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STEADY STATE AND DYNAMIC ANALYSIS OF AN LNG SUPPLY SYSTEM FOR SHIPPING INDUSTRY

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Resumo

O objetivo desta dissertação é a simulação em estado estacionário e em estado dinâmico de um sistema de abastecimento a gás natural liquefeito (LNG) para aplicação na indústria naval, utilizando a aplicação informática *Aspen HYSYS*®. A motivação para a realização deste trabalho passa pela necessidade de novas soluções de combustível e sistemas de abastecimento do mesmo face ao problema da poluição atmosférica. O sistema estudado inclui um tanque do qual saem duas correntes: uma corrente de *boil-off gas* (BOG) gasoso e uma corrente de LNG líquido que é bombeada para um permutador de calor para ser vaporizada. É este o combustível que é utilizado nos motores do navio, sendo que as condições de entrada deste são especificadas e devem ser controladas: pressão entre 6,7 e 9,0 bar e temperatura de cerca de 25 ºC.

Em primeiro lugar, foram efetuados testes para determinação das propriedades deste combustível: entalpia mássica, entalpia de vaporização, viscosidade do BOG e do LNG, massa volúmica do BOG e do LNG, temperaturas de ebulição e orvalho do LNG, coeficiente de expansão adiabático para o BOG e o LNG e por fiz o fator Z para o gás. Com estes testes, foi permitido definir que as pressões de armazenamento indicadas para este combustível seriam entre 1,25 e 5,0 bar. Foi então efetuada a simulação em estado estacionário e foram testados 18 casos de estudo variando a pressão do tanque entre os valores indicados anteriormente, o caudal aos motores (variando entre o caudal mínimo de 150 $kg \cdot h^{-1}$ e o caudal máximo de 1090 $kg \cdot h^{-1}$) e a utilização ou não da caldeira para queima de BOG. Foi possível concluir que, para todos os casos analisados, as condições de entrada de combustível nos motores cumpriam os requisitos de pressão e de temperatura estipulados.

De seguida, foi necessária a realização de um modelo dinâmico para simulação e controlo do variador eletrónico de frequência da bomba de LNG. Este modelo incluiu as curvas características da bomba e as respetivas curvas de eficiência para as diferentes velocidades de rotação. Foram implementados dois sistemas de controlo: PID e *split-range*. Para o PID foram estudados 6 cenários variando a pressão do tanque entre os limites apontados anteriormente e variando também o caudal aos motores. Para todos os casos testados e para os parâmetros determinados para o controlador (K_c de $7{,}7 \times 10^{-2}$ e T_i de $7{,}0 \times 10^{-2}$ s), a pressão de entrada aos motores encontra-se dentro do intervalo esperado. O modelo de controlo split-range inclui três controladores: um controlador de velocidade para a bomba, um controlador para o caudal ao vaporizador e também um controlador *split-range* que define os pontos estabelecidos para os anteriores. Este sistema foi implementado, não tendo sido testado devido a limitações do software.

Palavras-chave (tema): Simulação estacionária; Simulação dinâmica; *Aspen HYSYS*®; Indústria Naval; Gás Natural Liquefeito.

Abstract

The main goal of this dissertation is the steady state and dynamic simulation of a liquified natural gas (LNG) supply system to be applied in the shipping industry, using *Aspen HYSYS®*. The main motivation for this work is the need for new fuel solutions as well as new supply system's options to face the problem of air pollution. The studied system includes a tank, from which two streams come out: a boil-off gas (BOG) stream and a LNG stream, that is pumped into a heat exchanger to be vaporized. LNG is the fuel used in the ship's engines and the entry conditions of this fuel are specified and must be controlled: pressure between 6.7 and 9.0 bar and temperature of about 25 ºC.

Firstly, some tests were carried out to determine the properties of this fuel: mass enthalpy, enthalpy of vaporization, BOG and LNG viscosity, BOG and LNG volumetric mass density, LNG boiling and dew points, BOG and LNG adiabatic expansion coefficients and BOG Z factor. With these tests, it was possible to conclude that the indicated storage pressures for the fuel should be between 1.25 and 5.0 bar. The steady state simulations were then performed, and 18 scenarios were tested by varying the tank pressure between the previously stated values, the engine flow (ranging from a minimum flow of 150 $kg \cdot h^{-1}$ to a maximum flow of 1090 $kg \cdot h^{-1}$) and whether the BOG heater was used. It was concluded that, for all cases analysed, the fuel inlet conditions met the stipulated pressure and temperature requirements.

Next, it was necessary to create a dynamic model and control system of the variable frequency drive used in the LNG pump. This model included the pump characteristic curves as well as the efficiency curves for the different rotational speeds. Two control systems were implemented: PID and split-range. For the PID, 6 scenarios were studied, varying the tank pressure between the previously mentioned limits and varying the flow to the engines. For all tested cases and for the parameters set for the controller (K_c of 7.7×10^{-2} and T_i of 7.0×10^{-2} s), the inlet pressure to the engines is within the expected range. The split-range model includes three controllers: a pump speed controller, a vaporizer flow controller and a split-range controller that acts by providing the setpoints for the previous ones. This system was only implemented and not tested due to software limitations.

Keywords (theme): Steady state simulation; Dynamic simulation; *Aspen HYSYS®*; Shipping industry; Liquefied Natural Gas.

Declaration

I hereby declare, on my word of honour, that this work is original and that all non-original contributions were properly referenced with source identification.

Maria João Carvalho Santos

September 30th, 2019

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Notation and Glossary

Upper-case Roman

A	Area	m^2
BOR	Boil-off rate	$kg \cdot h^{-1}$
C_{Heater}	Capacity of the BOG Heater	kW
C_p	Specific heat capacity	$kJ \cdot kg^{-1} \cdot K^{-1}$
C_v	Thermal capacity at constant volume	$kJ \cdot kg^{-1} \cdot K^{-1}$
$C_{\mathrm{vaporizer}}$	Capacity of the LNG Vaporizer	kW
$Fouling_{\mathrm{plate}}$	Fouling on the plate side of a Heat Exchanger	$m^2 \cdot K \cdot W^{-1}$
H_{70}	Pump head at the speed if 70 Hz	m
H_{80}	Pump head at the speed if 80 Hz	m
H_{90}	Pump head at the speed if 90 Hz	m
H_{100}	Pump head at the speed if 100 Hz	m
H_{110}	Pump head at the speed if 110 Hz	m
H_{122}	Pump head at the speed if 122 Hz	m
H_{LNG}	Mass enthalpy of LNG	$kJ \cdot kg^{-1}$
$H_{\mathrm{O,A}}$	Mass enthalpy of the mixture of oxygen and argon	$kJ \cdot kg^{-1}$
H_{pump}	Pump head	m
K_{d}	Derivative gain	
K_{i}	Integrative gain	
K_{p}	Proportional gain	
M	Molar mass	$kg \cdot mol^{-1}$
$MARVS$	Maximum allowable relief valve setting	bar
MCR	Maximum continuous rating of the engine	kW
MIE	Minimum ignition energy at 25 ºC	mJ
OP	Manipulated variable	
P	Pressure	bar
$P_{\mathrm{dis,des}}$	Design pressure of the discharge side	bar
PC	Power consumption	kW
PC_{est}	Estimated power consumption	kW
P_{Feed}	Gas feed pressure	bar
$P_{\mathrm{in,des}}$	Inlet design pressure	bar
$P_{\mathrm{out,des}}$	Outlet design pressure	bar
$P_{\mathrm{plate,des}}$	Design pressure of the plate side	bar
P_{sat}	Saturation pressure	bar
$P_{\mathrm{shell,des}}$	Design pressure of the shell side	bar
$P_{\mathrm{suc,des}}$	Design pressure of the suction side	bar
P_{tank}	Pressure in the tank	bar
PV	Controlled variable	
Q	Flow rate	$l \cdot min^{-1}$
$\dot{Q}_{\mathrm{design}}$	Design flow rate	$m^3 \cdot h^{-1}$
R	Ideas gas constant	$J \cdot K^{-1} \cdot mol^{-1}$
T	Temperature	ºC
$T_{\mathrm{1\ atm}}$	Temperature at 1 atm	ºC
T_{AI}	Temperature of autoignition	ºC
T_{b}	Bubble point	ºC
T_{CIT}	Cold impact test	ºC
$T_{\mathrm{cold,in}}$	Inlet temperature of the cold fluid	ºC
$T_{\mathrm{cold,out}}$	Outlet temperature of the cold fluid	ºC
T_{d}	Dew point	ºC
T_{design}	Design temperature	ºC

T_{flame}	Flame temperature	ºC
T_{Feed}	Gas feed temperature	ºC
T_{flash}	Flash point	ºC
$T_{\text{hot,in}}$	Inlet temperature of the hot fluid	ºC
$T_{\text{hot,out}}$	Outlet temperature of the hot fluid	ºC
$T_{\text{plate,des}}$	Design temperature of the plate side	ºC
$T_{\text{pump,des}}$	Design temperature of the pump	ºC
$T_{\text{shell,des}}$	Design temperature of the shell side	ºC
$T_{\text{ST,des}}$	Design temperature of the storage tank	ºC
$T_{\text{valve,des}}$	Design temperature of the fluid passing the valve	ºC
U	Global heat transfer coefficient	$W{\cdot}m^{-2}{\cdot}K^{-1}$
V	Volume	m^3
V_{BT}	Volume of the buffer tank	m^3
W_{vessel}	Dry weight of the vessel	kg
Z	BOG Z factor	

Lower-case Roman

k_{BOG}	BOG adiabatic expansion coefficient	
k_{LNG}	LNG adiabatic expansion coefficient	
$k'_{\text{v,des}}$	Valve design constant	
$\dot{m}$	Mass flow	$kg{\cdot}h^{-1}$
$\dot{m}_{\text{cold}}$	Mass flow of the cold fluid	$kg{\cdot}h^{-1}$
$\dot{m}_{\text{engines}}$	Mass flow to the engines	$kg{\cdot}h^{-1}$
$\dot{m}_{\text{gas}}$	Gas consumption	$kg{\cdot}h^{-1}$
$\dot{m}_{\text{hot}}$	Mass flow of the hot fluid	$kg{\cdot}h^{-1}$
q	Heat transferred	kW
t	Time	s

Upper-case Greek

ΔH_{vap}	Enthalpy of vaporization for LNG	$kJ{\cdot}kg^{-1}$
ΔP_{des}	Design variation of pressure	bar
Δt	Time step size	s
ΔT	Difference of temperature	ºC
ΔT_{log}	Logarithmic mean temperature	ºC

Lower-case Greek

η_{70}	Hydraulic efficiency of the pump at the speed of 70 Hz	%
η_{80}	Hydraulic efficiency of the pump at the speed of 80 Hz	%
η_{90}	Hydraulic efficiency of the pump at the speed of 90 Hz	%
η_{100}	Hydraulic efficiency of the pump at the speed of 100 Hz	%
η_{110}	Hydraulic efficiency of the pump at the speed of 110 Hz	%
η_{122}	Hydraulic efficiency of the pump at the speed of 122 Hz	%
μ_{BOG}	BOG Viscosity	cP
μ_{LNG}	LNG Viscosity	cP
ρ	Volumetric mass density	$kg{\cdot}m^{-3}$
ρ_{-161^oC}	Volumetric mass density of LNG at -161 ºC and 1 atm	$kg{\cdot}m^{-3}$
ρ_{BOG}	BOG volumetric mass density	$kg{\cdot}m^{-3}$
$\rho_{\text{1atm,vap}}$	Volumetric mass density of LNG gaseous at 1 atm	$kg{\cdot}m^{-3}$

$\rho_{\mathrm{LNG,des}}$	Design volumetric mass density of LNG	kg·m^{-3}
ρ_{LNG}	LNG volumetric mass density	kg·m^{-3}
τ_{d}	Derivative time	s
τ_{i}	Integrative time	s
ω_{max}	Pump maximum speed	rpm

Indexes

k	Counter

List of Acronyms

ECA	Emission Control Area
EU	European Union
FGSS	Fuel Gas Supply System
GCS	Gas Control System
GSS	Gas Safety System
HFO	Heavy Fuel Oil
IAS	Integrated Automation System
IMO	International Maritime Organization
LNG	Liquefied Natural Gas
MARPOL	International Convention for the Prevention of Pollution from Ships
MDO	Marine Diesel Oil
MGO	Marine Gas Oil
NG	Natural Gas
PID	Proportional-integral-derivative control
PM	Particulate Matter
PTS	Port-to-Ship bunkering
RAM	Random Access Memory
SCR	Selective Catalytic Reduction
STS	Ship-to-Ship bunkering
TTS	Truck-to-Ship bunkering
VFD	Variable frequency drive

1 Introduction

1.1 Framing and presentation of the work

One of the main challenges that the shipping industry is facing nowadays is protecting the ocean and the environment. The reason for this is that the ship pollution constitutes about 3 % of the global air pollution [1, 2]. Other issue is the increasing maritime traffic (up to 4 % per year during the last decades [1]) that leads to an even more concerning volume of air pollutants' emissions [3].

Heavy Fuel Oil (HFO) is a fuel with high density and viscosity and that have large percentages of heavy molecules such as long-chain hydrocarbons and aromatics with long-branched side chains [4]. It is currently the most used fuel in the marine engines since it is a cost-effective option [3]. Despite being economic, these HFO are a big environmental problem: they contain high levels of carbon residues, sulphur, metallic compounds and asphalt. These characteristics lead to the production of significant amounts of pollutants such as nitrogen oxides (NO_x), carbon monoxide (CO), carbon dioxide (CO_2) and sulphur oxides (SO_x) and particulate matter (PM) when they are burned in the engines [1, 3].

The regulation of shipping emissions is carried out by the International Maritime Organization (IMO) and the limits on sulphur dioxide and nitrogen oxides emissions are set by the International Convention for the Prevention of Pollution from Ships (MARPOL), particularly in the Annex VI that the organization adopted [5, 6] . This convention also prohibits deliberate emissions of ozone-depleting compounds. Since there are areas more affected by these issues, both IMO and the European Union (EU) acknowledge protected some areas which are called Emission Control Areas (ECA) [3]. These zones have more strict control of the NO_x and SO_x emissions than the global waters and include the Baltic Sea, the North Sea, The North American ECA and the US Caribbean ECA [1, 5, 7].

Sulphur emissions are proportional to the content of sulphur compounds in the fuel, so the main solution to mitigate the emissions of SO_x is to decrease the content of sulphur in the fuel [8], making HFO no longer an alternative as a fuel [9]. Some low sulphur content fuels such as Marine Diesel Oil (MDO) and Marine Gas Oil (MGO) are already being used in ships worldwide, especially in ECA zones, besides being a more expensive option [3, 6].

The limits for the content of sulphur in the marine fuels are represented in Figure 1.1.

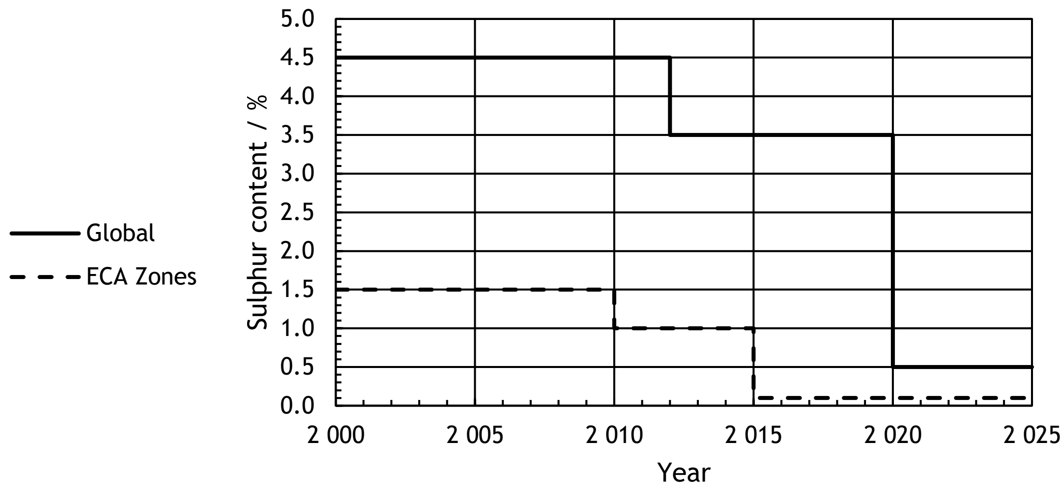

Figure 1.1 – Limits of SO_x emissions set up by MARPOL Annex VI [1, 7, 9].

NO_x emissions can be reduced with engine modifications such as exhaust gas recirculation, internal engine modifications, humid air motors, or with gas after-treatments such as Selective Catalytic Reduction (SCR) that can reduce the content of NO_x in more than 80 % [3].

Due to all these constraints, a possible and viable alternative would be Natural Gas (NG). Besides being used normally in the gaseous state on land, it's a better option to use liquified natural gas (LNG) in the maritime applications since it's been used for years as fuel in LNG carriers and for propulsion on traditional boiler or steam turbine systems. LNG has also been used in dual fuel diesel engines. Recently, the research has focused on using this fuel in ships other than LNG carriers and some ships are already operating on LNG [3, 6, 8]. LNG consists mainly of methane and is a better alternative because it has lower sulphur and carbon content and causes lower NO_x emissions than the traditional HFO [10, 11]. The disadvantages of this fuel are mainly the high initial investment cost and the lack of infrastructures for fuelling the ships [11]. However, the LNG itself has a competitive price in comparison to conventional fuels [3, 5, 8, 12].

This dissertation presents a simulation study of an LNG supply system for the shipping industry, namely a project system from Høglund Gas Solutions AS. The simulation was carried using *Aspen HYSYS®* software and taking advantage of its steady and dynamic state valences with and without control. Using a computational tool allows the study of several scenarios, reducing the time needed to test the operation and control of the system. Thus, the aim of the present project is to develop several simulations for studying the behaviour of a fuel supply system using LNG as fuel in a marine propulsion engine, in response of changes in system pressure or load changes in the gas consumers. It's also a goal of this work the development of a control system to assure the right conditions of the fuel delivery.

1.2 Presentation of the company

Høglund Gas Solutions AS (logo in Figure 1.2) is a member of Høglund Marine Solutions group, which is a well-established and well-reputed marine solution company, delivering high quality products and services to all types of marine installations worldwide.

Founded in 1991, this company delivers automation and power management systems, being their main areas:

- Supplying maritime automation products such as Integrated Automation systems (IAS), seismic control and fuel oil consumption calculations;
- Delivering power management control systems;
- Providing solutions for data logging, alarm/event logging and process control;
- Delivering customized competitive operator station interface;
- Consulting services within marine automation.

Høglund Gas Solutions AS was funded in 2017 and is a provider of gas handling systems to the maritime industry, specially using LNG as fuel and transportation or storage across the LNG supply chain.

(a) (b)

Figure 1.2 – Logo of Høglund Marine Solutions (a) and Høglund Gas Solutions AS (b).

1.3 Contributions of the Work

The work I have performed under this dissertation began with my interest in simulation under Project of Engineering classes of my masters' degree and the consequent opportunity to do my dissertation in this field. The first step of the work was an intensive research of the state of art and the literature to acquire the knowledge in the marine industry and the new solutions for the environmental challenges. Since *Aspen HYSYS®* is not the software used in classes during the degree (we use *Aspen Plus®*), I had to dedicate some time to learning to operate the software for performing the simulations.

Then, the simulation process began. The first step was the property study, and that was a request of the company that I performed. Then, I started by studying the system's provisional P&ID of the project provided by the company, and then I developed a model to simulate the system in the steady state, making the appropriate adaptations and assumptions. The next step was to determine the scenarios needed to be made, and that was also my decision: I tried to include all the situations and all the boundary cases in an expeditious set of scenarios. The system was then simulated, and the results analysed.

After, the dynamic simulation was performed, followed by the attempts of control system implemented. In this step, I had to make the decision to simplify the project, since the dynamic model was too complex and didn't stabilize. This decision was also made after an extensive review of literature to be sure I am in the correct way and the studies of dynamic simulations referred in literature using *Aspen HYSYS®* were much simpler than the system here featured.

Finally, the PID controller was idealized and implemented, and them the tests were performed by me. Similarly, to the steady state control, the scenarios tested were defined by me. The split-range controller, despite not being tested, was also designed and implemented by me with the help of the supervisors.

1.4 Organization of the thesis

This present dissertation is divided into 6 main Chapters: Introduction (Chapter 1), context and state of art (Chapter 2), materials and methods (Chapter 3), results and discussion (Chapter 4), conclusions (Chapter 5) and assessment of the work done (Chapter 6).

In Chapter 1, there's a brief exploration of the motivation to do an LNG supply system simulation and why is LNG a good solution for the environmental problems that marine industry faces nowadays. This subject is approached with more detail in Chapter 2, where other background needed is also introduced: how an LNG supply system works and the important equipment, the different types of simulation, control theory and different types of control.

Chapter 3 describes briefly the hardware and software used to perform this study. The system model, as well as the technical specifications of all the equipment used in the model is also featured in this chapter.

Chapter 4 contains the results of the work done in the simulation systems and the results' analysis and discussion. The first part of the chapter discusses the properties of LNG at saturation pressure, then it proceeds to the analysis of the steady state outline of experiences and results. The third part of the chapter describes the dynamic model and the results of the PID controller and implementation of a Split-Range solution. Finally, the reliability of a new solution of heating media system is analysed.

The conclusions of the work are presented in Chapter 5. A brief introspect of the objectives achieved, limitations and future work, and a final assessment of the work done is presented in Chapter 6.

2 Context and State of the art

2.1 Alternatives to HFO in the marine industry

Crude oil is a mixture of different types of hydrocarbons, and the crude oil refinery takes advantage of the different molecular weights, volatilities and temperatures to separate in different final products. The column has a temperature profile, being the higher temperature at the bottom. This means that the more volatile compounds evaporate and ascend in the distillation column, while the heavy compounds are left in the bottom part of the column [13]. This process is called fractional distillation and it is represented schematically in Figure 2.1.

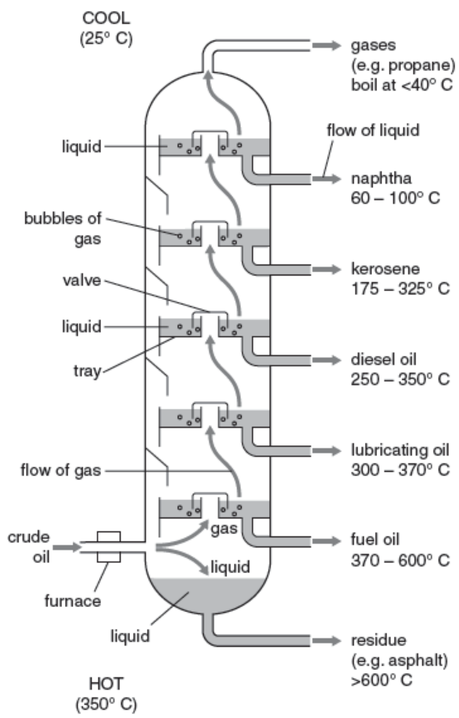

Figure 2.1 – Fractional distillation of crude oil. [13]

Around the world, HFO is used at large scare as main fuel in the marine industry. Fuel oil is a part of the base product in the fractional distillation, with boiling temperature of 600 ºC, meaning that it has a high content in heavy hydrocarbons and aromatics. This type of fuel is highly pollutant and IMO (with MARPOL convention) set new levels for legal emissions of pollutants in the marine industry such as NO_x and SO_x [5, 8]. With this new challenge, new solutions must be created to lower these pollutant emissions. The three main solutions found were the use of lower sulphur content fuels such as MGO, as discussed in Section 1.1, the use of scrubbers and SCR and the use of LNG as fuel [1, 14, 15].

2.1.1 Scrubber and SCR

A scrubber is an equipment that can remove the sulphur in the exhaust gases. However, this solution is not enough, since it doesn't remove the NO_x, so a combined SCR needs to be implemented. This technology allows the shipping companies to continue using HFO as fuel [10].

There are two types of scrubber. The wet scrubber, as seen in Figure 2.2, uses alkaline water to spray the exhaust gases. The sulphur oxides react with the water, forming sulphur acid, which is neutralized with the alkaline compounds in the water. The sulphur is retained in the water, that is treated afterwards in the proper unit, and the exhaust gases are emitted with a much lower sulphur content. There are three types of circuit: open, closed and hybrid. The open circuit uses as alkaline water sea water, that, after treatment, is returned to the ocean. In the closed circuit, the water is mixed in the ship with an alkaline agent, mainly caustic soda, to adjust de pH. The hybrid circuit allows the ship to operate in an open loop and switch to closed loop if the conditions are ideal. The water is treated afterwards, being taken to a settling tank where the particles settle. A cyclone then purifies the water using the centrifugal force, and the particles are then separated to a waste tank [15-19].

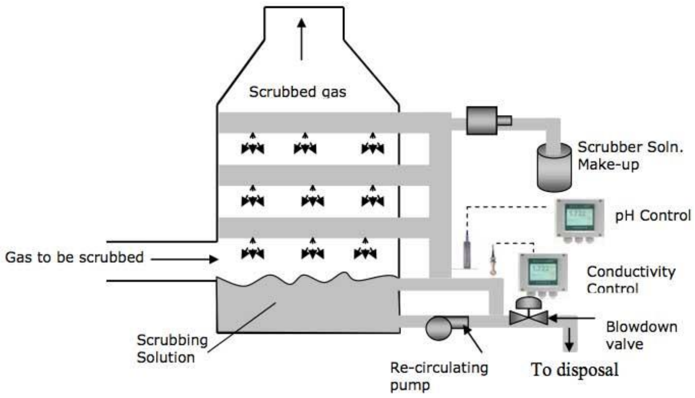

Figure 2.2 – Wet open scrubber system [20].

The dry scrubber uses a salt (normally hydrated lime, soda ash or sodium bicarbonate) to remove the sulphur from the exhaust gases. The high temperature of the gases (between 240 ºC and 400 ºC) allows the salt to absorb the sulphur, forming calcium sulphate. The dry system is more energy efficient, since it doesn't need water pumps, but, on the other hand the equipment is heavier and needs more storage capacity for the reagents and the reaction products [10].

The SCR consists in injecting ammonia or urea (reducing agents) in the exhaust gases. Then, these gases go through a catalytic reactor with temperature between 250 ºC and 500 ºC. From the reaction of NOx and the reducing agent, water and diatomic nitrogen is formed [10, 18].

The combination of these two methods allow the emission of exhaust gases with pollutant contents below the legal limits [14]. Figure 2.3 represents the normal operating configuration of a combined solution of a scrubber and SCR.

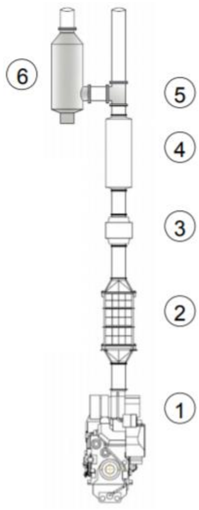

Figure 2.3 – Combined system with a scrubber and SCR. (1) Diesel engine; (2) SCR; (3) Exhaust gas boiler; (4) Silencer; (5) By-pass valve; (6) Scrubber unit [17].

2.1.2 Liquefied Natural Gas

The use of LNG as fuel in the marine industry is currently one of the most studied options to face the challenges of MARPOL. LNG is a mixture of hydrocarbons, mainly methane (and smaller fractions of ethane, propane and nitrogen). LNG vapour is flammable within specific concentration limits, and since it consists of mainly methane, the flammability limits of methane are used as an estimate (4.5 – 16.5 vol.%) [10, 21, 22]. The properties of LNG are listed in Table 2.1.

Table 2.1 – LNG properties [22]

LNG properties	
Physical state	Liquid
$T_{1\,\mathrm{atm}}$ (°C)	-161
$\rho_{-161^o\mathrm{C}}$ (kg·m^{-3})	448
$\rho_{\mathrm{1atm,vap}}$ (kg·m^{-3})	0.55
T_{flash} (°C)	-1.75
T_{AI} (°C)	540
MIE (mJ)	0.29
T_{flame} (°C)	1875

The use of LNG as ship fuel can reduce the sulphur emissions by 90 %, which makes the ships emissions within the legal limits set by IMO. LNG has a lower carbon content, which enables reductions of 20 % of carbon dioxide emissions. Also, LNG is less costly than MGO, which is the option to be applied if no other technical and operational measures are implemented to reduce sulphur emissions. LNG has a price comparable to HFO, being this a great cost advantage comparing with the other options [14, 21].

The bunkering of LNG to ships can be executed in three different ways: truck-to-ship (TTS), ship-to-ship (STS) and port-to-ship (PTS). TTS consists in LNG delivering to the receiving vessels by another ship or boat and it's mainly used for high volume bunkering. It can be performed at anchor or at station and requires efficient safety measures in the mooring. STS is ideal for low volumes since the trucks only can carry volumes between 40 and 80 m^3. The PTS option has logistic limitations since ports don't have, normally, a fixed point for ship bunkering. On the other hand, it's a good option because it can be used for high volume bunkering [10, 22-24]. The different types of bunkering are illustrated in Figure 2.4.

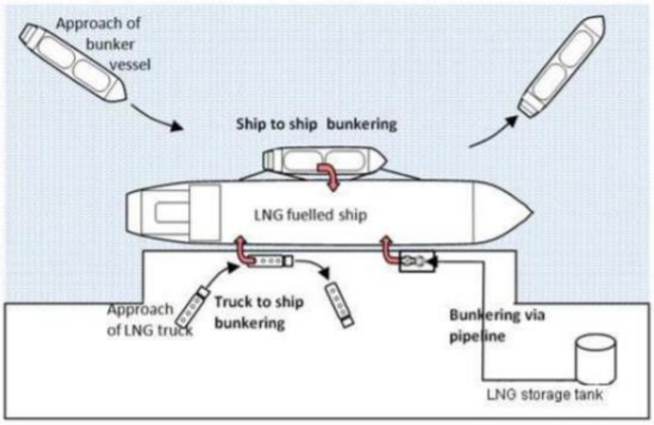

Figure 2.4 – Types of LNG Bunkering. [23]

Section 2.2 includes a scope of the typical types of LNG supply systems with applicability in the marine industry, since the system used in this work uses LNG as fuel.

2.2 LNG supply systems

After bunkering, LNG is taken to a tank directly through a single wall pipe if it's placed on deck or through a double wall pipe if it's placed under the deck. There are three types of tanks used in LNG fuelled ships: the IMO A, B or C. The type A IMO tank is projected as part of the ship, taking its shape, while IMO B tank has a prismatic or spherical shape. Both tanks work at atmospheric pressure, with its maximum being 1.7 bar. The IMO type C is a pressurized tank with capacity up to 3500 m^{3,} allowing high bunkering rates. Because of the conditions of pressure and temperature in the vessel, there are inevitable evaporation of LNG forming boil-off-gas (BOG). All the current applications of LNG as fuel in ships use IMO C tanks to store LNG [10, 25].

Since the storage of LNG must be done at cryogenic temperatures and can damage the ship structure, these tanks must be isolated. The second reason for this isolation is to avoid high rates of BOG formation. The isolation can be done either with isolating foams or with vacuum isolation (meaning that the tank has a double wall with vacuum in the middle) [22]. Then, the gas must enter the engines at gaseous state and at the right P and T. There are two types of engine: lower pressure engines, which have an inlet pressure of around 6.5 bar, and the high-pressure engines with inlet pressures of around 300 bar. Both engines work at temperatures between 0 and 60 ºC. To achieve these conditions, an evaporator must be placed after the tank. This equipment has the goal of heating and evaporate the liquid NG, bringing it from around -160 ºC to ambient temperature [10, 24]. Figure 2.5 represents a generic handling system for LNG in an LNG fuelled ship.

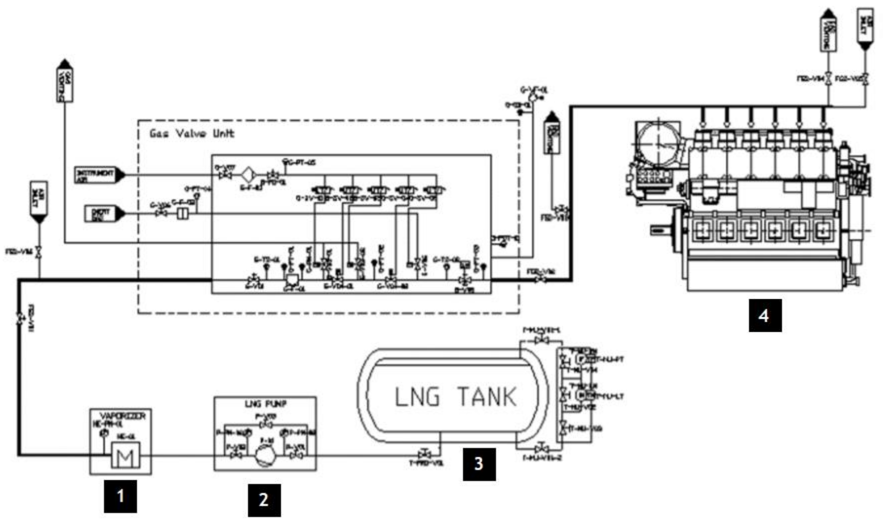

Figure 2.5 – Typical LNG supply system. (1) LNG vaporizer; (2) LNG pump; (3) LNG Tank; (4) Engine [26].

There is the need of instrumentation (valves, automatic control) that depend on the different applications, from ship to ship. These systems must be operated with maximum caution and with strict safety measures since there are many consequences of accidents with LNG fueled ships. A leak in the tank can cause a fracture in the ship structure due to the cryogenic temperatures and causing also injuries on the ship occupants. Another issue is that LNG can be suffocating if the proper ventilation is not assured. There can also happen a fast phase transition, if the LNG contacts with water at higher temperature, generating a shock wave like an explosion without combustion [8, 22, 27, 28]. For these reasons, all ship operators must receive safety training as well as training for operating and handling the gas [29].

2.3 Process Simulation

The operations in industry facilities or in processes in general can be simulated using computers and suitable software to imitate the real-world. The process of interest is called system and, to study these systems, a set of assumptions must be made (for example mathematical relationships to constitute a model that represents the behaviour of the system). If the correlations are simple, for example using algebra and calculus, it is possible to obtain an analytic solution. For other cases, that are most of the real-world systems, there's the need for simulations because of the complexity of the systems. In simulation, software is used to analyse the system numerically and to estimate the characteristics of the model [30-33].

There are several applications for simulations in the industry, being one of the most used techniques in research and management. First, simulation is helpful to design and analyse manufacturing systems in the production industries such as chemical industries. But simulation can be used in other areas such as information technology (for evaluating hardware and software requirements for a given system), economic areas (to evaluate financial and economic state) and even in the logistics area (to study inventories and transportation). Other advantages of simulation study are the opportunity of study systems that don't exist in the real world and access their viability and the opportunity of study long time frame systems in a short period of time (for example, the impact of a variable change in the years to come) [31, 34, 35].

There are, on the other hand, some setbacks on simulation use. Models for large-scale systems are, normally, very complex making arduous the task of creating programs to simulate them. However, nowadays we have programs that automatically provide many features needed in the models (such as *Aspen HYSYS*® and Aspen Plus® for the chemical engineering area [36]). Simulations can be also expensive and time-consuming to develop, since software is expensive, and the model design can take a lot of time until it represents the system in study correctly [32].

In order to build and analyse a simulation in the most correct way, Figure 2.6 shows some indicative steps to follow. Obviously, not all simulations need to follow these steps, but it's a simple way to start. The first step is the formulation of the problem and the planification of the study, meaning that the first step must be the outline of the objectives and issues and the design of the system alternatives. Following this step, one must collect data and/to define the model itself and to define operating procedures and variables. Then, it is necessary to analyse if the model is valid or not, for example, by getting in touch with people that have more insight in the operation system or testing the system for other states already known. The computer model is then ready to be build and verified, using whichever language is suitable (or even a software more targeted to the field of studies) and pilot runs are done to verify the model. If the model is valid, then the experiments must be designed: the experiences needed, the initial conditions, number of independent simulations and other variables. Then, the results of the experiments must be analysed with statistical techniques and documented later. The goal of a simulation is to be useful in the real-world system, so the solutions must be implemented [30, 34, 37].

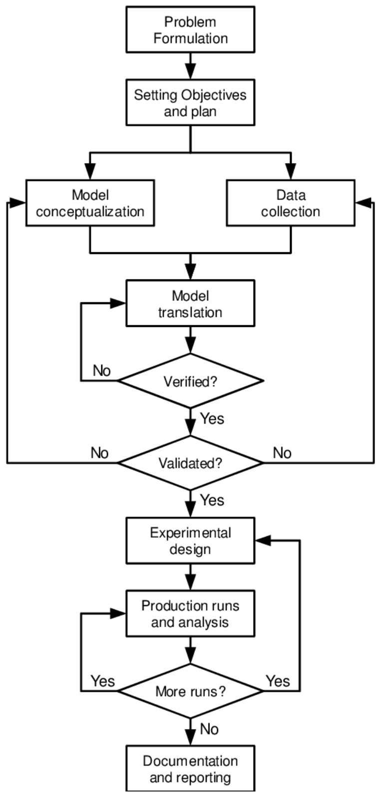

Figure 2.6 – The steps of a simulation study [30].

In the case of chemical engineering modeling, there are some equations and balances that are essential. First, mass and energy balance equations are needed. Then, other relations can be added such as kinetic equations, rates of heat and mass transfer, system property changes, phase equilibrium and control. There are also a set of steps for these simulations in particular: first, there's the need of selecting the chemical components that exist on the system and the thermodynamic model required to simulate the behavior of all the species in study. Then, the typology of the flowsheet must be designed, followed by the input of properties of the streams and the specifications of the equipment. Finally, the convergence method can be selected and the simulation run [36].

Steady state vs Dynamic mode simulations

A steady state model does not consider the time passing, being a representation of a system in a particular time (or the simulation represent a system in which time doesn't play a role). This means that this system has no time derivatives and that the mass balance and energy balance are just equations that equal what goes in and out of the system [33].

The dynamic model establishes relationships between variables as they change with time, commonly called dynamic characteristics [37]. Unlike traditional steady state simulations, dynamic simulations for chemical engineering analysis are much more recent, thanks to the development of new powerful software packages [34]. In order to build a dynamic model, just like steady state models, one of the basic principles is the conservation of mass or matter. In real situations, the simple equation used in as mass balance in steady state is not applicable because conditions change with respect of time. In that case (in the case of a dynamic study), Equation (2.1) is used [31, 33-37].

$$\begin{pmatrix}\text{Rate of accumulation of}\\ \text{mass in the system}\end{pmatrix} = \begin{pmatrix}\text{Rate of}\\ \text{mass flow in}\end{pmatrix} - \begin{pmatrix}\text{Rate of}\\ \text{mass flow out}\end{pmatrix} \qquad (2.1)$$

As can be concluded, at the steady state the accumulation factor is zero, so, the equation simplifies to the one showed before in this section. If the system has more than one chemical species, there’s the need to perform a mass balance individually to each one of them as seen in Equation (2.2) [31, 34].

$$\begin{pmatrix}\text{Rate of accumulation}\\ \text{of mass of component}\\ \text{in the system}\end{pmatrix} = \begin{pmatrix}\text{Mass flow of}\\ \text{the component}\\ \text{into the system}\end{pmatrix} - \begin{pmatrix}\text{Mass flow of}\\ \text{the component}\\ \text{out of the system}\end{pmatrix} \qquad (2.2)$$

However, if a reaction takes place, a reaction rate term is added to the balance equation, (Equation (2.3)). The mass balance equations can be applied to different systems, and each system has a different mathematical expression for each parcel. There’s the need to study the dynamics of the systems individually [31, 34].

$$\begin{pmatrix}\text{Rate of accumulation}\\ \text{of mass of component}\\ \text{in the system}\end{pmatrix} = \begin{pmatrix}\text{Mass flow of}\\ \text{the component}\\ \text{into the system}\end{pmatrix} - \begin{pmatrix}\text{Mass flow of}\\ \text{the component}\\ \text{out of the system}\end{pmatrix} + \begin{pmatrix}\text{Rate of production}\\ \text{of the component}\\ \text{by the reaction}\end{pmatrix} \qquad (2.3)$$

The second important balance is the energy balance. They are needed whenever temperature changes, for example if a reaction causes a change in the temperature of the system. The energy balance is made using the same guidelines from material balance, as enunciated in Equation (2.4). The equation represents a balance for an open system with energy exchange across boundaries [31, 33-35, 37, 38].

$$\begin{pmatrix}\text{Rate of}\\ \text{accumulation}\\ \text{of energy}\end{pmatrix} = \begin{pmatrix}\text{Rate of}\\ \text{energy input}\\ \text{due to flow}\end{pmatrix} - \begin{pmatrix}\text{Rate of}\\ \text{energy output}\\ \text{due to flow}\end{pmatrix} + \begin{pmatrix}\text{Rate of}\\ \text{energy input}\\ \text{due to transfer}\end{pmatrix} - \begin{pmatrix}\text{Rate of work}\\ \text{done by the system}\\ \text{on the surroundings}\end{pmatrix} \qquad (2.4)$$

There are other important correlations such as momentum balances and kinetic equations, but they are more individualized for the systems in study.

2.4 Process control systems

Nowadays, industrial processes are predominantly continuous and have high performance specifications, being important to minimize the impact that disturbances have in the system. With this challenge in mind, automatic control was implemented in chemical processes, bringing operation costs to a minimum, since the production became more efficient. Chemical processes are inherently dynamic, meaning that the variables are changing with time, creating the need for measuring, monitoring and inducing change in the process variables of interest. A control system is needed to keep variables within their operating ranges, keeping the stability of the process and optimizing the system [38, 39].

There are different types of variables associated with a process: input and output variables.

Input variables are the ones that independently stimulate the system and can induce change in the internal conditions. One type of input variable are the manipulated (or control) variables, that are the variables adjusted by some control mechanism (manual or automatic). Other type of input variables are the disturbance variables, in which values are not adjusted or cannot be decided at will [38].

Output variables are those that can give information about the internal state of the process. Measured and non-measured variables are output variables, and the designation depend or whether the variable is available to be directly measured or not. Figure 2.7 schematizes all the designations for the variables aforementioned [39-41].

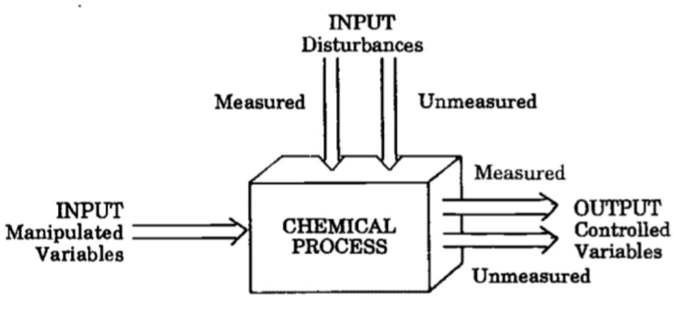

Figure 2.7 – The different types of process variables.

To keep these variables at proper ranges, different control systems can be implemented, being feedback and feedforward control modes the most common. The control system compares the information of the variables transmitted to them with the target value (set point) for the same variable and then activates an action in the system in order to keep the variables in the proper range.

The most popular system is the feedback control, also designated by feedback control loop, as represented in Figure 2.8. This control system works by giving process output information to the controller, being decisions made based on information that leaves the system. The advantage of this system is that it's simple and compensates for all the disturbances. Not needing to know the source of the disturbances, it only tries to maintain the control variable at the set point. The disadvantage is that it can only compensate a disturbance once the controlled variable already had deviated from the set point [38, 40].

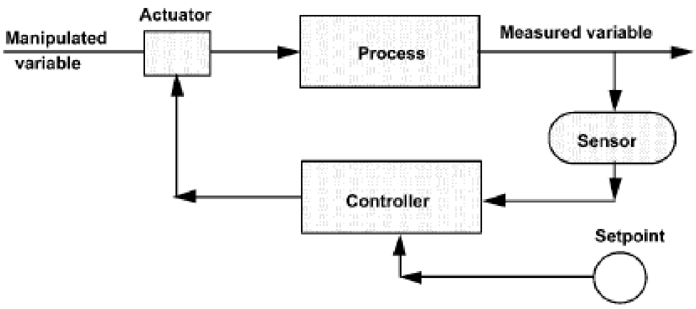

Figure 2.8 – Feedback control configuration [34].

Feedforward control is also very common, and it consists in a situation where the measured variable is the disturbance variable and not the output variable. This has the advantage of the decision of the controller being taken before the process is affected by the disturbance. Figure 2.9 represents a scheme of this method [41].

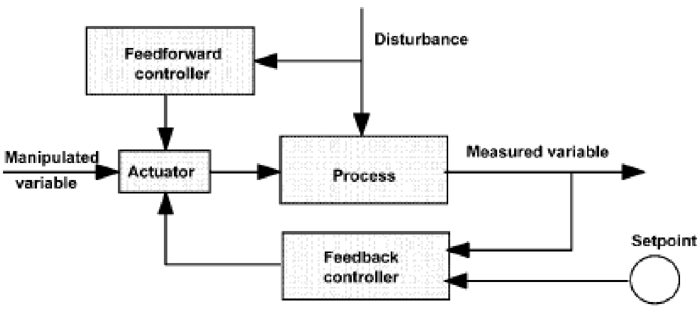

Figure 2.9 - Feedforward control configuration [34].

In the system studied in the present work, two types of controller were used: PID controller and split range controller.

2.4.1 PID controller

In feedback control, the goal is to reduce the error signal to zero. Equation (2.5) represents the mathematical expression used to calculate the error.

$$e(t) = y_{\mathrm{sp}}(t) - y_{\mathrm{m}}(t) \tag{2.5}$$

Where $e(t)$ is the error signal, $y_{\mathrm{sp}}(t)$ is the set point and $y_{\mathrm{m}}(t)$ is the measured value of the controlled variable (or equivalent signal). The set point, while indicated to be time-varying, in many systems it is kept constant for most of the time [42].

The proportional-integral-derivative (PID) control is the most popular feedback controller used in process industries. It's a robust model and it's easy to understand and implement, managing to adapt at most dynamic process plants. This type of control involves three types of actions, with different parameters and behaviour. The proportional action (P) depends on the present error, the integrative (I) action depends on the accumulation of the past errors and the derivative action (D) is the parcel that predicts future errors because it is based on the rate of change. These actions can be combined in many ways, depending on the needs of the system, being the more used controllers based on this method the P, the PI and the PID controllers [40-44].

The traditional form of the expression that models the PID control method is Equation (2.6) [40].

$$u(t) = K_{\mathrm{p}}e(t) + K_i \int_0^t e(t)\,dt + K_d \frac{de(t)}{dt} = K_{\mathrm{p}} \times \left[e(t) + \frac{1}{T_{\mathrm{i}}} \int_0^t e(t)\,dt + T_{\mathrm{d}} \frac{de(t)}{dt} \right] \tag{2.6}$$

Where $u(t)$ is the control signal, K_{p} is the proportional gain, K_i is the integrative gain, K_d is the derivative gain, T_{i} is the integral time and T_{d} is the derivative time [40]. However, the most used model in simulations of the PID control is the velocity form of the digital PID, in Equation (2.7) [42]. This is the form that *Aspen HYSYS®* uses.

$$u(t) = u(t-1) + K_{\mathrm{p}} \left[(e(t) - e(t-1)) + \frac{\Delta t}{T_{\mathrm{i}}} \times e(t) + \frac{T_{\mathrm{d}}}{\Delta t} \times (e(t) - 2 \times e(t-1) + e(t-2)) \right] \tag{2.7}$$

Where $u(t-k)$ is the value of the control signal k sampling periods before, $e(t-k)$ is the value of the error signal k sampling periods before and Δt is the sampling period [42].

2.4.2 Split-range controller

A split range controller is used when there is just measured variable but there are more than one manipulated variable. This is used in a lot of cases, being one example the pressure control in a tank, where the pressure can be manipulated either by the inlet or the outlet valve. If the pressure rises, the inlet valve will close. In the case of the inlet valve is completely closed but the pressure is still rising, the outlet valve can open [45].

Another example is the temperature control in a tank, where the control elements are the inlet vapor valve (heating valve) and the inlet cold water valve (cooling valve). The situation is really similar to the one exemplified before: if the temperature increases, the heating valve will close to decrease the temperature to a more proper value. If the valve is closed and the temperature is still high, the cooling valve can open to control the temperature [45, 46].

There's a need for a split of the controllers' range, as said by the name of this method. In the case of Figure 2.10, an example of a split-range application is illustrated. The output (from 3 to 9 psig) is given to both valves in parallel. From 3 to 9 psig, only the cooling valve has a response as it closes. From 9 to 15 psig, only the heating valve has a response as it opens. This shows that the cooling valve has an operating split-range of 0-50 % and the heating valve has an operating split-range of 50-100 % [46].

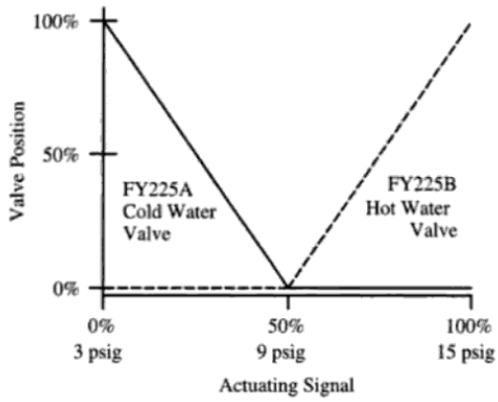

Figure 2.10 – Controller output and valve position for a split-range controller [45].

The split doesn't need to be done in a 0-50 % and 50-100 % way. There's always a need to adapt this method to the system in use, since valves (or other control elements) can be different in their dynamic and action. For example, if the flow rate that goes through each valve is the same, but the temperature differential is higher in the heating fluid, then the process gain is much greater too. In that case, the split should be 0-30 % for the cooling valve and 30-70 % for the heating valve, for example. Another split point can be chosen experimentally or with the aid of simulation programs.

There are also advantages in overlapping the ranges: the switch between the two control elements are much smoother and there are no dead band between operation of the valves. The disadvantage is the additional cost of utilities (or other costs, for another split-range control applications) [45, 46].

3 Materials and Methods

3.1 Hardware used

The hardware used to perform the simulations in *Aspen HYSYS®* was a personal computer HP, model ENVY 13-ad065nr. The processor of this computer is an Intel® Core™ i5-7200U and it has 8 GB of RAM Memory (8 GB LPDDR3-1866 SDRAM). This computer also has an integrated graphics card which is an Intel® HD Graphics 620. The operating system used was Windows 10. To use all the software needed (namely Aspen Tech's programs) it was necessary to access through a VPN connection. This VPN connection allowed to enter a private server at FEUP (salasapps.fe.up.pt). The remote desktop used had a processor Inter Xeon CPU E5-2650 v2 and 64 GB of RAM Memory.

3.2 Aspen HYSYS®

Aspen HYSYS® is a software of the *Aspentech®* company (logo is represented in Figure 3.1) and it is a chemical process simulator used to model chemical processes. It's the leading simulator software for oil and gas, refining and engineering processes [47]. It performs core calculations automatically, such as mass and energy balance, vapor-liquid equilibrium, heat and mass transfer, chemical kinetics and pressure drop. It provides information about physical, chemical and thermodynamic properties, process equipment, equipment sizing and capital and revolving investment. It is extensively used in industry to perform steady state and dynamic simulations, process design, performance modelling and optimization [47-49].

Figure 3.1 - Aspentech®'s logo [49].

In the present work, *Aspen HYSYS®* v9 was used to perform the calculations and simulations of the LNG supply system. Section 3.3 explores the architecture and equipment of the LNG supply system.

3.3 LNG supply system studied in Aspen HYSYS®

The goal of this work is to study the behaviour of a fuel supply system using LNG as fuel in a marine propulsion engine, in response to the load changes in gas consumer.

The Fuel Gas Supply System (following FGSS) in study in this thesis is intended to be installed on board of a marine vessel to use LNG as fuel. The LNG is stored at cryogenic temperatures (around -162 ºC), to be evaporated. Then, the fuel is stored in a buffer tank on board to be used as fuel. This storage shall be done at such conditions that the gas is delivered at a desired temperature and flow. The system can be summarized as in Figure 3.2.

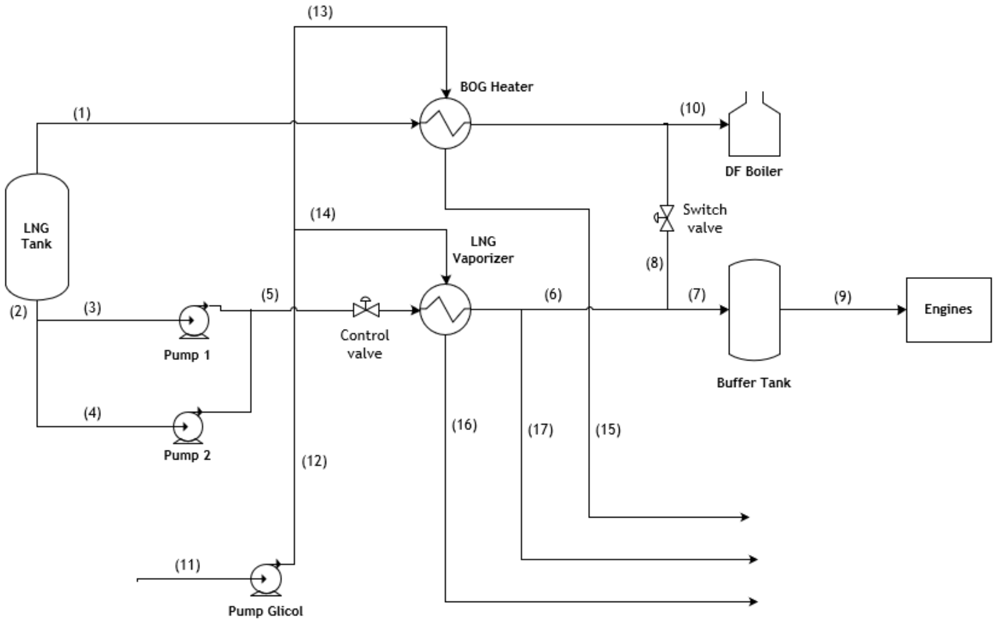

Figure 3.2 - Simplified scheme of the simulated system.

There's a tank with capacity of 440 m^3 and 95 % vol. of liquid inside. From this tank come out a liquid stream (2) and a gas stream (1). The liquid stream (2) goes into Pump 1 to elevate its pressure before entering the LNG Vaporizer. Pump 2 is a backup pump, only used when needed, so stream number (4) only exists in that case. After the vaporizer, the stream (6) is now in the gaseous state and at 25 ºC, being redirected to the buffer tank before entering the two engines (stream (9)). The gas stream (1) goes into a BOG Heater, where the temperature is raised up to, approximately, 25 ºC before entering the boiler (stream (10)). The exchanging fluid used is a mixture of ethylene glycol and water at 45 ºC, represented by the stream (11). This stream is pumped to the BOG Heater and the LNG Vaporizer, but the heating system used to set the temperature of this fluid is not part of this study. The switch valve is used to define the origin of the gas for the boiler. If the pressure is lower or equal to 3 bar, the gas is fed by the LNG Vaporizer, if the pressure is higher than 3 bar, the gas is fed by the BOG Heater. Stream (17) is the recycle stream, used when the flow that needs to be led to the engines are lower than the minimum flow of the pump. The control valve is used in the dynamic state to adjust the flow that goes to the engines. The philosophy of control in which this valve is used is specified in the Section 3.4.2..

This system was then simulated in the software *Aspen HYSYS®*, being the final simulations for the stationary study and dynamic study represented in Figure 3.3 and Figure 3.4, respectively.

Appendix A describes in detail the streams in both flowsheets. The simulation was carried out using Peng-Robson as the thermodynamic package, since it's has a robust database and equations, generating better predictions of equilibrium for hydrocarbon systems [50].

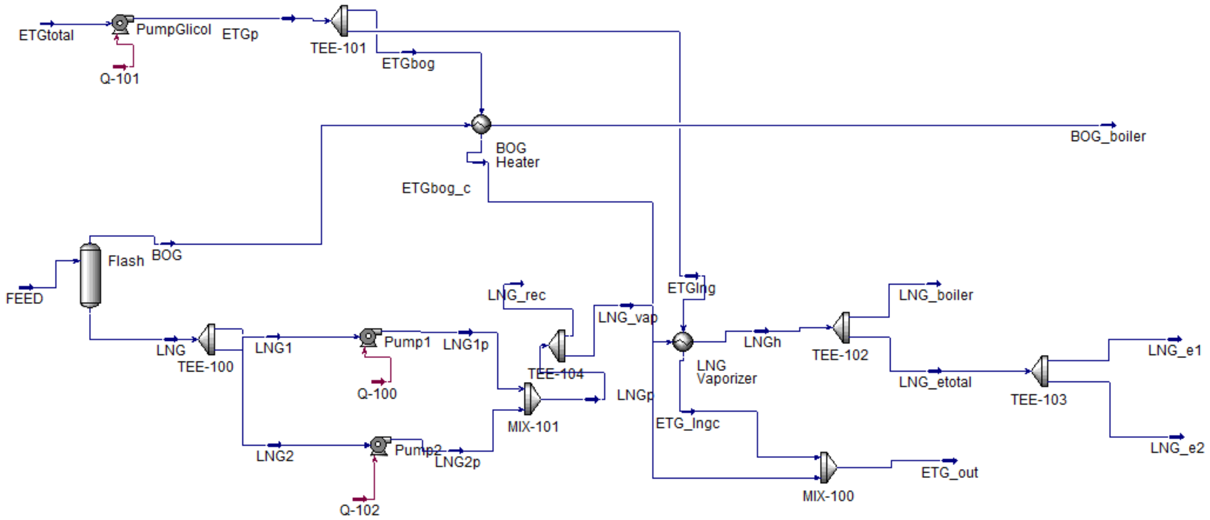

Figure 3.3 - Aspen HYSYS® simulation used to perform the stationary case studies.

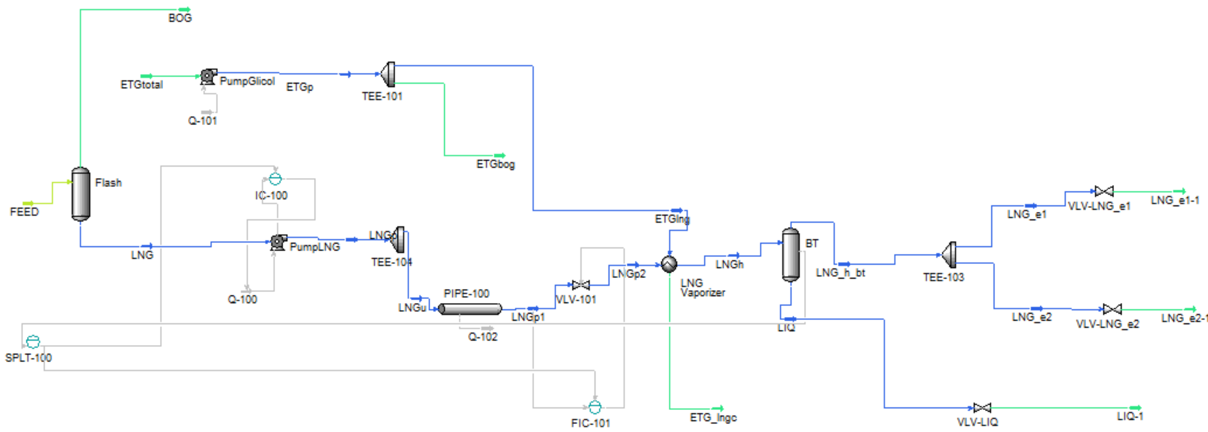

Figure 3.4 - Aspen HYSYS® simulation used to perform the dynamic analysis.

In the dynamic model, there was a need to simplify the simulation. Therefore, the handling of the BOG and the stand-by pump were supressed. On the other hand, the piping system was included to calculate the pressure drop in the system where the LNG is transported from the tank to the heater. It is included in the next sections the design properties of the main equipment used in the system above and the tools used in the software to simulate the different equipment.

3.3.1 LNG Storage Tank

The tank used in the simulation is a horizontal IMO type-C tank with a capacity of 440 m^3. The design parameters of this equipment are listed in the Table 3.1.

Table 3.1 – LNG Storage Tank design data

Property	Design value or designation
Fluid	LNG
Tank arrangement	Horizontal
Location	Below deck
$MARVS$ (bar)	6.0
$T_{\mathrm{ST,des}}$ (°C)	-165 to 45
T_{CIT} (°C)	-196
$\rho_{\mathrm{LNG,des}}$ (kg·m^{-3})	500
Material of the Inner Vessel	Stainless steel (SS 304)
BOR (kg·h^{-1})	66

This tank shall be monitored for vapor pressure, temperature and level. The level measurement shall be done for level monitoring with an acoustic radar and for the overfill protection with an ultrasonic level switch.

In the simulation model, a flash unit was used to simulate the LNG tank present in the handling system (Flash), being the origin of the LNG stream for the engines or the BOG stream for the boiler. It was considered a vapor fraction of 0.053, taking in consideration that the liquid level inside the tank should be approximately 95 %. The composition of both streams is contemplated in Table 3.2.

Table 3.2 – Composition of the streams of LNG and BOG

Compound	LNG Composition (wt.%)	BOG Composition (wt.%)
Methane	93.8	97.3
Ethane	4.96	0.03
Propane	0.86	-
Butane	0.21	-
Nitrogen	0.17	2.67

3.3.2 Fuel pumps

There are two pumps inside the tank to send the LNG to the vaporizer. These pumps are installed in the bottom of the tank and are completely submerged in LNG. The pumps are to be started, stopped or ramped up or down automatically, using the gas control system (GCS). They may also be started or stopped manually or by the gas safety system (GSS) in case of need. The design properties of the pumps are listed in Table 3.3.

Table 3.3 – LNG Pump design data

Property	Design value or designation
Type	Submerged centrifugal
$P_{\mathrm{suc,des}}$ (bar)	5.0
$P_{\mathrm{dis,des}}$ (bar)	16.0
$T_{\mathrm{pump,des}}$ (°C)	-165 to 45
Material	Stainless steel (SS 304)
$\dot{Q}_{\mathrm{design}}$ ($m^3 \cdot h^{-1}$)	3.1
H_{pump} (m)	240
ω_{max} (rpm)	6900
PC_{est} (kW)	3.7
W_{vessel} (kg)	50
Starting method	Variable frequency drive

In the simulation model for the steady state, since the pumps used are variable frequency pumps, it was necessary to adjust the pump head curves, and, after that, determine what curve led to the adequate pressure raise (taking into consideration that the pressure entering the engines should be as close as possible to 6.7 bar). Figure 3.5 shows the pump head curves and Equations (3.1) to (3.6) present the respective linear equations used in *Aspen HYSYS®*.

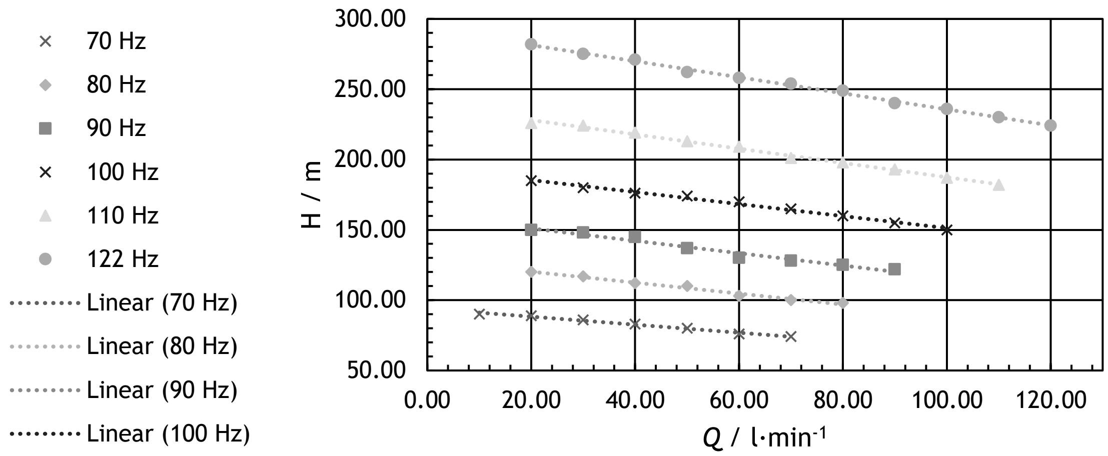

Figure 3.5 – Pump head curves for different frequencies.

$$H_{70} = -2.86 \times 10^{-1} Q + 9.40 \times 10^{1} \tag{3.1}$$

$$H_{80} = -3.89 \times 10^{-1} Q + 1.28 \times 10^{2} \tag{3.2}$$

$$H_{90} = -4.39 \times 10^{-1} Q + 1.60 \times 10^{2} \tag{3.3}$$

$$H_{100} = -4{,}27 \times 10^{-1} Q + 194 \times 10^{2} \tag{3.4}$$

$$H_{110} = -5.08 \times 10^{-1} Q + 2.38 \times 10^{2} \tag{3.5}$$

$$H_{122} = -5.71 \times 10^{-1} Q + 2.93 \times 10^{2} \tag{3.6}$$

It was also necessary to determine the adiabatic efficiency of the pump for each case, considering the flow rates in study. Figure 3.6 represents the pump efficiency curves, and Equation (3.7) to (3.12) are the result adjusted equations implemented in HYSYS®.

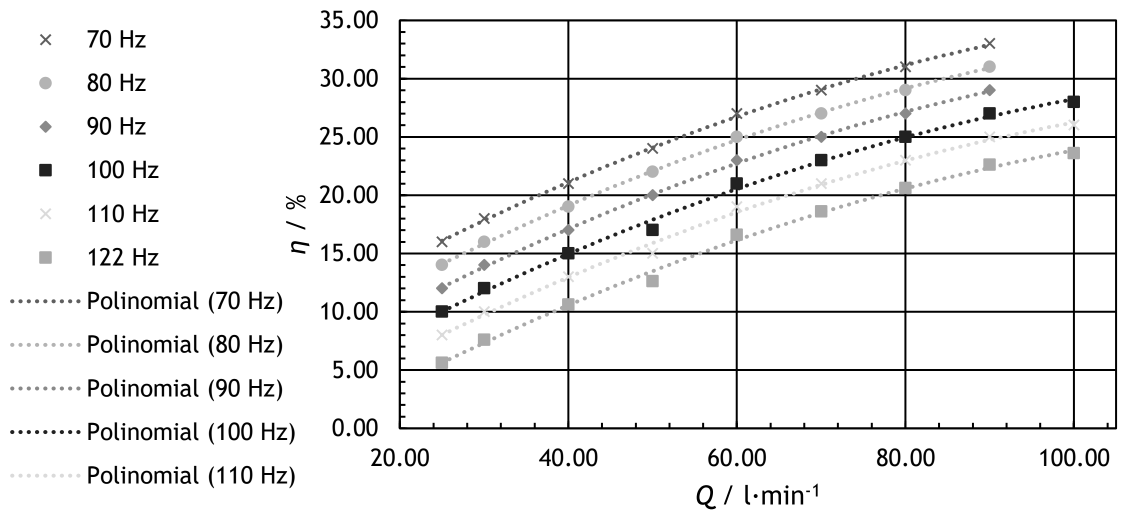

Figure 3.6 - Pump efficiency curves for different frequencies.

$$\eta_{70} = -1.52 \times 10^{-3} Q^2 + 4.33 \times 10^{-2} \times Q + 6.21 \quad (3.7)$$

$$\eta_{80} = -1.52 \times 10^{-3} Q^2 + 4.33 \times 10^{-2} \times Q + 4.21 \quad (3.8)$$

$$\eta_{90} = -1.52 \times 10^{-3} Q^2 + 4.33 \times 10^{-2} \times Q + 2.21 \quad (3.9)$$

$$\eta_{100} = -1.44 \times 10^{-3} Q^2 + 4.23 \times 10^{-1} \times Q + 3.42 \times 10^{-1} \quad (3.10)$$

$$\eta_{110} = -1.44 \times 10^{-3} Q^2 + 4.23 \times 10^{-1} \times Q - 1.66 \quad (3.11)$$

$$\eta_{110} = -1.44 \times 10^{-3} Q^2 + 4.23 \times 10^{-1} \times Q - 4.06 \quad (3.12)$$

These equations were used both in the steady state and in dynamic mode. In the dynamic mode analysis, these equations were used as characteristic curves of the pump (both for head and hydraulic efficiency).

3.3.3 LNG vaporizer and BOG heater

The vaporizer used in the system is a Shell and Plate heat exchanger. The goal of this vaporizer is to take the LNG and heat it from a temperature of around -160 °C to 25 °C. The design data of this equipment is listed in Table 3.4.

Table 3.4 – LNG vaporizer design properties

Property	Design value or designation
Type	Shell & Plate
Arrangement	Horizontal
$C_{\text{vaporizer}}$ (kW)	370
$P_{\text{shell,des}}$ (bar)	11
$P_{\text{plate,des}}$ (bar)	11
$T_{\text{shell,des}}$ (°C)	-165 to 45
$T_{\text{plate,des}}$ (°C)	-165 to 45
Materials	Stainless steel (SS 304)

The system also uses a BOG heater, which is a shell and tube heat exchanger, with the goal of warming up BOG to be send to the boiler. The main design data to this equipment is listed in Table 3.5.

Table 3.5 – BOG heater design properties

Property	Design value or designation
Type	Shell & Tube
Arrangement	Horizontal
C_{Heater} (kW)	30

The LNG vaporizer is a Shell and Plate heat exchanger, but in this case, it was approximated to a Shell and Tube equivalent heat exchanger. The BOG boiler was also simulated as an equivalent Shell and Tube heat exchanger. To perform the simulation of the heat exchangers, the area was maintained constant (22.1 m^2 for the LNG vaporizer and 2.5 m^2 for the BOG Heater) and the global heat transfer coefficient, U, was recalculated through Equation (3.13).

$$q = \dot{m} \times C_p \times \Delta T = U \times A \times \Delta T_{log} \tag{3.13}$$

In Table 3.6 and 3.7 are contemplated the different U for each flow for both heat exchangers. Table 3.8 presents the variables used to simulate the design heat exchangers used as a base for the calculations of the different U in each case. The heat exchanging fluid, in both cases the hot fluid, is a mixture of 50 vol.% of water and 50 vol.% ethylene glycol.

Table 3.6 – Global heat transfer coefficients considered for the LNG Vaporizer.

Flow (kg·h^{-1})	U (W·m^{-2}·K^{-1})
1300.0	16040
1090.0	13449
755.0	9316
545.0	6725
360.0	4442
150.0	1852

Table 3.7 - Global heat transfer coefficients considered for the BOG heater.

Flow (kg·h^{-1})	U (W·m^{-2}·K^{-1})
57.1	348
30.4	185

Table 3.8 – Design variables of the LNG Vaporizer and the BOG Heater.

Parameter	LNG Vaporizer	BOG Heater
A (m^2)	22.1	2.5
ΔT_{log} (°C)	101.4	73.1
$Fouling_{\mathrm{plate}}$ ($m^2 \cdot K \cdot W^{-1}$)	0.0034	0.0044
$T_{\mathrm{cold,in}}$ (°C)	-162	-160
$T_{\mathrm{cold,out}}$ (°C)	25	25
$T_{\mathrm{hot,in}}$ (°C)	45	45
$T_{\mathrm{hot,out}}$ (°C)	35	28
$\dot{m}_{\mathrm{cold}}$ ($kg \cdot h^{-1}$)	1300	210
$\dot{m}_{\mathrm{hot}}$ ($kg \cdot h^{-1}$)	34422	1115

A stream for the recirculation of the LNG is considered (LNGrec), since the pump has a minimum flow of 22 $l \cdot min^{-1}$ (581 $kg \cdot h^{-1}$). In the cases of minimum flow, it is necessary to separate this stream after the pump for adjusting the flow for the engines. The heat exchanging fluid, corresponding to the hot fluid in both cases, is a mixture of 50 vol.% of water and 50 vol.% ethylene glycol. The flows for the heat exchanger are fixed, 34422 $kg \cdot h^{-1}$ for the LNG vaporizer and 1115 $kg \cdot h^{-1}$ for the BOG heater, with fixed initial temperature of 45,0 ºC.

In the dynamic mode, an average U was admitted for all the simulations, namely 7404 $W \cdot m^{-2} \cdot K^{-1}$ for the LNG Vaporizer and 267 $W \cdot m^{-2} \cdot K^{-1}$ for the BOG Heater.

3.3.4 Buffer tank

A buffer tank is needed in this system to buffer and distribute the fuel gas supply downstream of the LNG vaporizer. This tank is made from stainless steel and follows the design data presented in Table 3.9.

Table 3.9 – Buffer tank design properties

Property	Design value or designation
Arrangement	Vertical
V_{BT} (m^3)	1.5

This equipment was not simulated in the steady state, since no accumulation is considered. In the simulation of the dynamic state, this equipment was simulated using a flash with a percentage of liquid equal to 0 %, acting like a manifold.

3.3.5 Engines and gas delivery

The two consumers in the system are the two engines, powered by LNG, and the boiler that uses BOG. The main parameters for both consumers are listed in Tables 3.10 and 3.11. Despite the ranges for the inlet pressure and temperature for both the LNG vaporizer and the boiler, the pressure of delivery for the LNG vaporizer must be the closest possible to 6.5 bar and 25 ºC since the engines can only react to slow and small changes in the operating conditions.

Table 3.10 – LNG engines design properties

Property	Design value or designation
Type	Rolls Royce B36:46L6PG
MCR (kW)	3600
$\dot{m}_{\mathrm{gas}}$ (kg·h^{-1})	545
P_{Feed} (bar)	6.7 to 9.0
T_{Feed} (ºC)	5.0 to 45.0

Table 3.11 – Gas boiler design properties

Property	Design value or designation
Q_{C} (kg/h)	210
P_{Feed} (bar)	1.0 to 2.0
T_{Feed} (ºC)	5.0 to 45.0

3.3.6 Heating media system

The heat required for the evaporation of the LNG (and for the BOG Heater) are provided directly from the ship's engine oil circuit and a back-up electrical heater. A water-glycol circuit in closed loop is the media used to perform the transfer from the lube oil (or the back-up electrical heater) to the vaporizer and the boiler. A pump is used to run the system and to transport the fluid to a plate type heat exchanger, where the water-glycol heats up to 45 ºC. The heat exchangers used in this section were not studied in the simulations. The pump design data is listed in Table 3.12.

Table 3.12 – Water-glycol pump design data

Property	Design value or designation
Type	Centrifugal
T_{design} (ºC)	-10.0 to 45.0
$\dot{Q}_{\text{design}}$ ($m^3 \cdot h^{-1}$)	65.0
H_{Pump} (m)	50
PC (kW)	18

3.4 Gas control system

The system in study must be automatically operated, in the sense that in shall continuously perform its purpose with little to no operator actions. The gas control system (GCS) is intended to be used for all normal operation requirements to assure that the measured values are kept within the normal range. This cover, among others, the valve positions, temperatures and pressures within the FGSS. There are two automatic control loops within the supply system. One of them has the goal of controlling the source of the feed to the gas boiler and the other has the aim of assuring the adequate conditions of the fuel that enters the engines.

3.4.1 C1 – Gas feed to boiler

Since the gas boiler on board can run with fuel from vaporized LNG or from heated BOG, an automatic control loop needs to be installed. The operation logic shall be:

- If tank pressure is greater or equal to 3.5 bar, then the source the fuel is the BOG heater, therefore a control valve placed after the LNG vaporizer must be completely closed and a control valve placed after the BOG heater must be completely open.
- If the pressure in the tank is lower than 3.5 bar, then the source of the fuel is the LNG vaporizer, therefore a control valve placed after the LNG vaporizer must be completely open and a control valve placed after the BOG heater must be completely closed.

This philosophy of control wasn’t implemented in the dynamic model, but was used manually in the steady state simulation.

3.4.2 C2 – Fuel supply control philosophy

The main purpose of the control system is to ensure that a steady flow of gas at the adequate temperature is provided to the gas consumers at a steady pressure, regardless of the fuel consumption. This is mostly achieved by monitoring the pressure in the buffer tank and maintaining it at a fixed value. A control loop is necessary; however, the final philosophy of this loop is yet in study being the following two proposals the ones that are being tested.

In the dynamic study of the work presented in this document, only P1 was attempted since P2 has the potential to be highly unstable due to the dynamic of the valves.

P1. Pressure in the buffer tank is measured by a pressure transmitter and is controlled to a fixed set-point by adjusting the running pump speed (through the pump variable frequency drive, VFD). The pump frequency range is limited within the acceptable operation profile of the pumps, therefore a split range (or equivalent logic) is implemented so that, when the pump frequency is at its minimum but the pressure of the buffer tank is still above the set point, a control valve placed before the LNG vaporized starts closing.

P2. Pressure in the buffer tank is measured by a pressure transmitter and is controlled by a cascade of control loops. The feed of fuel is controlled with a valve placed before the LNG vaporizer (the VFD control is only there to ensure the pump is operating in an efficient way). A third loop will regulate the opening of the recycle valve so that the combining openings of both valves guarantees the pump minimum flow rate.

The design data of the control valve meant to be used in the control philosophy P1 is presented in Table 3.13.

Table 3.13 – Design data of the control valve.

Property	Design value or designation for mean flow
$T_{valve,des}$ (°C)	-160
$P_{in,des}$ (bar)	8.0
$P_{out,des}$ (bar)	2.9
ΔP_{des} (bar)	5.1
$k'_{v,des}$	0.45
Valve type	Straight globe valve
Behaviour type	Equal percentage
Valve size	DN 20

4 Results and discussion

This chapter of the dissertation is divided in four major sections. The first one, Section 4.1, is dedicated to the results of the determination of important LNG properties. In the second section, Section 4.2, the results from the steady state case studies are analysed. The next section, Section 4.3, are the dynamic analysis with PID control and with the implementation of the split-range model. Finally, section 4.4 includes a study of the enthalpy of a new solution to be used as heating media in the system.

4.1 LNG properties calculation

The properties of the LNG are an important concern in study, so calculators were created to perform easy and quick property studies if the saturation pressure was altered. The properties enrolled in this study were:

1. Mass enthalpy, H_{LNG};
2. Enthalpy of vaporization, ΔH_{vap};
3. BOG viscosity, μ_{BOG};
4. LNG viscosity, μ_{LNG};
5. BOG volumetric mass density, ρ_{BOG};
6. LNG volumetric mass density, ρ_{LNG};
7. Bubble point, T_{b};
8. Dew point, T_{d};
9. LNG adiabatic expansion coefficient, k_{LNG};
10. BOG adiabatic expansion coefficient, k_{BOG};
11. BOG Z factor, Z.

To perform this study a flash was created, as seen in Figure 4.1. FEED represents a stream of LNG with a vapor fraction of 10%, and streams LNG and BOG represent the streams of boil-off-gas and liquid LNG respectively. A sensitivity study was conducted between 1.1 and 36.0 bar, keeping the saturation conditions for each case, and a graph for each property was created.

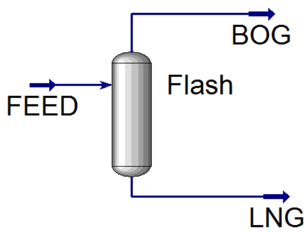

Figure 4.1 – Flowsheet of the simulation used for the LNG property study.

The first chart, Figure 4.2, represents the mass enthalpy of the LNG with alteration of the pressure in the vessel. It is observed that the higher the pressure the lower is the specific enthalpy. The enthalpy of vaporization in Figure 4.3 also decreases with the increase of the pressure. This impact is expected, since increasing the pressure has the overall effect of reducing the heat needed to the phase change, until it reaches 0 at the critical point.

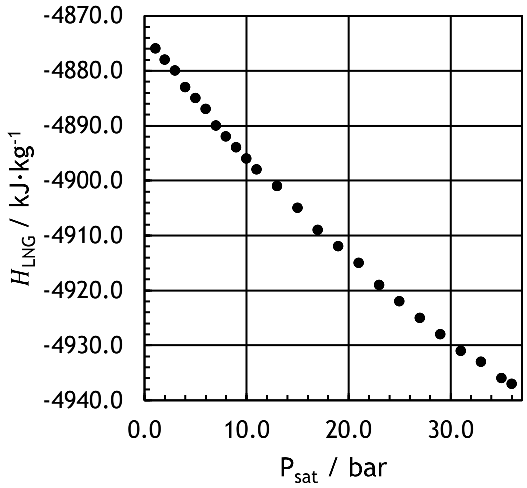

Figure 4.2 – Saturation pressure effect on mass enthalpy.

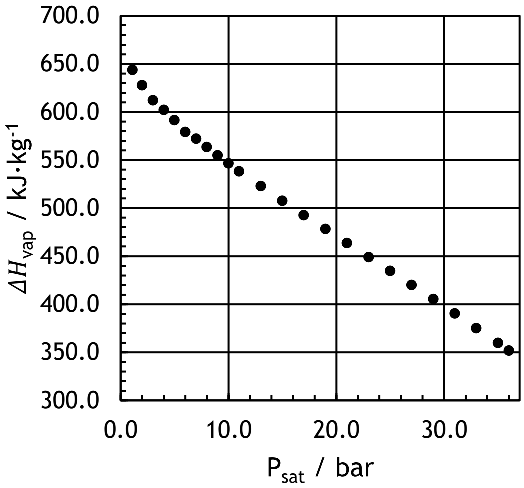

Figure 4.3 – Saturation pressure effect on enthalpy of vaporization.

Figure 4.4, plots the evolution of the bubble temperature with pressure. The behaviour of this chart is expected, since the bubble point increases with increased pressure up to the critical point, where the gas and liquid properties become identical. The dew point dependency with the pressure is represented in Figure 4.5. Increasing the pressure increases the dew point, which means that at higher pressures it is needed a higher temperature for the condensed phase to start separating from the gas phase.

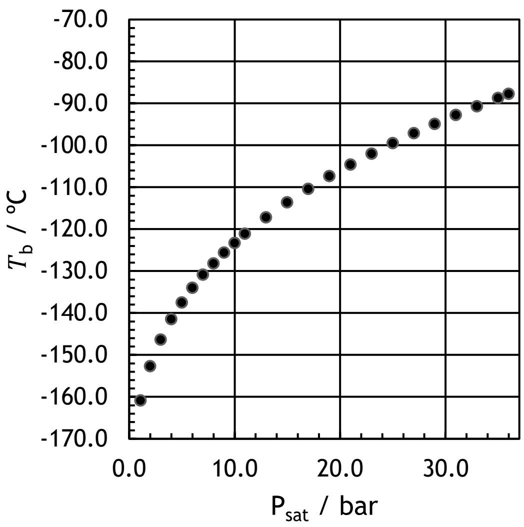

Figure 4.4 – Saturation pressure effect on bubble temperature.

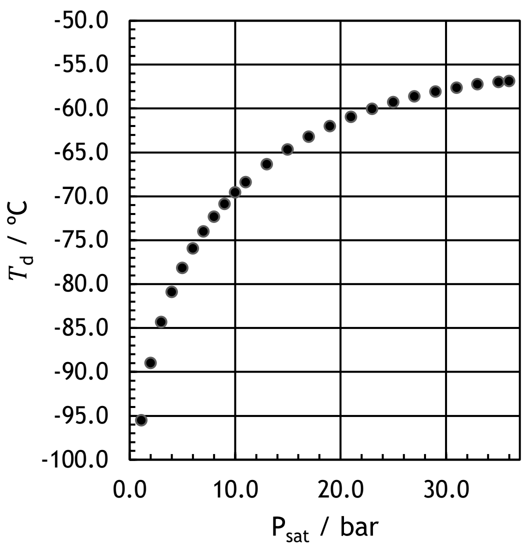

Figure 4.5 - Saturation pressure effect on the dew point.

Figure 4.6 and Figure 4.7 represent the dynamic viscosity and the volumetric mass density of both LNG and BOG, respectively. LNG viscosity is higher than the BOG viscosity since it is a liquid, the same happens for the volumetric mass density (liquids' volumetric mass density is, normally, a thousand times higher than gases').

Dynamic viscosity is more dependent of the temperature than of the pressure. So, keeping the saturation conditions means that, increasing the pressure, we are also increasing the temperature in the vessel leading to a decrease of LNG's dynamic viscosity with pressure. On the other hand, gases' viscosity increases with an increase of temperature, thus the slight increase of the BOG's dynamic viscosity for higher pressures. Following the same line of thought, the volumetric mass density of the liquid decreases with the increasing of temperature (and pressure). For the BOG, it is necessary to use the ideal gas law, presented in Equation (4.1). This equation can be written in function of the density, as represented in Equation (4.2).

$$P \times V = n \times R \times T \tag{4.1}$$

$$\frac{P}{T} = \frac{R}{M} \times \rho \tag{4.2}$$

Where P is the pressure, T is the temperature, v is the volume, R is the ideal gas constant, M is the molar mass and ρ is the volumetric mass density. Since the ratio between the pressure and the temperature is increasing, the volumetric mass density of BOG is increasing too.

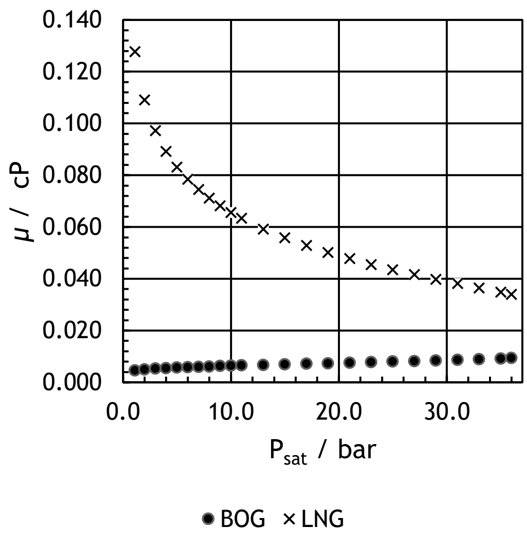

Figure 4.6 - Saturation pressure effect on the dynamic viscosity of BOG and LNG.

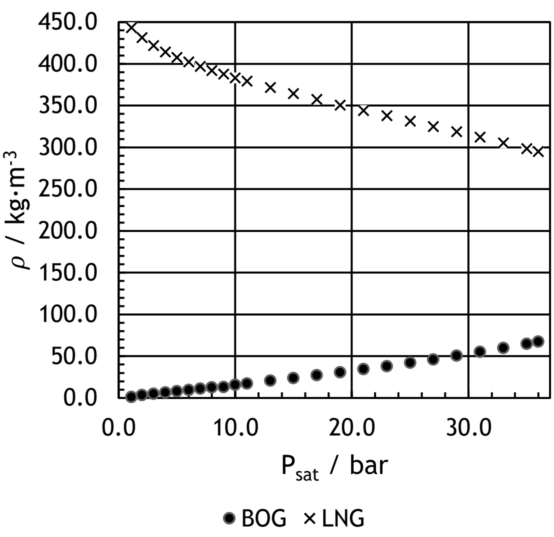

Figure 4.7 - Saturation pressure effect on the volumetric mass density of BOG and LNG.

Z is the compressibility factor of a gas and it describes the deviation of a real gas from the ideal gas behaviour. Figure 4.9 represents the evolution of Z with the increasing of the saturation pressure. Since LNG is 94.0% of methane, it can be concluded that it has the same tendency and that the deviation from ideal gas must increase with increasing pressure. Finally, the adiabatic expansion coefficient decreases for both LNG and BOG with the increasing saturation pressure. This constant, k, is the ratio between the thermal capacity at constant pressure, C_p, and the thermal capacity at constant volume, C_v, and it's represented in Figure 4.10..

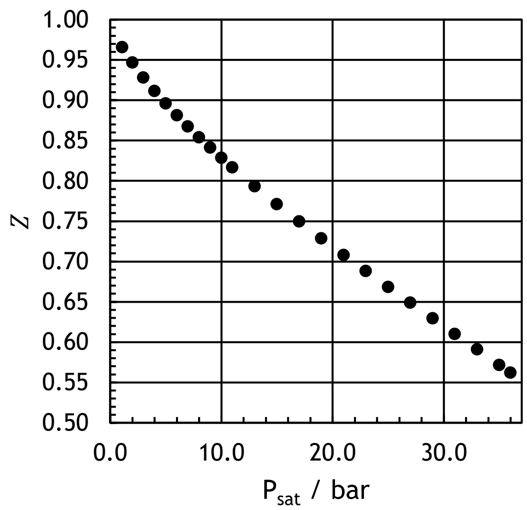

Figure 4.8 – Saturation pressure effect on the compressibility factor.

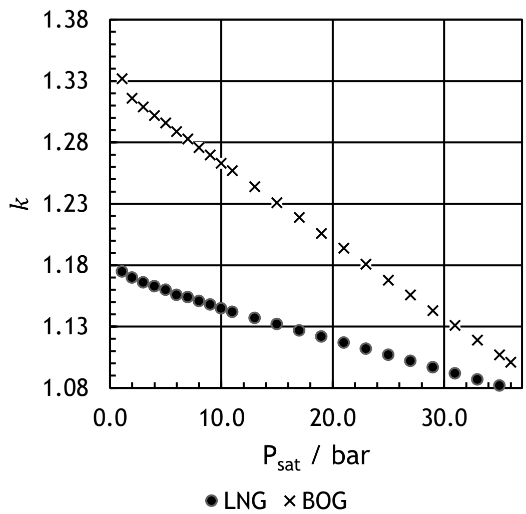

Figure 4.9 - Saturation pressure effect on the adiabatic expansion coefficient.

For each property, an appropriate equation has been set: then a Excel file was created with automatic calculators for use in Høglund Gas Solutions AS as a work tool, since this is quite useful to predict properties without the need of the construction of new simulations for each new scenario that must be studied. Equations (4.1) to (4.12) were the expressions generated to predict properties within the indicated range. The data used in this study can be consulted in Appendix A.

$$H_{\mathrm{LNG}} = 1.64 \times 10^{-2} \times {P_{sat}}^{2} - 2.34 \times P_{sat} - 4.87 \times 10^{3} \tag{4.3}$$

$$\Delta H_{\mathrm{vap}} = -4.00 \times 10^{3} \times {P_{sat}}^{3} + 2.68 \times 10^{2} \times {P_{sat}}^{2} - 1.28 \times 10^{1} \times P_{sat} + 6.51 \times 10^{2} \tag{4.4}$$

$$MW_{\mathrm{BOG}} = 3.83 \times 10^{-7} \times {P_{sat}}^{4} - 3.32 \times 10^{-5} \times {P_{sat}}^{3} + 1.18 \times 10^{-3} \times {P_{sat}}^{2} - 1.87 \times 10^{-2} \times P_{sat} + 1.63 \times 10^{1} \tag{4.5}$$

$$T_{\mathrm{b}} = 2.40 \times 10^{-3} \times {P_{sat}}^{3} - 1.82 \times 10^{-1} \times {P_{sat}}^{2} + 5.56 \times P_{sat} - 1.63 \times 10^{2} \tag{4.6}$$

$$\mu_{\mathrm{BOG}} = 1.14 \times 10^{-7} \times {P_{sat}}^{3} - 7.17 \times 10^{-6} \times {P_{sat}}^{2} + 2.54 \times 10^{-4} \times P_{sat} + 4.52 \times 10^{-3} \tag{4.7}$$

$$\mu_{\mathrm{LNG}} = 3.38 \times 10^{-7} \times {P_{sat}}^{4} - 3.01 \times 10^{-5} \times {P_{sat}}^{3} + 9{,}60 \times 10^{-4} \times {P_{sat}}^{2} - 1.41 \times 10^{-2} \times P_{sat} + 1.36 \times 10^{-1} \tag{4.8}$$

$$\rho_{\mathrm{BOG}} = 1.57 \times 10^{-2} \times {P_{sat}}^{2} + 1.26 \times P_{sat} + 1.03 \tag{4.9}$$

$$\rho_{\mathrm{LNG}} = -3.54 \times 10^{-3} \times {P_{sat}}^{3} + 2.46 \times 10^{-1} \times {P_{sat}}^{2} - 8.53 \times P_{sat} + 4.47 \times 10^{2} \tag{4.10}$$

$$T_{\mathrm{d}} = -1.14 \times 10^{-4} \times {P_{sat}}^{4} + 1.03 \times 10^{-2} \times {P_{sat}}^{3} - 3.47 \times 10^{-1} \times {P_{sat}}^{2} + 5.59 \times P_{sat} - 9.94 \times 10^{1} \tag{4.11}$$

$$Z = -4.33 \times 10^{-6} \times {P_{sat}}^{3} + 3.37 \times 10^{-4} \times {P_{sat}}^{2} - 1.82 \times 10^{-2} \times P_{sat} + 9.81 \times 10^{-1} \tag{4.12}$$

$$k_{\mathrm{LNG}} = 1.17 \times 10^{-5} \times {P_{sat}}^{2} - 3.06 \times 10^{-3} \times P_{sat} + 1.18 \tag{4.13}$$

$$k_{\mathrm{BOG}} = 1.52 \times 10^{-5} \times {P_{sat}}^{2} - 6.94 \times 10^{-3} \times P_{sat} + 1.33 \tag{4.14}$$

The information of these charts was used to confirm the pressure of the tank. The interval of pressures chosen were between 1.25 and 5.0 bar, which means that the gas has a lower deviation from the ideal gas behaviour. Defining these properties over a wide range of pressures is important to understand the system's behaviour and defining the proper operating conditions to ensure system stability and to meet the requirements of the industrial equipment used.

4.2 Steady state

The first step to simulate dynamically the system in study was to simulate its behaviour in the steady state as a first approximation. The goal of this analysis was to determine if the pressure of the streams entering the engines was within the range of suitable values (preferably between 6.7 and 10.0 bar).

4.2.1 Scenarios to study in steady state

The list of scenarios performed is represented in Table 4.1. There were 18 scenarios studied, varying the pressure in the tank, the flow to the engines and the season of the year.

The pressures considered were 1.25 bar, 3.0 bar and 5.0 bar. For each pressure were tested 3 different inlet engine flows:

- The maximum flow was considered equal to both engines working at maximum flow, which is 545 $kg \cdot h^{-1}$ of LNG for each engine.
- The average flow was considered half of the maximum flow, representing 273 $kg \cdot h^{-1}$ for each engine.
- The minimum flow was considered 150 $kg \cdot h^{-1}$ (the minimum allowed to go into each engine), but with only one engine working.

The last variable was the season of the year, which interfered with the boiler flow. In the summer, the boiler doesn't work most of the times, as the heat from it is not needed. In this case, the flow is equal to zero. In winter, it was considered that the boiler worked at maximum flow. The maximum flow for the boiler was defined in a different way for the different pressures:

- For the pressures 1.25 and 3.0 bar, is the LNG vaporizer that sends the gas stream for the boiler. In this case, it was considered maximum flow equal to 210 $kg \cdot h^{-1}$.
- For the higher pressure, 5.0 bar, is the BOG heater that sends the gas stream to the boiler. In this case, the maximum flow is limited by the level of liquid in the tank, so it depends from case to case, as indicated in Table 4.1.

Table 4.1 – List of the tests that were performed in the steady state simulation.

Case	Tank Pressure (bar)	Flow to the engines ($kg \cdot h^{-1}$)	Season	Flow to boiler ($kg \cdot h^{-1}$)
1	1.25	Maximum 1090	Summer	-
2			Winter	210
3		Average 545	Summer	-
4			Winter	210
5		Minimum 150	Summer	-
6			Winter	210
7	3.0	Maximum 1090	Summer	-
8			Winter	210
9		Average 545	Summer	-
10			Winter	210
11		Minimum 150	Summer	-
12			Winter	210
13	5.0	Maximum 1090	Summer	-
14			Winter	57.1
15		Average 545	Summer	-
16			Winter	30.4
17		Minimum 150	Summer	-
18			Winter	30.4

4.2.2 Results of the steady state simulations

The most important result that can be analysed from the simulation illustrated above is the pressure of the stream that enters both engines. In fact, this pressure must be within the interval of 6.7 to 9.0 bar, as indicated in the *Technical Specification* document [51].

The pressure, in bar, of the streams that enter the engines as well as its flows and temperatures can be consulted in Table 4.2.

Table 4.2 – Conditions of the engines' inlet stream for the 18 scenarios of study.

Case	Total engine flow ($kg\cdot h^{-1}$)	Pressure (bar)	Temperature (°C)
1	1090	7.0	26.0
2	1090	6.8	25.1
3	545	7.0	28.2
4	545	7.0	27.3
5	150	7.0	29.8
6	150	7.0	29.0
7	1090	7.2	25.8
8	1090	7.1	25.6
9	545	7.5	28.5
10	545	7.4	27.7
11	150	6.9	29.9
12	150	7.5	29.9
13	1090	7.8	26.7
14	1090	7.8	26.7
15	545	8.1	28.8
16	545	8.1	28.8
17	150	8.3	28.8
18	150	8.3	28.8

It can be concluded that, in all the case studies, the engines inlet pressure is within the appropriate range. It can be also observed that the pressure was not kept at 6.7 bar (the design pressure), as desired. This happened because the pump curves used were only the curves represented in Figure 3.5, instead of varying the frequency of the pump to achieve the desired pressure (and consequently, the desired curve). This simplification was performed in the steady state analysis since it's not possible to design a variable frequency drive or to input pump characteristic curves in a simulation that doesn't take time as a variable.

The temperature of the stream that enters the boiler should be within 0 and 45 ºC, preferably the closest as possible to 25 ºC. The temperature, pressure and flow of the streams entering the boiler can be consulted in Table 4.3.

Table 4.3 - Conditions of the boiler's inlet stream for the 9 cases of study.

Case	Origin	Flow ($kg \cdot h^{-1}$)	Pressure (bar)	Temperature (°C)
2	LNG Vaporizer	210	6.8	25.1
4			7.0	27.4
6			7.0	29.0
8			7.1	25.6
10			7.4	27.7
12			7.5	29.3
14	BOG Heater	57.1	5.0	32.7
16		30.4	5.0	31.8
18		30.4	5.0	33.0

It can be concluded that, for all the study cases, the temperature entering the boiler is suitable for the application in this system. Additional table results for all the case studies can be consulted in the Appendixes C, D and E.

4.3 Dynamic mode and control system

The goal of the control system implemented in the dynamic simulation was the control of the inlet pressure of the fuel into the engines in an automatic form. To do a more accurate approximation of the system, characteristic curves were used in the pump (PumpLNG) and a VFD was simulated.

4.3.1 PID controller of the VFD

To do that, the first approach was the control of the VFD with a PID controller, as illustrated in Figure 4.10. The simplified system has the Flash to separate the LNG and the BOG, the LNG pump and the separator for the recycle. Since the pressure drop in the LNG vaporizer and the rest of the system is very low (because the fuel is now in the gaseous state) it can be neglected. The parameters of the PID controller (VFD_controller) are indicated in Table 4.4. This controller adjusts the pressure by varying the speed of the pump.

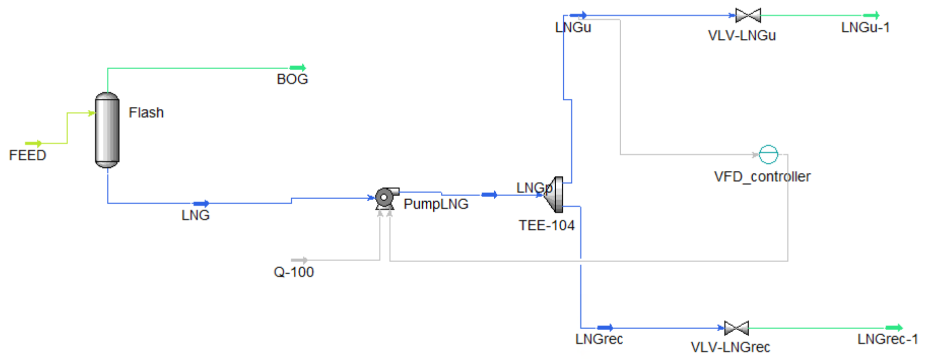

Figure 4.10 - System used to test the PID control of the VFD.

Table 4.4 - Data of the PID controller used

VFD_controller	
PV	Pressure of LNGu
OP	Speed of PumpLNG
Action	Reverse
Algorithm type	Hysys
Algorithm subtype	PID Velocity Form
K_c	7.7×10^{-2}
T_i (s)	7.0×10^{-2}

PV is the controlled variable and OP is the manipulated variable. The parameters used in this controller were obtained using the function *Autotuning* in the software used. This system was tested for the conditions enumerated in Table 4.5.

Table 4.5 - Scenarios tested for the PID controller model.

Case	P_{tank} (bar)	$\dot{m}_{pump}$ (kg·h^{-1})	PV interval (bar)
1	1.25	1090	6.0 - 7.0
2		580.0	6.2 - 6.8
3	3	1090	6.5 - 7.5
4		580.0	6.5 - 7.5
5	5	1090	8.0 - 9.0
6		580,0	8.0 - 9.0

The PV interval for each case was adjusted considering the pressure adjusted by the pump for each case. The flow was changed using the *Adjust* function in the software. Figures 4.11, 4.12 and 4.13 represent the results for case 1, 3 and 5, respectively. The red line represents the profile of setpoint implemented (SP), the green line represents the controlled variable (PV), in this case the pressure in the buffer tank, and the blue line represents the controller action (OP).

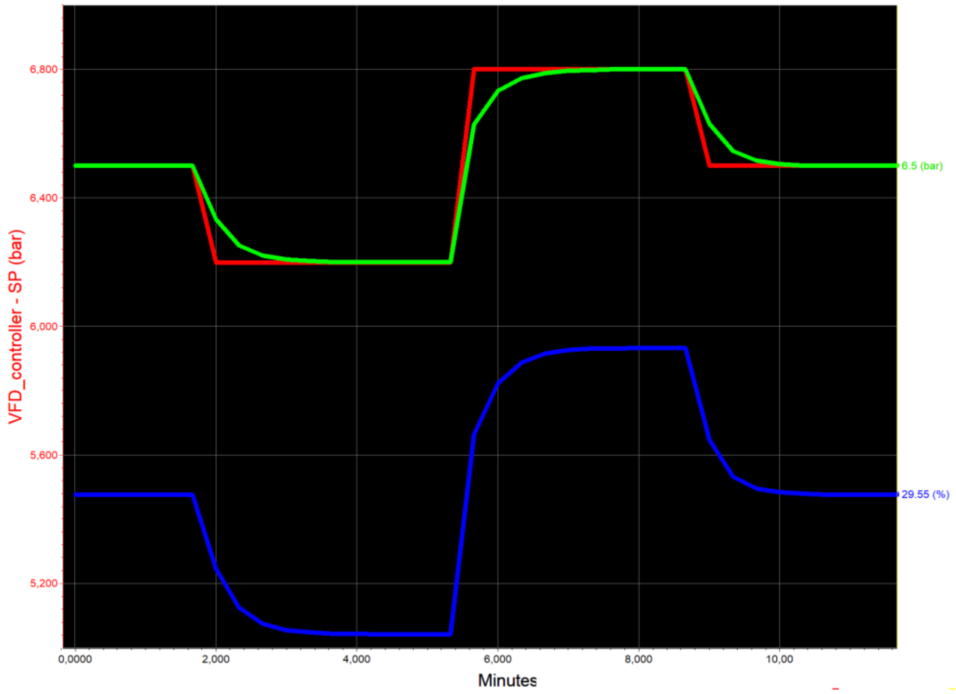

Figure 4.11 – Stripchart of case 1.

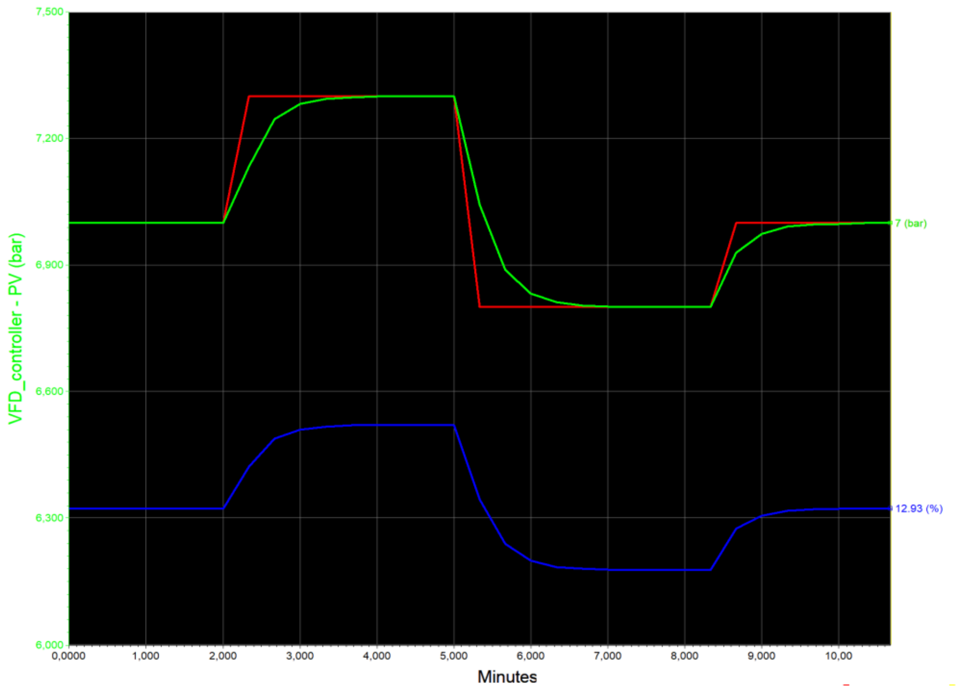

Figure 4.12 – Stripchart of case 3.

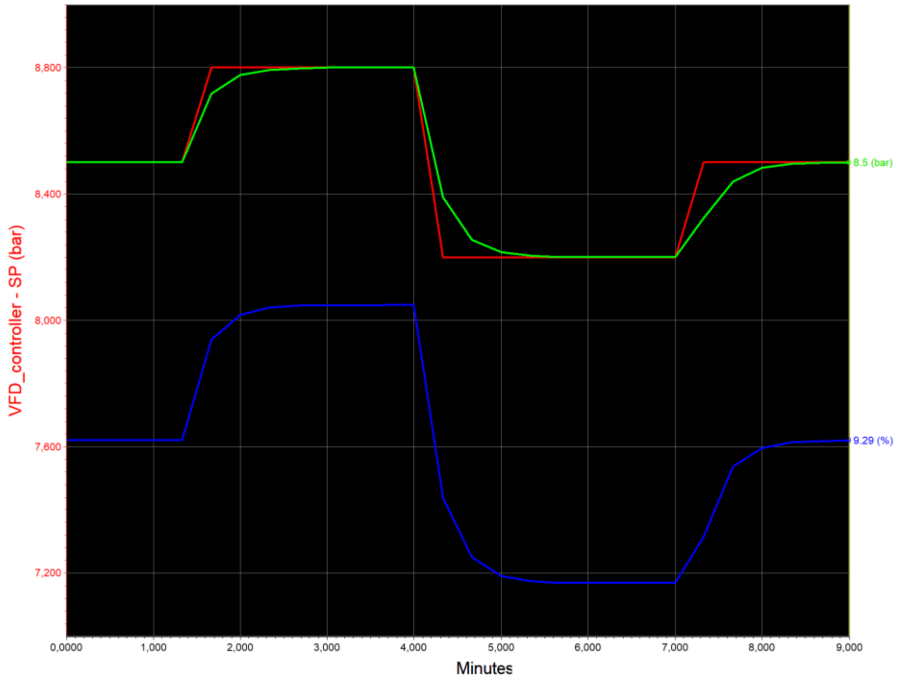

Figure 4.13 – Stripchart of case 5.

As can be seen, the controller used is very responsive to sudden modifications of pressure in the indicated stream, being a PI approach (with no derivative control) enough to provide a stable response of the system. The action of the controller, OP, (meaning the speed of the pump) also follows the tendency of the controller, showing that the VFD is working properly. For the lower pressure in the tank, the velocity is higher, meaning that the variation of pressure involved in the operation is high. For the high pressure in the tank, the velocity of the pump is lower since the pressure variation needed to achieve the wanted result is lower. All the scenarios studied lead to inlet pressures within the proper interval (between 6.7 bar and 9.0 bar), meaning that this system is effective in controlling the pressure of the stream.

4.3.2 Split-Range Controller of the VFD

In order to fully control the system, a split-range controller philosophy was idealized and implemented in the software. However, to make the calculations more precise with this model, a pipe was added to consider the pressure drop between the tank of LNG and the vaporizer. The results of the calculation as well as the data of the pipe shown in Table 4.6.

Table 4.6 – Design data and pressure drop of the piping

Property	Value or designation
Length (m)	40.0
Elevation change (m)	0
Pipe schedule	Schedule 40
Nominal diameter (mm)	20
Roughness (m)	4.6×10^{-5}
Pressure drop (bar)	0,14

As can be observed from the result obtained, the pressure drop in the pipe system is low despite the fluid being a liquid, where normally the pressure drop is much higher and relevant than with gases. The split-range system was idealized as two controllers in cascade mode, where the both set points were given by a third controller (a split-range controller). Figure 4.14 represents the system as it was simulated. This control strategy was developed considering the philosophy P1 mentioned in Section 3.4.2., in which the speed of the pump is adjusted to control the pressure in the buffer tank and, if the pump is working at minimum speed but the pressure is still too high, the valve will adjust the pressure by closing.

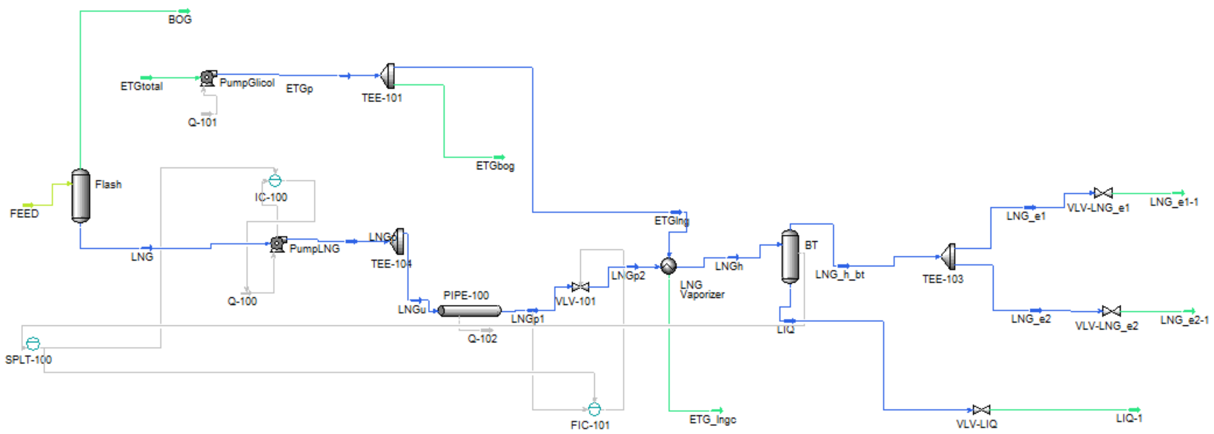

Figure 4.14 – Split-range controller implementation.

As can be seen in the figure above, there are two PID controllers operating in cascade mode: IC-100 (that controls the speed of the pump) and PIC-101 (that controls the opening of the control valve, VLV-101). Both controllers' setpoints are given by a third controller, SPLT-100, that measures the pressure in the buffer tank, BT, and gives information to both controllers mentioned (speed data to IC-100 and opening valve percentage to FIC-101). Since the actions of the controllers are not meant to happen at the same time, the split-range split itself must be adjusted. The desired action is if the action takes place first in the IC-100 and, if not enough, FIC-101 adjusts the valve position. The split can have an overlap to guarantee a smoother switch between both controllers. The data for the three controllers is summarized in Table 4.7.

Table 4.7 – Setup data for the three controllers used in the split-range implementation

Data	IC-100	FIC-101	SPLT-100
Type of control	PI	PI	Split-range
PV	Speed of PumpLNG	Mass flow of LNGp1	Pressure of BT
OP	Power to the pump (Q-100)	Actuator desired position of VLV-101	Setpoint of IC-100 and FIC-101
SP	Remote by SPLT-100	Remote by SPLT-100	6.7 bar
Action	Reverse	Reverse	Reverse
Mode	Casc	Casc	Auto
PVmin	4200 rpm	150 $kg \cdot h^{-1}$	6.7 bar
PVmax	6900 rpm	1090 $kg \cdot h^{-1}$	9.0 bar
Algorithm type	Hysys	Hysys	Hysys
Algorithm subtype	PID Velocity Form	PID Velocity Form	PID Velocity Form
Split	0 – 70 %	60 – 100 %	-

Unfortunately, this possibility wasn't tested to any scenario, being only implemented as a concept due to the limitations of the software package. The work implies the development of software modules, involving the connection between *Aspen HYSYS®* and *Visual Basic* scripts, which is out of scope of the work.

4.4 New possibility as heating media

The final goal of this thesis was the testing of the viability of a new heating media to optimize the heat exchange in the LNG vaporizer for future supply systems. The fluid shall have a dew point of around -165 ºC and must be non-toxic and non-flammable.

The fluid considered to perform this study was a mixture of 17 wt.% of oxygen and 83 wt.% of argon. To test if this fluid can perform the heat transfer needed to evaporate the cryogenic LNG, a study of properties of the mixture was performed and a P-H diagram (Figure 4.15) was created. The full data used to create the chart is presented in Appendix F.

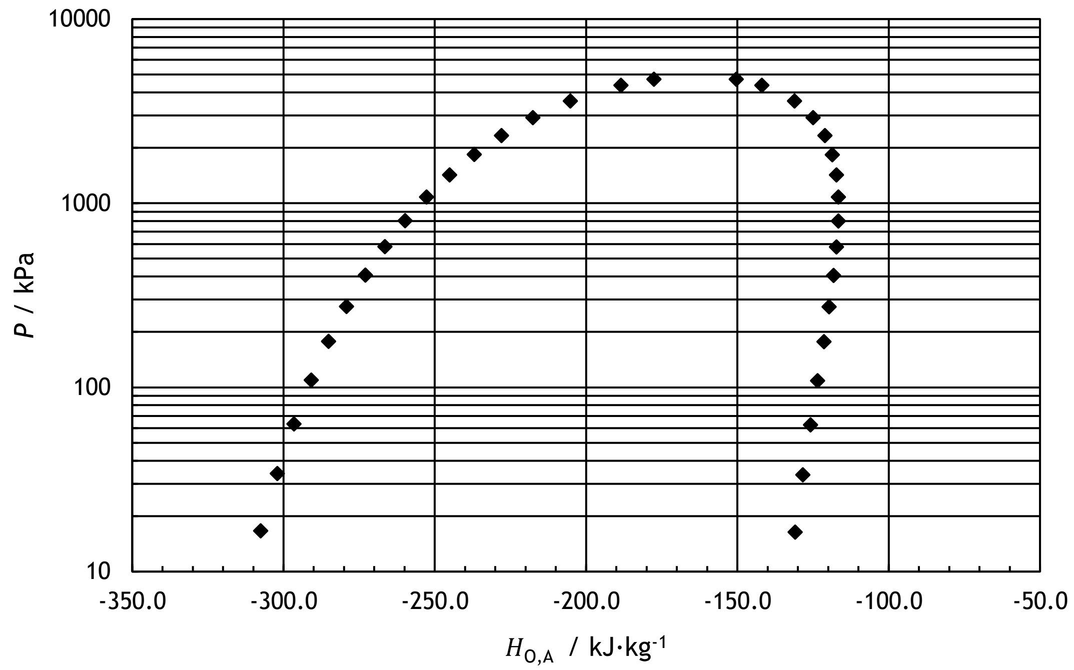

Figure 4.15 – P-H diagram for a mixture of oxygen and argon.

From the diagram and the simulation performed it can be concluded that the mixture of oxygen and argon, at 5.8 bar, has a dew point of around -165 ºC, which is adequate for the application. The pressure determined is also reasonable for a system for evaporation of LNG. The enthalpy of vaporization is around one third of the methane enthalpy of vaporization, meaning that the flow needed of this mixture to perform the heat transfer needed must be 3 times higher than the flow of LNG. Despite being a high flow rate, it’s a big improvement compared to the ethylene glycol and water mixture used in the system tested in this dissertation.

The mixture responds also to the needs indicated initially: is non-toxic and non-flammable, meaning that is safe to be transported in closed spaces inside of a ship.

Based on these conclusions, it is safe to say that the mixture of oxygen and argon considered in this section is a promising heating media is an alternative for the glycol water mixture already in use.

5 Conclusions

The estimation of LNG properties are an important starting point to the definition of the system's conditions to operate an LNG supply system, so a series of properties were studied for the pressure range between 1.1 and 36.0 bar. Mass enthalpy and vaporization enthalpy of LNG decrease with the increasing of the saturation pressure, however, the dew and bubble points increase with the increasing of the same parameter. Viscosity and volumetric mass density of LNG decrease with the increasing of the saturation pressure but, on the other hand, viscosity and volumetric mass density of BOG increase with the increasing of the saturation pressure. The adiabatic expansion coefficient decreases for both BOG and LNG with the increasing of saturation pressure and the BOG Z factor decreases with the increasing of the saturation pressure. It can be concluded that the lower range of pressures is adequate to set as operation pressure in the LNG storage tank.

The steady state scenarios tested would vary on the tank pressure (between 1.25 and 5.0 bar), on the flow to the engines (from 150 $kg \cdot h^{-1}$ to 1090 $kg \cdot h^{-1}$) and on the use or not of the BOG boiler (with a flow of 210 $kg \cdot h^{-1}$). For all 18 cases tested, the pressure of the flow to the engines were between 6.8 and 8.3 bar, which is within the appropriate range of 6.7 to 9.0 bar. The temperature of the fuel to the engines were between 25.1 and 29.9 ºC, which is also within the desired range (it must be as closest as possible to 25 ºC, but between 0 and 45 ºC). So, it can be concluded that, for all cases, the conditions of the fuel entering the engines are suitable for the application in this system.

The PID controller, used in the simplified system to control the pressure in the buffer tank (by varying the speed of the VFD), was tested for 6 scenarios, varying the pressure in the tank (between 1.25 and 5.0 bar) and varying the flow that goes through the pump (1090 $kg \cdot h^{-1}$ is the maximum and 580.0 $kg \cdot h^{-1}$ is the minimum, respectively). The parameters adjusted for this controller were a K_c of 7.7×10^{-2} and a T_{i} of 7.0×10^{-2}. It can be concluded from the stripcharts that the controller parameters used are well adjusted to the system since the answer of the controller is fast and stable. It can also be concluded that, for the simplified system, this controller can ensure the correct pressure of the buffer tank and consecutively the adequate pressure for the fuel inlet to the engines.

The split-range controller was implemented, and it is possibly the correct way to control all the system (as it can control both the speed of the pump and the flow to the vaporizer). However, this system wasn't tested due to limitations of the software.

Finally, the mixture of 17 wt.% of oxygen and 83 wt.% argon is promising as heating system to future LNG supply systems as it is a safe fluid to be transported in ships. It has a dew point of -165 ºC approximately at 5.8 bar, which makes it adequate for the application.

6 Assessment of the work done

6.1 Objectives Achieved

Different properties of LNG were tested to determine which are the better operating conditions for an LNG supply system, carried out in the marine industry. The steady state study was performed and analysed, enabling the conclusion that the conditions of the fuel delivery were appropriate for the ranges defined in the design specifications. The dynamic simulation was carried out in a simplified manner, with some equipment left out due to difficulties of convergence. However, the PID controller was implemented and it can be concluded that the pressure of fuel delivery is within the defined range. A split-range control approach in the dynamic mode was also implemented but not tested. Finally, the study of a new heating media system took place and the results were compared to the already used mixture of ethylene glycol and water. So, the objectives for this dissertation were achieved, however, due to several limitations, it was not possible to tune and test the split-range controller.

6.2 Limitations and Future Work

The main limitation of this work was the conceptualization of the system itself, due to all the equipment needed to be simulated. Another limitation was the utilization of a VFD to adjust the pump speed: this equipment was particularly hard to simulate since the software wasn't accepting the characteristic curves given as input. There was the need to contact *Aspentech®* support team, but it was a slow process that stole time to perform the tests needed. Another challenge was the utilization of a new software as it was a time-consuming task to learn how to use *Aspen HYSYS®* and to get used to all the tools available. For the future, it is important to tune and test the split-range controller with the goal of control the all system for all scenarios needed. There, the system should be completely automatic. It would be also interesting to make an economical study to access if the mixture of oxygen and argon is economically viable to be used as alternative to the current heating media.

6.3 Final Assessment

Process simulation is a powerful tool to study systems and to test different scenarios, being *Aspen HYSYS®* the main software used for oil and gas industries. This work allowed me to consolidate my knowledge in the field of Separation Processes, Transfer Phenomena and Thermodynamics and provided a great opportunity to learn a new simulation software used widely in the industry. It was also interesting to get to know the shipping industry more deeply and to create a solution for a current challenge: the air pollution.

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Appendix A. Description – Flowsheet streams

Table A.1 – Description of the streams used in the flowcharts of the simulations (steady state and dynamic mode)

Name	Description
ETGtotal	Stream of the mixture of ethylene glycol and water for the BOG heater and the LNG vaporizer.
ETGp	Stream of the mixture of ethylene glycol and water for the BOG heater and the LNG vaporizer that has been pressurized by PumpGlicol until a pressure of about 6.0 bar.
ETGbog/ETGlng	After the split (TEE-101), these streams are the fraction of ETGp that goes into the BOG heater or the LNG vaporizer, respectively.
ETGbogc/ETGlngc	Mixture of ethylene glycol and water cold from the BOG heater or the LNG vaporizer, respectively.
ETG_out	Mixture of ethylene glycol and water from the BOG heater and the LNG vaporizer after being mixed in MIX-100.
FEED	LNG stream that goes into the flash to be separated into BOG and LNG.
BOG	Gaseous fraction of the FEED. It is directed to the BOG heater.
LNG	Liquid fraction of the FEED.
LNG1/LNG2	Only one of these streams is in operation, since only one of the pumps is also in operation.
LNGp	Pressurized LNG after Pump1 or Pump 2.
LNGrec	Part of the stream of LNG that is recycled.
LNG_vap	Part of the stream of LNG redirected to the LNG vaporizer.
LNGh	Heater and vaporizer LNG, now in the gaseous state.
LNGboiler	In some cases, the boiler is run by LNG. In this case, this stream is used.
LNG_etotal	LNG that goes into the engines.
LNG_e1/LNG_e2	After split TEE-103, these streams are the inlet fuel streams for each engine.

Appendix B. LNG's properties - full data

Table B.1 – Properties of LNG between 1.10 and 36.0 bar

P (bar)	Mass enthalpy (kJ·kg^{-1})	Heat of vaporization (kJ·kg^{-1})	BOG Viscosity (cP)	LNG Viscosity (cP)	LNG Density (kg·m^{-3})	BOG Density (kg·m^{-3})	Bubble Point (°C)	Dew Point (°C)	LNG Cp(Cp-R)	BOG Cp(Cp-R)	BOG Z factor
1.10	-4876	643.4	0.00	0.13	443.1	1.2	-160.9	-95.5	1.18	1.33	0.97
2.00	-4878	627.6	0.00	0.11	431.3	3.4	-152.7	-89.0	1.17	1.32	0.95
3.00	-4880	611.8	0.01	0.10	421.9	5.0	-146.4	-84.3	1.17	1.31	0.93
4.00	-4883	601.9	0.01	0.09	414.3	6.5	-141.5	-80.9	1.16	1.30	0.91
5.00	-4885	591.3	0.01	0.08	407.8	8.0	-137.5	-78.2	1.16	1.30	0.90
6.00	-4887	578.9	0.01	0.08	402.1	9.5	-134.0	-75.9	1.16	1.29	0.88
7.00	-4890	572.0	0.01	0.07	396.9	11.1	-130.9	-74.0	1.15	1.28	0.87
8.00	-4892	563.2	0.01	0.07	392.0	12.6	-128.2	-72.4	1.15	1.28	0.85
9.00	-4894	554.6	0.01	0.07	387.5	13.1	-125.6	-70.9	1.15	1.27	0.84
10.00	-489	546.3	0.01	0.07	383.2	15.7	-123.3	-69.6	1.15	1.26	0.83
11.0	-4898	538.2	0.01	0.06	379.1	17.3	-121.1	-68.4	1.14	1.26	0.82
13.0	-4901	522.6	0.01	0.06	371.4	20.4	-117.2	-66.4	1.14	1.24	0.79
15.0	-4905	507.4	0.01	0.06	364.2	23.7	-113.6	-64.7	1.13	1.23	0.77
17.0	-4909	492.6	0.01	0.05	357.3	27.1	-110.4	-63.2	1.13	1.22	0.75
19.0	-4912	478.0	0.01	0.05	350.6	30.6	-107.4	-62.0	1.12	1.21	0.73
21.0	-4915	463.5	0.01	0.05	344.1	34.2	-104.6	-61.0	1.12	1.19	0.71
23.0	-4919	449.0	0.01	0.05	337.7	38.0	-102.0	-60.1	1.11	1.18	0.69
25.0	-4922	434.5	0.01	0.04	331.4	42.0	-99.5	-59.3	1.11	1.17	0.67
27.0	-4925	419.9	0.01	0.04	325.0	46.1	-97.2	-58.6	1.10	1.16	0.65
29.0	-4928	405.2	0.01	0.04	318.6	50.3	-94.9	-58.1	1.10	1.14	0.63
31.0	-4931	390.3	0.01	0.04	312.1	54.9	-92.8	-57.6	1.09	1.13	0.61
33.0	-4933	375.1	0.01	0.04	305.3	59.7	-90.7	-57.3	1.09	1.12	0.59
35.0	-4936	359.5	0.01	0.03	298.4	64.8	-88.8	-57.0	1.08	1.11	0.57
36.0	-4937	351.6	0.01	0.03	294.8	67.5	-87.8	-56.9	1.08	1.10	0.56

Appendix C. Complete material streams for the stationary case studies

Tables C.1 to C.18 summarize the complete material streams results for all the stationary state simulations that were studied, being all the tables retrieved from the software Aspen HYSYS®. The nomenclature used is the enunciated in Figure 3.3 of Chapter 3.3.

Table C.1 – Material streams for case 1

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	25.98	31.53	-158.6	-156.4	-158.6	-158.6	-156.6	-156.4	45.00
Pressure	bar	1.250	1.250	6.956	1.213	1.250	7.365	1.250	1.250	7.450	7.365	1.000
Molar Flow	kgmole/h	67.39	3.571	63.81	3.571	63.81	63.81	63.81	0.0000	0.0000	63.81	1235
Mass Flow	kg/h	1149	58.71	1090	58.71	1090	1090	1090	0.0000	0.0000	1090	3.554e+004
Liquid Volume Flow	m3/h	3.729	0.1892	3.540	0.1892	3.540	3.540	3.540	0.0000	0.0000	3.540	33.71
Heat Flow	kcal/h	-1.452e+006	-6.718e+004	-1.155e+006	-6.165e+004	-1.385e+006	-1.383e+006	-1.385e+006	0.0000	0.0000	-1.383e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	38.91	36.94	45.05	25.98	25.98	25.98	25.98	36.87	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	6.956	6.956	6.956	6.956	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	51.52	25.76	25.76	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	880.0	440.0	440.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6820	2.858	1.429	1.429	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.012e+006	-9.604e+007	-9.280e+007	-2.226e+005	-9.327e+005	-4.663e+005	-4.663e+005	-9.303e+007	

Table C.2 – Material streams for case 2

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	25.11	30.81	-158.6	-156.6	-158.6	-158.6	-156.6	-156.6	45.00
Pressure	bar	1.250	1.250	6.805	1.213	1.250	7.214	1.250	1.250	7.450	7.214	1.000
Molar Flow	kgmole/h	80.37	4.260	76.11	4.260	76.11	76.11	76.11	0.0000	0.0000	76.11	1235
Mass Flow	kg/h	1370	70.02	1300	70.02	1300	1300	1300	0.0000	0.0000	1300	3.554e+004
Liquid Volume Flow	m3/h	4.448	0.2256	4.222	0.2256	4.222	4.222	4.222	0.0000	0.0000	4.222	33.71
Heat Flow	kcal/h	-1.732e+006	-8.013e+004	-1.378e+006	-7.355e+004	-1.652e+006	-1.649e+006	-1.652e+006	0.0000	0.0000	-1.649e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	37.75	35.38	45.05	25.11	25.11	25.11	25.11	35.30	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	6.805	6.805	6.805	6.805	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	63.81	31.91	31.91	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	1090	545.0	545.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6820	3.540	1.770	1.770	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.013e+006	-9.608e+007	-9.280e+007	-2.227e+005	-1.156e+006	-5.779e+005	-5.779e+005	-9.307e+007	

Table C.3 – Material streams for case 3

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	28.22	30.66	-158.6	-154.4	-158.6	-158.6	-156.6	-154.4	45.00	45.05
Pressure	bar	1.250	1.250	7.041	1.213	1.250	7.730	1.250	1.250	7.450	7.450	1.000	6.100
Molar Flow	kgmole/h	35.91	1.903	31.91	1.903	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	612.1	31.28	545.0	31.28	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.987	0.1008	1.770	0.1008	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.737e+005	-3.580e+004	-5.770e+005	-3.286e+004	-7.379e+005	-7.358e+005	-7.379e+005	0.0000	0.0000	-7.358e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	41.79	41.00	45.05	28.22	28.22	28.22	28.22	40.98	-154.4	-154.4	
Pressure	bar	6.100	6.086	5.795	6.100	7.041	7.041	7.041	7.041	5.795	7.450	7.450	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	31.91	15.95	15.95	1196	2.096	31.91	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	545.0	272.5	272.5	3.442e+004	35.80	545.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	1.770	0.8850	0.8850	32.65	0.1163	1.770	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.592e+007	-9.280e+007	0.0000	-5.770e+005	-2.885e+005	-2.885e+005	-9.291e+007	-4.536e+004	-6.905e+005	

Table C.4 – Material streams for case 4

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	27.36	32.29	-158.6	-155.3	-158.6	-158.6	-156.6	-155.3	45.00
Pressure	bar	1.250	1.250	7.041	1.213	1.250	7.605	1.250	1.250	7.450	7.450	1.000
Molar Flow	kgmole/h	46.68	2.474	44.20	2.474	44.20	44.20	44.20	0.0000	0.0000	44.20	1235
Mass Flow	kg/h	795.7	40.67	755.0	40.67	755.0	755.0	755.0	0.0000	0.0000	755.0	3.554e+004
Liquid Volume Flow	m3/h	2.583	0.1310	2.452	0.1310	2.452	2.452	2.452	0.0000	0.0000	2.452	33.71
Heat Flow	kcal/h	-1.006e+006	-4.654e+004	-7.997e+005	-4.269e+004	-9.592e+005	-9.571e+005	-9.592e+005	0.0000	0.0000	-9.571e+005	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	40.78	39.44	45.05	27.36	27.36	27.36	27.36	39.40	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.041	7.041	7.041	7.041	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	31.91	15.95	15.95	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	545.0	272.5	272.5	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6820	1.770	0.8850	0.8850	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.010e+006	-9.597e+007	-9.280e+007	-2.224e+005	-5.772e+005	-2.886e+005	-2.886e+005	-9.296e+007	

Table C.5 – Material streams for case 5

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	29.80	31.96	-158.6	-154.1	-158.6	-158.6	-156.6	-154.1	45.00	45.05
Pressure	bar	1.250	1.250	7.041	1.213	1.250	8.122	1.250	1.250	7.450	7.450	1.000	6.100
Molar Flow	kgmole/h	35.91	1.903	8.782	1.903	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	612.1	31.28	150.0	31.28	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.987	0.1008	0.4872	0.1008	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.737e+005	-3.580e+004	-1.587e+005	-3.284e+004	-7.379e+005	-7.357e+005	-7.379e+005	0.0000	0.0000	-7.357e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	41.77	43.86	45.05	29.80	29.80	29.80	29.80	43.93	-154.1	-154.1	
Pressure	bar	6.100	6.086	5.795	6.100	7.041	7.041	7.041	7.041	5.795	7.450	7.450	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	8.782	8.782	0.0000	1196	25.22	8.782	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	150.0	150.0	0.0000	3.442e+004	430.8	150.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	0.4872	0.4872	0.0000	32.65	1.399	0.4872	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.584e+007	-9.280e+007	0.0000	-1.587e+005	-1.587e+005	0.0000	-9.283e+007	-5.457e+005	-1.900e+005	

Table C.6 – Material streams for case 6

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0530	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-158.6	-158.6	28.96	31.97	-158.6	-154.1	-158.6	-158.6	-156.6	-154.1	45.00	45.05
Pressure	bar	1.250	1.250	7.041	1.213	1.250	8.122	1.250	1.250	7.450	7.450	1.000	6.100
Molar Flow	kgmole/h	35.91	1.903	21.08	1.903	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	612.1	31.28	360.0	31.28	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.987	0.1008	1.169	0.1008	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.737e+005	-3.580e+004	-3.810e+005	-3.284e+004	-7.379e+005	-7.357e+005	-7.379e+005	0.0000	0.0000	-7.357e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_vap	LNG_rec	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	41.77	42.34	45.05	28.96	28.96	28.96	28.96	42.36	-154.1	-154.1	
Pressure	bar	6.100	6.086	5.795	6.100	7.041	7.041	7.041	7.041	5.795	7.450	7.450	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	12.29	8.782	8.782	0.0000	1196	21.08	12.93	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	210.0	150.0	150.0	0.0000	3.442e+004	360.0	220.8	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.6820	0.4872	0.4872	0.0000	32.65	1.169	0.7171	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.588e+007	-9.280e+007	-2.222e+005	-1.587e+005	-1.587e+005	0.0000	-9.288e+007	-4.560e+005	-2.797e+005	

Table C.7 – Material streams for case 7

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	26.37	32.54	-146.0	-144.2	-146.0	-146.0	-143.9	-144.2	45.00
Pressure	bar	3.000	3.000	7.186	2.963	3.000	7.595	3.000	3.000	9.200	7.595	1.000
Molar Flow	kgmole/h	67.30	3.500	63.80	3.500	63.80	63.80	63.80	0.0000	0.0000	63.80	1235
Mass Flow	kg/h	1147	57.28	1090	57.28	1090	1090	1090	0.0000	0.0000	1090	3.554e+004
Liquid Volume Flow	m3/h	3.725	0.1858	3.539	0.1858	3.539	3.539	3.539	0.0000	0.0000	3.539	33.71
Heat Flow	kcal/h	-1.439e+006	-6.597e+004	-1.154e+006	-6.080e+004	-1.373e+006	-1.371e+006	-1.373e+006	0.0000	0.0000	-1.371e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	39.31	37.32	45.05	26.37	26.37	26.37	26.37	37.25	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.186	7.186	7.186	7.186	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	51.51	25.76	25.76	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	880.0	440.0	440.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6818	2.857	1.429	1.429	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.011e+006	-9.603e+007	-9.280e+007	-2.224e+005	-9.320e+005	-4.660e+005	-4.660e+005	-9.302e+007	

Table C.8 – Material streams for case 8

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	25.54	31.87	-146.0	-144.4	-146.0	-146.0	-143.9	-144.4	45.00
Pressure	bar	3.000	3.000	7.053	2.963	3.000	7.462	3.000	3.000	9.200	7.462	1.000
Molar Flow	kgmole/h	80.27	4.174	76.09	4.174	76.09	76.09	76.09	0.0000	0.0000	76.09	1235
Mass Flow	kg/h	1368	68.31	1300	68.31	1300	1300	1300	0.0000	0.0000	1300	3.554e+004
Liquid Volume Flow	m3/h	4.442	0.2216	4.221	0.2216	4.221	4.221	4.221	0.0000	0.0000	4.221	33.71
Heat Flow	kcal/h	-1.716e+006	-7.868e+004	-1.377e+006	-7.254e+004	-1.638e+006	-1.636e+006	-1.638e+006	0.0000	0.0000	-1.636e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	38.23	35.83	45.05	25.54	25.54	25.54	25.54	35.76	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.053	7.053	7.053	7.053	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	63.80	31.90	31.90	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	1090	545.0	545.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6818	3.539	1.769	1.769	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.012e+006	-9.607e+007	-9.280e+007	-2.225e+005	-1.155e+006	-5.774e+005	-5.774e+005	-9.306e+007	

Table C.9 – Material streams for case 9

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	28.52	31.73	-146.0	-142.7	-146.0	-146.0	-143.9	-142.7	45.00	45.05
Pressure	bar	3.000	3.000	7.510	2.963	3.000	7.919	3.000	3.000	9.200	7.919	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	31.90	1.865	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	611.3	30.52	545.0	30.52	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.985	9.901e-002	1.769	9.901e-002	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.668e+005	-3.515e+004	-5.766e+005	-3.241e+004	-7.316e+005	-7.300e+005	-7.316e+005	0.0000	0.0000	-7.300e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	42.01	41.19	45.05	28.52	28.52	28.52	28.52	41.16	-142.7	-142.7	
Pressure	bar	6.100	6.086	5.795	6.100	7.510	7.510	7.510	7.510	5.795	7.919	7.919	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	31.90	15.95	15.95	1196	2.096	31.90	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	545.0	272.5	272.5	3.442e+004	35.80	545.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	1.769	0.8847	0.8847	32.65	0.1162	1.769	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.592e+007	-9.280e+007	0.0000	-5.766e+005	-2.883e+005	-2.883e+005	-9.291e+007	-4.500e+004	-6.850e+005	

Table C.10 – Material streams for case 10

Material Streams												
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	27.70	32.51	-146.0	-143.4	-146.0	-146.0	-143.9	-143.4	45.00
Pressure	bar	3.000	3.000	7.399	2.963	3.000	7.808	3.000	3.000	9.200	7.808	1.000
Molar Flow	kgmole/h	46.62	2.424	44.19	2.424	44.19	44.19	44.19	0.0000	0.0000	44.19	1235
Mass Flow	kg/h	794.7	39.67	755.0	39.67	755.0	755.0	755.0	0.0000	0.0000	755.0	3.554e+004
Liquid Volume Flow	m3/h	2.580	0.1287	2.451	0.1287	2.451	2.451	2.451	0.0000	0.0000	2.451	33.71
Heat Flow	kcal/h	-9.967e+005	-4.569e+004	-7.992e+005	-4.211e+004	-9.510e+005	-9.494e+005	-9.510e+005	0.0000	0.0000	-9.494e+005	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	41.08	39.70	45.05	27.70	27.70	27.70	27.70	39.66	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.399	7.399	7.399	7.399	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	12.29	31.90	15.95	15.95	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	210.0	545.0	272.5	272.5	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.6818	1.769	0.8847	0.8847	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.010e+006	-9.596e+007	-9.280e+007	-2.223e+005	-5.769e+005	-2.884e+005	-2.884e+005	-9.295e+007	

Table C.11 – Material streams for case 11

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	29.95	32.95	-146.0	-143.4	-146.0	-146.0	-143.9	-143.4	45.00	45.05
Pressure	bar	3.000	3.000	6.490	2.963	3.000	6.899	3.000	3.000	9.200	6.899	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	8.780	1.865	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	611.3	30.52	150.0	30.52	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.985	9.901e-002	0.4870	9.901e-002	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.668e+005	-3.515e+004	-1.586e+005	-3.239e+004	-7.316e+005	-7.303e+005	-7.316e+005	0.0000	0.0000	-7.303e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	41.99	43.91	45.05	29.95	29.95	29.95	29.95	43.98	-143.4	-143.4	
Pressure	bar	6.100	6.086	5.795	6.100	6.490	6.490	6.490	6.490	5.795	6.899	6.899	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	8.780	8.780	0.0000	1196	25.22	8.780	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	150.0	150.0	0.0000	3.442e+004	430.8	150.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	0.4870	0.4870	0.0000	32.65	1.399	0.4870	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.584e+007	-9.280e+007	0.0000	-1.586e+005	-1.586e+005	0.0000	-9.283e+007	-5.417e+005	-1.886e+005	

Table C.12 – Material streams for case 12

		Material Streams											
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-146.0	-146.0	29.23	32.95	-146.0	-142.7	-146.0	-146.0	-143.9	-142.7	45.00	45.05
Pressure	bar	3.000	3.000	7.510	2.963	3.000	7.919	3.000	3.000	9.200	7.919	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	21.07	1.865	34.00	34.00	34.00	0.0000	0.0000	34.00	1235	1235
Mass Flow	kg/h	611.3	30.52	360.0	30.52	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.985	9.901e-002	1.169	9.901e-002	1.886	1.886	1.886	0.0000	0.0000	1.886	33.71	33.71
Heat Flow	kcal/h	-7.668e+005	-3.515e+004	-3.808e+005	-3.239e+004	-7.316e+005	-7.300e+005	-7.316e+005	0.0000	0.0000	-7.300e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_vap	LNG_rec	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	41.99	42.47	45.05	29.23	29.23	29.23	29.23	42.48	-142.7	-142.7	
Pressure	bar	6.100	6.086	5.795	6.100	7.510	7.510	7.510	7.510	5.795	7.919	7.919	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	12.29	8.780	8.780	0.0000	1196	21.07	12.92	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	210.0	150.0	150.0	0.0000	3.442e+004	360.0	220.8	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.6818	0.4870	0.4870	0.0000	32.65	1.169	0.7169	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.588e+007	-9.280e+007	-2.221e+005	-1.587e+005	-1.587e+005	0.0000	-9.287e+007	-4.525e+005	-2.775e+005	

Table C.13 – Material streams for case 13

		Material Streams										
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	26.72	32.65	-137.2	-135.9	-137.2	-137.2	-135.1	-135.9	45.00
Pressure	bar	5.000	5.000	7.835	4.963	5.000	8.244	5.000	5.000	11.20	8.244	1.000
Molar Flow	kgmole/h	67.29	3.499	63.79	3.499	63.79	63.79	63.79	0.0000	0.0000	63.79	1235
Mass Flow	kg/h	1147	57.11	1090	57.11	1090	1090	1090	0.0000	0.0000	1090	3.554e+004
Liquid Volume Flow	m3/h	3.724	0.1861	3.538	0.1861	3.538	3.538	3.538	0.0000	0.0000	3.538	33.71
Heat Flow	kcal/h	-1.431e+006	-6.606e+004	-1.154e+006	-6.107e+004	-1.365e+006	-1.363e+006	-1.365e+006	0.0000	0.0000	-1.363e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	39.51	37.59	45.05	26.72	26.72	26.72	26.72	37.53	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.835	7.835	7.835	7.835	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	0.0000	63.79	31.90	31.90	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	0.0000	1090	545.0	545.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.0000	3.538	1.769	1.769	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.011e+006	-9.602e+007	-9.280e+007	0.0000	-1.154e+006	-5.770e+005	-5.770e+005	-9.301e+007	

Table C.14 – Material streams for case 14

		Material Streams										
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	26.72	32.65	-137.2	-135.9	-137.2	-137.2	-135.1	-135.9	45.00
Pressure	bar	5.000	5.000	7.835	4.963	5.000	8.244	5.000	5.000	11.20	8.244	1.000
Molar Flow	kgmole/h	67.29	3.499	63.79	3.499	63.79	63.79	63.79	0.0000	0.0000	63.79	1235
Mass Flow	kg/h	1147	57.11	1090	57.11	1090	1090	1090	0.0000	0.0000	1090	3.554e+004
Liquid Volume Flow	m3/h	3.724	0.1861	3.538	0.1861	3.538	3.538	3.538	0.0000	0.0000	3.538	33.71
Heat Flow	kcal/h	-1.431e+006	-6.606e+004	-1.154e+006	-6.107e+004	-1.365e+006	-1.363e+006	-1.365e+006	0.0000	0.0000	-1.363e+006	-9.581e+007
		ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	
Temperature	C	45.05	45.05	39.51	37.59	45.05	26.72	26.72	26.72	26.72	37.53	
Pressure	bar	6.100	6.100	6.086	5.795	6.100	7.835	7.835	7.835	7.835	5.795	
Molar Flow	kgmole/h	1235	38.75	38.75	1235	1196	0.0000	63.79	31.90	31.90	1196	
Mass Flow	kg/h	3.554e+004	1115	1115	3.554e+004	3.442e+004	0.0000	1090	545.0	545.0	3.442e+004	
Liquid Volume Flow	m3/h	33.71	1.058	1.058	33.71	32.65	0.0000	3.538	1.769	1.769	32.65	
Heat Flow	kcal/h	-9.581e+007	-3.006e+006	-3.011e+006	-9.602e+007	-9.280e+007	0.0000	-1.154e+006	-5.770e+005	-5.770e+005	-9.301e+007	

Table C.15 – Material streams for case 15

		Material Streams											
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	28.79	31.83	-137.2	-134.9	-137.2	-137.2	-135.1	-134.9	45.00	45.05
Pressure	bar	5.000	5.000	8.073	4.963	5.000	8.482	5.000	5.000	11.20	8.482	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	31.90	1.865	33.99	33.99	33.99	0.0000	0.0000	33.99	1235	1235
Mass Flow	kg/h	611.2	30.43	545.0	30.43	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.984	9.915e-002	1.769	9.915e-002	1.885	1.885	1.885	0.0000	0.0000	1.885	33.71	33.71
Heat Flow	kcal/h	-7.623e+005	-3.520e+004	-5.764e+005	-3.255e+004	-7.271e+005	-7.259e+005	-7.271e+005	0.0000	0.0000	-7.259e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	42.11	41.32	45.05	28.79	28.79	28.79	28.79	41.29	-134.9	-134.9	
Pressure	bar	6.100	6.086	5.795	6.100	8.073	8.073	8.073	8.073	5.795	8.482	8.482	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	31.90	15.95	15.95	1196	2.095	31.90	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	545.0	272.5	272.5	3.442e+004	35.80	545.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	1.769	0.8845	0.8845	32.65	0.1162	1.769	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.591e+007	-9.280e+007	0.0000	-5.764e+005	-2.882e+005	-2.882e+005	-9.291e+007	-4.474e+004	-6.812e+005	

Table C.16 – Material streams for case 16

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	28.79	31.83	-137.2	-134.9	-137.2	-137.2	-135.1	-134.9	45.00	45.05
Pressure	bar	5.000	5.000	8.073	4.963	5.000	8.482	5.000	5.000	11.20	8.482	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	31.90	1.865	33.99	33.99	33.99	0.0000	0.0000	33.99	1235	1235
Mass Flow	kg/h	611.2	30.43	545.0	30.43	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.984	9.915e-002	1.769	9.915e-002	1.885	1.885	1.885	0.0000	0.0000	1.885	33.71	33.71
Heat Flow	kcal/h	-7.623e+005	-3.520e+004	-5.764e+005	-3.255e+004	-7.271e+005	-7.259e+005	-7.271e+005	0.0000	0.0000	-7.259e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	42.11	41.32	45.05	28.79	28.79	28.79	28.79	41.29	-134.9	-134.9	
Pressure	bar	6.100	6.086	5.795	6.100	8.073	8.073	8.073	8.073	5.795	8.482	8.482	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	31.90	15.95	15.95	1196	2.095	31.90	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	545.0	272.5	272.5	3.442e+004	35.80	545.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	1.769	0.8845	0.8845	32.65	0.1162	1.769	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.591e+007	-9.280e+007	0.0000	-5.764e+005	-2.882e+005	-2.882e+005	-9.291e+007	-4.474e+004	-6.812e+005	

Table C.17 – Material streams for case 17

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	30.27	33.03	-137.2	-134.7	-137.2	-137.2	-135.1	-134.7	45.00	45.05
Pressure	bar	5.000	5.000	8.328	4.963	5.000	8.737	5.000	5.000	11.20	8.737	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	8.779	1.865	33.99	33.99	33.99	0.0000	0.0000	33.99	1235	1235
Mass Flow	kg/h	611.2	30.43	150.0	30.43	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.984	9.915e-002	0.4869	9.915e-002	1.885	1.885	1.885	0.0000	0.0000	1.885	33.71	33.71
Heat Flow	kcal/h	-7.623e+005	-3.520e+004	-1.585e+005	-3.253e+004	-7.271e+005	-7.258e+005	-7.271e+005	0.0000	0.0000	-7.258e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	42.09	43.96	45.05	30.27	30.27	30.27	30.27	44.02	-134.7	-134.7	
Pressure	bar	6.100	6.086	5.795	6.100	8.328	8.328	8.328	8.328	5.795	8.737	8.737	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	8.779	8.779	0.0000	1196	25.21	8.779	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	150.0	150.0	0.0000	3.442e+004	430.8	150.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	0.4869	0.4869	0.0000	32.65	1.398	0.4869	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.584e+007	-9.280e+007	0.0000	-1.585e+005	-1.585e+005	0.0000	-9.283e+007	-5.384e+005	-1.875e+005	

Table C.18 – Material streams for case 18

Material Streams													
		FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Vapour Fraction		0.0520	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	C	-137.2	-137.2	30.27	33.03	-137.2	-134.7	-137.2	-137.2	-135.1	-134.7	45.00	45.05
Pressure	bar	5.000	5.000	8.328	4.963	5.000	8.737	5.000	5.000	11.20	8.737	1.000	6.100
Molar Flow	kgmole/h	35.86	1.865	8.779	1.865	33.99	33.99	33.99	0.0000	0.0000	33.99	1235	1235
Mass Flow	kg/h	611.2	30.43	150.0	30.43	580.8	580.8	580.8	0.0000	0.0000	580.8	3.554e+004	3.554e+004
Liquid Volume Flow	m3/h	1.984	9.915e-002	0.4869	9.915e-002	1.885	1.885	1.885	0.0000	0.0000	1.885	33.71	33.71
Heat Flow	kcal/h	-7.623e+005	-3.520e+004	-1.585e+005	-3.253e+004	-7.271e+005	-7.258e+005	-7.271e+005	0.0000	0.0000	-7.258e+005	-9.581e+007	-9.581e+007
		ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	0.0000	
Temperature	C	45.05	42.09	43.96	45.05	30.27	30.27	30.27	30.27	44.02	-134.7	-134.7	
Pressure	bar	6.100	6.086	5.795	6.100	8.328	8.328	8.328	8.328	5.795	8.737	8.737	
Molar Flow	kgmole/h	38.75	38.75	1235	1196	0.0000	8.779	8.779	0.0000	1196	25.21	8.779	
Mass Flow	kg/h	1115	1115	3.554e+004	3.442e+004	0.0000	150.0	150.0	0.0000	3.442e+004	430.8	150.0	
Liquid Volume Flow	m3/h	1.058	1.058	33.71	32.65	0.0000	0.4869	0.4869	0.0000	32.65	1.398	0.4869	
Heat Flow	kcal/h	-3.006e+006	-3.009e+006	-9.584e+007	-9.280e+007	0.0000	-1.585e+005	-1.585e+005	0.0000	-9.283e+007	-5.384e+005	-1.875e+005	

Appendix D. Compositions of the streams for the stationary case studies

Tables D.1 to D.18 summarize the complete composition results for all the stationary state simulations that were studied, being all the tables retrieved from the software Aspen HYSYS®. The nomenclature used is the enunciated in Figure 3.3 of Chapter 3.3.

Table D.1 – Compositions for case 1

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.2 – Compositions for case 2

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.3 – Compositions for case 3

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	0.9385	0.9385	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	0.0013	0.0013	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.4 – Compositions for case 4

Compositions											
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.5 – Compositions for case 5

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	0.9385	0.9385	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	0.0013	0.0013	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.6 – Compositions for case 6

	Compositions											
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9670	0.9385	0.9670	0.9385	0.9385	0.9385	0.9385	0.9385	0.9385	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0001	0.0496	0.0001	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0329	0.0013	0.0329	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_vap	LNG_rec	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9385	0.9385	0.9385	0.9385	0.0000	0.9385	0.9385	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0000	0.0013	0.0013	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.7 – Compositions for case 7

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.8 – Compositions for case 8

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.9 – Compositions for case 9

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	0.9382	0.9382	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	0.0017	0.0017	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.10 – Compositions for case 10

Compositions											
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.11 – Compositions for case 11

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	0.9382	0.9382	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	0.0017	0.0017	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.12 – Compositions for case 12

	Compositions											
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9731	0.9382	0.9731	0.9382	0.9382	0.9382	0.9382	0.9382	0.9382	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0003	0.0496	0.0003	0.0496	0.0496	0.0496	0.0496	0.0496	0.0496	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0266	0.0017	0.0266	0.0017	0.0017	0.0017	0.0017	0.0017	0.0017	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_vap	LNG_rec	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9382	0.9382	0.9382	0.9382	0.0000	0.9382	0.9382	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0496	0.0496	0.0496	0.0496	0.0000	0.0496	0.0496	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0017	0.0017	0.0017	0.0017	0.0000	0.0017	0.0017	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.13 – Compositions for case 13

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.14 – Compositions for case 14

	Compositions										
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442
	ETGp	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	

Table D.15 – Compositions for case 15

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	0.9380	0.9380	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	0.0495	0.0495	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	0.0019	0.0019	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.16 – Compositions for case 16

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	0.9380	0.9380	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	0.0495	0.0495	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	0.0019	0.0019	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.17 – Compositions for case 17

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	0.9380	0.9380	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	0.0495	0.0495	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	0.0019	0.0019	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Table D.18 – Compositions for case 18

Compositions												
	FEED	BOG	LNGh	BOG_boiler	LNG	LNG1p	LNG1	LNG2	LNG2p	LNGp	ETGtotal	ETGp
Comp Mole Frac (Methane)	0.9400	0.9769	0.9380	0.9769	0.9380	0.9380	0.9380	0.9380	0.9380	0.9380	0.0000	0.0000
Comp Mole Frac (Ethane)	0.0470	0.0005	0.0495	0.0005	0.0495	0.0495	0.0495	0.0495	0.0495	0.0495	0.0000	0.0000
Comp Mole Frac (Propane)	0.0080	0.0000	0.0084	0.0000	0.0084	0.0084	0.0084	0.0084	0.0084	0.0084	0.0000	0.0000
Comp Mole Frac (n-Butane)	0.0020	0.0000	0.0021	0.0000	0.0021	0.0021	0.0021	0.0021	0.0021	0.0021	0.0000	0.0000
Comp Mole Frac (Nitrogen)	0.0030	0.0226	0.0019	0.0226	0.0019	0.0019	0.0019	0.0019	0.0019	0.0019	0.0000	0.0000
Comp Mole Frac (H2O)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7558	0.7558
Comp Mole Frac (EGlycol)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2442	0.2442
	ETGbog	ETGbog_c	ETG_out	ETGlng	LNG_boiler	LNG_etotal	LNG_e1	LNG_e2	ETG_lngc	LNG_rec	LNG_vap	
Comp Mole Frac (Methane)	0.0000	0.0000	0.0000	0.0000	0.9380	0.9380	0.9380	0.9380	0.0000	0.9380	0.9380	
Comp Mole Frac (Ethane)	0.0000	0.0000	0.0000	0.0000	0.0495	0.0495	0.0495	0.0495	0.0000	0.0495	0.0495	
Comp Mole Frac (Propane)	0.0000	0.0000	0.0000	0.0000	0.0084	0.0084	0.0084	0.0084	0.0000	0.0084	0.0084	
Comp Mole Frac (n-Butane)	0.0000	0.0000	0.0000	0.0000	0.0021	0.0021	0.0021	0.0021	0.0000	0.0021	0.0021	
Comp Mole Frac (Nitrogen)	0.0000	0.0000	0.0000	0.0000	0.0019	0.0019	0.0019	0.0019	0.0000	0.0019	0.0019	
Comp Mole Frac (H2O)	0.7558	0.7558	0.7558	0.7558	0.0000	0.0000	0.0000	0.0000	0.7558	0.0000	0.0000	
Comp Mole Frac (EGlycol)	0.2442	0.2442	0.2442	0.2442	0.0000	0.0000	0.0000	0.0000	0.2442	0.0000	0.0000	

Appendix E. Energy streams for the stationary case studies

Tables E.1 to E.18 summarize the complete results of the energy streams for all the stationary state simulations that were studied. The nomenclature used is the enunciated in Figure 3.3 of Chapter 3.3.

Table E.1 - Energy streams for case 1

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2144	0.0000	5473

Table E.2 - Energy streams for case 2

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2311	0.0000	5473

Table E.3 - Energy streams for case 3

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2052	0.0000	5473

Table E.4 - Energy streams for case 4

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2106	0.0000	5473

Table E.5 - Energy streams for case 5

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2176	0.0000	5473

Table E.6 - Energy streams for case 6

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	2176	0.0000	5473

Table E.7 - Energy streams for case 7

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1684	0.0000	5473

Table E.8 - Energy streams for case 8

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1806	0.0000	5473

Table E.9 - Energy streams for case 9

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1628	0.0000	5473

Table E.10 - Energy streams for case 10

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1665	0.0000	5473

Table E.11 - Energy streams for case 11

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1290	0.0000	5473

Table E.12 - Energy streams for case 12

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1628	0.0000	5473

Table E.13 - Energy streams for case 13

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1230	0.0000	5473

Table E.14 - Energy streams for case 14

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1230	0.0000	5473

Table E.15 - Energy streams for case 15

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1192	0.0000	5473

Table E.16 - Energy streams for case 16

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1192	0.0000	5473

Table E.17 - Energy streams for case 17

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1279	0.0000	5473

Table E.18 - Energy streams for case 18

Energy Streams				
		Q-100	Q-102	Q-101
Heat Flow	kcal/h	1279	0.0000	5473

Appendix F. Full data used in P-H diagram for a mixture of oxygen and argon.

Table F.1 – P-H diagram data.

T (°C)	H_V (kJ·kg^{-1})	H_L (kJ·kg^{-1})	P_V (kPa)	P_L (kPa)	ΔH_V (kJ·kg^{-1})
-200	-130.9	-307.5	16.38	16.66	176.6
-195	-128.3	-302.0	33.55	33.98	173.7
-190	-125.8	-296.5	62.65	63.27	170.7
-185	-123.5	-290.8	108.5	109.4	167.3
-180	-121.4	-285.1	176.7	177.8	163.7
-175	-119.7	-279.1	273.3	274.6	159.4
-170	-118.2	-272.9	404.8	406.4	154.7
-165	-117.2	-266.5	578.1	579.9	149.3
-160	-116.6	-259.8	800.4	802.2	143.2
-155	-116.6	-252.7	1079	1081	136.1
-150	-117.2	-245.1	1421	1423	127.9
-145	-118.6	-236.9	1835	1837	118.3
-140	-121.0	-227.9	2328	2330	106.9
-135	-124.9	-217.6	2909	2910	92.7
-130	-131.1	-205.2	3586	3587	74.1
-125	-141.9	-188.4	4368	4369	46.5
-123	-150.4	-177.6	4713	4713	27.2