QATAR UNIVERSITY

COLLEGE OF ENGINEERING

THERMODYNAMIC ANALYSIS OF FULL SCALE BASELOAD LNG CHAIN

BY

ZINEB BOUABIDI

A Thesis Submitted to

the Faculty of the College of Engineering

in Partial Fulfillment of the Requirements for the Degree of

Masters of Science in Environmental Engineering

June 2019

COMMITTEE PAGE

The members of the Committee approve the Thesis of
Zineb Bouabidi defended on 21/04/2019.

_____
Easa I. Al-musleh
Thesis/Dissertation Supervisor

_____
Ramazan Kahraman
Thesis/Dissertation Supervisor

_____
Iftikhar Karimi
Committee Member

_____
Mohammed Ali H. Salah
Committee Member

Approved:

_____
Abdel Magid Hamouda , Dean, College of Engineering

ABSTRACT

BOAUBIDI, ZINEB BOUAZZA, Masters : June : 2019 Master of Science in Environmental Engineering

Title: Thermodynamic Analysis of Full Scale Baseload LNG Chain

Supervisor of Thesis: Easa I. Al-musleh.

: Ramazan Kahraman.

The world's growing energy demand and the need for low-carbon energy sources are key factors that have made natural gas (NG) an attractive energy source compared to other available fossil fuels (i.e. coal and oil). Being the most feasible NG transportation method over long distances, the liquefied NG (LNG) demand is significantly increasing. The LNG supply chain, consisting of production, liquefaction, shipping, and regasification, is, however, an energy-intensive and thereby emission intensive process. Therefore, the appropriate LNG production with least energy consumption and maximum energy efficiency is of high importance. Thus, optimization of LNG chains is essential from both economic and sustainability point of view. Amongst energy efficiency optimization approaches, exergy analysis, based on the second law of thermodynamics, is a powerful tool that has been widely used to quantify exergy destructions and to determine exergy efficiencies and thereby, identify process improvement opportunities.

In this thesis work, rigorous and detailed exergy analysis was performed on an entire baseload LNG chain that was simulated using ProMax® and Aspen Plus® simulation software for the delivery of 439 million standard cubic feet per day

(MMSCFD). A comparison of the losses across the various units with and without utilities was performed, and optimization opportunities within the chain were identified.

Findings of this study revealed that the LNG chain under consideration is associated with total loss of near 647 MW and 1054 MW during holding and loading operation modes, respectively. The main contributor to the exergy loss was found to be the utility section accounting for 61% of the total exergy loss. Within the LNG process, significant amounts of losses were found to occur in the sulfur recovery units, liquefaction unit, and sweetening processes; accounting for 38%, 30% and 24% of the total exergy loss, respectively. The compressors and their drivers (GTs), stream generators, LNG flashing and storage, columns (absorbers, distillations) and heat exchangers were found to be the main exergy consumers.

DEDICATION

Dedicated to my inspiring parents and sisters for their endless support and encouragement

ACKNOWLEDGMENTS

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

CHAPTER 1: INTRODUCTION

1.1 Introduction

This chapter defines the research problem, the objective of the thesis and briefly reviews the main aspects related to the research topic (i.e. NG, the market share, liquefied NG (LNG), and exergy and its importance), then describes the methodology and limitations and concludes with a thesis outline.

1.2 Motivation

With the predicted growth in the global energy demand that is driven by the economic growth and population increase and the continuous argument concerning the adverse environmental impact its consumptions brings along, NG is gaining a privileged position among other fuels as it is recognized as an environmentally friendly fuel [1-4]. The most feasible way to transport large quantities of NG over long distances, from producing regions to consumption regions, is via liquefaction at -161℃. This is owing to the fact that the liquefied NG (LNG) occupies about $(1/600)^{th}$ of the volume of NG in its gaseous state making it easier and cost efficient for shipment and storage. The LNG supply chain is, however, an energy-intensive process requiring tremendous amounts of energy in various forms (e.g. heating, cooling, and compression). Energy consumption and environmental impact are proportionally related. This means the higher the energy consumption is, the higher the greenhouse gas emissions (GHG), CO_2 in particular, are that makes energy consumption a major cause of global warming. Consequently, reducing energy consumption within the LNG chains is important in order to reduce its negative environmental impact.

Improving the performance and optimizing the energy efficiency of LNG production processes have been a popular subject in the academic society in the past

decades [5-11]. Among a large number of publications that deal with LNG chain, the majority are concerned with the design, operation, and optimization of the liquefaction process itself due to the fact that it is a major economic factor in the chain [6, 11-23]. Based on a thorough review, it was also found that the application of exergy analysis was limited to specific segments of the LNG chain only [24-30]. To the best of our knowledge, there is no comprehensive study that assesses a full-scale baseload LNG chain using exergy analysis approach. Exergy analysis is an important tool to evaluate the thermodynamic efficiency of various processes and it has been extensively considered as useful methods in the design, evaluation, optimization, and improvement of processing plants [31, 32].

Therefore, the ultimate objective of this thesis is to carry out rigorous exergy analysis for an actual baseload LNG chain, from the wellhead to the end users that delivers 439 MMSCFD of NG while taking into consideration the tanks/ships holding and loading operating modes. The analysis helped in identifying critical components that should be optimized to enhance chain performance.

1.3   NG and LNG chain backgrounds

With the current continuous population growth and economic development, the global energy demand is projected to grow persistently with near 1.3% per annum over the next decade [33]. While a wide variety of energy sources are available to support the different industrial sectors (oil, coal, biomass, nuclear, renewables, and NG), NG seems to have the highest growing demand share due to its availability and environmentally friendly characteristics.

Referring to Figure 1.1, NG, composed primarily of methane, is the cleanest burning fossil fuel, as its combustion to produce a given amount of energy produces the least amount of (CO_2) and other pollutants. For example, it has been estimated that

power generation plants combusting NG emits less than half and third the GHG emissions that would be produced coal and oil power plants, respectively [34]. NG also provides a sizable reduction in emissions when used in a variety of industrial applications. For example, it was reported in [35], that NG fired industrial boilers emits 20-35% less CO_2 compared to coal and oil boilers. For transportation, NG powered vehicles produces 22.5% and 7% less GHG emissions compared to gasoline and diesel engines, respectively [36].
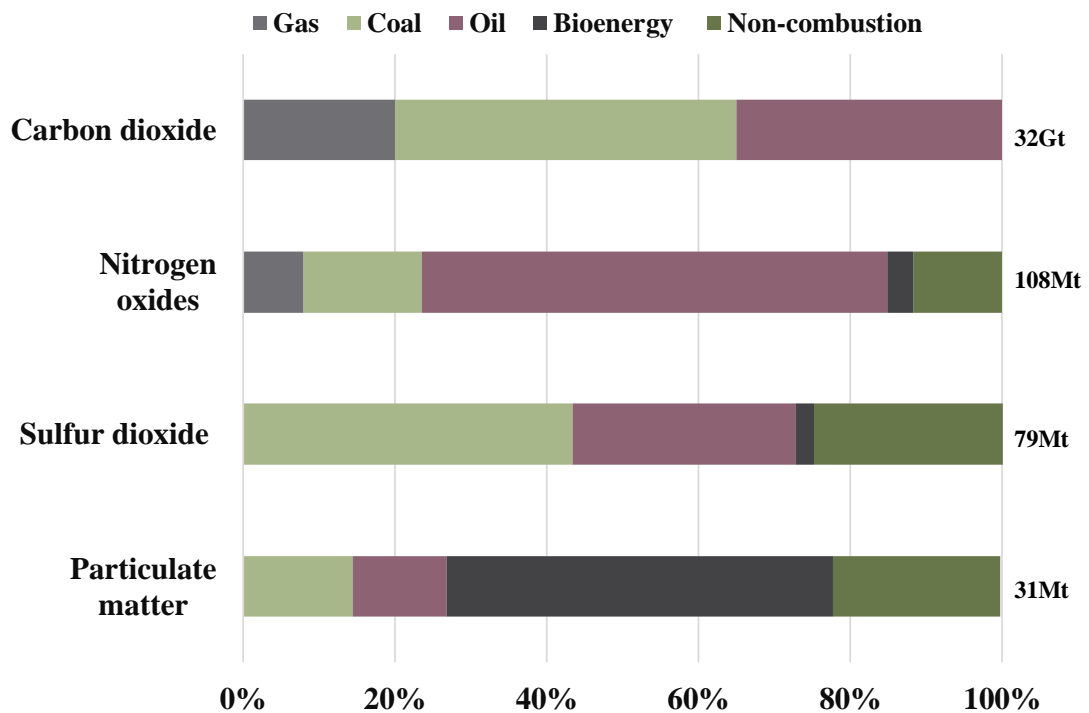
Figure 1. 1: Share of NG in total energy-related emissions on major air pollutants and CO_2, adapted from [34].

As a result, NG is gaining a privileged position among other fossil fuels, which may explain the continuous increase of its consumption. Figure 1.2 illustrates the

historical and projected electricity generation from selected fuels. The figure shows that electricity generation from both coal and nuclear energy is expected to decline, from 28% and 19% of the share of total generation in 2018 to 17% and 12% in 2050, respectively. On the other hand, the share of renewable generation and NG are expected to rise from 18 and 34% in 2018 to 31 and 39% in 2050, respectively.

Moving to Figure 1.3, NG consumption in the industrial and utility sectors, including the chemical industry (where NG is used as a feedstock), are expected to continuously increase. On the other hand, the rise in NG consumption in other sectors is insignificant.
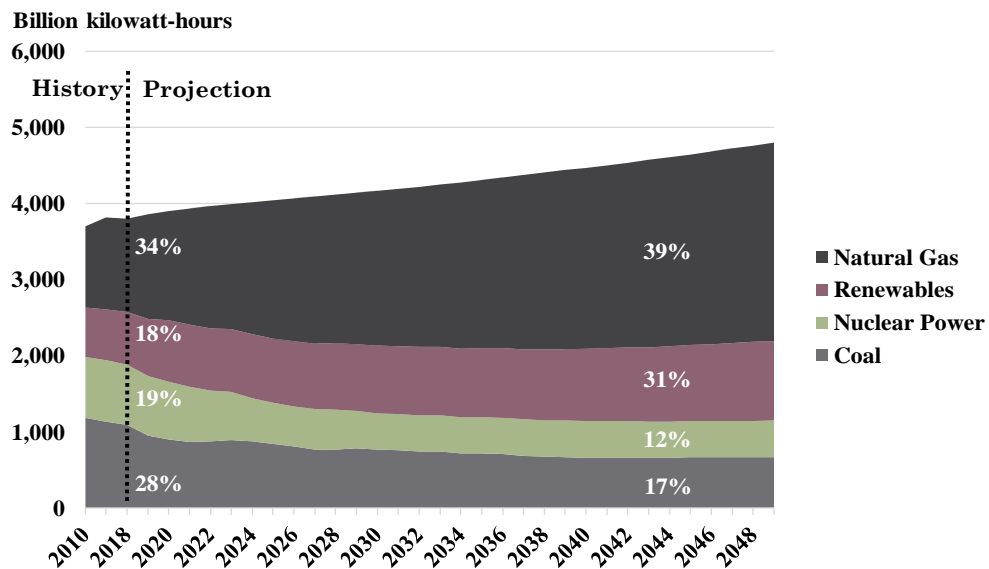
Figure 1. 2: Electricity generation from selected fuels, adapted from [37].
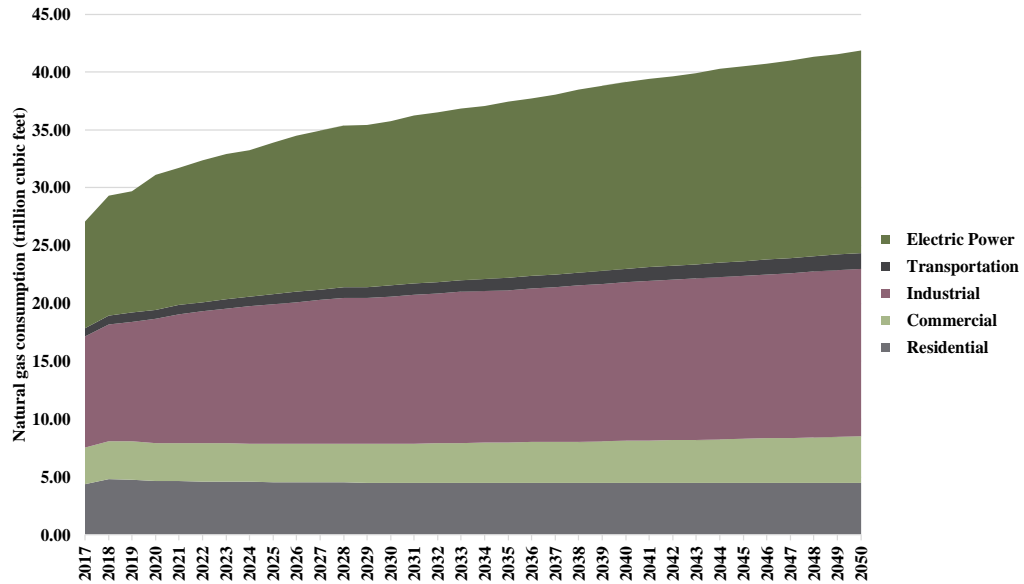
Figure 1. 3: NG consumption by sector, adapted from [37]

Supported by the growing NG demand, LNG trade is continuously increasing as it provides an economic mean of commercial transportation of NG over long distances owing to the fact that the volume of NG shrinks down 600 times when liquefied. According to ExxonMobil [38], liquefied NG trade supplies approximately one-third of NG demand from 2016 to 2040. About 32% of the global NG, corresponding to 346.6 billion cubic meters of NG (bcm) is moved as LNG while remaining is transported via pipelines in its gaseous state[39]. According to the future scenarios presented by the British Petroleum Company, LNG trade will account for approximately 15% of the globally traded NG in 2035, with 3.9% increase in the annual LNG trade[40]. In the same context, referring to Figure 1. 4, LNG supply form existing and under construction, facilities are expected to peak in the early 2020s and decline

thereafter due to expected declines in fields production. However, over 2030-2035, LNG demand growth is forecasted to accelerate as declining from existing suppliers which implies the need to construct new LNG plants to bridge the gap.
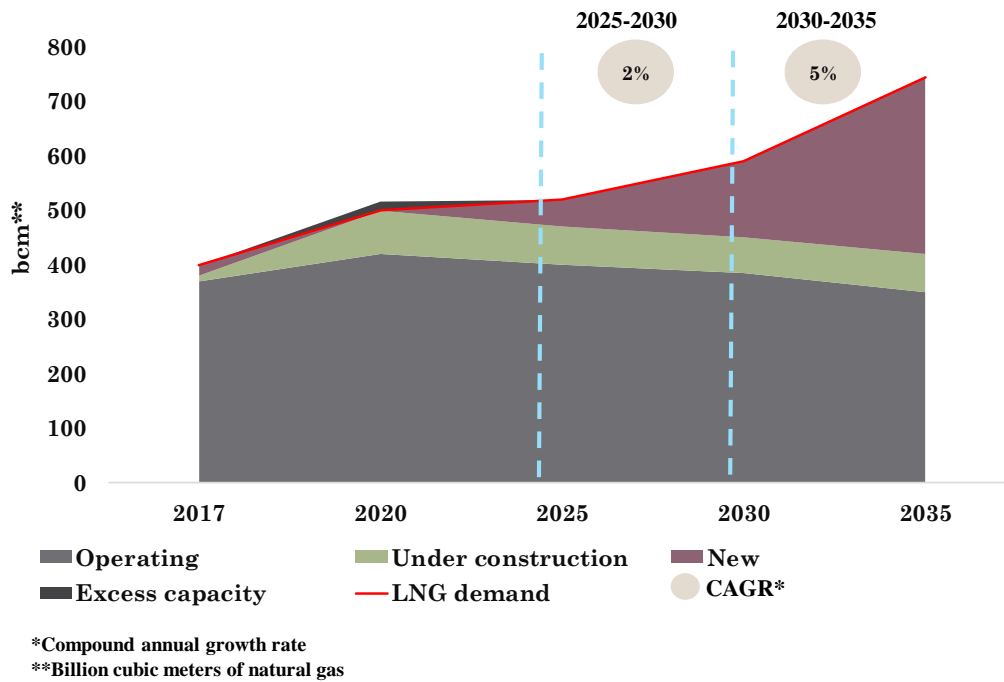
Figure 1. 4: Global LNG capacity and demand [38].

1.4 LNG chains and the environment

A typical LNG baseload supply chain is shown in Figure 1.5, can be divided into the following main segments: production wellhead, NG processing and liquefaction, LNG storage, the shipping, and LNG regasification in receiving terminal. In the second segment, where main processing takes place, NG from wellhead passes through a series of pretreatment stages (gas sweetening, mercury removal, and drying) to remove impurities such as CO_2, H_2S, mercury, and water. These are then followed

with the liquefaction process based on refrigeration means and separation of the sweet NG liquids (NGLs). Each of the chains is a made of a complicated process that consists of several integrated units consuming a considerable amount of energy in different forms. For example, liquefaction of NG is typically made of two integrated vapor compression cycles that typically require about 500 kWh electric energy per ton of LNG [41].
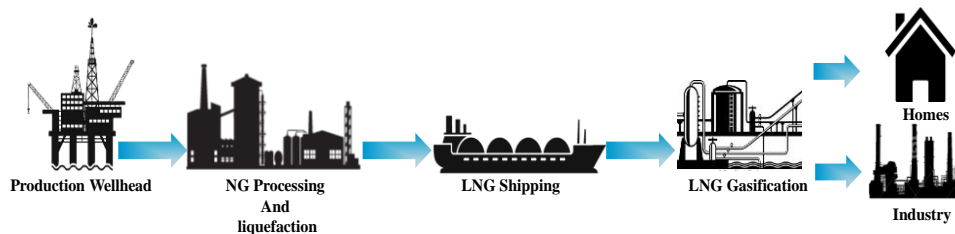
Figure 1. 5: LNG supply chain.

Thus, while NG is an environmentally friendly fuel, the substantial energy consumption (e.g. heat, steam, electricity, shaft work, or cooling water) in LNG production can have a significant impact on the environment mainly due to GHG emissions, primarily CO_2. For example, the average CO_2 emission for a liquefaction plant was estimated as 0.4 million metric ton (MMT) of CO_{2equ} (CO_2 equivalent emissions) per MTA of LNG [42]. In the U.S, methane emissions from LNG operations were estimated to be near 1.9 MMT CO_{2equ} in 2011[43]. Thereby, improving the energy efficiency of existing LNG plants and the development of more efficient systems are deemed to be essential given the expected growth of the LNG demand.

Considerable efforts have been dedicated to improving the energy efficiency of the LNG facilities using different techniques such as mathematical optimization and

process integration [9, 44-48]. While mathematical optimization is deemed to be effective for process enhancements, it can be a challenging task owing to the complexity of LNG systems. For example, a baseload LNG plant will typically have more than 30 operating variables available for optimization. Thermodynamic analysis can simplify this task, by identifying the most essential units that must be optimized to enhance the overall performance of the chain.

Analysis based on the first law of thermodynamics (i.e. energy and mass balance) is limited to energy quantity, not quality, where the latter represents how energy is degraded in the process. On the other hand, the second law of thermodynamics, entropy balance, takes into consideration the loss in energy quality by quantifying entropy generation arising from irreversibilities in a process. However, the entropy generation quantity does not provide an interpretation of the significance of the magnitude of degraded energy quality. Thus, it is more adequate to measure the lost work potential, exergy loss, owing to irreversibility, where this quantity is the product of entropy generation rate and environmental temperature. Consequently, the exergy and exergy analysis is practical representatives of entropy generation and entropy balance, respectively. Such an approach could be used either in combination with optimization to reduce the search space for optimization or stand-alone as pinpointing tools for performance evaluation.

1.5   Research methodology

The LNG supply chain simulation reported in[49], described in details in chapter 4, utilized as a basis for this research. Results of the simulation provided the characteristics of equipment and streams in the chain. Which are needed to carry out the rigorous exergy analysis to quantify the losses across the system. Thus, identifying the main units that should be optimized to achieve overall economic and environmental

benefits. Exergy loss across each equipment in the flowsheet was calculated using the steady-state exergy balance relation (see chapter 2), taking into consideration the three forms of exergy transfer (to or from the system) which are: exergy transfer by material flow, heat, and work. Taking into consideration all useful material streams out of the different processing units as valuable streams, exergy efficiency of each unit and the overall process was calculated.

1.6   Limitations

While exergy analysis pinpoints the processes and components with highest irreversibilities, provides some insight to where efforts should be focused to enhance the process, and where improvements should be allocated, it does not state whether the possible improvements are practical and feasible. This is owing to the fact that the exergy destruction in some of the plant components is unavoidable and limited by physical, economic and technological constraints. Advanced exergy analysis would be required to overcome these limitations as it takes into account splitting the exergy destruction in a component into exogenous and endogenous, and thus providing more insights about components interactions of the overall system. Advanced exergy analysis also takes into consideration splitting the exergy destruction into avoidable and unavoidable parts; thereby it shows the real potential of improvement of each component as well as for the overall system.

1.7   Thesis outline

This thesis divided into 6 chapters. Chapter 1 gives a general description of the main aspects related to the research topic. Chapter 2 describes the concept of exergy and related fundamental terms and presents the basic equations of exergy analysis. Chapter 3 provides a comprehensive review of literature related to the research problem

under consideration covering deficiencies of existing work on the application of exergy analysis on LNG processes. In chapter 4, a detailed description of the overall LNG chain is provided. Research findings, specifically the exergy losses across different units of the LNG chain under consideration, and an in-depth discussion of results are presented in chapter 5. Finally, chapter 6 summarizes the conducted work and provides conclusions, recommendations, and future work.

CHAPTER 2: EXERGY ANALYSIS

This chapter addresses the fundamentals of exergy analysis. It first defines the concept of exergy, then presents the different forms of exergy transfer (i.e. matter, heat, and work), followed by the benefits and needed equations. Essential concepts related to exergy analysis such as the reference environment state and exergy efficiency are also discussed.

2.1   Introduction

Thermodynamics play a dominant role in chemical engineering as it determines energy requirements, state of phase, physical and chemical equilibrium required for designing different units operations. Thermodynamics can also be used to assess the performance of a system in terms of energy. The first law of thermodynamics (law of energy conservation of energy balances) adapted for thermodynamic systems to identify input and output energy in a process. All processes must satisfy the energy balance where the total energy of a system remains constant, such that energy across a process is conserved even if it is transformed from one form to another. Energy balance, therefore, is an interesting way to compare processes in terms of required energy. The first law of thermodynamics, however, does not provide information on the energy efficiency of a process. In other words, it does not quantify irreversibilities that degrade the performance of the system. On the other hand, the second law of thermodynamics states that processes occur in a certain direction and energy has quantity as well as quantity [50]. Thus, it provides information on energy efficiency. The second law of thermodynamics is based on the concept of entropy, which is a measure of molecular disorder within a system, where a system with a high disorder degree such high-temperature gas, has high entropy and vice versa. Entropy also assists in explaining the natural direction of energy transformations and conversions. However, entropy is

practically not easy to use for interpretation consider for example its unit of energy/temperature. Exergy, the unit of energy/time; however, is more of a useful parameter.

2.2   Exergy concept

The term "exergy" was used for the first time by Rant in 1956 [51], referring to the Greek words ex(external) and ergos (work). Another term that is used describing exergy is "availability" referring to available energy [52]. Exergy is defined as the maximum theoretical useful work that can be obtained by bringing the system at a given state to thermodynamic equilibrium with the environment by means of a reversible process in which the system interacts only with its environment [53]. Where the state of a system is defined by its composition, pressure, and temperature. Exergy can be also defined as the minimum theoretical work required to bring the system from equilibrium with the environment to a given state [54]. Thus, exergy is a measure of the deviation of a system at a particular state from the state of the environment. This means that both the system and the environment attribute to exergy. It is to be noted that the environment is large resulting in almost a uniform temperature, pressure, and composition. Hence, the environment may be regarded as free of irreversibilities and thus there is no exergy destruction within the environment [55]. However, all significant irreversibilities are attributed to the system and its direct surrounding. Exergy numerical values are dependent upon the state of the system and the environment cannot be a negative value. When the process occurring between the environment and the system is reversible, exergy is conserved. Conversely, when the process is irreversible, exergy is destroyed. By definition, a system that is in equilibrium with its environment has zero exergy since it has no ability to generate work with respect to its environment, while a system that is not in equilibrium with its environment

has some quantity of exergy [50]. Consequently, exergy calculations will always be dependent on the state of the environment and of the system.

2.3   Exergy Reference Environment

Exergy analysis is a function of a chosen reference state that is described by intensive properties including temperature, pressure and chemical components composition of a reference environment. The reference state must be specified upon carrying on the exergy analysis such that exergy of a system or stream is always measured with respect to the reference environment in order to attain consistent analysis. The reference environment is defined as a large equilibrium system that acts as a source and sink for heat and materials. Only internally reversible processes are experienced in the reference environment where the state variables temperature T_0 , pressure P_0 and chemical potentials μ_{i0} (dependent upon the composition y_{i0}) of each of the i chemical components remain unaltered when heat and materials are exchanged between the environment and another system through a thermodynamic process [52]. When a system or a stream of interest is in global equilibrium (i.e. mechanical, thermal and chemical) with the reference environment, which means that the system is at rest relative to the environment, the system is said to be at dead state. When the system/stream is at dead state, there is no interaction between the system and the environment and, thereby, no potential to develop work (i.e. zero exergy). The reference environment properties are specified differently according to the application, one example is shown in Table 2.1. The reference environment here represents an ideal gas mixture modeling earth's atmosphere. Temperature and pressure of the environment are 298.15K and 1 atm, respectively. The table also presents the composition of the environment in terms of mole percent (y_0).

Table 2. 1: Exergy Reference Environment. Adapted From [55].

Temperature	298.15K (25^{0}C)
Pressure	1 atm

Composition	
Component	y_0 (%)
N_2	75.67
O_2	20.35
H_2O	3.12
CO_2	0.03
Other	0.83

2.4   Exergy transfer

Figure 2.1 illustrates an open system, also called control volume (CV), which has instantaneous properties such as internal energy U, volume V, and entropy S. Streams flow into the CV system at the initial state, go through a change of state and exits the system as outlet stream. The change of state process might accompany work

or heat transfer across the boundary of the system, where $\dot{Q}_{in}$ is heat input from a heat source (e.g. steam) at T_h , $\dot{Q}_{out}$ is the heat output to a heat sink (e.g. seawater or air) at T_c, W_{in} is the work supplied to the process (e.g. compression work) and W_{out} is the work generated (e.g. generated work in gas turbine). Mass, heat or work transfer across boundaries of a system are accompanied with exergy transfer, and they represents exergy gained or lost by the system during a process.
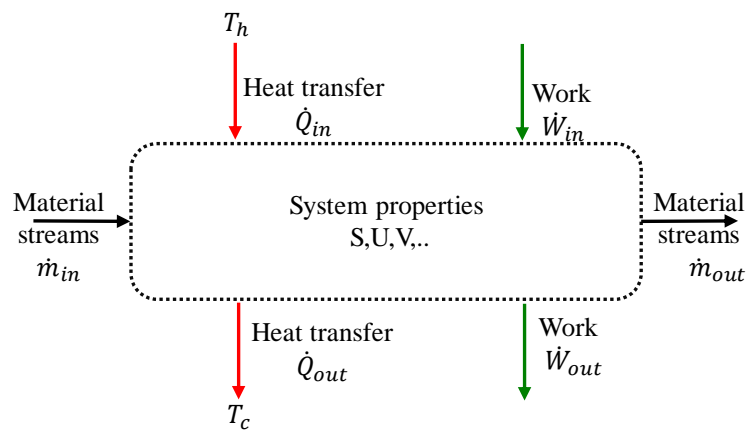
Figure 2. 1: an open system with different forms of flow across the system.

2.4.1   Exergy transfer by mass

Mass contains energy, entropy, and exergy, and it can transfer all these in or out of a system. The exergy transfer by mass mainly consists of physical and chemical exergies. Referring to Figure 2.2, and equation (1), the physical exergy is further classified into mechanical (i.e. Potential and kinetic) and thermomechanical (i.e. temperature and pressure based). On the other hand, chemical exergy consists of chemical reaction, mixing and separation parts.
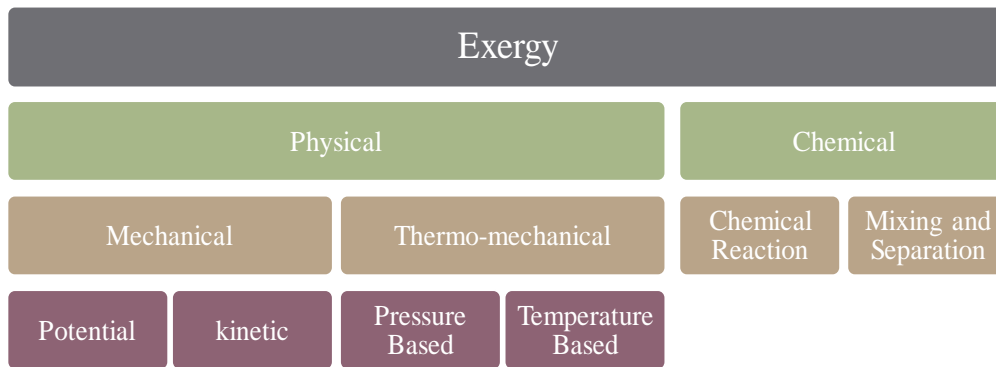
Figure 2. 2: Classification of exergy transferred by mass

$$Ex = ex^{Physical} + ex^{Chemical} \qquad (1)$$

The thermo-mechanical exergy is due to the deviation in temperature and pressure of the system from those of the environment, Also known as thermal and mechanical exergies, respectively. Thus, it might be defined as the work obtainable from taking the system from its initial state (T, P) to environmental state (T_0, P_0) through a set of reversible processes. As illustrated in Figure 2.3, different paths can be followed to generate the same amount of work. For path (e), temperature and pressure could change concurrently. In path (a+b), the temperature based exergy relates to first taking the system from T to T_0 while maintaining a constant pressure at P, whereas the pressure based exergy relates to bringing the system from P to P_0 at T_0. On the other hand, in the path (c+d), the pressure based exergy relates the change in pressure of the system from P to P_0 at constant temperature T, followed with a change in the temperature of the system from T to T_0 at constant pressure P, that corresponds to the temperature based exergy. The different paths would give different results for temperature based and pressure based components of the exergy. However, the exergy

16

is a property of the system, thus the total change in thermos-mechanical exergy is independent of the path.
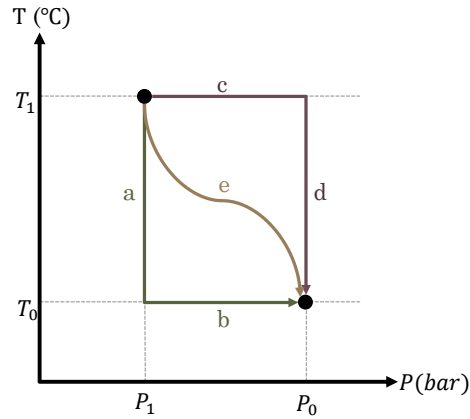
Figure 2. 3: Decomposition of thermomechanical exergy [56]

The thermo-mechanical exergy $(ex^{T,P})$ of a material stream is represented in terms of enthalpy and entropy as follows:

$$ex^{T,P} = m\big((h - h_0) - T_0(s - s_0)\big) \qquad (2)$$

Where h is the specific enthalpy at conditions of the process under consideration and h_0 is the specific enthalpy at the dead state temperature (T_0), s and s_0 are the specific entropy at conditions of the process under consideration and at reference state, respectively.

The second division of the physical exergy is the mechanical exergy consisting of kinetic and potential parts. Kinetic energy is a form of mechanical energy, thus it can be considered as work. Therefore, the exergy resulting from the kinetic energy due to

the system velocity measured relative to environment state is simply equal to kinetic energy itself regardless of the temperature and pressure of the system.

$$ex^{kinetic} = \frac{mV^2}{2} \tag{3}$$

Where V is the velocity of the system with respect to the reference environment. Similarly, potential energy is a form of mechanical energy and thereby can be considered as work. Thus, potential exergy corresponding to the potential energy of a system that results from the system height relative to dead state is equivalent to potential energy itself regardless of the temperature and pressure of the environment and is determined as follows:

$$ex^{potential} = mgz \tag{4}$$

Where g is the gravitational acceleration and z is the elevation of the process with respect to reference elevation in the environment. Therefore, the exergies associated with kinetic and potential energy are equivalent to themselves due to that they are available as work. On the other hand, the chemical exergy resulting from the deviation of the chemical composition of the system from the environmental composition is defined as the maximum work obtained when taking the substance under consideration from the environmental state (T_0, P_0) to the dead state (T_0, P_0, y_0) through a process that involves heat transfer and exchange of substances only with the environment. The chemical exergy of a system consists of two parts, reactive exergy that is associated with chemical reactions, and nonreactive exergy associated with non-reactive processes such as mixing and separation.

The chemical exergy can be determined according to equation 5 as the sum of a reactive $(\sum x_i \varepsilon_i)$ and nonreactive exergy $RT_o \sum x_i \, ln(x_i)$ [52]. Where the standard chemical exergy ε_i of components can be obtained from reference tables[31, 55].

Whereas in case of substances that are not tabulated in reference tables, the value of standard exergy could be determined by considering a stoichiometric reaction of the substance through which the substance decomposes to other substances whose standard chemical exergy is available in reference tables[57]. Method can be found in [31, 58]

$$ex^{Chem} = \dot{n} \left(\sum x_i \varepsilon_i + RT_o \sum x_i \ln(x_i) \right) \qquad (5)$$

Where ex^{Chem} is the chemical exergy of material stream, x_i is the composition of component i in the considered stream, ε_i is the standard chemical exergy, R is the universal gas constant $\left(8.314 \frac{J}{k.mol} \right)$, and T_o is Reference state temperature (25 ℃).

2.4.2   Exergy transfer by heat

According to Carnot cycle, during heat transfer, only a fraction of energy can be converted to useful work and this fraction of energy is the exergy. Thus, the exergy of heat supplied by a thermal energy reservoir (which can be a flowing stream going through temperature change) is equivalent to work output of a Carnot heat engine that rejects its heat to the environment [50, 59]. Thus, the maximum work that can be taken from a system supplying heat at T (i.e. thermal reservoir) is

$$W = Q \left(1 - \frac{T_0}{T_h} \right) \qquad (6)$$

2.4.3   Exergy transfer by work

Exergy is the useful work potential; thus, the exergy transfer by work (i.e. shaft or electrical work) is numerically equal to the work itself (W). It is important to note that exergy transfer by heat is zero for adiabatic systems, the exergy transfer by mass is zero for closed systems, which do not involve mass flow across their boundaries, and

exergy transfer by work is zero for isobaric systems. Accordingly, for isolated systems (i.e. no heat, work or mass transfer) the total exergy transfer is zero.

2.5 Exergy loss (destruction)

Exergy, unlike energy, is not conserved but destroyed within a system as a result of entropy generation due to irreversibilities. Irreversibilities might be classified as external or internal. External irreversibilities are due to heat transfer through a finite temperature system while internal irreversibilities include unrestrained expansion, chemical reaction, missing and friction. Exergy destruction within a system is proportional to the entropy increase. Exergy destruction represents the loss in potential work, and its value should be equal to or greater than zero. A positive value of exergy destruction indicates that the process is irreversible; here, the larger the exergy destruction the more irreversible the process is. In a reversible process, the exergy destruction is zero; however, no process is reversible in practice. On the other hand, negative exergy destruction indicates that the process is impossible. Hence, exergy destruction can be used to determine whether a process is reversible, irreversible or impossible.

As mentioned before, any process that generates entropy always results in exergy destruction within the process. Thus, exergy destruction is proportional to the entropy generated, and this is expressed in the following equation:

$$Ex_D = T_0\, S_{gen} \geq 0 \qquad\qquad (7)$$

Where $Ex_{Desroyed}$ the exergy is destructed within a process, T_0 is the temperature of the environment and S_{gen} is the generated entropy.

2.6   Exergy balances

Exergy concept is mainly applied to perform exergy balance for the analysis of the thermal and chemical systems. Exergy, unlike entropy, cannot be generated but can be destroyed on a process. Therefore, the change in total exergy of a system is equal to difference between the net exergy transferred into a system ($total\ exergy\ entering\ the\ system -\ total\ exergy\ leaving\ the\ system$) and exergy destroyed during the process due to irreversibilities. Then the exergy balance is expressed as follows:

$$\begin{pmatrix} Total \\ exergy \\ entering \end{pmatrix} - \begin{pmatrix} Total \\ exergy \\ leaving \end{pmatrix} - \begin{pmatrix} Total \\ exergy \\ destroyed \end{pmatrix} = \begin{pmatrix} Total \\ exergy \\ change\ of\ teh\ system \end{pmatrix} \tag{8}$$

Explicitly, exergy balance for any system going through any process is expressed as follows:

$$\sum Ex_{in} - \sum Ex_{out} - \sum Ex_D = \Delta Ex \tag{9}$$

Analysis of thermal plants and chemical systems in practice are mostly relevant to open systems (i.e. control volumes) analysis more than closed systems analysis. In addition, practically, most of the open systems in industrial processes such as compressors, heat exchangers, nozzles, turbines, etc., operate steadily. Which means they do not experience time variation in mass, energy, entropy, and exergy content. Therefore, for such systems, the amount of exergy in form of mass, heat or work, entering the system must be equal to the amount of exergy leaving the system plus the destroyed exergy (i.e. exergy loss).

$$\sum Ex_{in} = \sum Ex_{out} + \sum Ex_D \tag{10}$$

As mentioned earlier, exergy can be transferred to or from a system in three forms: mass flow, heat, and work. Thus, the exergy balance of an open steady state system can be expressed in the most general form as follows:

$$\sum Ex_{in}^m + \sum Q_{in}\left(1 - \frac{T_{ambinet}}{T_h}\right) + W_{in} = \sum Ex_{out}^m + \sum Q_{out}\left(1 - \frac{T_{ambinet}}{T_c}\right) \quad (11)$$

$$+ W_{out} + \Delta Ex_D$$

Table 2.2 shows exergy loss relations across different equipment, developed from the general exergy balance relation given by equation (11).

Table 2. 2: Exergy Loss Developed Equations

Component	Exergy loss	
Compressor	$\Delta Ex_D = \left(ex_{in}^{Phy} - ex_{out}^{Phy}\right) + W_{comp}$	(12)
Pump	$\Delta Ex_D = \left(ex_{in}^{Phy} - ex_{out}^{Phy}\right) + W_{Pump}$	(13)
Phase separator	$\Delta Ex_D = Ex_{in} - \sum Ex_{out}$	(14)
Heat exchanger	$\Delta Ex_D = \left(\sum ex_{in}^{Phy} - \sum ex_{out}^{Phy}\right)$	(15)
Cooler (exchanging heat with air)	$\Delta Ex_D = \left(\sum ex_{in}^{Phy} - \sum ex_{out}^{Phy}\right) - Q_{out}\left(1 - \frac{T_o}{T_c}\right)$	(16)
Expander	$\Delta Ex_D = \left(ex_{in}^{Phy} - ex_{out}^{Phy}\right) - W_{expander}$	(17)
Throttle valve	$\Delta Ex_D = \left(Ex_{in} - Ex_{out}\right)$	(18)

Component	Exergy loss	
Reactor	$\Delta Ex_D = \sum Ex_{in} - \sum Ex_{out}$	(19)
Heater exchanging heat with a hot utility of unknown flow	$\Delta Ex_D = \left(\sum ex_{in}^{Phy} - \sum ex_{out}^{Phy} \right) + Q_{in}\left(1 - \dfrac{T_o}{T_h}\right)$	(20)
Furnace with unknown fuel flow	$\Delta Ex_D = \left(\sum Ex_{in} - \sum Ex_{out} \right)$ $+ Q_{in}\left(1 - \dfrac{T_o}{T_{combustion}}\right)$	(21)
Furnace with known fuel flow	$\Delta Ex_D = \left(\sum Ex_{in} - \sum Ex_{out} \right)$	(22)

2.7 Exegetic efficiency

Exergy efficiency, often called second law efficiency, is widely used to quantify irreversibilities. Similar to energy efficiency, exergy efficiency represents the ratio of product to source in terms of exergy. Many definitions of exergetic efficiency for steady-state processes were presented and discussed in the open literature. Lior and Zhang [60] suggested that the exergy efficiency definitions can be divided into two main groups: the total, universal, overall, input-output exergy efficiency and the task,

consumed produced, functional, rational or utilitarian exergy efficiency. The first is defined as the ratio of all output exergy to input exergy flows across the system. Thus, for an open thermodynamic system at steady state, the total exergy efficiency is expressed as the ratio of all exergy outflows to inflows [60, 61]:

$$\eta = \frac{\sum Ex_{out}}{\sum Ex_{in}} = 1 - \frac{Ex_D}{\sum Ex_{in}} \tag{23}$$

The total exergy efficiency described above is considered to be appropriate when a major part of the exergy leaving the system is a useful product [60], the input and output exergy flows are transformed to other exergy forms [61] and for dissipative processes/devices [59, 60]. However, exergy outflows might include waste streams that represent external exergy losses such as material waste, emissions, and heat losses, which then implies that not all outputs of the system can be considered as useful outputs. And thus, the total exergy efficiency has been criticized in literature in that it takes into consideration all exergy flows without taking into account whether they are consumed in the thermodynamic conversion. Therefore, it is more adequate to consider the useable output exergy only to describe the efficiency of a system. The rational efficiency distinguishes the exergy flows that are not altered (i.e. neither used nor generated) from the exergy flows undergoing transformations. Thus, the rational exergy efficiency is claimed to be more precise to evaluate a system in that it is defined as the ratio of the produced exergy, that is exergy associated with generated products in the system, to the consumed exergy, which is exergy associated with resources used to produce these outputs. Rational efficiency can be then written:

$$\eta = \frac{\sum Ex_p}{\sum Ex_f} \tag{24}$$

Where Ex_p is the exergy of the desired product and Ex_f is the necessary input that is often named as exergetic fuel [62].

CHAPTER 3: LITRATURE REVIEW

This chapter provides a comprehensive review of the LNG chain exergy literature. The aim is to summarize the present status of the field.

3.1 Introduction

The LNG chain, which comprises of NG treatment and purification, liquefaction, storage, shipping, and regasification, is considered as one of the most energy-intensive industrial processes and a minor efficiency improvement could reduce the energy consumption significantly, which in turn may translate into huge economic profits and significant reduction in environmental impact associated with the energy consumption. It has been observed in the literature that there is a growing interest in process optimization within the LNG chain. Most optimization studies have been performed with the objective of improving the design, performance and minimizing power consumption (i.e. Maximizing exergy efficiency). Thermodynamic tools, such as exergy and pinch analysis, have been widely employed to assist the evaluation and optimization of different processes. These approaches might be employed as investigative tools for performance evaluation or in combination with optimization to minimize the search effort for optimization approaches. The focus here is, however, the exergy analysis. As mentioned earlier, all processes are, to some extent, irreversible in practice. These irreversibilities can be recognized and quantified, in the form of exergy destruction, using exergy analysis that was introduced by Kotas [31] and described in chapter 2. Consequently, exergy analysis is primarily an indicative tool used to pinpoint locations of highest irreversibilities, alternatively lowest exergy efficiency, and thus locating opportunities for process improvement.

3.2 Review

A selection of publications on exergy analysis of LNG chain processes is tabulated in table 1.3.

Table 3. 1: References Classified According to the Segment of the LNG Chain on Which Exergy Analysis Was Applied

The element of the LNG chain	References
Gas treatment and fractionation	[24-27] and [63-65]
Liquefaction	
	[66-85], [28, 29], [16], and [20]
Transport, storage and	[86-92] and [30]
regasification	

A review study of exergy analysis of the heat exchanger was conducted by Manjantah and Kaushik [93]. Different exergy analysis approaches have been proposed for process design purpose, where exergy analysis is combined with pinch analysis [94, 95]. Linhoff and Dhole presented low-temperature process design methods based on exergy and pinch analysis. Therein, exergy grand composite curve is used to approximate the change in shaftwok requirements of refrigeration system without re-simulation [96]. Marmolego and Gundersen proposed a graphical approach for energy

targeting, applicable for low-temperature processes where the proposed procedure combines exergy and pinch analysis. As a case study, the approach was used for a nitrogen expander process design [97]. Rian and Ertesvåg, performed exergy analysis for the aim of evaluating a complete LNG plant where a mixed-refrigerant cascade process was used for the liquefaction process [83]. Pellegrini et al. conducted energy and exergy analysis to evaluate the thermodynamic performance of two acid gas removal processes (MDEA and DPLT) in the LNG production chain. In this work, the authors also considered different NG liquefaction technologies: Mixed Fluid Cascade (MFC), Propane-Mixed Refrigerant (C3MR) and Single Mixed Refrigerant (SMR) [24]. Ghorbani et al. employed exergy analysis to evaluate NG liquids (NGLs) recovery integrated with NG liquefaction process. Here, replacing the pre-cooling stage of mixed fluid cascade refrigeration system with an absorption system was also investigated. Exergy analysis results showed that air coolers are associated with the highest exergy destruction prior and after installing the absorption refrigeration cycle at a rate of 56.21% and 42.72%, respectively [25]. Lee et al. Performed detailed exergy analysis for a cryogenic energy storage combined with LNG regasification process aiming to identify opportunities to enhance process efficiency, such as optimizing operating conditions or further possible cold energy recovery [98]. Morosuk et al. applied exergy analysis to evaluate a single mixed refrigerant process (PRICO), that is operated with compressors driven by a gas turbine system and is used for small scale LNG plant, and highlight improvement opportunities of the PRICO process [66]. Kanoğlu Performed exergy analysis on a multistage cascade refrigeration cycle that is used for NG liquefaction where exergy destruction and energetic efficiency for the main cycle components were developed [67]. Exergetic efficiency of the evaluated process is near 39% indicating a great improvement potential. Choi developed a new process for

reducing power consumption of the partial liquefaction process for LNG carriers using exergy analysis [68]. Tirandazi et al. Assessed a multistage refrigeration cycle in Ethane and heavier hydrocarbons (C2+) recovery plant in order to determine the components causing major exergy destruction. Exergy efficiency of the investigated refrigeration indicates a great potential of improvement, as it is determined to be near 43.5% [69]. Palizdar et al. conducted a thermodynamic evaluation based on conventional and advanced exergy analysis of three mini-scale nitrogen single expansion for natural liquefaction. The exergetic efficiency of the investigated process is nearly 44%, which indicates a great improvement potential [70]. Tesch et al. applied advanced and conventional exergy analysis to evaluate LNG regasification process integrated with an air separation process [86]. Vatani et al. investigated five conventional LNG process using exergy and energy analysis approaches. Namely, the single mixed refrigerant (Linde and Air Products and Chemicals Inc), the propane precooled mixed refrigerant, dual mixed refrigerant, and mixed fluid cascade [29]. Remeljej et al. evaluated three small scale LNG processes including two open loop expandesr, two-stage expanders nitrogen refrigerant, and single-stage mixed refrigerant. However, their work focused primarily on the cold section of small-scale plants[71]. Samimi et al. performed exergy analysis for sulfur recovery unit (SRU) aiming to allocate thermodynamic losses. Results of this work reveal that the exergetic efficiency could be improved by minimizing the exergy losses; such that the exergy efficiency in the Claus reaction furnace and incinerator could be increased from 11.22% and 9.9% to 14.96% and 12.58% correspondingly [26]. Derbal et al. performed exergy analysis of mixed refrigerant (MR) process in LNG plant to determine the performance of the equipment constituting the MR process (main cryogenic heat exchanger and centrifugal compressors) and control their operation [72]. Zargarzadeh et al. Developed

a tool, Olexan, using Visual Basic Application (VBA) to enable dynamic and online exergy analysis in an interactive manner at various levels of the equipment and plant [99]. Banat et al. Utilized exergy to show the units that are associated with high exergy destruction within a gas-sweetening unit; results revealed that the absorber, followed by heat exchangers and vessels were the units with the highest losses [63]. Khan et al. performed energy and exergy analysis to enhance the efficiency of a three-stage propane pre-cooling cycle, where the effect of changing operating conditions of evaporators was analyzed considering energy consumption of the process, and exergy losses and efficiencies were reported for several cases [73]. Sagia et al. carried out a study to analyze irreversibilities in a one-stage refrigeration process that uses vapor compression cycle; exergy balances were carried out on the system components, and losses were calculated by an algorithm developed on a thermodynamic basis. Results were presented through Grassmann diagrams while targeting the effect operating temperatures have on the cycle's exergy efficiency [28]. Chang et al. Investigated combined Bryton-JT cycles with single-component refrigerants NG liquefaction, namely N_2-Brayton cycle with C_2-JT and C_3-JT, aiming to high efficiency, large capacity and simple equipment. Therein, exergy efficiency was examined to evaluate different LNG process concepts, where the methane cycle in conventional cascade process is replaced with a nitrogen expander cycle [74]. Fábrega et al. Performed exergy analysis of the refrigeration cycles in the ethylene and propylene production process. In this study, the commercial simulator HYSYS$^{©}$ was used to generate process data. Exergetic analysis results showed that the highest exergy destruction occurs in the mixers, followed with compressors and then heat exchangers accounting for 40%, 22% and 20% respectively. New operational data were proposed aiming to reduce exergy losses, where 13% of the total exergy loss for the entire refrigeration system could be

reduced [100]. Gong et al. Performed exergy analysis to compare the performances of a single mixed-refrigerant process and an auto-cascade mixed refrigerant process. Results showed that the auto-cascade process provides better performance [75]. Tsatsaronis and Morosuk conducted a detailed advanced exergy analysis of a novel co-generation concept that combines LNG regasification with the generation of electricity. Results of the exergy analysis were confirmed with sensitivity analysis. This study also provides developed suggestions for improving the overall system efficiency [87]. Tsatsaronis and Morosuk in another work, analyzed the improvement potential and interactions among components in LNG-based cogeneration systems and showed the advanced exergy analysis over the conventional one[88]. Tsatsaronis and Morosuk also presented aan dvanced exergy-based evaluation of the same system, including advanced exergetic exergoeconomic and exergoenvironmntal analysis to evaluate the performance, economic and environmental aspects of the system. Further, the potenital to improve the thermodynamic efficiency, reduce the overall environmental impact and reduce overal cost was also determined. The main aim of this study is to demonstrate the potentai for electricity generation during LNG regasification [89]. Panjeshahi et al. have analyzed the south Pars NG Plant using exergy concept where the exergetic loss and exergetic efficiency have been determined for the major unit operations. The analyzed units involve the primary separation, sweetining, glycol recovery, condensate stabilization, dehydration, mercury removal, mercaptan removal, sulfur recovery, sour water treatment, liquid gas treatment, gas pressure compressopn, air centrifugal compression, steam production, power production, mercaptan removal. Therein, some modificatinos have been proposed aiming to lower the energy consumption [64]. Baccanelli et al. have been modeled three NG low-temperature purification techniques, namely: the dual pressure low-temperature distillation process, the anti-sublimation

process and the hybrid configuration. In this study, the exery analysis was employed to evaluate the performance of the modeled processes. Results of this work showed that the dual pressure low-temperatre distillation schem provided the highest thermodynaic performance (highest exergy efficiency) compared to other processes [65]. Ghorbani et al. investiated an integral NG liquids (NGL) and liquefied NG (LNG) using exergy and exergoeconomic analysis methods. Results reaveled that the air-coolers and multi-stream heat exchangers were of the highest exergy efficiency. It has been shown that the exergy destruction in the compressors directly affects their required power, thus, any reduction in the exergy destruction would significanlr reduce the operating cost [27]. Kanbur et al. have conducted exergy analysis for cthe onventional micro-cogeneration system and a micro-cogeneration system that utilizes the liquefied NG (LNG) aiming to detect the exergy destruction and loss ratios. Results showed that the minimum exergy dstruction is associated with pumps and LNG vaporizers [90].

Pinch Analysis (PA) has also gained a high level of industrial application and has been successfully applied for many process design aspects such as to improve heat recovery, design better power, utility and heat systems [101]. Aspelund et al. analyzed the design of sub-ambient processes using a new methodology based on pinch and exergy analysis called ExPAnd that was developed in the LNG industry, in order to minimize energy requirements in the NG liquefaction cycle. The efficiency of the ExPAnd approach was verified by applying it to design of LNG process where exergy efficiency could be increased from 49.7% to 85.7% compared to standard pinch analysis [16]. Al-Sobhi et al. utilized process integration techniques to optimize an LNG process without the sulfur recovery and detailed NG liquefaction systems. Pinch analysis was employed to identify and enhance the inefficiencies of the process. It was found that through heat integration, heating and cooling utilities could be reduced by

15% and 29%, respectively [102]. Malham et al. introduced a novel hybrid pinch-exergy integration methodology to couple both analysis methods and overcome their individual limitations. The proposed methodology is applied in a basic LNG process to assess exergy losses and determine needed changes in operating conditions and process structure [103]. Gourmelon et al. proposed a novel methodology that combines Pinch and exergy analysis. The approach, referred to as PiXAR, is used for analysis and retrofitting of existing industrial processes [104]. Wechsung et al. developed an optimization formulation model that combines pinch analysis, exergy analysis, and mathematical programming. Their model is used for the synthesis of heat exchanger networks [85].

3.3 Conclusions

The literature review indicates that the research activity in the field of LNG process optimization is increasing in last years, and is likely to continue in the future owing to the expected growth in the LNG trade. As powerful thermodynamic tools that aid the process optimization, exergy and pinch analysis have been widely employed to improve the performance of LNG chains. The majority of published studies; however, have been focused on the liquefaction process itself owing to its high capital and operating costs, which in turn have a significant impact on the economy of the LNG chain. On the other hand, thermodynamic analysis of other segments of the LNG chain is still lack of research, specifically for the hot section of the chain (i.e. upstream of the liquefaction process). Further, optimizing LNG baseload chain requires a comprehensive evaluation of all units, yet, among the available studies, no work has been considered a full LNG chain. This work, however, contributes to filling this gap.

CHAPTER 4: PROCESS DESCRIPTION

This chapter presents an overall description of the LNG chain used in this study which was adopted from [49]. Detailed flowsheets are presented in chapters with the permission of the authors.

4.1 LNG chain description

Figure 4.1 schematically illustrates the LNG supply chain considered in this work; the chain consists of the following interdependent segments: LNG main process, exporting terminal (i.e. LNG storage and loading), shipping and receiving terminal (i.e. unloading and regasification). The main LNG process could be further divided based on operating temperature into two sections, the hot section that requires energy in the form of high heat to remove and transform impurities from NG and the cold section that requires energy in the form of refrigeration. Raw NG enters the pre-separation unit for gas, water, and pentane plus (condensate) separation. Condensate is then sent to condensate stabilization for light hydrocarbon stripping. The aim of the condensate stabilization is primarily to lower the vapor pressure of the condensate liquids, thus avoid production of vapor phase upon flashing the condensate to atmospheric pressure at storage tank [105]. Stabilized condensate is stored and sold as crude oil [106]. NG leaving the pre-separation unit, referred to as sour gas that contains undesired impurities (CO_2, H_2S , mercaptans, Benzene-Toluene-Xylene (BTX), mercury and water) that should be removed for operation and products specifications purposes. The presence of high freezing point and corrosive components might cause blockage in the downstream cryogenic heat exchangers (used for NG liquefaction) which can cause major disturbances. In addition, LNG should not contain more than 4 ppm and 57 ppm of H_2S and CO_2, respectively [49]. Thus, the sour gas undergoes several treatment steps before liquefaction. Sour gas first enters the Acid Gas Removal Units (AGRUs) where

H_2S, CO_2, and mercaptans (known as acid gas) are removed by means of absorption processes with aqueous alkanoamine solutions. The considered LNG chain involves two amine systems, methyldiethanolamin (MDEA) and Sulfinol unit. The last is followed with an Acid Gas Enrichment Unit (AGEU) in order to increase the H_2S content in the acid gas to improve the subsequent sulfur recovery. Separated is sent to the sulfur recovery unit (SRU). To produce elemental sulfur from H_2S using the Claus chemistry. The LNG chain under consideration involves two sulfur recovery units, a Claus process for the acid gas leaving the MDEA unit and a SuperClaus process for the acid gas leaving the AGE unit. The sweet NG from the sulfinol process is fed to the dehydration unit where molecular sieves are used to remove water from the NG upon liquefaction. Once NG is dried to the required level, mercury is removed from the NG stream in the mercury removal unit to prevent its reaction with aluminum that is a common material used for the construction of liquefaction cryogenic heat exchangers [107]. After impurities are removed from the feed stream to the desired extent, NG is liquefied in the liquefaction process, using propane pre-cooled mixed refrigerant C3MR refrigeration cycle, integrated with the NGL recovery process. In the NGLs recovery unit, heavier hydrocarbon liquids including ethane, propane, and butanes, are separated by means of fractionation to yield a source of revenue, as NGLs are marketable products of great value [108]. On the other hand, NG stream is delivered to the liquefaction facility at elevated pressure, where it is subjected to a series of subsequent cooling stages by indirect heat exchange with refrigerant until it is completely liquefied. Here, both of the aforementioned units are cryogenic processes requiring refrigeration systems, thus integrating them allows sharing the refrigeration system, thereby, reducing the number of process equipment that results in economic advantage [109]. LNG is then sent to the nitrogen rejection unit integrated with the helium extraction

unit (NRU-HeXU). The purpose of NRU is to reduce the LNG nitrogen content to increase its lower heating value. Helium extraction unit, however, is required to meet product specification and to improve the economics of the plant as crude helium can be sold separately. LNG is then sent for a storage facility. The tanks have two modes of operation: holding (or LNG storage) and loading (ship loading). At the storage tank, a portion of LNG is evaporated forming tankage boil off gas (TBOG), due to LNG flashing to atmospheric pressure, heat leak through pipelines and tank walls, and vapor displacement [110]. LNG is then loaded to an LNG ship and BOG is generated due to the above-mentioned factors. The generated BOG during LNG loading process is known as jetty BOG (JBOG). During LNG holding, generated TBOG is utilized as a fuel. On the other hand, during the loading process, JBOG is generated in much higher amounts than TBOG. Thus, a portion of it will be used to supply the fuel requirement of the plant whereas excess JBOG is flared. Afterward, LNG is transported from the production facility to the final destination (import region). BOG is also generated during the shipping process. Once LNG carrier berth into the LNG importing terminal, the LNG is unloaded from cargo tank into cryogenic storage tanks. Similar to exporting terminal, BOG is also generated in the receiving terminal on the jetty area as well as in the storage tank. LNG is then regasified using open rack vaporizer (ORVs) system where LNG is vaporized using seawater. Gasified NG is then distributed to customers via the gas grids.

The LNG main process units require a substantial amount of shaft work, electrical power, cooling utility, and heating medium, all of which are generated and supplied through the utility section of the plant which in turn consumes the fuel generated in different units of the plant. Utilities in the utility section are generated using steam-power cogeneration system and standalone gas turbines (GTs). With

seawater and air being the main cooling media. For the receiving terminal, in addition to seawater electric power is needed and generated using standalone turbines.
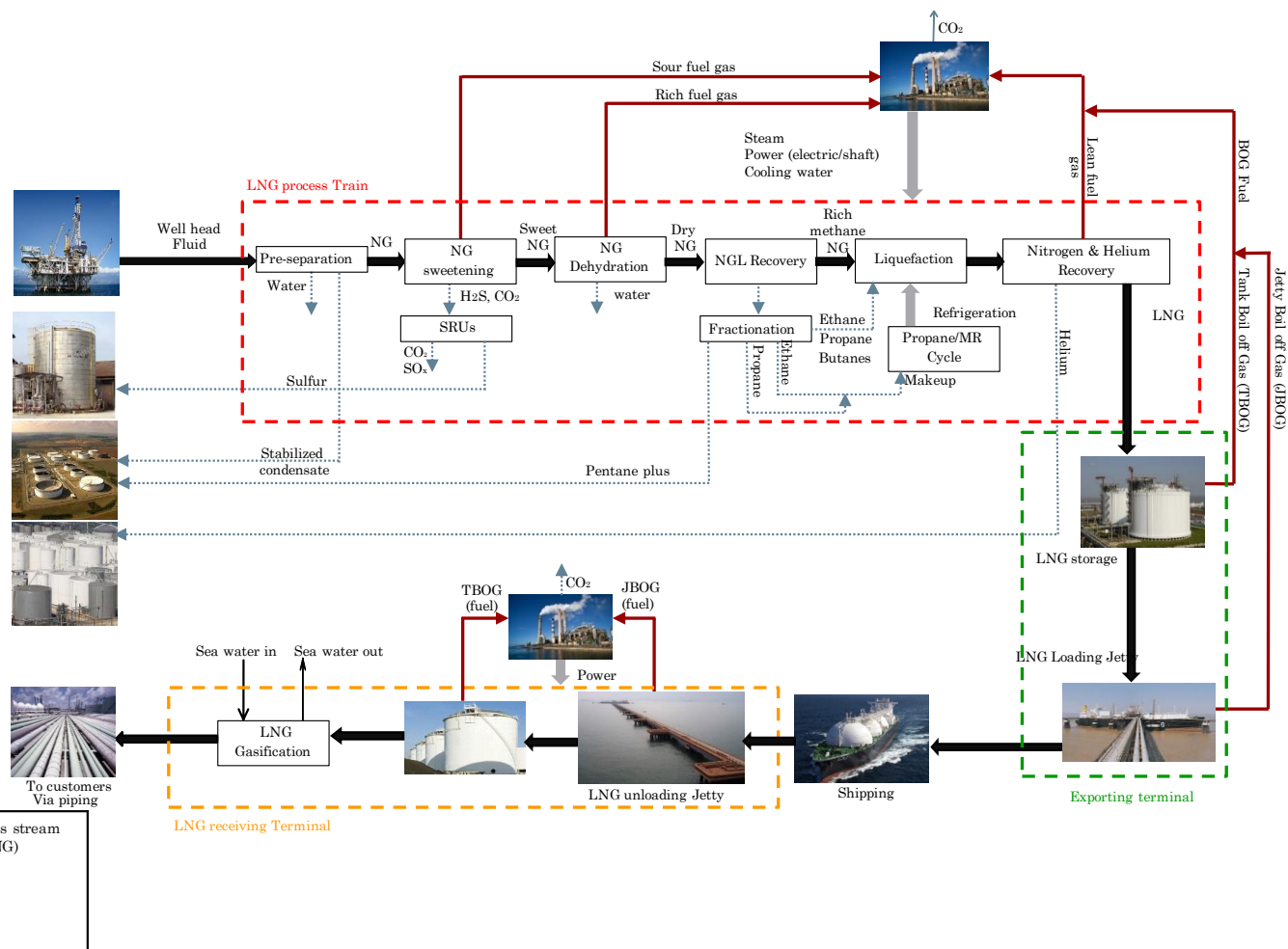
Figure 4. 1: block flow diagram for the chain considered in this work.

5.1 Introduction

Following the approach described in section 1.5, and fixing the environment reference temperature and pressure at 25℃ (298.15 K) and 1.01325 bar, respectively, exergy flows (i.e. material, heat, and work), exergy destructions within the plant, exergy losses to the environment, and exergetic efficiencies across the LNG chain units were calculated. In this chapter, the results of this exergy analysis are presented and discussed with reference to the objective of the study that is to quantify and allocate thermodynamic inefficiencies within the chain.

5.2 Overall results

Results of the exergy analysis including the exergy streams entering and leaving the chain as well as the utility plant are presented in Figure 5.1. The figure shows that the exergy content of the NG feed needed to deliver 439 MMSCFD of NG to the end user pipeline grid is approximately 7.81GW. This amount flows through the various processes making up the chain until it ends up in the final regasified LNG for which exergy content is 4.77 GW. Additional exergy of 1.73 GW also leaves the chain in the form of by-products (i.e. Helium, NGLs, Condensate, Fuel, etc.). Exergy flows into and out of the LNG chain units in the form of heat, mass, and work are also shown in the figure. The total exergy loss across the chain was identified to be almost 1.04 GW, corresponding to near 13% of the exergy input. This translates into an overall chain exergy efficiency of 87%. The total loss here combines the losses across the process equipment in the main process train, LNG storage, utility section (the GTs and cogeneration system), LNG shipping, and receiving terminal plus its utility (electric power). Figure 5. 1 shows the exergy flowrate of the main process stream (i.e. NG or

38

LNG) from different units of the LNG chain. Here, the gradual decrease in the exergy flow rate indicates the transformation of the NG feed exergy into losses and by-products (material and heat). On the same figure, the second curve shows that the total production (mainstream and by-products) is decreasing as we move towards the final product indicating that there is a portion of exergy that is being destructed either internally, due to irreversibilities (pressure drop) or externally due to the discharge of useless streams to the environment, such as flue gases.
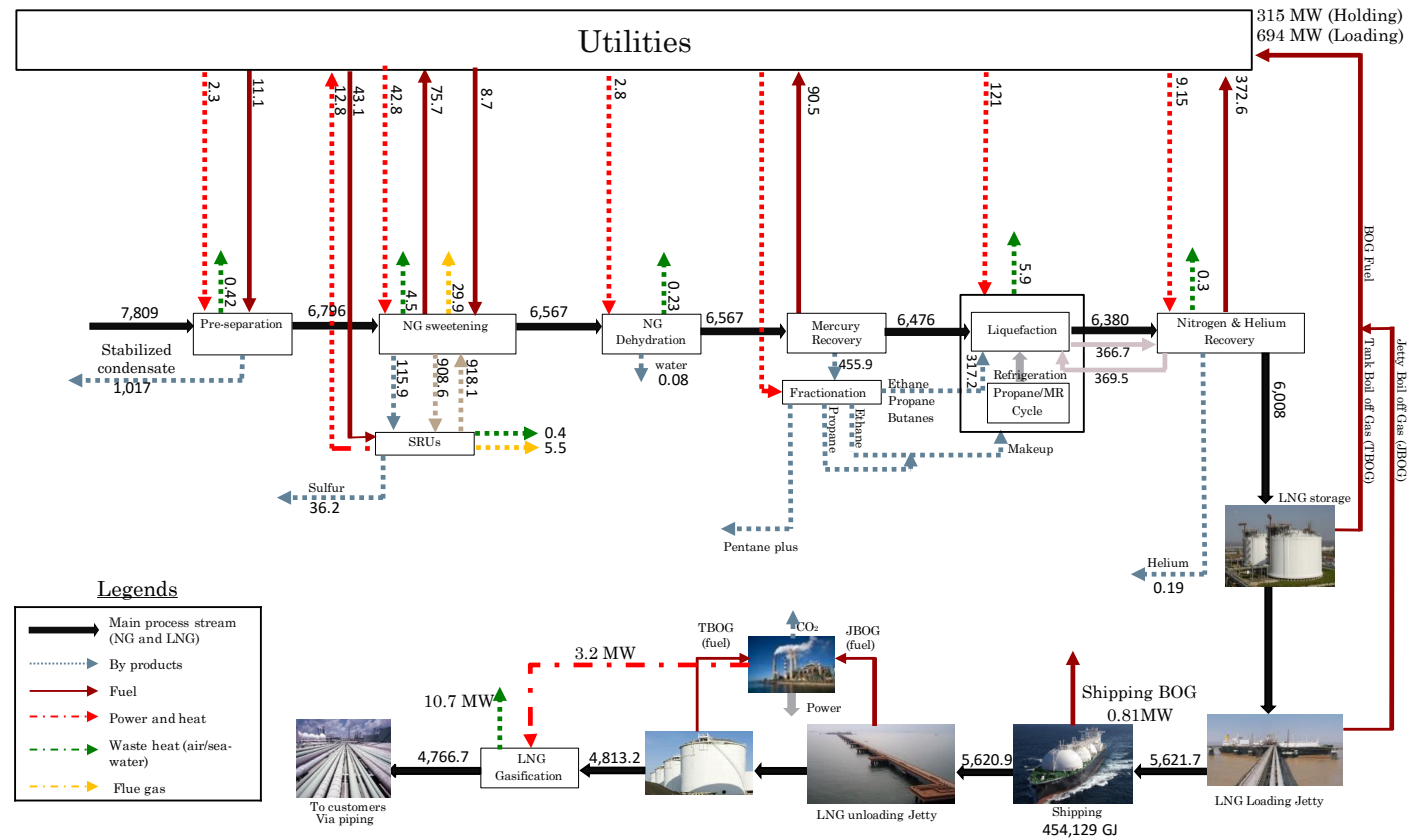
Figure 5. 1: Schematic overview of exergy streams entering and exiting the LNG chain.
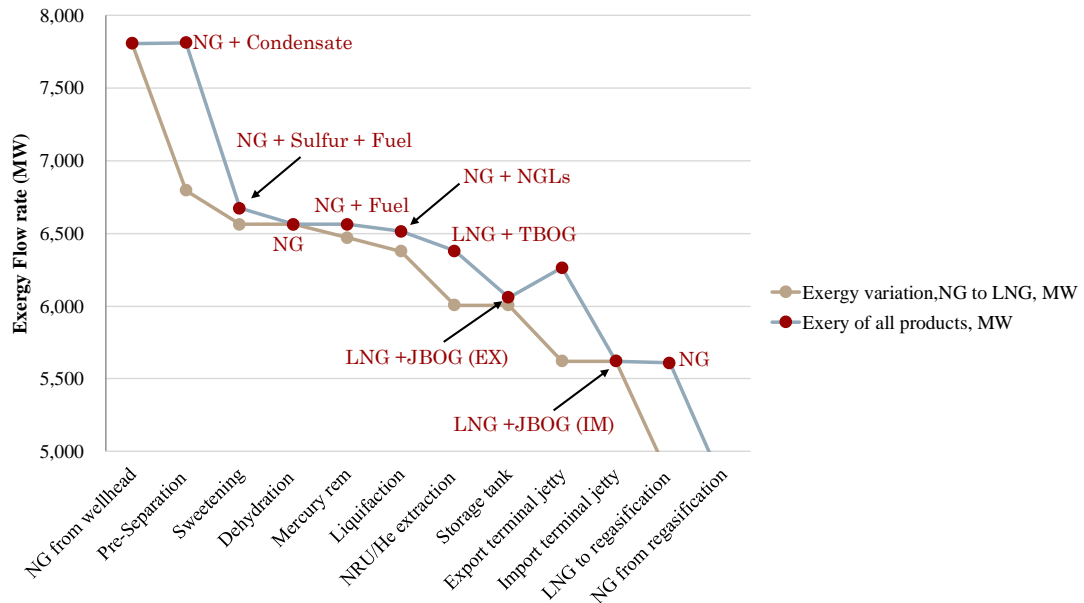
Figure 5. 2: Exergy flow rate variation across the chain

Results indicate that most of the losses occur in the main process train and its utility generation. Among the main process train units (without the LNG tank and utilities), the main contributors to the total loss are SRUs, followed by liquefaction and sweetening which are associated with 94, 73 and 59 MW of losses, respectively. Accounting for 39%, 30% and 24% of the total exergy loss, respectively (see Figure 5.3). However, when considering utilities, the losses across these units increases to 100, 237, and145 MW respectively. Here, each equipment loss in each unit is added to the loss across its relevant utility generator. Each MW of loss in these processes corresponds to approximately 0.1, 2.3 and 1.4 MW loss in the utility GTs and steam generators, respectively. The following subsections discuss the details of the analysis across the segments making up the LNG chain under consideration.
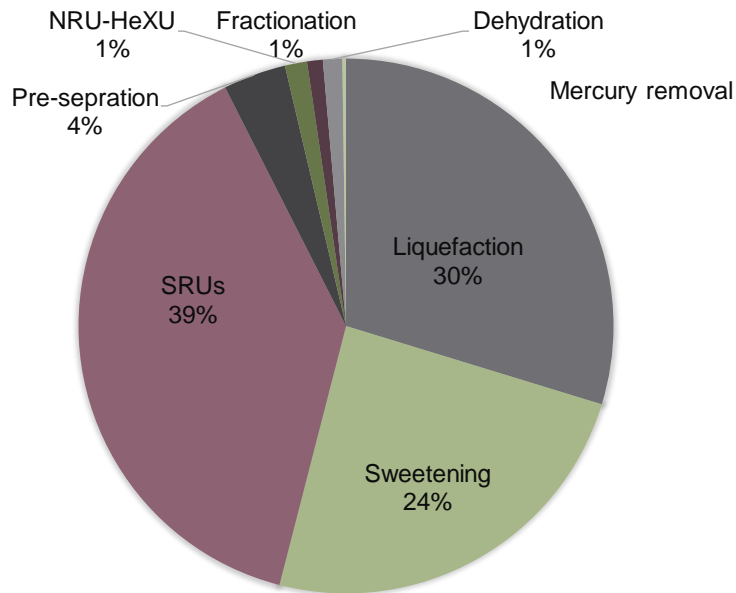
Figure 5. 3: Percentage exergy destruction in main process units, without utilities.

5.3   Liquefaction process (C3MR)

The C3MR and its integration with the NGLs recovery process are illustrated in Figure 5.4. The process comprises of two refrigeration cycles, pure propane pre-cooling cycle to pre-cool the treated feed gas and to partially liquefy the MR, and an MR cycle to complete the liquefaction and sub-cooling process in the main cryogenic heat exchanger (MCHE). The remarkably high exergy destruction in that unit (30% of total losses) is attributed to the C3MR cycle compressors, their drivers (i.e. GT's), and heat exchangers. Figure 5.5 shows the breakdown of the exergy destruction across the liquefaction unit (a) without and (b) with utility losses. When studying the main process without its associated utilities (Figure 5.5 a), it was found that the highest consumers are the heat exchangers, followed by compressors, representing 64% and 28% of the

total 73 MW losses, respectively. The exergy efficiency of the liquefaction unit is 41% which may indicate a potential for improvement. On the other hand, when considering utilities it was found that the process requires 422 MW of fuel exergy (consumed in GTs) to produce 120 MW compression power and 137 MW of combustion flue gas exergy. Thus, the total losses across the C3MR cycle while taking into consideration losses in utilities is 273 MW. Referring to (Figure 5.5 b), losses were the highest across the compressors and their drivers, followed by heat exchangers, with approximately 75.2% and 18.9% of the cycle's total loss, respectively. Losses across the compressors are almost 20 MW (28% of total 73MW), however, when considering the utilities (primarily 30% efficient simple GT cycle), losses across the compressors increased to 178 MW (75% of total 237MW). This means that the rotating equipment losses alone (158 MW) are equivalent to compression work needed to run a baseload liquefaction unit producing over 4 MTA of LNG.
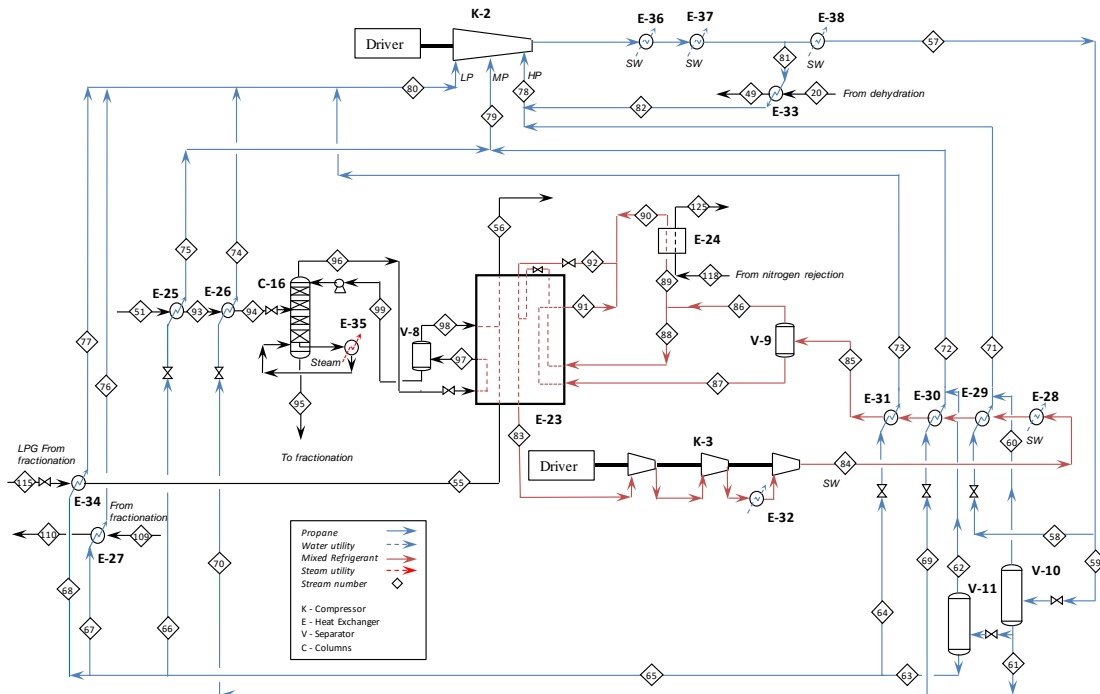
Figure 5. 4: NGL recovery unit integrated with the C3MR liquefaction process

schematic. Adapted from[49]with the kind permission of the authors
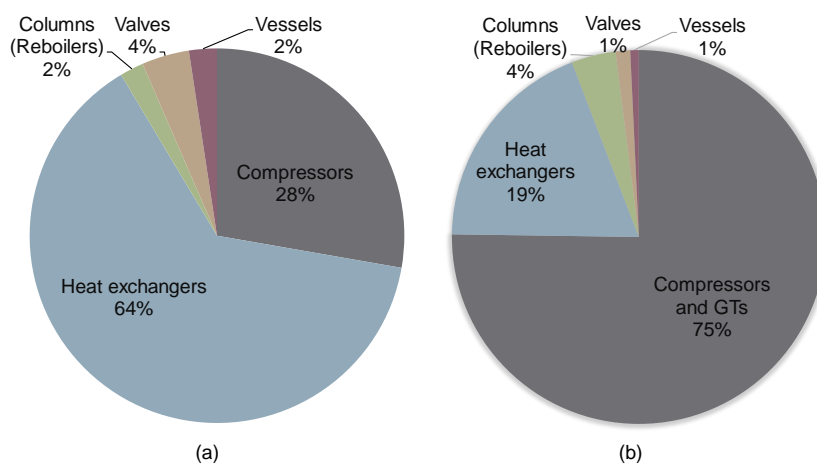
Figure 5. 5: Exergy destruction in the liquefaction unit. (a) without utilities losses, (b)

with utilities losses

Figure 5.6 shows the exergy losses across the heat exchangers in the liquefaction cycle. Of the 47 MW exergy destroyed in heat exchangers, 15.2 MW exergy is destroyed in the MCHE (E-23), this is followed by propane condenser E-37 (6.2 MW loss), and MR compressor inter and aftercooler, E-28, and E-32, respectively (combined loss of 11 MW). Our simulation shows that 1MTA of LNG requires near 40 MW of compression energy. Thus, the 47 MW total loss across the heat exchangers alone is equivalent to the power needed to produce near 1.175 MTA LNG. This highlights the importance of optimizing those heat exchangers.
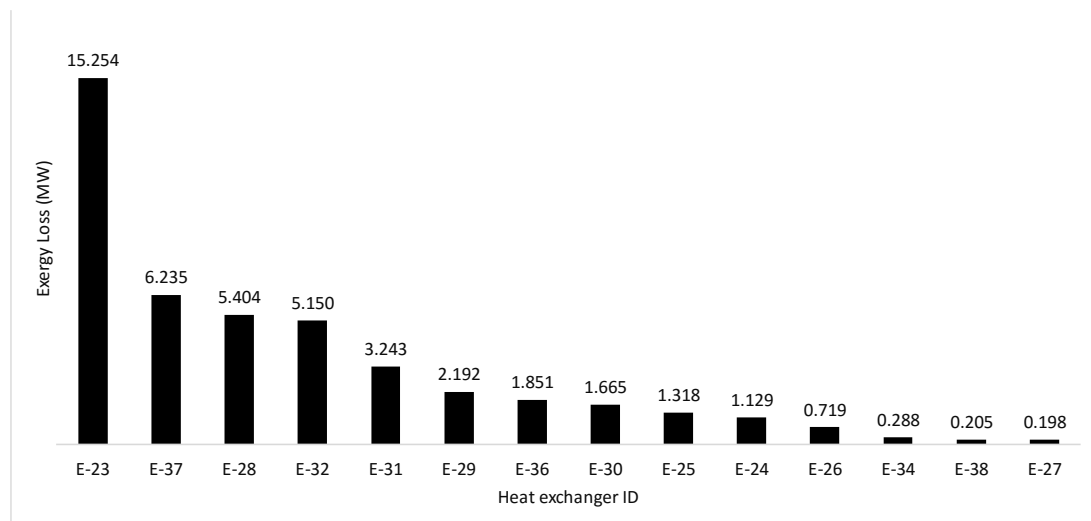
Figure 5. 6: Exergy destruction in heat exchangers

These findings emphasize the need for liquefaction cycles optimization. Minimizing compression power of baseload liquefaction system is; however, a complex none straightforward task. The process has many designs and operating variables (more

than 20) that are available for optimization. Some of these are MR refrigerant composition, heat exchangers minimum temperature approaches and UA. While many works have been done to optimize a given liquefaction cycle configuration, the majority of the works make use of sophisticated optimization algorithms that require using multiple software [21, 22, 111-113]. Such approaches may be found to be effective but are believed to be not suitable for conventional software users such as process engineers and students. Another issue, which is related to design, is choosing the proper heat exchangers minimum temperature approaches. In the baseload C3MR process, such as the one on hand, there are more than five heat exchangers that require deciding on their minimum temperature approaches. Large temperature approaches can save in capital at the expense of operating cost and vice versa. Given the fact that the process configuration is standard and very common in the expanding LNG industry, it may be of great benefit to identify optimum MTA for each exchanger. Optimizing energy consumption while minimizing capital cost is not widely discussed in the open literature and simple algorithms that can be utilized by non-advanced users are missing.

Large savings in fuel were observed if one can achieve small savings in the compression power. Thus, given the continuous growth of LNG demand, development of new liquefaction cycles for grassroots and retrofitting seems to be attractive options.

Compressors drivers (i.e. GTs) optimization and development should also be of great importance due to their current inefficiencies. Such inefficiencies are attributed to the thermodynamic limitations of the GT cycles, which use the Brayton cycle concept[50]. Brayton based cycles are associated with efficiencies that range between 30 to 55% (based on fuel LHV), with the latter being achieved using a combined cycle system[114, 115].

Rather than the 8:1 drivers to compressors loss ratio, using combined cycle would result in an almost 6MW loss in the combined cycle for every 1 MW loss in the refrigerant compressors. Furthermore, fuel cell systems should also be considered to provide shaft power to the process compression. Unlike the Brayton cycle, fuel cells efficiencies are not thermodynamically limited [116] and a state of the art combined cycle solid oxide fuel cell (SOFC), which can produce power with near 71% efficiency[114, 117], would result in a loss ratio as low as 3:1. Having said so, using such sophisticated drivers is expected to substantially increase the capital cost of the plant (grassroots or retrofitting); thus, there is a need to identify the tradeoff between the exergy and capital savings.

5.4   NG Sweetening and Acid Gas Enrichment

The sweetening unit comprising of two AGRUs and AGE are shown in Figure 5.7, 5.8 and 5.9. These processes mainly involve columns (i.e. absorbers and distillation columns plus their reboilers) and incinerators for waste disposal. Referring to Figure 5.10 (a), out of near 59 MW total exergy loss in the sweetening unit, more than 50% of the total losses take place in the units' columns while 35% is lost in the incinerators. The columns alone destroy nearly 30MW, which is almost 42% of the total exergy destruction in the liquefaction unit without utilities. Considering the utility side of the process Figure 5.10(b), mainly due to steam generation, it was found that exergy losses in columns and incinerators are the highest, accounting for 77% and 15 % of the total 145 MW loss, respectively. Columns total exergy losses increased to near 111 MW, which is close to the actual C3MR compression power. This means that near 81 MW is lost in the steam generation step only.
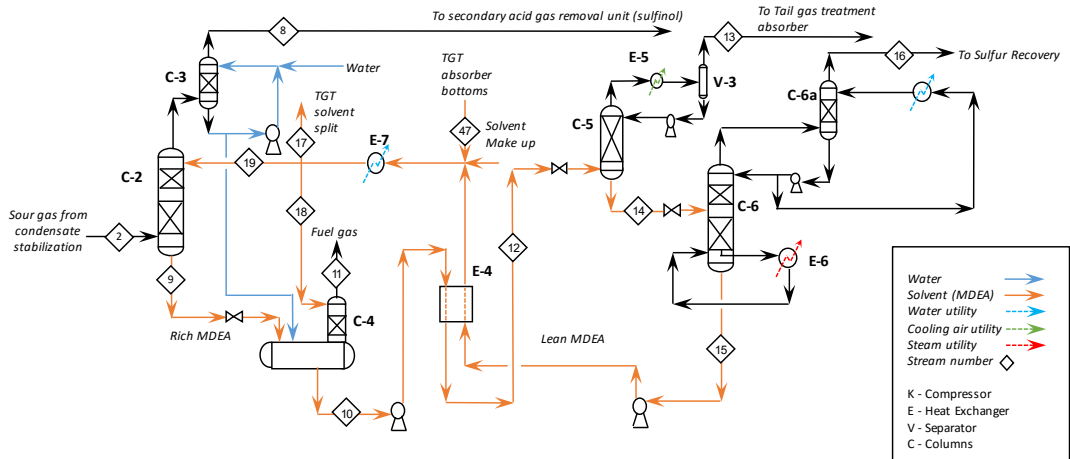
Figure 5. 7: Process scheme of the MDEA sweetening Unit (Step 1). Adapted from[49]with the kind permission of the authors
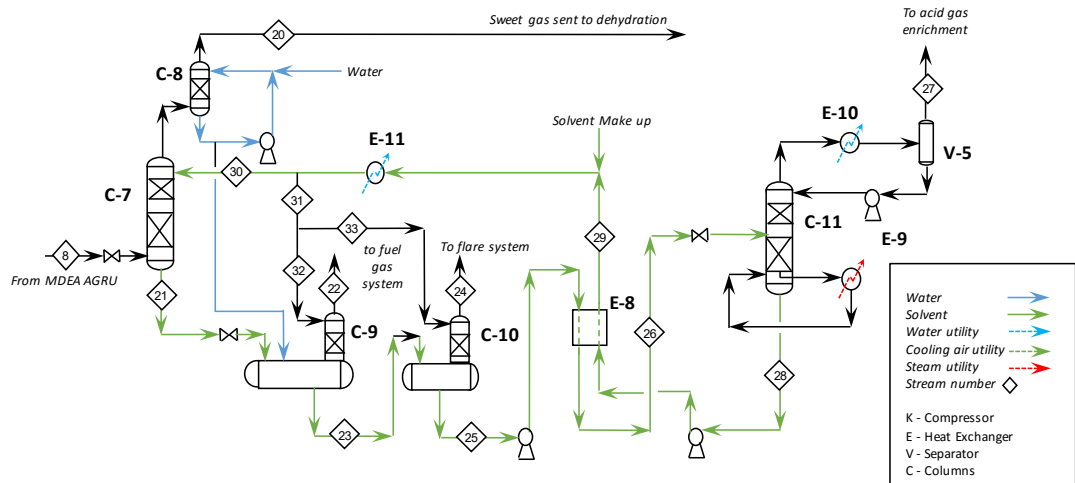

Figure 5. 8: Flow diagram of the Sulfinol® sweetening process (Step 2). Adapted from[49]with the kind permission of the authors
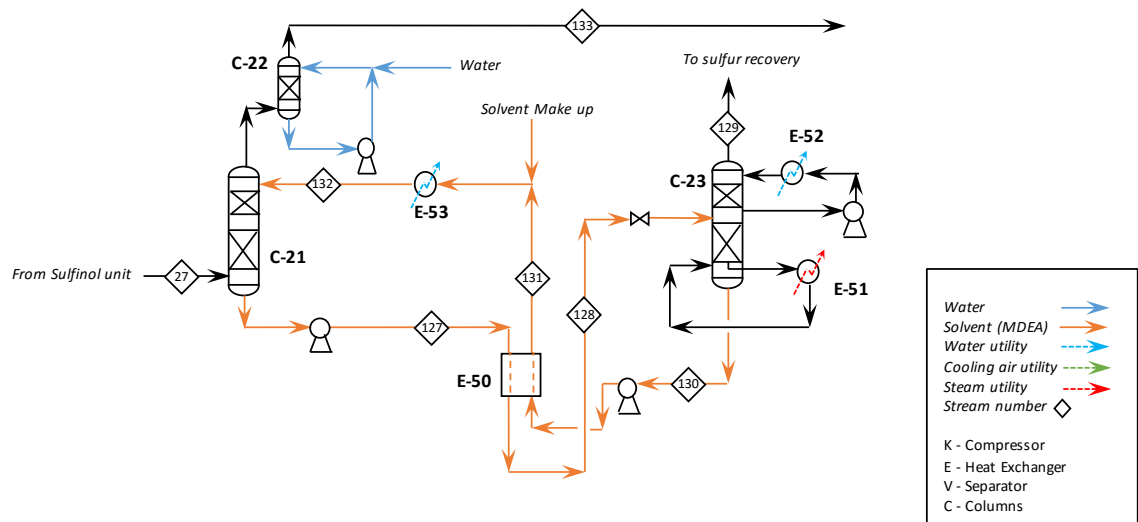
Figure 5. 9: Flow diagram of the (AGEU) process. Adapted from[49]with the kind permission of the authors
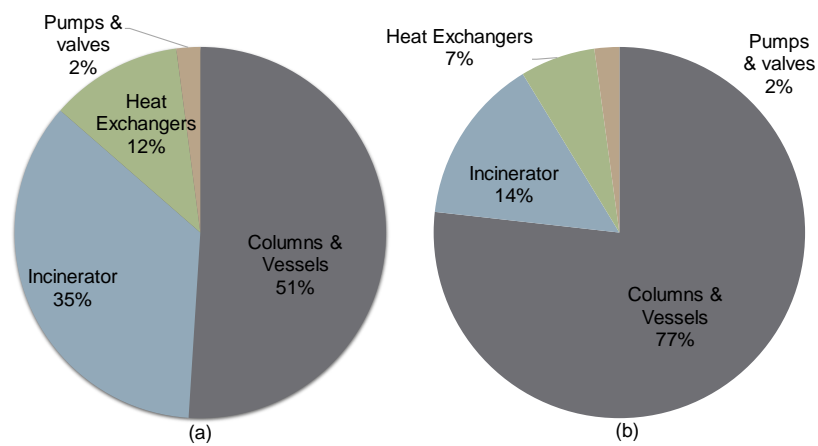
Figure 5. 10: Exergy destruction in sweetening unit. (a) Without utilities, (b) with utilities

Hence, a key option for enhancing the sweetening unit would be to minimize the utility steam (160 to 200˚C steam) consumption in the columns' reboilers. These reboilers consume around 36MW of exergy, which is equivalent to approximately 106MW of heat at an average temperature of 180˚C. The use of intermediate reboilers at proper location along the distillation columns worth investigating. Here, a portion of the column boil-up can be generated at a lower temperature; thus, reducing the need for high-quality heat at the column reboiler [118]. Another option that might be considered to reduce the reboilers' duties is the side-draw from regeneration column that is called as a split-loop arrangement [119], where a part of the amine solution is withdrawn from an intermediate stage of the regeneration column and recycled back to an intermediate stage of the absorption column. However, while this configuration of the process allows reducing reboiler duty, it comes at less acid gas absorption efficiency due to using semi-lean amine, which in turn translates to less sweet gas quality. Apart from developing new schemes, heat (or exergy) recovery is also an interesting option to consider to minimize the heating utility. One, however, should consider a number of constraints such as the physical distance between the heat sources and sinks, as it may not be feasible to recover heat from units that are far from each other. Figure 5.11 (a) shows the sweetening heat composite curves for all the streams that need to be heated and cooled, pinched at a temperature of 80˚C. The curves reveal that almost all of the columns reboilers require external heating utility (steam in this case) even at maximum heat recovery (i.e. pinched curves). It is also worth noting that the available heat that ends up in the cooling utility is at low temperatures (lower than 80˚C) with limited LNG applications. Consider for example Figure 5.11(b) which shows the pinched exergy composite curves for the unit's hot and cold streams. The overlap area between the hot and cold curves shows the maximum possible exergy recovery in the unit. While the

upper portion of the cold curve shows that around 50 MW of exergy need to be supplied, the lower portion of the hot curve shows nearly 10 MW of exergy is recoverable using a reversible Rankine cycle. However, an actual power cycle for such low temperature heats will typically be a 30% efficient Rankine cycle [50, 120]. Thus, only 3 MW of mechanical power can be generated which may not justify the economics. This emphasizes the need for new low-temperature cycles operating at higher efficiency. A heat supply option would be the SRU, as it is typically in close proximity to the sweetening unit. Pinch analysis, Figure 5.12 (a), reveals that near 50 MW of heat at a temperature higher than 238˚C is recoverable for steam generation or heating other streams in the process. As shown in the exergy composite curves of Figure 5.12 (b), the recoverable heat is equivalent to approximately 50MW of exergy. The issue here; however, is most of the streams in the SRU consist of highly corrosive components (i.e. H_2S, SO_x, S_x, etc.), thus aggressive heat recovery may require the use of expensive alloy exchangers.

In addition, operational changes in the gas-sweetening process might also enhance the overall performance. Various modifications can be carried out, including changing the feed gas pressure and temperature, amine temperature and concentration or a major change in the process are to switch to different solvents. These changes are typically attractive options for process retrofitting because in most cases little to no capital expenditure is required.
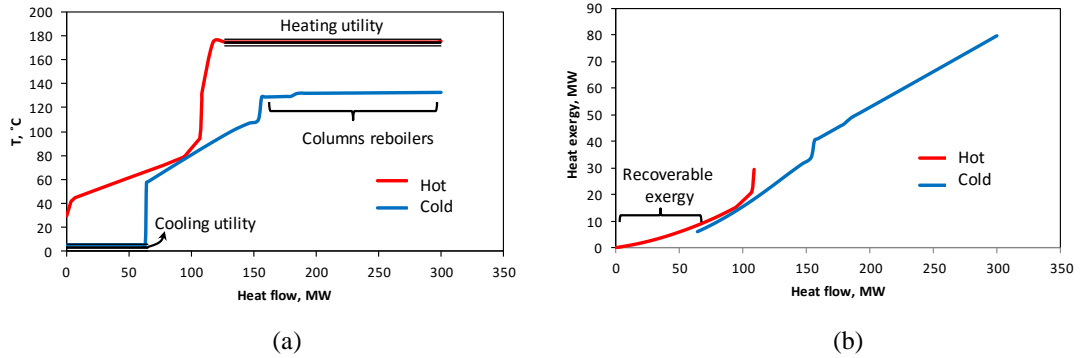
Figure 5. 11: Sweetening (a) heat and (b) exergy composite curves
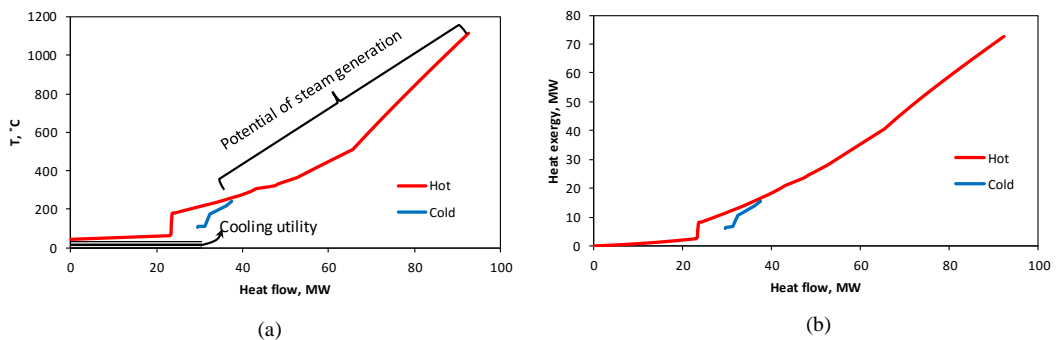
Figure 5. 12: SRU (a) heat and (b) exergy composite curves

5.5   SRU

Moving to the sulfur recovery unit, see Figure 5.13, where H_2S is converted to elemental sulfur through a Claus based process. The SRU consists of two main reaction steps. First, a portion of H_2S is burned in the reaction furnace (R-1) forming SO_2. Then, H_2S and SO_2 react in a series of catalytic reactors (R-2, R-3 and R-4) to form elemental

sulfur (Sx). Liquid sulfur is recovered in the condensers downstream each catalytic reactor while unconverted H_2S and SO_2 go to the following catalytic stage.

Near 94 MW of exergy is lost in the SRU without utilities. Referring to Figure 5.14 (a), most of the exergy loss takes place within the reaction stages, namely the reaction furnace (R-1), incinerator (R-6), catalytic reactors (R-2, R-3, and R-4) and the inline burner (R-5). Such high losses are attributed to the chemistry and combustion nature of these unit operations. In other words, chemical reactions in SRU convert higher chemical exergies materials to lower chemical exergies materials, while releasing a considerable amount of heat due to the fact that the Claus reactions are highly exothermic**.**
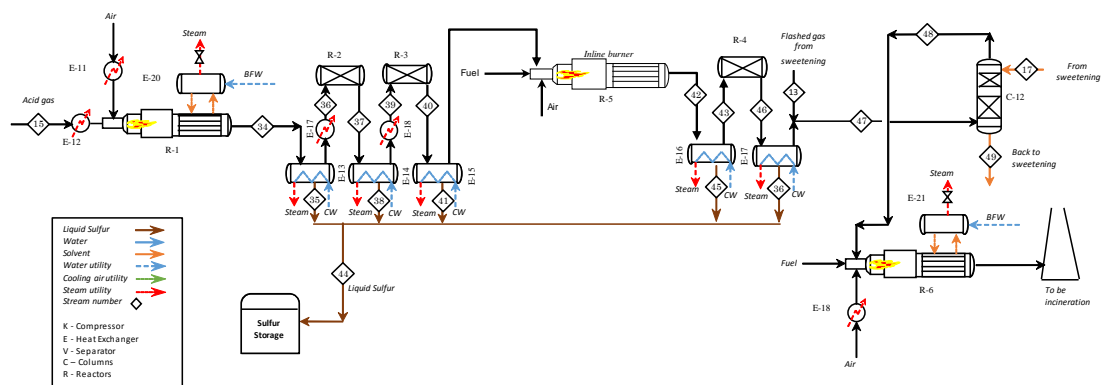
Figure 5. 13: Process flow diagram of sulfur recovery unit (SRU). Adapted from [49] with the kind permission of the authors.
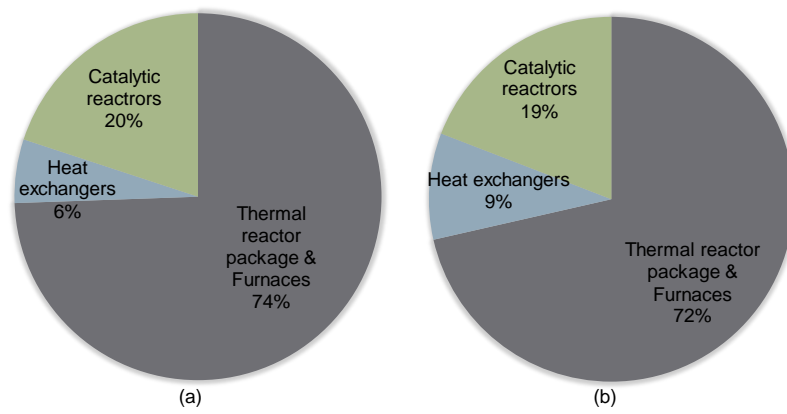
Figure 5. 14: Exergy destruction in SRUs. (a) Without utilities, (b) with utilities

Table 5.1 shows the percentage of reactants to products exergy transformation (i.e. exergy out/exergy in) of each reaction system in the SRU. In reaction furnace (R-1) near 61% of the feeds' chemical exergies are transformed into product chemical exergy. The remaining 39% is degraded to 11 MW physical exergy (i.e. heat exergy), at the furnace maximum temperature (i.e. 1116˚C). However, the process under consideration recovers the heat by generating 266˚C-saturated steam, which translates into 7 MW heat exergy. Which mean that near 5 MW of heat exergy is degraded due to the adopted heat recovery strategy. Similarly, in the inline burner, about 73% of the feed chemical exergy is transformed into product chemical exergy. The remaining 27% is lost in the form of physical exergy, accounting for 2 MW heat exergy at the maximum operating temperature of the burner (i.e. 1314 ˚C). The recovery of this heat according to the current process, results in generating 210℃ saturated steam of 0.7 MW heat exergy. In incinerator (R-6), only 4% of the feed is transformed into products chemical exergy while the rest, near 25 MW, is lost as heat exergy. However, considering the heat recovery strategy applied in the process, a 210℃ C saturated steam of 11 MW exergy is generated. Which means that about 14 MW is lost during the heat recovery process. Concurrently to steam generation, the SRU also involves steam consumers including feed acid gas preheater (E-12), process gas reheaters (E-17, E-18 and E-19), Tail Gas Treatment (TGT) unit reboiler (located in the MDEA process) and combustion air preheaters, all of which utilize HHS except for the TGT reboiler that requires LP steam. While the process is producing near 15.6 MW steam exergy, it only requires 1.7 MW. Thus, the SRU is essentially a net steam exporter, and excess generated steam is utilized outside the SRU, e.g. amine columns (i.e. absorbers and distillation columns) reboilers and steam turbines.

Table 5. 1: Exergy Transformation

Reactor	Exergy transformation
Reaction furnace (R-1)	61%
1st Catalytic reactor (R-2)	72%
2nd Catalytic reactor (R-3)	75%
3rd Catalytic reactor (R-4)	96%
Inline Burner (R-5)	73%
Incinerator (R-6)	4%

5.6　LNG storage, loading, and shipping

The exporting terminal, shipping and receiving terminal are illustrated in Figure 5.14. during holding mode near 5% of the produced LNG is lost as BOG because of heat ingress (pipelines, tank, etc.), LNG pumping and flashing to the storage pressure (atmospheric). Combining this BOG with the vapor displaced due to LNG build up in the tank, the total TBOG is associated with near 314MW exergy. Figure 5.16 illustrates the TBOG generation contribution of each factor. LNG flashing is of the highest contribution accounting for 79% of total generated TBOG.
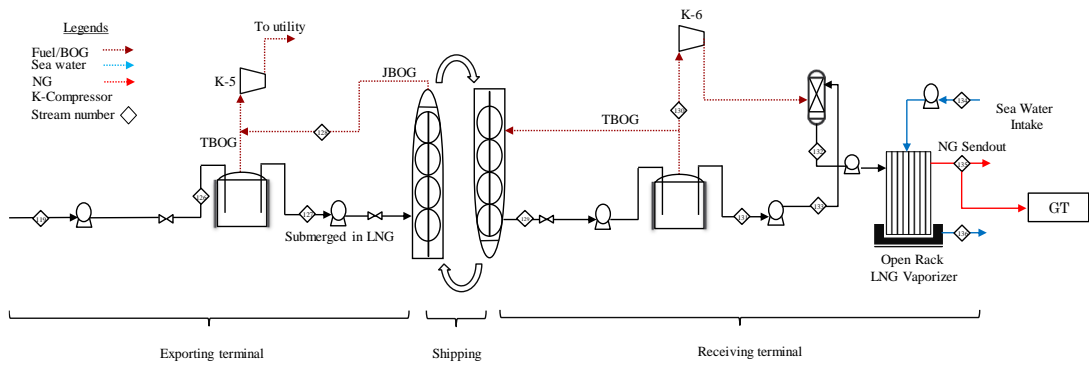
Figure 5. 15: Exporting terminal/shipping and receiving terminal. Adapted from [49] with the kind permission of the authors
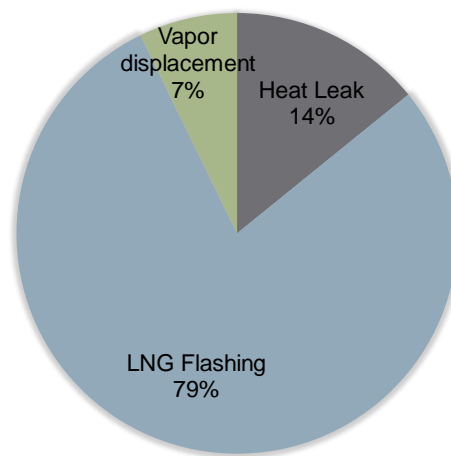
Figure 5. 16: BOG generation contribution due to each factor

This TBOG is recovered and used to supply a portion of the plant's needed fuel exergy. Therefore, while BOG recovery helps alleviate the issue of FG availability (by avoiding the need of using the additional feed for the utility section); it is associated

with losses embedded in the liquefaction step that needs to be taken into consideration. This is explained by considering the liquefaction cycle of the LNG chain under consideration. According to the simulated C3MR cycle (Figure 5.4) the production of 3.65 MTA LNG requires almost 120 MW of compression power. With BOG production being 5% of the total LNG produced, it would have consumed close to 6 MW of physical exergy (i.e. compression power) during liquefaction. Additionally, near 1 MW of exergy is lost from the storage tank. These losses amplified by 1.99 MW when BOG is compressed to the utility, as it would need to be supplied at almost 30 bar (typical FG pressures). The total exergy loss is, therefore, close to 8 MW.

During LNG loading, the TBOG generation is reduced by 83%. Thus, reducing the TBOG fuel exergy from 314 MW to 52 MW. The JBOG exergy, however, will be 46 % more than the plant fuel exergy requirement during loading. In this case, 11% of the produced LNG is lost thus wasting near 13 MW of C3MR compression power. Total loss from LNG tank, carrier, and compression is near 11 MW.

The same logic applies to the fuel generated in the NRU/He extraction. This fuel is sourced from the high-pressure LNG leaving the main cryogenic heat exchanger and accounts for almost 9.6% of the total flow. Liquefying this fuel portion consumes compression power of around 11.52 MW and the fuel leaves the nitrogen rejection column with physical exergy of 6.15 MW. Hence, by flashing the nitrogen-rich LNG in the NRU, a portion of the liquefaction compression power was wasted. While one might argue that not flashing the nitrogen-rich LNG while using fuel gas upstream of the liquefaction step could decrease the loss, it would come at the price of higher nitrogen content in the LNG and the need for higher pressure storage. These accentuate the trade-off between exergy losses, product specification, and storage cost when making such operational decisions. It also motivates the operators to investigate other

means of nitrogen removal from the LNG. In general, the avoidance of using fuel from the downstream of the cold section of the plant (liquefaction and NRU/He extraction) should result in lower exergy losses. The needed fuel may be taken before liquefaction where less energy-intensive processing had taken place. Having said so, one here needs to consider multiple issues including heavy hydrocarbon contents of the fuel. Selling these hydrocarbons may be more beneficial than saving on exergy losses.

The foregoing findings and discussions emphasize the need to reduce the impact of excessive BOG production, which in turn translates into more LNG production and reduced plant energy consumption. This could be achieved via an optimal design of LNG plant components to effectively handle the generated BOG during holding and loading operating modes or via altering process conditions to reduce the BOG generation rate.

During shipping, inevitable heat ingress to cargo tanks results in evaporating a portion of LNG, that is called shipping BOG. Typically, the BOG ranges from 0.1 to 0.15% of the full content per day, depends on tanks type (i.e. moss or membrane), where a ship's voyage is of 21 days [121-123]. Recently developed classes of LNG carriers, referred to as Q-Flex and Q-Max, have cargo capacity that varies from 210,000 to 270,000 cubic meter [110]. However, the filling capacity of each tank should be 98% of its total volume for safety concerns [124]. These requirements are taken into account to quantify the BOG generation during the ship's voyage for the LNG chain under consideration. Here, the total ship cargo capacity is 216,200 cubic meter. Thus, the delivery of 3.58 MTA LNG will require 38 shipments. About 0.81 MW of the shipped LNG exergy is lost as ship BOG; this means that near 31MW worth of BOG exergy is annually lost. BOG can be utilized as fuel for the propulsion system, burn in a gas combustion unit, vented to atmosphere or re-liquefied [121]. BOG has been utilized in

various proposed propulsion systems in the industry, where different prime movers are considered (i.e. steam turbine, gas turbine, diesel engine, and dual fuel engine) [125]. A steam turbine propulsion system (or non-ideal Rankine cycle) will typically be 20 to 30% efficient. Thus, out of the 31 MW of BOG exergy, 6.2 to 9.3 MW of BOG exergy will be utilized. If the propulsion system uses NGCC, with a 55% efficiency, the utilized BOG exergy will be near 17.01 MW. If the ship BOG is to be liquefied using a simple nitrogen expander cycle, total liquefaction power was identified to be 9kW/shipment. Consuming 15.9% of the BOG to power the cycle compressors using a 30% GT. In other words, onboard liquefaction of the ship BOG would consume near 7.6MW of BOG exergy annually. Accordingly, BOG re-liquefaction and direction back to the cargo tank leads to high efficiency and economy, which in turn enhance the profitability significantly.

5.7 LNG Regasification

Moving to the receiving terminal (see Figure 5.15) LNG pumped to the vaporizer, open rack vaporizer is considered here, for the regasification. As LNG is evaporated, near 46 MW of the physical exergy (LNG cold energy) is wasted into seawater due return the cold seawater from the vaporizer back to the sea without any means of energy recovery. Percentage contribution to exergy loss in the LNG regasification section is illustrated in Figure 5.17. Among the regasification components, the heat exchange segment contributes the largest portion of the total exergy loss, accounting for 82% of the total loss that corresponds to 38 MW exergy loss. Up to date, more than 90% of the LNG regasification plants operate without any mean of LNG cold exergy recovery concept [126]. However, different approaches have been proposed to utilize LNG cold exergy by integrating the LNG regasification process with other types of processes. Recovering this exergy could be achieved via

power generation through Brayton cycle [127], combined cycles [128-131], stiriling cycles [132, 133], absorption power cycle [134], Rankine cycle [135-137] and the direct expansion cycle [138].

For example, Reverse refrigeration cycles utilize the concept of Rankine cycle; however, these are typically the maximum work ($W_{max} = Ex_{in} - Ex_{out} + pumping\ power$) that can be generated from the evaporated LNG is around 13MW for every 1 MTA LNG. With a 30% efficiency, this power would be almost 3.6MW/MTA LNG. For a large importer like Japan (imported near 84 MTA LNG during 2017) [139], this would mean the generation of nearly 300 MW of shaft power. Figure shows how much power can be produced if the cycle efficiency improved from 30 to 50% with 53% being the thermodynamic efficiency limit that can be achieved with a reversible ideal cycle (*Max. work/heat from evaporation*). Use of LNG as cooling media may be another attractive option at receiving terminals [140-142]. Such practice should result in less exergy losses as the cold can be 100% recovered at slightly higher temperatures, unlike the power generation, which at best case scenario will achieve 53% cold to work recovery in a reversible ideal cycle. The recovery of the cold energy of LNG has been previously applied for other processes such as air separation process [143], the freeze desalination process [144], and for CO2 liquefaction.
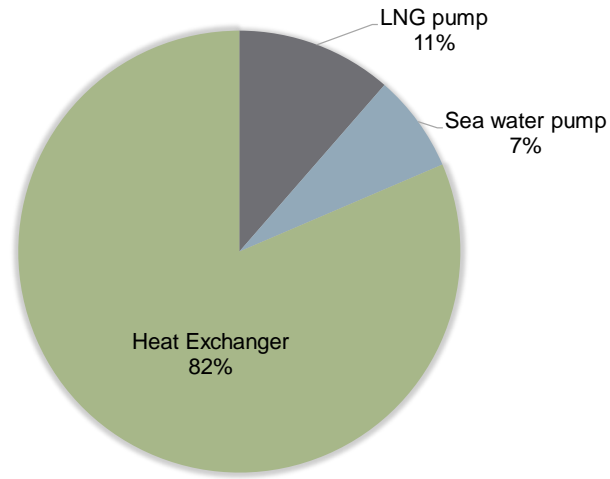
Figure 5. 17: Exergy destruction in the LNG regasification plant

5.8 Main process and export terminal utilities

As mentioned previously, exergy analysis in this study was not limited to main process units; rather it was extended to utility system that consists of simple gas turbine cycles (30% efficient) and steam simple cogeneration (78% efficient) for the generation of main process train required utilities (steam, shaft work, and electric power). Figure 5.18 illustrates the simulated utility plant serving the main process units and export terminal.

Following the approach explained in section 1.5, exergy losses across the utility section were found to be near 401 MW, where 72% of the total loss is attributed to GTs and remaining 28% is lost in the steam generators (Figure 5.19). For the utility GTs, losses were calculated by subtracting the shaft work generated by GTs and the flue gas from the fuel and air chemical exergy. Overall, 672 MW exergy of fuel is supplied to

GTs to generate total shaft work of 184 MW and combustion flue gas of 200 MW exergy. This means that near 287 exergy loss is attributed to utilities GTs. On the other hand, steam generators (boilers) losses were calculated by subtracting the generated steam exergy, which was found to be very close to heat exergy of the boilers, from the fuel and boiler feed water exergies. To steam generation systems, 152 MW, 4 MW, and 11 MW exergy of fuel, combustion air and boiler feed water (BFW), respectively, are fed to the steam generation system. Here, near 50MW exergy of steam is generated. Considering the combustion flue gas (2.8 MW exergy), exergy loss across the boilers is about 114 MW. Steam systems are also integrated into sulfur recovery units to handle

the waste heat generated in the sulfur unit and aid in controlling the overall system.
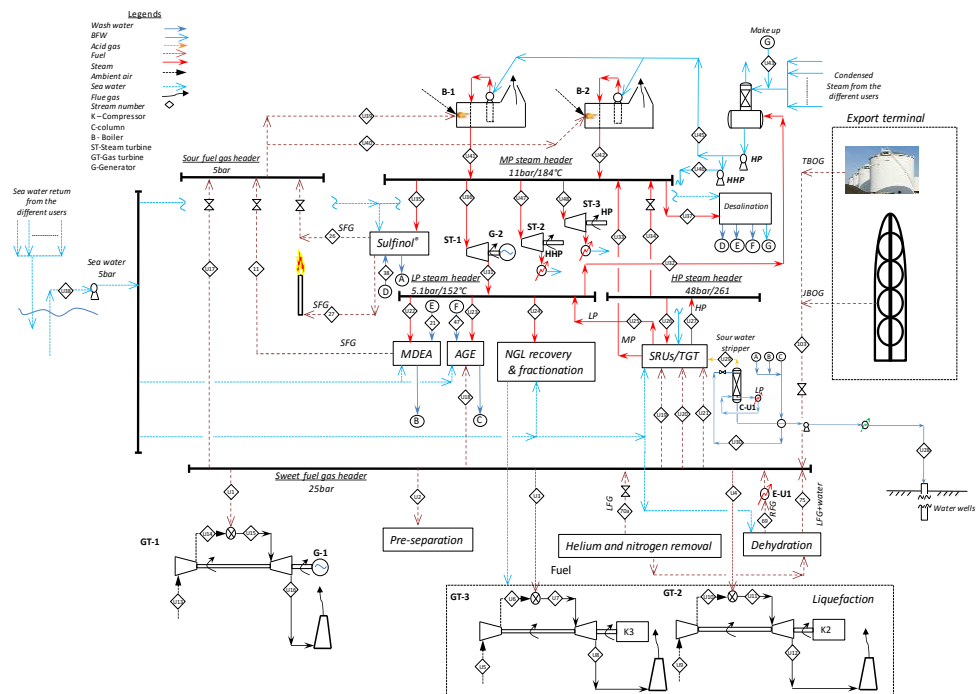
Figure 5. 18: Main process and export terminal utility. Adapted from [49] with the kind permission of the authors.
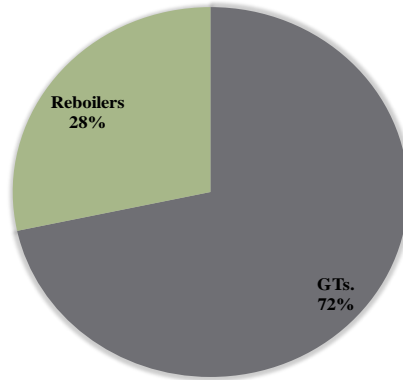
Figure 5. 19: Exergy breakdown in the main process and export terminal utility

CHAPTER 6: CONCLUSIONS AND RECOMMENDATIONS

Baseload LNG chain optimization can be very challenging due to the complexity of the systems making up the chain. Exergy analysis; however, can provide valuable insights into locations that should be targeted for improvement; thus, reducing the optimization efforts. In this thesis, rigorous exergy analysis was carried out to analyze a full LNG chain considering the performance of each individual component making up the chain. Exergy destruction equations for the equipment making up the LNG (such as heat exchangers, compressors, pumps, valves, condensers, etc.) were developed and utilized.

The study also shed light on the units with the highest exergy consumption, thereby highlighting optimization opportunities within the LNG chain. Results showed that the main contributors to exergy loss in the chain are the utility blocks, followed by, sulfur recovery, liquefaction, gas sweetening, and regasification processes; with losses of approximately 110, 26, 20, 16 and 13 MW/MTA LNG produced, respectively. Minimization of such losses is vital, as it translates to more work delivery to end users; thus, more environmental and economic advantages.

Potential improvements can be made by considering the following:

- The liquefaction process is an energy-intensive process owing to the cryogenic operating temperature and it is associated with the highest losses within the process. These losses were attributed to the compressors, their drivers and heat exchangers. It was seen that 1 MW loss in the compressors is associated with 8 MW of loss in the shaft-work generation step. Therefore, it is deemed essential to optimize the liquefaction section with the objective of minimizing the compression energy of the liquefaction process, which could

result in significant energy savings within the chain. Examples include optimization of MR refrigerant, compressor drivers, heat exchangers MTA and UA while considering the trade-off between capital and operational expenditure, CAPEX and OPEX, respectively, as it is vital to minimize the cost of the system for optimal NG liquefaction system. Results also motivate the development of new liquefaction cycle concepts for both existing (retrofitting) and grass-root plants.

- The NG sweetening unit is an energy-intensive process and is associated with high exergy loss of 59 MW. Therefore, optimization of the sweetening unit is of importance as it could result in remarkable minimization of energy consumption and thus great economic benefits to the LNG chain. Over 50% of the losses occur in the columns, this result in exergy destruction of around 30 MW, which compounds to 111 MW when considering utilities. Therefore, minimizing utility steam, via recovery/ integration, could be a key option for enhancing this process, while taking into consideration the physical distance limitations between the heat source and sink. The SRU could be an option for heat supply; however, most of its streams are considered highly corrosive, and heat recovery would require the use of expensive allow exchangers. This gives rise for the need of developing new and cost-effective materials of construction for such environments.

- The SRU is associated with significant exergy destruction of near 94 MW. The thermal and catalytic reaction stages with the values of 75% and 20% have the highest exergy destruction contribution in the unit respectively. The SRU is, in overall, a steam exporter unit such that steam is generated via utilizing generated heat due to the exothermic nature of the Claus reactions. In order to

improve the thermodynamic efficiency of the SRU, many heat recovery strategies could be applied. For example, steam superheating, steam superheating for power and heat cogeneration, and direct heat utilization to heat process streams. However, energy consumption, implementation cost and feasibly of these strategies should be studied.

- Almost 5% of the produced LNG is lost due to flashing to storage pressures, resulting in around 0.19 MW of exergy loss. While the BOG could be used as fuel gas in the plant, considerable amounts of exergy would be lost when compressing it to the desired pressure. BOG is also indirectly associated with losses in the liquefaction process as almost 7.4 MW exergy is lost due to its liquefaction as part of the feed NG. Therefore, minimizing the native impact of excessive BOG production is important as it can translate into more LNG production and reduced plant energy consumption. From an energy point of view, it would be desirable to reduce BOG formation. This could be achieved by further cooling the LNG leaving the MCHE; however, it comes at the price of additional compression power and increased nitrogen content in the final product. Evaluating other methods of nitrogen removal from the LNG could help alleviate this issue. The main process and export terminal utility section of the plant consumes 824 MW exergy of fuel generated in different units of the plant to generate 50MW steam exergy using steam-power cogeneration system and 184 MW compression work. While 203 MW is lost in the form of combustion flue gas, the total exergy destruction in utility section is 401 MW out of which 72% and 28% are lost in GTs and steam generators, respectively.

- In the receiving facilities, 46 MW exergy (4% of total losses) was lost into the sea due to LNG regasification indicating that there is a room for

improvements and optimization. This implies the need to analyze and investigate different regasification process configurations or to develop strategies for cold exergy recovery, all of which would result in minimizing exergy destruction and losses. LNG cold exergy recovery could be achieved via power generation through reverse refrigeration cycles or cold utilization for cooling applications. Enhancing the efficiencies of reverse refrigeration cycles could potentially result in over twice as much shaft power as that produced using current technologies.

Accordingly, the results of exergy analysis performed in this work provide a guideline to where the efforts should be focused in order to improve the LNG chain and provides insights to possible further work related to the topic in hand. It is also recommended to carry out advanced exergy analysis that aims to quantify avoidable and unavoidable exergy destruction in order to gain more insight into the actual potential of process improvements and to further develop the performance indicators.

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