical to assume constant molar flows by optimizing columns. One assumes $L_n = L_{n+1} = L$ *[mol/s]* and $V_{n-1} = V_n = V$ *[mol/s]*. This assumption is realistic for ideal mixtures where the components have a similar molar heat of vaporization. This simplifies calculation on molar flows and is presented as

$$y_{i,n} = \left(\frac{L}{V}\right) x_{i,n+1} + \frac{w_i}{V} \tag{12}$$

This fundamental background for column design provides a foundation for more detailed design. Next designing steps includes energy demand, number of stages, feed gas, composition, temperature and pressure, product specifications and accessible heat- and cooling fluids.

4.2. Basic Distillation design

The following subchapters present a description of important aspects to consider when designing a distillation column. These aspects include a calculation of material and energy balance, amount of stages in the column, feed inlet location and plate capacity limitations.

4.3.1. Material Balance

A material balance around the column can be given by equation 13 and gives values of concentrations of components in the distillate and bottom products. These concentrations are set by contractual specifications.

As shown on the equation, the distillate (D) and the bottoms product specifications set the overall material balance for the column. For sharp splits with $X_d \approx 1$ an $X_b \approx 0$, one gets that D = xF. So to meet the distillate and bottoms specification one adjusts the split D/F such that the distillate flow equals the amount of light component in the feed.

$$\frac{D}{F} = \frac{x_F - x_B}{x_D - x_B} \text{ and } \frac{B}{F} = \frac{x_F - x_D}{x_B - x_D} \tag{13}$$

Where

x = mole fraction of component in the feed
D,B,F = rate of flow of distillate, bottom and feed stream, respectively

A change in external flows (D/F and B/F) has a large effect on composition, at least for sharp splits, because any significant deviation in D/F from x implies large change in the composition.

4.3.2. Stages and Reflux Design

Once the foregoing preliminary calculations have been completed, the next step is to determine the combination of number of theoretical plates and reflux necessary to make the desired separation between key components. (Campbell, 1992).

A theoretical stage (plate) is an ideal stage from which the phases exist in equilibrium and is for a given separation obtained with infinite internal flows per unit feed. Similarly the minimum reflux rate occurs at that value of reflux with an infinite number of trays in the column. This can also be viewed in figure 13, where at a vertical asymptote the curve approaches the minimum reflux (L_m) and horizontal asymptote the curve approaches minimum amount of trays. (S_m)

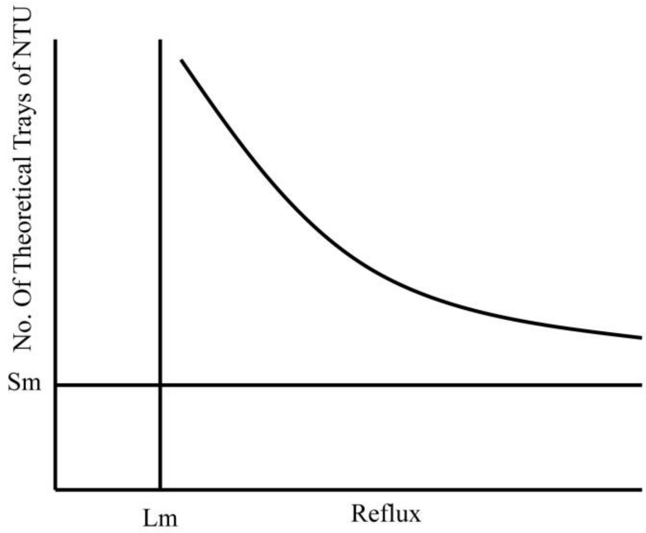

Figure 13 Theoretical trays vs. Reflux (Campbell, 1992)

The Fenske equation showed as followed offers a rapid and fairly accurate method of computing minimum trays. It can be applied to two components in a column at infinite reflux ratio. (Mokhatabm et al., 2006) For a column with N plates, repeated use of this relation gives directly the Fenske`s equation for the overall separation factor:

$$s_m = \frac{\left(\frac{x_L}{x_H}\right)_T}{\left(\frac{x_L}{x_H}\right)_B} = \alpha^N \tag{14}$$

To extend this expression to real columns with finite reflux one must assume constant molar flows, K- values and relative volatility. Most distillation columns operate at an actual reflux from 1.05 to 1.25 times the minimum, regulated to minimize heating and cooling costs from respectively condenser and reboiler. (Campbell, 1992)

The relation between theoretical and actual amount of plates gives the overall efficiency. It is expressed by the equation

$$Overall\ efficiency = \frac{No.of\ Theoretical\ Plates}{No.of\ Actual\ Plates} \tag{15}$$

Typically, distillation columns have efficiency between 50-80%. (Campbell, 1992)

4.3.4. Feed plate Location

The feed tray location should be determined so that the feed enters at a temperature so that no sharp change occurs in the column temperature gradient at the feed tray. (Mokhatabm et al., 2006) Additionally, when evaluating where to position the inlet location, one must take into account that the separation ability deteriorates the lower feed inlet. (*Distillation*, 2009)

There is no exact way to calculate the locate feed plate, but in general one can say that the optimal feed stage location is at the intersection of the two operation lines shown on figure 14. This diagram, the McCabe- Thiele diagram is an advanced diagram from figure 12, which was presented earlier.

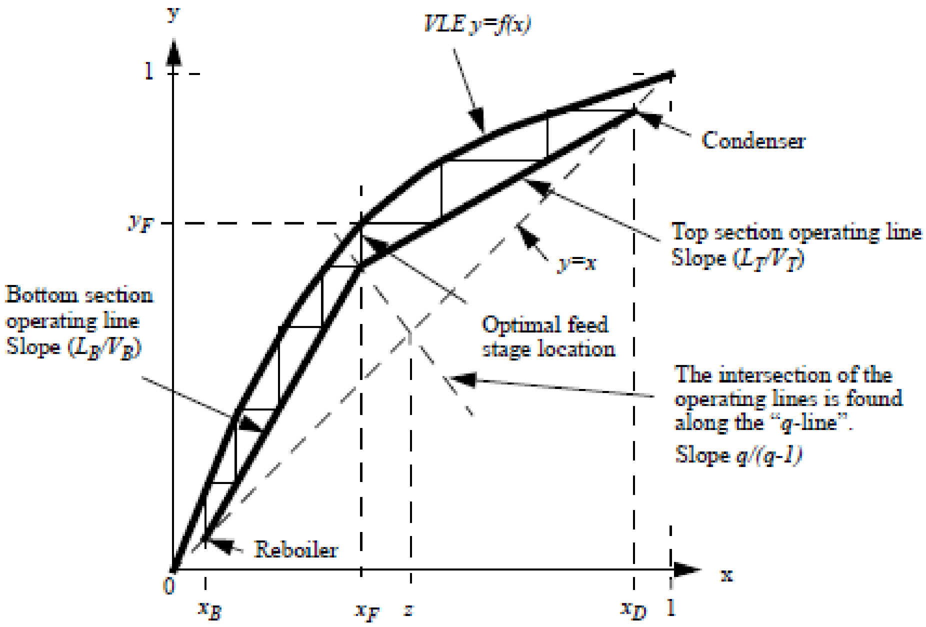

Figure 14 Optimally Located Feed (Halvorsen & Skogestad, 2000)

A way of calculating the feed plate location is by using an empirical correlation

$$\log\left(\frac{N}{M}\right) = 0.206\log\left[\left(\frac{Bx_{HF_k}}{Dx_{LK_F}}\right)\left(\frac{x_{LK_B}}{x_{HK_D}}\right)^2\right] \quad (16)$$

Where

D, B, F =	rate of flow of distillate, bottom and feed stream respectively
N =	no. theoretical plates in rectifying section
M =	no. theoretical plates in stripping section
X_{HK} =	Composition in heavy key, light key, in distillate, bottom and feed.

4.3.5. Plate Capacity Limits

The area of satisfactory operation in a distillation column is bond by tray stability limits. Stability limits are set by amount of liquid and vapour flow rate through the column. This can be viewed in figure 15 where area of normal operation is set by several limitation lines.

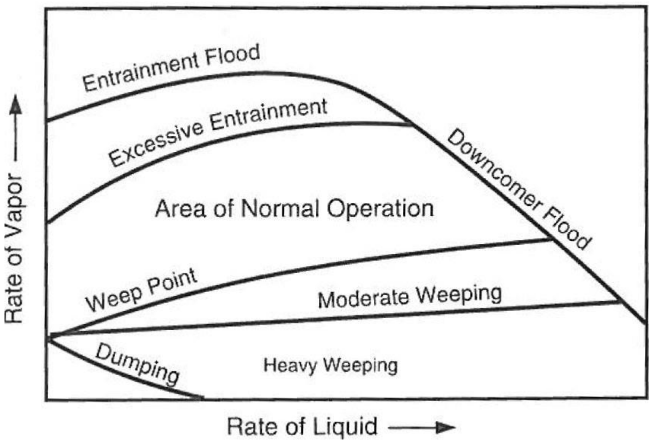

Figure 15 Plate Stability Diagram (Kister, 2000)

One of the complications which can occur in the column is flooding. Flooding is excessing accumulation of liquid inside the column. (Kister, 2000) This can, as shown on the figure, occur if there is great difference between vapour and liquid rate. If the liquid-to-vapour rate ratio is low, one can achieve entrainment, while the inverse relation can form downcomer[3] flood. For downcomer flooding the downcomer can no longer accommodate the liquid leaving the plate, while for entrainment the vapour velocity is too high. The gas space becomes full of aerated liquid or foam, so not all of the liquid can flow downwards by gravity. (Campbell, 2001)

Calculating flooding in a column depends on whether plate or packed columns are used. For trays, the flood point, where entrainment flooding can occur, can be expressed through the Fair`s correlation and Kister - Haas correlation as followed. (Kister, 2000)

---

[3] An opening in the tray which allows the liquid to flow down the column

$$u_{N,flood} = C_{SB}\left(\frac{\sigma}{20}\right)^{0.2}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} \tag{17}$$

$$C_{SB} = 0.144\left[\frac{d_H^2\sigma}{\rho_L}\right]^{0.125}\left[\frac{\rho_G}{\rho_L}\right]^{0.1}\left[\frac{S}{h_{ct}}\right]^{0.5} \tag{18}$$

Where

d_H = Hole diameter [in]
S = Tray spacing [in]
h_{ct} = Clear liquid height at transition from the froth to spray regime [in]
ρ= Density, liquid and vapour [ft/m^3]

In general a tower size should not exceed 75-80% of flood. (Campbell, 2001)

Another limitation in column design is to avoid weeping. If the vapour flow is to low, the pressure exerted by the vapour is insufficient to hold up the liquid on the tray. Therefore, liquid starts descending through the tray perforations (Kister, 2000). If the weeping is excessive, all the liquid on the trays will crash through the base of the column. In addition to efficiency loss, the column might have to be re-started.

Weeping might be difficult to predict, but a simple way to check if weeping might occur is to compare vapour velocity through the plate holes with the weeping factor. The minimum velocity must be greater than the weep velocity. (Kister, 2000)

4.3.6. Energy Balance

The last step in basic calculation scheme of the process design for a distillation column is to make an overall heat balance around the fractionator. (Campbell, 2001) This has importance since it gives a view of economic feasibility and an evaluation of troubleshooting.

It is given as

$$Q_B + Q_C = h_D D + h_B B - h_F F \tag{19}$$

Where

Q_B= Reboiler heat load [kW]
Q_C= Condenser heat load [kW]
h= Enthalphy [kJ/kgmole]
D,B,F= Rate of flow of distillate, bottom stream and feed stream, respectively [kgmole/h]

5. Simulation prerequisites

To be able to assess whether increased max GHV specification can allow more LPG rich feed gas in an LNG plant, two simulation models are established. The following chapter presents an introduction to the modelling and highlights several aspects that need to be considered when establishing the simulation models.

5.1. Feed gas

Feed gas entering onshore facilities can have various compositions depending on both well, reservoir and pipeline conditions. (Gudmundsson, 2012) Non-associated gas, i.e. gas coming from pure gas/condensate fields, normally enters as a dry gas with a relative low composition of heavy hydrocarbons. Associated gas on the other hand, includes tie- in gas from oil fields and has a higher composition of heavy hydrocarbons.

Feed gas compositions used in this thesis are fictitious designed, based on literary examples and designed to achieve the desired specifications. The composition is shown in table 2.

Table 2 Feed Gas [mole%]

	[mole%]
Nitrogen	1.01
Methane	79.98
Ethane	7.47
Propane	3.01
i-Butane	2.01
n- Butane	0.60
i-Pentane	1.95
n-Pentane	1.95
n-Hexane	1.95
n-Heptane	0.03
Octane	0.01
Benzene	0.01
Toluene	0.01

In addition various feed gases will be presented later on to assess feed gas flexibility in the LNG Plant.

5.2. Cases

Two different models are established in this thesis. The first model represents existing facilities, which are constructed based on process flow diagrams (PFDs) provided by the supervisor for this thesis.

The second model is a simplification where adjustments are done to mature the idea of increasing flexibility for handling richer gas. This model is simulated with various feed gases, respectively with an increase of propane and butane. An overview of the two cases is shown in table 3.

Table 3 Simulation Models

Design	Modification	Feed Gas	GHV [MJ/sm^3]
Existing Facilities	None	Case 1	40
Simplification of Existing Facilities	Removal of C_4/C_5 Reflux	Case 2 Case A-E	41

5.3. Heating Value Requirements

As presented in chapter 3.1, GHV has a value around 40 MJ/sm^3, which addresses both Asian, European and American markets. In addition, represented facilities are also designed for this heating value for handling equipment and transport safety. Selected specification for the LNG product in existing facilities is therefore set to be 40 MJ/Sm^3.

As previously mentioned US markets are of less interest, making it applicable to increase heating value. Since the main assignment of this thesis is to address feed gas flexibility by heating value control, modification of the existing facilities is designed with an upper heating value for 41 MJ/sm^3.

Both heating values are represented in table 3.

5.4. True Vapour Pressure

After separation of heavy hydrocarbons in the condensate stabilizer, the end product, condensate, is exported for storage and loading. To make sure the condensate product is fit for storage and loading, there are several specifications that need to be reached. One important specification is the vapour pressure on the product. This has to be taken into consideration to avoid damage to equipment and unsafe transportation.

True Vapour Pressure (TVP) defines the equilibrium partial pressure exerted by the volatile liquid. (EPA, 2013) It is a function of temperature and according to specifications should not exceed 14.7 psi at 100 F. (Nordstad, 2013). These values are shown in table 4.

Table 4 TVP Specifications on the Condensate Product

TVP at 100 F	< 14.7 psi
DVPE at 37.8 °C	< 0.782 atm

5.5. Specifications on Gas Material Flows

Further specifications are given. In addition to affecting WI and GHV, the gas composition affects important parameters regarding gas interchangeability, such as Sooting index[4], incomplete combustion factor, Yellow tip index[5], etc.

To make sure end products, respectively LPG and LNG, reach gas market specifications and ensure safe transportation, restrictions on mole composition is in the simulations in this thesis given as in table 5.

Table 5 Maximal Composition in End Products (Nordstad, 2013)

Gasoline to Storage	C1<1mole% C2<1mole% C5<2mole%
LNG Product	C5<1mole%

---

[4] Soot formation in hydrocarbon combustion (Levinsky, 2005)

[5] Defined by incomplete combustion where excess hydrocarbons could result in unacceptable levels of carbon monoxide. (Fredheim et al.2012)

5.6. Simulation Software

The simulation tool used in this thesis is Aspen HYSYS version 7.3, which is a software programme designed to enable process industry companies to optimize engineering, manufacturing and supply chain operations (Dhole et al., 2011) It can be used for both oil, gas and refinery and simulate both static and dynamic conditions.

The simulation software is constructed with standardized data for state equations for components, selected at start up. In this thesis Peng- Robinsons state equations are chosen to determine all thermodynamic relations. Based on this data collection HYSYS can incorporate mathematical equations which were presented in chapter 4. HYSYS calculates mass balances, energy use, equilibrium compositions and state changes.

5.7. Assumptions

To design a simulation model in HYSYS, several assumptions has been made. These are listed as follows.

- The model represents existing facilities and has starting conditions where initial pressure is 70 bars and initial temperature is -1°C. These conditions are given from PFDs

- To construct a plant with a size dimension similar to an actual plant, it is assumed that the feed gas entering the onshore facilities is approximately 20 million sm^3/day. This is equivalent to a mole stream on 35 234 kgmole/h.

- It is assumed that the gas entering the system already is dehydrated, which means MEG and H_2O is non-existent in the gas. A similar assumption is done for acid gases such as CO_2, H_2S and for Mercury. Since acid gases are excluded it has not been necessary to design separation equipment for sour gases in the following models. This is done because it is believed that including these gases in the feed gas has no effect on desired results.

- Condensate is removed in a front- end stabilizer and the end product remains constant after doing modification to process plant and feed gas.

- All components are designed without irreversible pressure drops. Heat exchangers are also designed with 100 % work efficiency.

- No heat leakage to the environment. All process equipment is considered thermally insulated.

- Compressors and expanders designed in the models are assumed to have an adiabatic 75 % efficiency.

- The split in columns are assumed to be ideal, which means that all butane and propane accumulated as NGL can be exported as LPG.

- Other utility systems such as flare and blow down and additional heating and cooling systems are excluded from the simulation models.

- Theoretical trays in both columns are designed with 100% efficiency, which implies that theoretical and actual amount of trays are the same.

- Some simplifications are also executed after the scrub column. The bottom product from the scrub column enters the fractionation system, where C_4/C_5 is recycled back to the scrub column. Since the fractionation system is a complex system, it is not included in the plant design. The recycle stream is therefore only represented by an external stream, C /C_5 reflux, with conditions and composition provided by supervisor. An important aspect to consider when simplifying this process step is that amount C_4/C_5 reflux cannot be higher than C_4/C_5 exiting as bottom product. This is taken into account.

- The scrub column is designed as a sieve tray column, since it will be working under high pressure without pressure loss.

5.8. Scrub Column Parameters

When constructing and optimizing the column, there are several parameters one can manipulate to get desired results. These parameters are presented in table 6.

Table 6 Scrub Column Parameters

Temperature after Condensation [°C]	-50
Temperature in Reboiler [°C]	56.8
Temperature exiting the Column [°C]	-27.5
C_1 at Bottom Flow [-]	0.15

The table presents ideal values, but to achieve various results and wanted specifications some deviations are accepted throughout the simulations.

5.9. Assumption summaries

A summary of assumptions and modelling specifications are presented in table 7.

Table 7 Summary of Modelling Specifications

Thermodynamic model	Peng- Robinson
Feed Gas conditions	P=70 bar T=-1°C $\dot{m}$ = 20 million sm^3/day
Condensate Stabilizer	P = 15.2 bar T_{bottom} = 215°C TVP<14 psi C1<1mole% n=20
Compressors/ Expanders	$\eta = 0.75$
Heavy Hydrocarbon Column	P = 60 bar $GHV_{with\ reflux}$ = 40 MJ/m^3 $GHV_{without\ reflux}$ = 41 MJ/m^3 $C5_{top\ product}$ <1mole% n=21 $T_{reboiler}$=56.8°C T_{top}= - 27.5°C Sieve trays
C_4/C_5 Reflux stream	P=60.8 bar T=-33.8°C $\dot{m}$= 348.2 kmole/h

6. Existing Facilities

The following chapter presents a description of the model which depicts molar and mass flow results as well as equipment behaviour and values. A more detailed view of the scrub column performance is also presented, where mass balances, mechanical design, composition, LPG recovery, temperature profile and energy performances are highlighted.

6.1. Description of Model

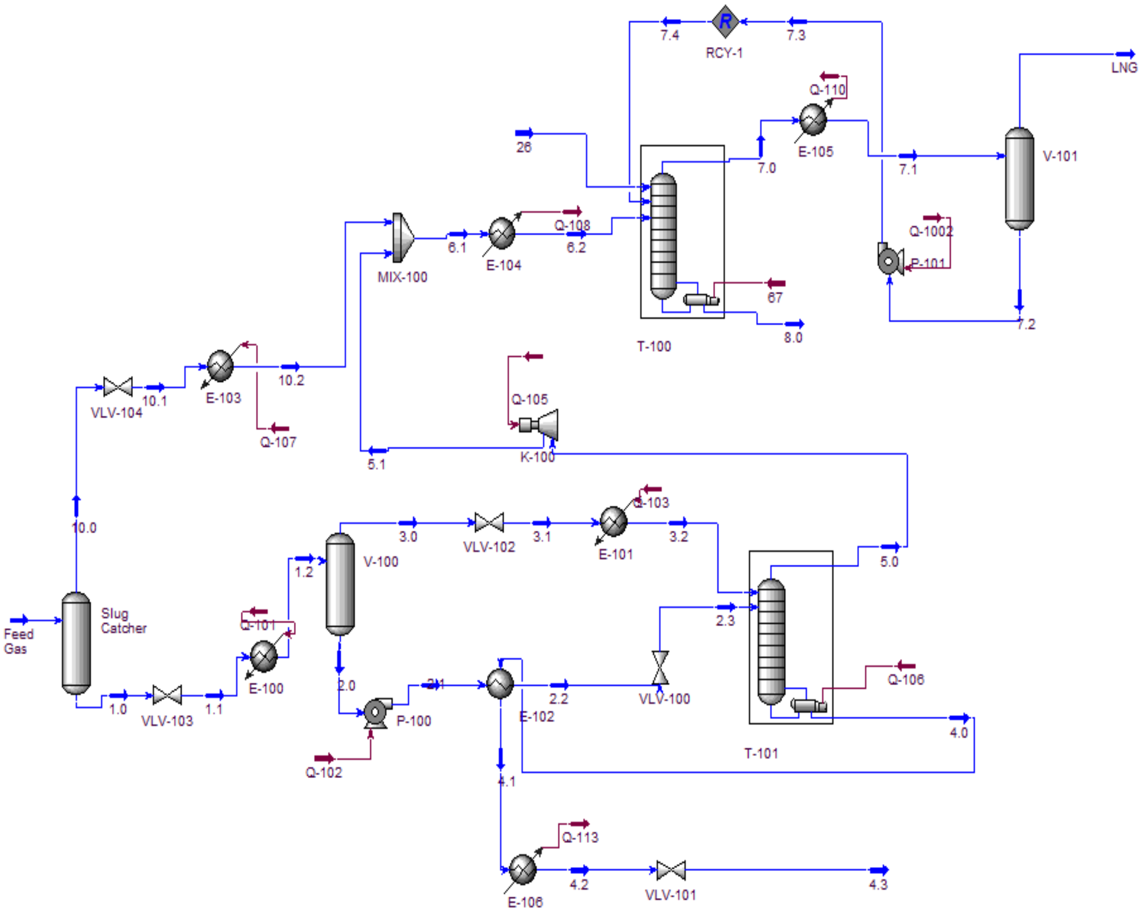

Figure 16 Existing Process Facilities from Reception to Liquefaction

A description of the simulated model can be seen figure 16, and will be further described in the following subchapters. A better review of the process with all condition values can be found in attachment II.E. In addition, a full scale figure can also be viewed in attachment II.A.

6.1.1. Slug Catcher and Inlet Facilities

A two phase feed flow on 20 million sm^3/day enters the slug catcher with a temperature -1°C and a pressure of 70 bar. As mentioned previously, the slug catcher is a three phase separator, but is for the simulation represented by a two phase separator, separating gas and condensate. Most of the feed gas exits as a top product going directly to further processing. The heaviest hydrocarbon exits as a bottom product, preparing for extraction in the condensate stabilizer.

6.1.2. Condensate Stabilizer

After the slug catcher the condensate bottom stream is expanded and heated up before it is separated in two streams. Both streams are then cooled before arriving the condensate stabilizer. The condensate stabilizer has 21 trays and an operating pressure at 15 bars and a temperature for 215.7 °C.

The simulation model represents the stabilizer as a reboiled absorber, consisting of a reboiler which increases separation in the column. The stabilizer extracts the heaviest hydrocarbons from the stream, where the bottom products, Condensate, is cooled by heat exchange with the inlet stream and send for transportation and storage with a RVP below 1 bar.

6.1.3. Recompression

The top product after leaving the condensate stabilizer have a pressure of 15 bars, which will then enter a compressor before returning back to the overhead gas stream from the slug catcher. This stream, now at a pressure at 70 bars will then get further heated before arriving the scrub column.

6.1.4. Heavy Hydrocarbon/ Scrub Column

The gas stream enters the scrub column at a pressure 60 bar and a temperature around – 30 °C. The scrub column separates the gas stream between propane and ethane, where propane and heavier hydrocarbons follow the bottom product to further fractionation.

Since the scrub column operates at such a high pressure it is very sensitive in operation. This makes converge challenging. To achieve converge in the column; it is designed as a Reboiled Absorber where condensation and separation is designed separately from the column. The condenser and separator send the top product back to the column, which acts like a reflux and improves separation in the column. The outer top product goes to liquefaction and will eventually end up as the final LNG product.

The bottom product which goes to fractionation originally first enters a Propanizer. This column separates C_3 and heavier hydrocarbons, where the bottom product are recycled and used as reflux in the scrub column. This recycle stream is represented as stream 26 in figure 16 and will in this thesis be expressed as C_4/C_5 reflux.

More details regarding design and behaviour of the scrub column will be presented in chapter 6.3.

6.2. Achieved Specifications

Achieved specifications are presented in table 8. Temperatures given are similar to data provided from supervisor.

Table 8 Achieved specifications

LNG Product	GHV [MJ/(sm3] ≈ 40	40.00
	$C_{4+}<0.01$	0.001
	T_{top}[°C] ≈ -50	- 50
To Fractionation	C1	0.14
	$C_2<1$mole%	0.1*
	$C_{5+}<2$mole%	0.11*
	T [°C]	102.7
Condensate Product	TVP at 37.8C [bar]	0.97
	T [°C]	115
Overhead gas after Condensate Stabilizer	T<20°C	69.97

* Prior to fractionation

6.3. Molar and Mass Balances

When presenting molar and mass flows, figure 17 shows inlet and outlet streams and are related to values given in table 9. Green lines on the block diagram represent recycled streams, which behave as reflux in the system.

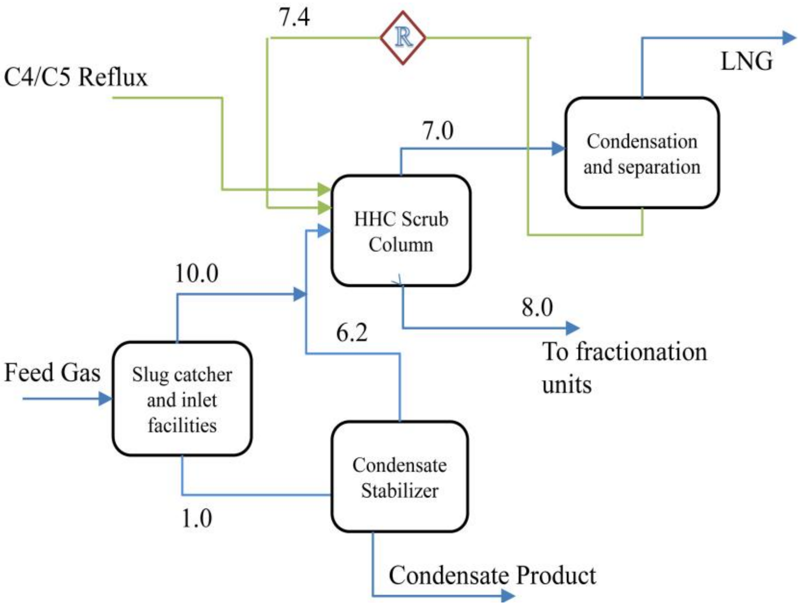

Figure 17 Block Diagram

An overview of molar and mass flows through the system is presented in table 9. Condensate and LNG, written in cursive in the table, leaves the system. Bottom products leaving after the fractionation system is not represented.

Table 9 Molar and Mass Flows from Reception to Liquefaction

	Molar Flow [kgmole/h]	**Mass Flow [tonnes/h]**
Feed Gas	35 240	871.7
1.0	9526	385.2
6.2	33 840	763.1
Condensate Product	*1400*	*108.6*
7.0	32 720	590.8
8.0	*3936*	*186.7*
C_4/C_5 Reflux	348.2	22.3
7.4	2468	61.9
LNG	*30 250*	*529.3*

6.4. Scrub Column Performance

There are several parameters in the scrub column which affects the separation results. Both amount of trays, tray inlet positions and size of reboiler and condenser can be varied when designing a distillation column. Based on design parameters given in subchapter 5.8 and table 6, the results of separation and internal flow balance are presented.

6.4.1. Column Flow Balance

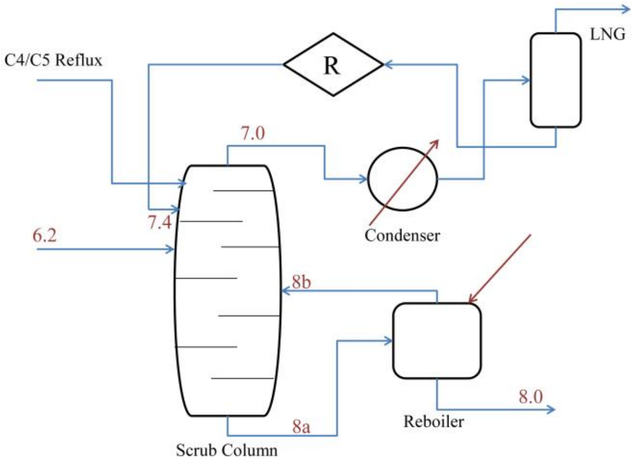

Figure 18 Process Block Diagram of Column Streams

Figure 18 shows flows entering the the vessel and flows connected to reboiler and condenser. Flows 8a and 8b represents flows to and from reboiler respectively. The column consist of an integrated reboiler and externally constructed condensation and separation. Originally the C_4/C_5 reflux comes from stream 8.0, but as previously mentioned, the process is simplified by designing the stream C_4/C_5 reflux externally. The internal flows are presented in table 10.

Table 10 Column Molar and Mass Flows

Stream	**[kmole/h]**	**[tonnes/h]**
6.2	33 840	686.1
8a	7843	331.0
8b	3908	144.2
8.0	*3936*	*186.7*
7.0	32 720	590.8
7.4	2468	61.9
LNG	*30 250*	*529.3*
C4/C5 Reflux	348.2	22.3

With a mass boil-up ratio at 1.159 and a reflux ratio 0.027 in the reboiler, the stream going to boil up is 331.0 tonnes/h. The bottom reflux flow, stream 8b, is 144.2 tonnes/h, consisting of mainly methane, ethane, propane and butane. A relation between the column flows can also be viewed on figure 19, showing final split in the column. 88% of the inlet streams ends up as an LNG product. This estimates as 529.3 tonnes/hour or 17.1 sm^3/day.

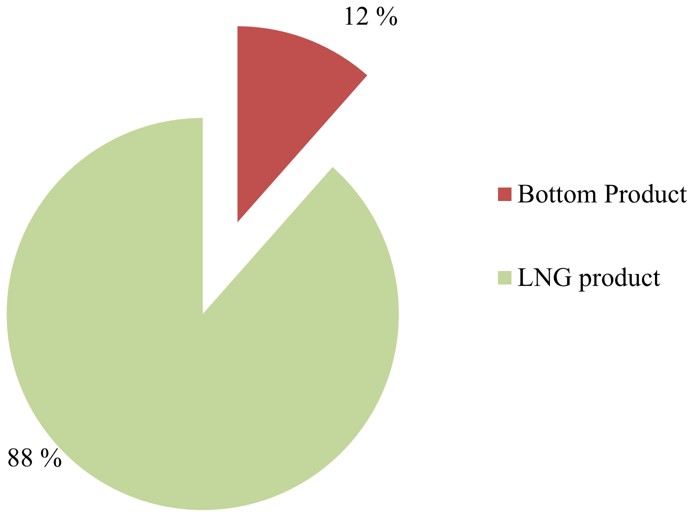

Figure 19 Mass Flow Products

Furthermore, figure 20, shows flow relations in the condenser and reboiler respectively. The exit product for the condenser is the LNG product, while the exit product for the reboiler, stream 8.0, goes to further fractionation.

As shown, the bottom reflux stream in the reboiler is much higher than top reflux after condensation. About 50% of the bottom stream leaving the column is reboiled and recycled vapour, while the reflux after condensation is only about 6%. Separation efficiency is mostly dependent on reboiler conditions. The reboiler streams depend on reboiler temperature, which will be limited by the mixture condition in the reboiler. I addition, the reflux temperature affects the reboiler temperature. The lower the reflux temperatures, the higher the reboiler temperatures will be and hence higher boil up streams.

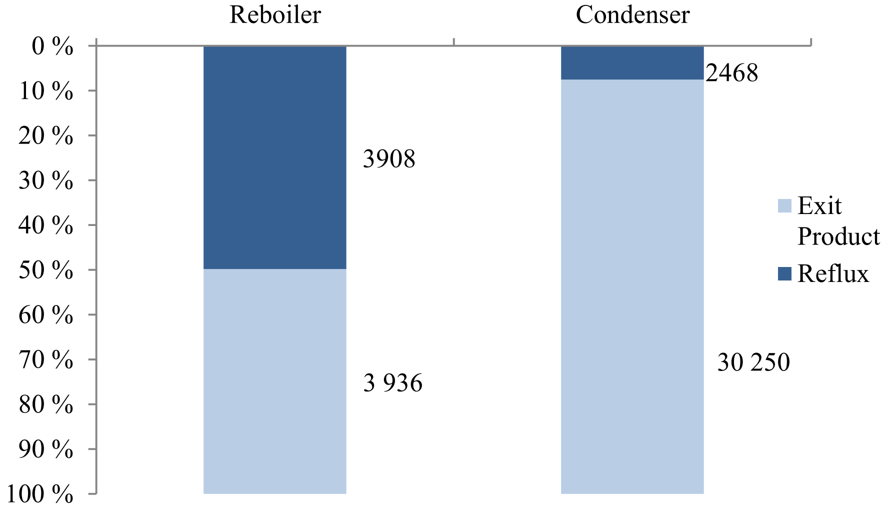

Figure 20 Flow Relations in Reboiler and Condenser

6.4.2. Mechanical Design

Based on HYSYS tray sizing utilities, the column geometry and hydraulic results are presented in table 11. Calculation methods are based on mathematic methods presented in chapter 4.3.5 and provided by tray/packing vendor Koch-Glitsch. As a default, HYSYS calculates with a flooding max estimation on 70%. However, the system can converge and handle flooding propabilities over 70% without system warnings.

Table 11 Column Geometry and Hydraulic Results

Jet Flooding Method	Glitsch
Section Diameter [m]	4.267
Hole Area [m2]	1.204
Tray Spacing [m]	0.609
Section Height [m]	12.80
Max Flooding [%]	70.49

6.4.3. Column Feed Composition

The effect of the feed tray inlet location can be viewed by figure 21, which shows the relation between the mole fraction in the internal flows in the column and amount of trays. The y-axis depicts the mole composition while the x-axis represents trays in the column, from bottom (0) to top (21).

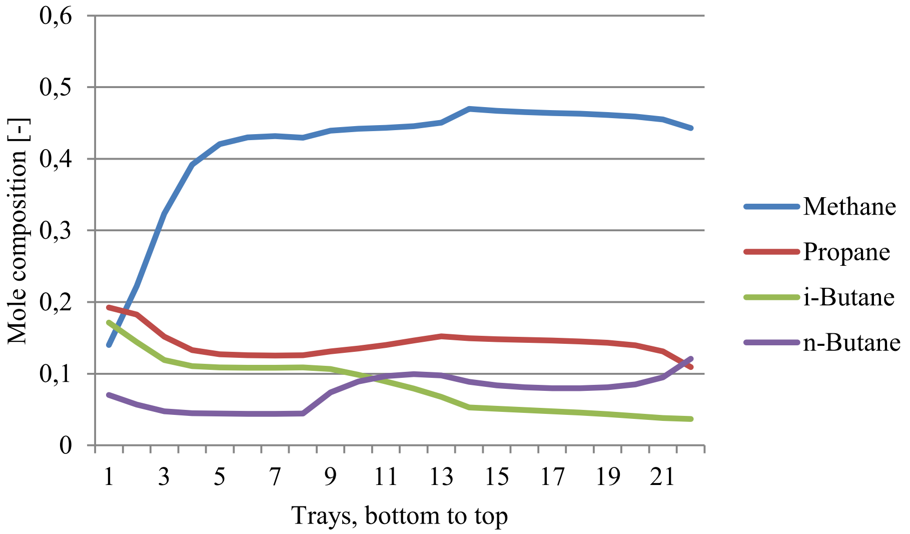

Figure 21 Composition through the Column, Bottom to Top

As shown, methane will have a composition increase across the column, from tray 0 to 21, bottom to top. Correspondingly, the propane and heavier hydrocarbon composition are highest at the bottom of the column. After tray 7, where the main gas stream enters, the composition of propane increases downwards, while stabilizing after tray 13 and upwards.

6.4.4. LPG Recovery

The split of components through the column is presented in table 12, where the table shows molar flows in flow 7.0 and 8.0, top and bottom respectively.

Table.12 LPG Split in the Column [kgmole/h]

	7.0	**8.0**
Propane	501.2	756.6
i- Butane	78.0	674.5
n- Butane	184.4	276.8
LPG	763.6	1707.9

Values from table 12 is also indicated in figure 22. The y-axis represents percentage of flow products, propane and butane respectively. The split between 7.0 and 8.0 expresses the split between top and bottom product. As shown, 60% of the propane is recovered in the bottom, while for butane 79% follows the bottom product. In combined, the scrub column recovers 1707 kgmole/h LPG or up to 70%, which goes to further fractionation.

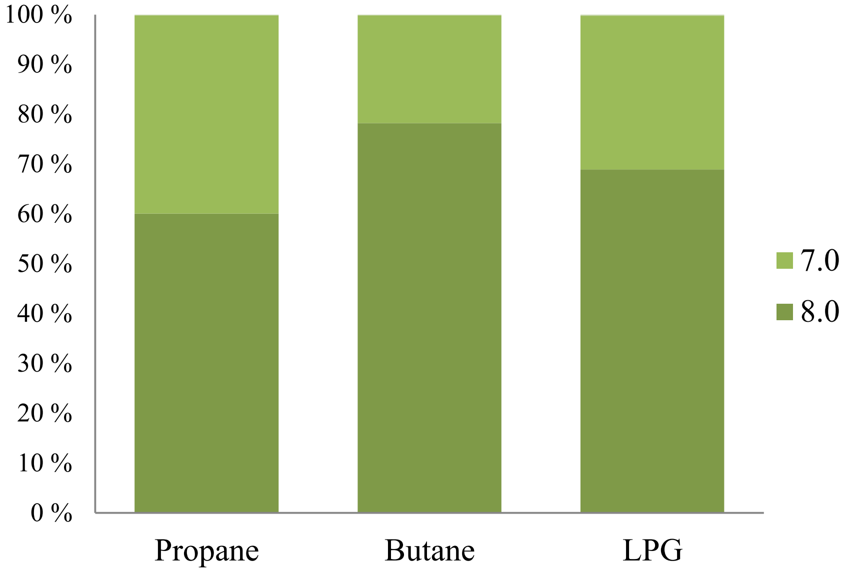

Figure 22 Propane and Butane Recovery in the Scrub Column

6.4.5. Temperature profile

Figure 23 shows the temperature profile through the column, where the x- axis represents amount of trays, 0- 21, bottom to top. Since there are temperature specifications that need to be met, the temperature across the column behaves as presented. As shown, the temperature is highest in the bottom, where temperature after reboiler is around 100 °C. At tray 7, where the main gas stream, temperature increases rapidly downwards towards the reboiler. From 7 to the top, temperature remains almost constant and exits the column at around -25 °C.

The temperature increase is highest from tray 5 to tray 0, which relates with the high reboiler stream entering at tray 1. At tray 13, where the internal reflux enters, the temperature increases downwards. The C_4/C_5 Reflux enters at the top of the column and contributes maintaining a temperature on top product exiting at -27°C.

Associated values to figure 23 can be found in attachement II.F.

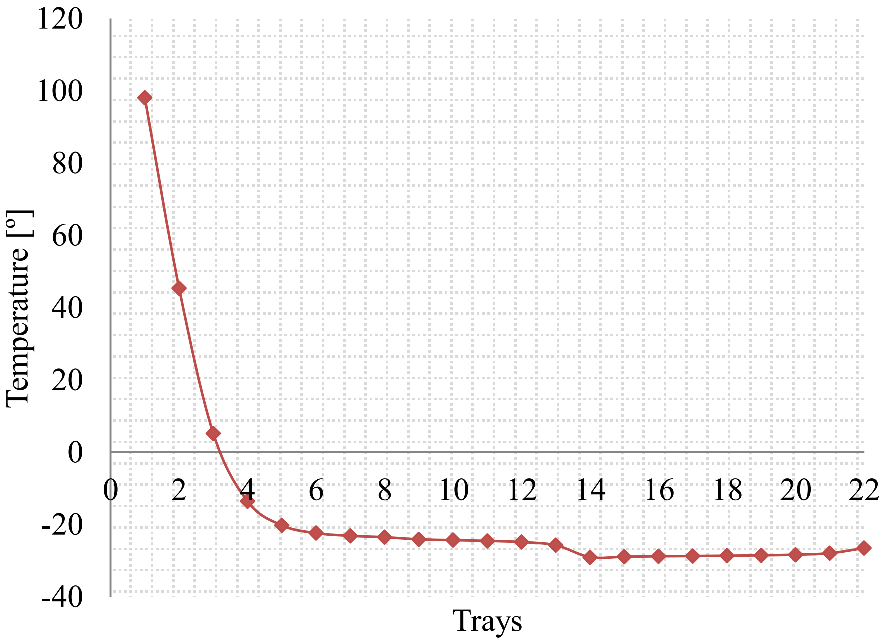

Figure 23 Temperature vs. Tray Position from Bottom

6.4.6. Vapour/Liquid Flow

As mentioned in chapter 4.3.5.the relation between vapour and liquid trough the column is important due to challenges with weeping and flooding. Figure 24 shows the vapour and liquid flow through the column. As shown, when feed gas enters at tray 7 and upwards the vapour-liquid molar flow ratio is at the highest. As discussed in chapter 4.3.5, great differences between vapour and liquid flow rates might induce flooding. After tray 15 the vapour flow is about 33 000 kgmole/h while the liquid flow rate is around 1000 kgmole/h. In this area the column is most vulnerable for unstability.

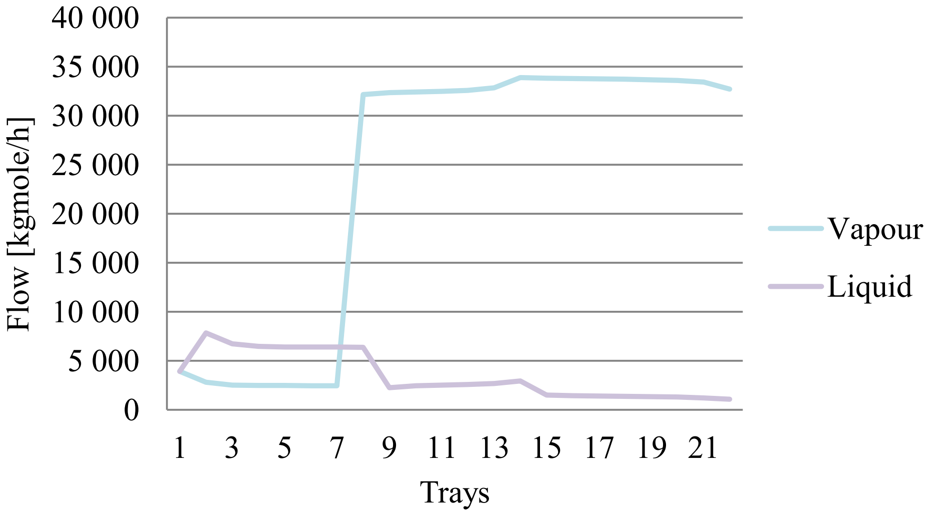

Figure 24 Flow Relations in the Column

As mentioned, HYSYS uses a default value on 70% flooding and a weeping factor 1 and as shown in table 11 in subchapter 6.4.2, mechanical design, the flooding range is below critical level with a flooding value 70.49% and within operation range.

6.4.7. Energy Balance

To achieve wanted separation in the scrub column, the column needs energy supply. Since all the vapour driving the separation comes from the reboiler, energy demand is highest at the bottom. The column uses hot oil to provide the correct temperature in the reboiler to vaporize the lightest components in the flow. Correspondingly, in order to cool down the top product, induce a reflux rate and improve separation energy is also needed treating the top product. A condenser with an external cooling medium is required to achieve a reflux and reducing the temperature from -27°C to -52°C. Consequently, energy in the form of heat is removed from the system. This is done by heat exchanging with a cooling medium. Ideal cooling medium is sea water, due to great availability and minimal additional costs. However, if excessive cooling is needed, recycled propane or other mixtures can be used.

The theoretical approach to the energy supply is presented in chapter 4.3.6. As previously presented, the energy demand is affected by volatility, feed rate, feed conditions and other column design factors. The energy supply in both reboiler and condenser is presented as red lines in figure 16, and with values in table 13.

Table 13 Energy Use in Reboiler and Condenser

	[kW]
Reboiler	20 170
External condenser	18 070

7. Simplification of Existing Facilities

To change GHV specifications there are several controlling methods, which can be applied to the existing facilities. An easy way to increase the heating value from 40 MJ/sm^3 to 41 MJ/sm^3, is as discussed in theoretical background, by doing modifications in the scrub column. This is executed in the following model and as shown on figure 25, simplifications are performed on circled area, where stream 26, C_4/C_5 reflux, is removed.

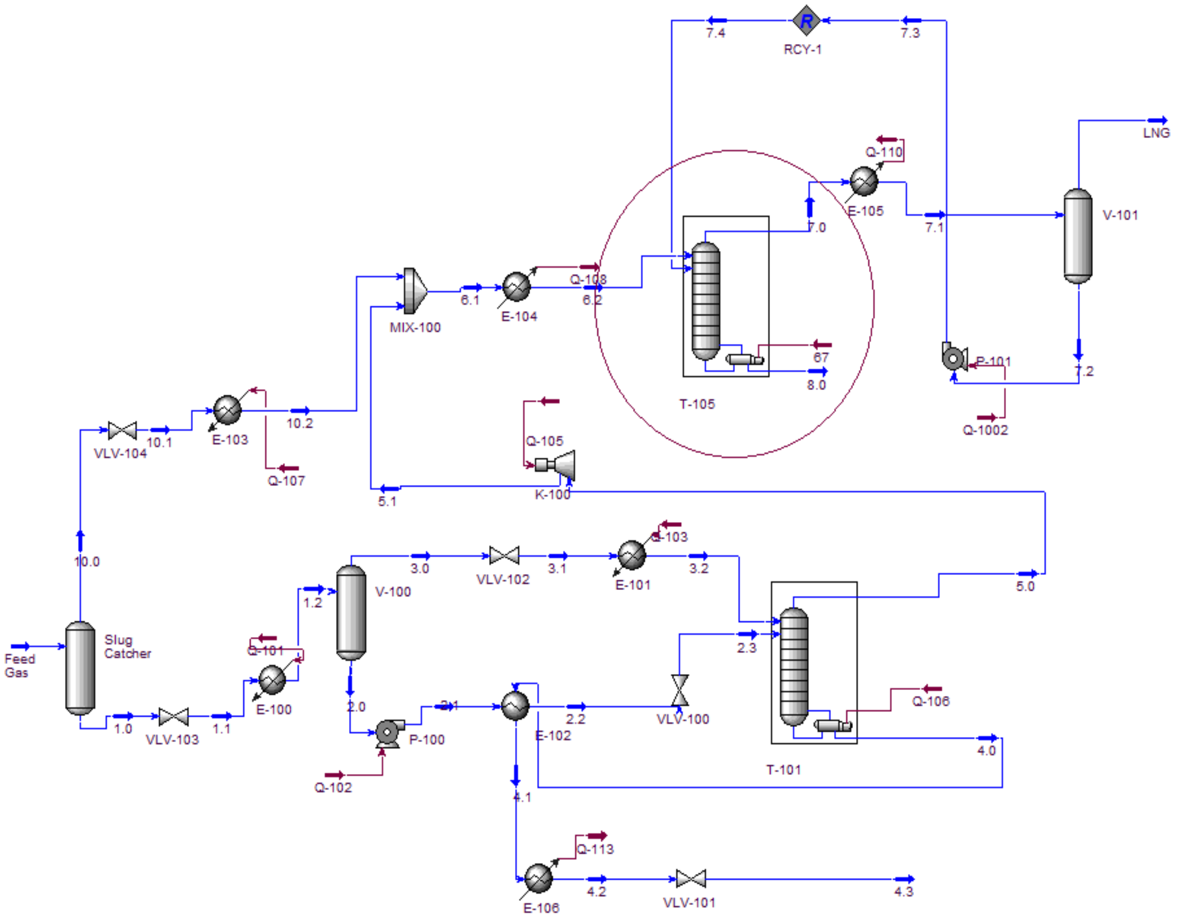

Figure 25 Modifications on Existing Facilities

Since no changes are done on facilities prior to the scrub column, this chapter will only present results on scrub column and after. Similar as for existing facilities, a more detailed presentation of figure, conditions and compositions can be found in attachment III, A-E respectively.

7.1. Description of Modifications

As mentioned, simplifications are done on the scrub column, where C_4/C_5 reflux is excluded. Instead of four inlet feeds at stage 1, 7, 13 and top inlet, there are only three inlet streams. These streams are feed gas from pre-treatment facilities and the internal reflux streams.

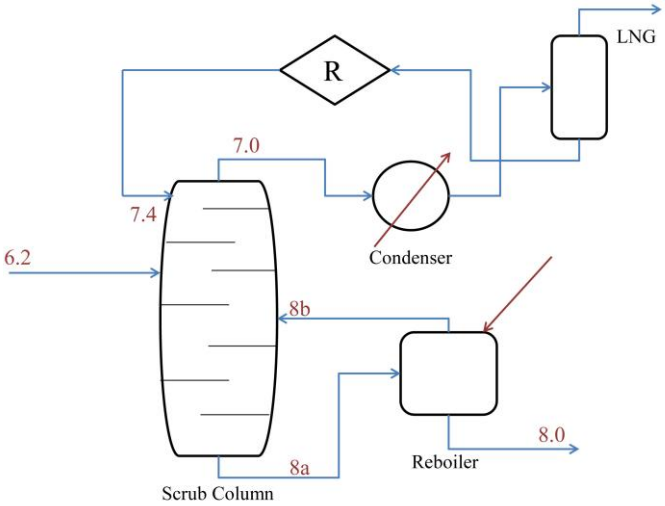

Figure 26 Modified Scrub Column

Furthermore, to have full functionality in the column and to achieve convergence in HYSYS, the internal reflux stream, 7.4, now enters at the top. New inlet and outlets can also be viewed in figure 26.

7.2. Achieved Specifications

The simulation is done based on desired increase in heating value and achieved specifications are presented in table 14. By removing the reflux stream, separation capacity decreases and more propane follows the top stream. This results in more heavy hydrocarbons in LNG product and a reached heating value on 40.45 MJ/sm^3.

Table 14 Acheived Specifications

LNG Product	GHV [MJ/sm^3] ≈ 41 T_{top}[°C] ≈ -50	40.45 -50
To Fractionation	C_1 T [°C]	0.14 100

To further achieve a heating value towards 41 MJ/sm^3, the top temperature in HYSYS can be adjusted. By lowering temperature after condensation fewer heavy hydrocarbons liquefy and reflux rate is reduced. However, to minimize amount of parameters, results are presented with a temperature similar to existing facilities.

7.3. Molar and Mass Balances

Table 15 shows molar and mass flows entering and exiting the column. Stream 6.2 remains similar before modification, while all other streams are adjusted.

Table 15 Molar and Mass Flows

Stream	[kmole/h]	[tonnes/h]
6.2	33 840	686.1
8a	6293	266.5
8b	3100	114.0
8.0	3192	152.5
7.0	35 180	648.9
7.4	4529	104.5
LNG	30 670	541.6

With a boil up ratio 0.9712 and a reflux ratio 0.072, the bottom product, stream 8.0, ends up at 152.5 tonnes/h. The top product, stream 7.0, is 648.9 tonnes/h and after condensation and separation the LNG product is 541.6 tonnes/h or 17.4 sm^3/day. With new specifications the end products are divided as shown in figure 27, where 91% ends up as LNG product. As shown, 9% of the inlet flows ends up as a bottom product. With a composition similar to existing plant, but with a reduced reflux rate, separation is less efficient than existing facilities.

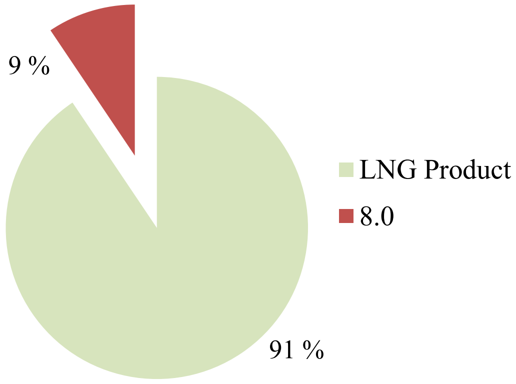

Figure 27 Split in the Column

Flow relations can also be viewed in figure 28, which presents flow amounts used as reflux. As seen, 50% is used as reflux in the reboiler. The top reflux stream, stream 7.0 respectively, is only about 10% of the top stream.

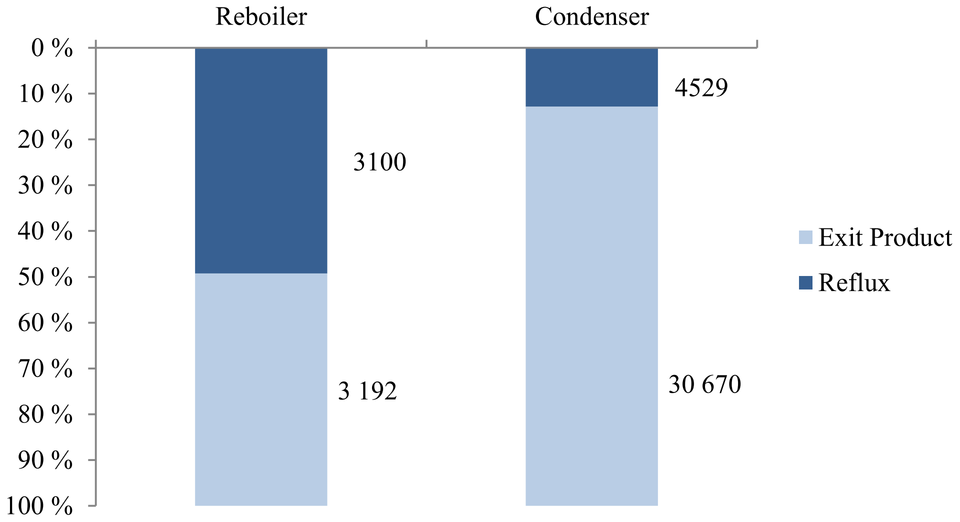

Figure 28 Molar Flow through Reboiler and Condenser

7.4. Mechanical Design

The mechanical design parameters are presented in table 16, and as shown max flooding occur at 73%. Although HYSYS operates with a flooding maximum on 70%, the system still converges without warnings. The scrub column therefore still operates in an acceptable area.

Table 16 Mechanical and Hydraulic Results

Jet Flooding Method	Glitsch
Section Diameter [m]	4.267
Hole Area [m2]	1.132
Tray Spacing [m]	0.609
Section Height [m]	12.80
Max Flooding [%]	73.66

Mechanical design results are values which correlate to actual facilities, but are presented as guidelines and can to some degree be used as boundary limits when doing changes within the scrub column. Therefore some deviations between existing facilities and modified system are accepted. However, when performing changes both section diameter and tray spacing remains similar.

7.5. Column Feed Composition

Figure 29 shows the composition through the column, from plate 0 to 21, bottom to top. As shown, the LPG composition remains under 0.2 mole, and is fractioned similarly both in top and bottom.

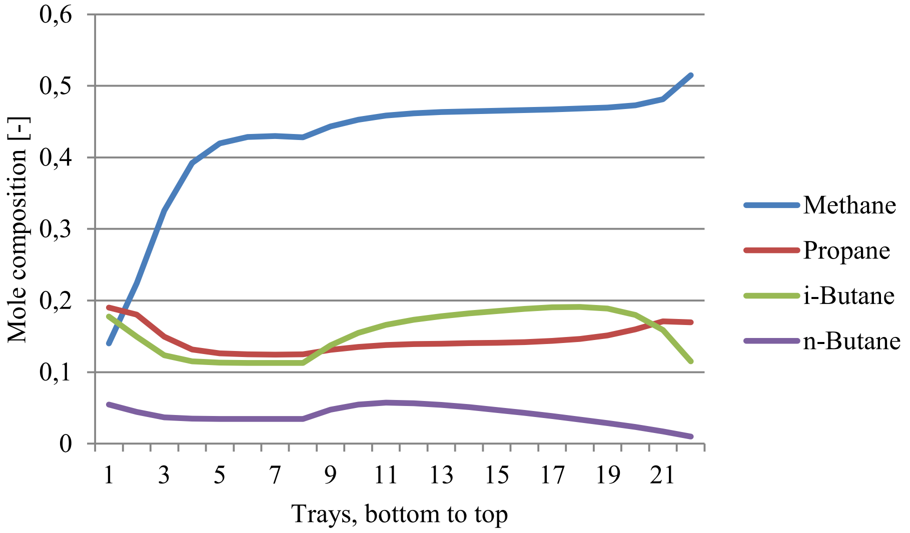

Figure 29 Composition through the Column

7.6. LPG Recovery

Table 17 shows propane and butane products in both top and bottom stream, respectively stream 7.0 and 8.0. Additionally, figure 30 shows recovery for these streams, given in percentage. The bottom stream 8.0 contains 607 kgmole/h propane, which results in a recovery percentage on 40%. Similarly, 742 kgmole/h butane i.e. 70% of the butane follows the bottom product. These bottom products in combined, results in an LPG recovery about 55%, or 1204 kgmole/h.

Table 17 LPG split in the Column [kgmole/h]

	7.0	**8.0**
Propane	890.0	607.5
i- Butane	290.8	567.8
n- Butane	23.4	174.94
LPG	1349.7	1204.2

Propane in the bottom is directly dependent on amount of C_4/C_5 through the column, and due to removal of this reflux, the propane recovery decreases. Instead, the amount is included in stream 7.0. Consequently, the increase of propane increases the GHV in the LNG product.

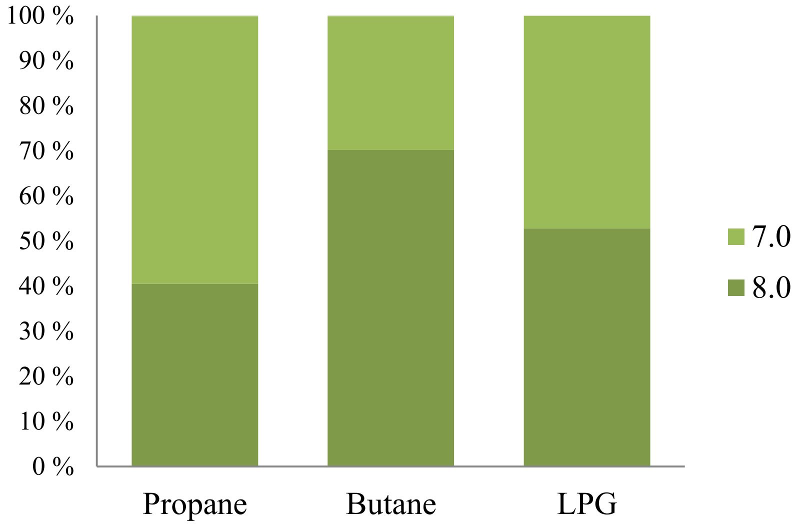

Figure.30 Propane and Butane Recovery

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7.7. Temperature Profile

The temperature profile for the modified scrub column is shown in figure 31, where the x-axis represent amount of trays, bottom to top, while the y-axis is temperature. Reboiler design temperature is initially 56.8 °C, but in order to achieve convergence for the column without C_4/C_5 reflux in HYSYS, some deviations have been accepted. Temperature in the bottom is therefore somewhat higher than in existing facilities, where the bottom product exits the system with a temperature around 100 °C. The temperature stabilizes from tray seven and upwards until exiting at – 30 °C.

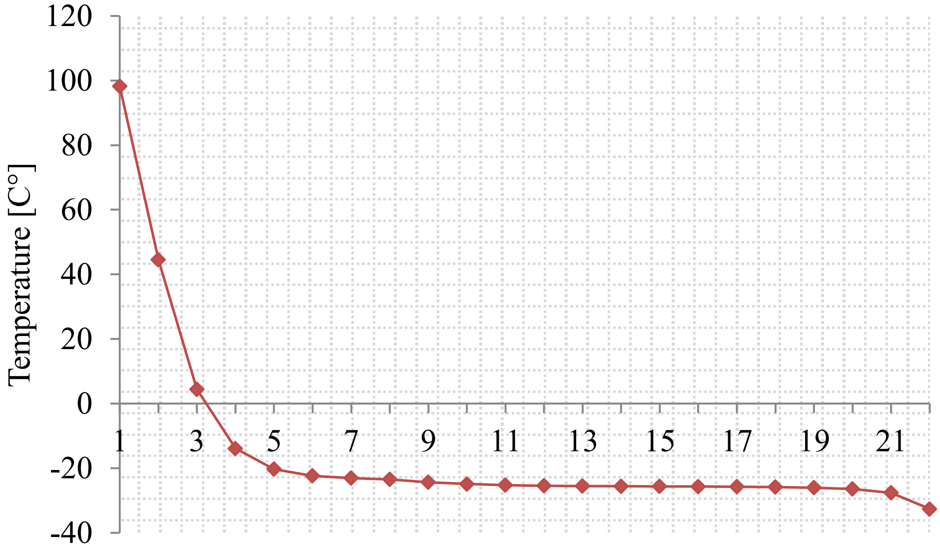

Figure 31 Temperature vs. Tray Position through the Scrub Column

7.8. Vapour/Liquid Flow

Figure 32 shows vapour and liquid flow in kgmole/h through the column from tray 0 to 21. As shown; at tray 8 and upwards, vapour flow is over 30 000 kgmole/h, while liquid flow is around 500 kgmole/h. Without C_4/C_5 reflux the vapour-liquid ratio increases, and after tray 8 and upwards towards the top, the ratio is highest. As for existing facilities this area is vulnerable for changes and without reflux the vapour- liquid rate ratio is higher through the entire column. This exposes the column for higher risk of flooding. However, as presented in table 16, flooding is measured at a maximum 73%. The column still operates within flooding range, and the simulations give results without system warnings.

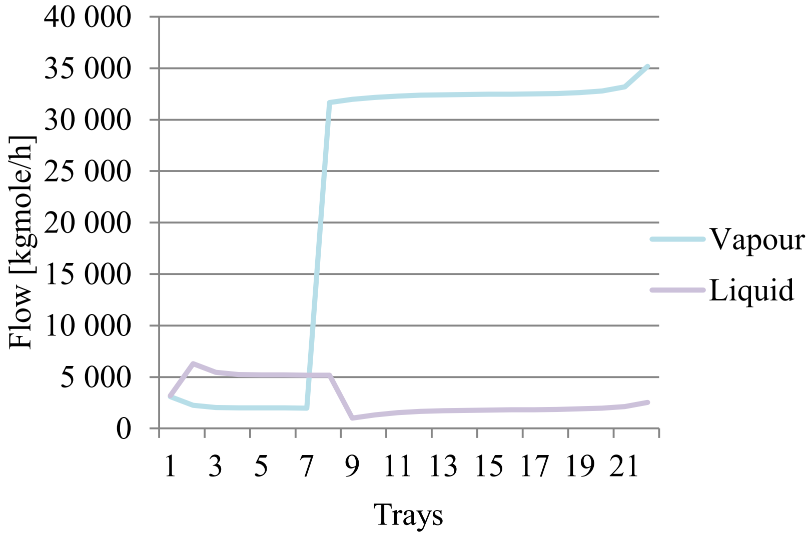

Figure 32 Vapour and Liquid flow through the Scrub Column

7.9. Energy Balance

Energy use in the column is as presented in table 18. Energy demand in condenser is higher than for the reboiler. This is a natural consequence of C_4/C_5 reflux removal, since less energy is needed to vaporize heavy hydrocarbons in the inlet gas stream. Similarly, boil up stream is less and energy requirements are lower in the reboiler.

Table 18 Energy Use in Reboiler and Condenser

	[kW]
Reboiler	16 380
External condenser	17 710

In the simulations, the condenser duty depends on inlet composition and outlet temperature. Outlet temperature can be manipulated, but are specified similar as for existing facilities, respectively -50 °C. Therefore, condenser duty will directly depend on inlet composition. In general a lean gas composition requires lower temperature and pressure to liquefy than a rich gas. In this modified case, the composition contains more heavy hydrocarbons than existing facilities. A consequence of this is as followed; Dew point temperature increases, less cooling medium are required to liquefy the gas hence less energy duty in the condenser.

8. Feed Flow Flexibility

To address the question whether an increased max GHV specification can be used to allow a more LPG rich feed in the simplified facilities, five new cases are calculated, completed and considered. These cases are constructed with different LPG amount in feed.

Based on these various cases, this chapter compare results in mass balances, LPG recovery, hydraulic results and energy use. These results will be used in the assessment on feed gas flexibility.

8.1. Feed Gases

Various LPG increase in the feed flow, is presented in table 19. LPG is increased from 20% to 200%, which makes a LPG feed increase on 27 0929 kg/h to 241 857 kg/h.

Table 19 Feed composition [mole%] with LPG increase

		LPG Increase [%]				
	Case 2	Case A 20%	Case B 60%	Case C 80%	Case D 100%	Case E 200%
Nitrogen	1.01	0.99	0.97	0.96	0.95	0.89
Methane	79.98	79.10	77.38	76.55	75.73	70.56
Ethane	7.47	7.39	7.22	7.15	7.07	7.03
Propane	3.01	3.57	4.66	5.18	5.70	8.00
i-Butane	2.01	2.38	3.11	3.46	3.80	6.40
n- Butane	0.60	0.71	0.93	1.04	1.14	1.91
i-Pentane	1.95	1.93	1.89	1.87	1.85	1.72
n-Pentane	1.95	1.93	1.89	1.87	1.85	1.72
n-Hexane	1.95	1.93	1.89	1.87	1.85	1.72
n-Heptane	0.03	0.03	0.03	0.03	0.03	0.03
Octane	0.01	0.01	0.01	0.01	0.01	0.01
Benzene	0.01	0.01	0.01	0.01	0.01	0.01
Toluene	0.01	0.01	0.01	0.01	0.01	0.01
LPG feed Increase [kgmole/h]	-	394	1186	1583	1978	4520
LPG feed increase [kg/h]	-	27 929	68 002	88 127	108 089	241 857

8.2. Adjustments

All cases are simulated with increased max GHV specification on 41 MJ/sm^3. Since heating value depends on flow composition, heating value increases when more LPG ends up in the LNG product. As presented in chapter 5.8 there are several parameters to manipulate to achieve desired results.

One adjustment which effect GHV results is the temperature in the top, prior to the reflux drum. As mentioned, by changing this parameter, amount of condensed hydrocarbons change. By increasing this temperature, fewer hydrocarbons condense and internal reflux decreases. When separating the flow, heavy hydrocarbons will follow the top product instead. Therefore, to achieve a GHV specification on 41 MJ/sm^3 with increased LPG in feed gas, the temperature has to be adjusted upwards.

To achieve convergence a new adjustment is done for Case D and E. Since LPG stream in feed gas is doubled, the column must operate under other specifications to function. To achieve convergence the methane composition in the bottom product is increased. As a result, more LPG follows the bottom product, which lowers the heating value in top product.

However, increasing methane also has its limitations. Since the fractionation system is an integrated and complex design system, it is sensitive for composition changes. If the amount of methane in the bottom is too high, there might be need for a demethanizer to extract additional methane.

8.3. Mass Balances

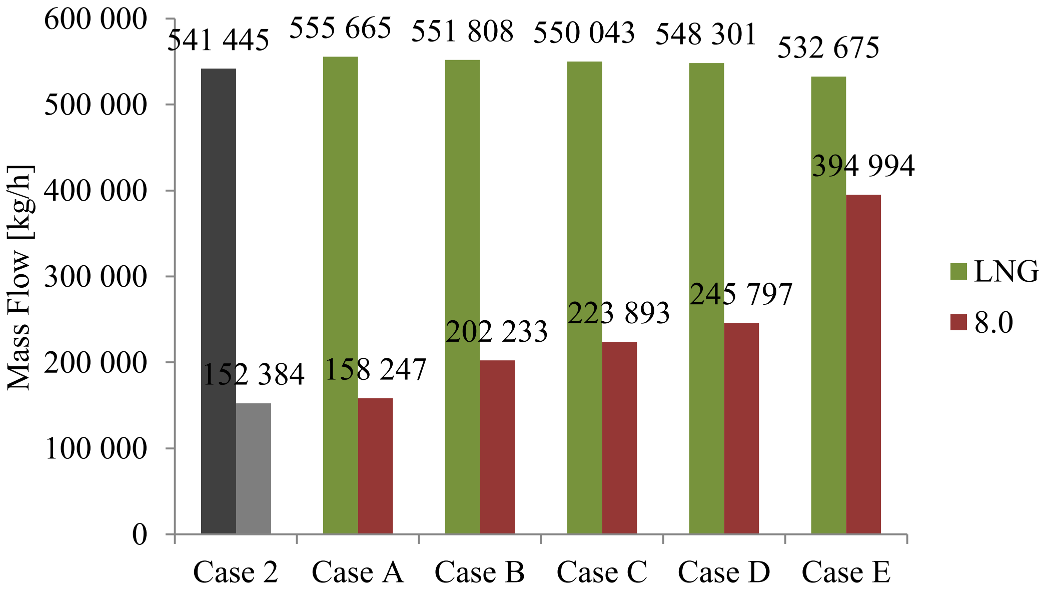

Figure 33 Mass Balance in Scrub Column

Figure 33 shows mass flows in the LNG product and in the bottom product, where black and grey column show flows before increasing LPG in feed. As shown on the figure the LNG product increases from 541 tonnes/h to more than 555 tonnes/h. This difference counts for added propane and butane in the LNG product. All values can be found in attachment IIII, A-H.

Furthermore, when increasing LPG in the feed, the LNG product remains somewhat similar despite of LPG feed increase. Correspondingly, bottom product, stream 8.0 increases. From this, it can be shown that most of the LPG increase will follow the bottom product. For Case E, where LPG in feed is increased with a factor two, the bottom product amount is almost the same as top product, where nearly 400 tonnes/h exit at the bottom.

8.4. LPG Bottom Recovery

Another way of viewing the scrub column performance is by comparing LPG Recovery when increasing LPG in feed gas composition. As shown on figure 34, recovery increases for both propane and butane. For Case E butane recovery is as high as 87%, which means that butane is only partly present in the top product. Same results in mass and molar flows are given in attachment III, D-H.

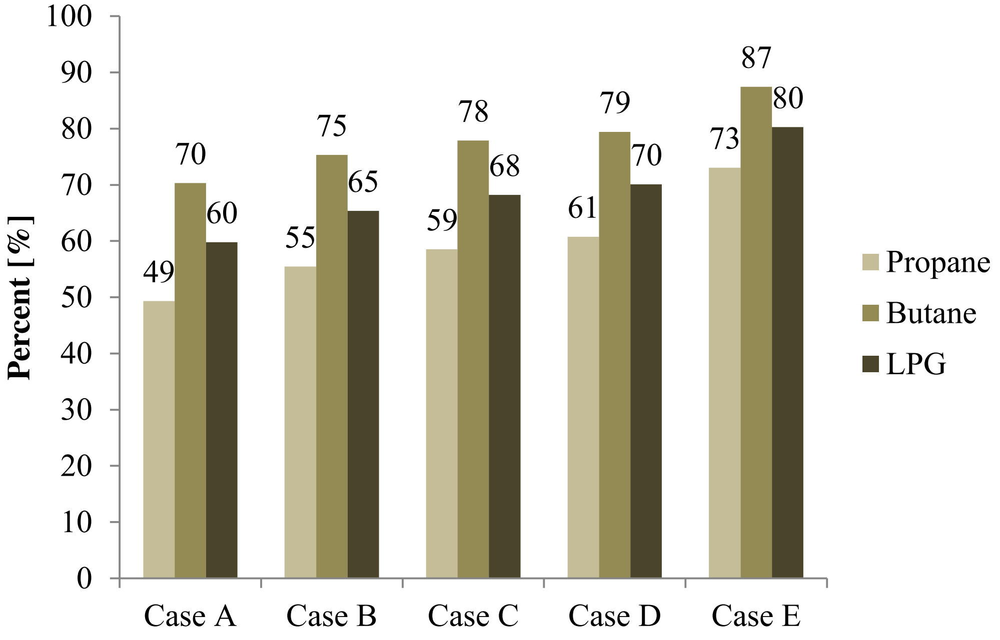

Figure 34 Propane, Butane and LPG Recovery in Bottom

8.5. Liquid and Vapour Flows

When adding LPG in the feed gas composition, the vapour/liquid fraction changes. As presented in chapter 4.3.5, limitations for the vapour/liquid fraction are set in the scrub column. Flooding can occur if there is great difference between vapour and liquid rate. Weeping can occur if vapour and liquid rate differences are small. Since Case A-E is constructed with a continuous liquid increase, operation limitations will be set by flooding. These challenges are highest at the top of the column and as shown on figure 35, flooding will limit the addition of LPG. Dotted red line on the figure represents the HYSYS flooding operation line, designed at 70%. As shown, case D and E exceed this line. Although, as mentioned in theoretical background, a scrub column can operate in a 60-80% flooding area, an LPG addition should not exceed immense amounts.

A result of the high liquid flow in the column is that head loss under downcomer is too large. This occurs due to oversizing of the column. If the head loss is low, liquid misdistribution occurs in the tray inlet. This leads to poor separation efficiency.

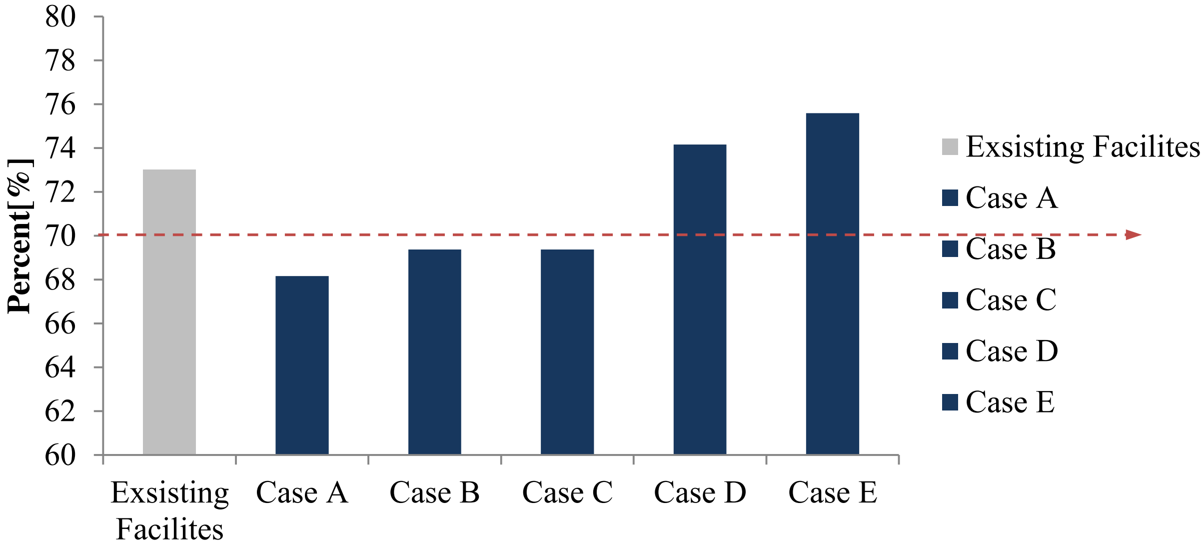

Figure 35 Flooding Probability

8.6. Energy Use

A consequence for increasing LPG in feed gas is a higher energy demand in the scrub column. As seen in table 20 and figure 36 energy use increases extensively in the reboiler, while hardly changes in the condenser. To cope with increasing bottom flow, temperature in reboiler must increase. By increasing temperature in reboiler, a bigger fraction of gas stream vaporizes and is recycled back to the column. To increase temperature more external heat must access the system. By supplementing additional hot oil, energy use increases and will ensure additional costs.

Additionally, for case E, amount of methane following the bottom product is increased and since the reboiler temperature is a variable depending on methane, energy use is noticeably higher.

Table 20 Energy Use in Case A-E [kW]

	Case A	Case B	Case C	Case D	Case E
Reboiler	17 170	21 610	25 410	28 320	46 190
Condenser	7 902	8 314	8 035	8 069	6 760

It should also be noted that energy use in condenser is evidently lower than in the reboiler. Since temperature drop before and after condensation is reduced, less cooling medium is required.

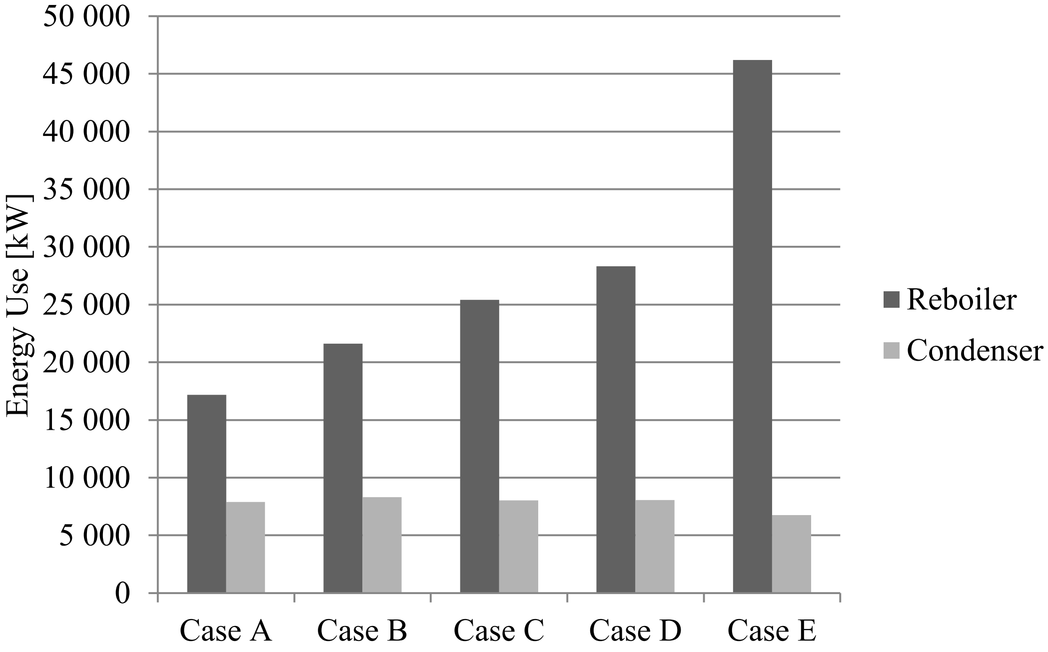

Figure 36 Energy Use in Case A-E

9. Discussion

9.1. GHV Control

As presented in literary survey, there are several ways to control GHV on the final product. Choosing an ideal method is based on important parameters such as separation efficiency, energy demand, installation work and gas flexibility.

By reducing the amount of nitrogen in the feed gas, heating value decreases. However, this requires additional installation and since heating value is much more dependent on LPG in feed gas, a better option to control heating value is by adjusting LPG amount in gas flow. The best way of GHV control and to lower the heating value, is by changing the amount of LPG through the LNG process. As mentioned, varying the LPG amount in the gas stream can be done both separately before liquefaction, integrated in the liquefaction step and front end. By installing an LPG extraction unit before liquefaction, such as a turbo-expander, a less complex LNG plant is obtained. For a developer of a gas-field, this leads to a clear split in responsibility for both design and operation. However, efficiency of installing a turbo-expander is limited and since this process requires high pressure and shows little room for flow flexibility it is not a preferable option.

The easiest way of manipulating and controlling the GHV is therefore by doing modification to already installed units. For an ordinary LNG plant this would be by manipulating the LPG recovery units. Without installing extra equipment, changing parameters in the scrub column, will affect the heating value. Both by varying reflux, plate inlet, pressure, reboiler and condenser temperature in a scrub column, separation varies and hence heating value on the LNG Product.

9.2. Effect of C4/C5 Reflux Removal

Two cases have been simulated in this thesis, with two different GHV specifications, 40 MJ/sm^3 and 41 MJ/sm^3 respectively. Results vary between these two models. By comparing results between the two models, effects of changing GHV specifications are apparent and will be discussed as followed.

The first case where designed with a GHV spefication on 40 MJ/sm^3. To achieve this result heavy hydrocarbons were extracted both front end at reception facilities and by a scrub column before the fractionation system. Since this case represented existing facilities, several system limits were already given. This reduced the degree of freedom when designing the simulation model.

At reception facilities the condensate stabilizer were designed to produce a condensate product which could not oversee a TVP at 1 bar by 37 °C. If lighter hydrocarbons such as propane and butane where to end up in the condensate product TVP would increase. But as mentioned in the literature survey, the TVP specification on the condensate product is important to achieve safe transportation without utility damage. Therefore adjustments done to the condensate stabilizer were limited.

When extracting hydrocarbons in the scrub column, both amounts of trays, feed locations, pressure, reboiler and condenser temperature could be manipulated to get an achieved max GHV at 40 MJ/sm^3. However, since feed flow composition where constructed based on literature examples and not actual existing facilities, this degree of freedom made it possible to achieve desired results without manipulating the scrub column parameters. Amount of trays, feed locations, feed flows and temperatures are therefore similar to existing facilities.

Results and relations between parameters from this case present a view of how GHV is controlled in a typical LNG plants and how specifications on the sales product plays a vital role when designing pre-treatment facilities. There is a fine balance between capital cost, production efficiency and operability of the units. Since treatment facilities consume a large amount of energy to liquefy the gas, optimizing the process efficiency is of great importance. Improving the efficiency of the process by limiting the natural gas conversion into power, will bring additional LNG to be sold and consequently additional revenues.

The second case, case 2, is designed based on a GHV specification at 41 MJ/sm^3, and the simplest and most efficient way to achieve this specification, is to change parameters in the scrub column. Process steps prior to the scrub column remain similar as to case 1 and it is assumed that a similar amount of condensate is extracted from the condensate stabilizer.

Since the column where designed with constant pressure and constant amount of trays, modification where done by removing the C_4/C_5 reflux. In case 1, this reflux assisted the scrub column operation. The reflux originally entered at the top of the scrub column and improved separation between vapour and liquid. When the C_4/C_5 reflux where removed, separation efficiency decreased and more heavy hydrocarbons followed the top product. A consequence of this adjustment, is a higher heating value in the LNG product.

The heating value did not reach maximum GHV specification, but increased to 40.45 MJ/sm^3. If additional parameters in the scrub column where changed, the GHV could become 41 MJ/sm^3. If the condenser temperature has been adjusted upwards, fewer hydrocarbons would liquefy and GHV would become higher in the LNG product. By changing this parameter internal reflux rate would also change, which again would affect scrub column performance. For comparison reasons, this temperature where held constant similar to case 1 at -50 °C.

The effect on removing this reflux can also be viewed in table 21, where results from existing facilities and simplified facilities, respectively Case 1 and 2, are compared.

Table 21 Comparison between Case 1 and Case 2

	Case 1 Existing Facilities	**Case 2** Simplification of Existing Facilities
GHV [MJ/sm^3]	40.00	40.45
LNG product [kmole/h]	30 250	30 670
Flooding probality [%]	70.49	73.66
LPG Recovery [%]	70	55
Energy Use [kW]	38 240	34 090

When limitations on GHV specifications were increased, some benefits became apparent. Energy demand is considerably reduced. This is shown in table 21 above, where differences count for 10%. Main differences are in the reboiler. When doing modifications, bottom product is reduced. Less energy is therefore needed to heat up the bottom product. Vaporized steam recycled back to the column can also decrease, since there are fewer liquid flows in the enriching section.

This sort of energy reduction can induce major cost savings. Since the liquefaction process is the largest cost component of the LNG cost train and counts for almost 30-45% of total costs (Kidnay et al., 2011), modifications done prior or in the liquefaction process has great cost saving potential. Costs are high due to the large amounts of cryogenic material required, and since modification done in case 2 requires less hot oil in reboiler, operational costs will consequently become lower.

However, an issue worth mentioning is the risk of flooding. As seen on table 21, flooding probability increases when C_4/C_5 reflux is removed. This can be better explained on figure 37, where liquid and vapour flows in both cases are compared. As seen vapour/liquid ratio is biggest at tray nine and upwards for both cases. However, at the top, where reflux C_4/C_5 initially enters, ratio between vapour and liquid is higher for case 2. Trays above feed and upwards are more vulnerable for mechanical problems. HYSYS operates with a limit 70%, but operation can normally be in area 60-80%, which means case 2 is still in a functioning area.

If flooding occurs, the scrub column performance decline and the column could become less operative.

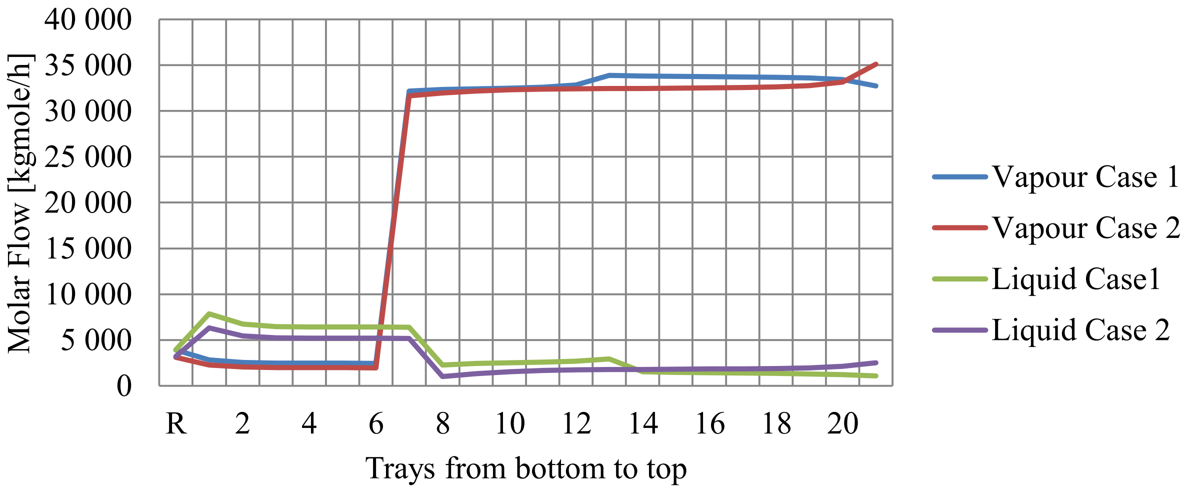

Figure 37 Vapour and Liquid Flows Case 1 and Case 2

When mechanical issues considered, results of plant design simplifications show great potential for including associated gas to pure gas fields. By increasing GHV specification and simplifying the scrub column installation, it would reduce operational costs, and achieve higher gains from operative petroleum fields.

9.3. Feed Gas Flexibility

By design, plants can handle different feed gas compositions. For short-term changes, such as interruption to upstream facilities, a degree of impact on production may be acceptable so long as product quality can be maintained. However, for long term operation changes, such as tie-ins of associated gas, measures may have to be taken to meet production requirements without compromising on quality.

To address whether simplifying the process and accepting a heating value on 41 MJ/sm^3 can allow more LPG rich feed to the LNG Plant, chapter 8 show results when simulation are done with various feed gases, from lean to rich gas respectively.

When richer gas is added, some design parameters are changed. Allthough the amount of trays and feed gas inlet locations remained constant, the condenser temperature were adjusted upwards to achieve a heating value up to 41 MJ/sm^3. When this temperature were adjusted, internal upper reflux were reduced and hence the temperature profile and bottom flows to the fractionation system increased. This adjustment reduces risk of bottlenecking the fractionation system.

Based on mass balances and LPG Recovery, the system showed great room for increasing LPG in feed gas. When increasing LPG in the feed stream from 20% and upwards, LPG bottom recovery improved as well. With a 20% increase of LPG in feed gas, LPG bottom recovery is correspondingly over 60%, while an 80% increase gives a LPG bottom recovery on 68%. Most of LPG added in feed gas will follow the bottom product.

However, as presented in chapter 8.5, LPG-increase is limited due to mechanical design and hydraulic performance. Since liquid flow rate increases, the probability increases for flooding. For case D and E, where LPG is increased 100% and 200% from original LPG feed, head loss under downcomer is too large. As presented in chapter 4.3.5, this problem occurs when the downcomer no longer can accommodate the liquid leaving the plate. This can cause loss of separation and negatively impact the performance and energy efficiency of the distillation process.

Other factors limiting LPG increase is bottom product composition. After leaving the distillation column, the bottom products enter the fractionation system. This integrated system is complex designed and is highly dependent on feed inlet composition. When increasing LPG in the scrub column feed inlet, design parameters must change. As presented, the column variables are methane leaving the column, and to achieve convergence for Case D and E, this variable must change. By increasing the amount methane following the bottom product, the column converges. However, if the methane concentration is too immense before entering the fractionation system, it can create problems for the fractionation system. One way of coping with this problem is as mentioned, by installing a Demethanizer prior to the original fractionation system. But when regarding cost efficiency and simplicity this alternative becomes excessive.

As presented in results, energy use increase from Case A to Case E. When the LPG amount increases in feed flow, heat required to vaporize the bottom product in reboiler also increases. For case E, where LPG in feed is increased with a factor three, energy use is 2.8 times higher. Based on the argument that a plant simplification should be energy saving and cost saving, option C and D becomes irrelevant. Additionally, by adjusting temperature after condenser upwards, less cooling medium is used, reducing energy demand. However, this temperature adjustment only varies energy use in a small degree.

Based on presented cases and comparisons an ideal LPG increase should result in a good LPG recovery, while avoiding challenging the scrub column performance. Energy use should also be kept relatively low, to maintain efficiency and cost gains from modifications. Based on these issues case D and E are excluded, and case A and B standout as ideal alternatives. These cases have a LPG feed increase from original feed gas of 20% and 60%, or 27 929 kg/h and 68 002 kg/h, respectively.

A summary of advantages and disadvantages can also be viewed in table 22.

Table 22 Case Comparisons

	Case A	Case B	Case C	Case D + E
Advantages	- No risk for flooding - Low energy use in reboiler	- Within flooding range - Increased LPG Recovery	- Good LPG bottom recovery	- Excellent LPG bottom recovery
Disadvantages	- Low LPG recovery	- Increased energy use	- increased methane in bottom - head loss under downcomer is too large	- High energy use - head loss under downcomer is too large

A new GHV specification and a system simplification show potential for increasing heavy hydrocarbons in a feed stream. Including associated gas from exsisting oil fieds, provides better utilization of oil fields, which again can increase petroleum revenues.

As a final note it should be commented that getting conclusive results by using HYSYS is limited. When simulating a plant similar to an existing plant, several simplifying assumptions where made. This can give results different to actual behaviour. Additionally, mechanical design of a scrub column is very complex. As presented in the literature survey, results depends of several parameters. Since all simulations are done with HYSYS standard design values, size parameters might not correlate to actual sizes. Mechanical and hydraulic results are therefore more indicative that tangible, and must be evaluated accordingly. As mentioned previously, condenser temperature and methane compositions are also two degrees of freedom, which are manipulated to get desired results.

10. Conclusion and Recommendations

During the last years, associated gas has become more of an issue in Norwegian Petroleum production. The energy efficiency and profitability can be increased by including gas from oil fields. The challenges faced, however, are the GHV specification defined by the international markets and wether the fractionation system can handle richer gas.

Based on these conditions, this thesis has presented a description of GHV control. This is done by a literature survey and by a HYSYS simulation process. First a process similar to existing facilities is simulated. Next simplifications of the process are done and lastly a feed flow flexibility assessment is performed. Conclusive arguments are presented as followed.

- One way of controlling the heating value is by controlling the amount of heavy hydrocarbons in the LNG product. A description on how GHV is controlled in a typical LNG plant and methods used to allow current heating value is presented in the literature survey and simulated in HYSYS. Results from simulations are presented in chapter 6, where a heating value of 40 MJ/sm^3 is achieved by adjusting the scrub column and controlling LPG following the bottom product. By extracting heavy hydrocarbons front end in a Condensate Stabilizer and in a Scrub Column, heating value is controlled. This gives a LNG end product of 529 tonnes/h or about 17.1 million sm^3/day. 70% of the LPG is extracted in Scrub Column, responding to 3916 sm^3/day.

- The higher concentration of heavy hydrocarbons, the higher heating value becomes. So when GHV is increased to 41 MJ/sm^3 in the LNG product, amount of heavy hydrocarbons in the product must increase. This is done by simplifying the simulated process. More specifically C_4/C_5 Reflux is removed. Separation efficiency decreases in the scrub column and instead of heavy hydrocarbons following the bottom product, they are included in the LNG product. LPG bottom recovery is reduced to 55%. This gives a LNG product on 541 tonnes/h or about 17.4 million sm^3/day. The added amount counts for heavy hydrocarbons, mainly LPG, which increases heating value.

- An assessment of how the increased max GHV specification can be done to improve feed gas flexibility is considered in chapter 8 where five different feed gas compositions are introduced. These cases hardly change LNG product, since most of added LPG follows the bottom product. However, the scrub column performance decreases when LPG added exceed 80%, due to higher risk for flooding. Energy use also increase for added LPG, which limits possibilities for adding LPG. Based on these limitations, an ideal increase would be up to 60% increased LPG to the feed gas. When exceeding an amount on 68 002 kg/h LPG to the feed gas, gains decline.

The results are based on several assumptions, which make it difficult to get fully conclusive results. However, despite uncertainties, all results indicate that GHV adjustments provide an opportunity to include associated gas in the future.

11. Further Work

During the work of this thesis several new concerns have appeared. In general both improvements and new assessments could be performed, to ease the decision making of adding associated gas to an existing LNG plant. Several aspects need to be evaluated and the following considerations represent tasks which could be further looked into:

- As mentioned in chapter 5, several simplifications are done to construct a simple process of the LNG pretreatment facilities prior to the fractionation system. To achieve more accurate results a process where these simplifications are included should be simulated and evaluated. Additional streams, which exist in actual plants, should be included. Water and sour gases normally exist in a feed gas and should also be included (and separated) in the process.

- Since simulations are performed with HYSYS standard values for the column, they will not be fully representative for an actual system. A preliminary design of the scrub column should therefore be considered. This would consist of a design method where amount of trays, tray inlet location, reboiler and condenser sizing should be performed. Sizing of column height, diameter, hole area and tray size should also be performed.

- There are other parameters in a scrub column varying the GHV on the end product. As mentioned in the literature survey, tray inlet location, amount of trays, internal reflux effect separation performance. It could be interesting to evaluate cases where these parameters have varied, to evaluate new possibilities for feed gas flexibility.

- Heavy hydrocarbons are also separated front end of the scrub column, and by doing adjustments to equipment before the scrub column, interesting results could be relevant. Adjustments done to the condensate stabilizer could increase separation performance and effect results on the LNG product.

- Challenges when adding additional heavy hydrocarbons to a feed gas are apparent in the fractionation system. Therefore, a natural next step when assessing feed gas flexibility would be to evaluate consequences for the fractionation system. i.e evaluating the bottom product leaving the scrub column to new distillation columns.

- This thesis has not included economic aspects, and although energy use is a good indication on expenses, it is not sufficient for decision making. Therefore it would be interesting performing an economic analysis when changes are done to the system. On this note, a more incisive gas market analysis could also contribute whether LPG increase is profitable.

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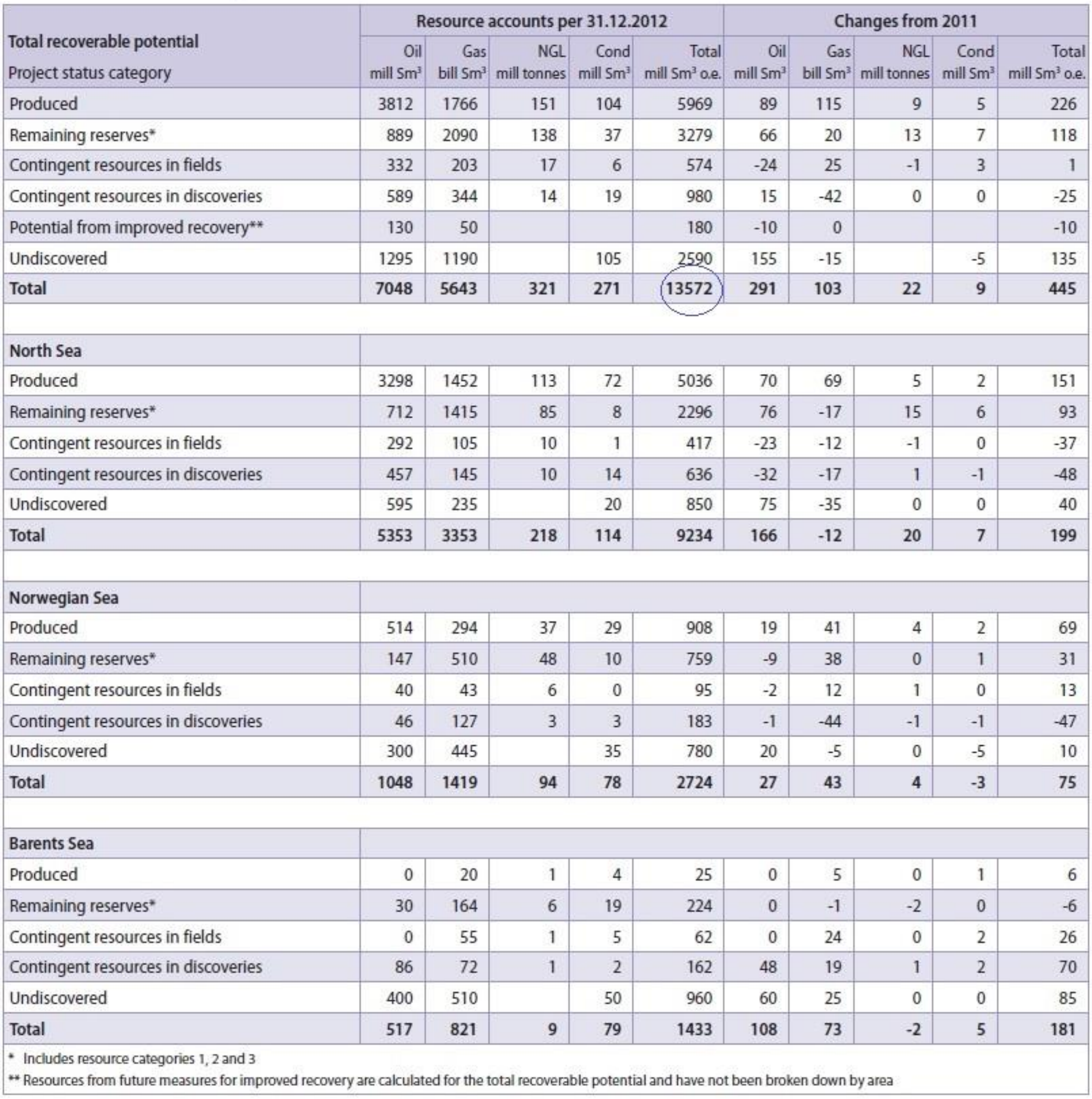
Attachements

I. Petroleum Resources and Reserves



Total recoverable potential	Resource accounts per 31.12.2012					Changes from 2011				
Project status category	Oil mill Sm^3	Gas bill Sm^3	NGL mill tonnes	Cond mill Sm^3	Total mill Sm^3 o.e.	Oil mill Sm^3	Gas bill Sm^3	NGL mill tonnes	Cond mill Sm^3	Total mill Sm^3 o.e.
Produced	3812	1766	151	104	5969	89	115	9	5	226
Remaining reserves*	889	2090	138	37	3279	66	20	13	7	118
Contingent resources in fields	332	203	17	6	574	-24	25	-1	3	1
Contingent resources in discoveries	589	344	14	19	980	15	-42	0	0	-25
Potential from improved recovery**	130	50			180	-10	0			-10
Undiscovered	1295	1190		105	2590	155	-15		-5	135
Total	**7048**	**5643**	**321**	**271**	**13572**	**291**	**103**	**22**	**9**	**445**
North Sea										
Produced	3298	1452	113	72	5036	70	69	5	2	151
Remaining reserves*	712	1415	85	8	2296	76	-17	15	6	93
Contingent resources in fields	292	105	10	1	417	-23	-12	-1	0	-37
Contingent resources in discoveries	457	145	10	14	636	-32	-17	1	-1	-48
Undiscovered	595	235		20	850	75	-35	0	0	40
Total	**5353**	**3353**	**218**	**114**	**9234**	**166**	**-12**	**20**	**7**	**199**
Norwegian Sea										
Produced	514	294	37	29	908	19	41	4	2	69
Remaining reserves*	147	510	48	10	759	-9	38	0	1	31
Contingent resources in fields	40	43	6	0	95	-2	12	1	0	13
Contingent resources in discoveries	46	127	3	3	183	-1	-44	-1	-1	-47
Undiscovered	300	445		35	780	20	-5	0	-5	10
Total	**1048**	**1419**	**94**	**78**	**2724**	**27**	**43**	**4**	**-3**	**75**
Barents Sea										
Produced	0	20	1	4	25	0	5	0	1	6
Remaining reserves*	30	164	6	19	224	0	-1	-2	0	-6
Contingent resources in fields	0	55	1	5	62	0	24	0	2	26
Contingent resources in discoveries	86	72	1	2	162	48	19	1	2	70
Undiscovered	400	510		50	960	60	25	0	0	85
Total	**517**	**821**	**9**	**79**	**1433**	**108**	**73**	**-2**	**5**	**181**

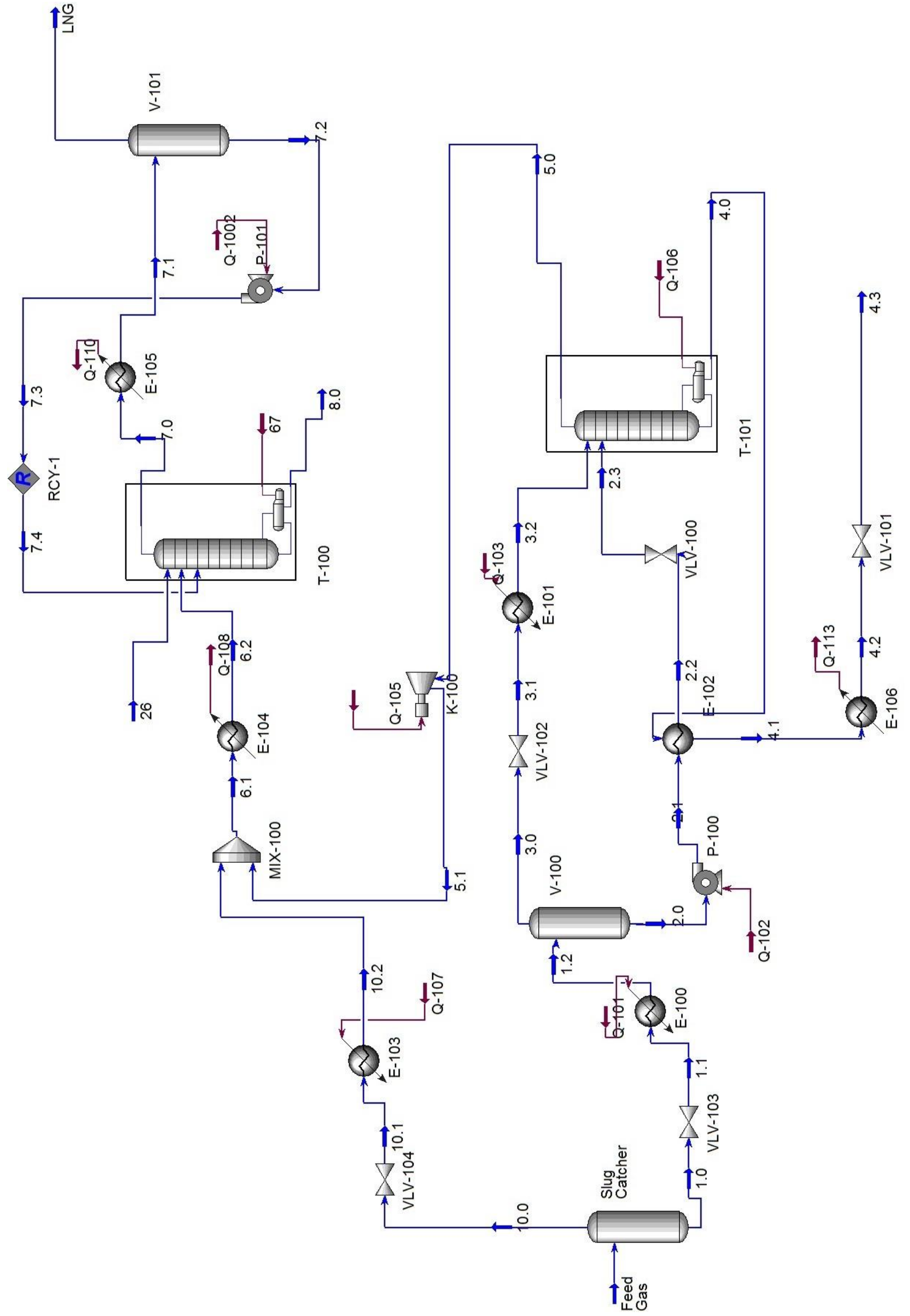
* Includes resource categories 1, 2 and 3

** Resources from future measures for improved recovery are calculated for the total recoverable potential and have not been broken down by area

Circled area: Total estimated resources reffered in chapter 1.1

II. Case 1 Existing Facilities

A. figure



B. Mole Compositions [-]

	Feed Gas	1.0	1.1	1.2	2.0	2.1	2.2	2.3
Nitrogen	0,010	0,0019	0,0019	0,0019	0,0001	0,0001	0,0001	0,0001
Methane	0,800	0,3785	0,3785	0,3785	0,0509	0,0509	0,0509	0,0509
Ethane	0,075	0,1127	0,1127	0,1127	0,0557	0,0557	0,0557	0,0557
Propane	0,030	0,0862	0,0862	0,0862	0,0895	0,0895	0,0895	0,0895
i-Butane	0,020	0,0784	0,0784	0,0784	0,1163	0,1163	0,1163	0,1163
n-Butane	0,006	0,0259	0,0259	0,0259	0,0422	0,0422	0,0422	0,0422
i-Pentane	0,020	0,0984	0,0984	0,0984	0,1900	0,1900	0,1900	0,1900
n-Pentane	0,020	0,1024	0,1024	0,1024	0,2044	0,2044	0,2044	0,2044
n-Hexane	0,020	0,1120	0,1120	0,1120	0,2427	0,2427	0,2427	0,2427
n-Heptane	0,000	0,0018	0,0018	0,0018	0,0040	0,0040	0,0040	0,0040
n-Octane	0,000	0,0006	0,0006	0,0006	0,0014	0,0014	0,0014	0,0014
Benzene	0,000	0,0006	0,0006	0,0006	0,0013	0,0013	0,0013	0,0013
Toluene	0,000	0,0006	0,0006	0,0006	0,0014	0,0014	0,0014	0,0014
Total	1,000	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000

	3.0	3.1	3.2	4.0	4.1	4.2	4.3	5.0
Nitrogen	0,0033	0,0033	0,0033	0,0000	0,0000	0,0000	0,0000	0,0025
Methane	0,6345	0,6345	0,6345	0,0000	0,0000	0,0000	0,0000	0,5009
Ethane	0,1572	0,1572	0,1572	0,0000	0,0000	0,0000	0,0000	0,1491
Propane	0,0836	0,0836	0,0836	0,0002	0,0002	0,0002	0,0002	0,1140
i-Butane	0,0488	0,0488	0,0488	0,0247	0,0247	0,0247	0,0247	0,0958
n-Butane	0,0132	0,0132	0,0132	0,0228	0,0228	0,0228	0,0228	0,0269
i-Pentane	0,0268	0,0268	0,0268	0,2483	0,2483	0,2483	0,2483	0,0499
n-Pentane	0,0226	0,0226	0,0226	0,2861	0,2861	0,2861	0,2861	0,0430
n-Hexane	0,0099	0,0099	0,0099	0,4038	0,4038	0,4038	0,4038	0,0177
n-Heptane	0,0001	0,0001	0,0001	0,0071	0,0071	0,0071	0,0071	0,0001
n-Octane	0,0000	0,0000	0,0000	0,0025	0,0025	0,0025	0,0025	0,0000
Benzene	0,0001	0,0001	0,0001	0,0021	0,0021	0,0021	0,0021	0,0001
Toluene	0,0000	0,0000	0,0000	0,0024	0,0024	0,0024	0,0024	0,0000
Total	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000

	5.1	**10.0**	**10.1**	**10.2**	**6.1**	**6.2**	**26**	**8.0**
Nitrogen	0,003	0,0116	0,0116	0,0116	0,0105	0,0105	0,0000	0,0000
Methane	0,501	0,8817	0,8817	0,8817	0,8329	0,8329	0,0000	0,1400
Ethane	0,149	0,0673	0,0673	0,0673	0,0778	0,0778	0,0000	0,2022
Propane	0,114	0,0192	0,0192	0,0192	0,0313	0,0313	0,0090	0,1922
i-Butane	0,096	0,0088	0,0088	0,0088	0,0199	0,0199	0,0993	0,1714
n-Butane	0,027	0,0022	0,0022	0,0022	0,0053	0,0053	0,4716	0,0703
i-Pentane	0,050	0,0042	0,0042	0,0042	0,0101	0,0101	0,2670	0,1068
n-Pentane	0,043	0,0034	0,0034	0,0034	0,0085	0,0085	0,1510	0,0853
n-Hexane	0,018	0,0016	0,0016	0,0016	0,0036	0,0036	0,0020	0,0314
n-Heptane	0,000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0001	0,0002
n-Octane	0,000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Benzene	0,000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0002
Toluene	0,000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Total	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

	7.0	**7.1**	**7.2**	**7.3**	**7.4**	**LNG**
Nitrogen	0,0111	0,0111	0,0035	0,0035	0,0035	0,0117
Methane	0,8939	0,8939	0,6543	0,6543	0,6541	0,9135
Ethane	0,0691	0,0691	0,1729	0,1729	0,1726	0,0607
Propane	0,0153	0,0153	0,0787	0,0787	0,0788	0,0101
i-Butane	0,0024	0,0024	0,0176	0,0176	0,0178	0,0011
n-Butane	0,0056	0,0056	0,0469	0,0469	0,0471	0,0023
i-Pentane	0,0018	0,0018	0,0178	0,0178	0,0179	0,0004
n-Pentane	0,0008	0,0008	0,0082	0,0082	0,0082	0,0002
n-Hexane	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Heptane	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Octane	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Benzene	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Toluene	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Total	1,0000	1,0000	1,0000	1,0000	1,0000	1,0000

C. Molar Flows [kmole/h]

	Feed Gas	**1.0**	**1.1**	**1.2**	**2.0**	**2.1**	**2.2**	**2.3**
Nitrogen	354	11	11	11	0	0	0	0
Methane	28 189	2 170	2 170	2 170	128	128	128	128
Ethane	2 632	646	646	646	140	140	140	140
Propane	1 061	494	494	494	225	225	225	225
i-Butane	709	449	449	449	292	292	292	292
n-Butane	213	149	149	149	106	106	106	106
i-Pentane	688	564	564	564	478	478	478	478
n-Pentane	688	587	587	587	514	514	514	514
n-Hexane	688	642	642	642	610	610	610	610
n-Heptane	11	10	10	10	10	10	10	10
n-Octane	4	3	3	3	3	3	3	3
Benzene	4	3	3	3	3	3	3	3
Toluene	4	3	3	3	3	3	3	3
Total	35 243	5 732	5 732	5 732	2 514	2 514	2 514	2 514

	3.0	**3.1**	**3.2**	**4.0**	**4.1**	**4.2**	**4.3**	**5.0**
Nitrogen	11	11	11	0	0	0	0	11
Methane	2 042	2 042	2 042	0	0	0	0	2 170
Ethane	506	506	506	0	0	0	0	646
Propane	269	269	269	0	0	0	0	494
i-Butane	157	157	157	35	35	35	35	415
n-Butane	42	42	42	32	32	32	32	117
i-Pentane	86	86	86	348	348	348	348	216
n-Pentane	73	73	73	401	401	401	401	186
n-Hexane	32	32	32	565	565	565	565	77
n-Heptane	0	0	0	10	10	10	10	0
n-Octane	0	0	0	3	3	3	3	0
Benzene	0	0	0	3	3	3	3	0
Toluene	0	0	0	3	3	3	3	0
Total	3 218	3 218	3 218	1 400	1 400	1 400	1 400	4 332

	5.1	10.0	10.1	10.2	6.1	6.2	26	8.0
Nitrogen	11	343	343	343	354	354	0	0
Methane	2 170	26 019	26 019	26 019	28 189	28 189	0	551
Ethane	646	1 986	1 986	1 986	2 632	2 632	0	796
Propane	494	566	566	566	1 060	1 060	3	757
i-Butane	415	259	259	259	674	674	35	674
n-Butane	117	64	64	64	181	181	164	277
i-Pentane	216	124	124	124	341	341	93	420
n-Pentane	186	101	101	101	288	288	53	336
n-Hexane	77	46	46	46	123	123	1	124
n-Heptane	0	0	0	0	1	1	0	1
n-Octane	0	0	0	0	0	0	0	0
Benzene	0	0	0	0	1	1	0	1
Toluene	0	0	0	0	0	0	0	0
Total	4 332	29 511	29 511	29 511	33 843	33 843	348	3 936

	7.0	7.1	7.2	7.3	7.4	LNG
Nitrogen	363	363	9	9	9	354
Methane	29 252	29 252	1 615	1 615	1 614	27 637
Ethane	2 262	2 262	427	427	426	1 835
Propane	501	501	194	194	194	307
i-Butane	78	78	44	44	44	34
n-Butane	184	184	116	116	116	69
i-Pentane	57	57	44	44	44	13
n-Pentane	25	25	20	20	20	5
n-Hexane	0	0	0	0	0	0
n-Heptane	0	0	0	0	0	0
n-Octane	0	0	0	0	0	0
Benzene	0	0	0	0	0	0
Toluene	0	0	0	0	0	0
Total	32 723	32 723	2 469	2 469	2 468	30 254

D. Mass Flows [kg/h]

	Feed Gas	**1.0**	**1.1**	**1.2**	**2.0**	**2.1**	**2.2**	**2.3**
Nitrogen	9 925	308	308	308	7	7	7	7
Methane	452 233	34 810	34 810	34 810	2 053	2 053	2 053	2 053
Ethane	79 138	19 421	19 421	19 421	4 214	4 214	4 214	4 214
Propane	46 766	21 787	21 787	21 787	9 928	9 928	9 928	9 928
i-Butane	41 187	26 124	26 124	26 124	16 990	16 990	16 990	16 990
n-Butane	12 371	8 632	8 632	8 632	6 168	6 168	6 168	6 168
i-Pentane	49 654	40 685	40 685	40 685	34 470	34 470	34 470	34 470
n-Pentane	49 654	42 337	42 337	42 337	37 086	37 086	37 086	37 086
n-Hexane	59 308	55 334	55 334	55 334	52 595	52 595	52 595	52 595
n-Heptane	1 066	1 034	1 034	1 034	1 014	1 014	1 014	1 014
n-Octane	405	400	400	400	397	397	397	397
Benzene	277	260	260	260	247	247	247	247
Toluene	327	319	319	319	313	313	313	313
Total	802 314	251 449	251 449	251 449	165 481	165 481	165 481	165 481

	3.0	**3.1**	**3.2**	**4.0**	**4.1**	**4.2**	**4.3**	**5.0**
Nitrogen	301	301	301	0	0	0	0	308
Methane	32 757	32 757	32 757	0	0	0	0	34 810
Ethane	15 206	15 206	15 206	0	0	0	0	19 421
Propane	11 859	11 859	11 859	12	12	12	12	21 775
i-Butane	9 134	9 134	9 134	2 010	2 010	2 010	2 010	24 114
n-Butane	2 465	2 465	2 465	1 859	1 859	1 859	1 859	6 774
i-Pentane	6 215	6 215	6 215	25 078	25 078	25 078	25 078	15 606
n-Pentane	5 251	5 251	5 251	28 897	28 897	28 897	28 897	13 440
n-Hexane	2 739	2 739	2 739	48 718	48 718	48 718	48 718	6 615
n-Heptane	20	20	20	1 003	1 003	1 003	1 003	31
n-Octane	3	3	3	398	398	398	398	2
Benzene	13	13	13	229	229	229	229	31
Toluene	5	5	5	311	311	311	311	7
Total	85 968	85 968	85 968	108 514	108 514	108 514	108 514	142 935

	5.1	**10.0**	**10.1**	**10.2**	**6.1**	**6.2**	**26**	**8.0**
Nitrogen	308	9 617	9 617	9 617	9 925	9 925	0	1
Methane	34 810	417 423	417 423	417 423	452 233	452 233	0	8 840
Ethane	19 421	59 718	59 718	59 718	79 138	79 138	0	23 925
Propane	21 775	24 979	24 979	24 979	46 754	46 754	138	33 362
i-Butane	24 114	15 063	15 063	15 063	39 177	39 177	2 009	39 201
n-Butane	6 774	3 739	3 739	3 739	10 513	10 513	9 545	16 091
i-Pentane	15 606	8 970	8 970	8 970	24 576	24 576	6 708	30 329
n-Pentane	13 440	7 318	7 318	7 318	20 758	20 758	3 794	24 212
n-Hexane	6 615	3 974	3 974	3 974	10 590	10 590	61	10 650
n-Heptane	31	32	32	32	63	63	2	65
n-Octane	2	5	5	5	7	7	0	7
Benzene	31	17	17	17	48	48	0	48
Toluene	7	8	8	8	16	16	0	16
Total	142 935	550 865	550 865	550 865	693 800	693 800	22 258	186 747

	7.0	**7.1**	**7.2**	**7.3**	**7.4**	**LNG**
Nitrogen	10 168	10 168	243	243	243	9 925
Methane	469 287	469 287	25 914	25 914	25 894	443 373
Ethane	68 020	68 020	12 836	12 836	12 807	55 185
Propane	22 102	22 102	8 568	8 568	8 572	13 533
i-Butane	4 536	4 536	2 532	2 532	2 551	2 004
n-Butane	10 719	10 719	6 731	6 731	6 752	3 989
i-Pentane	4 134	4 134	3 175	3 175	3 179	959
n-Pentane	1 806	1 806	1 465	1 465	1 466	342
n-Hexane	11	11	10	10	10	1
n-Heptane	0	0	0	0	0	0
n-Octane	0	0	0	0	0	0
Benzene	0	0	0	0	0	0
Toluene	0	0	0	0	0	0
Total	590 783	590 783	61 473	61 473	61 474	529 310

E. Conditions

	Feed Gas	1.0	1.1	1.2	2.0	2.1
Vapour/Phase Fraction [-]	0,84	0,00	0,38	0,56	0,00	0,00
Temperature [°C]	-1,00	-1,00	-19,96	40,00	40,00	45,85
Pressure [bar]	70,00	70,00	15,00	15,00	15,00	100,00
Std Ideal Liq. Flow[sm3/h]	2 183	494	494	494	275	275
Molar Enthalpy [kJ/kgmole]	-87 471	-127 580	-127 580	-119 240	-160 857	-159 577
Molar Entropy [kJ/kgmoleC]	136,66	103,63	107,74	137,13	97,27	98,41
Density [kg/m3]	4,65	11,71	80,31	46,30	582,56	590,87
GHV vol. bas [MJ/sm3]	50,75	94,95	94,95	94,95	143,68	143,68

	2.2	2.3	3.0	3.1	3.2	4.0
Vapour/Phase Fraction [-]	0,00	0,07	1,00	1,00	0,97	0,00
Temperature [°C]	80,00	76,89	40,00	40,00	29,00	156,05
Pressure [bar]	100,00	15,00	15,00	15,00	15,00	15,00
Std Ideal Liq. Flow[sm3/h]	275	275	219	219	219	169
Molar Enthalpy [kJ/kgmole]	-154 041	-154 041	-86 725	-86 725	-88 001	-156 495
Molar Entropy [kJ/kgmoleC]	114,88	117,80	168,28	168,28	164,13	148,53
Density [kg/m3]	552,77	282,08	16,70	16,70	17,89	472,39
GHV vol. bas [MJ/sm3]	143,68	143,68	59,23	59,23	59,23	171,87
TVP AT 37.5°C [bar]						**0**,9865

	4.1	4.2	4.3	5.0	5.1	10.0
Vapour/Phase Fraction [-]	0,00	0,00	0,00	1,00	1,00	1,00
Temperature [°C]	113,95	17,40	17,84	60,64	162,00	-1,00
Pressure [bar]	15,00	15,00	4,60	15,00	62,00	70,00
Std Ideal Liq. Flow[sm3/h]	169	169	169	325	325	1689
Molar Enthalpy [kJ/kgmole]	-166 435	-184 829	-184 829	-93 879	-87 975	-79 680
Molar Entropy [kJ/kgmoleC]	124,19	69,87	70,29	170,57	175,68	143,07
Density [kg/m3]	538,56	644,78	643,12	19,80	66,41	77,74
GHV vol. bas [MJ/sm3]	171,87	171,87	171,87	72,15	72,15	42,37

	10.1	**10.2**	**6.1**	**6.2**	**26**	**8.0**
Vapour/Phase Fraction [-]	1,00	1,00	1,00	0,88	0,00	0,00
Temperature [°C]	-1,00	15,00	43,42	-23,20	-33,80	98,11
Pressure [bar]	62,00	62,00	62,00	60,20	60,80	60,00
Std Ideal Liq. Flow[sm3/h]	1 689	1 689	2 014	2 014	37	363
Molar Enthalpy [kJ/kgmole]	-79 680	-78 591	-79 792	-84 444	-168 291	-119 564
Molar Entropy [kJ/kgmoleC]	143,07	147,75	152,52	136,09	33,54	139,88
Density [kg/m3]	77,74	59,41	57,70	92,03	658,36	356,65
GHV vol. bas [MJ/sm3]	42,37	42,37	46,14	46,14	139,13	102,72

	7.0	**7.1**	**7.2**	**7.3**	**7.4**	**LNG**
Vapour/Phase Fraction [-]	1	0,92	0,00	0,00	0,00	1,00
Temperature [°C]	-26,53	-50,00	-50,00	-50,00	-50,00	-50,00
Pressure [bar]	60	60,80	60,80	60,80	60,80	60,80
Std Ideal Liq. Flow[sm3/h]	1851	1851	163	163	163	1688
Molar Enthalpy [kJ/kgmole]	-79 829	-81 817	-96 827	-96 827	-96 855	-80 592
Molar Entropy [kJ/kgmoleC]	139,96	131,39	109,96	109,96	109,93	133,14
Density [kg/m3]	76,02	111,02	380,54	380,54	380,54	102,58
GHV vol. bas [MJ/sm3]	41,17	41,17	55,54	55,54	55,54	40,00

F. Energy Balances

	Q-101	**Q-102**	**Q-106 Reboiler**	**Q-113**	**Q-103**
Heat Flow [kJ/h]	47 801 800	3 219 340	44 679 043	25 752 515	-4 103 010
Power [kW]	13 280	899	12 410	7 153	-1 140

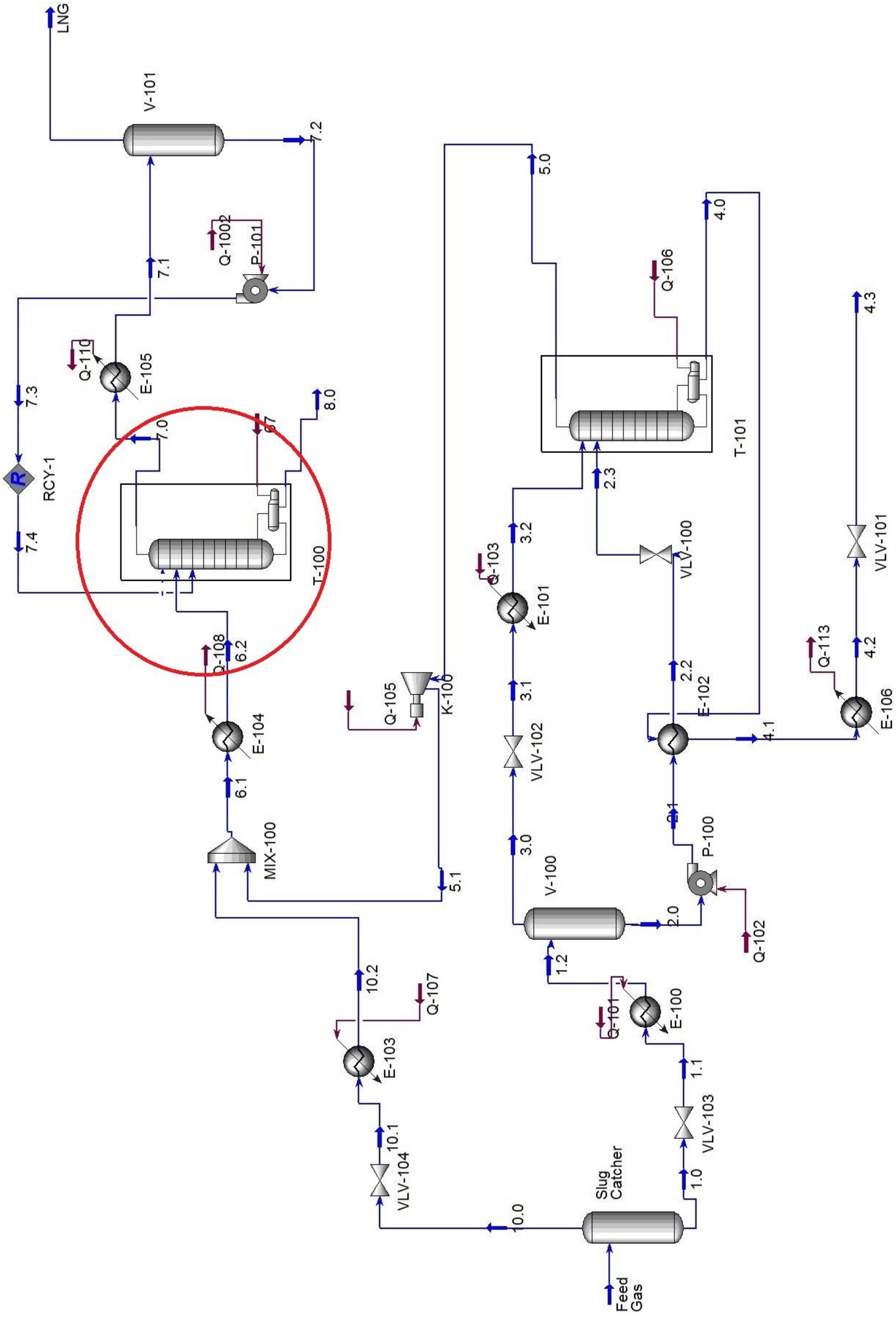
	Q-105	**Q-107**	**Q-108**	**Q-110 Condenser**	**67 Reboiler**
Heat Flow [kJ/h]	25 575 545	32 138 859	157 433 034	65 058 770	72 623 277
Power [kW]	7 140	8 927	43 730	18 070	20 170

G. Scrub Column Performance

Trays	Temperature [°C]	Pressure [bar]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]
21	-26,53	60	1 075	-
20	-27,95	60	1 231	33 450
19	-28,35	60	1 303	33 606
18	-28,53	60	1 347	33 678
17	-28,64	60	1 382	33 722
16	-28,72	60	1 417	33 757
15	-28,80	60	1 459	33 792
14	-28,90	60	1 518	33 834
13	-29,04	60	2 941	33 893
12	-25,75	60	2 694	32 848
11	-24,87	60	2 588	32 601
10	-24,53	60	2 525	32 495
9	-24,33	60	2 452	32 432
8	-24,11	60	2 261	32 359
7	-23,53	60	6 384	32 168
6	-23,13	60	6 409	2 448
5	-22,40	60	6 420	2 474
4	-20,25	60	6 427	2 484
3	-13,60	60	6 473	2 491
2	5,19	60	6 750	2 537
1	45,48	60	7 850	2 814
Reboiler	98,11	60	-	3 915

III. Case 2 Simplification of Exsisting Facilitites

A. Figure



B. Composition [-]

	7.4	6.2	7.0	8.0	LNG
Nitrogen	0,0041	0,0105	0,0106	0,0000	0,0116
Methane	0,6843	0,8329	0,8766	0,1401	0,9050
Ethane	0,1703	0,0778	0,0785	0,2012	0,0649
Propane	0,0969	0,0313	0,0254	0,1902	0,0148
i-Butane	0,0417	0,0199	0,0084	0,1778	0,0034
n-Butane	0,0028	0,0053	0,0005	0,0547	0,0002
i-Pentane	0,0000	0,0101	0,0000	0,1068	0,0000
n-Pentane	0,0000	0,0085	0,0000	0,0902	0,0000
n-Hexane	0,0000	0,0036	0,0000	0,0385	0,0000
n-Heptane	0,0000	0,0000	0,0000	0,0002	0,0000
n-Octane	0,0000	0,0000	0,0000	0,0000	0,0000
Benzene	0,0000	0,0000	0,0000	0,0002	0,0000
Toluene	0,0000	0,0000	0,0000	0,0001	0,0000
Total	1,0000	1,000	1,000	1,000	1,000

C. Molar Flows [kgmole/h]

	7.4	6.2	7.0	LNG
Nitrogen	18	354	373	354
Methane	3 102	28 189	30 844	27 745
Ethane	772	2 632	2 762	1 991
Propane	439	1 060	893	455
i-Butane	189	674	296	105
n-Butane	13	181	19	6
i-Pentane	0	341	0	0
n-Pentane	0	288	0	0
n-Hexane	0	123	0	0
n-Heptane	0	1	0	0
n-Octane	0	0	0	0
Benzene	0	1	0	0
Toluene	0	0	0	0
Total	4 533	33 843	35 186	30 657

D. Mass Flows [kg/h]

	7.4	6.2	7.0	8.0	LNG
Nitrogen	515	9 925	10 439	1	9 926
Methane	49 758	452 233	494 822	7 169	445 114
Ethane	23 217	79 138	83 057	19 298	59 870
Propane	19 360	46 754	39 359	26 756	20 075
i-Butane	10 988	39 181	17 212	32 958	6 128
n-Butane	728	10 515	1 108	10 135	332
i-Pentane	0	24 580	0	24 580	0
n-Pentane	0	20 761	0	20 761	0
n-Hexane	0	10 592	0	10 592	0
n-Heptane	0	63	0	63	0
n-Octane	0	7	0	7	0
Benzene	0	48	0	48	0
Toluene	0	16	0	16	0
Total	104 566	693 815	645 997	152 384	541 445

E. Conditions

	7.0	7.1	7.2	7.3
Vapour/Phase Fraction [-]	1,00	0,87	0,00	0,00
Temperature [°C]	-32,61	-50,00	-50,00	-50,00
Pressure [bar]	60,00	60,80	60,80	60,80
Std Ideal Liq. Flow[sm3/h]	2 009	2 009	291	291
Molar Enthalpy [kJ/kgmole]	-80 671	-82 485	-93 324	-93 291
Molar Entropy [kJ/kgmoleC]	138,48	130,57	114,44	114,47
Density [kg/m3]	85,48	119,80	341,00	340,68
GHV vol. bas [MJ/sm3]	41,82	41,82	51,82	51,79

	7.4	8.0	LNG
Vapour/Phase Fraction [-]	0,00	0,00	1
Temperature [°C]	-50,00	98,15	-50
Pressure [bar]	60,80	60,00	60,8
Std Ideal Liq. Flow[sm3/h]	291	296	1 718
Molar Enthalpy [kJ/kgmole]	-93 291	-120 192	-80 884
Molar Entropy [kJ/kgmoleC]	114,47	140,22	132,95676
Density [kg/m3]	340,68	356,95	106,46
GHV vol. bas [MJ/sm3]	51,79	103,41	40,35

F. Scrub Column Performance

Trays	Temperature [°C]	Pressure [bar]	Net Liquid [kgmole/h]	Net Vapour [kgmole/h]
21	-32,61	60	2 537	<empty>
20	-27,64	60	2 123	33 190
19	-26,45	60	1 971	32 777
18	-26,05	60	1 899	32 625
17	-25,87	60	1 858	32 553
16	-25,77	60	1 831	32 512
15	-25,70	60	1 808	32 484
14	-25,65	60	1 786	32 462
13	-25,60	60	1 759	32 440
12	-25,53	60	1 717	32 412
11	-25,43	60	1 645	32 370
10	-25,24	60	1 518	32 299
9	-24,91	60	1 307	32 172
8	-24,33	60	1 006	31 961
7	-23,45	60	5 154	31 660
6	-23,06	60	5 174	1 964
5	-22,36	60	5 182	1 985
4	-20,32	60	5 188	1 993
3	-13,97	60	5 222	1 998
2	4,28	60	5 433	2 032
1	44,30	60	6 269	2 243
Reboiler	98,15	60	<empty>	3 079

III. Feed Gas Flexibility

A. Feed Flow Composition [-]

Feed Gas	20% LPG Increase	60% LPG Increase	80% LPG Increase	100% LPG Increase	200% LPG Increase
Nitrogen	0,010	0,010	0,010	0,010	0,009
Methane	0,791	0,774	0,765	0,757	0,706
Ethane	0,074	0,072	0,071	0,071	0,070
Propane	0,036	0,047	0,052	0,057	0,080
i-Butane	0,024	0,031	0,035	0,038	0,064
n-Butane	0,007	0,009	0,010	0,011	0,019
i-Pentane	0,019	0,019	0,019	0,018	0,017
n-Pentane	0,019	0,019	0,019	0,018	0,017
n-Hexane	0,019	0,019	0,019	0,018	0,017
n-Heptane	0,000	0,000	0,000	0,000	0,000
n-Octane	0,000	0,000	0,000	0,000	0,000
Benzene	0,000	0,000	0,000	0,000	0,000
Toluene	0,000	0,000	0,000	0,000	0,000
Total	1,000	1,000	1,000	1,000	1,000

B. Feed Molar Flow [kgmole/h]

	20% LPG Increase	60% LPG Increase	80% LPG Increase	100% LPG Increase	200% LPG Increase
Nitrogen	354	354	354	354	353
Methane	28 189	28 189	28 189	28 189	28 055
Ethane	2 632	2 632	2 632	2 632	2 797
Propane	1 272	1 696	1 908	2 120	3 180
i-Butane	850	1 133	1 274	1 416	2 544
n-Butane	255	339	383	424	759
i-Pentane	688	688	688	688	685
n-Pentane	688	688	688	688	685
n-Hexane	688	688	688	688	685
n-Heptane	11	11	11	11	11
n-Octane	4	4	4	4	4
Benzene	4	4	4	4	4
Toluene	4	4	4	4	4
Total	35 637	36 429	36 826	37 221	39 763

C. Feed Mass Flow [kg/h]

	20% LPG Increase	60% LPG Increase	80% LPG Increase	100% LPG Increase	200% LPG Increase
Nitrogen	9 925	9 925	9 925	9 925	9 878
Methane	452 233	452 233	452 233	452 233	450 080
Ethane	79 138	79 138	79 138	79 138	84 094
Propane	56 091	74 789	84 137	93 486	140 228
i-Butane	49 382	65 843	74 073	82 304	147 867
n-Butane	14 801	19 716	22 261	24 645	44 116
i-Pentane	49 654	49 654	49 654	49 654	49 418
n-Pentane	49 654	49 654	49 654	49 654	49 418
n-Hexane	59 308	59 308	59 308	59 308	59 025
n-Heptane	1 066	1 066	1 066	1 066	1 061
n-Octane	405	405	405	405	403
Benzene	277	277	277	277	276
Toluene	327	327	327	327	325
Total	822 264	862 336	882 461	902 423	1 036 191

D. Increase Case A

		7.4			7.0	
20%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,003	3	71	0,011	357	9 996
Methane	0,531	512	8 214	0,886	28 235	452 964
Ethane	0,165	159	4 777	0,067	2 145	64 499
Propane	0,140	134	5 931	0,022	713	31 436
i-Butane	0,102	99	5 726	0,009	300	17 456
n-Butane	0,027	26	1 499	0,002	65	3 753
i-Pentane	0,024	24	1 696	0,001	41	2 994
n-Pentane	0,008	8	574	0,000	13	973
n-Hexane	0,000	0	0	0,000	0	0
n-Heptane	0,000	0	0	0,000	0	0
n-Octane	0,000	0	0	0,000	0	0
Benzene	0,000	0	0	0,000	0	0
Toluene	0,000	0	0	0,000	0	0
Total	1,000	964	28 488	1,000		584 070

		8.0			LNG	
20%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,000	0	1	0,011	354	9 925
Methane	0,140	466	7 483	0,897	27 725	444 789
Ethane	0,194	646	19 417	0,064	1 987	59 743
Propane	0,208	693	30 575	0,019	579	25 541
i-Butane	0,183	611	35 489	0,007	203	11 774
n-Butane	0,054	180	10 458	0,001	39	2 267
i-Pentane	0,098	326	23 552	0,001	18	1 275
n-Pentane	0,086	285	20 585	0,000	5	352
n-Hexane	0,037	123	10 557	0,000	0	0
n-Heptane	0,000	1	60	0,000	0	0
n-Octane	0,000	0	7	0,000	0	0
Benzene	0,000	1	48	0,000	0	0
Toluene	0,000	0	15	0,000	0	0
Total	1,000	3 332	158 247	1,000	30 909	555 665

E. Increase Case B

		7.4			7.0	
60%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,003	3	91	0,011	358	10 016
Methane	0,541	649	10 404	0,885	28 230	452 886
Ethane	0,154	185	5 553	0,063	2 024	60 866
Propane	0,157	188	8 290	0,026	839	37 007
i-Butane	0,106	127	7 385	0,011	339	19 677
n-Butane	0,027	32	1 852	0,002	71	4 134
i-Pentane	0,012	14	1 038	0,001	24	1 711
n-Pentane	0,002	2	140	0,000	3	224
n-Hexane	0,000	0	0	0,000	0	0
n-Heptane	0,000	0	0	0,000	0	0
n-Octane	0,000	0	0	0,000	0	0
Benzene	0,000	0	0	0,000	0	0
Toluene	0,000	0	0	0,000	0	0
Total	1,000	1 200	34 753	1,000	31 887	586 520

		8.0			LNG	
60%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,000	0	1	0,012	354	9 925
Methane	0,140	608	9 751	0,899	27 583	442 503
Ethane	0,182	792	23 825	0,060	1 840	55 324
Propane	0,241	1 045	46 060	0,021	652	28 741
i-Butane	0,203	881	51 233	0,007	212	12 315
n-Butane	0,059	258	14 983	0,001	39	2 285
i-Pentane	0,079	341	24 630	0,000	9	644
n-Pentane	0,068	295	21 250	0,000	1	71
n-Hexane	0,028	120	10 376	0,000	0	0
n-Heptane	0,000	1	55	0,000	0	0
n-Octane	0,000	0	6	0,000	0	0
Benzene	0,000	1	48	0,000	0	0
Toluene	0,000	0	14	0,000	0	0
Total	1,000	4 342	202 233	1,000	30 689	551 808

F. Increase Case C

	7.4			7.0		
80%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,003	3	87	0,011	357	10 011
Methane	0,537	615	9 865	0,886	28 126	451 230
Ethane	0,148	170	5 110	0,061	1 945	58 477
Propane	0,164	188	8 270	0,027	869	38 320
i-Butane	0,108	123	7 177	0,011	340	19 771
n-Butane	0,027	31	1 778	0,002	71	4 112
i-Pentane	0,011	13	934	0,001	21	1 502
n-Pentane	0,003	3	241	0,000	4	288
n-Hexane	0,000	0	0	0,000	0	0
n-Heptane	0,000	0	0	0,000	0	0
n-Octane	0,000	0	0	0,000	0	0
Benzene	0,000	0	0	0,000	0	0
Toluene	0,000	0	0	0,000	0	0
Total	1,000	1 146	33 462	1,000	31 733	583 710

	8.0			LNG		
80%LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,000	0	1	0,012	354	9 923
Methane	0,140	677	10 868	0,900	27 507	441 294
Ethane	0,177	857	25 772	0,058	1 774	53 332
Propane	0,253	1 226	54 077	0,022	680	29 991
i-Butane	0,210	1 017	59 137	0,007	216	12 527
n-Butane	0,062	298	17 330	0,001	40	2 306
i-Pentane	0,071	345	24 925	0,000	8	577
n-Pentane	0,061	297	21 425	0,000	1	93
n-Hexane	0,025	119	10 238	0,000	0	0
n-Heptane	0,000	1	52	0,000	0	0
n-Octane	0,000	0	6	0,000	0	0
Benzene	0,000	1	47	0,000	0	0
Toluene	0,000	0	13	0,000	0	0
Total	1,000	4 839	223 893	1,000	30 580	550 043

G. Increase Case D

	7.4			7.0		
100% LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,003	3	93	0,011	358	10 017
Methane	0,539	654	10 497	0,886	28 096	450 734
Ethane	0,144	175	5 249	0,060	1 888	56 758
Propane	0,170	207	9 130	0,029	913	40 242
i-Butane	0,110	134	7 776	0,011	353	20 518
n-Butane	0,027	33	1 908	0,002	73	4 238
i-Pentane	0,007	8	589	0,000	14	1 012
n-Pentane	0,001	1	65	0,000	1	97
n-Hexane	0,000	0	0	0,000	0	0
n-Heptane	0,000	0	0	0,000	0	0
n-Octane	0,000	0	0	0,000	0	0
Benzene	0,000	0	0	0,000	0	0
Toluene	0,000	0	0	0,000	0	0
Total	1,000	1 215	35 306	1,000		583 614

	8.0			LNG		
100% LPG Increase	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen	0,000	0	1	0,012	354	9 924
Methane	0,140	748	11 997	0,900	27 442	440 245
Ethane	0,172	919	27 629	0,056	1 713	51 513
Propane	0,265	1 414	62 364	0,023	706	31 133
i-Butane	0,217	1 156	67 216	0,007	219	12 755
n-Butane	0,063	337	19 610	0,001	40	2 324
i-Pentane	0,065	350	25 236	0,000	5	377
n-Pentane	0,056	299	21 553	0,000	0	30
n-Hexane	0,022	117	10 075	0,000	0	0
n-Heptane	0,000	0	50	0,000	0	0
n-Octane	0,000	0	6	0,000	0	0
Benzene	0,000	1	47	0,000	0	0
Toluene	0,000	0	13	0,000	0	0
Total	1,000	5 341	245 797	1,000	30 480	548 301

H. Increase Case E

		7.4			7.0		
200% Increase	**LPG**	**[-]**	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen		0,003	3	73	0,012	355	9 948
Methane		0,524	513	8 232	0,892	27 252	437 201
Ethane		0,125	122	3 677	0,052	1 589	47 790
Propane		0,180	177	7 788	0,030	904	39 862
i-Butane		0,132	129	7 496	0,013	382	22 212
n-Butane		0,032	32	1 844	0,003	78	4 530
i-Pentane		0,004	4	278	0,000	7	539
n-Pentane		0,000	0	29	0,000	1	53
n-Hexane		0,000	0	0	0,000	0	0
n-Heptane		0,000	0	0	0,000	0	0
n-Octane		0,000	0	0	0,000	0	0
Benzene		0,000	0	0	0,000	0	0
Toluene		0,000	0	0	0,000	0	0
Total		1,000	980	29 416	1,000		562 135
		8.0			**LNG**		
200% Increase	**LPG**	[-]	Molar Flow	Mass Flow	[-]	Molar Flow	Mass Flow
Nitrogen		0,000	0	3	0,012	353	9 875
Methane		0,150	1 316	21 111	0,904	26 739	428 966
Ethane		0,152	1 330	39 980	0,050	1 467	44 111
Propane		0,280	2 453	108 149	0,025	728	32 082
i-Butane		0,257	2 252	130 911	0,009	253	14 722
n-Butane		0,075	657	38 172	0,002	46	2 679
i-Pentane		0,041	361	26 055	0,000	3	222
n-Pentane		0,035	303	21 867	0,000	0	18
n-Hexane		0,011	100	8 652	0,000	0	0
n-Heptane		0,000	0	39	0,000	0	0
n-Octane		0,000	0	4	0,000	0	0
Benzene		0,000	1	41	0,000	0	0
Toluene		0,000	0	10	0,000	0	0
Total		1,000	8 773	394 994	1,000	29 588	532 675

I. Scrub Column Vapour Flows [kgmole/h] Case A-E

	Case A	Case B	Case C	Case D	Case E
Vapour Flow	**20% Increase**	**60% Increase**	**80% Increase**	**100% Increase**	**200% Increase**
21	31 869	31 869	31 733	31 694	30 569
20	31 547	31 549	31 429	31 397	30 384
19	31 463	31 449	31 344	31 313	30 340
18	31 428	31 392	31 297	31 262	30 315
17	31 411	31 355	31 265	31 224	30 295
16	31 403	31 331	31 244	31 195	30 278
15	31 398	31 314	31 228	31 173	30 264
14	31 395	31 303	31 217	31 156	30 252
13	31 392	31 294	31 209	31 143	30 242
12	31 390	31 287	31 202	31 132	30 233
11	31 388	31 281	31 196	31 123	30 225
10	31 384	31 274	31 190	31 114	30 217
9	31 376	31 265	31 181	31 103	30 208
8	31 360	31 248	31 165	31 086	30 196
7	31 330	31 215	31 133	31 052	30 173
6	2 063	2 771	3 122	3 477	5 677
5	2 085	2 800	3 154	3 512	5 734
4	2 094	2 810	3 165	3 523	5 748
3	2 100	2 817	3 171	3 530	5 742
2	2 140	2 883	3 255	3 636	5 889
1	2 387	3 310	3 799	4 319	7 119
reboiler	3 369	5 016	5 964	7 032	12 713

J. Scrub Column Liquid Flows [kgmole/h] Case A-E

	Case A	**Case B**	**Case C**	**Case D**	**Case E**
Liquid Flow	**20% Increase**	**60% Increase**	**80% Increase**	**100% Increase**	**200% Increase**
21	641	862	842	918	795
20	558	761	757	834	751
19	523	704	709	782	725
18	506	668	678	744	705
17	497	644	656	716	689
16	492	627	641	694	675
15	489	615	630	677	663
14	487	607	622	664	653
13	485	599	615	653	644
12	483	593	609	643	636
11	478	586	602	634	628
10	470	577	593	623	619
9	454	561	578	606	606
8	424	527	545	573	584
7	5 395	7 113	7 961	8 818	14 450
6	5 417	7 141	7 993	8 853	14 507
5	5 426	7 152	8 004	8 865	14 521
4	5 432	7 158	8 010	8 871	14 514
3	5 472	7 225	8 094	8 977	14 662
2	5 719	7 652	8 638	9 660	15 891
1	6 701	9 357	10 803	12 373	21 486
reboiler	3 332	4 342	4 839	5 341	8 773