NTNU – Trondheim
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Helium Extraction from LNG End Flash

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Submission date: June 2014
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Norwegian University of Science and Technology
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NTNU
Norwegian University
of Science and Technology

Department of Energy
and Process Engineering

EPT-M-2014-54

MASTER THESIS

for

Student Donghoi Kim
Spring 2014

Helium extraction from LNG end flash

Gjenvinning av helium fra end flash i LNG-anlegg

Background and objective

Natural gas is today one of the largest sources for helium. Helium in natural gas is usually found together with nitrogen, and one attractive way of recovering helium is to extract it from the end flash gas in LNG production facilities, where there is already a cryogenic process facility that can be adapted to helium recovery. The LNG plant operator then normally sells a mixture of nitrogen and helium (raw helium) to an offsite helium producer, which does the final purification, liquefaction and storage of the purified He product. The raw helium gas consists of at least 40% helium where the other components are nitrogen and small amounts of methane, and in some cases also hydrogen, argon and neon.

The challenge for helium recovery from LNG end flash is to design a cryogenic process that is able to deliver

- LNG (with sufficiently low nitrogen content – usually below 1 mol%),
- Nitrogen to stack (with sufficiently low C_1 content to minimize hydrocarbon losses), and
- Raw helium (i.e. N_2 with more than 40 mol% He).

The process need to be integrated into the cryogenic end flash stage of the LNG facility, as well as into the liquefaction process, with a sensible process configuration to minimize power demands and complexity of design and operation. There seems to be few publications on this subject, but some basis may be found in the literature (including patent publications).

The main objective of the Master thesis will be to propose, analyze and compare various process configurations for "raw" helium extraction from LNG end flash gas, and to provide recommendations for process selection and further work in this area. Further processing of the raw helium is not within the scope of the work.

The following tasks are to be considered:

1. Literature survey of helium extraction and recovery processes from LNG end flash and natural gas/nitrogen gas sources, including patent literature. Selection of process options for further study and evaluation.

2. Establishment of design and modelling basis, and preparation of process/flow sheet models for relevant helium extraction and processing schemes.

3. Analysis and comparison of process options for helium extraction from LNG end flash, including evaluation of separation efficiency and energy losses, complexity, equipment/design issues, and operational issues.

4. Overall evaluation, conclusion and recommendations from the study, including generic learnings and observations from the work and suggested way forward.

-- " --

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

Department of Energy and Process Engineering, 14. January 2014

Olav Bolland
Department Head

Jostein Pettersen
Academic Supervisor

Research Advisor: Professor Truls Gundersen, NTNU

Preface

This master thesis has been written at Department of Energy and Process Engineering, Norwegian University of Science and Technology (NTNU), as a final part of the international master's degree programme in Natural Gas Technology.

First, I would like to thank my supervisor Jostein Pettersen and co-supervisor Truls Gundersen from Department of Energy and Process Engineering, NTNU, for their dedicated support and valuable feedback on my project. I really appreciate that I could have enough time to be guided by having many meetings.

Second, I would like to extend my gratitude to Rahul Anantharaman, SINTEF, for helping me whenever I had troubles with modelling work in HYSYS.

Finally yet importantly, I really appreciate my friends in Norway and South Korea who supported me during this semester.

Trondheim, 10th June 2014

Donghoi Kim

Abstract

Helium is an invaluable element as it is widely used in industry such as cryogenics and welding due to its unique properties. However, helium shortage is expected in near future because of increasing demand and the anxiety of supply. Consequently, helium production has attracted the attention of industry. The main source of He is natural gas and extracting it from LNG end-flash is considered as the most promising way of producing crude helium. Thus, many process suppliers have proposed process configurations for this and there are mainly three types of helium extraction process, flashing-based, distillation-based and the integration of flashing and distillation. Therefore, the objective of this thesis is to conduct a comparative evaluation of the proposed helium extraction processes for LNG plants and give a guideline for a proper selection. This evaluation was performed by simulating each process through Aspen HYSYS. The simulation result was then analysed, focusing on various criteria.

The result indicates that all the helium extraction process studied could extract most of helium contained in feed gas except the Technip Distillation process. Regarding LNG production, the integration of the Re-boiled Distillation and a LNG process gives the lowest LNG production specific power. In terms of fuel gas, only the APCI Distillation process could generate fuel gas having less than 40 mole% N_2, which is upper limit for industrial gas turbines, even with He and thus N_2 rich feed gas. Between others, only the Technip Distillation and the ExxonMobil Integration process could produce liquid N_2 and the amount was enough to be used for further processing of crude helium. Concerning freezing of impurities such as CO_2, the ExxonMobil Integration process displayed an excellent performance, recording the warmest working temperature. When considering complexity, the APCI Distillation process was the simplest one, while the two integration-based processes from Linde and ExxonMobil required a lot of equipment. Finally, economic evaluation showed that all the integrations of each the helium extraction process studied and a LNG plant create more economic value than a LNG plant alone. Nevertheless, the difference was just ca. 1 % with current helium price.

The selection of the most suitable helium extraction process for a LNG plant completely depends on the situation as each process possesses its own characteristic. Thus, proper technical choices have to be made to achieve a successful helium extraction project.

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Nomenclature

Abbreviations

ASU	Air separation unit
CAPEX	Capital expenditure
DMR	Dual mixed refrigerant
HC	Hydrocarbon
HE	Heat exchanger
HeRU	Helium recovery unit
HeXU	Helium extraction unit
J-T	Joule-Thomson
LHe	Liquefied helium
LMTD	Log mean temperature difference
LN_2	Liquefied nitrogen
LNG	Liquefied natural gas
MBTU	Million British thermal units
MRI	Magnetic resonance imaging
MTPA	Million tonne per annual
NRU	Nitrogen removal unit
OPEX	Operating expenditure
PFHE	Plate fin heat exchanger
PSA	Pressure swing adsorber
STHE	Shell and tube heat exchanger
SWHE	Spiral wound heat exchanger

1 Introduction

In most people's mind, helium is not regarded as an invaluable element. However, this noble gas is a vital component widely used for industry like cryogenics, semiconductors and welding. In addition, the fact that this element is irreplaceable due to its unique properties makes it more important. However, helium shortage is expected in near future because of increasing demand and the anxiety of supply. Accordingly, helium production has attracted the attention of industry.

The main sources of this noble gas are air and natural gas. Between them, natural gas is today the largest source for helium since extracting helium from air is technically uneconomical. Helium in natural gas is usually found together with nitrogen, and one attractive way of recovering helium is to extract it from the end-flash gas in LNG production facilities, where there is already a cryogenic process facility that can be adapted to helium extraction. The LNG plant operator then normally sells a mixture of nitrogen and helium (crude helium) to an offsite helium producer, which does the final purification, liquefaction and storage of the purified He product. The crude helium gas consists of at least 50% helium where the other components are nitrogen and small amounts of methane, and in some cases also hydrogen, argon and neon.

The helium extraction process need to be integrated into the cryogenic end-flash stage of the LNG facility, as well as into the liquefaction process, with a sensible process configuration to minimize power demands and complexity of design and operation. Thus, many process suppliers and oil & gas operators have proposed process configurations for helium extraction from LNG end-flash to optimise the integration, while reducing energy consumption for producing crude helium.

It is challenging to find the most suitable helium extraction process to be integrated to a LNG plant since it depends on various parameters, such as helium extraction efficiency and the effect on LNG production rate. However, there is no technical paper, which evaluates and compares the processes for helium extraction from LNG end-flash. Therefore, it is worth assessing them to find out the most appropriate technology to be added to a LNG process.

Thus, the main objective of this master thesis is to analyze and compare various process configurations for helium extraction from LNG end-flash, and to provide recommendations for process selection and further work in this area. Further processing of the crude helium is not within the scope of the work.

This assessment was performed by modelling and simulating each helium extraction process through Aspen HYSYS (shortly HYSYS), which is a commercial simulation tool widely used for hydrocarbon processes. The design basis for modelling was carefully set in order to achieve a fair comparison among the different types of the processes. The simulation result was then analysed by focusing on the performance of crude helium and LNG production, the quality of fuel gas and liquid nitrogen, helium extraction unit working conditions, economic evaluation, and equipment count. Such detailed comparison will guide us to find a proper solution for extracting helium from LNG end-flash.

This report explores the topic of this thesis project through following chapters. Chapter 2 explains where helium is used, why there is a need for more helium production facilities and what type of helium extraction process is promising. Chapter 3 discusses the principle of helium extraction process, including the latest technologies evaluated in this project. Chapter 4 describes the design basis for a fair comparison, containing simulation scope, conditions for modelling, etc. Chapter 5 explains the procedures for modelling of helium extraction processes and provides the simulation result of the models by using HYSYS. Chapter 6 analyses and evaluates the simulation result based on key parameters for helium extraction processes. Chapter 7 conclude the analysis result and provides recommendations. Chapter 8 introduces several important issues for the future work.

2 General Information about Helium

2.1 Properties and Applications

Helium is one of noble gases having no colour, taste and odour. Due to its unique physical and chemical properties, helium is widely used for industrial applications. First, helium has the lowest boiling point of 4.2 K among any other elements and it does not freeze even at zero K under standard pressure (Lemmon, Huber, & McLinden, 2013). This characteristic enables helium to be used for cryogenic processes, such as cooling of superconductors in magnetic resonance imaging (MRI) equipment and condensing of hydrogen and oxygen for rocket fuel, which requires extremely low temperatures. In addition, helium is used to purge fuel tanks and fuel delivery systems, for example rocket propulsion systems (Smith, Goodwin, & Schillinger, 2004). As fuel is stored and delivered at very low temperatures, helium is a suitable substance for purging since it does not form solid at the low temperatures.

Helium is an inert gas, having the second lowest reactivity. This feature helps helium to be used in manufacturing processes when an inert atmosphere around products is needed, such as semiconductors. Helium is also widely used as an ideal shielding gas for welding like aluminium, stainless steel and magnesium alloys due to its inert properties, which prevent materials from oxidation during welding (RasGas, 2005).

Moreover, the high thermal conductivity of helium allows this noble gas to be used for metallurgical processes, for example optical fibre production, as this type of processes requires rapid cooling. Together with radiation tolerance, the high thermal conductivity also makes helium to be an optimal heat transfer substance for nuclear power plants (Das, Kumar, Mallik, & Bhandari, 2012).

Another striking feature of helium is its lightness. Helium gas is the second lightest gas, followed by hydrogen. In conjunction with its non-flammability, this feature allows helium to be used to lift airships and balloons. Even though hydrogen gas has the lowest density, it is not preferred due to its flammability and explosiveness (Smith et al., 2004).

The molecular size of helium is extremely small so that it can be used for leak detection for various systems. Due to the small size of helium atoms, helium escapes through very small gaps in equipment demanding a high degree of

sealing. Thus, helium leak test is performed for valves, pipelines, heat exchangers and so forth to make sure it is sealed thoroughly (RasGas, 2005). The small molecular size also makes helium to be more diffusive than air so helium is utilised as breathing mixture for divers (Johnson, 2012).

Figure 2.1 summarises the main end uses of helium and its share in 2011 in U.S. This figure clearly indicates that helium is mainly used for cryogenics, controlled atmosphere, purging and welding.

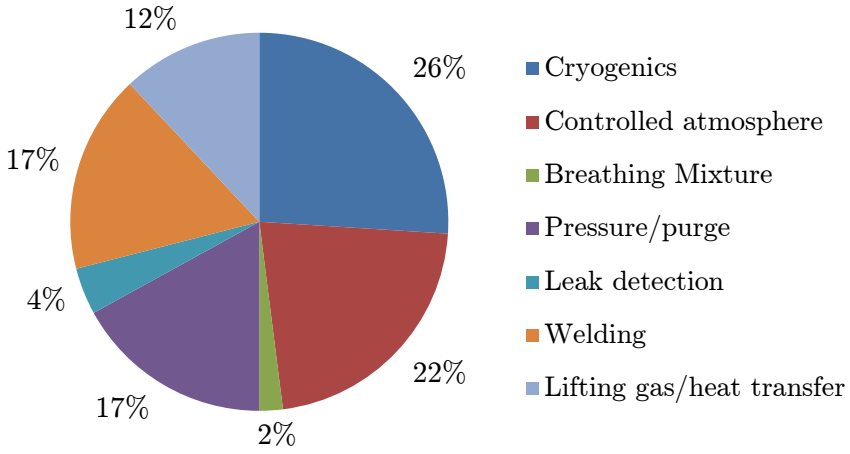


Figure 2.1 Estimated Helium Consumption, by End Use, in U.S. in 2011 (Peterson & Madrid, 2013)

2.2 Main Sources of Helium

Helium is an invaluable element to industries as explained Chapter 2.1. Helium is naturally produced in the earth's crust by alpha decay of uranium and thorium, which are radioactive elements. The produced helium is then leaked to the atmosphere and finally released to space. The amount of helium production balances with the one of helium escaped from the earth, resulting in the constant helium fraction in the atmosphere (Haussinger et al., 2000). Therefore, helium is a non-renewable resource.

In addition to the atmosphere, such the invaluable element, helium, was also found in a natural gas field in U.S. in 1906 (Nuttall, Clarke, & Glowacki, 2012). After that, helium is believed to be existed in all natural gases (Broadhead, 2005). Consequently, there are two conventional sources of helium, the atmosphere and natural gases. However, extracting helium from the

atmosphere is not regarded as an attractive option. Together with the high volatility, the low concentration of helium in air (0.056 mole% (Haussinger et al., 2000)) makes extracting helium from the atmosphere uneconomical (Ernst&Young, 2012). Therefore, only natural gases have been the major source of helium. Most of helium reserves dissolved in natural gases is located in only few countries in the world as seen in Figure 2.2. Especially, the United States accounts for almost 50 % of the total helium reserves in the world.

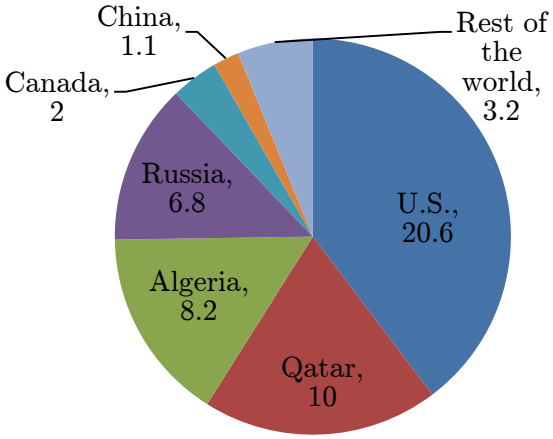


Figure 2.2 World Helium Reserves (Unit: Billion Cubic Meters) (Hamak, 2014)

There are also unconventional helium sources. Except natural gases, helium is dissolved in monazite mineral and thermal springs (Mukhopadhyay, 1980). The noble gas is sometimes discovered in natural CO_2 reserves. Recently, a facility for separating helium from the carbon dioxide gases is expected to be built by Air Products and Chemical (APCI) in the United States (Bomgardner, 2013).

2.3 Helium Shortage

Helium is mainly produced by the United States. As Figure 2.3 shows, U.S. accounts for around 75% of world helium production. Thus, global helium supply hugely relies on the United States. In particular, about 40% of helium produced in U.S. comes from government owned helium storage facilities, which are operated by the Bureau of Land Management (BLM) (Hamak, 2014). The U.S. government has bought and stored tremendous helium produced from helium bearing natural gas fields especially after Helium Act Amendments was passed in 1960s (Smith et al., 2004). However, Helium Privatization Act was approved in 1996. According to the law, BLM started sell-

ing the government helium stockpile to pay the debt incurred by buying helium in the past. This amount of helium has been playing an important role as it provides 30% of world helium supply up to now.

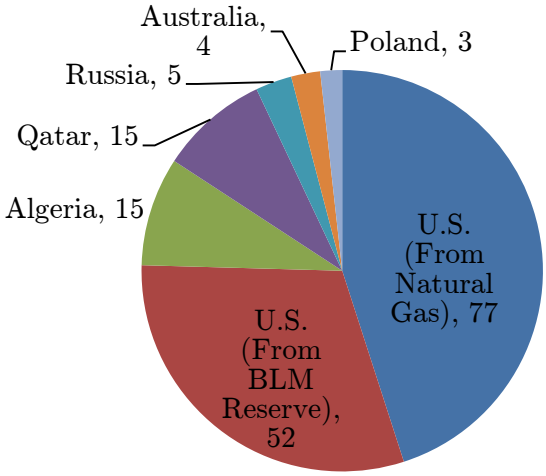


Figure 2.3 World Helium Production in 2013 (Unit: Billion Cubic Meters) (Hamak, 2014)

However, the BLM helium reserve was planned to be shutdown in mid-2013 as the debt was predicted to be paid off at the time (Kynett, 2012). Thus, a significant shortage of helium was expected in U.S. and thus the world, causing a price rise in helium. To prevent such catastrophe, the U.S. congress passed Helium Stewardship Act on October 2013, allowing BLM to sell 60% of the rest of the helium stockpile until 2015 (Esneault, 2013). However, this act could not be the permanent solution as BLM is going to shut down their reserve in 2015, while losing 40% of world helium supply. In addition to this, the fact that the production capacity of helium-bearing natural gas fields in U.S. has been declining by 10 to 15 % per year tells the total helium production capacity of the United States will clearly decrease after 2015, while the amount of decrease in helium production is offset by helium coming from BLM reserve until 2015 (Smith et al., 2004).

In contrast to the fact that helium production in U.S. is stable, world helium demand continues rising. China, South Korea and Taiwan are regarded as the main contributors to the increasing helium demand as their electronics industry is rapidly growing, such as semiconductors (Nuttall et al., 2012). It means there must be new helium projects. Fortunately, Qatar launched a new helium production plant on December 2013, solely meeting 8% of world helium de-

mand (John, 2013). Moreover, Russia and Algeria are planning to build helium production facilities in near future (Ernst&Young, 2012; Shiryaevskaya, 2011). Despite the world helium supply is going to increase thanks to Qatar, Algeria and Russia, some experts anticipated that the global demand for helium would be outnumbered the helium production in ten years (Ernst&Young, 2012).

There is also uncertainty of helium supply. Since there are only six players (Air Liquide, APCI, Linde, Messer, Matheson and Praxair) in helium sector in the world, some of the companies have complete control of helium supply in certain area, giving a doubt that they will respond change in helium demand in a timely manner (Nuttall et al., 2012). Moreover, helium supply is easily disrupted when one of any helium production plants in the world is going offline, as there are only few. Actually, shutdown of Algerian helium plant in 2004 and Wyoming helium plant in U.S. in 2011 had a significant effect on helium market, a shortage of helium supply for around two months (Kynett, 2012). Thus, increasing helium demand and the anxiety of helium supply has led rise in helium price as shown in Figure 2.4. This figure shows that helium price has been quadrupled for the last fifteen years in private sectors and this trend is expected to be continued (Cai, Clarke, Glowacki, Nuttall, & Ward, 2010; Garvey, 2011)

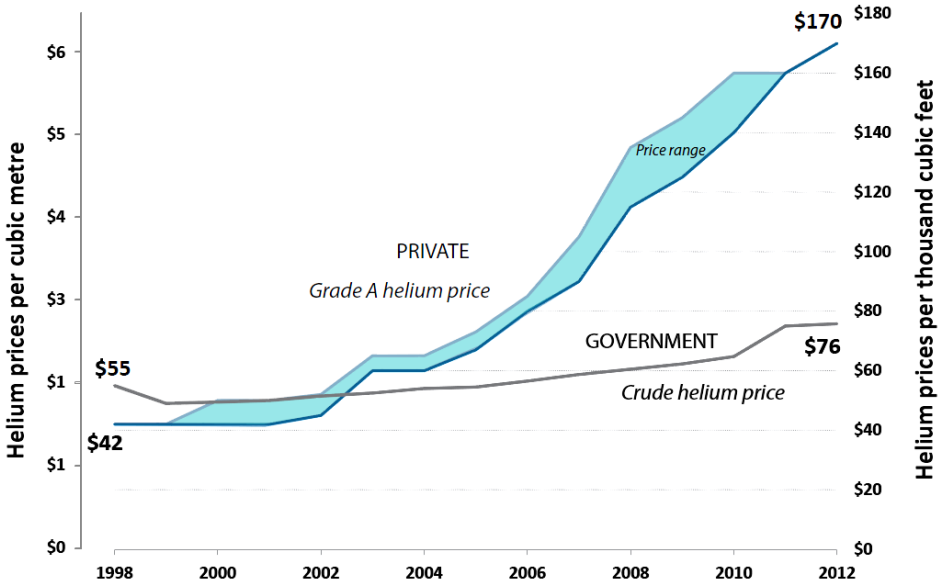


Figure 2.4 Price Increases in Grade-A and Crude Helium between 1999 and 2011 (Johnson, 2012)

Thus, more helium projects are required to resolve the tight helium supply situation. APCI also predicted that a new helium production plant is needed for every 20 months to meet the increasing helium demand (Smith et al., 2004)

2.4 Trends in Helium Extraction Facilities

As explained in Chapter 2.3, there is a need for more helium plants and the high price of helium has made a helium project more attractive to natural gas/LNG producers. In the past, this noble gas could be commercially extracted from gas processing facilities only when natural gas holds at least 0.1-0.5 vol% of helium (Gottier, 1991; Handley & Miller, 1992; Haussinger et al., 2000; Smith et al., 2004). It was mainly due to the low temperature level (ca. -185°C), which is required extracting helium from natural gas. To cool a natural gas stream down to the required temperature level, an external refrigeration system is typically installed, consuming a large amount of energy and thus reducing economic feasibility of helium production in natural gas processing plants (Oelfke & Victory, 2013).

However, this limitation has been extended downwards to ca. 0.05 mole% He in natural gas by extracting it from LNG end-flash, making more natural gas fields economically viable to produce helium. There are also more advantages from a process point of view. First, helium is already concentrated in LNG end-flash in some degree, making extracting helium in downstream easier, compared to extracting it from helium-bearing natural gas (Haussinger et al., 2000). In addition to that, an external refrigeration system required for helium extraction in gas processing facilities is not needed as LNG liquefaction process has already low enough temperature level to recover helium (Oelfke & Victory, 2013). Moreover, there is no need for extra gas treatment facilities since natural gas is thoroughly treated in a LNG plant. As helium extraction is conducted at a low temperature, there is a high chance of freezing impurities during the extraction. So more attention had to be given to natural gas processing facilities to reduce the impurities before the gas enters a helium extraction step (Haussinger et al., 2000).

Therefore, integrating helium extraction process into a LNG plant has been drawing attention from industry as a new source of helium, helping LNG project economics by producing helium as a by-product (Cai et al., 2010). Consequently, state-of-the arts have been newly proposed by process providers to extract helium from LNG end-flash and these technologies are explained in the next chapter.

3 Helium Extraction from LNG

3.1 An Overview of a Helium Production Process

The schematic of a helium extraction system from an LNG stream and the scope of work for this thesis are depicted in Figure 3.1. This is mainly divided into five steps. First, natural gas containing helium undergoes pre-treatment facilities in order to remove heavy hydrocarbons, water and sour gases. It is a substantially important step for LNG processes since the impurities may freeze out and cause plugging in cryogenic heat exchangers at a low temperature (Fredheim, Solbraa, Pettersen, & Bolland, 2012).

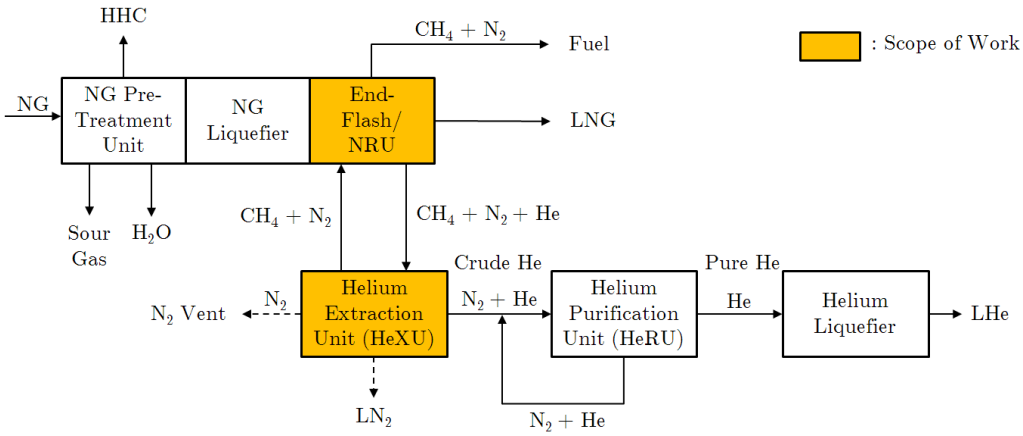


Figure 3.1 Block Diagram of Helium Production from LNG

The allowable quantities of the impurities are even more severely restricted when adding a helium extraction process to LNG plants due to its lower working temperature than LNG processes (Haussinger et al., 2000). At a lower temperature, the contaminants, especially CO_2, become solid at lower concentration (Mokhatab, Mak, Valappil, & Wood, 2014; Oelfke & Victory, 2013). Consequently, they have to be removed to lower level for helium extraction, compared to typical LNG processes. It may need higher specifications of the treatment facilities, increasing its capital and operating cost. Therefore, the working temperature of a helium extraction unit (HeXU) is of great concern as the temperature decides the specifications of the upstream pre-treatment facilities (Haussinger et al., 2000; Mokhatab et al., 2014).

After pre-treatment, the feed gas is liquefied and sub-cooled before being expanded to intermediate pressure to generate a multiphase stream. For the expansion, Joule-Thomson valves or hydraulic turbines are usually applied. The multiphase stream is separated into a gas and liquid in a flash tank ("end flash"). Through this, most of the helium and some of nitrogen contained in the feed stream are split into the gas. Consequently, helium-free LNG with reduced nitrogen content is produced and it is depressurised to around atmospheric pressure for storage.

The helium separation is generally conducted as part of the nitrogen removal process. Helium bearing natural gas tends to have a high nitrogen fraction (Chiu & Sheu, 2011; Handley & Miller, 1992; Paradowski & Vovard, 2011). The nitrogen content in natural gas has to be reduced to increase LNG heating value and meet the transport specification (Mokhatab et al., 2014; Vovard, Bladanet, & Cook, 2011). So it is commonly found that a helium extraction process is integrated with the nitrogen removal process (Haussinger et al., 2000). If the nitrogen concentration is too high in the feed gas, for example over 5 mole%, this flashing based separation may be replaced by a nitrogen removal unit (NRU), which is a distillation based process (Jacks & McMillan, 1978). This is to fulfil the nitrogen content requirement of LNG, which is generally lower than 1 mole% N_2 (Mokhatab et al., 2014).

The helium concentrated gas from a flash tank or a nitrogen removal unit in an LNG plant is then fed to a HeXU where crude helium is produced by cryogenic separation. Normally a HeXU consists of a phase separation drum and a heat exchanger as illustrated in Figure 3.2. The helium rich gas still possesses large amounts of N_2 and CH_4. Therefore, the gas is partially condensed through the heat exchanger so that nitrogen and methane can be separated into the condensate. As the boiling point of helium is lower than the other two components, most of the helium is left in the gas phase while nitrogen and methane is condensed. (Froehlich & Clausen, 2007; Haussinger et al., 2000). This results in a high concentration of helium in the crude helium stream, which is normally higher than 50 mole% and the rest is mainly nitrogen (Daly, 2005). The cooling duty of the heat exchanger is supplied by expanding the condensed nitrogen and methane mixture through a Joule-Thomson valve. Thus, this is an auto-refrigeration process. Depending on the type of helium extraction processes, the mixture may consist of almost pure nitrogen. Therefore, one can vent it to the atmosphere or collect the liquefied nitrogen for other process steps.

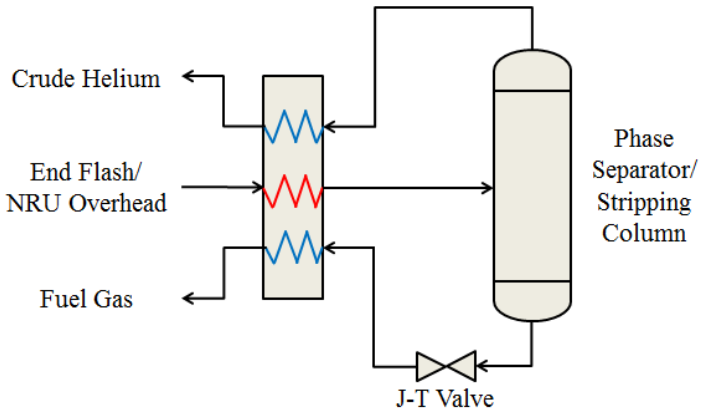


Figure 3.2 An Example of a Typical Helium Extraction Unit

The crude helium is then purified up to 99.999 vol% by a pressure swing adsorber (PSA) in a helium recovery unit (HeRU) (Daly, 2005; Haussinger et al., 2000). If necessary, H_2 and CO_2 removal steps can be added to the HeRU as illustrated in Figure 3.3. It is essential that any impurities are reduced to a very low level in this stage. Otherwise, they will freeze out in the helium liquefier, which has an extremely low working temperature. Before entering the PSA, nitrogen content in the raw helium gas is reduced by cryogenic separation, which utilises liquid N_2 as a refrigerant. Even though the PSA can purify a raw helium gas having higher than 50 vol% of nitrogen and generate pure helium, the HeRU requires the cryogenic nitrogen separation step for economic reasons as explained in the following.

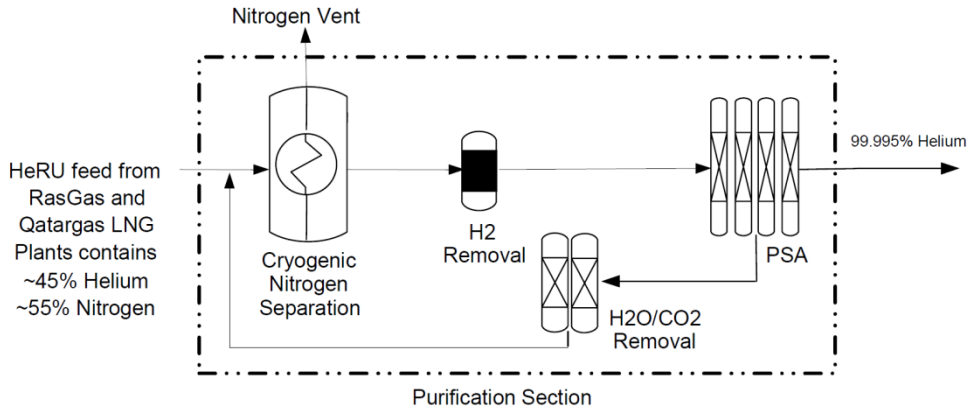


Figure 3.3 An Example of a Typical Helium Recovery Unit (Daly, 2005)

The tail gas rich in nitrogen from the PSA still holds considerable amounts of helium so it is recompressed and recycled to the upstream of the HeRU (Agrawal, Herron, Rowles, & Kinard, 2000; Haussinger et al., 2000). If the gas entering the PSA has a higher content of nitrogen, it causes a larger amount of the tail gas, increasing capital and operating cost for the tail gas recompression facility. The recommended nitrogen fraction of the gas entering the PSA is regarded as 5-15 vol% (Froehlich & Clausen, 2007; Haussinger et al., 2000).

Finally, the pure helium gas discharged from the PSA is liquefied in a helium liquefier unit. Linde, Claude and Collins processes are the main technologies applied for helium liquefaction (Haussinger et al., 2000). All the processes provide the cold duty by expanding helium itself through J-T valves and/or hydraulic turbines. A point to be considered is that J-T valves can only be used when the J-T coefficient is positive at a given condition so that one can have temperature reduction via this device. Again, cryogenic adsorbers are installed in this unit to make sure that all the impurities are removed before helium is fully liquefied (Chiu & Sheu, 2011; Froehlich & Clausen, 2007; Haussinger et al., 2000). Then helium is cooled to ca. 4 K at around atmospheric pressure before being stored as a liquid.

As mentioned in Chapter 2.4, the main difference between helium production from natural gas and from LNG is the way crude helium is generated. The rest of the system is almost the same and it is well known technologies. Therefore, this report focuses on the technologies for extracting crude helium from LNG. The following helium extraction methods have been suggested recently in order to be integrated with an LNG plant.

3.2 Technologies for Helium Extraction from LNG Streams

3.2.1 Principle of Helium Extraction Processes

Generally, cryogenic separation is used for extracting helium from LNG. The processes for He extraction are firstly categorised by how to separate helium from LNG before the separated helium rich gas is fed to a HeXU to produce crude helium. The principle of cryogenic separation for the mixture is similar to nitrogen removal from LNG, these are sorted by two principles; flashing & phase separation (shortly named as flashing in this paper) and distillation (Chiu & Sheu, 2011; Gottier, 1991). The main difference between these tech-

nologies is that the helium rich gas has different conditions resulting from each separation method such as composition, temperature and pressure. This variation has a large impact on the performance of crude helium production in a HeXU.

In addition, type of a HeXU can be divided into a typical HeXU and a HeXU including a distillation column. Therefore, there are mainly five types of helium extraction processes as shown in Figure 3.4. Especially, the combination of flashing cycles and a HeXU with a distillation column is named as integration based process in this report.

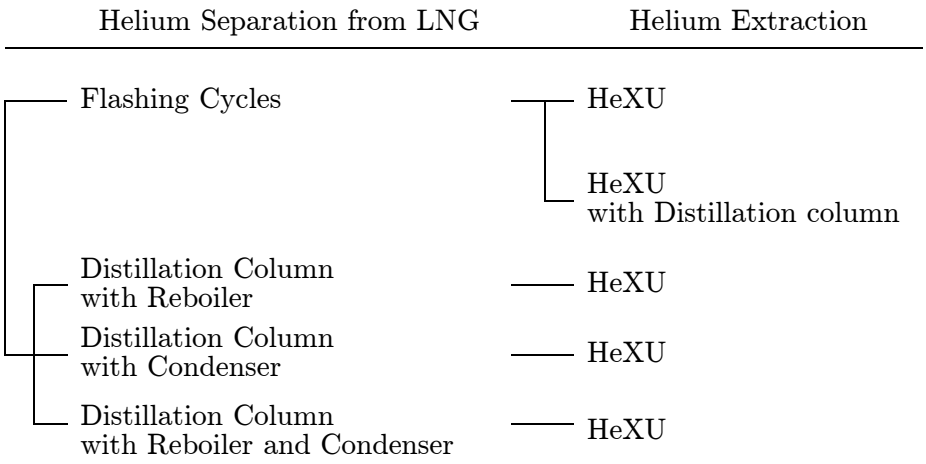


Figure 3.4 Type of Helium Extraction Processes from LNG

The simplest way of separating helium from liquefied natural gas is J-T expansion and phase separation, called flashing. By depressurising LNG, the dew point of lighter components in LNG such as He becomes very low and the lighter components escape from the liquid as vapour phase, which is called end-flash. However, sharp separation of mixtures cannot be achieved by this method. Figure 3.5 shows that the end-flash contains not only He but also large amounts of N_2 and CH_4. To achieve a sharper separation of helium, higher-pressure end-flash is required as indicated in Figure 3.5. This, on the other hand, results in collecting a smaller proportion of helium molar flow in LNG into the flash gas as illustrated in Figure 3.6. Therefore, a relatively low-pressure flash gas (in this example, 5 bara) is normally extracted from LNG to produce an enough amount of crude helium in a downstream process, which is a HeXU.

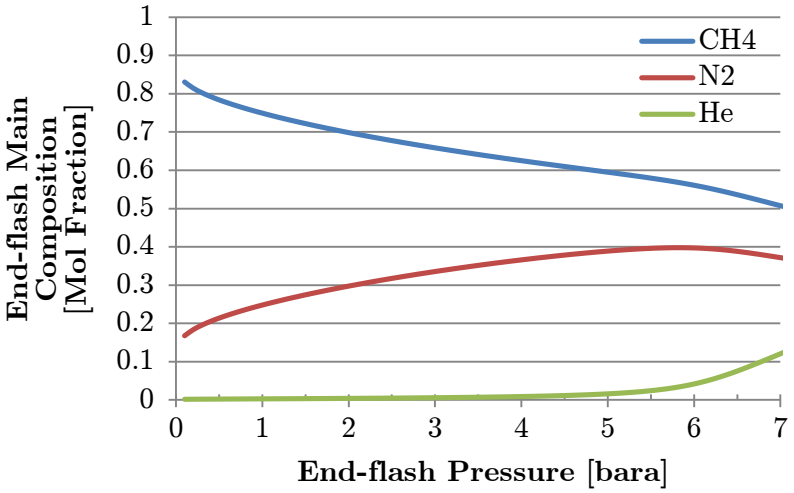


Figure 3.5 An Example of End-flash Composition Concerning End-flash Pressure Levels

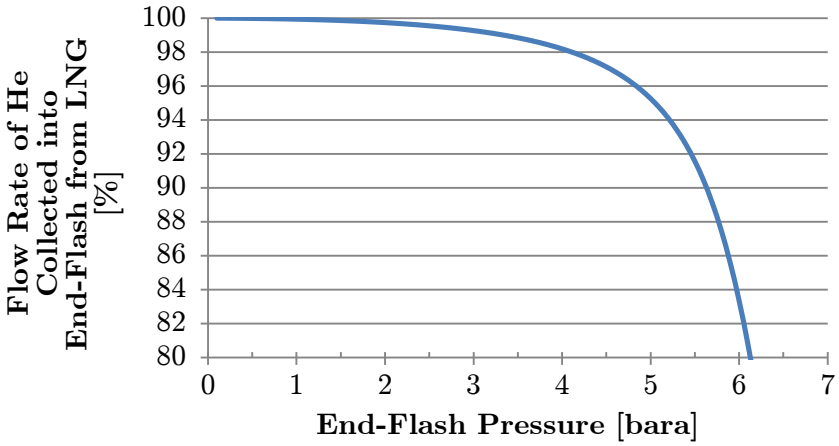


Figure 3.6 An Example of Helium Collection Rate Concerning End-flash Pressure Levels

One thing that has to be considered during J-T expansion is whether the expansion has a cooling effect via a J-T valve. J-T expansion is an isenthalpic depressurising process. During this expansion, it has a heating effect until a certain pressure level (Dots in Figure 3.7) as the J-T coefficient has a negative value. Below the pressure level, it finally has a cooling effect, having the J-T coefficient larger than zero. The solid line made by connecting the dots in Figure 3.7 is called the inversion curve. Thus, J-T expansion has to end up within the inversion curve in order to have a cooling effect. Otherwise, a stream passing through a J-T valve may be heated up, wasting its cold energy.

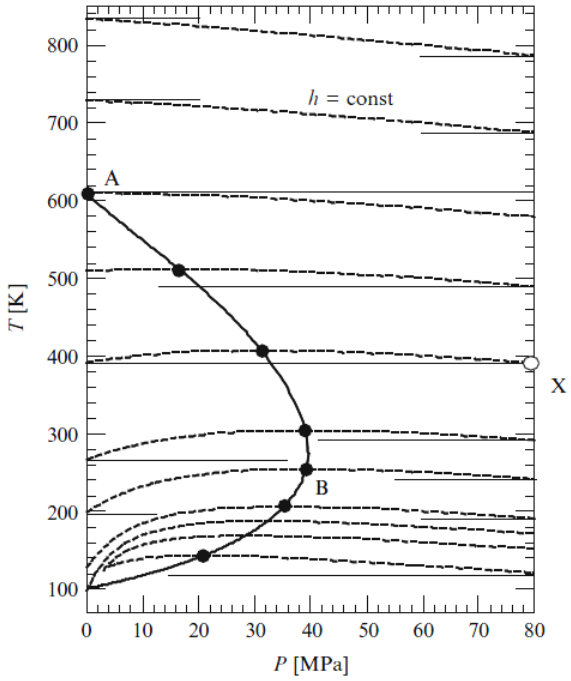


Figure 3.7 An Example of the Inversion Curve (Solid Line) of Nitrogen (Maytal & Pfotenhauer, 2013)

The helium-containing flash gas is then sent to a HeXU to increase its He purity by partial condensation, which is another typical method of cryogenic separation. As explained in Chapter 3.1, crude helium typically holds at least 50 mole% He. To possess such mole fraction, the flash gas has to be partially condensed to reduce CH_4 and N_2 content by lowering its temperature. Figure 3.8 indicates the relationship of He content in crude helium and the end-flash condensing temperatures. The conditions of the feed used for this and following figures were adopted from Chapter 4.1.1. By referring this figure, the 5 bara flash gas has to be cooled to ca. -180°C to achieve 50 mole% He. One can even have higher helium content by further lowering the condensing temperature. However, this causes a smaller amount of He in crude helium as shown in Figure 3.9. To avoid this situation, the temperature has to be even lower than -190°C, which is unfavourable when considering the freezing of impurities in a HeXU.

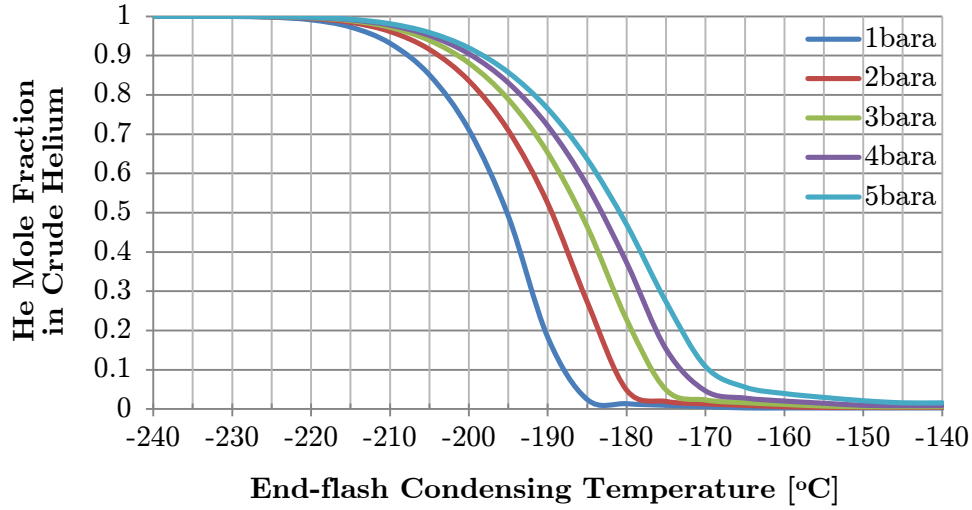


Figure 3.8 He Mole Fraction versus End-flash Condensing Temperature

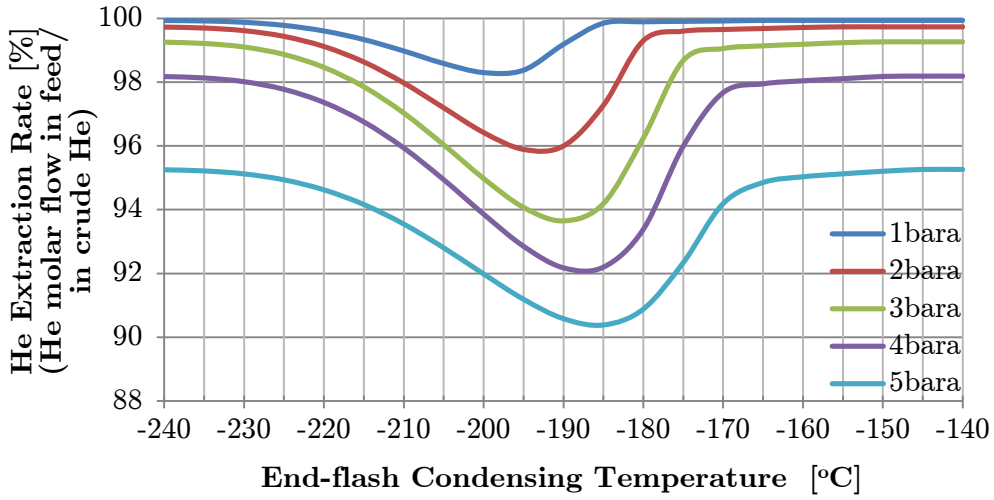


Figure 3.9 He Extraction Rate versus End-flash Condensing Temperature

To further increase helium molar flow in crude helium, one may reduce end-flash pressure. With a specific condensing temperature, lower-pressure end-flash delivers a higher amount of He in crude helium as seen in Figure 3.9. However, it requires lower condensing temperature to reach a specific helium mole fraction in crude helium compared to higher-pressure end-flash as indicated in Figure 3.8. For example, 5 bara end-flash needs to be chilled to around -181°C to have 50 mole% He in crude helium and 3 bara end-flash has to be cooled to ca. -190°C to reach the same helium mole fraction. This is a

very undesirable condition for HeXUs. First, the lower condensing temperature makes impurities easy to be frozen in a HeXU. In addition, it may need more compression energy for crude helium and fuel gas due to the lower pressure.

Therefore, choosing an end-flash pressure level and its condensing temperature is a matter of economic evaluation whether the benefit from selling more crude helium with lower end-flash pressure exceeds the increased operating cost of crude helium and fuel gas compressors. To overcome this offsetting matter and make flashing based processes more efficient, various helium extraction methods have been suggested. Thus, two processes between them were chosen for this project and explained in Chapter 3.2.2. Later, their performances are evaluated in Chapter 5 and 6.

Another main principle of helium extraction processes is distillation. As discussed earlier, flashing cannot achieve sharp helium separation from LNG. To have shaper separation of helium, a distillation column can be used instead of a J-T valve and a phase separator. By doing this, one can extract a larger amount of helium from LNG into the column overhead product at a specific pressure level, compared to the flashing method. It results in producing more crude helium and reducing power consumption of crude helium and fuel gas compressors. The main problem of such cryogenic distillation is the difficulty of finding proper cooling media for the condenser of the column (Windmeier & Barron, 2000). Therefore, many applications have been developed to fulfil the condenser duty by heat integrating the column with the whole system. In Chapter 3.2.3, two of the configurations are introduced and further analysed in Chapter 5 and 6.

There are also various configurations of helium extraction processes, which apply both flashing and distillation. First, a flashing step is used to obtain helium-enriched end-flash. Then, unlike flash-based separation in a HeXU, a distillation column is included in the unit. By partial condensation of end-flash in a HeXU, one can only produce a limited amount of crude helium with a specific helium mole fraction. However, the distillation column enables higher crude helium production, while keeping the pressure level high and helium content of crude helium up. Thus, two processes using both separation principles are presented in Chapter 3.2.4 and further evaluated in Chapter 5 and 6.

3.2.2 Flashing Based Processes

With the trend that more helium is produced from LNG streams, the world largest helium producer, APCI, invented a flashing based helium extraction process (Roberts & Repasky, 2007). This process is designed to recover helium from feed streams to an LNG plant that has less than 0.1 vol% He in the feed, which makes it suitable for LNG plants where LNG streams typically have a low concentration of He. The schematic of the process is illustrated in Figure 3.10.

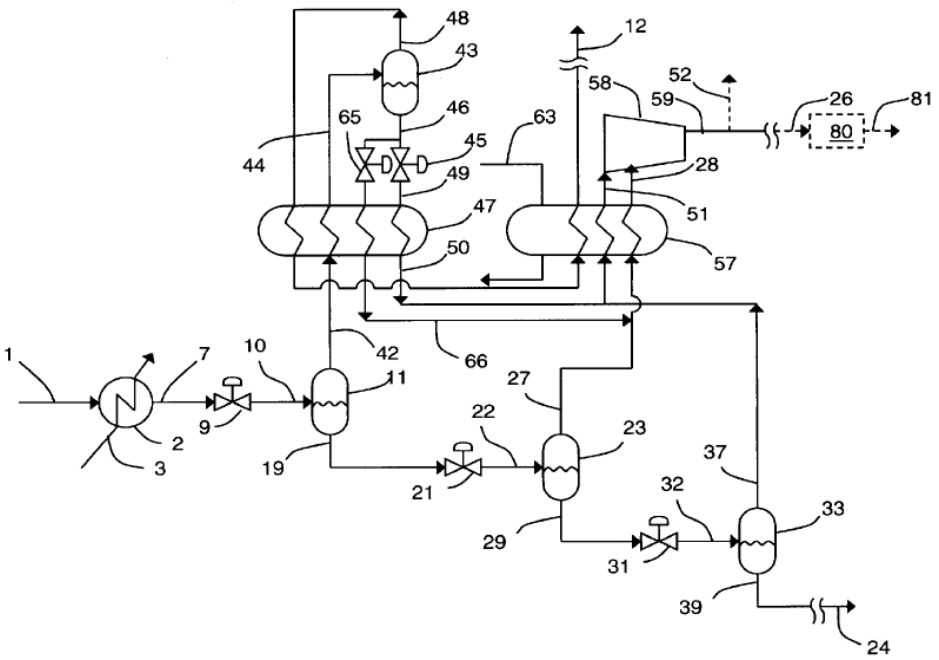


Figure 3.10 Process Flow Diagram of the APCI Flashing Process (Roberts & Repasky, 2007)

1	: Feed gas
2	: LNG liquefier
7	: Sub-cooled LNG
11,23,33	: Phase-separator
12	: Crude helium
24	: Final LNG product
43	: HeXU phase separator
47	: HeXU heat exchanger
57	: Cold Recovery heat exchanger
81	: Fuel gas

The process consists of three flashing stages (11, 23 and 33 in Figure 3.10), a HeXU (a heat exchanger 47 and a phase separator 43) for producing crude helium (12) and a heat exchanger (57) for cold recovery to produce extra LNG (63). The cooler (2) indicates an LNG liquefier and fuel gas (81) is recovered after being pressurised by a compressor (58). In this process, only the first flash vapour (42) is used for generating crude helium via a HeXU. Then the flash gas typically contains around 90% of the helium in a feed gas with large amounts of N_2 and CH_4. Other flash gases are utilised as fuel gas after having their cold energy recovered to produce extra LNG from the feed stream.

APCI suggested various configurations of the process in their patent in order to improve process efficiency. Figure 3.10 shows the most efficient configuration among others, which is the fifth embodiment of the patent. The main difference compared to other embodiments is found in its HeXU. The cold duty of the HeXU is supplied by a couple of pressure levels of condensed nitrogen and methane (46) through J-T valves (45 and 65) while other embodiments apply only one pressure level. This results in minimising temperature difference in the cryogenic heat exchanger in the HeXU, reducing its exergy loss. The fifth embodiment (Figure 3.10) is named as "APCI Flashing process" in this report for the sake of convenience.

Linde, who is the second largest helium manufacturer in the world, also developed a multi-stage flash cycle process for helium production, with a patent application (Bauer, Gwinner, & Sapper, 2009). The schematic of this technology is illustrated in Figure 3.11 and named as "Linde Flashing process" in this report.

Similar to the APCI Flashing process, the Linde process consists of three flashing stages (D1, D3 and D4 in Figure 3.11) and a HeXU (one heat exchanger (E) and two-phase separators (D5 and D7)). A LNG liquefier and a cold recovery system of gas products are not included in the configuration (Those were built in the HYSYS model for the Linde process). Stream 1 and 17 are sub-cooled LNG and final LNG product, respectively. The difference to the APCI process is that all the flash gases (2 and 6) except the last stage flash vapour are sent to a HeXU to recover crude helium. Linde claims that almost 100% of helium in the feed is collected and it helps to produce more and richer crude helium by using this measure (Bauer et al., 2009). Chapter 6.1 evaluates the Linde process and the APCI process in terms of crude helium production.

Another advantage of this technology is a sharp separation of helium and methane compared to multiple flash cycles. This is achieved by introducing a couple of extra flash cycles between the first and the last flash tank (D2). This leads to lower concentration of methane in crude helium compared to other flash-based helium extraction technologies. It is a favourable condition for the downstream process from an environmental point of view as less methane is purged to the atmosphere (Bauer et al., 2009). However, it is not viable to prevent the methane leakage to a very low level since a stripping step is required for deep separation of methane and helium.

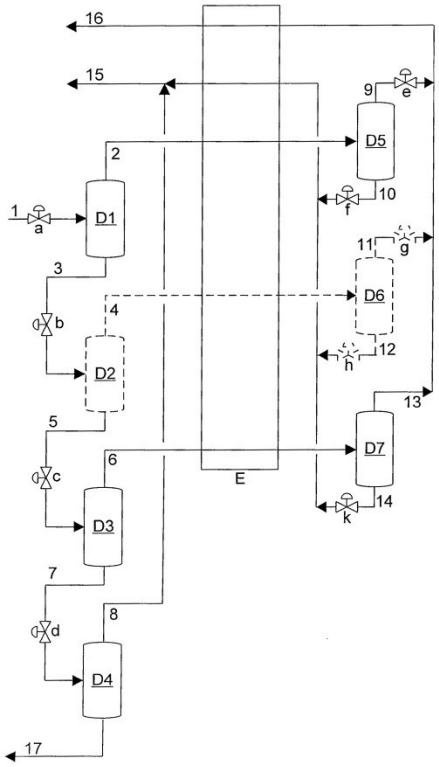


Figure 3.11 Process Flow Diagram of the Linde Flashing Process (Bauer et al., 2009)

1	: Sub-cooled LNG
2	: LNG liquefier
15	: Fuel gas
16	: Crude helium
17	: Final LNG product
D1-4	: Phase-separator
D5-7	: HeXU phase separator
E	: HeXU heat exchanger

In this project, the technologies from APCI and Linde were simulated and analysed to see their performances as representatives of helium extraction process using flash cycles.

3.2.3 Distillation Based Processes

APCI also developed a helium extraction process based on distillation with a condenser, filed as US 2007/0245771 A1 (Spilsbury, 2007). This is labelled as "APCI Distillation process" in this thesis. The schematic of this process is shown in Figure 3.12.

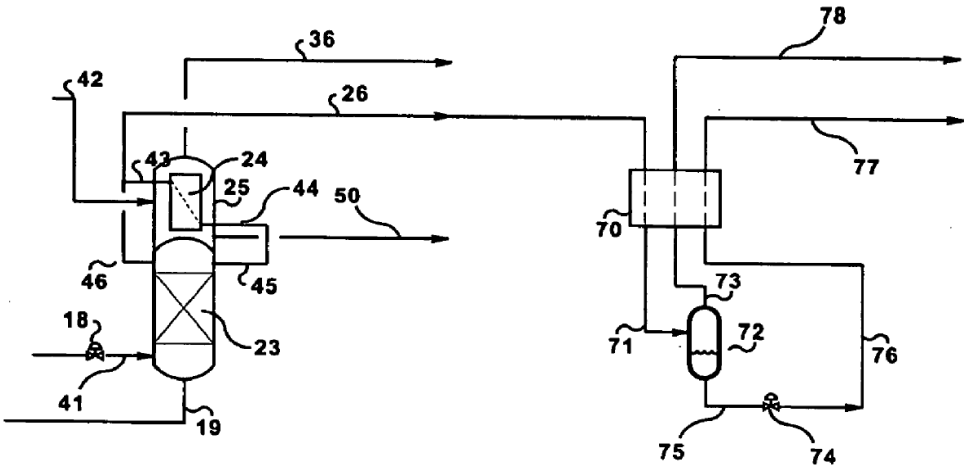


Figure 3.12 Process Flow Diagram of the APCI Distillation Process (Spilsbury, 2007)

23 : Dual-stage NRU stripping column
24 : Dual-stage NRU stripping column condenser
25 : Dual-stage NRU flash drum
26 : Dual-stage NRU overhead gas
36 : Fuel gas
41 : Throttled Sub-cooled LNG
42 : Further sub-cooled LNG
50 : Final LNG product
70 : HeXU heat exchanger
72 : HeXU phase separator
78 : Crude helium

It is a system where a dual-stage nitrogen removal column (23 and 25 in Figure 3.12) is integrated with a HeXU (one heat exchanger (70) and one phase separator (72)). Stream 41, 42 50 are depressurised LNG, sub-cooled LNG and final LNG product, correspondingly. This process is connected respectively to

an LNG plant as follows (see Figure 3.13). Relatively high temperature LNG (17) extracted from an LNG liquefier (18) is throttled to an intermediate pressure and passed through the first stage of the NRU (23), which is a distillation column to strip off helium and nitrogen. Next, the bottom product of the column is returned to the liquefier to be sub-cooled before being depressurised to near atmospheric pressure. The depressurised LNG enters the second stage of the NRU (25), which is a phase separator. It also works as a condenser for the distillation column (23). Then, the final LNG product (50) is produced from the separator after supplying the cold duty of the NRU.

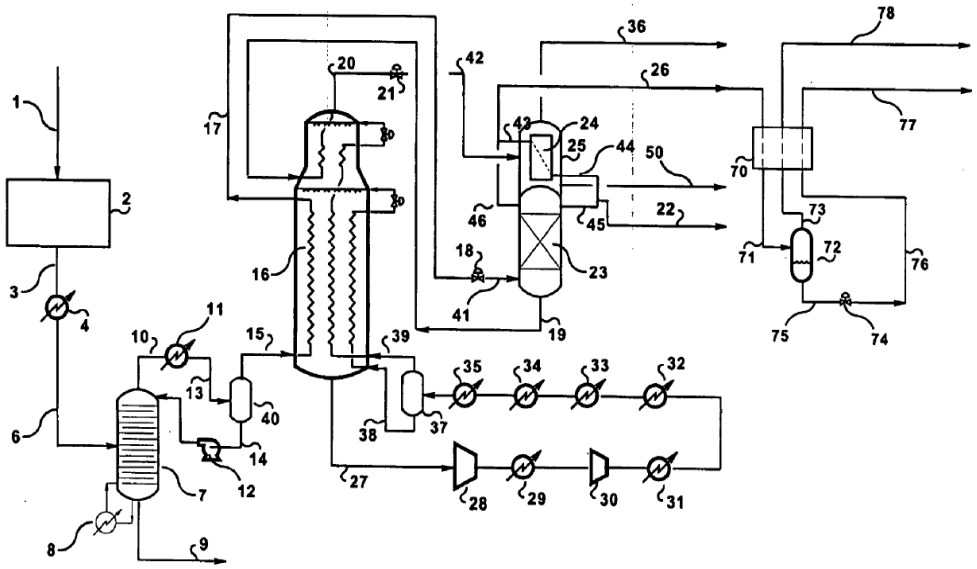


Figure 3.13 An Example of the Integration of an LNG Liquefier, NRU and HeXU

As explained above, the dual stage column is a system that a stripping column (23) and a flash tank (25) are attached in one column while having different pressure levels in each vessel. From the stripping column (23), which is a high-pressure section, helium and nitrogen rich gas (26) is collected as gas phase. This overhead product is then delivered to a HeXU where crude helium (73) is extracted. Unlike flash-based helium extraction processes, the helium rich gas (46) have very small amounts of methane. Consequently, the condensed liquid (75) in the HeXU is almost pure nitrogen and it can be vented (77) to the atmosphere after being used as a refrigerant.

The flash tank (25), which is a low-pressure section, works as a condenser for the high-pressure distillation column. Through the condenser, a portion of the nitrogen rich overhead (43) from the stripping column is condensed and refluxed to the column. Optionally, a portion of the liquefied nitrogen (LN_2) rich stream is extracted from stream 44. This LN_2 is an extra benefit as it is used as a refrigerant for helium purification/liquefaction and as a shielding material for liquefied helium transportation (Al-Harbi, 2014; Schmidt, 2009; Spilsbury, 2007). To produce liquefied N_2, normally an additional air separation unit is needed (Schmidt, 2009). The cooling duty of the condenser is supplied by sub-cooled LNG (42) and a small portion of the LNG is evaporated, which is utilised as fuel gas (36). Thus, there is no need for an external refrigeration cycle for the condenser.

This process is primarily designed for deep N_2 stripping from LNG and venting pure nitrogen. In general, the overhead gas from N_2 rejection columns are used as fuel gas for gas turbines and it normally contains a high concentration of nitrogen. The N_2 fraction becomes even higher when the feed gas has more nitrogen, reducing the heating value of the fuel gas and even exceeding the limitation of N_2 content in fuel gas (Rufford et al., 2012). Moreover, there is a demand for lowering the nitrogen content of a fuel gas for low NO_x burners in gas turbines, installed due to environmental regulations (Paradowski & Vovard, 2011; Spilsbury, 2007). However, the APCI Distillation process achieves N_2 lean fuel gas by removing a large amount of nitrogen from LNG by venting it to air and extracting it as liquid. Therefore, this invention may be an environmental friendly technology while achieving a proper heating value of the fuel gas.

Technip applied for a further improved helium extraction process based on distillation with two condensers and one reboiler (Paradowski & Vovard, 2011), called "Technip Distillation process" in this paper. Technip stated the fraction of N_2 in the fuel gas from the APCI Distillation process is still high as it is near the limitation of N_2 content in industrial gas turbines. Thus, they claimed their process produces a fuel gas containing less nitrogen. In addition, a N_2 vent stream from this process has a smaller amount of methane than the application from APCI, reducing the greenhouse effect of CH_4 and losses of CH_4 product. The schematic of the process is depicted in Figure 3.14.

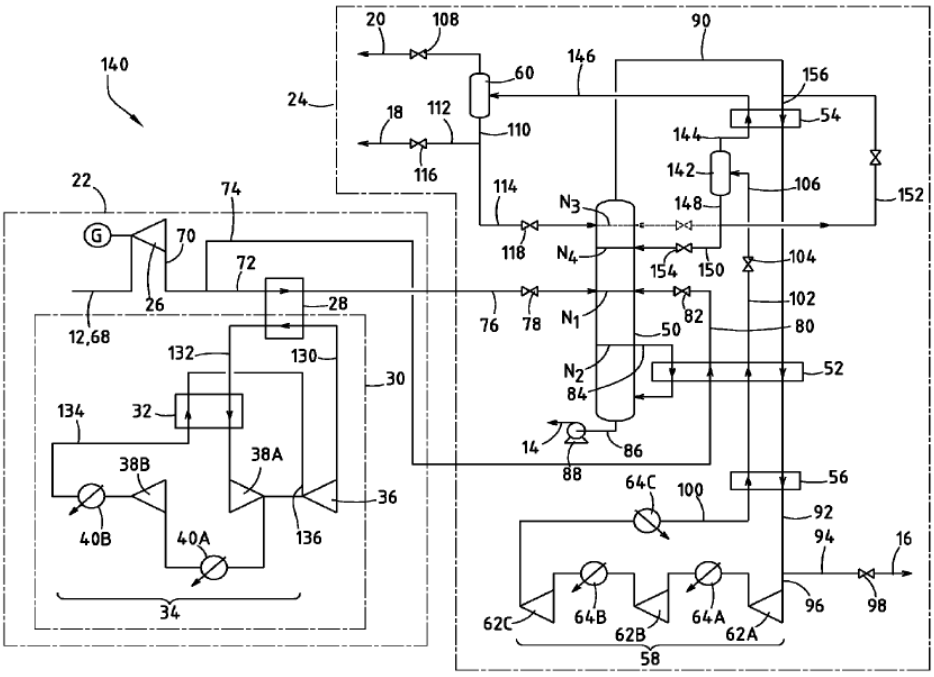


Figure 3.14 Process Flow Diagram of the Technip Distillation Process (Paradowski & Vovard, 2011)

12	: Sub-cooled LNG
14	: Final LNG product
16	: N_2 vent
18	: Liquid N_2
20	: Crude helium
30	: External refrigeration cycle
50	: NRU
52	: NRU reboiler heat exchanger
54	: NRU condenser heat exchanger
58	: NRU overhead gas recompression system
60,142	: NRU condenser phase separator

In contrast to the APCI Distillation process, it consists of a single column NRU (50 in Figure 3.14) with a reboiler (52) and two condensers (60 and 142). The overhead vapour (90) from the single column is heat exchanged through exchanger 54, 52 and 56 before a part of the vapour is vented via stream 94. The rest of the gas (96) is then compressed and partially condensed to supply the first reflux stream (148) to the column across the first condenser (142). The vapour (144) from the first condenser is again partially condensed through 54, which works as a heat exchanger in a HeXU. Then, the two-phase mixture is separated through the second condenser (60), producing crude helium gas (20) and condensed nitrogen (110), which is fed to the distillation col-

umn as another reflux streams. Thus, this separator plays the role of a phase separator in a HeXU as well.

The duties of the reboiler and the condensers are covered by complicated heat integration throughout the system as presented in Figure 3.14. However, unlike the APCI Distillation process, this invention requires extra refrigeration (30) for increasing the degree of LNG subcooling and a compression stage (58) for downstream processing. Thus, it offsets the advantages of this application by demanding more capital cost and consuming more energy.

The last helium extraction process based on distillation is a system utilising a distillation column with a reboiler. As there is no patent application for such process, the simulation model for this process was devised by author based on the APCI flashing process. This process is named as "Re-boiled Distillation process".

The process flow diagram of this technology is shown in Figure 3.15. Sub-cooled LNG coming from a LNG liquefier passes through another heat exchanger, which works as a reboiler for a distillation column in this process. After giving heat energy to the exchanger, the further cooled LNG stream is then depressurised by a liquid expander. Next, the two-phase stream leaving from the expander is supplied to the distillation column. The column produces a helium-rich overhead vapour, which is fed to a HeXU, and a helium-lean bottom liquid, which becomes final LNG product. Before the bottom liquid goes through two flashing cycles, the liquid is heated up by the reboiler where its heat duty is supplied by the sub-cooled LNG stream. The rest part of the Re-boiled Distillation process is identical to the APCI flashing process.

The reason for using the liquid expander instead of a J-T valve is to reduce the amount of vapour produced after depressurising the sub-cooled LNG stream. The vapour is the feed stream to a HeXU in this system and the reboiler already contributes generating a large amount of vapour as the feed to the HeXU. Therefore, if a J-T valve is used with the reboiler, a significant portion of the sub-cooled LNG will be evaporated, resulting in less final LNG production. Other features of this process will be handled in Chapter 5 and 6 after simulating this technology.

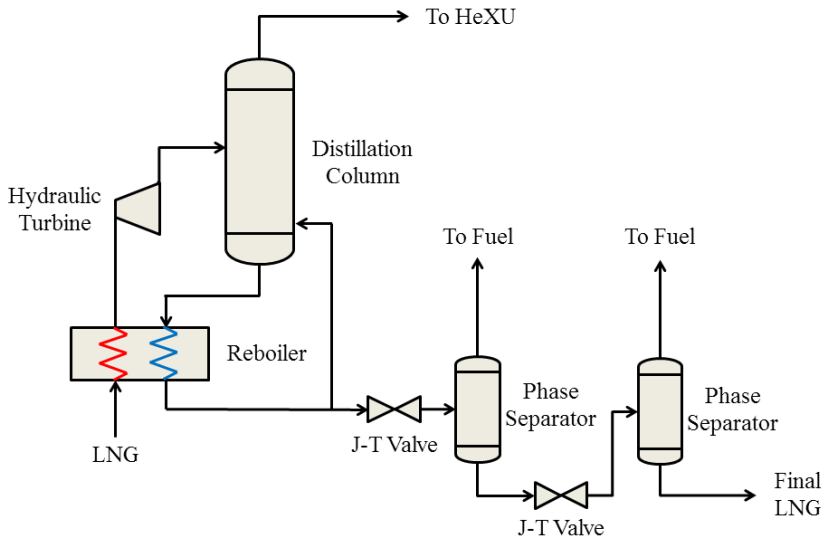


Figure 3.15 Process Flow Diagram of the Re-boiled Distillation Process

The three distillation based helium production processes introduced here largely differ from each other. Therefore, they are thoroughly simulated and evaluated in order to understand their strengths and weaknesses in depth, especially the Re-boiled Distillation process, and to compare with the flash-based inventions.

3.2.4 The Integration of Flashing and Distillation

To improve the performance of flashing based He extraction processes, Linde suggested a new configuration which integrates flashing with distillation (Schmidt, 2009). It is labelled as "Linde Integration process" in this report and depicted in Figure 3.16. This technology consists of a HeXU (a heat exchanger (E), a phase separator (D) and a distillation column (T)) for crude helium production and a heat exchanger (E') for LN_2 production.

Stream 1 is a flash gas from throttled sub-cooled LNG to an intermediate pressure. If the gas has lower than 15 bara, a compression stage is suggested before the flash gas is fed to this system. Stream 7 is a reflux to stream 1 to maximise crude helium production. Stream 13, 19 and 21 are fuel gas, N_2 vent and LN_2 product, respectively. Flashing of LNG, a LNG liquefier and a cold recovery system of the N_2 vent stream (19) and fuel gas (13) are not included in the configuration but they were built in the HYSYS model for the Linde distillation process.

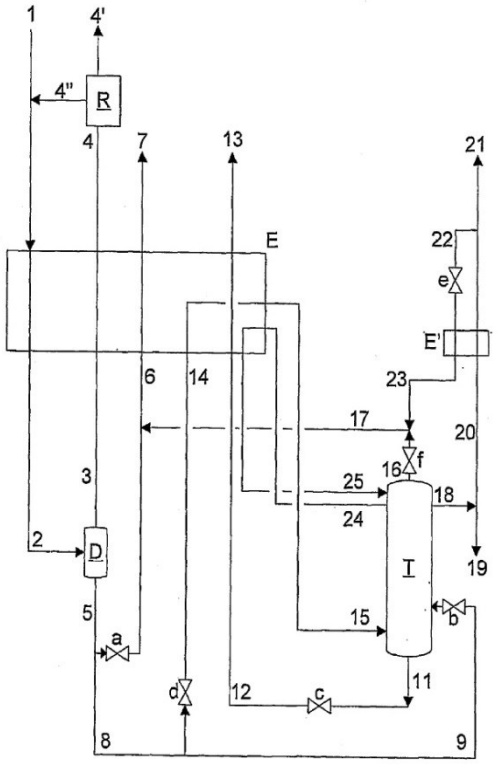


Figure 3.16 Process Flow Diagram of the Linde Integration Process (Schmidt, 2009)

1	: End-flash from an LNG plant
4	: Crude Helium
7	: Reflux to 1
13	: Fuel gas
18	: Liquid N_2
19	: N_2 vent
21	: Sub-cooled N_2
D	: HeXU phase separator
E	: HeXU heat exchanger/stripping column reboiler
E'	: Liquid N_2 sub-cooler
T	: HeXU stripping column

The main difference to typical flashing based processes is that this technology puts an extra distillation column in a HeXU. In case of a normal HeXU, the whole nitrogen and methane mixture condensed through a heat exchanger in the unit is sent back to the exchanger, working as a refrigerant. Meanwhile, this process conveys a portion of the condensate (8 in Figure 3.16) to a rectification column (T), which is heat-integrated with a heat exchanger (E) in or-

der to produce LN_2 (21), and a N_2 vent stream (19), which cannot be obtained by usual flashing based processes.

LN_2 is a valuable by-product because the liquefied inert gas is used for downstream processes as explained in the last paragraph on page 16. Thanks to this feature, the application does not require an extra ASU (air separation unit) to generate liquefied nitrogen from air. Linde also claims that this technology obtains a large enough quantity of liquefied nitrogen to liquefy the crude helium stream. It is achieved by subcooling the produced liquefied nitrogen before it is depressurised to near atmospheric pressure, thus minimising the amount of end-flash gas (Schmidt, 2009). Thus, the amount was checked and compared with other process producing LN_2 in Chapter 6.3.

Being able to discharge almost pure N_2 to the atmosphere is also a favourable characteristic. Similar to distillation based processes, it will help to obtain a fuel gas having less nitrogen content, thus producing less NOx emissions released from gas turbines. Besides, Linde indicates that this invention boosts crude helium production by recycling a portion of a nitrogen and methane rich stream after being used as refrigerant in a HeXU because the stream holds a considerable amount of helium.

One disadvantage of this technology is that the helium concentrated gas (1) entering this process has to be compressed to a high pressure, which needs extra compressors and electric power. However, it increases the operating pressure of the HeXU and the temperature needed for the partial condensation of a helium-enriched gas entering the unit. Therefore, the HeXU becomes more tolerant to larger amounts of impurities like CO_2 without freezing. This results in lower specifications of natural gas pre-treatment facilities, improving the project economics (Bauer et al., 2009; Oelfke & Victory, 2013).

Recently, ExxonMobil also published a patent application for producing crude helium, which combines flashing with distillation (Oelfke & Victory, 2013). It is named as “ExxonMobil Integration process” in this paper. The schematic of this technology is shown in Figure 3.17. This process is made up of two HeXUs, which are heat-integrated with each other. The first one includes a heat exchanger (202) and a distillation column (206), producing a gas (406) having higher helium concentration than the feed to this process and fuel gas (214). The second one consists of a heat exchanger (404) and a phase separator (414), producing crude helium (422) and fuel gas (416).

The feed stream (122) of this process is a helium-enriched flash gas from sub-cooled LNG throttled. Similar to the Linde Integration process, the feed gas needs to be compressed to 40 bara as the process provider advices. Thus, a multi-stage compressor was put to pressurise the feed gas when modelling this process in HYSYS. A LNG liquefier and the flashing of LNG are not contained in the configuration so these are also included in the HYSYS model of this process.

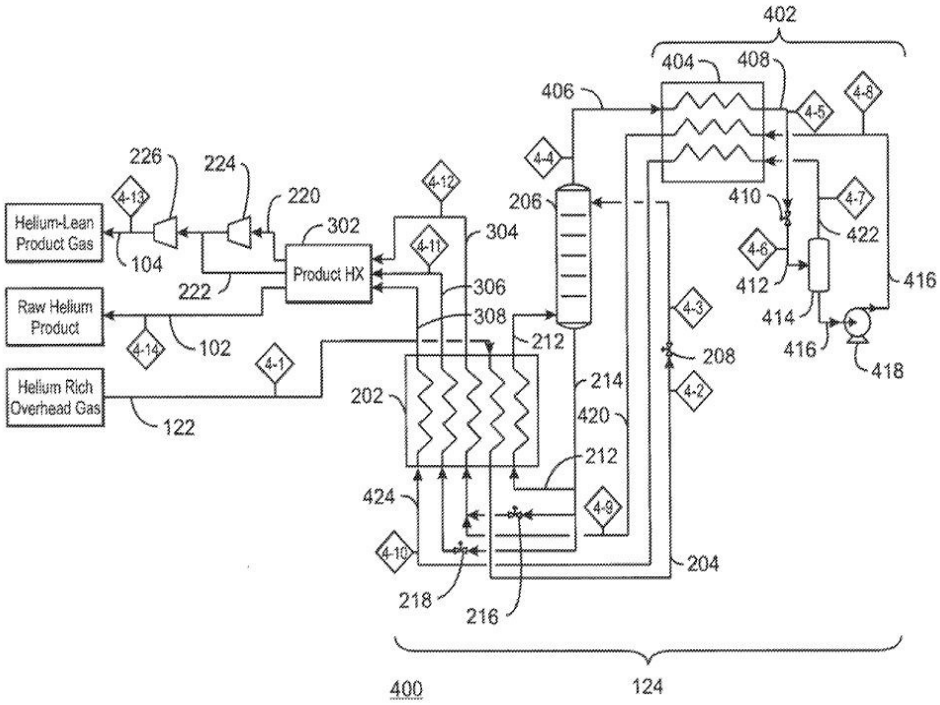


Figure 3.17 Process Flow Diagram of the ExxonMobil Integration Process (Oelfke & Victory, 2013)

102, 422	: Crude Helium
104	: Fuel gas
122	: End-flash from an LNG plant
124	: HeXU
206	: HeXU stripping column
202, 404	: HeXU heat exchanger/stripping column reboiler
302	: Cold recovery heat exchanger
414	: HeXU phase separator

This configuration shows different way of combining the two working principles to the APCI Integration process. First, the compressed feed gas to this process is partially condensed through a heat exchanger (202), like the Linde Integration process. However, the two-phase mixture (204) is delivered to a distillation column instead of being sent to a phase separation tank, hence obtaining an overhead product (406) in which helium is more concentrated. Afterwards the overhead gas is transported to the second HeXU and the vapour having a higher concentration in helium enables the unit to recover more crude helium.

As the whole process is fully heat-integrated, it does not require external heat input or refrigeration for the stripping column. In addition, ExxonMobil indicates that this integration improves the thermodynamic performance of the heat exchanger (202). Another striking feature claimed by ExxonMobil is that the working pressure of the whole process is higher, especially in the distillation column (206) and it increases the operating temperature of this system. It helps the column to accept more impurities before they freeze, which is an advantageous condition when designing natural gas pre-treatment units. Thus, this process was thoroughly analysed to verify the two features claimed by this technology provider in Chapter 6.

The two processes explained here are different in a way that flashing and distillation steps are merged, showing distinctive characteristics. Therefore, the two process applications were selected as representatives of this type of systems. Consequently, they were simulated and assessed to find the extra benefits that the integration-based technologies bring in comparison with flashing and distillation based ones.

4 Design Basis

When it comes to an objective evaluation of processes, it is vital to have the same basis of comparison for the processes studied as the conditions affect their performance. Especially when dealing with different types of processes, care should be taken to set the basis. Due to the characteristics of each process, it is not simple to put them on an equal playing field. Thus, the following text introduces the comparison basis applied in this project. As the principles of the six processes are different from each other, the set of conditions was manipulated to have a fair and consistent assessment.

First, the scope of the simulation work was limited by the purpose of this study. The system for producing liquefied helium from LNG consists of a number of process units as shown in Figure 3.1. However, the aim of this paper is to compare only the helium extraction related parts. Accordingly, some assumptions were made and the simulation scope was limited to focus on the relevant parts as follows.

- Natural gas pre-treatment step is outside of the scope.
- Helium purification and liquefaction steps are excluded.
- Natural gas liquefaction and subcooling is within the scope
- Flashing step or a NRU is included.
- Helium extraction unit is within the scope.
- Crude helium & fuel gas cold recovery are included.
- Crude helium & fuel gas re-compression are also within the scope
- Utility production is within the scope.

The LNG liquefaction and subcooling step are not of primary interest in this work. However, helium extraction processes are integrated with an LNG liquefier and the performance of a HeXU is influenced by the conditions of the LNG coming from the liquefier. As a result, a liquefier was included in the simulation models studied in the form of a simple cooler. Crude helium and fuel gas cold recovery and recompression stages were added to the models because the pressure and temperature levels of the products from a HeXU differ between inventions and this difference affects the efficiency of the helium extraction system. The cold energy exploited from the products was employed in producing sub-cooled LNG in the simulation work.

4.1 General Conditions

4.1.1 Feed Gas Conditions

Helium content in the feed gas was carefully chosen since it is one of the main factors that affects the performance and economics of a helium extraction facility. As explained in Chapter 2.4, helium was typically produced from natural gas containing 0.1-0.5 vol% of helium. In contrast, helium can be commercially produced by using a gas from a flashing step or an NRU in an LNG liquefier even though the feed stream has a very low He fraction. The flash gas having ten times higher helium concentration than the feed gas makes it possible. As an example, the two helium production plants recently built and integrated with an LNG plant in Qatar utilises a feed gas containing just 0.04 mole% He (Al-Harbi, 2014; Daly, 2005). Therefore, 0.05 mole% He was chosen as feed gas condition. This was to see whether the helium extraction processes studied are able to produce crude helium with such low content of He as other helium plants do in reality in Chapter 6.1. 0.02 mole% and 0.5 mole% He were also selected for the sensitivity analysis in also Chapter 6 as He-lean case and He-rich case, respectively.

Nitrogen concentration in the feed was decided by referencing the data about various helium bearing natural gases in the world. The N_2 content in the feed is also an important performance variable for a helium extraction process. Natural gases containing helium typically have a high concentration of N_2 as explained in Chapter 3.1. The fraction of nitrogen is roughly proportional to the helium fraction as indicated in Table 4.1. Thus, the N_2 fractions of the three feed cases were set with consideration for its helium mole fraction as shown in Table 4.2.

In addition, the feed gas was regarded as sweetened, dehydrated natural gas, containing small amounts of heavier hydrocarbons. Consequently, extra natural gas liquid (NGL) extraction unit and pre-treatment facilities were not required when modelling the helium extraction processes studied. While supplying the feed gas, it is also assumed that there is no change in pressure, temperature and composition. The feed gas had a fixed molar flow as well in order to know how much LNG each process could produce while recovering crude helium. The fixed amount is based on the required feed gas molar flow rate to produce 3MTPA LNG in APCI DMR process simulated in previous work (Kim, 2014). The detailed conditions of the feed are tabulated in Table 4.2.

Table 4.1 Composition of Various Helium-bearing Natural Gases (Al-Muhannadi, Okuyama, & Durr, 2001; Bouzid, Roche, & Coyle, 2010; Deaton & Haynes, 1961; Mukhopadhyay, 1980; Stolypin, Shakhov, Stolypin, & Mnushkin, 2006)

Location	Type of Plants	Feed Gas Composition [mole%]	
		N_2	He
Ras Laffan (Qatar)	LNG Production	4	0.04
Orenburg(Russia)	Gas Processing	4.8	0.05
Skikda(Algeria)	LNG Production	6	0.2
Odolanow (Poland)	Gas Processing	35	0.4
Otis (US)	Gas Processing	12.7	1.4
Keyes (US)	Gas Processing	35.8	3.1

Table 4.2 Conditions of Feed Gas

Property	Lean case	Base case	Rich case	Unit
Helium	0.02	0.05	0.10	mole%
Nitrogen	2.00	5.00	10.00	mole%
Methane	90.65	87.85	83.17	mole%
Ethane	4.88	4.73	4.48	mole%
Propane	1.69	1.64	1.55	mole%
n-Butane	0.40	0.34	0.32	mole%
i -Butane	0.35	0.38	0.36	mole%
i-Pentane	0.01	0.01	0.01	mole%
Pressure		60		bara
Temperature		35		°C
Flow rate		235,000		kgmole/hr

4.1.2 Ambient Air Temperature

An air cooling system was assumed to be used for heat rejection in this simulation work as more LNG base load plants have adopted this system instead of water cooling (Durr, Coyle, Hill, & Smith, 2005). The power consumed for fans on the exchangers was disregarded to make this project work simple. For this study, air-cooling system is assumed to be used.

The ambient air temperature was set to be 27°C as the helium extraction facility was assumed to be located in a warm region (Kusmaya, 2013). The air cools the closed loop coolant, which is water, down to 30°C through air-cooled heat exchangers so that the exchangers have 3 K of minimum temperature approach. Then the chilled closed loop coolant is applied to cool crude helium and the fuel gas down to 35°C via coolers. This results in a minimum pinch of 5 K in the coolers. The 5 K temperature approach is adopted from previous work (Kusmaya, 2013). Pressure drop across the coolers were neglected in this project. Detailed conditions of the cooling system are summarised in Table 4.3.

Table 4.3 Cooling System Conditions

Item	Value
Ambient air temperature	27°C
Closed loop coolant supply temperature	30°C
(Inter)Cooler discharge temperature	35°C
Min. temp approach of coolers	5 K
Pressure drop in coolers	0 bar

4.1.3 Driver Solution

All compressors in each process evaluated were assumed to be driven by electric motors. The fact that electric motor power output is not influenced by ambient air temperature eases the analysis of the extraction processes. It is also assumed that the fuel gas produced in the extraction processes, boil off gas and other fuel gas sources in the LNG facility is sufficient to supply the electricity needed for the electric motors and the LNG plant connected to a HeXU.

The power supplied to the motors is normally from gas turbine generators and the capacity and efficiency of these machines is strongly affected by the variation of ambient temperature. In spite of the importance of the temperature, the effect was disregarded in this project by fixing ambient temperature to avoid building excessively complex simulation models.

4.1.4 Plant Availability

Availability of the whole plant including an LNG plant and a helium extraction unit was set to 90%, which is 330days per year for all the processes studied. In reality, the availability may differ for the different process concepts. It depends on process configurations and the type of process equipment. However, this project does not aim to carry out an economic evaluation. Thus, the availability was not considered as a design factor. In addition, as mentioned in the previous chapter, compressor drivers were assumed to be electric motors and these have the highest availability among other drivers like gas turbines and steam turbines. Therefore, the availability of 90% would be a reasonable and even conservative value for this project.

4.2 Conditions for Modelling

4.2.1 Simulation Package and Equation of State

Aspen HYSYS V8.3 (HYSYS) was selected to simulate the proposed extraction processes. Peng-Robinson (PR) equation of state (EoS) property package was applied to model feed gas in HYSYS. The enthalpy calculation was performed by Lee-Kesler equation in the simulation tool.

4.2.2 Product Specifications

50 mole% helium concentration in a crude helium stream was the primary production target of this simulation work for a HeXU. Many references define that crude helium possess at least 40 mole% He, preferably more than 50 mole% He. (Al-Harbi, 2014; Daly, 2005; Emley & Maloney, 1997; Handley & Miller, 1992). If the noble gas fraction is lower than 50 mole%, the downstream helium purification facility may have too high load to handle larger amount of impurities, especially nitrogen as explained in Chapter 3.1. This causes the purification plant to consume more energy (Emley & Maloney, 1997).

The minimum helium extraction rate of 90% was also selected as crude helium production specification. In this thesis, helium extraction rate is defined as the amount of helium molar flow in crude helium divided by the one in the feed. The extraction rate varies depending on the type of He extraction processes. Most of the He extraction processes studied here are claimed to have a very high helium extraction rate. Thus, the extraction rate of 90% may be a good index to evaluate whether the technologies could achieve this rate with the base case feed gas condition introduced in Chapter 4.1.1, which is a similar to the feed gas conditions in the patent publications.

Depending on the feed gas conditions and the type of helium extraction process, it may be difficult to meet both the crude helium specification and crude helium production specification mentioned above. Thus, the top priority was given to achieving 50 mole% of helium in crude helium. It is due to the fact that the downstream process, which is helium purification step, is more sensitive to the change of helium mole fraction than crude helium flow rate as the process is normally designed for handling fixed compositions of a feed gas (Pettersen & Gundersen, 2014).

In case of the crude helium pressure level, 25 bara was assumed based on the helium extraction unit in Qatar, which has similar operating conditions to this project (Al-Harbi, 2014). This high pressure is suitable for the helium purification process as explained in Chapter 3.1. As high purification of helium is mainly achieved by a PSA, the crude helium needs to be pressurised before entering the purification step. Hence, the simulation models for the helium extraction processes studied include multi-stage crude helium compressors with intercoolers to meet this specification. Therefore, the pressure level of crude helium coming from a HeXU is an important performance parameter of the unit as it affects the amount of work needed for crude helium compression.

Crude helium temperature was set to 35°C in this project. Even though the raw gas is compressed after being extracted from a HeXU, the compressor discharge temperature keeps a relatively low level. It is thanks to the near isothermal compression of oil-flooded screw type compressors, which are typically used for helium compression. Detailed description is indicated in Chapter 4.2.3. In Table 4.4, the crude helium specifications are summed up.

Table 4.4 The Specifications of Crude Helium

Property	Value
Crude helium pressure	25 bara
Crude helium temperature	35°C
Helium content in crude helium	0.5 mole%
Helium extraction rate	≥ 90%

The sub-cooled LNG discharged from an LNG liquefier was assumed to have a temperature of -142°C. This temperature was acquired by simulating APCI DMR process to achieve a certain amount of end flash gas, which has around 10% of feed gas chemical energy. It was performed by referencing previous work (Kim, 2014). It is a typical amount of end flash, which is an enough quantity of fuel to provide the electricity need for an LNG plant (Durr et al., 2005). This temperature, however, was freely modified in each helium extraction process to meet the specifications of other products.

Together with the sub-cooled LNG, the final LNG product conditions are tabulated in Table 4.5. Final LNG was assumed to have 1.3 bara since the helium ripped LNG stream is J-T throttled to near atmospheric pressure for storage. As the characteristics of the He extraction processes are different from each other, the temperature of final LNG may not be identical for all the processes simulated. The temperature in Table 4.5 was based on the simulation result of the APCI Flashing process.

Another specification of final LNG is that nitrogen mole fraction has to be lower than 1% to prevent roll-over during transporting it by LNG carriers (Mokhatab et al., 2014). To meet this specification, some of helium extraction processes were allowed to modify their structure and stream conditions during simulation, for example the degree of LNG subcooling. Further details are explained in Chapter 6.

Table 4.5 The Specifications of Sub-cooled and Final LNG Products

Property	Value
Sub-cooled LNG pressure	60 bara
Sub-cooled LNG temperature	-142 °C
Final LNG pressure	1.3 bara
Final LNG temperature	-160.7 °C
N_2 content in final LNG	≤ 1 mole%

Fuel gas was assumed to have 20 bara and less than 130°C in this project. These are the typical values for industrial gas turbines as a fuel gas (Pettersen & Gundersen, 2014). Thus, multi-stage compression was added to pressurise the low-pressure fuel gas generated from a helium extraction process for the pressure specification. In addition, an extra cooler was placed at the outlet of the compression stage in the simulation models to reduce the fuel gas temperature to 35°C.

Nitrogen mole fraction in a fuel gas was also limited to maximum 40%, which is a typical design criterion for industrial gas turbine fuel (Bouzid et al., 2010; Toci, Nibbelke, & Bowtell, 2010). Higher N_2 fraction than this value may cause excessive NOx emissions from gas turbines, violating the environmental regulation as explained in Chapter 3.2.3. Thus, the concentration of N_2 in a fuel gas was taken as a performance indicator to figure out which process is the more environmentally friendly technology among other helium extraction processes. On the other hand, GE claims that the higher N_2 fraction a fuel gas has, the larger gas turbine power output one can have (Toci et al., 2010). Table 4.6 sums up the conditions for a fuel gas.

Table 4.6 The Specifications of Fuel Gas

Property	Value
Fuel gas pressure	20 bara
Fuel gas temperature	35 °C
N_2 content in fuel gas	≤ 40 mole%

Some of the helium extraction processes studied here releases a high purity nitrogen gas to the atmosphere and this wasted gas was controlled to have lower than 1 mole% HC in this project. This is a typical value to meet the environmental regulation, which is 0.5-1 mole% HC in vent gases (Wilkinson & Johnson, 2010). The needed pressure for the inert gas was set to higher than 1.3 bara in order to vent the gas to air. However, there was no limitation to the temperature level of the gas since it may vary a lot, depending on the structure of each process.

In addition, hydrocarbon content in LN_2 generated from a couple of the He extraction processes was limited to lower than 1 mole% as well. The reason for limiting hydrocarbon content in liquid nitrogen is to avoid the solidification of HC in the low temperature processes where LN_2 is used. In case of LN_2 pressure, it was let down to 1.3 bara to know that which process generates the most liquid nitrogen at its storage pressure between the helium extraction processes. The corresponding temperature of the stored LN_2 may be around the boiling point of pure nitrogen at that pressure, which is -193.62°C (Lemmon et al., 2013)

The purity of the liquid nitrogen was set to 99 mole%. This purity is lower than commercial LN_2 products as the produced LN_2 in this project was not meant to be sold commercially. Generally higher than 99.99 vol% of nitrogen is used as the specification for liquid N_2 in the market (Praxair, 2013). Table 4.7 summarises the conditions for the two nitrogen-based products.

The production rate of LN_2 was restricted to be higher than 5,500 liter/day. The helium liquefaction plant in Qatar consumes 15,000 liter/day LN_2 as a shielding material of helium containers for transport and a refrigerant for the helium treating process. As this plant handles around three times larger helium molar flow rate than the feed gas used in this project, 5,500 liter/day may be a proper production rate to cover the need for LN_2 for this project.

Table 4.7 The Specifications of N_2 Vent and LN_2

Property	Value
N_2 vent pressure	≥ 1.3 bara
N_2 vent temperature	N/A
Hydrocarbon in N_2 vent	≤ 1 mole%
LN_2 pressure	1.3 bara
LN_2 temperature	~ -193.62 °C
Hydrocarbon in LN_2	≤ 1 mole%
Purity of LN_2	≥ 99 mole%

4.2.3 Conditions for Equipment

Reasonable polytropic efficiencies were applied to compressors depending on the type of the machines. First, fuel gas compressors were assumed a centrifugal radial type, having polytropic efficiency of 78%. This is a slightly conservative value since the efficiency of centrifugal compressors deployed in cryogenic industry currently is reported to reach over 80% (Ransbarger, 2007).

For crude helium compression, oil-flooded rotary screw compressors were chosen. Its isothermal efficiency was set to 60% based on a reference (Arenius et al., 2006). The low molecular weight of helium makes centrifugal compressors less feasible, with a large number of compression stages required to reach a desirable pressure ratio. Hence the positive displacement type compressors have been the common solution for helium compression up to 24 bara (Koelet, 1997). Recent helium extraction facilities in Algeria and Qatar adopted this type of machine to compress crude helium (Al-Harbi, 2014; Froehlich & Clausen, 2007).

Before introducing how the screw compressor is modelled in HYSYS, it may helpful to explain about the working principle of oil-flooded screw compressors. The cold crude helium coming from a HeXU is mixed with oil in the compressor. The oil works as a lubricant, a sealant and a coolant in the compressor. The crude helium is heated up to an intermediate temperature due to compression and the temperature is kept constant by cooling from oil until the crude helium is discharged. This is thanks to the oil, which takes heat from the compressed crude helium and keeps the gas cool. Therefore, the oil allows

this compression to approach an isothermal process. (Koelet, 1997). After being discharged from the compressor, the oil is separated from the gas and it is cooled in an oil-cooler and re-injected to the compressor. The schematic is depicted in Figure 4.1.

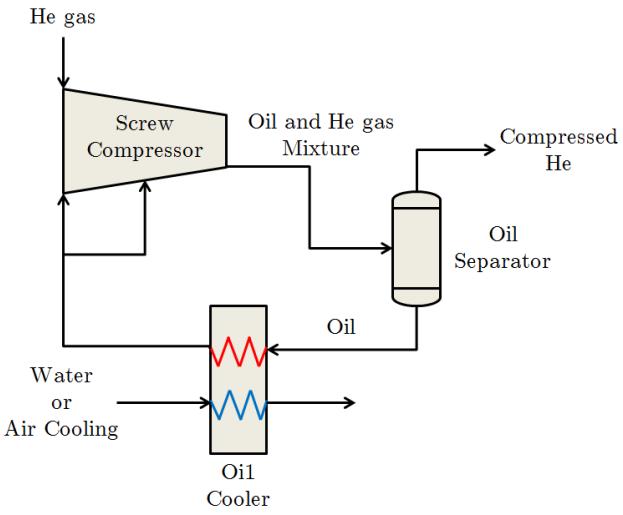


Figure 4.1 The Schematic of an Oil-flooded Screw Compressor

Such an isothermal compression process cannot be modelled straight away in HYSYS so it was roughly approximated by the following method. First, the cold crude helium coming from a HeXU is heated up to an intermediate temperature through a heater. This means the heating up of the oil and gas mixture in the compressor before the temperature is stabilised because of the oil cooling. Thus, the heat duty of the heater is ignored. The intermediate temperature was set to 60°C. Typically, oil-flooded screw compressors have a discharge temperature limit of 100°C due to material expansion problems (Koelet, 1997). Moreover, in this project, the compression ratio was expected to be just around two. Therefore, 60°C might be a reasonable value.

Then, the warmed crude helium is compressed through a compressor, which is modelled as a reciprocating type applying an adiabatic efficiency. The pressurised crude helium from the compressor is chilled via a cooler so that the chilled crude helium can have the same temperature as the compressor inlet temperature (60°C), which means isothermal compression. Here, the cooler works as the oil in the screw compressor. The coolers were counted as heat exchangers in this project as the heat absorbed in the oil has to be removed to ambient, eventually requiring extra coolers (Al-Harbi, 2014; Wilbert, 1998).

The adiabatic efficiency for the reciprocating compressor model was derived from the isothermal efficiency of 60%. Isothermal compression work at 60°C was calculated by using the isothermal efficiency. Next, the adiabatic efficiency of the reciprocating compressor was manipulated to have its compressor work the same as the isothermal compression work. Then, ca. 70% adiabatic efficiency for the reciprocating compressor was acquired. Several compressor and cooler stages may be added depending on the crude helium pressure level at the compressor inlet and the desirable discharge pressure. Furthermore, a cooler was put at the compressor outlet to cool the compressed crude helium. As mentioned in 4.1.2, this cooler is connected to an air-cooling system, which reduces the crude helium temperature to 35°C.

In case of other turbo machinery, a polytropic efficiency of 85% and an adiabatic efficiency of 75% were used for expanders and liquid turbines, respectively. Further assumptions were made to model hydraulic turbines in HYSYS. As the simulation tool does not have a model for hydraulic turbines, an expander model was used for it instead. The specification of the rotating equipment is summarised in Table 4.8.

Table 4.8 The Efficiency of Rotating Machinery

Equipment	Modelled As	Efficiency	
Fuel gas compressor	Centrifugal	Polytropic	78%
Crude helium screw compressor	Reciprocating	Isothermal	60%
Gas expander	Centrifugal	Polytropic	85%
Hydraulic turbine	Centrifugal	Adiabatic	75%

Table 4.9 introduces the types of cryogenic heat exchangers used. The heat exchangers were chosen as suggested by the process providers. In case of the Linde Flashing, APCI Distillation and Technip Distillation process, there is no suggestion for the type of exchangers. So, plate fin heat exchangers (PFHE) were assumed to be used for these processes. Since very small minimum temperature approach is needed (Wadekar, 2000). Coil wound heat exchangers could also be an alternative, but from a modelling point of view, this would not make any difference (as long as pressure drops are not included).

Table 4.9 Restrictions and Types of Cryogenic Heat Exchangers

Process	Type of heat exchangers	Min. Approach
APCI Flashing	PFHE/STHE	3 K
Linde Flashing	PFHE (Assumed)	3 K
APCI Distillation	PFHE (Assumed)	3 K
Technip Distillation	PFHE (Assumed)	3 K
Re-boiled Distillation	PFHE (Assumed)	3 K
ExxonMobil Integration	PFHE/STHE/SWHE	3 K
Linde Integration	PFHE	3 K

Minimum temperature approach in cryogenic heat exchangers was assumed 3 K for all the processes. The lower minimum temperature approach, the higher thermodynamic efficiency will achieved in cryogenic separation processes (Agrawal et al., 2000). However, it is not possible to have a heat exchanger having zero K pinch in reality. Zero K approach means the thermal effectiveness of the heat exchanger is equal to one. This requires infinite overall heat transfer coefficient, heat transfer area and pure counter current flow, which is infeasible to generate in reality (Wadekar, 2000). Therefore, the temperature approach in cryogenic heat exchangers was constrained to be 3 K for this project. This value was derived from previous work (Kim, 2014). In some cases, lower temperature approach may be applied for heat exchange at very low temperature level, e.g. in helium processes.

4.2.4 Other Constraints

To make simulation work simple, additional assumptions were introduced. First, pressure drop thorough interconnecting piping was neglected. Moreover, heat leak into heat exchangers and piping was assumed zero by supposing fully insulated pipelines and a HeXU coldbox.

5 Modelling and Results of Helium Extraction Processes

This chapter introduces the procedures for building simulation models for the processes studied. It also provides a process description for each technology, while indicating constraints and free variables for designing. Based on the variables, the methodologies for optimising the systems are presented as well. Patent applications of the processes were thoroughly reviewed in order to develop models in HYSYS. The conditions and assumptions used in the papers were revised and adapted into this project in order to have the same design basis for all the liquefaction systems. Chapter 4 explains the basis. The structures of the processes and equipment as illustrated in their patents were kept in the simulation models as far as possible.

5.1 Flashing Based Processes

5.1.1 APCI Flashing Process

The HYSYS model for the APCI Flashing was built by referring to the patent application, US 2007/0157662 A1 and Figure 3.10 (Roberts & Repasky, 2007). This process model was developed in HYSYS based on fifth embodiment in the application, which is the most efficient configuration. So far, there have been no technical papers about this system and detailed data showing the process conditions searchable yet. Due to the lack of information about the process, many design parameters used in the model were reasonably assumed.

The HYSYS model of the APCI Flashing process is illustrated in Figure 5.1. In this modelling work, base case feed gas indicated in Chapter 4.1.1 was used as feed. First, treated feed gas at 35°C and 60 bara (Stream "Feed" in Figure 5.1) is split into stream F1 and R1. Stream F1, which has most of the feed molar flow, is sent to an LNG liquefier (LNG Liquefier). Through the liquefier, sub-cooled LNG (F2) is produced at -145.5°C. This temperature was initially -142°C as design basis mentioned in Chapter 4.2.2. Then it was adjusted to have 1 mole% N_2 in the final LNG product (LNG) after passing through a multi-stage flashing step in this process.

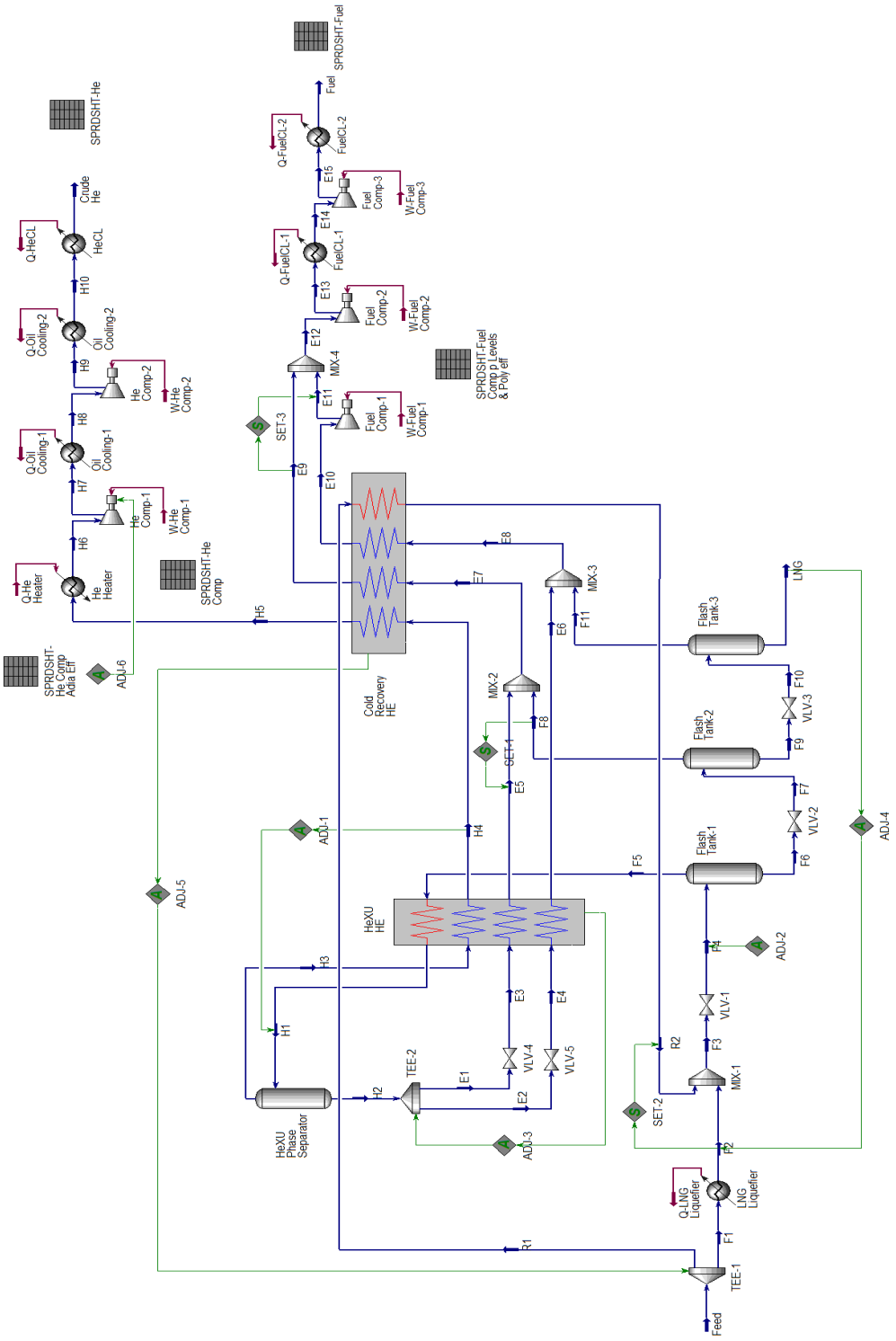


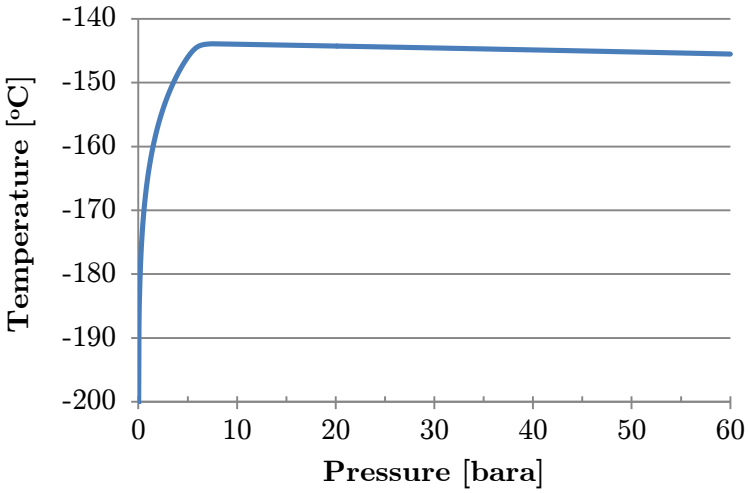
Figure 5.1 The HYSYS Model for the APCI Flashing Process

Stream R1 having a small portion of the feed is piped to a heat exchanger for cold recovery (Cold Recovery HE). Then, stream R1 is liquefied and sub-cooled to the same temperature as F2 in the heat exchanger. The amount of stream R1 was manipulated to meet the 3 K minimum temperature approach of the exchanger. The amount of stream R1 was around 5% of the feed molar flow and thus stream F1 had ca. 95% of the feed molar flow. As this process exploits cold from the gas products of this process, crude helium and fuel gas, the cold energy is used for liquefying stream R1.

Next, the two sub-cooled streams, F2 and R2, are merged in stream F3 and it is throttled to 5.084 bara by the first J-T valve (VLV-1) to generate first end-flash (F5), which is the feed for a HeXU. The depressurised LNG is then separated into helium-stripped LNG (F6) and end-flash (F5) in a phase separator (Flash Tank-1). The outlet pressure of VLV-1 was controlled to reach a helium extraction rate of 90%, which is one of the specifications introduced in Chapter 4.2.2.

The outlet pressure of VLV-1 is an important factor when it comes to J-T effect. Depending on the pressure level, one could have either a heating or a cooling effect through the J-T valve. Figure 5.2 shows the effect of the isenthalpic change in the pressure of F3 stream on its temperature level. The isenthalpic pressure change means a Joule-Thomson process, which is the working principle of J-T valves.

Based on Figure 5.2, the -145.5°C and 60 bara sub-cooled LNG (F3) may have a cooling effect through a J-T valve if it is depressurised to lower than 5.22 bara. Otherwise, the expanded sub-cooled LNG may be heated up even though its pressure level is decreased. In this simulation model, the sub-cooled LNG is throttled to lower than that pressure to meet the specification of helium extraction rate, thus having a cooling effect via VLV-1. Nevertheless, the J-T valve may not deliver a cooling effect depending on the process conditions such as changed feed gas composition.



*J-T Valve Inlet condition: -145.5°C at 60.00bara
*J-T Valve Cooling Effect: from 0 bara to 5.5bara

Figure 5.2 Constant Enthalpy line of the Sub-cooled LNG (Stream F3)

The helium-stripped LNG (F6) is depressurised to 3.1 bara by the second J-T valve (VLV-2) and the second flash gas (F8) is gathered from the expanded stream (F7) through Flash Tank-2. In this process, the outlet pressure of the VLV-2 is only one free variable that affects the efficiency of this process. By performing a case study with this independent variable, 3.1 bara was selected as it gives the lowest energy consumption for the crude helium and fuel gas compression. The detailed information is given in Appendix A. Finally, the last flash gas (F11) and final LNG product (LNG) are acquired from the third phase separator (Flash Tank-3) after the LNG (F9) from the second phase separator is expanded to 1.3 bara, which is a typical pressure of LNG run-down.

Looking at the HeXU part, the first flash gas from Flash Tank-1 is cooled down to -180.9°C and partially condensed via a heat exchanger in the HeXU (HeXU HE). The condensing temperature was decided to achieve the helium mole fraction of 50% in the vapour of the partially condensed stream (H1). Afterwards, the condensed stream goes through a phase separator (HeXU Phase Separator), generating a helium-rich gas (H3) and a helium-lean condensate (H2). Stream H3 is directly re-sent to HeXU HE to supply the cold duty of the heat exchanger. In case of stream H2, it is split into two streams, E1 and E2, and they are depressurised to different pressure levels before being piped and exploited their cold energy to the heat exchanger. Stream E1 is

combined with the second flash gas (F8) after being throttled. Thus, Stream E1 is depressurised down to the pressure level of stream F8, which is 3.1 bara. Similarly, expanded stream E2 is merged with the third flash gas (F11) so stream E2 is throttled to 1.3 bara, which is the pressure level of stream F11.

By controlling the flow split between stream E1 and E2, 3 K minimum pinch of HeXU HE was obtained. It was possible as the two expanded liquids (E3 and E4) have different temperatures by being expanded to distinctive pressure levels. In addition, these two cold energy sources for the heat exchanger with different temperature levels result in a smaller gap between the cold composite and the hot composite curve of the heat exchanger, compared to typical HeXUs using an identical pressure level of the liquids as coolant. In this simulation model, the split ratio was 83% and 17%, respectively. The corresponding LMTD was 6.2 K. In contrast, the LMTD of 9.5 K was recorded in the configuration utilising only one pressure level of the two liquids for HeXU HE cooling duty.

Next, the warmed helium-rich gas (H4) and the two warmed expanded condensates streams (E5 and E6) discharged from HeXU HE are delivered to another heat exchanger (Cold Recovery HE). This exchanger is for recovering their remaining cold energy and for sub-cooling a part of the feed gas (R1). Before being transported to Cold Recovery HE, stream E5 and E6 join with the second and the third flash gas, producing stream E7 and E8, respectively. In contrast, stream H4 is headed to the heat exchanger directly. As mentioned earlier, the constraint of 3 K minimum temperature approach for the heat exchanger was fulfilled by adjusting the molar flow rate of stream R1.

After collecting the rest of the cold energy in Cold Recovery HE, the helium rich gas is compressed from 5.1 bara to 25 bara, which is the crude helium specification. This compression is performed by a two-stage screw compressor and theoretical compression pressure ratio per stage[1] that gives the smallest compression work was calculated based on the two pressure levels. In this case, the pressure ratio was 2.2 per stage. Finally, crude helium is produced by chilling the compressed helium-rich stream to 35°C via a cooler (HeCL). The detailed description of this compression process is explained in Chapter 4.2.3.

[1] Theoretical Compression Ratio per Stage
= (Highest Pressure/Lowest Pressure)$^{(1/\text{Compression Stage})}$

The two streams E7 and E8 are also warmed up through Cold Recovery HE and they are combined in a mixer (Mix-4) after the warmed E8 is pressurised by a compressor (Fuel Comp-1) to reach the pressure level of E7, which is 3.1 bara. The mixture of warmed stream E7 and E8 is used as fuel gas in this project. To meet the fuel gas specifications, the mixture (E12) is compressed to 20 bara through two-stage compression. The compression ratio per stage is also obtained by the theoretical compression ratio equation. The ratio was 2.6 in case of this process.

5.1.2 Linde Flashing Process

The simulation model of the Linde Flashing process was also established by merely referring to their patent application, DE 10 2007 047 147 A1 and Figure 3.11 (Bauer et al., 2009). As Linde does not provide all the stream conditions and design constraints of their process, this process was modelled by using the design basis adopted in this thesis. Base case feed gas was also utilised as feed for this HYSYS model. The schematic of the HYSYS model is shown in Figure 5.3. The configuration of this process is similar to the APCI Flashing process.

First, feed gas is separated into stream F1 and R1 before they are liquefied and sub-cooled in LNG Liquefier and Cold Recovery HE, respectively. The separate ratio between F1 and R1 was manipulated to have 3 K minimum temperature approach in Cold Recovery HE. The heat exchanger is for cold recovery of the low temperature fuel gas and crude helium produced from a HeXU in this process. The stream R1 fraction was 5.2% of the feed stream. The two sub-cooled streams were set to have the same temperature of -145.7°C and they are merged in stream F3. This temperature was controlled to achieve the typical LNG transport specification, 1 mole% N_2 in LNG in this simulation model.

Then, the sub-cooled feed stream F3 passes through three flashing steps before being stored at near atmospheric pressure. Firstly, F3 is expanded through a J-T valve (VLV-1) to 4.9 bara, producing the first flash gas (F5) and the first helium-ripped LNG (F6). This pressure was set to have 2 mole% He in the first flash gas as it gives the best performance of this system. The mole fraction is only one free variable of this process, affecting the performance. Thus, a case study with the variable was performed and it is explained in Appendix B.

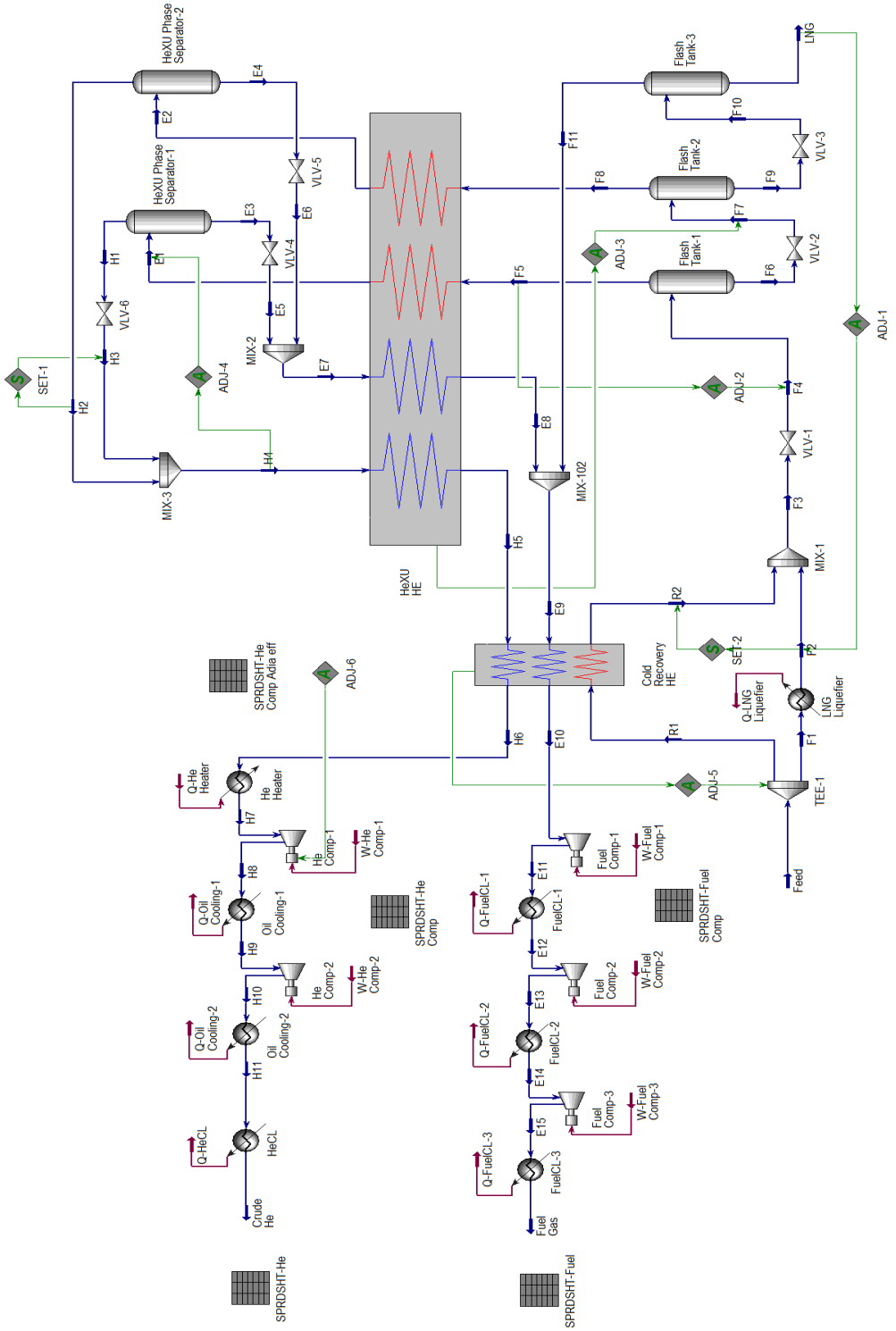


Figure 5.3 The HYSYS Model for the Linde Flashing Process

Stream F6 is again J-T expanded to 1.9 bara by VLV-2 and the throttled stream is separated into the second flash gas (F8) and the second helium-ripped LNG (F9) via Flash Tank-2. The outlet pressure of VLV-2 was controlled to have a minimum pinch of 3 K in a heat exchanger in the HeXU (HeXU HE). Finally, stream F9 is throttled to 1.3 bara to produce a part of fuel gas (F11) and LNG for storage.

Next, both the two flash gas streams (F5 and F8) are delivered to the HeXU to produce crude helium (the APCI Flashing process sends only the first flash gas to the HeXU). Stream F5 and F8 are chilled to -183.3°C in the heat exchanger in the HeXU (HeXU HE). With the temperature, the streams are partially condensed and the each stream is separated into a helium-rich gas and a helium-lean liquid by HeXU Phase Separator-1 and 2. Then the first helium-rich gas (H1) coming from HeXU Phase Separator-1 is depressurised to the pressure level of the second helium-rich gas (H2), which is discharged from HeXU Phase Separator-2.

Afterwards, the expanded stream H3 is combined with stream H2 in a mixer (MIX-3) to produce crude helium (H4). Thus, the condensing temperature of -183.3°C was selected to reach 50 mole% He in stream H4 as a crude helium specification introduced in Chapter 4.2.2. Stream H4 is then returned to HeXU HE to supply its cold duty. In case of the two helium-lean liquids (E3 and E4), they are expanded to 1.3 bara and re-sent to HeXU HE after being mixed as stream E7. The reason for throttling the two streams is to provide the cold duty of HeXU HE and to be merged with the last flash gas (F11), which has the pressure level of 1.3 bara.

In HeXU HE, the cold energy of the helium-rich gas (H4) and the helium-lean liquid (E7) are exploited and stream E7 is completely evaporated. The evaporated stream E8 is then blended with the last flash gas (F11) to form fuel gas (E9). The fuel gas and the helium-rich gas (H5) warmed in HeXU HE are delivered to Cold Recovery HE. The cold energy of stream E9 and H5 is used for liquefying and sub-cooling a portion of the feed (R1). As explained earlier, the amount of the feed sent to this exchanger was adjusted to fulfil the specification of cryogenic heat exchangers used in this project, 3 K of minimum temperature approach.

Finally, the helium-rich gas (H6) discharged from Cold Recovery HE is compressed from 1.9 bara, which is the pressure level of the second flash gas to the pressure level of crude helium specification, to 25 bara. It is conducted by a two-stage screw compressor to have a proper compression ratio, which gives high compressor efficiency. The pressure ratio of the each stage was calculated by the theoretical compression ratio equation. The ratio was 3.59, which is within the typical boundary of pressure ratio for screw compressors. Thus, if a single-stage screw compressor is used, it may require too large compression ratio, resulting in significant reduction in compressor efficiency. As the compressed crude helium has a high temperature at 60°C, it is chilled down to 35°C by a cooler (HeCL).

Similarly, the fuel gas (E10) coming from Cold Recovery HE also passes through a three-stage centrifugal compressor to be pressurised from 1.3 bara to 20 bara, which is one of the fuel gas specifications set in this paper. The pressure ratio for each compression stage, which is acquired by the theoretical compression ratio equation, was 2.49 in this process. Between compression stages, intercoolers (FuelCL-1 and 2) were assumed and an extra cooler (FuelCL-3) was installed at the outlet of this compression to cool the compressed fuel gas down to 35°C.

5.2 Distillation Based Processes

5.2.1 APCI Distillation Process

The APCI's crude helium production system based on distillation was developed by referring to their patent application US 2007/0245771 A1 and Figure 3.12 (Spilsbury, 2007). As APCI provides various embodiments for nitrogen and helium extraction in the application, the fifth embodiment was chosen because this embodiment is the only one producing crude helium. The design conditions used in the patent application are different from this thesis. Thus, various assumptions were made and the design basis of this project was applied to build a HYSYS model for the APCI Distillation process. The schematic of this process is depicted in Figure 5.4.

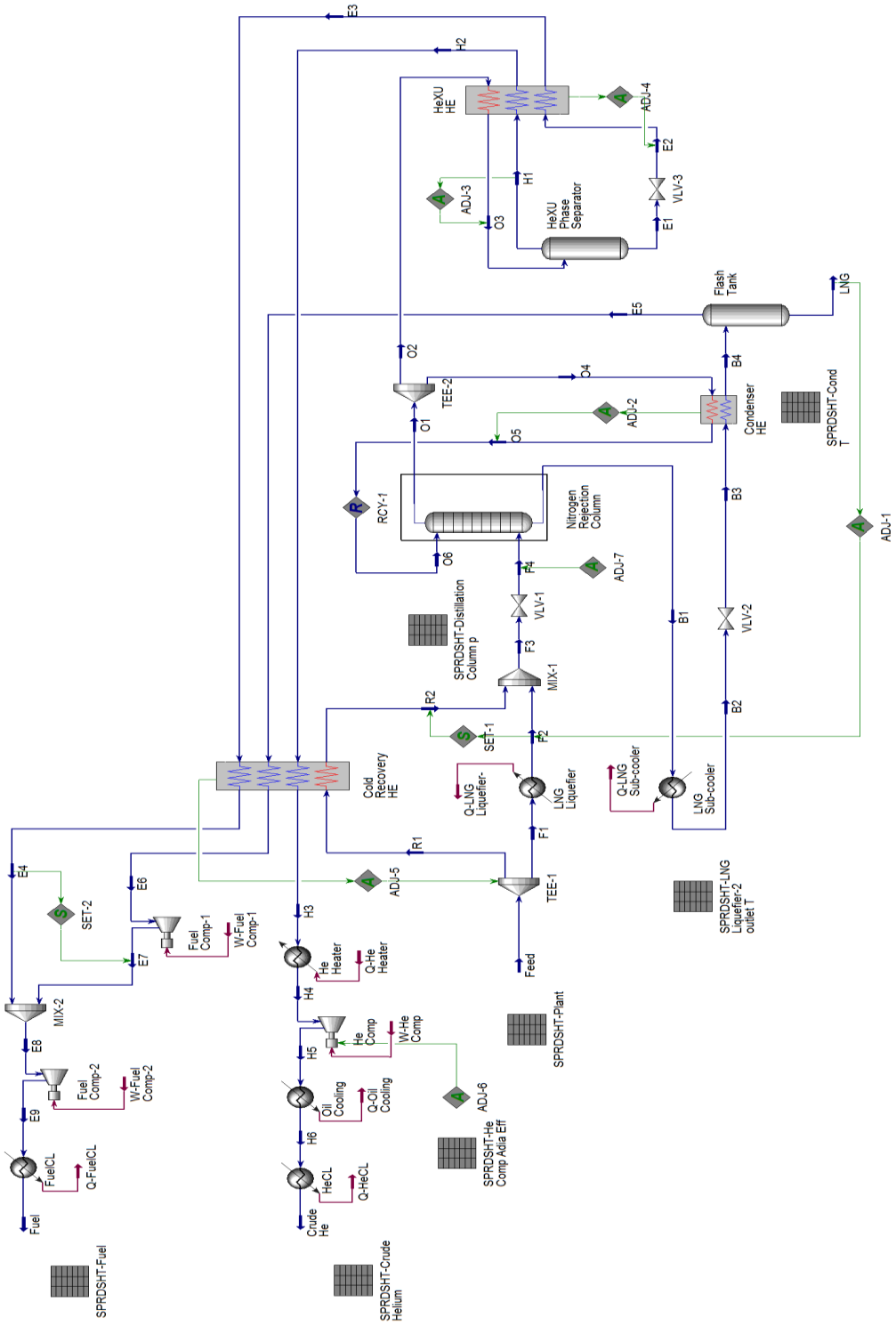


Figure 5.4 The HYSYS Model for the APCI Distillation Process

Similar to the two flashing based processes studied in previous chapters, this process starts with splitting a feed stream (stream "Feed" in Figure 5.4) into stream F1 and R1. Stream F1 is delivered to a cooler (LNG Liquefier) working as the pre-cooling, liquefaction and sub-cooling part of an LNG liquefier and discharged at -137°C to be sub-cooled. Stream R1 is also chilled to the discharge temperature of LNG Liquefier before being mixed with stream F2 leaving LNG Liquefier. The discharge temperature of the cooler was controlled to obtain 1 mole% N_2 in the final LNG product (LNG).

The mixed stream F3 is then expanded to 4.5 bara via a J-T valve (VLV-1) to provide a feed stream for a nitrogen rejection column in this HYSYS model (Nitrogen Rejection Column). Consequently, the working pressure of the column was 4.5bara. The outlet pressure of VLV-1 was adjusted to achieve a helium extraction rate of 90%. Even though APCI proposes 15 bara for the outlet pressure of VLV-1, one cannot have LNG containing less than 1 mole% N_2 with this pressure level. Therefore, the outlet pressure of VLV-1 was further reduced than what APCI suggests to flash off more nitrogen in stream F3.

The throttled stream F4 is separated to a bottom liquid B1 and an overhead vapour O1 in the N_2 rejection column. The overhead stream O1 is split to two streams, stream O2 and O4. Stream O2 is used as feed for a HeXU in this process and the rest of the overhead vapour (O4) is sent back to the column after being condensed in a heat exchanger (Condenser HE). The condensing temperature of stream O4 was adjusted to have 3 K minimum temperature approach of Condenser HE and the temperature was -158.6°C. The split between stream O2 and O4 is the main free variable of this process. Therefore, a case study was conducted to find the optimal split. The result of the case study indicated that the split ratio of 50% and 50% gives the best performance of this process (Therefore, the split ratio of fifty-fifty was selected for this process model). The detailed analysis was explained in Appendix C.

The bottom product B1 is sent to another cooler (LNG Sub-cooler), which works as the sub-cooling part of an LNG liquefier and the sub-cooled stream B2 leaves the cooler at -155°C. The sub-cooling temperature was the value APCI suggests in their application. Next, the sub-cooled stream B2 is depressurised to 1.3 bara by a J-T valve (VLV-2) and delivered to Condenser HE to supply the cold duty of the condenser. The warmed stream B4 through the condenser is finally transported to a phase separator (Flash Tank) to generate an LNG product (LNG) and end-flash (E5).

Stream O2 is a helium-containing vapour, which is utilised for producing crude helium in downstream processing. First, stream O2 passes through a heat exchanger (HeXU HE) to be partially condensed. The condensed stream O3 is discharged from the exchanger at -182.8°C and it is separated to a helium-rich vapour H1 and a helium-lean liquid E1 in HeXU Phase Separator. The outlet temperature of HeXU HE was controlled to achieve 50 mole% He in the helium-rich gas H1. Stream H1 returns to HeXU HE in order to cool stream O2. The helium-lean liquid E1 is also sent back to HeXU HE to supply the cold duty after being throttled to 1.7 bara by J-T expansion. The throttling pressure was manipulated to obtain 3 K minimum pinch of HeXU HE.

The two warmed streams H2 and H3 enter another heat exchanger (Cold Recovery HE) via HeXU HE with the end-flash stream E5 to liquefy a part of the feed stream (R1). The amount of stream R1 was adjusted to fulfil the 3 K minimum temperature approach specification for heat exchangers in this project. The flow rate of stream R1 was 5.1% of the feed in this simulation model, reducing the cold duty of LNG Liquefier.

Stream E4 and E6 leaving Cold Recovery HE form fuel gas by several compression stages. First, stream E6 goes through a single-stage centrifugal compressor (Fuel Comp-1) to reach the pressure level of stream E4. Stream E7 discharged from Fuel Comp-1 is then mixed with stream E4 and the mixture E8 is again compressed to 20 bara by another single-stage centrifugal compressor (Fuel Comp-2), which is the specification of fuel gas. The compressed fuel gas E9 is cooled down to 35°C through a cooler (FuelCL-3) as the temperature is another specification of fuel gas in this project.

Stream H3 discharged from Cold Recovery He is also compressed to 25 bara by a single-stage oil-flooded screw compressor (He Comp) and chilled to 35°C via a cooler (HeCL) to become a crude helium stream (Crude Helium).

5.2.2 Technip Distillation Process

The HYSYS model for the Technip Distillation process was developed by referring to the patent application US 2011/0226009 A1 and Figure 3.14 (Paradowski & Vovard, 2011). Especially this simulation model was built based on the second installation in the application, which is the simplest helium extraction method between alternative configurations. As this process is based on a sophisticated distillation column, most of the stream conditions and design factors that Technip offers in the installation were directly adopt-

ed to build this simulation model. However, some design features were adjusted due to the different feed gas conditions. The schematic of this process is illustrated in Figure 5.5.

In this process, a feed gas stream (Feed in Figure 5.5) at 35°C and 60 bara passes through an LNG liquefier (LNG Liquefier) to be liquefied and sub-cooled to -149.5°C. The sub-cooling temperature was adopted from the patent application. Then, the sub-cooled stream F1 is let down to 6 bara by a liquid expander (Liquid Expander), producing electrical energy. The outlet pressure level of the liquid expander was also the value used in the patent application. The expanded stream F2 is separated to stream F3 and R1. Again, stream F3, which is the larger portion of stream F2 enters another refrigerator (External Refrigerator) and stream F4 leaves at -160°C, which is a design condition suggested by Technip. The lesser portion of stream F2 (R1) is conveyed to a heat exchanger (Main HE) and stream R2 is discharged at the temperature level of stream F4.

The two cold streams F4 and R2 are combined as stream F5, which is a main feed stream to a nitrogen rejection column (Nitrogen Rejection column) in this process. Before being fed to the N_2 rejection column, stream F5 is throttled to 1.3 bara, which is the working pressure of the column. The working pressure is the value Technip suggests. The throttled stream F6 is then introduced the intermediate tray of the column and separated to the bottom liquid B1 and the overhead vapour O1.

The bottom product B1, which is LNG, is slightly re-heated in Main HE to boil off nitrogen content down to 1 mole% before it is stored. The re-warmed stream B2 is separated in a tank (Reboiler Phase Separator) to a boil-off stream B3 and a final LNG product (LNG). The boil-off stream B3 is re-introduced to the nitrogen rejection column to strip nitrogen and helium in stream F5, which is the feed stream to the column. Thus, Main HE and Reboiler Phase Separator work as a reboiler of the nitrogen rejection column in this process. The re-boiling temperature that is the temperature level of stream B2 was adjusted to make the final LNG product to have 1 mole% nitrogen. In this process, the temperature was -160.8°C.

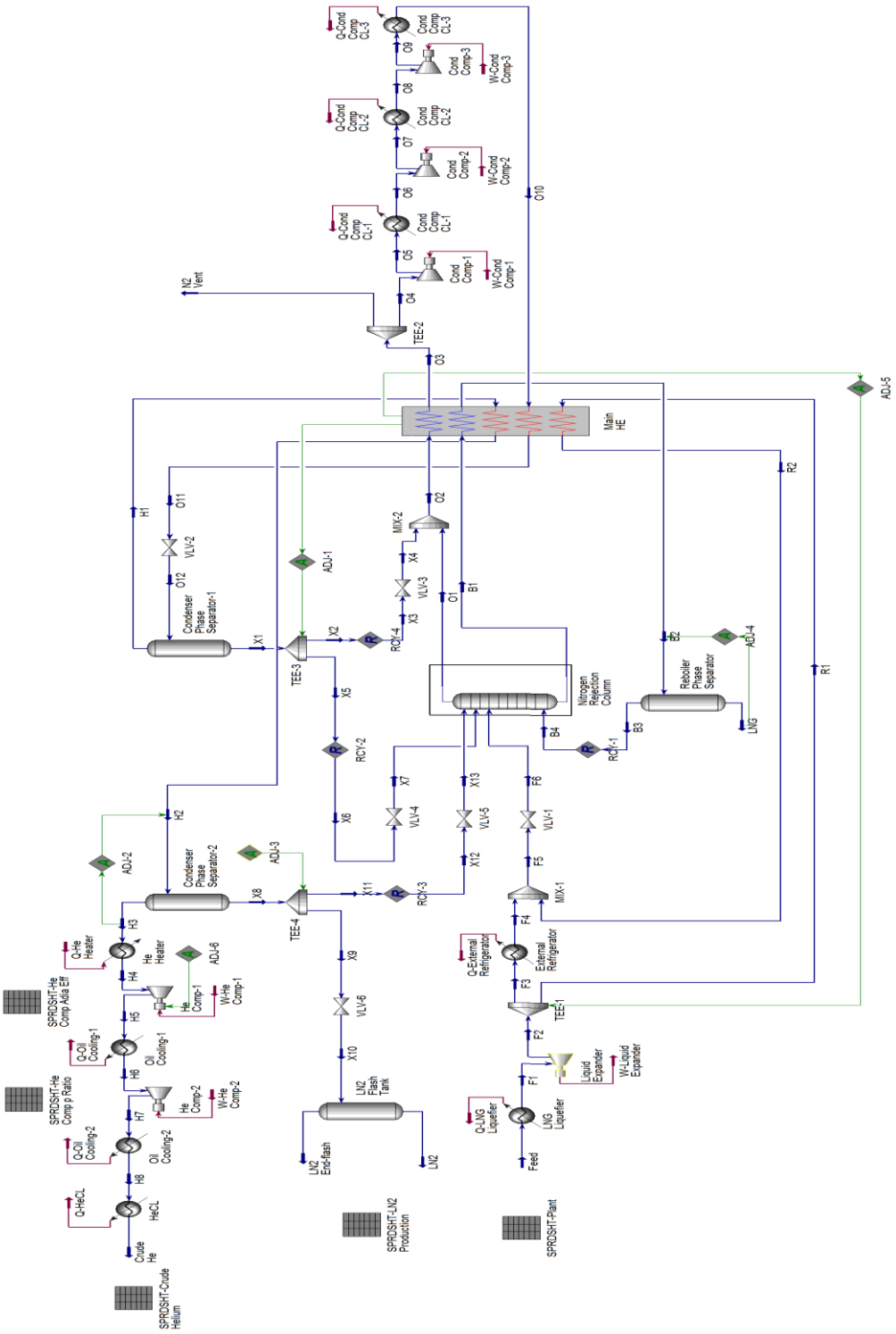


Figure 5.5 The HYSYS Model for the Technip Distillation Process

The overhead stream O1 is utilised as a reflux stream to the N_2 rejection column and a source for producing crude helium. First, stream O1 is mixed with stream X4 coming from the bottom of the first condenser phase separator (Condenser Phase Separator-1) and the admixed stream O2 is delivered to a heat exchanger (Main HE) to take charge of the cold duty of the lowest temperature area in the exchanger. Main HE was originally three individual heat exchangers in the patent application. However, the three exchangers were merged into one exchanger in this HYSYS model for the convergence of the simulation model.

The warmed stream O3 in the exchanger is divided into two streams, N2 Vent and O4, by TEE-2. Stream N2 Vent is purged to the atmosphere as a nitrogen vent stream. Stream O4 is pressurised by a three-stage centrifugal compressor (Cond Comp-1, 2 and 3) with intercoolers (Cond Comp CL-1, 2 and 3) before it is sent back to Main HE. The Cond Comp-3 outlet pressure of 19.7 bara was adopted from the patent application of this process. The pressure ratio of the each stage was obtained by the theoretical compression ratio equation, recorded as 2.5.

The split between stream N2 Vent and O4 is the only one free variable in this process. As stream N2 Vent, which contains a large amount of helium, is wasted to atmosphere, the split ratio affects the helium extraction rate of this system. Moreover, change in the flow rate of stream O3 influences the compression work of Cond Comp-1, 2 and 3. Therefore, a case study was performed to find the optimal split, indicating that the optimal split between stream N2 Vent and O4 is 30.5% and 69.5%, respectively (See Appendix D). This value is within the range that Technip proposes in the patent application. The proposed typical flow ratio of N2 Vent by Technip is ranged from 10% to 50%.

The compressed stream O10 leaving from Cond Comp-3 is cooled down to -170°C via Main HE to be fully condensed before it is used as a reflux stream to the N_2 rejection column. The condensing temperature of -170°C was adopted from the patent application of this process. The fully condensed stream O11 is then depressurised to 5 bara by a J-T valve (VLV-2) to reach -179.5°C, which is the temperature level that Technip suggests. The throttled stream O12 is then conveyed to a phase separator (Condenser Phase Separator-1) to generate a flash gas (H1) and a condensate (X1).

The condensate (X1) coming from the bottom of the phase separator is split to stream X2 and X5. Stream X5 becomes the first reflux stream to the nitrogen rejection column after being throttled to the working pressure level of the column. The throttled stream X7 supplies cold energy to condense hydrocarbons evaporated in the column. Thus, Main HE and Condenser Phase Separator-1 work as a condenser of the column. In contrast, the rest of the condensate (X2) returns to Main HE to supply the cold duty of the heat exchanger. Therefore, the split between X2 and X5 was controlled to achieve 3 K minimum pinch in Main HE. The flow split between stream X2 and X5 was 5.4% and 94.6%, respectively. This is within the range Technip proposes in the patent application, as the suggested flow ratio of X2 is from 0% to 10%.

Finally, the flash gas H1 discharged from Condenser Phase Separator-1 is used to produce crude helium in downstream processes. Stream H1 is partially condensed in the Main HE at -187.9°C and the two-phase mixture H2 is divided into a helium-rich gas (H3) and helium-lean liquid (X8) through a phase separator (Condenser Phase Separator-2). The condensing temperature was manipulated to obtain 50 mole% helium in stream H3. Then, stream H3 is pressurised to 25 bara by a two-stage screw compressor (He Comp-1 and 2). The compressed helium-rich stream H8 is chilled to 35°C via a cooler (HeCL) to generate crude helium. Therefore, the Main HE and the Condenser Phase Separator-2 play the role of a HeXU in this process.

The bottom product of the Condenser Phase Separator-2 (X8) is separated to stream X9 and X11 to generate liquid nitrogen and another reflux stream required in the N_2 rejection column. Stream X9, which becomes LN_2, is throttled to 1.3 bara and divided into an end-flash vapour stream (LN2 End-flash) and a final LN_2 product (LN2). Thus, the split between stream X9 and X11 was controlled to produce 5,500 liter/day of LN_2. The flow ratio of stream X9 was 3% of stream X8. Technip suggests that stream X9 is limited to less than 10% of stream X8. Thus, this simulation model fulfils the restriction of the patent application.

The rest of the stream X8 (X11) is also depressurised to 1.3 bara, which is the working pressure of the nitrogen rejection column, and introduced in the column to provide cold energy. As stream X11 has the lowest temperature between other streams entering the column, it is fed to the top tray of the column.

5.2.3 Re-boiled Distillation Process

As there is no patent application for helium extraction processes utilising a distillation column with a reboiler, a process model was devised by author based on the APCI flashing process (US 2007/0157662 A1) and Figure 3.15. To simplify the simulation work, a HYSYS model for a distillation column with a reboiler was used and the duty of the reboiler was supplied by a cooler model in HYSYS, which extracts heat energy from a sub-cooled LNG stream. This process is named as "Re-boiled Distillation process". The schematic of this process is shown in Figure 5.6.

First, a feed gas ("Feed" in Figure 5.6) is split into stream F1 and R1. Stream F1 is then sent to a cooler (LNG Liquefier) to be liquefied and sub-cooled at -144.0°C. The rest of the feed stream (R1) is also delivered to a heat exchanger (Cold Recovery HE) to be chilled to the same temperature of the sub-cooled stream F1. As this process generates gas products having low temperatures, the products are used to sub-cool a portion of the feed stream in order to reduce the duty of LNG Liquefier. The split between stream F1 and R1 was manipulated to achieve 3 K minimum temperature difference of Cold Recovery HE. The ratio was 95.1% and 4.9% in this simulation work.

The two sub-cooled streams, F2 and R2 are merged and conveyed to a cooler (Reboiler HE) where the heat of admixed stream F3 is exploited and delivered to the reboiler in a distillation column (Nitrogen Removal Column). To balance the two duties of the cooler and the reboiler in the column, the sub-cooling temperature of the feed gas was controlled. In addition, the outlet temperature of Reboiler HE was adjusted to have 3 K minimum pinch between the hot stream passing through Reboiler HE and the cold stream in the reboiler in Nitrogen Removal Column.

The heat exploited stream F4 is expanded to 4 bara before being fed to the column. Thus, 4 bara was the working pressure of Nitrogen Removal Column. Unlike other processes, the expansion is done by a liquid expander (Liquid Exp). It is to reduce the amount of the overhead vapour of Nitrogen Removal Column as an enough amount of the overhead product is already produced by the reboiler. The outlet pressure of the expander was selected after performing a case study. The detailed information is addressed in Appendix E.

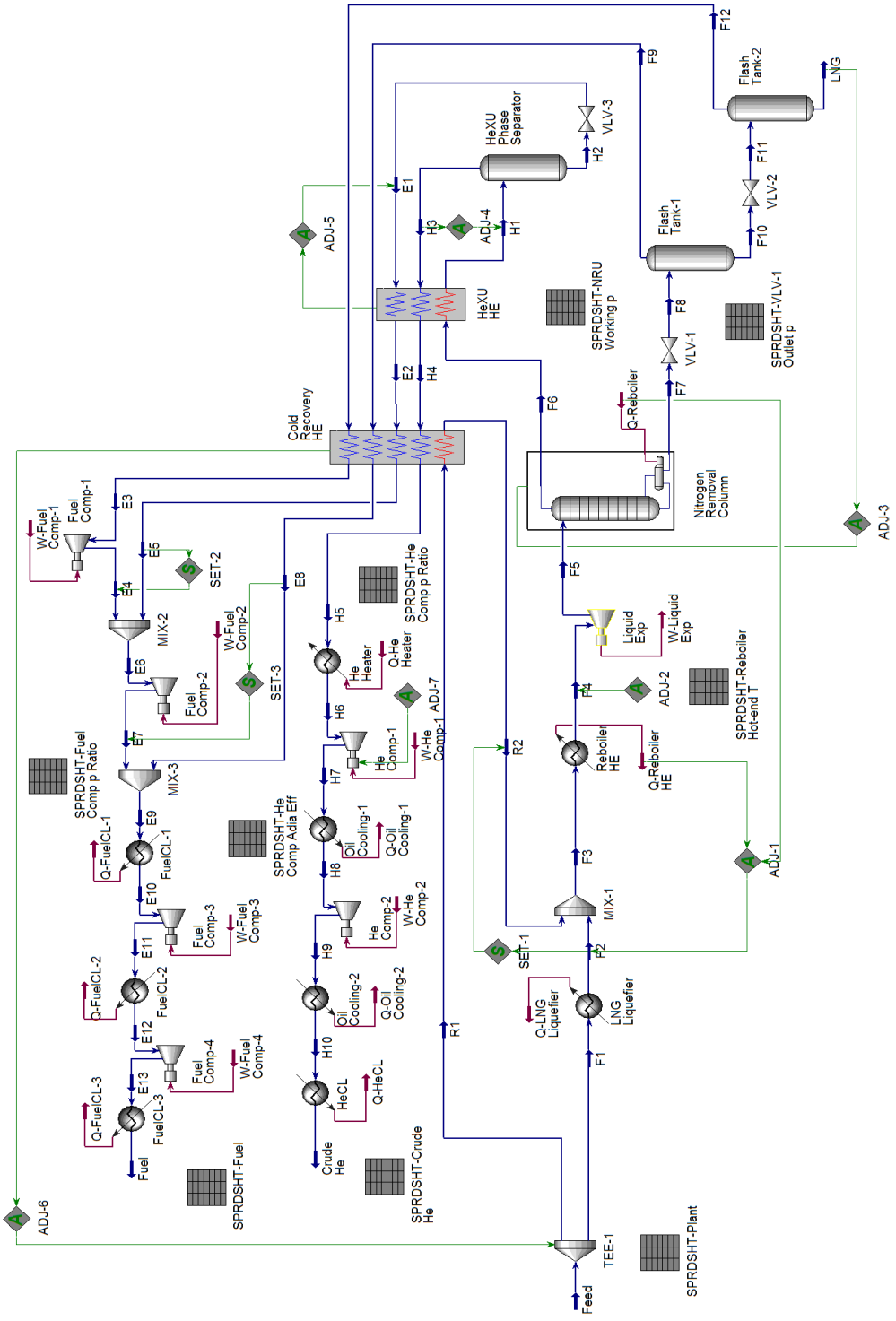


Figure 5.6 The HYSYS Model for the Re-boiled Distillation Process

In the distillation column (Nitrogen Removal Column), the expanded stream (F5) is separated to stream F6 and F7. The only one specification of the column was the nitrogen mole fraction of the bottom product (F7). It was controlled to achieve 1 mole% N_2 in final LNG product (LNG) in this process. The value was 3.5 mole% N_2.

The bottom product (F7) is then throttled to 2.7 bara via VLV-1 and again to 1.3 bara via VLV-2, while producing two flash gases (F9 and F12) and final LNG product (LNG). The outlet pressure of VLV-2 was selected since it is a specification of LNG product. In case of the outlet pressure of VLV-1, the average value between the distillation working pressure and the outlet pressure of VLV-2.

The overhead product (F6) is delivered to HeXU HE as a feed stream for a HeXU and stream F6 is condensed at -184.2°C in the exchanger. The condensed stream H1 is then separated to a helium-rich gas (H3), which becomes crude helium, and a helium-lean liquid (H2) in a separator (HeXU Phase Separator). Thus, the condensing temperature was controlled to obtain 50 mole% He in stream H3 to fulfil the specification of crude helium. The cold duty of HeXU HE is supplied by returning stream H2 and H3. In case of stream H2, it is throttled to 1.5 bara by a J-T valve (VLV-3) to achieve 3K minimum temperature approach in HeXU HE.

As the two warmed streams (E2 and H4) in HeXU HE still have a low temperature level, these streams are conveyed to another heat exchanger (Cold Recovery HE) for cold recovery. The two flash gases are also sent to the heat exchanger for exploiting their cold energy to sub-cool a part of the feed gas (R1).

Next, stream H5 coming from Cold Recovery HE goes through a two-stage oil-flooded screw compressor (He Comp-1 and 2) to be pressurised to 25bara. The pressure ratio between stages was 2.2 and it was obtained by the theoretical pressure ratio equation. The high-pressure helium-rich stream H10 is then finally cooled to 35°C by a cooler (HeCL) to fulfil one of the crude helium specifications.

The rest of the cold streams leaving from Cold Recovery HE are merged and compressed to 20 bara to form fuel gas. First, stream E3, which is warmed the last flash gas (F12), is compressed to the pressure level of stream E5 and mixed with it. The mixed stream E6 is again pressurised to 2.7 bara, which is the pressure level of stream E8. The pressurised stream E7 is then combined

with stream E8 and the mixture is chilled to 35°C via FuelCL-1 before entering a two-stage centrifugal compressor with an intercooler (FuelCL-3) to reach 20 bara. The pressure ratio for each stage was calculated by the theoretical pressure ratio equation to have the smallest compressor work. The ratio was 2.5bara in this simulation work. The high-pressure stream (E13) discharged from the compressor is then cooled to 35°C as it has too high temperature to be provided to gas turbines as fuel.

5.3 Integration of Flashing and Distillation

5.3.1 Linde Integration Process

This Helium extraction process from Linde was modelled by referring to the patent application, US 2009/0013718 A1 and Figure 3.16 (Schmidt, 2009). In this application, Linde does not offer any detailed stream conditions. Therefore, many design factors were reasonably assumed to make its simulation model. The schematic of this HYSYS model is shown in Figure 5.7.

This process starts with flashing of a sub-cooled LNG stream like typical flashing based helium extraction processes. First, a feed gas is split into stream F1 and R1 and stream F1 is liquefied and sub-cooled in LNG Liquefier, reaching at -145.2°C. The sub-cooling temperature was controlled to have less than 1 mole% N_2 in LNG after the sub-cooled stream is throttled three times. In contrast to stream F1, stream R1 is sent to Cold Recovery HE to be cooled to the sub-cooling temperature. The cold duty of this heat exchanger is supplied by gas products generated in this process. The amount of R1 was adjusted to have 3 K minimum pinch in Cold Recovery HE. The split between stream F1 and R1 was 95.2% and 4.8%, respectively.

The two sub-cooled streams are then merged as stream F3 and it is J-T expanded to 5.5 bara via VLV-1, having a cooling effect on the merged stream. Afterwards, the expanded stream is separated into the first end-flash (F5) and helium-lean LNG (F9) through a phase separator (Flash Tank-1). The outlet pressure of VLV-1 was adjusted to have a helium extraction rate of 90%.

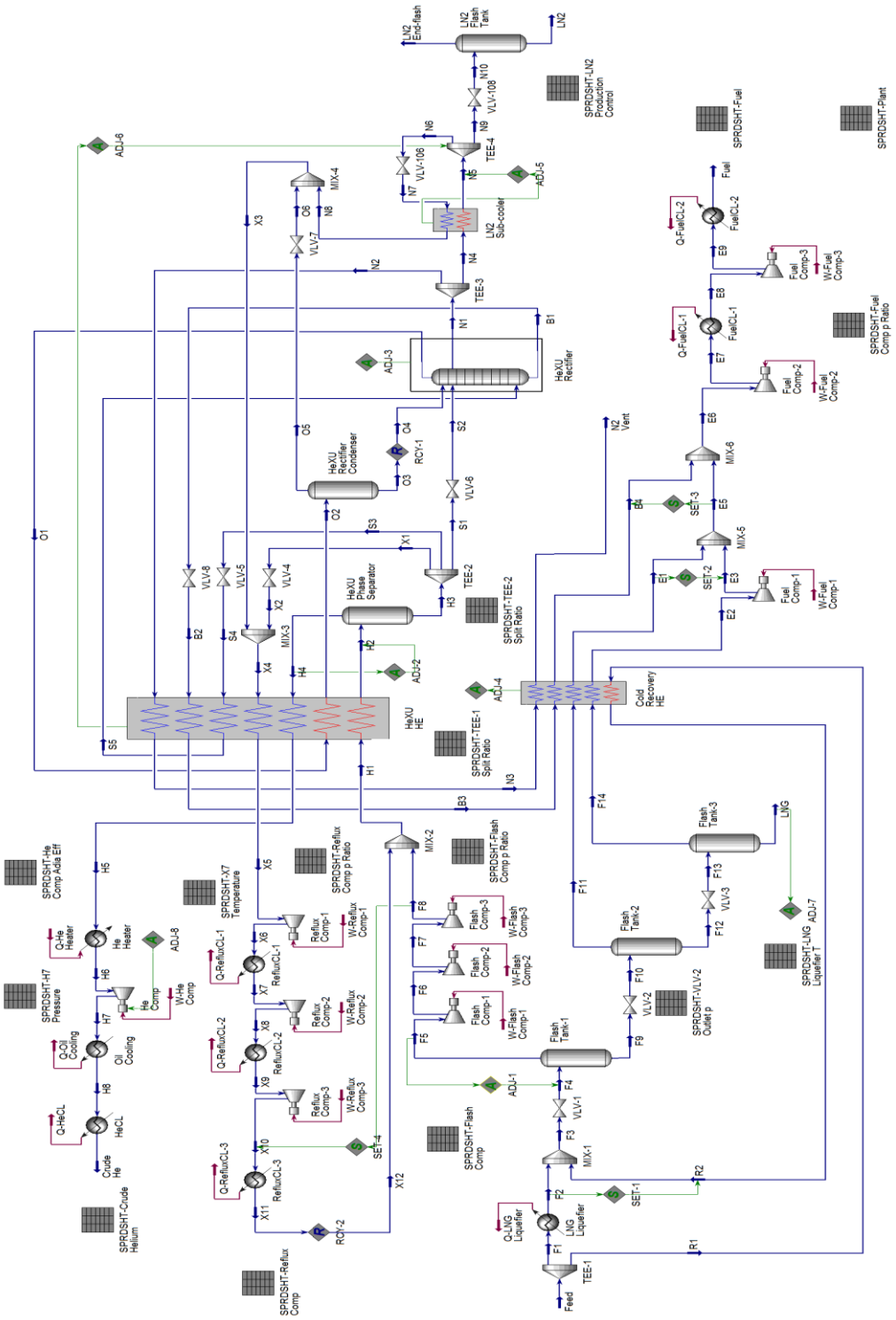


Figure 5.7 The HYSYS Model for the Linde Integration Process

Again, the helium-lean LNG experiences J-T throttling to 3.4 bara by VLV-2, producing the second end-flash (F11) and nitrogen-stripped LNG (F12). Finally, stream F12 is depressurised to 1.3 bara to generate LNG for storage while releasing the last end-flash (F14). The outlet pressure of VLV-2 was derived from the mean value of the VLV-1 outlet pressure and the LNG storage pressure. In case of the two flash gases (F11 and F14), they are piped to Cold Recovery HE to give their cold energy before becoming fuel gas.

Unlike other helium extraction processes, the HeXU feed stream F5 is compressed to 15 bara by a three-stage compressor (Flash Comp-1, 2 and 3). The outlet pressure of the compressor is the value suggested by Linde in the patent application of this process. The pressure ratio of the each stage calculated by the theoretical compression ratio equation was 1.4. The outlet temperature of the each compression stage was low enough due to the low temperature of stream F5 and the low compression ratio. Thus, intercoolers between the stages were not installed in this process. The final discharge temperature of the compressor was -91.8°C at 15 bara.

The pressure level of stream F8 is one of free variables in this process. Thus, a case study with this variable was conducted to find the pressure level, while fixing other free variables at specific values. As indicated in the patent application, the recommended pressure level of F8 is ranged from 15 bara to 30 bara. Therefore, the case study was performed within the range, indicating that 15 bara delivers the least energy consumption for this process. In Appendix F, the result of the case study was explained in detail.

Then, the compressed first end-flash (F8) is combined with a reflux stream (X12) coming from the HeXU and sent to HeXU HE. In the heat exchanger, the admixed stream is partially condensed at -166.6°C. Afterwards, the condensed stream is transported to HeXU Phase Separator to produce helium-rich gas (H4) and helium-lean liquid (H3). The condensing temperature was controlled to possess 50 mole% He in the helium-rich gas before it is sent back to HeXU HE to be exploited its cold energy.

In case of stream H3, it is split into three streams by TEE-2. One of the three steams (X1) are returned and merged with the HeXU feed stream (H1). The others (S1 and S3) are transported to HeXU Rectifier. The split ratio of TEE-2 is also another free variable of this application so a case study was conducted with the variable to optimise this process. Other free variables in this process were fixed at specific values during the case study. As the TEE-2 split

ratio is composed of three parameters, the flow ratio of stream X1, S1 and S3, the number of cases may be too large. Accordingly, it is assumed that the flow ratio of stream S1 and S3 are the same. Therefore, there was only one parameter in this case study, the flow ratio of stream X1 as the flow ratio of stream S1 and S3 are dependant to the ratio. The result of the case study in Appendix F revealed the optimal flow ratio of stream X1, S1 and S3 are 32%, 34% and 34%, respectively.

As mentioned earlier, stream X1 is combined with the compressed first end-flash (F8). Before being merged with stream F8, stream X1 is expanded to 1.3 bara, reaching at -186.5°C. Then, it is combined with another cold stream X3, producing stream X4. Stream X3 is a mixture of stream O6 and N8. Stream O6 is the expanded overhead vapour of HeXU Rectifier Condenser, having 1.3 bara. Stream N8 is the cooling medium of LN2 Sub-cooler, warmed through the cooler. Since stream N8 has the pressure level of 1.3 bara, stream X1 and O6 that are connected to stream N8 had to be depressurised to the pressure level of the cooling medium. Then, the mixed cold stream X4 is sent to HeXU HE to take charge of the duty for the coldest part of the heat exchanger. After supplying cold energy, the warmed stream X5 is discharged from the exchanger at around -50.9°C.

Then, stream X5 is recompressed from 1.3 bara to 15 bara to be mixed with the compressed first end-flash (F8). The compression was performed by a three-stage centrifugal compressor (Reflux Comp-1, 2 and 3) with coolers at the outlet of the each stage (RefluxCL-1, 2 and 3). The pressure ratio of 3.4 for each stage was acquired by the theoretical compression ratio equation. The discharge temperature of the coolers was 35°C as indicated in Chapter 4.1.2. A point to be considered is that the discharge temperature of the first compression stage through Reflux Comp-1 may be lower than the cooler outlet temperature of 35°C depending on the temperature level of the Reflux Comp-1 inlet stream X5. Thus, the HYSYS model was controlled not to use RefluxCL-1 when the outlet temperature of the first compression stage was lower than 35°C.

S1, another stream coming from TEE-2, is throttled to 7 bara before being sent to HeXU Rectifier as a main feed stream for the column. Thus, the working pressure of the rectifier was set to 7 bara. Lower throttling pressure decreases the working temperature of the rectifier as the throttled stream has also lower temperature. In addition, the lower working temperature of the rectifier requires the condenser of this rectifier (HeXU Rectifier Condenser) to

have lower temperature level. This imposes larger cold duty on HeXU HE and reduces its working temperature as the exchanger works as a condenser for the rectifier. Lower than 7 bara causes HeXU HE to have too low working temperature. Therefore, the throttling pressure level of 7 bara was selected.

The remaining stream coming from TEE-2 is stream S3 and this stream works as the re-boiler of HeXU Rectifier. The helium-lean liquid stream (S3) is first expanded to the working pressure of the rectifier to provide the cold duty of HeXU HE. Through the heat exchanger, stream S3 is partially condensed at -156.0°C and sent to the bottom stage of the rectifier. The condensing temperature was selected to delivers enough heat to the column to strip helium and nitrogen in the stream S2 and O4. This stripped helium is the one that are not recovered via HeXU Phase Separator and left in the helium-lean liquid (H3).

The overhead vapour of the rectifier (O1), which contains the stripped helium, is sent to HeXU HE and condensed at -179°C. The heat exchanger plays the role of a condenser for the rectifier. By this condensation, a large amount of N_2 in stream O1 is removed into the condensate (O3), while making the condenser overhead product (O5) richer in He. As the rectifier produce liquid N2 as a side stream drawn, the condensing temperature was set to produce high purity liquid N_2 in the column. Then the helium-rich stream O5 is merged with the feed stream of HeXU (F8). Thus, it allows this system to minimise helium leakage to other products like LN_2, N_2 vent and fuel gas. The helium-lean condenser bottom product is delivered to the rectifier as reflux to condense methane and nitrogen in the vapour stream in the column.

At the bottom of the rectifier, CH_4 and N_2 rich liquid (B1) is extracted from the column and used as fuel after being utilised its cold energy via HeXU HE and Cold Recovery HE. First, the bottom product that has the working pressure of the rectifier is depressurised to the pressure level of the second end-flash as they join in downstream. Then, HeXU HE exploits the cold energy from the throttled stream B2 and discharges the warmed stream B3.

One thing that is special about HeXU Rectifier is that a portion of the liquid on the top tray in the column is extracted to produce liquid nitrogen in this process. The extracted liquid stream (N1) is almost pure nitrogen and its fraction depends on from which tray it is extracted. Higher nitrogen fraction in stream N1 is a favourable condition as one can produce purer LN_2. The LN_2 produced in this process is used for downstream helium purification and lique-

faction process, which have low working temperature. Thus, purer LN_2 holding a smaller amount of impurities like CO2 may have a lower chance to freeze in the processes. For this reason, the nitrogen-rich liquid is extracted from the top tray in the rectifier to acquire the liquid as pure in nitrogen as possible.

Before the extracted liquid is sent to downstream process for producing LN_2, it is divided into two streams, stream N2 and N4. Stream N2 takes the role of venting nitrogen from this process as it is discharged to the atmosphere after it passes through HeXU HE and Cold Recovery HE to provide cold energy. Thus, only stream N4 is utilised to generate LN_2. In this simulation model, most of the nitrogen-rich stream is decided to be used for producing LN_2 rather dumping it to the atmosphere. Therefore, the flow ratio between stream N2 and N4 was set to 10% and 90% of stream N1.

As the N_2-rich liquid (N4) has to be depressurised to near atmospheric pressure for storage, it is sub-cooled to -190.8°C via LN2 Sub-cooler to maximise LN_2 production. The sub-cooling temperature was adjusted to have 3 K temperature approach of the heat exchanger. A part of the sub-cooled stream (N6) is throttled to 1.3 bara and sent back to the exchanger to supply the cold duty of LN2 Sub-cooler. The rest of the sub-cooled stream (N9) is finally depressurised to 1.3 bara, producing LN_2 for storage. The flow ratio between stream N6 and N9 are controlled to achieve 3 K minimum pinch in HeXU HE. The ratio in this HYSYS model was 70.8% and 21.2% of stream N5, respectively.

In this process, fuel gas is formed by mixing three streams; the second end-flash (F11), the last end-flash (F14) and the warmed HeXU rectifier bottom product (B3). They all pass through Cold Recovery HE to sub-cool a portion of the feed stream (R1). The amount of feed sent to the exchanger was manipulated to have 3 K minimum approach in the heat exchanger. After the exchanger, stream F14, which has the lowest pressure level, is compressed by Fuel Comp-1 to the pressure level of the second end-flash. Then the two end-flash streams are mixed with stream B3 as all the three streams have the same pressure level. The admixed stream E6 goes through a two-stage centrifugal compressor (Fuel Comp-2 and 3) to reach the fuel specification pressure, 20 bara. The pressure ratio of 2.4 for the each stage was obtained by the theoretical pressure ratio equation to minimise the compression work. Between the stages, an intercooler (Fuel CL-1) was put to cool the compressed stream (E7) down to 35°C. An additional cooler was also installed at the outlet of the last compression stage to chill the compressed fuel gas to 35°C.

5.3.2 ExxonMobil Integration Process

The HYSYS model for the ExxonMobil Integration process was developed based on the patent application WO 2013/015907 A1 and Figure 3.17 (Oelfke & Victory, 2013). Particularly, this simulation model was built by referring to the fourth drawing in the application. The configuration of this process is illustrated in Figure 5.8.

First, a portion of a feed gas is delivered to a heat exchanger (LNG Liquefier in Figure 5.8) and the rest of the feed gas is sent to another heat exchanger (Cold Recovery HE). The two feed stream are then both liquefied and sub-cooled to -146.1°C in the heat exchangers and merged together in Mix-1. Next, the admixed stream F3 goes through three flashing steps to produce a helium-rich stream (F5), fuel gas streams (F12 and F15) and final LNG product (LNG). The sub-cooling temperature was adjusted to have less than 1 mole% N_2 in the LNG produced in this system.

The pressure level of the each flashing step is 3.9, 2.6 and 1.3 bara and all the flashing processes were conducted by J-T valves. The first flashing pressure level was selected to achieve a helium extraction rate of 90%. The last flashing pressure was set for LNG storage. The second flashing pressure was simply the average value of the first and the last flashing pressure.

The first end-flash (F5) is compressed to 40 bara through a three-stage centrifugal compressor (Flash Comp-1, 2 and 3), while the Linde Integration process pressurises it to 15 bara. The pressure level was the value ExxonMobil suggests in their patent application. Thanks to the low temperature of the first end-flash, intercoolers and a cooler at the outlet of the compressor were not needed. The pressure ratio of the each stage was 2.2 and it was calculated by the theoretical pressure ratio equation.

The compressed first end-flash (F8) is then sent to a HeXU, which is mainly composed of two heat exchangers (HeXU HE-1 and 2), a distillation column with reboiler (HeXU Distillation Column) and a phase separator (HeXU Phase Separator). First, stream F8 is delivered to the first heat exchanger (HeXU HE-1) to be fully condensed at -128°C. The condensing temperature was the value ExxonMobil suggests in the patent application of this process. The condensate (F9) is then delivered to the top tray of the distillation column to strip helium.

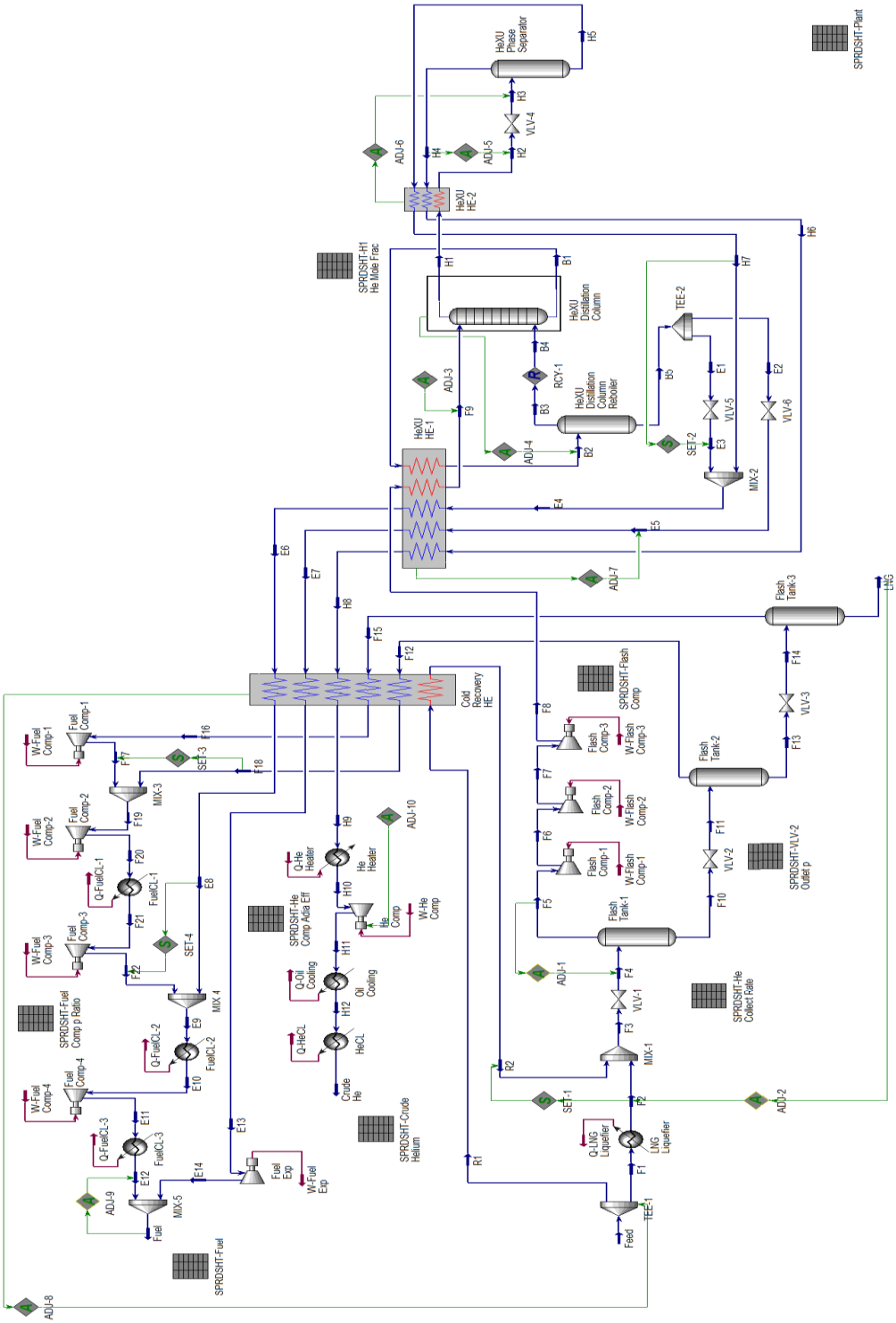


Figure 5.8 The HYSYS Model for the ExxonMobil Integration Process

The heat duty for stripping helium from the condensate (F9) in the distillation column is supplied by a reboiler, which consists of HeXU HE-1 and HeXU Distillation Colum Reboiler in this process. The bottom product (B1) of the column is sent back to HeXU HE-1 and boiled up to -117.3°C, making stream B1 partially evaporated. Then, the two-phase mixture (B2) is split into a boil-up vapour (B3) and a reboiler bottom liquid (B5) via a phase separator. Finally, stream B3 is fed to the bottom tray of the column to deliver heat for evaporating helium in the condensate (F9). Accordingly, the overhead product of this column can hold a high concentration of helium. ExxonMobil indicates in their patent application that the mole fraction of the overhead of the column is ten times higher than the feed stream to the distillation column. Thus, the reboiler temperature was controlled to increase the helium mole fraction in the overhead product (H1) tenfold, compared to the feed stream to the column (F9).

Next, the helium-concentrated vapour (H1) from the distillation column is purified by partial condensing to obtain a gas richer in helium. First, stream H1 is transported to another heat exchanger in the HeXU (HeXU HE-2) to be cooled down to -165.6°C while being partially condensed. The chilled helium-rich stream (H2) through HeXU HE-2 is then depressurised to 15.1 bara by a J-T valve (VLV-4). The throttled stream (H3) is phase-separated in a tank (HeXU Phase Separator), generating a helium-rich gas (H4) and a helium-lean liquid (H5). As stream H4 becomes crude helium, the vapour has to contain 50 mole% He. Thus, the condensing temperature of stream H1 was manipulated to achieve the helium mole fraction of 0.5 in stream H4. The outlet pressure of VLV-4 was also controlled to have 3 K minimum temperature approach in HeXU HE-2. The cold duty of the heat exchanger is supplied by stream H4 and H5.

Together with the bottom product (B5) from the reboiler of HeXU Distillation Column, the warmed two streams (H6 and H7) coming from HeXU HE-2 are used to provide the cold duty of HeXU HE-1. Before being sent to HeXU HE-1, Stream B5 is split into two streams (E1 and E2) via TEE-2. Then, stream E1 is mixed with the warmed helium-lean liquid (H7) after being depressurised to the pressure level of stream H7. The mixture of stream E1 and H7 has a temperature of -137.9°C and it is fed to HeXU HE-1. The rest of the bottom product (E2) is also delivered to the heat exchanger after being throttled to 30.0 bara via VLV-6, reaching to -123.7°C. The outlet pressure level of VLV-6 was adjusted to achieve 3 K of minimum temperature approach in HeXU HE-1. Therefore, there are three distinctive temperature levels of cold

streams entering HeXU HE-1, including the warmed helium-rich vapour (H6), which has the temperature level of -139.9°C.

The split of TEE-2 between E1 and E2 is the only one free variable of this process so a case study was performed to find the optimal value. The split between E1 and E2 affects the performance of HeXU HE-1 as the amount of the second and the third coldest stream entering the heat exchanger (E4 and E5 respectively) are changed. Consequently, the outlet pressure of VLV-6 has to be adjusted to change the temperature level of the third coldest stream (E5) corresponding to the changed flow rate. It is to meet the specification of 3 K minimum pinch in HeXU HE-1. Later, stream E5 is compressed to be a fuel gas. Therefore, change in the split ratio of TEE-2 brings the change in the pressure level of stream E5 and thus compression work of the fuel gas. The optimal split between E1 and E2 was 20% and 80% respectively, based on the result of the case study. This is the value delivering the smallest compression work of the fuel gas. The result of the case study is analysed in Appendix G in detail.

After supplying their cold energy in HeXU HE-1, the three cold streams (H6, E4 and E5) are sent to Cold Recovery HE to liquefy and sub-cool a part of the feed gas (R1). The second and the last end-flash (F12 and F15) are also flowed into this exchanger to take charge of a part of the cold duty needed. The amount of stream R1 was controlled to have 3 K minimum temperature approach in Cold Recovery He. The amount of stream R1 was‘ 3.4% of the feed flow rate.

Finally, the helium-rich vapour discharged from Cold Recovery HE (H9) is compressed to 25 bara by a single-stage compressor and chilled to 35°C through a cooler (HeCL) to be crude helium. As stream H9 has a high-pressure level, multi-stage compression for this stream was not needed to reach 25 bara and the compression ratio of the single-stage compressor was just recorded as 1.7.

Other four cold streams discharged from Cold Recovery HE (F16, F18, E8 and E13) are used to form fuel gas in this process. First, the warmed last end-flash (F16) is pressurised to the pressure level of the second end-flash by Fuel Comp-1 and mixed with the heated second end-flash (F18). Then, the mixture of two flash gases goes through a two-stage centrifugal compressor (Fuel Comp-2 and 3) with an intercooler (FuelCL-1) to be pressurised to the pressure level of stream E8. The pressure ratio between the stages was 2.4 and it

was calculated by the theoretical compression ratio equation. The pressurised flash gas mixture (F22) is merged with stream E8, which has a lower temperature than stream F22. Still the mixture of stream F22 and E8 (E9) has a high temperature. Consequently, stream E9 passes through another intercooler (FuelCL-2) to be chilled to 35°C. Afterwards, the stream leaving from FuelCL-2 (E10) is compressed to 20 bara via Fuel Comp-4 and chilled to 54.2°C in FuelCL-3. The stream discharged from FuelCL-3 (E12) is then combined with stream E14. The mixture of stream E12 and E14 is fuel gas of this process.

As the temperature of stream E14 is lower than 35°C, stream E12 does not need to be cooled to 35°C to produce 35°C fuel gas. Thus, the outlet temperature of FuelCL-3 was adjusted to make the mixture of stream E12 and E14 to be 35°C. Stream E14 is generated by expanding stream E13. As stream E13 has a higher-pressure level than the fuel gas specification, it is depressurised to 20 bara by a gas expander (Fuel Exp), while producing work.

6 Evaluation of Helium Extraction Processes

In this chapter, the simulation result of the integration between helium extraction processes and the APCI DMR process was assessed and compared with the one of the APCI DMR process alone. The APCI DMR process was chosen as a representative of an onshore LNG plant. It is to see how much helium extraction process integrated to a LNG plant will affect the performance of LNG production, compared to a typical LNG plant. Even though the APCI C3MR process is more relevant for a base-load LNG plant, the DMR process was selected since the HYSYS model was available and the thermodynamic efficiency of the APCI DMR and C3MR process are similar. The HYSYS model of the APCI DMR process was built by Maya (Kusmaya, 2013) and modified to add a system for cold recovery of fuel gas and three flashing cycles for nitrogen removal (See Appendix H).

To see the operational flexibility of each process a sensitivity analysis was also conducted. As feed gas conditions are not always stable, a helium extraction process has to have a good performance with off design conditions. Thus, flexibility is deemed to be an important performance parameter when choosing a helium extraction process. In this chapter, feed gas composition was varied and corresponding performance was evaluated. Other design basis was kept to be identical to the base cases of the processes evaluated in this project. Feed gas could be leaner or richer in helium, depending on the conditions of well streams.

Thus extra simulation work was conducted with He-lean feed gas case (0.02% He) and He-rich feed gas case (0.1% He) to compare with the result from the base case (0.05% He). The composition of each gas is tabulated in Table 4.2 in Chapter 4.1.1. One notable thing is that the Technip Distillation process was not fit for the sensitivity analysis with the changed feed gas composition due to the complexity of the simulation, especially the distillation part.

6.1 Production of Crude Helium

Main performance index for production of crude helium is helium extraction rate. It is to see whether a helium extraction process could extract helium from a LNG stream having a low helium fraction and could reach helium ex-

traction rate of 90% as most of the He extraction processes studied here are claimed to have a very high helium extraction rate. However, there were some processes, which could not reach helium extraction rate of 90%. In this case, the processes were pushed to come close to the value. The mole fraction of the produced crude helium was 50% for all the helium extraction processes as it was one of the specifications for crude helium.

Process performance is also measured by specific power. In crude helium production systems, specific power can be described as the power needed per unit volume of crude helium produced as crude helium is sold by volume base. With plant availability of 330 days per year, specific power for each process was calculated. The result is illustrated in Figure 6.1.

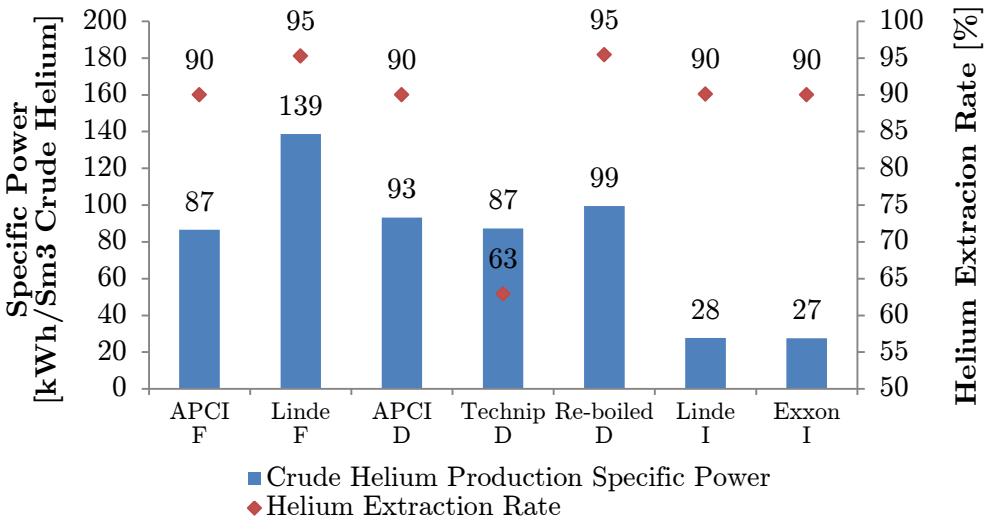


Figure 6.1 Crude Helium Production Specific Power with Helium Extraction Rate (F: Flashing, D: Distillation, I: Integration)

Figure 6.1 shows that all the helium extraction processes are able to extract helium from a LNG stream as they all have helium extraction rate higher than 0%. Especially, the Linde Flashing process and the Re-boiled Distillation process show excellent performance of extracting helium. They recorded helium extraction rate of 95%, which is the lowest possible value for the two processes. The Linde Flashing process could achieve such high helium extraction rate as it uses two end-flash streams as feed for the HeXU in the process. The two end-flash streams always have more than 99.9% of helium in the feed gas, while meeting other design specifications, i.e. 3K minimum temperature approach in the HeXU heat exchanger. In contrast, the APCI Flashing process

only uses the first flash gas stream as feed for the HeXU and the stream has only about 93% of helium in the feed gas. Thus, if the APCI Flashing process wants to extract more helium, this process needs to throttle a sub-cooled LNG more to separate helium from the LNG stream, while losing more hydrocarbons through the first flash gas. The Re-boiled Distillation process also produces an overhead stream from the distillation column, containing 100% of helium in the feed. It is thanks to the reboiler, evaporating helium left in a LNG stream in the column. As the overhead stream is used as feed for the HeXU, this process could have a higher helium extraction rate than others.

In case of the Technip Distillation process, it could not achieve a helium extraction rate of 90%. The maximum achievable helium extraction rate was around 63%. The main reason for such low helium extraction rate of this process is the nitrogen vent stream, which contains a large quantity of helium. In this project, ca. 34% of helium molar flow in the feed stream is wasted through the vent stream. In addition, a portion of the feed stream for the HeXU is also re-introduced to the nitrogen rejection column, reducing the amount of helium entering the HeXU in this process. However, it is expected value as Technip claims around helium extraction rate of 60% in their patent application. Except these three processes, others could be manipulated to obtain a helium extraction rate of 90%, which means more operational flexibility of the process when it comes to extracting helium.

Regarding specific power, Figure 6.1 indicates that the Linde Integration process and the ExxonMobil Integration process are the most efficient processes. Unlike other processes, the feed stream for the HeXU in these integration-based processes is pressurised to 15 bara and 30 bara, respectively. Thus, the high pressure feed for the HeXU increases the pressure level of crude helium generated in the HeXU and it reduces the amount of energy used for compressing crude helium to 20 bara, which is one of the specification of crude helium in this project. Even though the Linde Flashing process and the Re-boiled process have a higher helium extraction rate, they tend to have higher specific power than other processes. It is because that the pressure level of crude helium leaving from the HeXU is lower than other processes, requiring more compression energy in downstream. Particularly, the Linde Flashing process has the highest specific power as the crude helium stream is produced at 1.3 bara in the HeXU, which is near the atmospheric pressure. The rest of the processes showed similar specific power of around 90 kWh/Sm^3. It is because that their crude helium produced in the HeXU has similar pressure level of near 5 bara.

When it comes to feed gas composition change, all the process could meet a helium extraction rate of 90% for the three feed gas cases except the Linde Flashing and the Re-boiled Distillation process as seen in Figure 6.2. For the Linde Flashing and the Re-boiled Distillation process, the helium extraction rate tends to decrease with increasing helium fraction in feed gas. It is mainly because that their HeXU feed gas contains more nitrogen than the base case. This reduces the relative volatility between helium and nitrogen, resulting in lower separation efficiency between them in the HeXU. As mentioned earlier, the two processes are less flexible than other helium extraction processes. Thus, it is hard to change the composition of the HeXU feed gas to increase helium extraction rate.

However, the Linde Flashing process still shows an outstanding performance of extracting helium with a helium-rich feed gas. This process recorded around helium extraction rate of 91%, which is the lowest possible value for the process. On the contrary, the Re-boiled Distillation process had even lower than helium extraction rate of 90% with a helium-rich feed gas, recording ca. 87%. Therefore, the Linde Flashing process shows the best performance with a change in helium fraction in feed gas, while the Re-boiled Distillation process has the poorest performance.

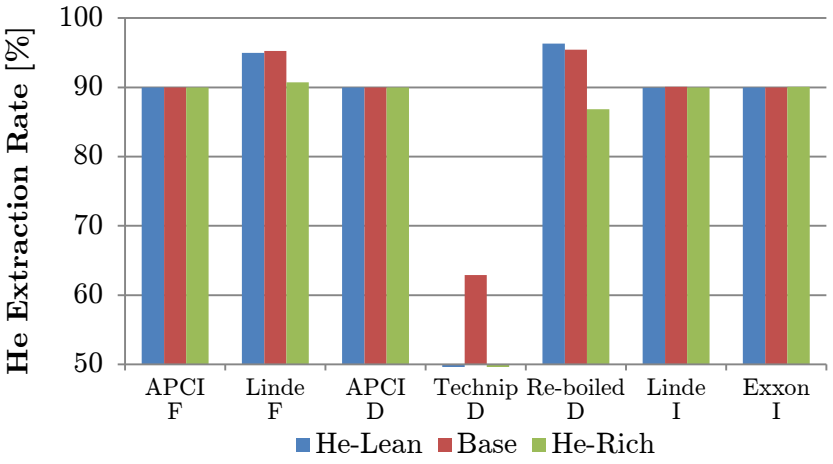


Figure 6.2 The Effect of Feed Gas Composition on Helium Extraction Rate (F: Flashing, D: Distillation, I: Integration)

Regarding helium production specific power, Figure 6.3 indicates that still the two integration-based processes display the lowest specific power thanks to the compressed HeXU feed gas. However, the trend of specific power with a

change in helium concentration in a feed gas was irregular for each process. As helium production specific power is sensitive to the operating conditions of the HeXU, the variation of the specific power depends on the process characteristic. Nevertheless, one can conclude that the Re-boiled Distillation and the Linde Integration process have the best operational flexibility because their specific power is almost stable with a change in helium fraction in a feed gas.

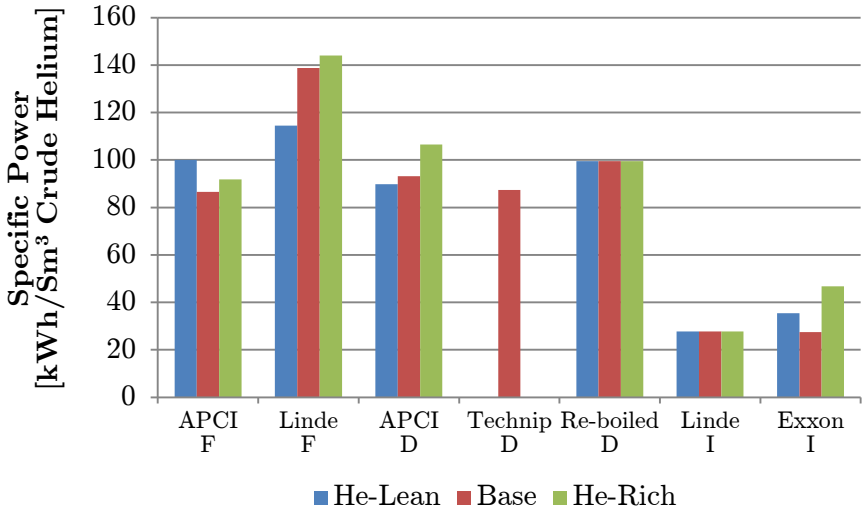


Figure 6.3 The Effect of Feed Gas Composition on Crude Helium Production Specific Power
(F: Flashing, D: Distillation, I: Integration)

6.2 Production of LNG

Crude helium is obtained as a by-product when integrating helium extraction process into a LNG plant. Therefore, it is important to know how the helium extraction process added to a LNG plant will affect the performance of LNG production since LNG is still the main product bringing the most economical value to the plant. This was assessed by two factors, specific power and production rate of LNG. The specific power was defined as the power needed per unit mass of LNG produced and the production rate was expressed as million tonne of LNG produced per year. Based on the plant availability of 330 days per year, the two values were calculated. However, the HYSYS models for the helium extraction processes do not give the power needed for their LNG liquefiers as they were simply modelled as cooler in HYSYS. Thus, the HYSYS model for the APCI DMR process was connected to the coolers to calculate how much compression energy is required to fulfil the cold duty of the coolers.

The performance of the helium extraction processes integrated with the APCI DMR process was compared with the one of the APCI DMR process alone as baseline process for LNG production. All the processes were also controlled to have 1 mole% N_2 in LNG product to meet the specification of LNG introduced in Chapter 4.2.2. The result is depicted in Figure 6.4.

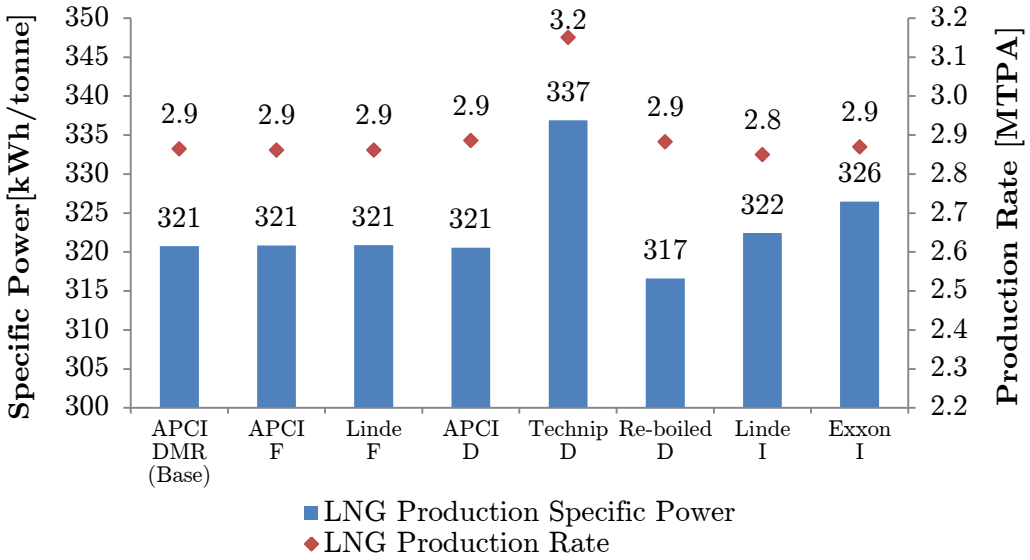


Figure 6.4 LNG Production Specific Power and LNG Production Rate (F: Flashing, D: Distillation, I: Integration)

Figure 6.4 clearly suggests that the Re-boiler distillation process is the most efficient process to be added to a LNG plant when it comes to LNG production. The specific power of the integration of the Re-boiled Distillation and APCI MDR process is even lower than the baseline process, the APCI DMR process. It is mainly because that a part of LNG sub-cooling duty is supplied by the cold bottom stream of the distillation column via the reboiler, decreasing the LNG liquefier cold duty.

The Technip Distillation process shows the poorest efficiency when this is added to the APCI DMR process and the specific power was 5% higher than the baseline process. To run the distillation column in this process, it requires compression of the overhead stream of the column, consuming a large amount of energy and thus increasing the specific power. The combination of the ExxonMobil Integration and the APCI DMR process gives the second largest specific power. All the processes simulated in this project utilise the cold energy of gas products generated from their system in order to sub-cool a portion of

the feed gas, lowering the cold duty of the LNG liquefier. However, the ExxonMobil Integration process has the smallest cold energy of gas products compared to other processes. It resulted in giving more load to the LNG liquefier, increasing the specific power of LNG production. Nevertheless, the difference to the specific power of the baseline process was just 1.5%.

Other integrations of helium extraction processes and the APCI DMR process tend to have similar specific power to the baseline process at around 321 kWh/Sm^3. It means there is no penalty of extracting helium on producing LNG when integrating the helium extraction process and a LNG plant.

When it comes to LNG production rate, the integration of the Technip Distillation process shows the largest production capacity, reaching 3.2 MTPA. It was achieved by condensing a large amount of hydrocarbon vapour existing in the distillation column. The condensing duty is provided by two reflux streams to the column from two condensers in the process. In addition, the refluxing streams to the column also contain a small amount of methane, thus helping maximising LNG production. Due to this feature, this process does not produce fuel gas, which means a part of the feed gas has to be used as fuel for this system. Except the Technip Distillation process, other helium extraction processes simulated gives almost the same LNG production rate as the baseline process when being added to the APCI DMR process. Therefore, the Re-boiled Distillation process may be the optimal process to be integrated to a LNG plant since it produces the same amount of LNG per year as the baseline process, while consuming less energy than others do.

The performance of LNG production for each helium extraction process was also measured for a helium-lean feed gas case and a helium-rich feed gas case and compared with the base case. The result is illustrated in Figure 6.5, Figure 6.6 and Figure 6.7. Figure 6.5 indicates the trend that feed gas richer in helium causes an increase in specific power and vice versa regardless of the type of the helium extraction processes studied. In case of helium-rich feed gas, the boiling point is reduced since a portion of hydrocarbons in the gas was replaced with nitrogen and helium, which have a lower boiling point than the hydrocarbons. Thus, it requires larger cold duty for sub-cooling the feed gas in a LNG liquefier, resulting in increasing the LNG production specific power.

Especially, the specific power of the combination of the ExxonMobil Integration process and the APCI DMR process soars up when the helium fraction in the feed gas rises. In this process, the cold energy acquired from the gas products for sub-cooling a part of the feed gas is around a half, compared to integrations of other helium extraction processes and the APCI DMR process. Thus, it imposes more cold duty on the LNG liquefier, raising the LNG production specific power. Unlike others, the integration of the APCI Distillation and the APCI DMR process shows only a small fluctuation in the specific power, indicating that the APCI Distillation process has the best operational flexibility when being added to a LNG plant.

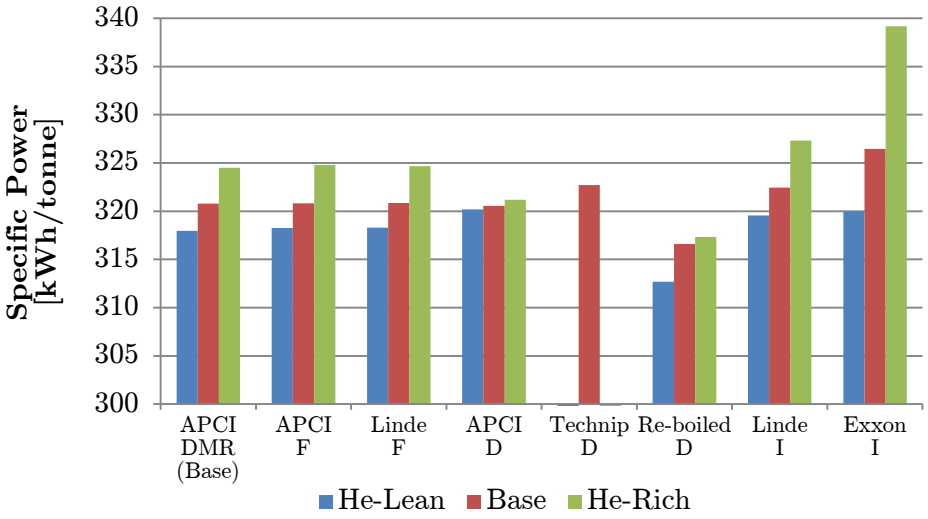


Figure 6.5 The Effect of Feed Gas Composition on LNG Production Specific Power
(F: Flashing, D: Distillation, I: Integration)

Regarding LNG production rate, feed gas having smaller helium fraction allows all the processes simulated to produce more LNG as seen in Figure 6.6. In a LNG plant, high pressure feed gas is sub-cooled via a LNG liquefier and throttled to around 1.3 bara to produce final LNG product. In this procedure, helium-rich feed gas will produce more end-flash and thus less LNG, compared to helium-lean feed gas if the sub-cooling temperature is the same for both feed gas conditions. This is due to the lower boiling point of the helium-rich feed gas than the helium-lean feed gas. The sub-cooling temperature for each helium extraction process was almost the same for the three feed gas cases in each the helium extraction process.

Especially, the APCI Distillation process shows a significant increase in LNG production rate with helium-lean feed gas. As the APCI Distillation process has a condenser integrated into the distillation column, this unit condenses a large amount of the evaporated methane in the column, raising LNG production. On the contrary, the Re-boiled process recorded a small increase in LNG production with helium-lean feed gas. It is because that the evaporated hydrocarbons via the reboiler in the distillation column in this system directly escape through the overhead stream.

In addition, these two distillation-based processes had a poor LNG production capacity with helium-rich gas. To reduce N_2 content in LNG down to 1 mole%, these two system had to flash a LNG stream relatively more than other helium extraction processes, leading to smaller LNG production rate.

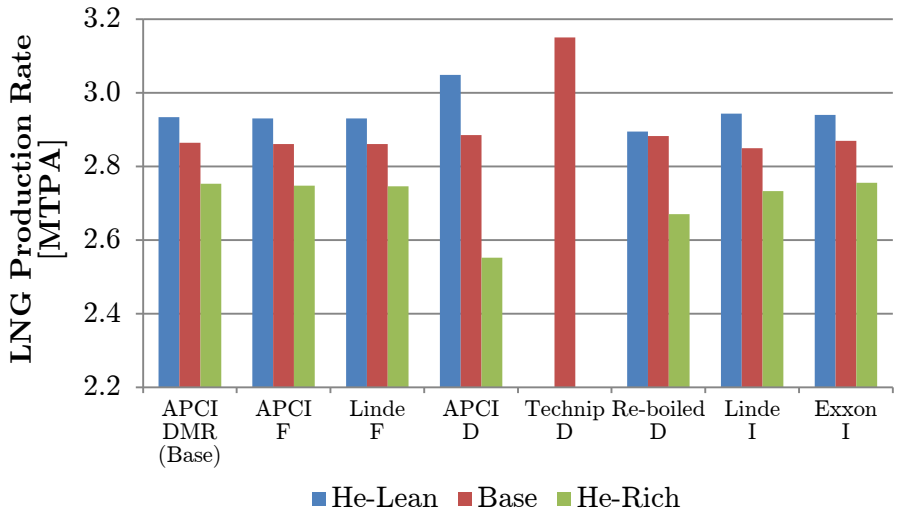


Figure 6.6 The Effect of Feed Gas Composition on LNG Production Rate (F: Flashing, D: Distillation, I: Integration)

Figure 6.7 also shows the effect of change in feed gas composition on nitrogen content in LNG product. As one of LNG specifications, N_2 content in LNG has to be always lower than 1 mole% regardless of change in feed gas composition. When feed gas is rich in helium and thus nitrogen, only the distillation-based helium extraction processes could succeed in reducing nitrogen content in LNG to 1 mole%, while achieving the crude helium specification of 50 mole% He and a helium extraction rate of 90%. A distillation column with reboiler / condenser or both enables the distillation-based processes to control nitrogen and helium content in LNG, resulting in lower nitrogen content in final LNG product.

If the two flashing-based helium extraction processes desire to have such low nitrogen fraction in LNG product, they may have to produce more end-flash to flash off nitrogen content in LNG. However, it will vastly reduce the production rate of LNG as the produced end-flash may contains a large amount of methane. Therefore, the two flashing-based processes would eventually require a use of NRU.

The two integration-based processes also shows similar trend as the flashing-based processes since this process also consists of three flashing steps for reducing nitrogen content in LNG. Therefore, the flashing-based and the integration-based processes may require extra nitrogen removal unit at the end of flashing steps to reduce N_2 content in final LNG product when feed gas is rich in helium and nitrogen. This will make these processes more complex and expensive. In case of helium-lean feed gas, all the processes simulated generate final LNG product having lower than 1 mole% N_2.

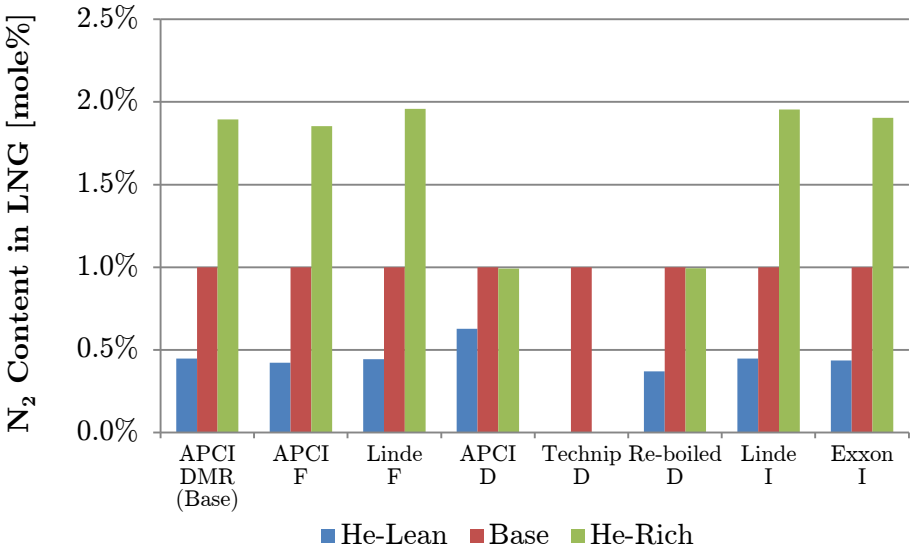


Figure 6.7 The Effect of Feed Gas Composition on N_2 Content in LNG (F: Flashing, D: Distillation, I: Integration)

6.3 Quality of Fuel Gas, LN_2 and N_2 Vent

The nitrogen content in fuel gas is also an important criterion when evaluating helium extraction processes as it has to be lower than 40 mole% to be used for industrial gas turbines. The result is shown in Figure 6.8. As the Technip Distillation process does not produce fuel gas, this process was not included in the result.

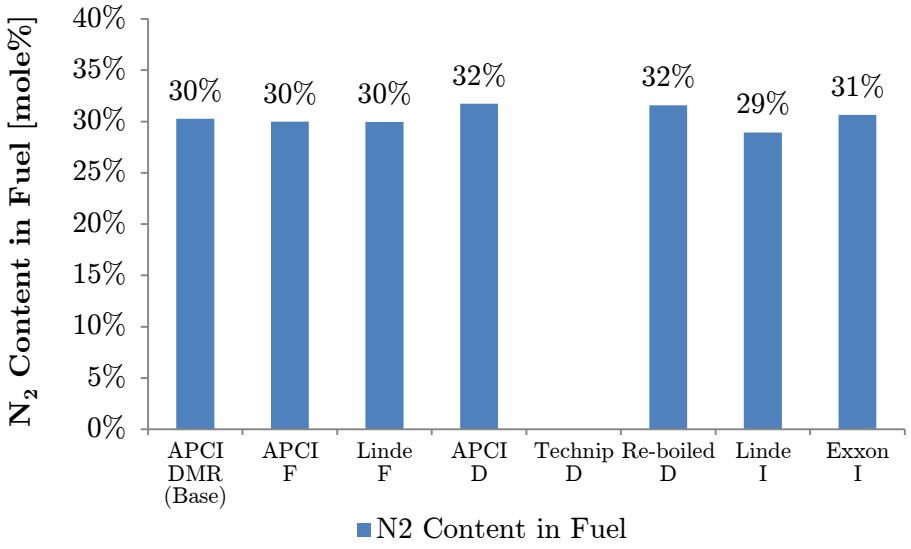


Figure 6.8 Nitrogen Content in Fuel Gas
(F: Flashing, D: Distillation, I: Integration)

Figure 6.8 indicates that all the helium extraction processes produce fuel gas having lower than 40 mole% N_2. Including the baseline process, all the helium extraction processes have around 30 mole% N_2 in fuel gas. Especially, the Linde Integration process achieved the lowest nitrogen content in fuel gas compared to others. The fact that this process vents pure nitrogen from the system to the atmosphere enables to have such low nitrogen content in fuel gas. However, the amount of nitrogen vent stream is small so the effect of the venting was minor, showing that only 1 or 2 mole% difference to other processes.

When the composition of feed gas changes, the nitrogen fraction in fuel gas tends to rise with the increasing helium and nitrogen fraction in feed gas and vice versa as illustrated in Figure 6.9. Especially in case of helium-rich feed gas, all the helium extraction processes except the APCI Distillation process generate fuel gas having more than 40 mole% N_2, which is the upper limit of nitrogen concentration for industrial gas turbines. The APCI Distillation process has the HeXU feed gas containing lower nitrogen content than any other processes. As a large portion of the HeXU feed gas becomes fuel gas, it allows this process to have lower N_2 content in fuel gas, compared to others.

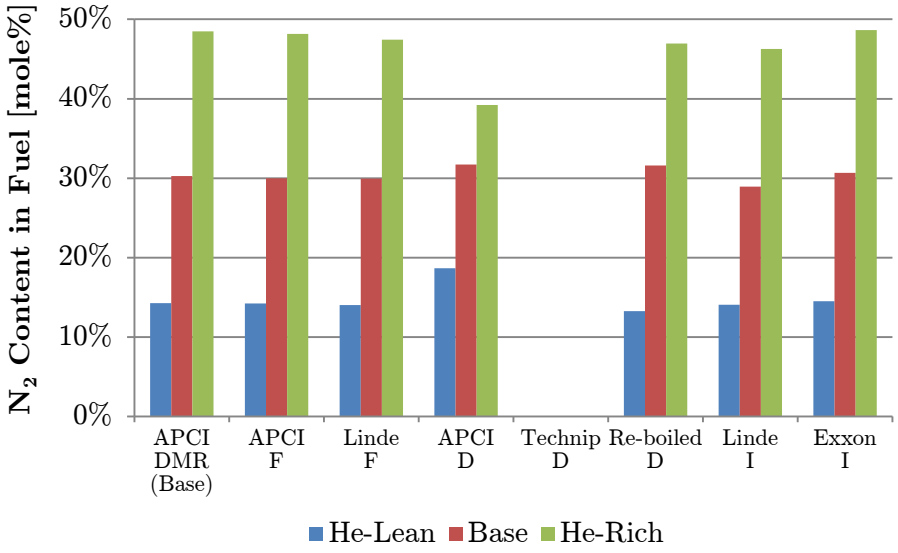


Figure 6.9 The Effect of Feed Gas Composition on N_2 Content in Fuel (F: Flashing, D: Distillation, I: Integration)

Another by-product of helium extraction process is liquid nitrogen and there are only two processes producing LN_2, the Technip Distillation process and the Linde Integration process. Thus, the production rate and purity of LN_2 were measured to evaluate the performance of LN_2 production between the two processes. In addition, a sensitivity analysis with different feed gas composition was conducted for the Linde Integration process to see how the variation of feed gas composition affects the performance of LN_2 production. As mentioned at the beginning of this chapter, the sensitivity analysis was not able to conduct for the Technip Distillation process due to the process complexity. The result is summarised in Table 6.1.

Table 6.1 LN_2 Production Rate and Composition

LN_2	Unit	Technip Distillation	Linde Integration		
		Base	He-lean	Base	He-rich
Volume Flow	liter/day	5,500	4,087	5,500	5,500
N_2 Content	mole%	99.92%	99.76%	99.94%	99.97%
He Content	mole%	0.02%	0.00%	0.00%	0.00%
CH_4 Content	mole%	0.06%	0.23%	0.06%	0.03%

Table 6.1 shows that both the two helium extraction processes with base case feed composition could produce 5,500 liter/day LN_2, which is the assumed minimum required quantity for downstream process. Consequently, the two processes may not need extra ASU to produce liquid nitrogen, which are used as a shielding material of helium containers for transport and a refrigerant for the helium treating process. In addition, they generate high purity LN_2 having higher than 99.9 mole% N_2. Consequently, LN_2 from both the processes contains extremely small amount of helium and methane, which will help reducing hydrocarbon loss from the system and maximising helium extraction.

However, the LN_2 production rate tends to fall with decreasing helium and nitrogen content in feed gas. In case of the Linde Integration process, the maximum possible LN_2 production rate with helium-rich feed gas was 4,087 liter/day, while this process could be controlled to produce 5,500 liter/day LN_2 with helium-rich feed gas. As helium-lean feed gas contains a small amount of nitrogen, it is inevitable to have reduction of LN_2 production rate. Additionally, Table 6.1 indicates that more helium and thus nitrogen content in feed gas enable to produce purer LN_2, which is a favourable condition for the performance of this process.

The two processes generating LN_2 are also the only ones venting pure nitrogen gas to the atmosphere. Thus, it is important to know whether the nitrogen vent stream has lower than 1 mole% CH_4, which is a typical environmental requirement for vent gases. Table 6.2 compares the composition of N_2 vent stream from the each process and indicates none of the both processes emits nitrogen gas containing higher than 1 mole% CH_4. However, nitrogen vent stream from the Technip Distillation process tend to have significantly more methane content and molar flow, compared to the Linde Integration process. Therefore, care has to be taken for the Technip process since the CH_4 content may exceed 1 mole% when handling a feed gas, which is richer in nitrogen than the base feed conditions used for this project. This will cause a considerable amount of hydrocarbon leak to the atmosphere.

One noticeable thing is that the methane content in the N_2 vent stream emitted from the Linde Integration process has a tendency to increase with decreasing the fraction of helium and nitrogen in feed gas. Thus, one has to pay more attention to a N_2 vent stream when having feed gas leaner in helium and nitrogen to meet the environmental regulation.

Table 6.2 N_2 Vent Production Rate and Composition

N_2 Vent	Unit	Technip Distillation	Linde Integration		
		Base	He-lean	Base	He-rich
Molar Flow	kgmole/hr	937.7	9.0	2.8	4.7
N_2 Content	mole%	99.08%	99.74%	99.88%	99.92%
He Content	mole%	0.46%	0.04%	0.06%	0.05%
CH_4 Content	mole%	0.46%	0.23%	0.06%	0.05%

6.4 HeXU Working Conditions

Helium extraction process works at a very low temperature area, typically around -188°C. Accordingly, impurities entering this system will easily freeze especially in a HeXU. Thus, increasing the working temperature level is an important target. Otherwise, one has to put more attention to reduce the amount of impurities flowing into helium extraction system, which requires additional cost and energy. Thus, HeXU working conditions of each helium extraction process were measured to see which process brings the most favourable condition when it comes to the freezing of impurities. HeXU working temperature and pressure were defined as the lowest temperature and pressure level found in the HeXU in this project. The result is illustrated in Figure 6.10.

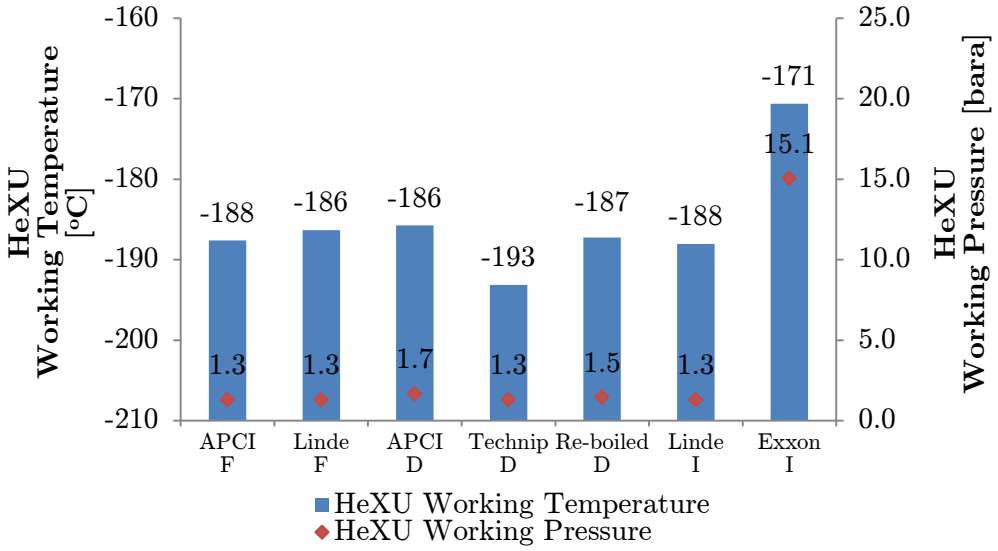


Figure 6.10 HeXU Working Temperature and Pressure
(F: Flashing, D: Distillation, I: Integration)

Figure 6.10 shows that the ExxonMobil Integration process is the least sensitive process to the solidification of impurities as this system has the warmest HeXU working temperature at -171°C. It is thanks to the compressed HeXU feed gas. HeXUs are auto-refrigerated and the HeXU feed gas is let-down in pressure typically to around the atmospheric pressure in order to supply the cold duty of the HeXU. Due to the depressurising, HeXUs normally have low working temperature. However, the high pressure HeXU feed gas from the ExxonMobil process does not need to be depressurised to a low-pressure level to supply the cold duty of the HeXU, increasing the HeXU working temperature.

However, another integration-based process, the Linde Integration process, recorded the second lowest working temperature even though this technology compresses the HeXU feed gas. It is because that the HeXU feed gas has to be depressurised to a very low level (1.3bara in this project) to sub-cools liquid nitrogen produced in the HeXU, resulting in a low HeXU working temperature at -188°C. The Technip Distillation process has the lowest HeXU working temperature, compared to others. Similar to the Linde Integration process, this process also produces liquid nitrogen. Therefore, the temperature level of the distillation column has to be decreased significantly to generate almost pure nitrogen overhead stream as it is used for producing LN_2. Except these three processes, other helium extraction processes showed similar HeXU working temperature at around -187°C.

When the composition of feed gas changes, the HeXU Working temperature tends to rise with the increasing helium and nitrogen fraction in feed gas and vice versa as illustrated in Figure 6.11. The HeXU feed gas extracted from helium-rich feed gas has to be condensed at a lower temperature and phase separated in a HeXU to produce crude helium gas containing 50 mole% He, compared to the case of helium-lean feed gas. The helium and nitrogen content in the HeXU feed gas rise with increasing helium and thus nitrogen fraction in feed gas and the increased amount of He and N2 in the HeXU feed gas reduces the dew point. Therefore, this requires a lower HeXU working temperature to condense the HeXU feed gas to generate crude helium having 50 mole% He.

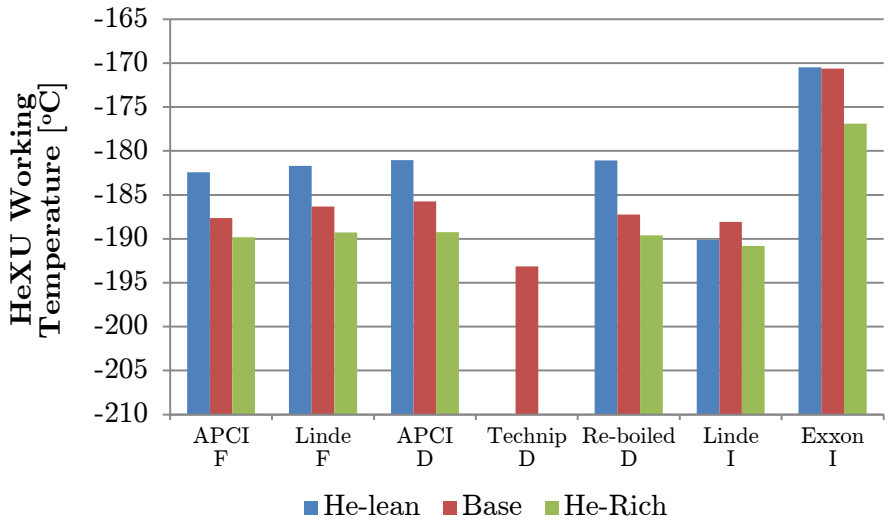


Figure 6.11 The Effect of Feed Gas Composition on HeXU Working Temperature (F: Flashing, D: Distillation, I: Integration)

6.5 Economic Evaluation

As crude helium and LNG production vary depending on the type of helium extraction process, a simplified economic evaluation was performed to see which process creates the most profit. This was measured by summing the value of LNG and crude helium produced per year in each process. The unit price of LNG was set to \$9.46 per million British thermal units (MBTU), which was the annual price of natural gas in 2012 from UK Heren NBP Index (BP, 2014). As LNG price is typically similar to natural gas price, it may be a reasonable approximation. The unit price of crude helium was set to \$3.5 per standard cubic meter. This unit price was derived from the price of Grade-A gaseous helium, which is \$7.21 per standard cubic meter. This was estimated price by U.S. Geological Survey for private industry's Grade-A gaseous helium (Hamak, 2014). Grade-A gaseous helium is defined as high purity helium (99.95 %). Since the crude helium produced in this project has the purity of 50%, the unit price of the crude helium was roughly a half of the Grade-A helium price based on the purity. The price for purifying helium from 50% to 99.95% was disregard when calculating the unit price of crude helium.

Fuel gas generated in the helium extraction processes was assumed to be used for industrial gas turbines to generate electricity for all the rotating machines in a process. The efficiency of industrial gas turbines was reasonably set to 30%. If the electricity converted from fuel gas via industrial gas turbines ex-

ceeds the amount of energy required for a process, the surplus of fuel gas was assumed to be sold at the same unit price of natural gas. Since the unit price of natural gas is the same as LNG in this project, the surplus of the fuel gas was simply added to the LNG chemical energy value. Thus, the economic value for LNG includes the fuel gas balance. If the process demands more power than what is produced from the gas turbines with the fuel gas produced in each process, the shortage was assumed to be covered by buying natural gas and supplying it to the gas turbines with the fuel gas. As the price of natural gas and LNG were the same in this project, it will eventually reduce LNG economic value by the amount of natural gas bought. Another assumption is that the economic value of produced LN_2 in the Technip Distillation process and the Linde Integration process was not included in this evaluation as the price of LN_2 is regarded as relatively cheaper than LNG and crude helium. Detailed procedures are introduced in Appendix I.

The result of the economic evaluation of helium extraction processes were compared with the baseline LNG process, the APCI DMR process. It is to know whether a helium extraction process added to a LNG plant creates more profits, compared to a typical LNG plant, the APCI DMR process in this project. The result is shown in Figure 6.12.

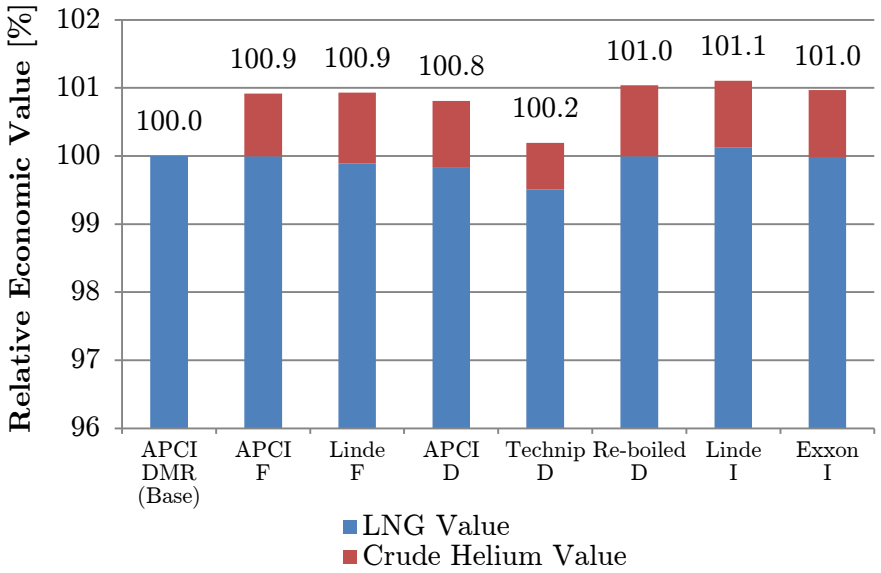


Figure 6.12 Economic Evaluation of Helium Extraction Processes Based on Equal Feed Gas Flow
(F: Flashing, D: Distillation, I: Integration)

Figure 6.12 indicates that all the helium extraction processes integrated in a LNG plant deliver more economic value than a simple LNG plant, which is the baseline process. It means adding a helium extraction process to a LNG plant may bring more profits to the LNG project.

Nevertheless, the extra profits that are generated by helium extraction processes were relatively small, compared to the LNG value. On average, the crude helium value was around 1% of the LNG value produced by the APCI DMR process. It is mainly due to the small amount of crude helium produced and its low price. (See Appendix I for detailed values)

The Technip Distillation process offers just 0.2% more economic value than the baseline process. Even though this process has the largest LNG production rates as seen in Figure 6.4, a large amount of the feed gas has to be extracted as fuel, reducing the economic value of LNG. It is because that this helium extraction process does not produce fuel gas, while consuming the largest amount of energy to produce LNG and crude helium, compared to other processes.

As the economic evaluation is affected by crude helium produced, a sensitivity analysis was performed with different feed gas compositions, helium-lean and helium-rich feed gas. The result is displayed in Figure 6.13. This figure clearly indicates that the higher helium content feed gas has, the more value a helium extraction process can create. With helium-lean feed gas, all the processes could generate less than 1% more value than the APCI DMR process, which is the baseline.

In contrast, the Re-boiled Distillation process could create maximum 2% more value with helium rich feed gas than the baseline. Therefore, one can conclude that more helium content in feed gas makes integrating a helium extraction process into a LNG plant slightly more economically feasible.

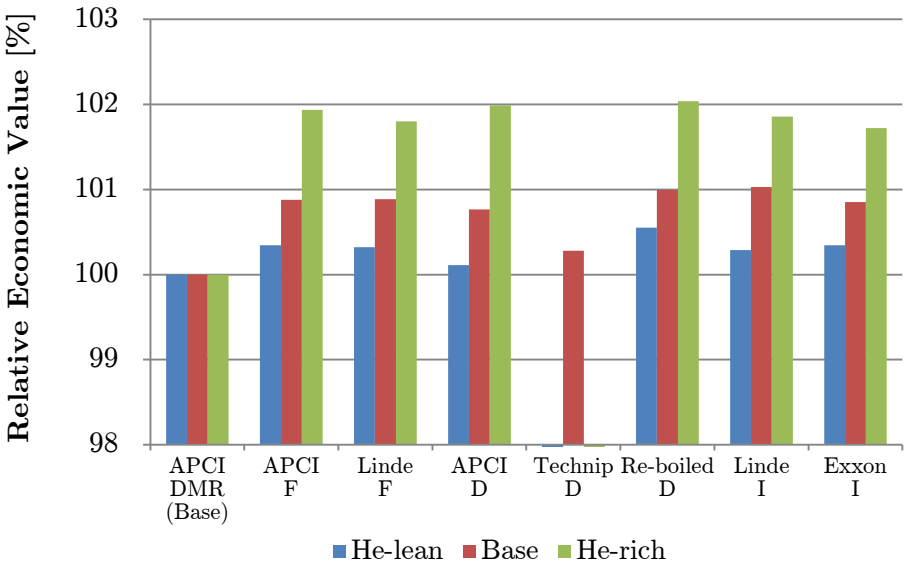


Figure 6.13 The Effect of Feed Gas Composition on Economic Evaluation (F: Flashing, D: Distillation, I: Integration)

6.6 Number of Equipment and Complexity

Required number of equipment for each process studied was counted by referring to the simulation models of the helium extraction processes as illustrated in Chapter 5. Each component shown in the models was considered as one unit. For example, although two compressors in the models mean a multi-stage compressor in reality, it was counted as two units in this work. In addition, size of equipment was not considered since it may lead a unit to be separated in several units in parallel due to its capacity limit. Compressor drivers and valves were not counted in this work. The result is summarised in Table 6.3.

This table indicates that the APCI Distillation process has only 12 units, which is the lowest number of equipment in comparison with the others. A striking feature of this process is that it has also the lowest number of rotating machines. This needs just three compressors. Minimising the number of rotating machinery is one of the design criteria for plant availability. This type of machines is the main cause of plant down time, thus increasing plant unavailability and operating expenditure (OPEX). Another helium extraction process from APCI, the APCI flashing process, also recorded a relatively small number of units, thanks to the simplicity of the process.

Table 6.3 Number of Equipment for Helium Extraction Processes

Unit	APCI Flashing	Linde Flashing	APCI Distillation	Technip Distillation	Re-boiled Distillation	Linde Integration	ExxonMobil Integration
No. of Rotating Units	5	5	3	6	6	10	9
- Compressor	5	5	3	5	5	10	8
- Gas Expander	0	0	0	0	0	0	1
- Liquid Expander	0	0	0	1	1	0	0
No. of Static Units	10	12	9	14	12	17	14
- Cryogenic H.E.	2	2	3	3**	3***	3	3
- Condenser/Cooler	3	4	2	4	4	6	4
- LNG Liquefier*	1	1	1	1	1	1	1
- External Refrigerator	0	0	0	1	0	0	0
- Distillation Column	0	0	1	1	1	1	1
- Phase Separator	4	5	2	4	3	6	5
Total No. of Units	15	17	12	20	18	27	23

* A LNG liquefier originally consisting of a large number of equipment was counted as one unit in this work.
**The HYSYS model for the Technip Distillation process was built with only one cryogenic heat exchanger for simplicity of simulation work even though there are three cryogenic heat exchangers in the patent application. Thus, it was counted as three exchangers in this work.
*** The reboiler of the distillation column, which was modelled as a cooler in the HYSYS model, was counted as one cryogenic heat exchanger in this work.

Unlike the APCI flashing process, the Linde Flashing process has an intermediate number of units, which are similar to the Technip and the Re-boiled Distillation process. One notable thing is the Technip Distillation process requires an external refrigeration system apart from a LNG liquefier as the system has to cool a stream down to below -160°C, which is too low temperature to be covered by typical LNG liquefiers. The external refrigeration system may need a number of units such as refrigerant compressors and cryogenic heat exchangers. However, the system was simply modelled as a cooler in HYSYS and counted as just one unit. Therefore, the Technip Distillation process may require the largest number of units in reality, compared to any other processes.

In contrast, the Linde Integration process has the largest number of units, requiring 12 units more than the APCI Distillation process. Mainly the number of compressors contributes it. Unlike other helium extraction processes, this technology compresses the first flash gas before it is used as a feed stream for the HeXU. Thus, it requires extra compressors for the re-compression of the first flash gas. In addition, this process returns one of the streams coming from the HeXU to the HeXU feed stream, which has a higher-pressure level than the stream leaving from the HeXU. Hence, it is an inevitable to put compressors for pressurising the return stream to the pressure level of the HeXU feed in order to be merged with it. Thus, it has the largest number of rotating machines even though compression stages can be combined. Consequently, more maintenance work is expected for this technology, decreasing the plant availability.

Similarly, another integration-based process from ExxonMobil has the second largest number of units. It is also mainly due to the extra compressors required to recompress the first flash gas, which becomes the HeXU feed. It was obvious that the two integration-based processes have the largest number of units as they combine two types of separation systems, flashing and distillation.

Care should be taken that the number of units alone does not tell which process has higher complexity. Complexity is also related to a process point of view. However, in general, the more units a process contains, the higher complexity it can have. Therefore, from this point of view, the APCI flashing and distillation process were the simplest ones and the Linde Integration and ExxonMobil Integration process were the most complex systems. Even though the Technip Distillation process has a relatively small number of units, the fact

that crude helium production cannot be decoupled from LNG production makes this system difficult to operate. Except this process, all the other helium extraction process could stop producing crude helium without any negative effect on LNG production.

7 Conclusion and Recommendation

Through this report, various process configurations for crude helium production from LNG end-flash gas has been studied and evaluated to see which technology is the most suitable system to be integrated into a LNG plant.

First, when it comes to the performance of helium extraction from LNG, the Linde Flashing process achieved the highest helium extraction rate regardless of change in helium content in feed gas. Even though all the helium extraction processes simulated were pushed to have a helium extraction rate of 90%, the Linde Flashing process recorded 95%, which was the minimum possible value. Although the Re-boiled Distillation process also has the same minimum possible helium extraction rate as the Linde Flashing process, this process showed poor helium extraction rate when feed gas is rich in helium.

From an energy efficiency point of view, the two integration-based processes, the Linde Integration and the ExxonMobil Integration process displayed the lowest crude helium production specific power, which was around one-fifth of other processes. However, the total amount of energy used for helium compression was ranged from 20 to 150 kW. This amount is relatively small, compared to the energy used for a LNG liquefier or fuel compressors. Therefore, crude helium production specific power does not affect a selection of a helium extraction process.

Regarding LNG production, the Re-boiled Distillation process was the most efficient system with a whole range of feed gas composition change. The specific power was even lower than the baseline process, the APCI DMR process. The ExxonMobil Integration process showed the highest LNG production specific power regardless of change in helium content in feed gas. The Technip Distillation process had the biggest LNG production rate, which was 0.3MTPA larger than the baseline process as this process does not produce fuel gas. Nevertheless, this advantage is offset by the second largest LNG production specific power of this technology.

When the composition of feed gas changes, especially helium and thus nitrogen content, one of the most important things is to meet the LNG specification, 1 mole% nitrogen in LNG. Regarding this, only the APCI Distillation and the Re-boiled Distillation process could achieve less than 1 mole% N_2 in LNG when feed gas is rich in helium and nitrogen. Even though a case study with the variation of helium and nitrogen content in feed gas was not able to perform for the Technip Distillation process, this process is also expected to meet the LNG specification when having helium-rich feed gas thanks to the distillation column.

An attention has to be paid to nitrogen content in fuel gas as gas turbines, especially industrial type used in this project, can accept maximum 40 mole% N_2 in fuel. In that sense, all the processes simulated produced fuel gas having nitrogen content less than 40 mole% with the base feed gas and helium-lean feed gas. However, only the APCI Distillation process could generate fuel gas having less than 40 mole% nitrogen when feed gas is rich in helium and thus nitrogen. Therefore, the APCI Distillation process was superior to other helium extraction processes regarding the nitrogen fraction in fuel gas.

In terms of LN_2 production, only the Technip Distillation and the Linde Integration process could generate it and the quality of the produced liquefied nitrogen was similar each other. These two processes were also only the systems having a nitrogen vent stream and the both processes met the environmental regulation, which is less than 1 mole% methane in vent streams. However, the methane concentration in N_2 vent stream from the Technip Distillation process with base feed gas case reached around 0.5 mole%. Thus, care has to be taken when feed gas becomes leaner in helium and nitrogen as it may increase the CH_4 fraction in the N_2 vent stream.

Concerning freezing of impurities in helium extraction process, the ExxonMobil Integration process displayed an excellent performance, recording the warmest HeXU working temperature due to the high HeXU working pressure. Therefore, this may result in less attention to reduce the amount of impurities flowing into helium extraction system, which requires additional cost and energy. In contrast, the Technip Distillation and the Linde Integration process had the lowest HeXU working temperature, which is a required temperature level for LN_2 production.

When considering complexity, the APCI Distillation process was superior to other processes, having only 12 units. The two integration-based processes from Linde and ExxonMobil required a lot of equipment for helium extraction, 27 and 23 units respectively. Especially the number of rotating machines was around two times bigger than other processes, meaning lower plant availability due to the high failure rate of rotating units. One noticeable thing is that only the Technip Distillation process could not decouple crude helium production from LNG production, this will make this process difficult to operate.

Finally, the economic evaluation of helium extraction processes indicates that all the crude helium production technologies produce more economic value than the APCI DMR process even when feed gas has low helium content ca. 0.02 mole%. Nevertheless, the difference was just around 0.5 %, compared to the economic value created by the APCI DMR process. When the helium content increases to 0.05 mole%, which is typical helium fraction in LNG streams, the difference rises to about 1 %. In helium-rich feed gas case (0.1 mole% He), the difference goes up to around 2%. Thus, the economic feasibility of helium extraction process highly depends on the helium concentration in feed gas and the unit price of crude helium. The price of crude helium is highly likely to rise steeply in near future since the helium storage in U.S., which vastly affects helium price, is going to be run out. Therefore, adding a helium extraction process to an existing LNG plant or integrating it into a new LNG plant project, will be a more economically feasible, creating extra profits.

In conclusion, the selection of the most suitable helium extraction process completely depends on the situation. As each process has distinct features to others, the choice has to be made based on the situation of a helium extraction project like helium content in feed gas. Nevertheless, the Re-boiled Distillation process may be the best option regardless of helium and nitrogen fraction in feed gas. It is mainly thanks to the highest helium extraction rate and the lowest LNG production specific power, while having relatively simple process structure. In addition, even though feed gas becomes richer in helium and nitrogen, this process could reduce nitrogen content down to 1 mole%. One disadvantage of the Re-boiled Distillation process is that it tends to give smaller LNG production rate measured in unit mass when having helium-rich feed gas. However, it is compensated by a larger LNG heating value, increasing LNG economic value. Another drawback is that this process has a relatively low helium extraction rate when feed gas is rich in helium. Nevertheless, the effect of decrease in helium extraction rate on plant economics is small enough to ignore.

8 Further Work

As this project aimed at a primary screening of helium extraction configurations for a LNG plant, more detailed analysis may be required to have better understanding about each process. First, sensitivity analysis to helium and nitrogen content in feed gas for the Technip Distillation could be performed to have more fair comparison with other processes since it could not be conducted during this project. In addition, gas turbine systems can be modelled and integrated with each HYSYS model for the helium extraction processes as it is omitted in this project. It is to know how the composition change in fuel gas, occurred by helium extraction process, will affect the performance of gas turbines.

Some of helium extraction processes produce more fuel gas than what they need. Thus, a system for returning a portion of fuel gas to a feed stream could also be modelled in HYSYS in order to measure exact process performance. Investigation of exergy analysis could be useful to find where the largest exergy loss of each process come from and see if there is a chance to improve process efficiency. To have precise economic evaluation of helium extraction process, calculating the size of equipment in the HYSYS models could be of interest. Plant availability for each helium extraction process could also be measured to understand how much the extraction process added to a LNG plant will affect the reliability of the total system.

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Appendices

Appendix A
Optimisation of APCI Flashing Process

To optimise this process, the outlet pressure of the second J-T valve (VLV-2), the only one free variable of this process, was manipulated from 2 bara to 4 bara in a case study. The outcome of the case study indicates 2.7 bara is the pressure level that brings the best performance of this technology. One noticeable thing is that all the graphs acquired as the outcome show discontinuous trend of curves in the graphs due to numerical tolerance of the simulation model for this process.

Figure A-1 indicates that a lower outlet pressure of the second J-T valve brings a larger LNG molar flow, meaning a greater LNG production rate. The largest LNG molar flow of 20280kgmol/hr was recorded in a span from 2.3 bara to 2.7 bara. Similarly, the duty of LNG liquefier is increased with decreasing the pressure level, having the largest duty in a range from 2.3 bara to 2.7 bara. However, the increased duty is compensated by the larger LNG production rate. As seen in Figure A-1, the change in LNG liquefier cold duty is exactly proportional to the change in LNG molar flow, meaning the constant specific power for producing a unit molar flow throughout the whole range of the second J-T valve outlet pressure. Thus, the outlet pressure level of the valve has to be within the range from 2.3 bara to 2.7 bara to maximise LNG production. As this process is structured to have a constant crude helium production rate, a helium extraction rate of 90% in this paper, crude helium molar flow was disregarded when choosing this pressure level.

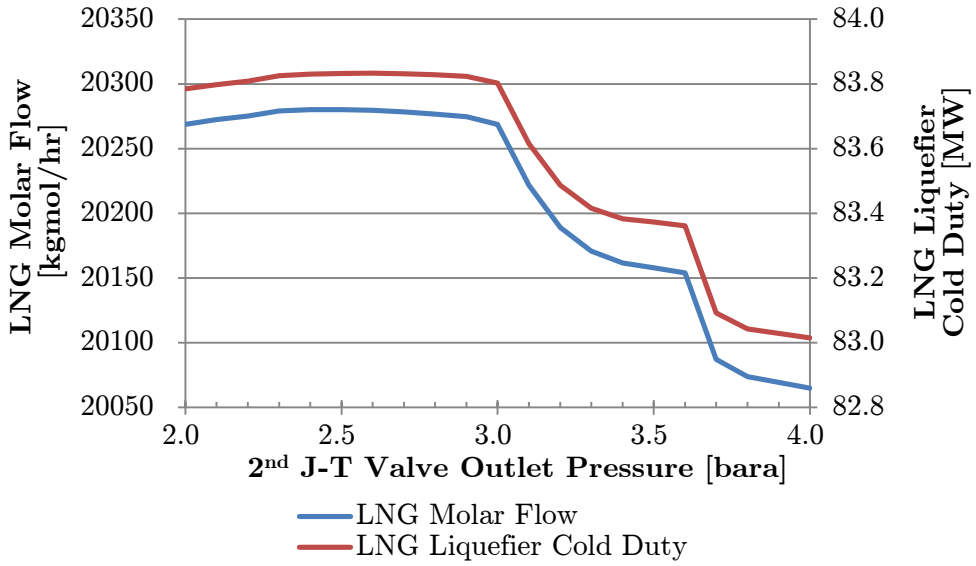

Figure A-1 The Effect of Change in the Second J-T Valve Outlet Pressure on LNG Molar Flow and LNG Liquefier Cold Duty

Thus, the optimal outlet pressure of the second J-T valve was selected to have the minimum compression work for fuel and crude helium. As illustrated in Figure A-2, the smallest compression work is found at 2.7 bara. In addition, the variation of the LNG Liquefier cold duty is almost offset by the change in the compression work, making total energy used for this process constant.

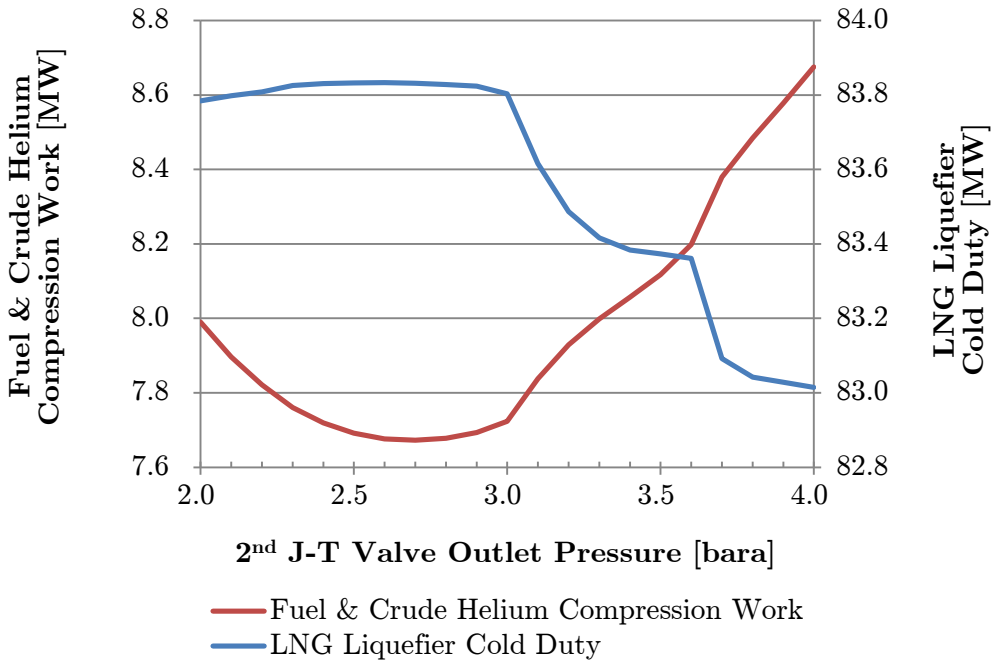

Figure A-2 The Effect of Change in the Second J-T Valve Outlet Pressure on Fuel & Crude Helium Compression Work

In contrast, the second J-T valve outlet pressure of 2.7 bara brings inferior performance of HeXU, compared to the case of higher pressure level than that. As seen in Figure A-3, LMTD of the HeXU HE becomes smaller when increasing the outlet pressure of the second J-T valve, lowering exergy loss of this heat exchanger. Moreover, the inlet temperature of the phase separators in the HeXU (HeXU Phase Separator) increased with the outlet pressure level, which makes the HeXU to hold larger amount of impurities. However, in this optimization work, the difference of the LMTD and the phase separator inlet temperature between the case of 2.7 bara and 4.0 bara which is the highest pressure in the case study are not significant. Thus, the outlet pressure level of the second J-T valve was finally set as 2.7 bara, ignoring the small difference.

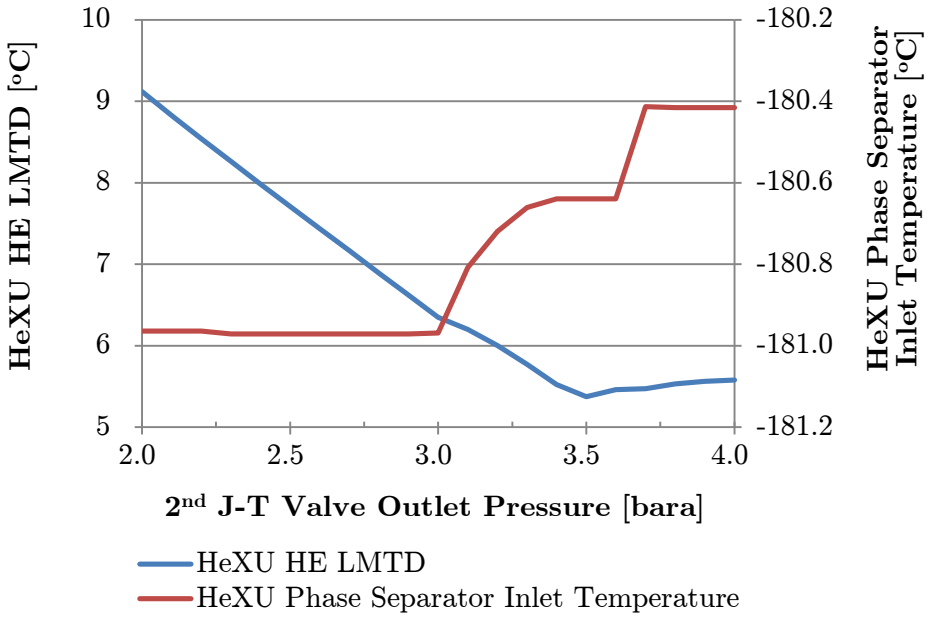

Figure A-3 The Effect of Change in the Second J-T Valve Outlet Pressure on HeXU HE LMTD and HeXU Phase Separator Inlet Temperature

Appendix B
Optimisation of Linde Flashing Process

A case study was performed to find the optimal helium mole fraction in the first flash gas (F5). Linde suggests in their patent application that the helium mole fraction typically ranges from 2% to 20%. Thus, the case study was performed by changing the helium mole fraction in the flash gas from 2% to 20%. To change the helium content, the outlet pressure of VLV-1 was adjusted. The result of the case study indicates 2 mole% He in the first end-flash brings the best performance of this process. One noticeable thing is that all the graphs acquired as the result show small fluctuations of curves in the graphs due to numerical tolerance of the simulation model for this process.

Figure B-1 shows that a lower He mole fraction in the gas gives a larger LNG molar flow, which means a larger production rate of LNG. When evaluating this case study, a crude helium production rate is also counted as a performance index. Unlike the APCI Flashing process, this process was not controlled to have a constant crude helium production rate, denoted as helium extraction rate of 90% in this report. The crude helium production rate also follows the trend of LNG production in a range from 5 mole% to 20 mole%. From 5 mole% to 2 mole%, it is decreased. Thus, 5 mole% He in the first end-flash may be the best choice when it comes to maximum crude helium production. However, if the helium mole fraction is increased from 2 mole% to 5 mole%, LNG production rate is decreased by 426 kgmol/hr, while crude helium production rate is increased by only 0.2 kgmol/hr. Therefore, 2 mole% He in the first end-flash may be the proper value unless the economic value of 0.2kgmol crude helium is larger than the value of 426kgmol LNG.

On the other hand, Figure B-2 shows that the lower helium mole fraction of the first end-flash (F5) results in larger cold duty of the LNG liquefier. This, however, is almost offset by smaller fuel and crude helium compression work with decreasing helium mole fraction of F5. In addition, the increasing cold duty with the He mole fraction is also compensated with growing LNG production rate as indicated in Figure B-1. As this gives the constant value of the cold duty per LNG molar flow, one may disregard the change in the LNG Liquefier cold duty and select 2 mole% to have the lowest Fuel and crude helium compression work.

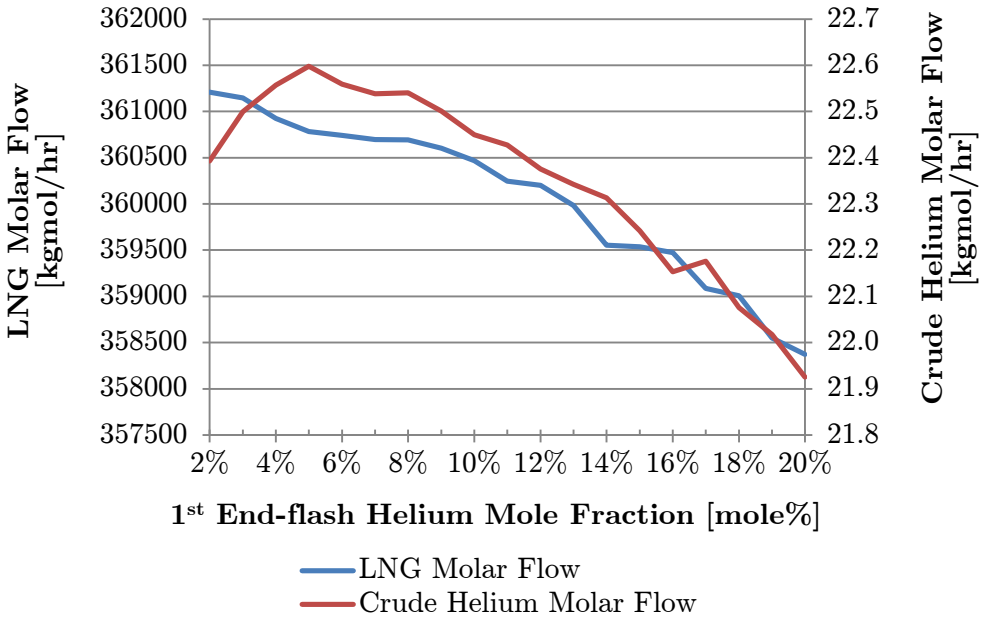

Figure B-1 The Effect of Change in the Helium Mole Fraction of the First End-flash on LNG and Crude Helium Production Rate

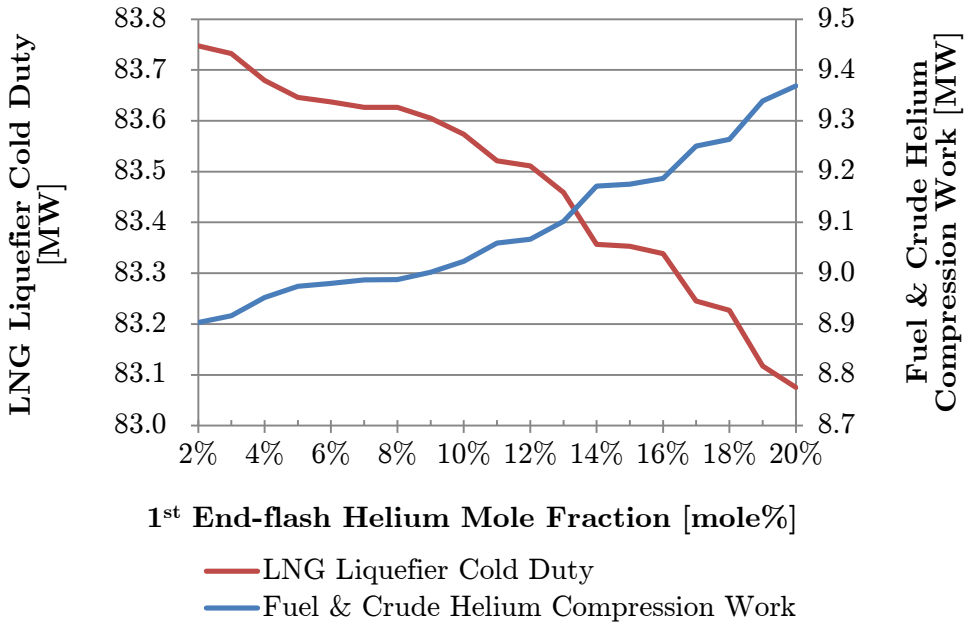

Figure B-2 The Effect of Change in the Helium Mole Fraction of the First End-flash on LNG Liquefier Cold Duty and Fuel & Crude Helium Compression Work

Producing the first end-flash (F5) that has 2 mole% He is also a reasonable choice as only this case has the cooling effect through the first J-T valve (VLV-1). Figure B-3 indicates that the sub-cooled stream F3 have a heating effect via the first J-T valve from 3 mole% to 20 mole%, making the first end-flash warmer. It may increase the cold duty of the heat exchanger in the HeXU as the first end-flash has to be cooled down to a specific temperature in the exchanger to produce He-rich crude helium. This heating effect also increases the exergy loss of the valve.

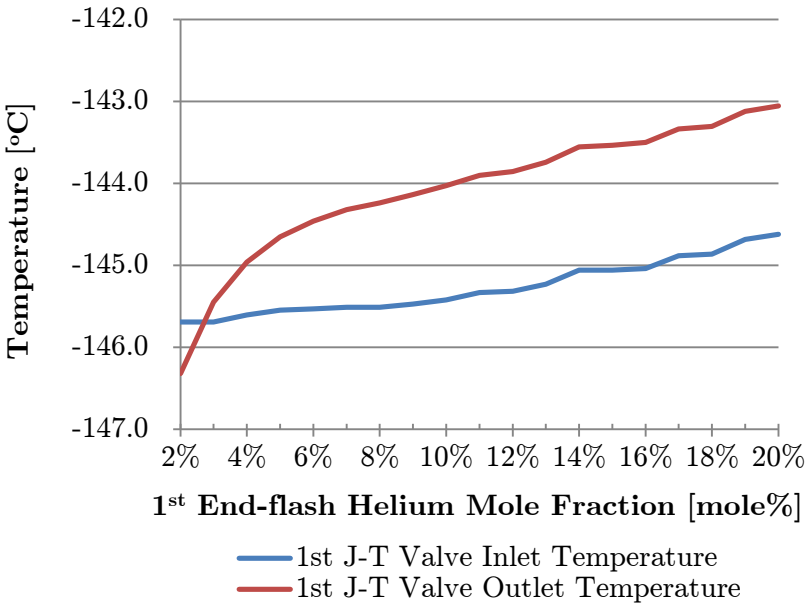

Figure B-3 The Effect of Change in the Helium Mole Fraction of the First End-flash on Temperature Change via the First J-T Valve

Other indexes also show that lower helium mole fraction of the first end-flash is more favourable for a HeXU. As illustrated in Figure B-4, the leaner in helium the first flash gas become, the smaller LMTD the HeXU HE in this process has. The smaller LMTD gives lesser exergy loss through the heat exchanger. In addition, the inlet temperature of the phase separators in the HeXU (HeXU Phase Separator-1 and 2) decreases with the helium mole fraction of the first end-flash. Thus, it raises the working temperature of the HeXU and allows the HeXU to handle a larger amount of impurities.

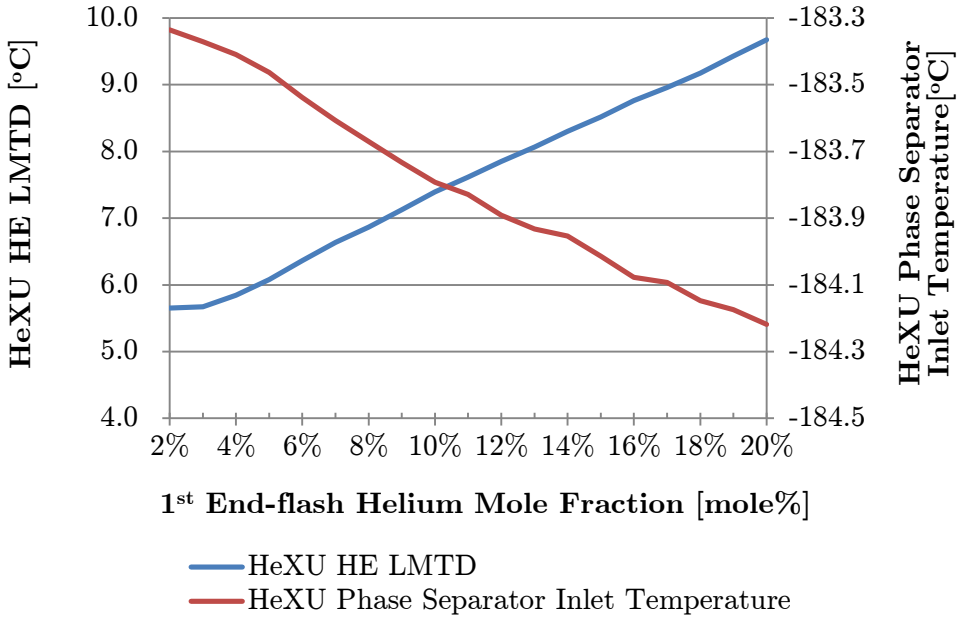

Figure B-4 The Effect of Change in the Helium Mole Fraction of the First End-flash on HeXU HE LMTD and HeXU Phase Separator Inlet Temperature

Appendix C
Optimisation of APCI Distillation Process

In the APCI Distillation process, there is only one free variable affecting the performance of the process, the split ratio of the nitrogen rejection column overhead product between stream O2 and O4. As stream O4 is fully refluxed to the column after being condensed in Condenser HE, the split ratio affects the performance of the column and the composition of the overhead product, affecting the performance of the HeXU. Therefore, a case study was conducted by changing the flow ratio of stream O4. The result of the case study shows that the flow ratio of 0.5 for stream O4 is the optimal value for this process.

In case of the total LNG liquefier duty, which is the sum of LNG Liquefier and LNG Sub-cooler cold duties, it rises with the flow ratio of stream O4 but it is offset by increasing LNG production, which is indexed as LNG chemical energy in Figure C-1. The trend of change in the total LNG liquefier duty is almost the same as the change in LNG chemical energy, meaning that the value of produced LNG per unit cold duty of the LNG liquefier is nearly constant. Therefore, the increasing total LNG liquefier duty was disregarded when choosing the optimal value of the flow ratio of stream O4.

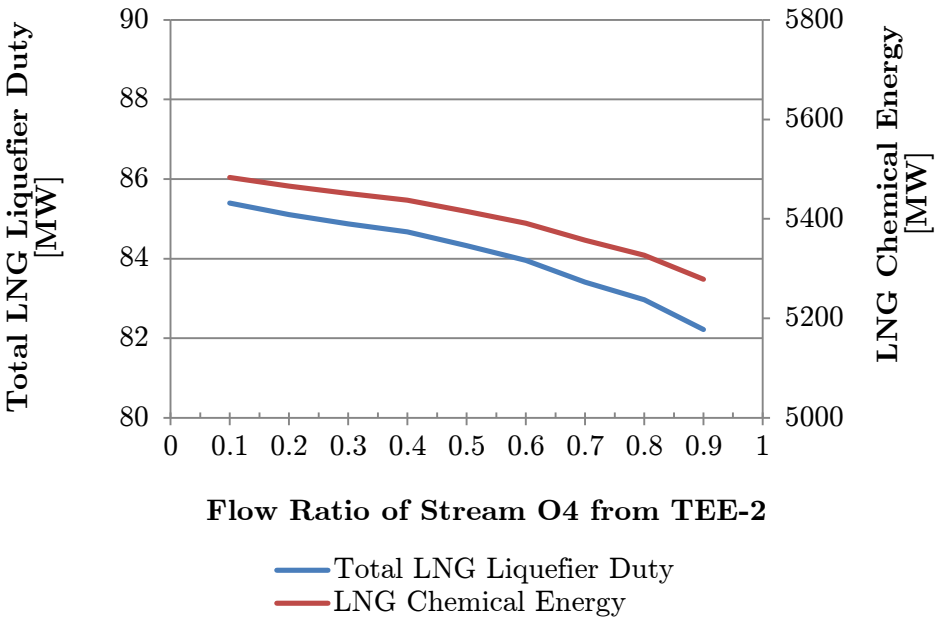

Figure C-1 The Effect of Change in the Flow Ratio of Stream O4 from TEE-2 on Total LNG Liquefier Duty and LNG Chemical Energy

Figure C-2 shows that a lower flow ratio of stream O4 gives a larger reflux flow rate of the distillation column condenser, which will increase the size of the distillation column. In addition, the temperature of the distillation column overhead vapour falls with decreasing the flow ratio of stream O4. It is an unfavourable condition for the column when it comes to the freezing of impurities. Therefore, a higher flow ratio of stream O4 brings better process performances as it gives a smaller size and a higher working temperature of the column.

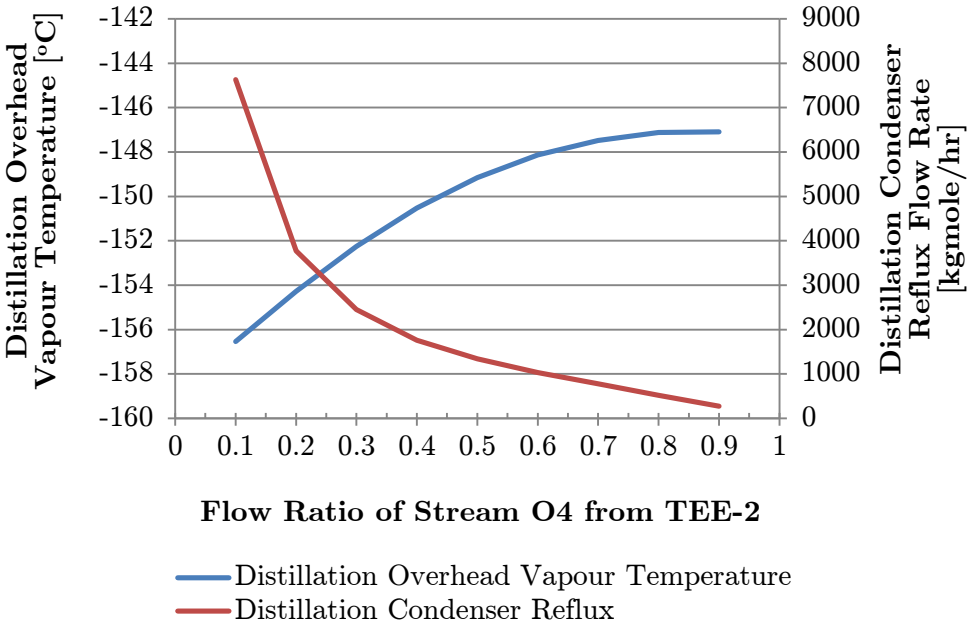

Figure C-2 The Effect of Change in the Flow Ratio of Stream O4 from TEE-2 on Distillation Overhead Vapour Temperature and Condenser Reflux Flow Rate

Figure C-3 indicates that the total compression work, which is the sum of fuel gas and crude helium compression work, gradually increases with the flow ratio of stream O4 up to 0.5. Afterwards, the total compression work increases significantly. As mentioned in the previous paragraph, one has to set the flow ratio of stream O4 as a high value. However, a higher flow ratio of stream O4 gives a large increase in the total compression work especially after the ratio of 0.5. Therefore, the flow ratio of 0.5 for stream O4 may be the best value as it delivers a relatively low compression work before it soars up with the increasing flow ratio.

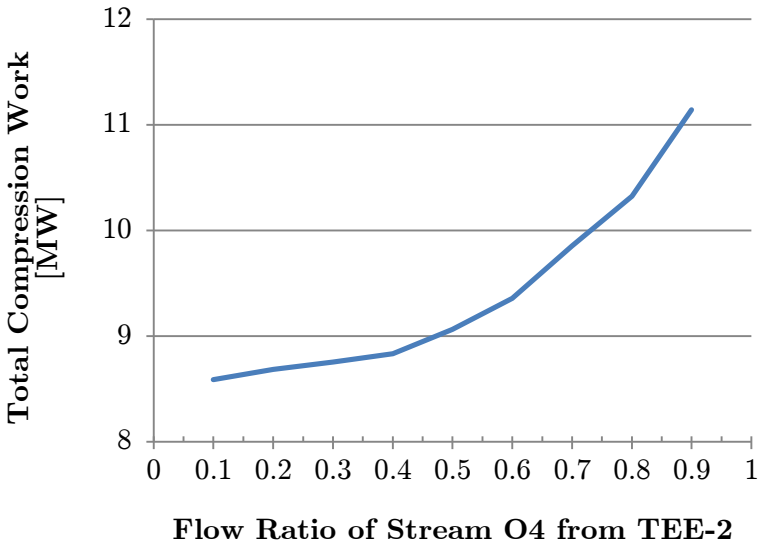

Figure C-3 The Effect of Change in the Flow Ratio of Stream O4 from TEE-2 on Total Compression Work

Appendix D
Optimisation of Technip Distillation Process

A case study was performed to optimise the Technip Distillation process with the only one free variable, the split ratio of TEE-2 between stream N2 Vent and O4 in Figure 5.5. As the split ratio affects the amount of the two reflux streams X7 and X13 in Figure 5.5, the split ratio only can be manipulated in a limited boundary. Otherwise, the nitrogen rejection column could not be converged in HYSYS. The limited range of the stream N2 Vent flow ratio, which makes the column converged in this simulation work, was from 30.5% to 32% of stream O3. Thus, the case study was performed with the range and the result indicates that the N2 Vent stream flow ratio of 0.305 was finally chosen as the optimal value.

Figure D-1 indicates that helium extraction rate of this system decreases with the flow ratio of stream N2 Vent. As helium is dumped to the atmosphere through stream N2 Vent, a larger flow rate of N2 Vent means more leakage of helium from the system. As seen in Figure D-1, the total compression work used for this process also has the same trend as helium extraction rate. Thus, from an energy consumption point of view, the largest flow ratio of stream N2 Vent may be the best choice. However, the flow ratio of 0.305 was selected as the optimal flow ratio of stream N2 Vent. Since this process could not achieve a helium extraction rate of 90%, the flow ratio of stream N2 Vent was used to maximise the extraction rate.

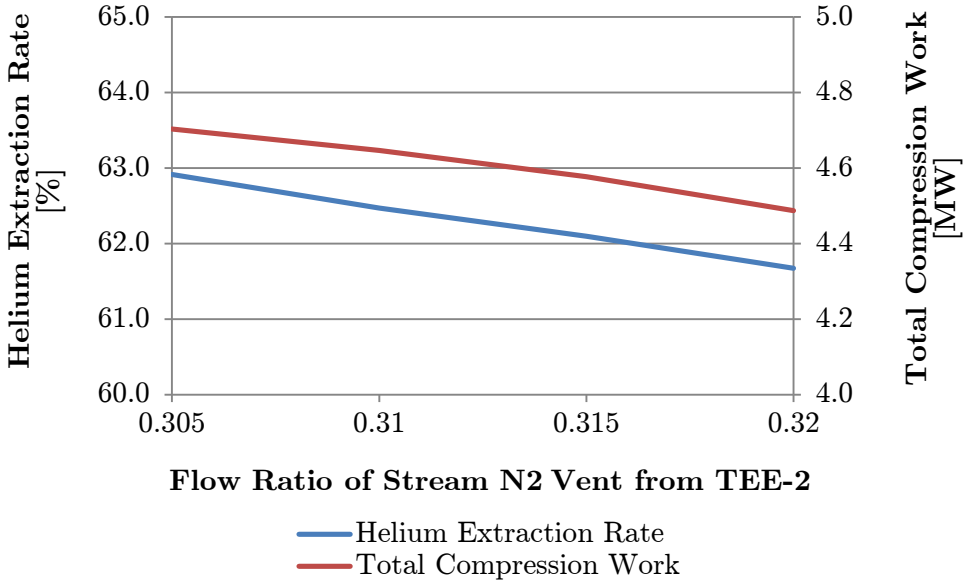

Figure D-1 The Effect of Change in the Flow Ratio of Stream N2 Vent from TEE-2 on Helium Extraction Rate and Total Compression Work

On the other hand, a smaller flow ratio of stream N2 Vent causes the decreasing temperature of the nitrogen rejection column overhead vapour (O1 in Figure 5.5) in this process as seen in Figure D-2. The lower temperature level of the overhead gas means a lower working temperature of the nitrogen rejection column, which is unfavourable condition when it comes to freezing of impurities.

Thus, a higher flow ratio of stream N2 Vent has to be chosen if impurities in the process are increased. However, there is no contamination in the feed gas in this project so this factor was disregarded to concentrate on rising helium extraction rate of this process. Therefore, the flow ratio of 0.305 was finally chosen as the optimal value of this simulation model.

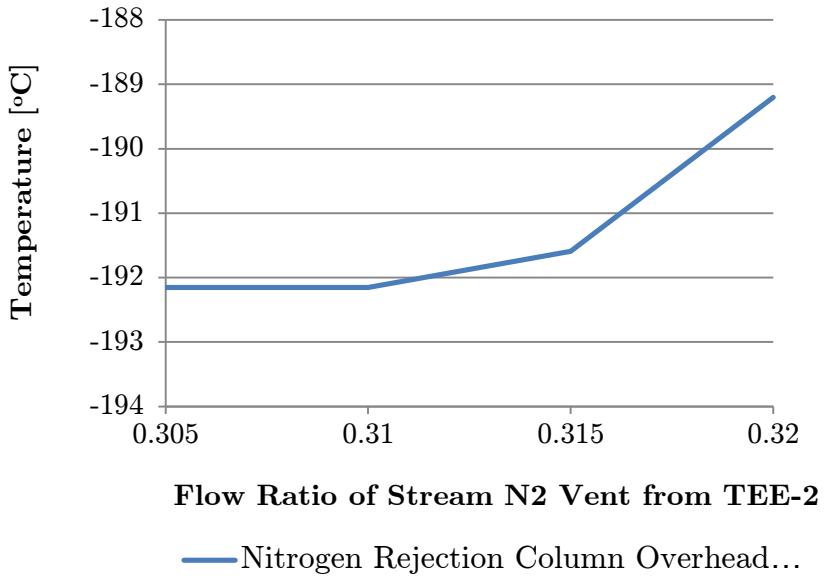

Figure D-2 The Effect of Change in the Flow Ratio of Stream N2 Vent from TEE-2 on the Temperature of Nitrogen Rejection Column Overhead Vapour

Appendix E
Optimisation of Re-boiled Distillation Process

To optimise this process, the outlet pressure of the liquid expander (Liquid Exp in Figure 5.6), which is only one free variable, was manipulated in a case study. The result of the case study shows 4 bara is the value delivering optimal performance of this process.

First, Figure E-1 indicates that the LNG liquefier duty decreases with increasing outlet pressure of the liquid expander. At the same time, LNG production, which is indexed as LNG chemical energy in Figure E-1 also falls. The trend of change in the total LNG liquefier duty is almost the same as the change in LNG chemical energy, meaning that the value of produced LNG per unit cold duty of the LNG liquefier is nearly constant. Therefore, these two indexes were disregarded when choosing the optimal value of the liquid expander outlet pressure.

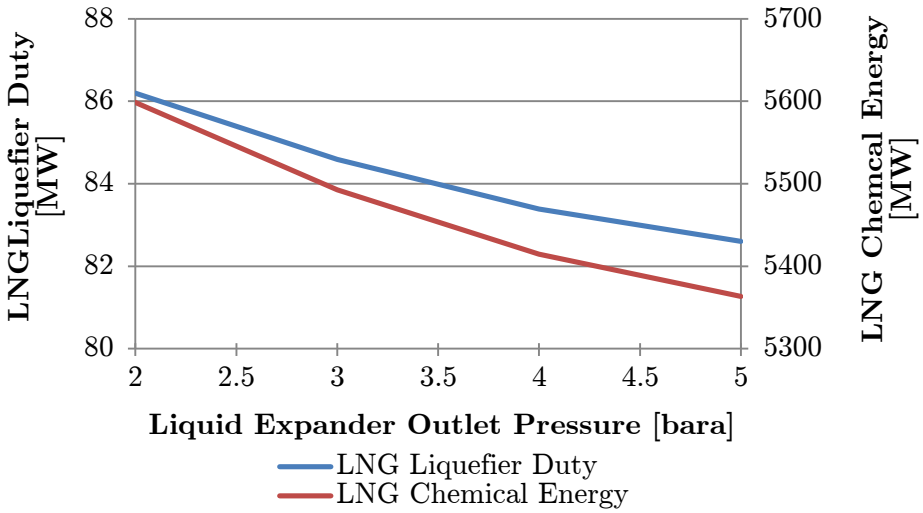

Figure E-1 The Effect of Change in the Liquid Expander Outlet Pressure on LNG Liquefier Duty and LNG Chemical Energy

Figure E-2 shows that the working temperature and pressure of the HeXU in this process rise with increasing liquid expander outlet pressure. The working temperature/pressure of the HeXU means the temperature/pressure of a stream, which has the lowest temperature/pressure level in the HeXU. When considering freezing of impurities in the HeXU, a higher pressure may be the optimal value. In addition, HeXU working pressure goes down to under atmospheric pressure below the liquid expander outlet pressure of 3.5 bara, which may give extra attention about permeation of air into the HeXU. Therefore, the outlet pressure of the liquid expander has to be at least equal or larger than 4 bara.

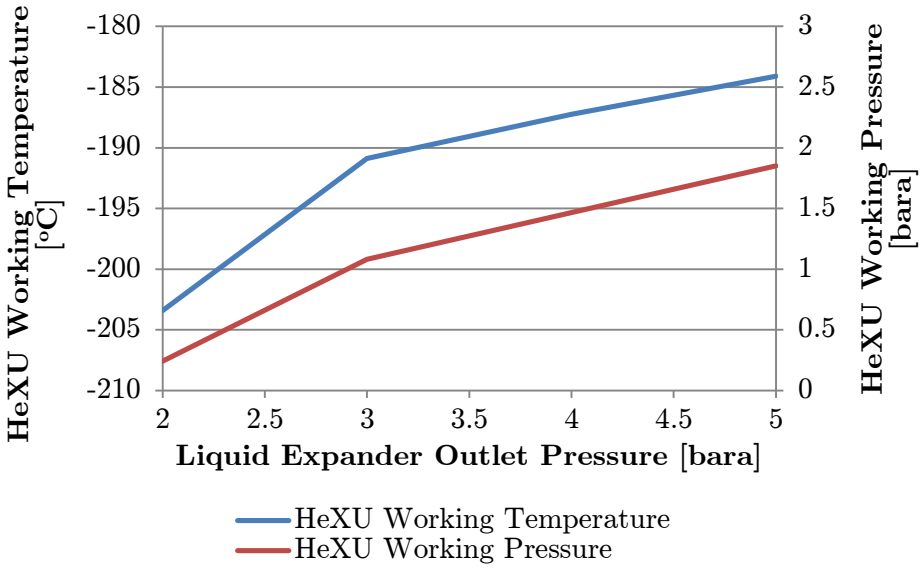

Figure E-2 The Effect of Change in the Liquid Expander Outlet Pressure on HeXU Working Temperature and Pressure

Over the liquid expander outlet pressure of 4 bara, helium extraction rate decreases as shown in Figure E-3. In conclusion, the liquid expander outlet pressure of 4 bara was selected as the optimal value to maximise helium extraction rate of this process, while keeping HeXU working pressure above atmospheric pressure.

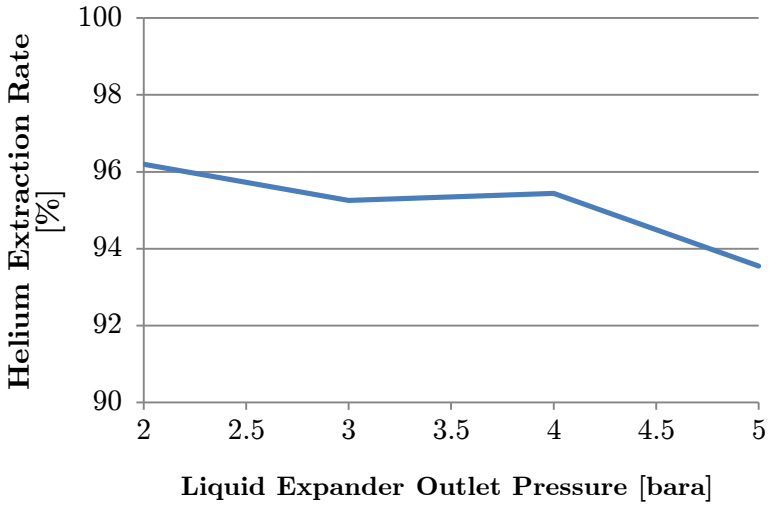

Figure E-3 The Effect of Change in the Liquid Expander Outlet Pressure on Helium Extraction Rate

Appendix F
Optimisation of Linde Integration Process

As there are two free variables in this process, a case study was performed with the each free variable to find proper values of the variables, bringing the best performance. First, the effect of change in the pressure of the compressed first end-flash (F8) was addressed in Figure F-1 and Figure F-2. As suggested by Linde, the range of the pressure level was limited from 15 bara to 30 bara. Other variable, the flow ratio of X1 from TEE-2, was set to 0.3. The outcome of the case study concludes 15 bara is the optimal value for the pressure of the compressed first end-flash.

Figure F-1 shows that the total compression work increases with the pressure level. The total compression work means the sum of the first end-flash compression, the HeXU reflux stream compression and fuel compression work. Thus, 15 bara, which is the lowest level of the pressure range, may be the best pressure value to minimise the energy consumption of this process.

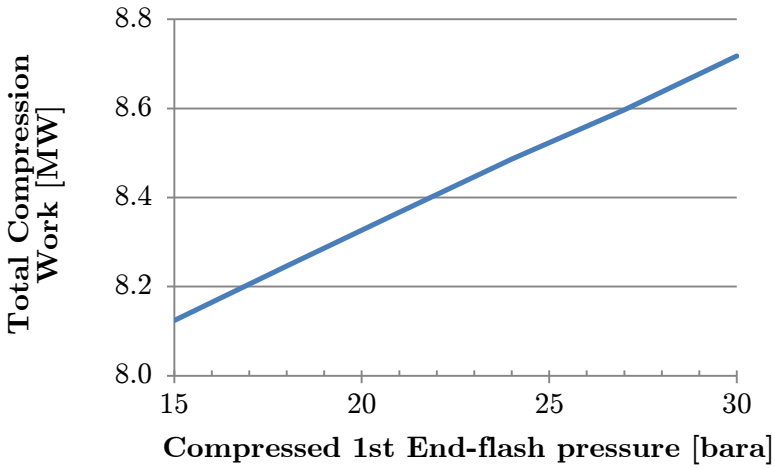

Figure F-1 The Effect of Change in the Pressure of the Compressed First End-flash on Total Compression Work

In contrast, a higher-pressure level of stream F8 leads to a warmer HeXU Phase Separator inlet temperature, thus increasing the working temperature of the HeXU as indicated in Figure F-2. Due to the high pressure, methane and nitrogen in the compressed first end-flash is condensed at a higher temperature, making the flash rich in He. On the other hand, the LMTD of HeXU HE rises with the pressure level, causing more exergy loss. Therefore, 15 bara was chosen as it gives the smallest LMTD of HeXU HE. The inlet temperature of the phase separator was disregarded during this evaluation as the case of 15 bara already has a high enough temperature, which can avoid freezing of impurities.

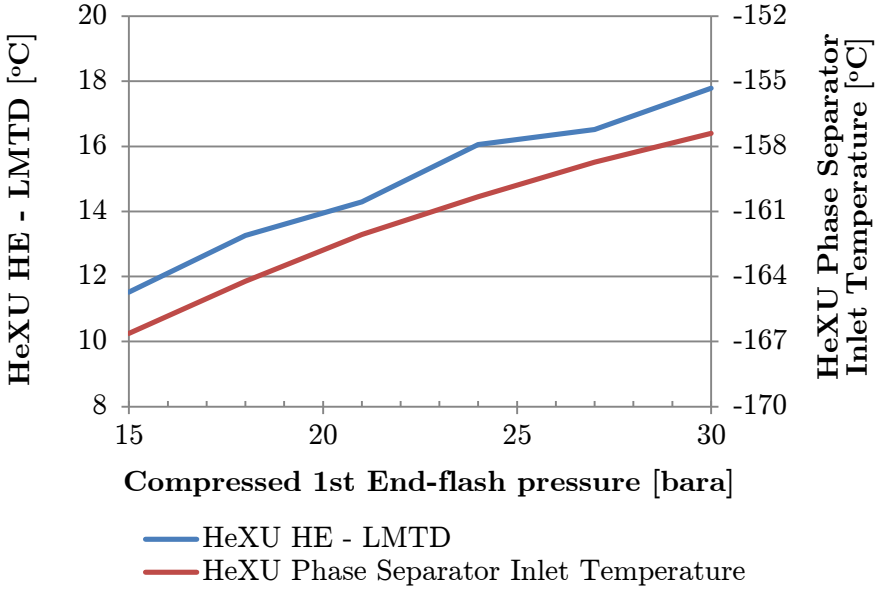

Figure F-2 The Effect of Change in the Pressure of the Compressed First End-flash on the LMTD of HeXU HE and HeXU Phase Separator Inlet Temperature

Next, another variable, the flow ratio of X1 from TEE-2, was case-studied with a range from 0.2 to 0.4. The pressure level of the compressed first end-flash was set to 15 bara, which is the optimal value. The result of the case study was introduced by Figure F-3, Figure F-4 and Figure F-5 and it indicates that the flow ratio of 0.32 is the optimal value. First, Figure F-3 indicates that the total compression work rises with increasing the flow ratio of X1. As stream X1 means the reflux stream to the HeXU feed (F8), a larger flow ratio of X1 imposes more duty on the reflux compressor (Reflux Comp-1-3). Thus, this figure tells the lowest flow rate, 0.2 is the optimal value when only considering the total compression work.

Figure F-4 shows there is no change in LMTDs of HeXU HE and Cold Recovery HE with flow ratio of X1. Therefore, the LMTD values were disregarded when choosing the optimal value of the flow ratio.

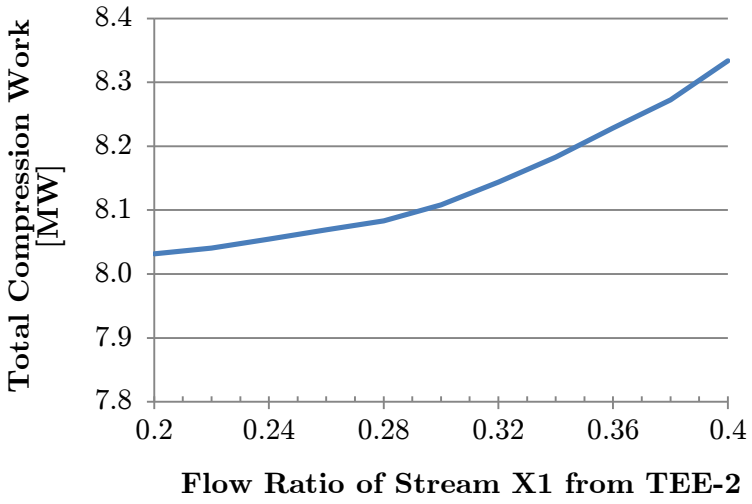

Figure F-3 The Effect of Change in the Flow Ratio of Stream X1 from TEE-2 on Helium Extraction Rate and Total Compression Work

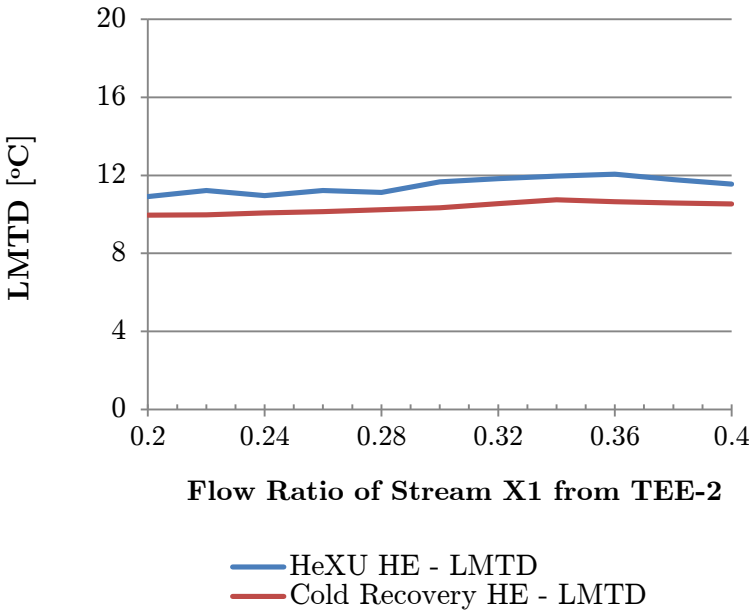

Figure F-4 The Effect of Change in the Flow Ratio of Stream X1 from TEE-2 on the LMTD of HeXU HE and Cold Recovery HE

Figure F-5 indicates that a smaller flow ratio of X1 causes a lower working temperature of HeXU Phase Separator and HeXU Rectifier, which is an unfavourable condition. To increase the working temperature of the HeXU, the flow ratio has to be larger than 0.2, which was the selected value based on the

total compression work. Therefore, the flow ratio of 0.32 was chosen to rise the working temperature of the HeXU as this value delivers the highest working temperature and the relatively smaller compression work.

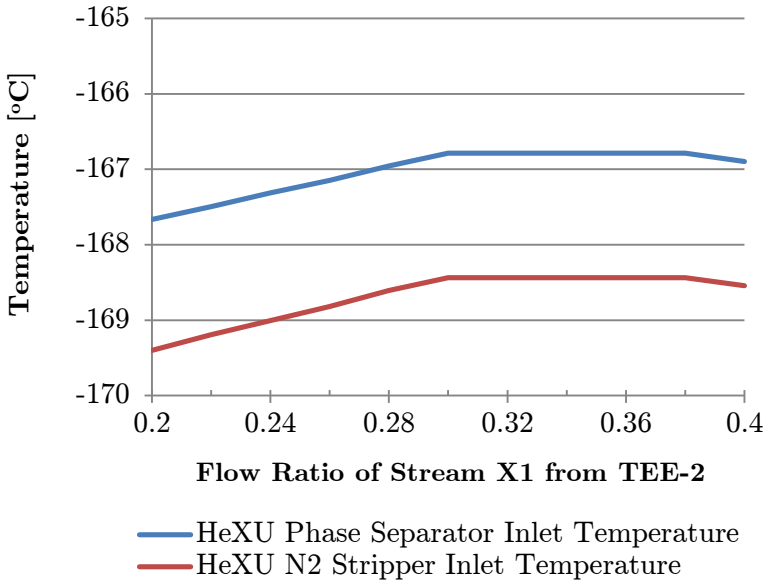

Figure F-5 The Effect of Change in the Flow Ratio of Stream X1 from TEE-2 on HeXU Phase Separator Inlet Temperature and HeXU N2 Stripper Inlet Temperature

Appendix G
Optimisation of ExxonMobil Integration Process

To optimise this process, a case study was conducted by changing the split ratio of TEE-2 between stream E1 and E2, which is only one free variable of this technology. The result of the case study reveals that the flow ratio of 0.8 is the optimal value for stream E2.

As shown in Figure G-1, the total compression work reduces with the flow ratio of E2 until 0.8. The total compression work means the sum of the first end-flash compression, the crude helium compression and fuel compression work minus the work produced by the gas expander, Fuel Exp. The trend is because that an increasing flow ratio of E2 reduces the flow rate of the other stream coming from TEE-2 (E1), which is throttled to a low-pressure level and re-compressed by a fuel compressor. Thus, an increase in the flow ratio of E2 means reduced fuel gas compression work. In addition, a larger flow rate of E2 brings more work produced through a gas expander as the high-pressure stream E2 is depressurised to 20 bara, which is the fuel gas specification.

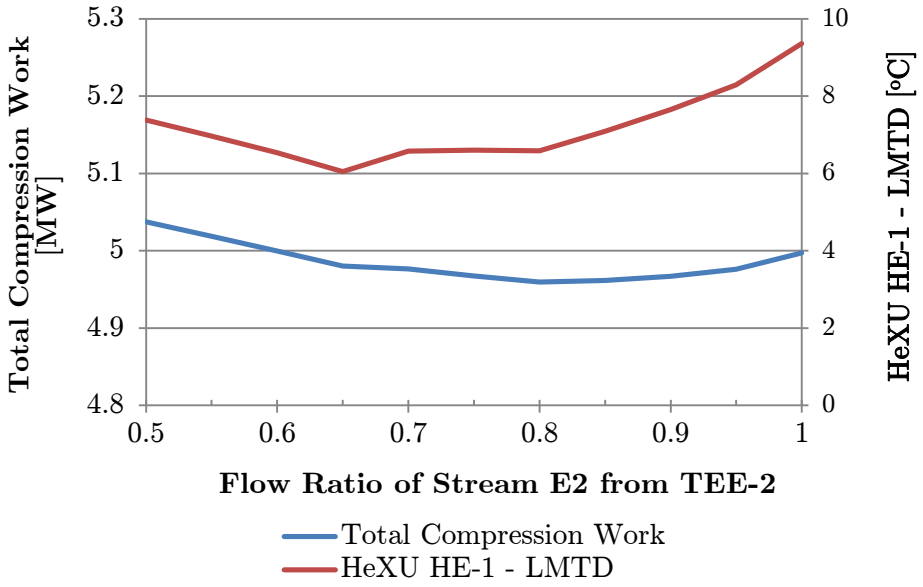

Figure G-1 The Effect of Change in the Flow Ratio of Stream E2 from TEE-2 on the Total Compression Work and the LMTD of HeXU HE-1

Afterwards, the compression work starts increasing with the flow ratio of E2. Even though the flow rate of E2 is increased, the pressure difference between the inlet and outlet of the gas expander becomes very small, thus producing smaller amount of work. Therefore, the optimal value of the flow ratio of E2 may be 0.8 from an energy point of view. The trend of the LMTD of HeXU

HE-1 is also similar to the one of the total compression work, showing that the LMTD at the flow ratio of 0.8 is the second lowest value throughout the whole range of the flow ratio. It means relatively small exergy loss is generated from HeXU HE-1. Therefore, the flow ratio of 0.8 was chosen as it gives a good performance of the heat exchanger and delivers the lowest compression work.

Appendix H
HYSYS Model for APCI DMR Process

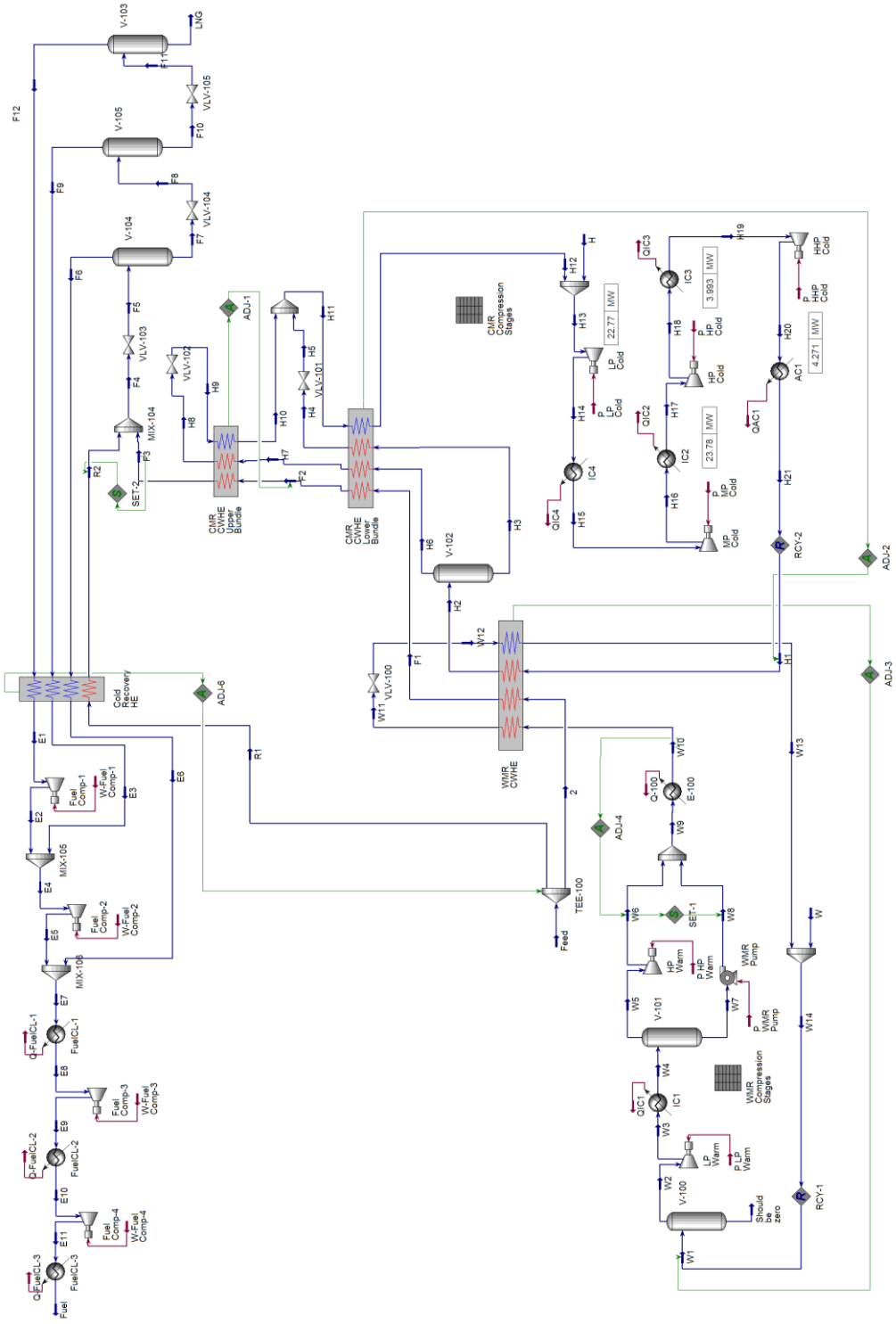

Figure H-1 HYSYS Model for the APCI DMR process

Appendix I
Economic Evaluation of Helium Extraction Processes

		Unit	APCI DMR (Base)	APCI F	Linde F	APCI D	Tech-nip D	Re-boiled D	Linde I	Exxon I
Power needed for LNG Liquefier	A	MW	116.0	115.9	115.9	116.8	124.6	115.2	116.0	118.3
Fuel compression Work	B	MW	7.0	7.7	8.8	9.0	0.0	7.8	5.0	5.0
Crude helium compression work	C	MW	0.0	0.0	0.1	0.0	0.0	0.1	0.0	0.0
Additional power needed	D	MW	0.0	0.0	0.0	0.0	5.7	0.0	0.0	0.0
Total power needed	E=A+B+ C+D	MW	123.0	123.6	124.8	125.9	130.3	123.1	121.0	123.3
GT efficiency	F		0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Needed fuel chemical energy	G=E/F	MW	410.1	412.0	416.0	419.5	434.4	410.2	403.3	410.8
Produced fuel chemical energy	H	MW	546.4	553.3	552.5	506.4	0.0	511.1	574.4	535.2
Net needed fuel chemical E	I=G-H	MW	-136.3	-141.3	-136.5	-86.9	434.4	-100.9	-171.1	-124.4
Produced LNG chemical E	J	MW	5379.5	5374.2	5373.2	5419.3	5923.0	5414.7	5351.4	5390.5
Net LNG chemical E	K=J-I	MW	5515.8	5515.5	5509.8	5506.2	5488.6	5515.6	5522.5	5514.9
LNG unit price	L	$/MBTU	9.46	9.46	9.46	9.46	9.46	9.46	9.46	9.46
LNG value	M*=K×L	$/s	49.5	49.5	49.4	49.4	49.2	49.5	49.5	49.5
		%**	100.0	100.0	99.9	99.8	99.5	100.0	100.1	100.0

*1MJ=948.45BTU, **100%=LNG value of the APCI DMR process

Table I-1 Economic Evaluation of Helium Extraction Processes (Continue) (F: Flashing, D: Distillation, I: Integration)

		Unit	APCI DMR (Base)	APCI F	Linde F	APCI D	Tech-nip D	Re-boiled D	Linde I	Exxon I
Produced crude helium volume flow	N	Sm^3/hr	0.0	469.9	529.5	500.1	349.6	530.4	500.1	500.1
Crude helium unit price	O	\$/ Sm^3	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5
Crude helium value	P=N×O	\$/s	0.0	0.5	0.5	0.5	0.3	0.5	0.5	0.5
		%**	0.0	0.9	1.0	1.0	0.7	1.0	1.0	1.0
Total value	Q=M+P	\$/s	49.5	49.9	50.0	49.9	49.6	50.0	50.0	50.0
		%**	100.0	100.9	100.9	100.8	100.2	101.0	101.1	101.0

Table I-1 Economic Evaluation of Helium Extraction Processes
(F: Flashing, D: Distillation, I: Integration)