Process Modelling and Optimization of Liquefied Petroleum Gas (LPG) Production in a Gas Processing Unit

By

Usama Talat Abbasi

(Registration No: 00000328766)

Department of Chemical Engineering

School of Chemical and Materials Engineering

National University of Sciences and Technology (NUST)

Islamabad, Pakistan

2024

Process Modelling and Optimization of Liquefied Petroleum Gas (LPG) Production in a Gas Processing Unit

By

Usama Talat Abbasi

(Registration No: 00000328766)

A thesis submitted to the National University of Sciences and Technology, Islamabad,

in partial fulfillment of the requirements for the degree of

Master of Science in
Process Systems Engineering

Supervisor: Dr. Muhammad Ahsan

School of Chemical and Materials Engineering

National University of Sciences & Technology (NUST)

Islamabad, Pakistan

(2024)

THESIS ACCEPTANCE CERTIFICATE

Certified that final copy of MS thesis written by Mr **Usama Talat Abbasi** (Registration No 00000328766), of School of Chemical & Materials Engineering (SCME) has been vetted by undersigned, found complete in all respects as per NUST Statues/Regulations, is free of plagiarism, errors, and mistakes and is accepted as partial fulfillment for award of MS degree. It is further certified that necessary amendments as pointed out by GEC members of the scholar have also been incorporated in the said thesis.

Signature: ________________
Name of Supervisor: **Dr Muhammad Ahsan**
Date: 25-JUN-24

Signature (HOD): ________________
Date: 26/6/24

Signature (Dean/Principal): 26/6/24
Date: ________________

Form TH-1

National University of Sciences & Technology (NUST)
MASTER'S THESIS WORK
Formulation of Guidance and Examination Committee

Name: **Usama Talat Abbasi**
Department: **SCME**
Credit Hour Completed: **24**

NUST Regn No: **00000328766**
Specialization: **Process Systems Engineering**
CGPA: **3.69**

Course Work Completed

S/No	Code	Title	Core/Elective	CH	Grade
1	PSE-801	Process Systems Theory	Core	03	B+
2	PSE-852	Process Modeling and Simulation	Core	03	B+
3	PSE-802	Optimization and Decision Analysis	Core	03	A
4	PSE-823	Advanced Process Dynamics and Control	Core	03	B+
5	TEE-820	Process Intensification	Elective	03	A
6	CSE-801	Computational Fluid Dynamics	Elective	03	A
7	EME-902	Numerical methods for CHE	Elective	03	B+
8	EME-809	Wastewater Treatment and Design	Elective	03	B
9	RM-898	Research Methodology	Core	02	Q

Date: 08/06 /2022

Student's Signature Usama

Thesis Committee

1. Name: **Dr. Muhammad Ahsan (Supervisor)**
   Department: SCME
2. Name: **Dr. Iftikhar Ahmad**
   Department: SCME
3. Name: **Dr. Muhammad Nouman Aslam Khan**
   Department: SCME

Signature: ____________

Signature: ____________

Signature: ____________

Date: 21/6/22

Signature of Head of Department: ____________

APPROVAL

Date: 21.6.2022

Dean/Principal

Distribution
1x copy to Exam Branch, HQ NUST
1x copy to PGP Date, HQ NUST
1x copy to Exam branch, respective institute

Form: TH-04

National University of Sciences &Technology (NUST)

MASTER'S THESIS WORK

We hereby recommend that the dissertation prepared under our supervision by

Regn No & Name: **00000328766 Usama Talat Abbasi**

Title: **Process Modeling and Optimization of Liquefied Petroleum Gas (LPG) Production in a Gas Processing Unit.**

Presented on: **04 Jun 2024** at: **1430 hrs in SCME (Seminar Hall)**

Be accepted in partial fulfillment of the requirements for the award of Master of Science degree in **Process System Engineering**.

Guidance & Examination Committee Members

Name: **Dr Iftikhar Ahmad** Signature:

Name: **Dr M. Nouman Aslam Khan** Signature: Nouman.

Supervisor's Name: **Dr Muhammad Ahsan** Signature:

Dated: 06.06.2024

Head of Department
Date 6/6/24

Mazoor
Dean/Principal
Date 6/6/24

School of Chemical &Materials Engineering (SCME)

AUTHOR’S DECLARATION

I **Usama Talat Abbasi** hereby state that my MS thesis titled **“Process Modelling and Optimization of Liquefied Petroleum Gas (LPG) Production in a Gas Processing Unit**is my own work and has not been submitted previously by me for taking any degree from National University of Sciences and Technology, Islamabad or anywhere else in the country/ world.

At any time, if my statement is found to be incorrect even after I graduate, the university has the right to withdraw my MS degree.

Name of Student: Usama Talat Abbasi

Date: 07 July 2024

PLAGIARISM UNDERTAKING

I solemnly declare that the research work presented in the thesis titled **"Process Modelling and Optimization of Liquefied Petroleum Gas (LPG) Production in a Gas Processing Unit"** is solely my research work with no significant contribution from any other person. Small contribution/ help wherever taken has been duly acknowledged and that complete thesis has been written by me.

I understand the zero-tolerance policy of the HEC and the National University of Sciences and Technology (NUST), Islamabad towards plagiarism. Therefore, I as an author of the above-titled thesis declare that no portion of my thesis has been plagiarized and any material used as reference is properly referred/cited.

I undertake that if I am found guilty of any formal plagiarism in the above-titled thesis even after the award of MS degree, the University reserves the right to withdraw/revoke my MS degree and that HEC and NUST, Islamabad have the right to publish my name on the HEC/University website on which names of students are placed who submitted plagiarized thesis.

Student Signature: Usama

Name: Usama Talat Abbasi

DEDICATION

I dedicate this work to my loving family; their unwavering support and encouragement have been a source of strength for me.

ACKNOWLEDGEMENT

All praises be to ALLAH, the mightiest, the most compassionate and the wisest, whose boundless blessings have granted me the strength to complete this work. HIS divine guidance has been a source of light throughout my journey. I would have accomplished nothing without the unfailing support, love and constant encouragement of my parents, which was vital in pushing me to complete my thesis.

I want to extend sincere and heartfelt gratitude to my supervisor, **Dr. Muhammad Ahsan**, for his invaluable guidance, support and the patience to bear with me. His consistent encouragement and uplifting, even when I was absent for long durations, was monumental in helping me achieve the completion of this thesis. His help and advice were colossal in navigating the difficulties I encountered during the study.

I would also like to offer my earnest gratitude and respect to **Dr. Iftikhar Ahmed** and **Dr. Muhammad Nouman Aslam,** members of my Guidance and Examination Committee, for their valuable support and guidance throughout my study.

Lastly, I would like to acknowledge and recognize the efforts of my elder brother, **Engr. Masab Javed,** an alumnus of NUST-SEECS, whose motivation and backing were instrumental in lifting me whenever I got off track during the research. I would also like to thank my friends and colleagues, who had a vital role in allowing me to compile and conclude my work.

Usama Talat Abbasi

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

NOMENCLATURE

NGPP	Natural Gas Processing Plant
LPG	Liquefied Petroleum Gas
PA	Pinch Analysis
AEA	Aspen energy analyzer
APEA	Aspen Process Economic Analyzer
NG	Natural Gas
NGL	Natural Gas Liquids
MOO	Multi Objective Optimization
E	Exergy
E_{PH}	Physical exergy
E_{PT}	Potential exergy
E_{KN}	Kinetic exergy
E_{CH}	Chemical exergy
Ex_{in}	Inlet Streams Exergy
Ex_{out}	Outlet Streams Exergy
I	Irreversibility
IP	Improvement Potential
E_d	Exergy destruction
T	Temperature
T_o	Reference temperature
S_g	Entropy generated
η	Exergy efficiency
EOS	Equation of State
MEA	Monoethanolamine
DEA	Diethanolamine
TEG	Triethylene glycol

ABSTRACT

In this work, Techno-Thermoeconomic Analysis has been performed for an entire Natural Gas Processing Plant (NGPP) to analyze and evaluate the potential scenarios for improving the economic and energy conditions of the unit. The interest is driven by the need to optimize the quality and yield of Liquefied Petroleum Gas (LPG) without hampering the Sales of gas conditions. Sales gas is the primary consumer product of the GPP. The steady-state model has been developed and validated using ASPEN HYSYS. ASPEN Economic Analyzer has been employed to identify cost intensive areas of the unit and to explore options to maximize profits from the Plant. ASPEN Energy Analyzer (AEA) was then utilized to determine energy exhaustive operating units of the plant, and alternatives to reduce energy consumption were suggested based on rigorous Pinch Analysis (PA). Exergy Analysis is executed to determine the exergy efficiency, which, being an indicator of thermal performance, assists in locating the in-efficient operating units having the highest exergy destruction. Economic Analysis indicates improvement in overall product sales when feed gas temperatures are decreased for the NGL Recovery section. Similarly, Exergy analysis showcases significant efficiency improvement potential in the NGL Fractionation train. The exergy efficiency of the base case can be improved by 68%, and the exergy destruction is reduced by 62% when the stream entering the recovery section is pre-heated.

Keywords: Natural Gas Processing, LPG Production Optimization, Techno-commercial Analysis, Heat Integration, Exergy efficiency, Exergy destruction, Exergy Improvement Potential, Pinch Analysis, NGL Fractionation Train, Heat Conservation

CHAPTER 1: INTRODUCTION

1.1 Background and Context

The environmental concerns associated with conventional fossil fuels, along with the depleting reserves, have triggered interest in the exploration of alternative and sustainable energy sources. As the global energy demand continues to rise, the world is making a transition towards carbon-free fuels. Since the development of the proposed transition is hindered by time constraints, policy makers and energy market leaders have looked for a “bridging fuel” to reduce carbon emissions and environmental impacts by 2030. Liquefied Petroleum Gas (LPG) has been termed the “fuel of the era”, a potential bridging fuel, as it burns cleaner, has a higher calorific value than Natural gas (NG) and is conveniently transportable, making it easily accessible. LPG Releases 18% and 30% lower carbon contents than oil and coal respectively [1]. An increase in the demand for LPG has been reported over the years with projections of the global market size to grow over USD 245.72 Billion in 2032, exhibiting a strong growth of 72% from 2022 (Fig 1).

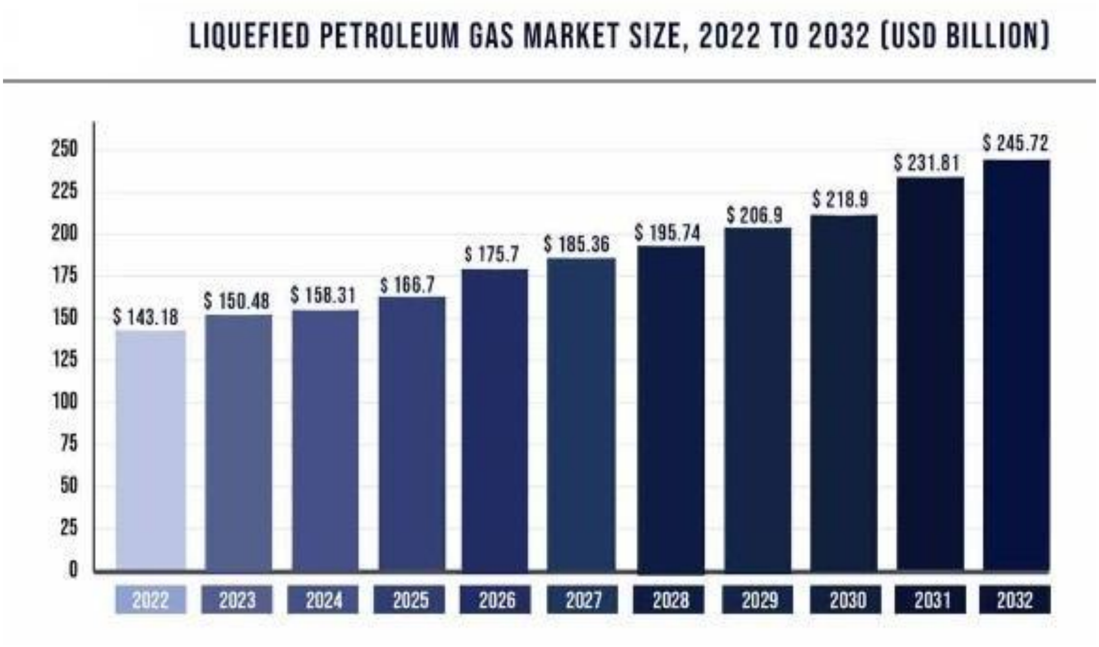


Figure 1.1: LPG Global Market Size Prediction 2022-2032

The global increase in LPG demand is attributed to the growth of its consumption in domestic as well as industrial applications. The increase is most notably observed in Europe, where adaption towards cleaner fuel for residential heating and automotive fuel is accelerated. A sector-classed breakdown of global LPG demand is demonstrated in Figure 2.

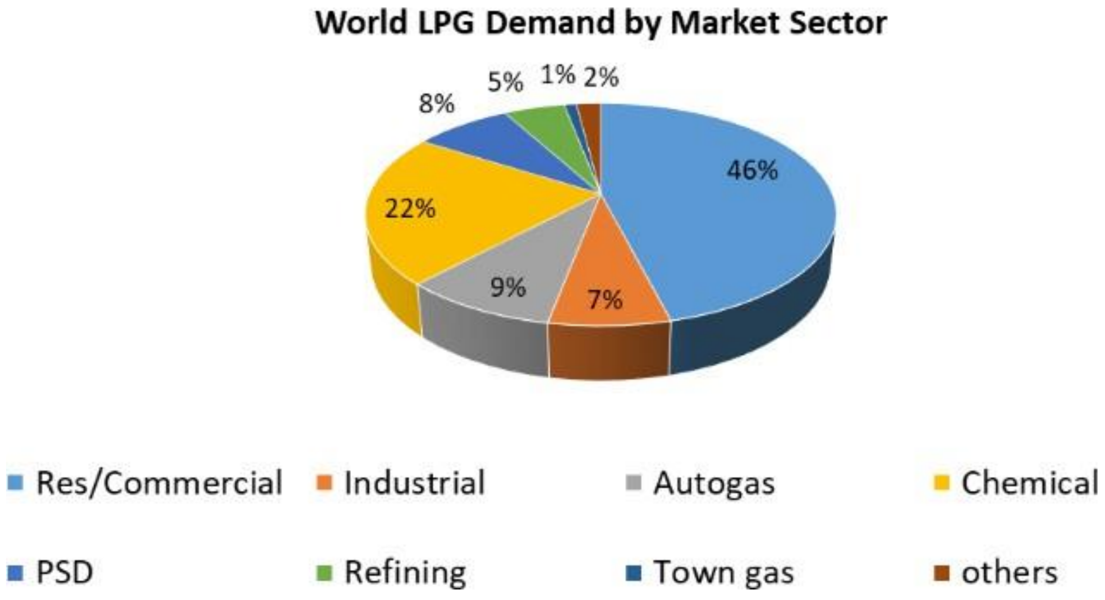


Figure 1.2: LPG Demand by Market Sector

LPG is a by-product of oil and natural gas production and petroleum refining, being produced in a highly purified state. Around 60% of the LPG comes from natural gas processing facilities [2], and the figure is expected to rise as LPG supplies from natural gas production worldwide are expected to increase. Although globally a surplus of LPG exists, there is a scarcity at domestic levels, and many countries need to import it. Several parameters affect the quality of LPG being produced at the NG processing plants.

The LPG process starts with Natural gas extraction. Natural gas is found deep underground in reserves at high pressure. Its composition and properties are highly dependent on the unique geography, type, depth and rock formation of the particular reserve. Generally, Natural gas extracted from the earth comprises lighter hydrocarbons such as methane, ethane, propane, and butane. Along with these light hydrocarbons, it is also coupled with the presence of heavier hydrocarbons such as pentane and hexane. Nonhydrocarbon contents such as Nitrogen, Carbon dioxide, hydrogen and Hydrogen sulfide also commonly occur within raw natural gas. In order to meet sales gas specifications and satisfy transportation requirements, Natural gas needs to be treated. The composition of the treatment section largely depends upon the impurities contained in the natural gas. However, a Natural Gas Processing Plant (NGPP) usually comprises four major sections: Sweetening, Dehydration, Refrigeration and NGL Fractionation [3].

NG processing, fractionation and the subsequent production of LPG consume high amounts of energy. This energy intensiveness is majorly attributed to the refrigeration loop and fractionation areas of the process, where compressor, condenser and reboiler duties add to the cost and energy of the plant [4]. This necessitates a need for a rigorous thermodynamic analysis to improve energy efficiency. Refined energy efficiency not only contributes to the environmental, economic and technical efficiency of the plant but also improves plant life. To preserve the characteristics of LPG, which contains a blend of propane and butane, various studies have been conducted to examine the trade-off points between product quality, energy efficiency and maximum profitability so that the plant is running at optimum conditions. The summary of Sales LPG Characteristics is described in the table below:

Table 1.1: LPG Consumer Market Target Specifications

SR #	PROPERTY	TARGET SPECIFICATION
1	Vapor Pressure	124 psig at 70F 192 psig at 100F 286 psig at 130F
2	Specific Gravity	0.509 at 60F
3	Initial Boiling Point	-51F at 1 bar
4	Dew Point	-46F at 1 bar
5	Specific Heat	0.588 Btu/lb at 60F 2.462 KJ/kg at 15.6C
6	Lower Limit of Flammability	2.4 vol% gas in air
7	Upper Limit of Flammability	9.6 vol% gas in air
8	Latent Heat of Vaporization	185 Btu/lb 430.3 KJ/kg
9	Gross Heating Value (Liquid)	21,550 Btu/lb 50,125 KJ/kg
10	Gross Heating Value (Gas)	2,560 Btu/ft^3 9,538 KJ/m^3

While some studies have focused on observing the effects of the main parameters of the feed stream (such as Temperature, pressure, and Flow rate) on energy consumption of the heat exchange equipment, others have concentrated on determining the retrofit solutions or proposed different refrigeration schemes to optimize the efficiency of LPG recovery from Natural gas. Relocation of the process equipment has proven to be an efficient method of NGPP Optimization. It has been demonstrated that repositioning critical equipment results in the reduction of capital costs by up to 20% and has the potential to increase LPG yield by 5%-10% [5]. Pinch Analysis has long been used as a technique to reduce energy consumption of the process by tracking energy flows. The Pinch Analysis (PA) can be extended to determine energy integration opportunities from heat exchanger networks to low-temperature processes in the process plant. It examines the consumption of hot and cold utilities and assists in identifying opportunities to reduce them. However, the major limitation of pinch analysis is that it only considers the temperature of the streams as a parameter for calculation. This is where exergy analysis proves to be instrumental. Exergy essentially is the available energy for useful work. Hence, it can be used to examine the energy performance of all process streams, considering all the critical properties such as temperature, pressure, and composition. Hence, a rational approach is to use a combination of Pinch and exergy analysis to identify thermodynamically under-performing streams and unit operations and devise strategies for improvement [4], [6].

1.2 Objectives

The objective of this thesis is to apply pinch-based energy, exergy, and economic analysis to optimize the energy consumption of a distillation column in a petroleum refinery. The study aims to design an energy-efficient distillation column that meets the product specifications while minimizing energy consumption.

- To review the literature on production optimization, economic profitability, and energy efficiency of Natural Gas Processing Plant with a focus on LPG.
- To develop a rigorous steady-state model of the Natural Gas Processing Unit in ASPEN HYSYS. Consequently, validation of the model to affirm consistency.
- To perform an economic analysis of the plant utilizing ASPEN HYSYS to identify cost-intensive process unit operations.
- To perform Energy Analysis using ASPEN HYSYS Energy Analyzer (AEA). Employing Pinch Analysis to identify Minimum Energy requirements and potential for energy savings.

- To carry out exergy analysis to identify exergy destruction in process units. Determine exergy efficiency for the process and locate areas where thermodynamic performance can be enhanced.
- To provide recommendations for future research on techno-thermo economic analysis of Natural gas processing Plants.

1.3 Organization of the Thesis

This work is arranged into five chapters, the details and short description of which can be found in the table below: -

Table 1.2:Thesis Organization

Chapter 1	**Introduction** This chapter acts as a preamble to this research work. The chapter includes the motivation, background, previous research history, and objectives and underscores the work carried out during the research.
Chapter 2	**Literature Review** Chapter 2 details the literature study to understand model simulation, energy and economic optimization, production efficiency, pinch and exergy analysis.
Chapter 3	**Methodology** Chapter 3 illustrates the approach adopted to accomplish objectives. It dawns with a detailed description of the process, development of process simulation and its validation, ASPEN Process Economic analyzer and ASPEN Energy Analyzer characterization, exergy and pinch analysis.
Chapter 4	**Results and Discussions** In this chapter, the results of the energy analysis and economic analysis are discussed.
Chapter 5	**Conclusion** The last chapter discusses the summary of the thesis along with future recommendations.
The references and appendices are provided at the end of the thesis.	

CHAPTER 2: LITERATURE REVIEW

2.1 Literature review

In the global energy market, Natural gas and associated Natural Gas Liquids (NGL) are emerging as a desirable fuel option due to their clean burning properties, high heating value (9400 kcal/m^3) and lower environmental impact. Along with the increasing domestic use, there is a growing demand for natural gas utilization as feedstock and fuel gas for chemical and petrochemical industries. Natural gas meets around 27.2% of the total global energy demands [7].

Natural gas is found in deep underground reservoirs as an amalgam of various hydrocarbons and nonhydrocarbon contents. In order to bring it to the use of domestic and industrial consumers, natural gas has to be treated to meet unique sales specifications as well as stringent environmental parameters. The treatment of natural gas is designed based on the content of foreign undesirable impurities that come with the raw gas. Typically, Natural gas needs to be treated for the removal of water (H2O) to avoid problems posed by water vapor, such as pipeline plugging, corrosion, decreased combustion efficiency and formation of hydrates. Similarly, H2S and CO2 content needs to be brought within acceptable ranges, 4ppm and 5%, respectively, to avoid harmful environmental impact and corrosion in the downstream equipment. Naturally, this necessitates that the Natural gas Treatment plant consists of four sub-systems.

i) A Natural Gas Sweetening Unit: To remove H2S and CO2 (acidic) content by contacting alkaline amines.
ii) Natural Gas Dehydration Unit: To remove water content from the gas by utilizing Glycol absorption.
iii) NGL Fractionation Unit: To recover valuable products known as Natural gas liquids (NGL) from dry gas, as these products contribute to the economy by acting as a feedstock in refineries and other process industries. Another vital reason to recover these products is to maintain the dew point of sales gas to avoid hydrate formation.
iv) Refrigeration Cycle/Gas Compression Cycle.

To carry out this treatment, the gas processing facility comprises a number of various processes and equipment, such as distillation towers, separators, pumps, valves, exchangers and compressors. A need for heat integration to improve the thermal efficiency of the process arises since the facility is operated at high pressures and sub ambient temperatures and there exists an opportunity to reduce energy consumption across the process flow.

To boost economic benefits associated with the production of Natural gas and its associated liquids, researchers have long exhibited interest in investigating the parameters that lead to the performance improvement of the process. Several studies have been made to minimize energy consumption and costs associated with the processing of natural gas. Natural gas fractionation train is the most energy-intensive process in the Natural gas processing. The distillation methods employed for the recovery of Natural gas Liquids are studied in detail with the purpose of gauging the effect of main feed parameters on energy consumption and plant profitability. In research conducted by **I. B. Tuychiyev et al.** [8]**,** they considered minimum energy consumption for reboilers and condensers of the propane-butane fractionation within the fractionation train and using this as an objective, calculated optimum reflux ratios and NGL product purity.

Sensitivity analysis for all four treatment sections in the NGPP, namely, Sweetening, dehydration, NGL Recovery and sales gas compression, was performed by **Mozammel Mazumder et al.** [9]. Distillation columns were selected to carry out sensitivity analysis separately for the four sub-systems. The optimized designs revealed an increase in the economic performance of the plant by 20%. Most researchers have focused on studying optimization opportunities within a single subsystem of an NGPP rather than targeting the optimization of the plant as a whole. R.K Abdul Rehman [3] discussed different amine blends and properties, such as amine circulation rates and amine concentration, to analyze economic benefits in a Gas Sweetening Unit. If cold energy utilization is maximized and variables related to high-pressure absorbers are adjusted accordingly, economic profits owing to the recovery of 90% NGL can be achieved. Cooling of Natural gas is directly proportional to the recovery of NGL products. Processes based on Joule-Thompson expansion, turbo-expansion and external refrigeration have been widely investigated to achieve the objective of efficient cooling. To rigorously evaluate the thermal performance of the process, exergy is used as an efficient tool as it combines both the first and second laws of thermodynamics to determine the quantity and quality of energy utilization. Distillation columns contribute to the maximum exergy destruction in the process owing to the mass and heat transfer occurring within them, and the thermal performance can be improved by varying inlet temperature and pressure [10]. Takwa et al. have aimed to use exergy as a crucial parameter for evaluating the thermo-economic performance of the LPG Production process. The results pinpointed major exergy destruction for rectification section columns. Suggestions were made to improve the performance by decreasing reboiler temperatures, resulting in an improvement in exergy efficiency [11].

ASPEN HYSYS is effectively used to simulate the process and apply optimization strategies allowing to analyze results to produce better efficiencies. Maximizing heat integration within the process by controlling the outlet temperatures of refrigeration cycle exchangers and reboilers in the fractionation section gave a positive propane recovery of 85%. The cost estimation of the plant was then run in ASPEN Process Economic Analyzer to give a capital cost of 49 million USD [12]. In another work, Usman Ali et al.[5] in a bid to enhance LPG production yield various schemes of resizing and repositioning of critical equipment as well as rerouting of cooling streams within the flowsheet. The three simulation cases developed in ASPEN HYSYS indicated a decrease in hot side utilities by 10%,14% and 20%, contributing to the decrease in overall operating cost. Another superior strategy for augmenting plant profitability while preserving product quality is to use multi-objective optimization (MOO). This approach makes use of a simulation package such as ASPEN HYSYS and an optimization-based simulation-tracking tool in the form of MATLAB or Python to produce insightful results under uncertain conditions [13]. Similarly, the built-in optimizer tool in the Aspen Hysys module has also been effectively used to improve plant economy by increasing recovery rates and reducing energy consumption. The study underscores the crucial nature of the temperature and pressure of streams entering the fractionation unit and their significant impact on recovery rates of C^{3+} components, hence intensifying economic performance.

In the context of efficient energy performance, Pinch analysis has been used as an important tool to determine energy-saving opportunities in the gas processing plant. It can be used to compare the energy performance of various optimized configurations of the NGPP, as discussed above. One coherent method of determining the configuration is the “driving force method”. The driving force method bases its sequencing technique on the difference in properties of the components in a mixture. Essentially, it is the best approach while modeling separation equipment. When used in conjunction with the thermal Pinch analysis, studies have reported a strong 48% energy savings and 58% capital cost savings compared to simple configuration [14]. Although Pinch Analysis has been employed to determine energy efficiency and saving potential for heat exchanger networks accurately, it limits itself only to a few thermodynamic analyses. Hence, many authors have undertaken studies encompassing a combined pinch-exergy analysis to optimize work and thermodynamic efficiency [15].

Although extensive work has been done to investigate optimum operating conditions and the effects of varying parameters on products' recovery rate and energy consumption of the plant, however, these studies have been limited in their scope to increase the sensitivity of the analysis, i.e., these works have focused on a single sub-system of the natural gas processing plant. In this work, thermo economic analysis for the plant wide natural gas processing has been presented.

CHAPTER 3: METHODOLOGY

3.1 Process Description

The process of treating Natural gas and recovering valuable liquids from it begins with the extraction of raw gas from the well. Before entering the stream directly into the process equipment, a general analysis of its composition is performed to measure the quantity of liquid gas and water contents. Hence, a saturator is first placed where the stream is saturated with water so that moisturization downstream can be prevented and the system does not run in an in-efficient state. The stream then flows through a three-phase separator where water, heavier hydrocarbons and gas are separated. The gas then flows into four treatment sections, namely natural gas sweetening, dehydration section, refrigeration and NGL recovery section. A general flow diagram of the process is presented in Figure 3.1, following which the process occurring in the four sections is briefly discussed.

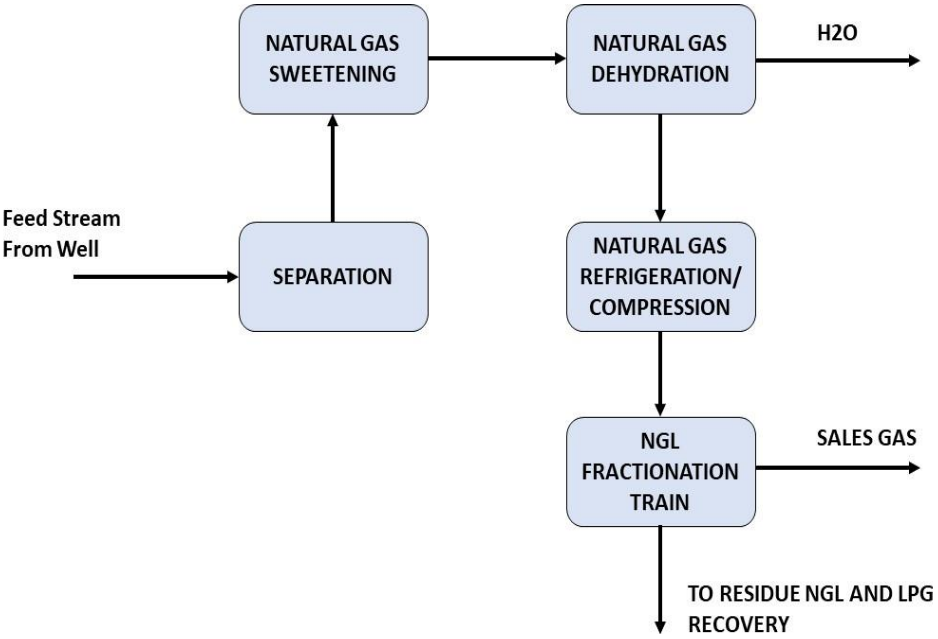


Figure 3.1: Block Flow Diagram of NGPP

Natural Gas Sweetening

The raw natural gas from the wellhead is a blend of numerous organic and inorganic contents. Among these, H_2S and CO_2 need to be removed to prevent corrosion in the downstream equipment (caused by acid formation when these gases react with water) and minimize environmental impacts. Corrosion in the equipment or pipeline can lead to stress cracking. Several processes exist in the industry for the removal of these acid gases, such as the solid bed process or the chemical solvent process. However, the Amine scrubbing process is the most frequently used method with desired success. The Sour gas (Containing H_2S and CO_2) enters the sweetening unit, where it is treated with alkanamine chemical solvents, causing the affinity of amines to absorb acid gases from it. The most commonly used amine solutions to achieve the desired purity of the sour gases are Monoethanolamine (MEA) and Diethanolamine (DEA). A typical sweetening unit consists of a Sour gas absorber or a contact tower, an amine regenerator and heat-exchanging units. The typical flow diagram can be seen in Figure 3.2.

Natural Gas Dehydration

After it goes under treatment in the NG Sweetening unit, Natural gas still contains moisture. In order to protect the downstream equipment from damage and to meet transportation requirements (Residual Humidity < 7 $lb/MMsft^3$), it needs to be dehydrated. The selection of the appropriate dehydration method is dependent on the initial water content present in the feed gas and the required water dew point downstream. While molecular sieve adsorption dehydration is preferred where very low dew point is targeted (ranges$<-50^oC$), economically, it is more viable to select adsorption by Tri-ethanol glycol (TEG) due to its high hygroscopicity and low solubility in natural gas. Similar to the sweetening unit, the natural gas dehydration system also comprises a dehydration absorber, a glycol regeneration tower and a gas stripper. The glycol solution absorbs water and then sinks to the bottom of the contactor, and dry, dehydrated gas is then transported out for further processing.

Natural Gas Refrigeration

In order to achieve maximum recovery of liquid hydrocarbons from natural gas, it is pertinent to perform cooling at cryogenic temperatures ($<-100^oF$). Cooling may be achieved using various techniques such as the Joule-Thomson cooling process, which utilizes a turbo expander to cool the gas stream or mechanical refrigeration, which makes use of chillers for achieving the required temperatures. The refrigerants employed are usually propane-butane or a mixed refrigerant. In general, a combination of these cooling strategies is employed together. The chilled gas, now having major liquid content, is sent to a separator where liquid NGL is separated from natural gas and sent for further processing to the NGL Fractionators.

NGL Fractionation

Natural Gas Liquids (NGLs) are valuable hydrocarbons like crude oil or natural gas and are mainly comprised of carbon and hydrogen. These components are present in the liquid state within gas either from the start of the process or are liquified during the refrigeration cycle. NGLs are mainly ethane, propane, butane, or heavier hydrocarbons. All these components pose as a valuable product since these can be sold as fuel to domestic or industrial users. Hence, the separation of these components is vital to increase the profitability of the plant. The process of separating these hydrocarbons from a mixed NGL stream is NGL fractionation. It uses a train of distillation columns to separate each component based on its volatility. Each distillation column is named after the product it gives off, such as de-ethanizer and de-propanizer.

A comprehensive process flow demonstrating the equipment involved in gas processing is shown in Figure 3.2

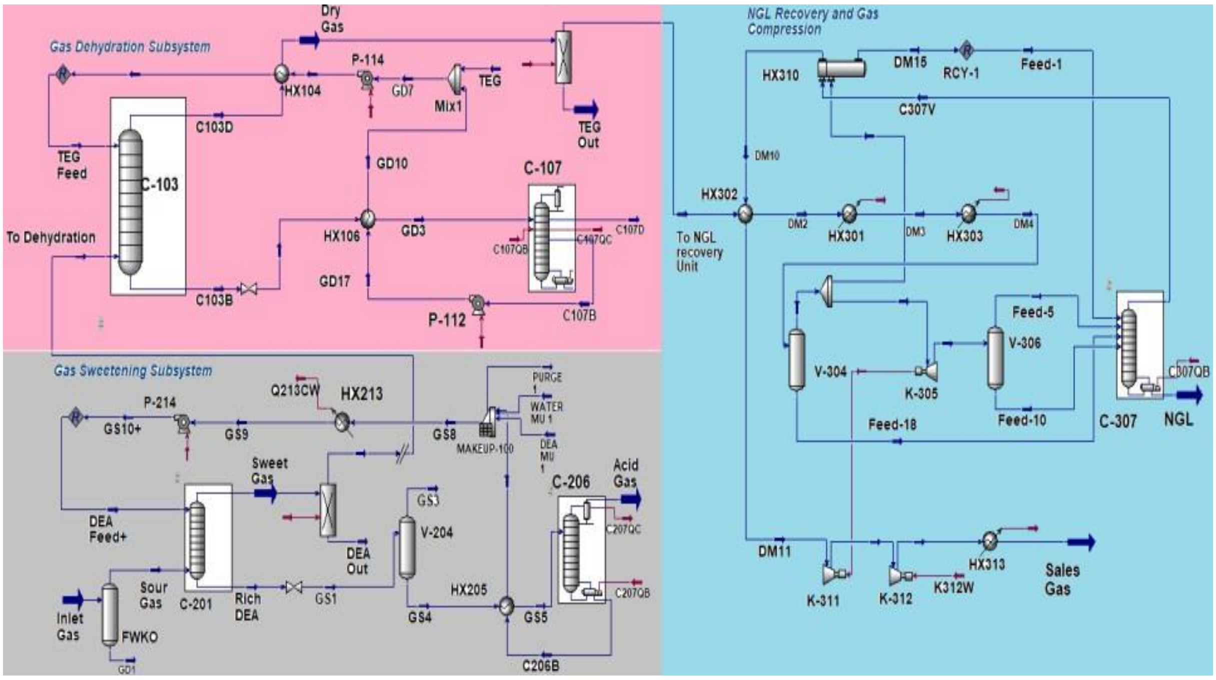


Figure 3.2: Process Flow Diagram of NGPP

3.2 Methodology

The research work is carried out by executing four major steps. The Initial step is to produce a rigorous steady-state model of the Natural gas processing plant using ASPEN HYSYS. The model is validated against real time data to verify the functioning of the model. The next step is to perform an energy analysis in the ASPEN Energy Analyzer environment, where pinch analysis is utilized to determine minimum energy requirements. Exergy analysis, which is an efficient indicator of thermal performance, is also performed using ASPEN-Excel functions to identify locations where energy consumption can be improved. The third step is to optimize the energy and NGL recovery efficiency of the plant while keeping a constraint on the product quality. The last step revolves around comparing the economic performance of the base and optimized case by utilizing the ASPEN Process Economy Analyzer (APEA). This process is also graphically demonstrated in Figure 3.3

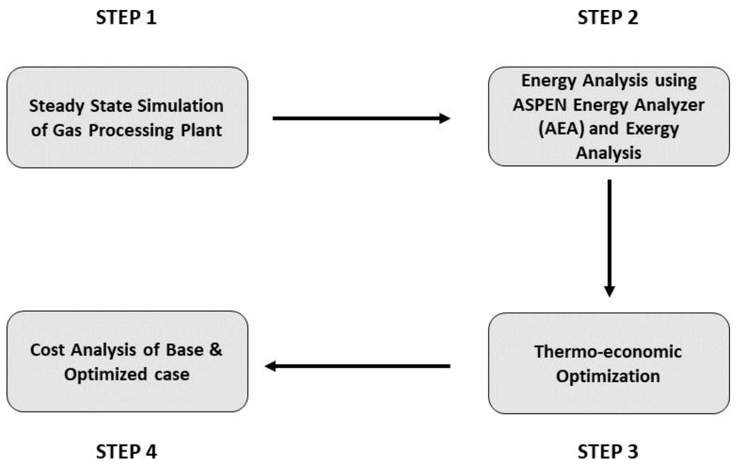


Figure 3.3: Flow Construct of the Analysis

3.3 Process Simulation

In this study, Aspen HYSYS Version 10 is used for the steady-state representation and analysis of the gas processing plant. To begin with, the feed stream coming from the well head is at a high pressure of about 61 bar and has a vapor content of 0. 9821. As the gas first enters the sour gas sweetening section, the “Acid gas – Chemical solvents” property package in the fluid basis is selected to calculate thermodynamic properties. It provides thermodynamics data based on the Electrolyte Non-Random Two-Liquid (Electrolyte NRTL) model and considers necessary aqueous-phase equilibrium and kinetic reactions essential for precise calculations. It incorporates the Peng-Robinson Equation of state for the estimation of vapor-phase properties. The feed stream composition and other crucial properties are demonstrated in Table 3.1 and Table 3.2 below.

Table 3.1: Feed Stream Composition

Component	Mole Fraction	Liquid Phase	Aqueous Phase
Nitrogen	0.0010	0.0010	0.0000
CO2	0.0287	0.0292	0.0007
H2S	0.0157	0.0160	0.0010
Methane	0.7520	0.7657	0.0009
Ethane	0.0970	0.0988	0.0001
Propane	0.0640	0.0652	0.0000
i-Butane	0.0105	0.0107	0.0000
n-Butane	0.0065	0.0066	0.0000
i-Pentane	0.0030	0.0031	0.0000
n-Pentane	0.0015	0.0015	0.0000
n-Hexane	0.0008	0.0008	0.0000
n-Heptane	0.0006	0.0006	0.0000
H2O	0.0187	0.0009	0.9973
DEAmine	0.0000	0.0000	0.0000

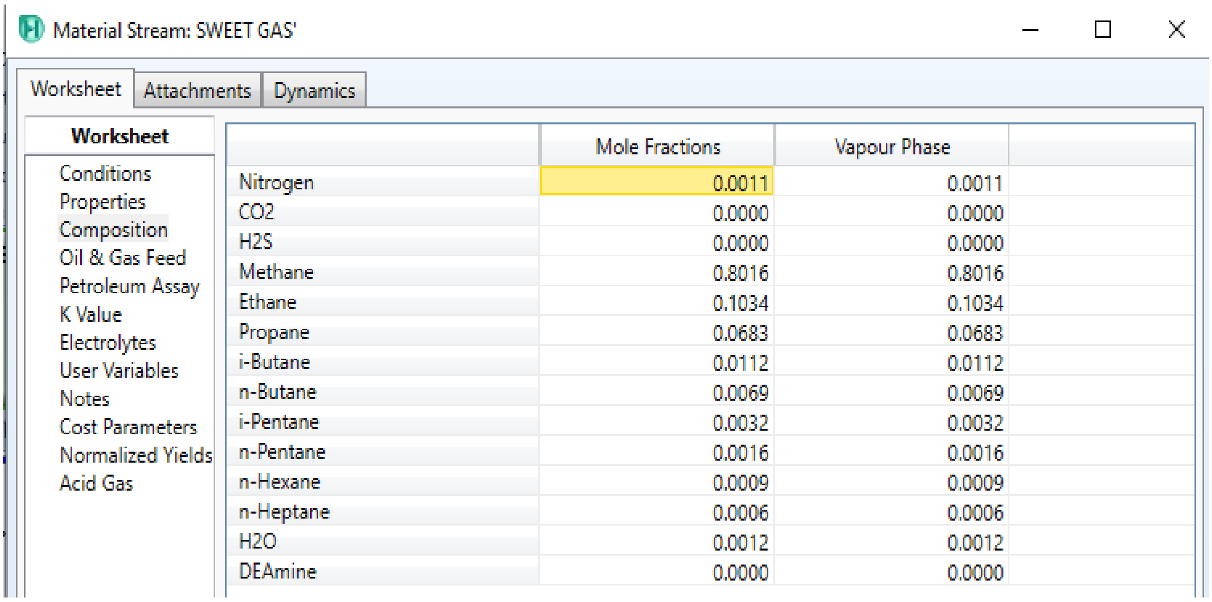
Table 3.2: Feed Stream Conditions

Property Name	Feed Gas from Wellhead	Vapour Phase	Aqueous Phase
Temperature [C]	29.44	29.44	29.44
Pressure [bar_g]	61.04	61.04	61.04
Molar Flow [kgmole/h]	12700	12473.10797	226.8920279
Mass Flow [kg/h]	271881.4995	267786.2053	4095.294136

The feed gas is first passed through a feed water knockout drum modeled as a separator in ASPEN HYSYS. The purpose is to remove tangible water content from the feed stream. The Outlet stream from the separator shows that 99 % of the water content is removed. Then, the stream is entered into a T-100 Amine contactor, which is described as a conventional absorber. The absorber is a trayed tower with 20 stages with a valve tray configuration. The sour gas enters from the bottom stage (20th stage), whereas the Lean DEA solution (3.69% concentration) enters from the top inlet stage with a recirculation rate of 64304 barrels/day. The Acid gas loading for the amine solution is 0.021 (moles of acid gas/mole of amine), which is sufficient for the given composition of the feed stream. The pressure drop within the column is 0.69 bar.

A Rich DEA stream leaves the bottom of the contactor tower after absorbing acid gases from natural gas. The stream contains 1.8% CO_2 and 0.9% H_2S and leaves at a pressure of 60.35 bar and 50.90°C. It is always economical to regenerate DEA from this stream after stripping acid gases and recycling the DEA to an amine contactor. Prior to regeneration, a separator is installed to separate Light Hydrocarbons from the Rich DEA Stream. For this purpose, a valve is installed just upstream of a separator for flash separation. The high-pressure drop in the valve brings the stream from 60.35 to 5.190 bar, and hence, the vapor phase content (which is mostly light HCs) increases. The Overhead stream coming out of the separator has 66% methane and a flowrate of 597kg/hr. The Rich DEA stream coming out from the bottom, which is now devoid of valuable hydrocarbons, is sent to a heat exchanger to raise its temperature. The heat exchanger is modeled as a process-process shell and tube-type heat exchanger (E-100) where the temperature of the Rich DEA stream is raised from 50.90 °C to 93.33 °C. This stream then enters a distillation tower, which represents the DEA Regenerator. The stripping of acid gases from the DEA solution happens in an 18-stage column, where a rich DEA stream enters at the 4th inlet stage. A full reflux configuration is selected at the top with a reflux ratio of 0.505, meaning the overhead product of the regenerator is an acid gas containing 25% H_2S and 46% CO_2. This acid gas stream will be sent to the sulfur recovery unit for further treatment, which, being out of the scope of this study, will not be discussed further. A regenerated DEA stream (RDEA), coming out of the bottom of the column and having a 3% concentration of DEA is employed as a hot stream in the shell side of the exchanger. The temperature of RDEA is lowered from 124 °C to 84.9 °C. The temperature is further lowered to 34.30 °C to be able to recycle this lean Regenerated DEA back to the Amine contactor (T-100). Figure 3.5 displays the process flow diagram of the natural gas sweetening unit simulated in ASPEN HYSYS.

The sweet gas stream leaves the contactor (T-100) as an overhead vapor product at 35.17°C and a pressure of 59.66 bar. The composition of sweet gas is shown in Figure 3.4, which shows the successful removal of acid gases from natural gas.



	Mole Fractions	Vapour Phase
Nitrogen	0.0011	0.0011
CO2	0.0000	0.0000
H2S	0.0000	0.0000
Methane	0.8016	0.8016
Ethane	0.1034	0.1034
Propane	0.0683	0.0683
i-Butane	0.0112	0.0112
n-Butane	0.0069	0.0069
i-Pentane	0.0032	0.0032
n-Pentane	0.0016	0.0016
n-Hexane	0.0009	0.0009
n-Heptane	0.0006	0.0006
H2O	0.0012	0.0012
DEAmine	0.0000	0.0000

Figure 3.4: Sweet Gas Stream Composition

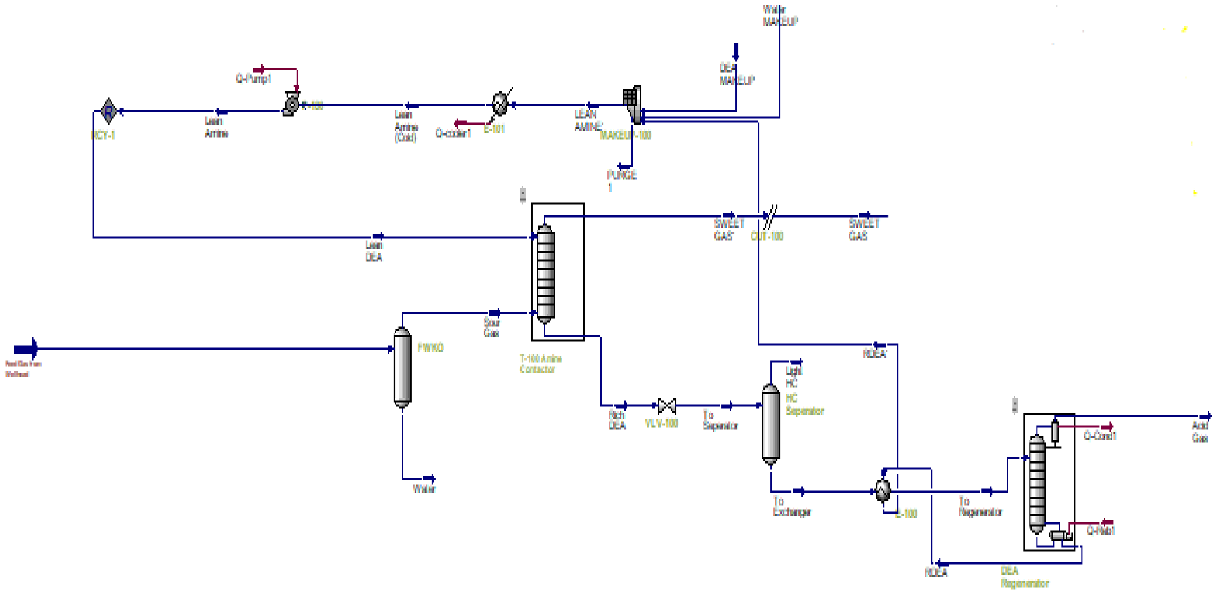
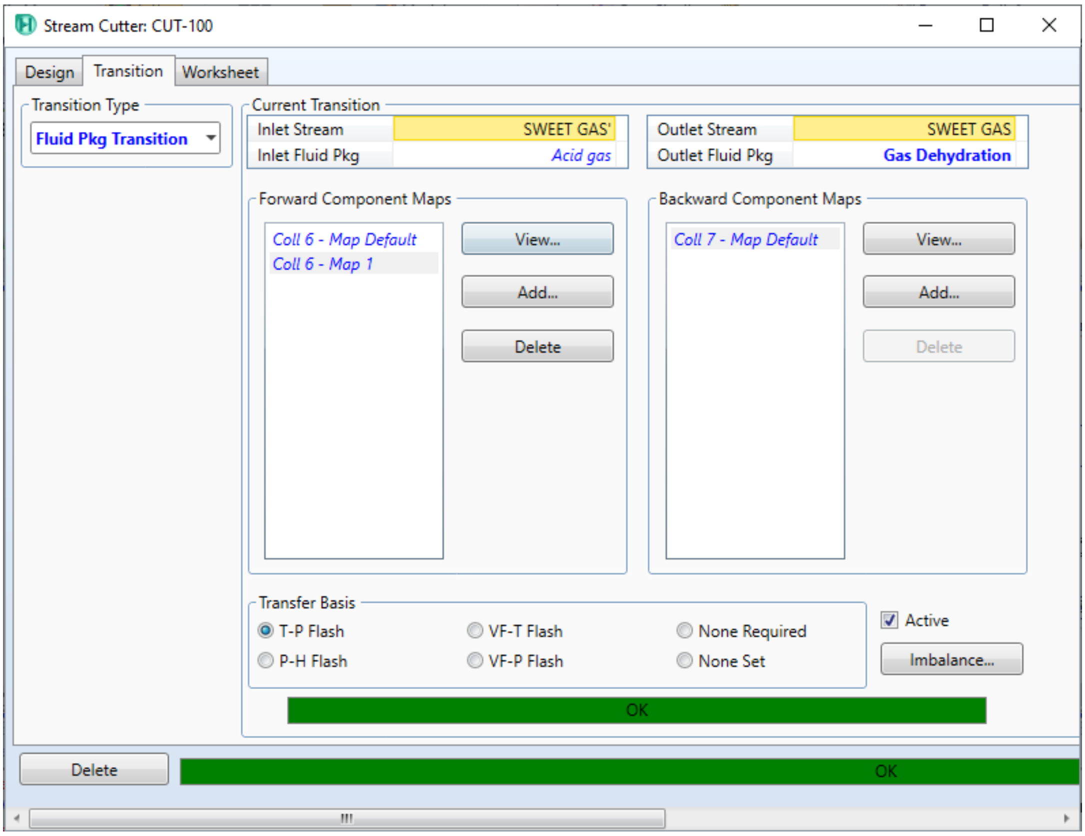


Figure 3.5: Natural Gas Sweetening PFD in ASPEN HYSYS

From the sweetening unit, the sweet gas stream has to go through a fluid package transition from the acid gas package to a more suitable dehydration package, which can represent the thermodynamic conditions and kinetic reactions more accurately. For this purpose, a glycol package from the HYSYS fluid packages' library is selected. Although Peng Robinson EOS can also be used to model dehydration effectively, the glycol package gives a slight advantage by calculating dry gas water content more precisely. A stream-cutter is available in the model palette used to apply this transition. The stream cutter environment can be seen in Figure 3.6.


Stream Cutter: CUT-100

Transition Type: Fluid Pkg Transition

Current Transition

Inlet Stream: SWEET GAS'
Inlet Fluid Pkg: Acid gas
Outlet Stream: SWEET GAS
Outlet Fluid Pkg: Gas Dehydration

Forward Component Maps
- Coll 6 - Map Default
- Coll 6 - Map 1

Backward Component Maps
- Coll 7 - Map Default

Transfer Basis
● T-P Flash
○ VF-T Flash
○ None Required
○ P-H Flash
○ VF-P Flash
○ None Set

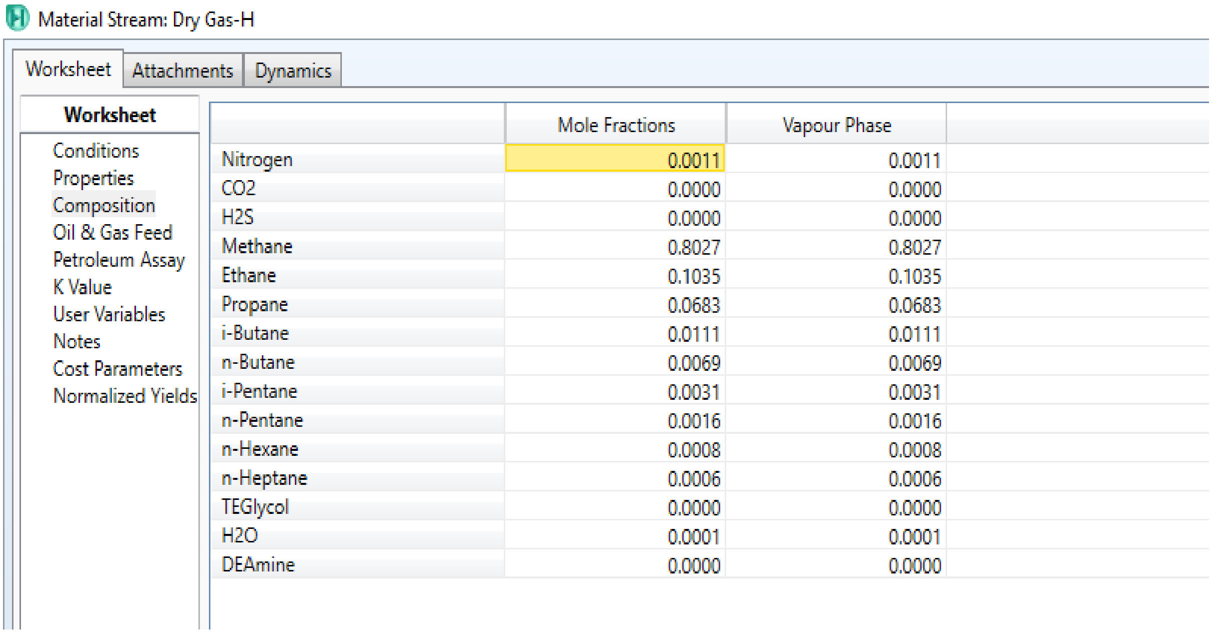
☑ Active

Figure 3.6: Stream Cutter- Fluid package Transition (Acid Gas to Glycol

Then, similar to the sweetening process, the stream is sent to a glycol contactor, where, along with a TEG stream, it gives off a dry gas stream as overhead product and a hydrated TEG stream from the bottom. The column is considered a valve-type 14-stage absorber, where the Aqueous TEG stream is entered through a top stage inlet while the sweet gas enters from the bottom stage inlet. The dry gas composition verifies that the H_2O content of the natural gas is reduced to 3.669 lb/MMSCF from 55.26 lb/MMSCF, which is well within the technical pipeline specifications of the gas. The Dry gas is heated from 43.44 ^{o}C to 53.31 ^{o}C through a process-process shell and tube heat exchanger (E-103). The tube-side heating fluid, in this case, is the regenerated TEG, which needs to be cooled down to 48.89 ^{o}C to enable recycling to the glycol contactor again. The hydrated glycol, which is the bottom product of the glycol contactor absorber, is at a temperature of 42.39 ^{o}C and pressure of 59.66 bar. The pressure needs to be decreased so that the vapor content within this stream is maximized to achieve maximum regeneration. A linear valve is installed which decreases the pressure to 0.779 bar To achieve this. After that, the low-pressure hydrated glycol stream is heated up to a temperature of 104 °C by means of a shell and tube heat exchanger (E-102). This further increases the vapor content. Increasing the vapor content is vital since it will make regeneration more profitable by reducing the size and minimizing the energy consumption of the equipment required for regeneration. The Hot stream then enters a TEG Regenerator, which is modeled as a distillation column. It is a 2-stage total reflux distillation column, where the Hot Aqueous TEG stream enters at the 1st tray.

A condenser temperature of 104 °C and a reboiler temperature of 234 °C are used as active specifications to converge the column successfully. The overhead product from the condenser is a saturated stream containing 68% water and 11% Methane. The bottom product of the regenerator is a super-heated 93.9% pure TEG stream at a temperature of 234 °C. This stream is then passed through the shell side of heat exchangers E-102 and E-103 to reduce the temperature to 48.89 °C. The pressure also needs to be increased to recycle the regenerated TEG to the Glycol Contactor. For this reason, a pump P-102 is installed upstream of the tube side of E-103. The pump raises the pressure to 61.03 bar and consumes a power of 38.87 KW.

Figure 3.7 shows the composition of the product of the Natural gas dehydration unit.



	Mole Fractions	Vapour Phase
Nitrogen	0.0011	0.0011
CO2	0.0000	0.0000
H2S	0.0000	0.0000
Methane	0.8027	0.8027
Ethane	0.1035	0.1035
Propane	0.0683	0.0683
i-Butane	0.0111	0.0111
n-Butane	0.0069	0.0069
i-Pentane	0.0031	0.0031
n-Pentane	0.0016	0.0016
n-Hexane	0.0008	0.0008
n-Heptane	0.0006	0.0006
TEGlycol	0.0000	0.0000
H2O	0.0001	0.0001
DEAmine	0.0000	0.0000

Figure 3.7: Dry Natural Gas Stream Composition

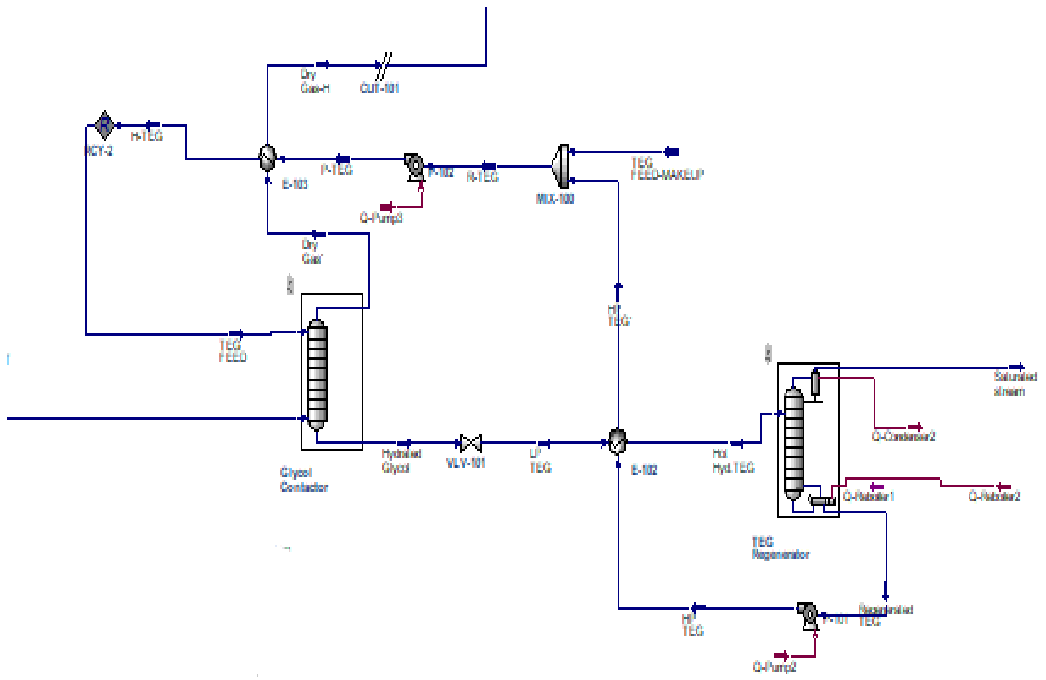


Figure 3.8: Natural Gas Dehydration PFD in ASPEN HYSYS

Similar to the earlier fluid package transition, the dry gas, which was previously used in the glycol fluid package undergoes a change of fluid package to Peng-Robinson Equations of State (PR-EOS). Since oil & and gas processing is a complex process containing stream interactions in single, two- or three-phase systems, the PR equation is one of the most favored in the oil gas industry. It gives an efficient, reliable, and accurate solution of these solutions over a wide range of operating conditions (T > -271°C and P < 100 bar). For cryogenic systems (temperatures < -271 °C) or high-pressure systems, a modified version of PR namely Peng-Robinson-Stryjek-Vera (PRSV) is employed for calculating thermodynamic properties. The PR-EOS is stated as below:

$$p = RT\, V - b - a\, V(V + b) - b(V - b) \qquad (3.1)$$

where,

$$a = \sum_{i=1}^{N} \sum_{j=1}^{N} x_i x_j (a_i a_j)^{0.5} (1 - k_{ij}) \qquad (3.2)$$

$$b = \sum_{i=1}^{N} x_i b_i \qquad (3.3)$$

The PR equation can also be expressed in the form of compression factors:

$$Z^3 - (1 - B)Z^2 + (A - 3B^2 - 2B)Z - (AB - B^2 - B^3) = 0 \qquad (3.4)$$

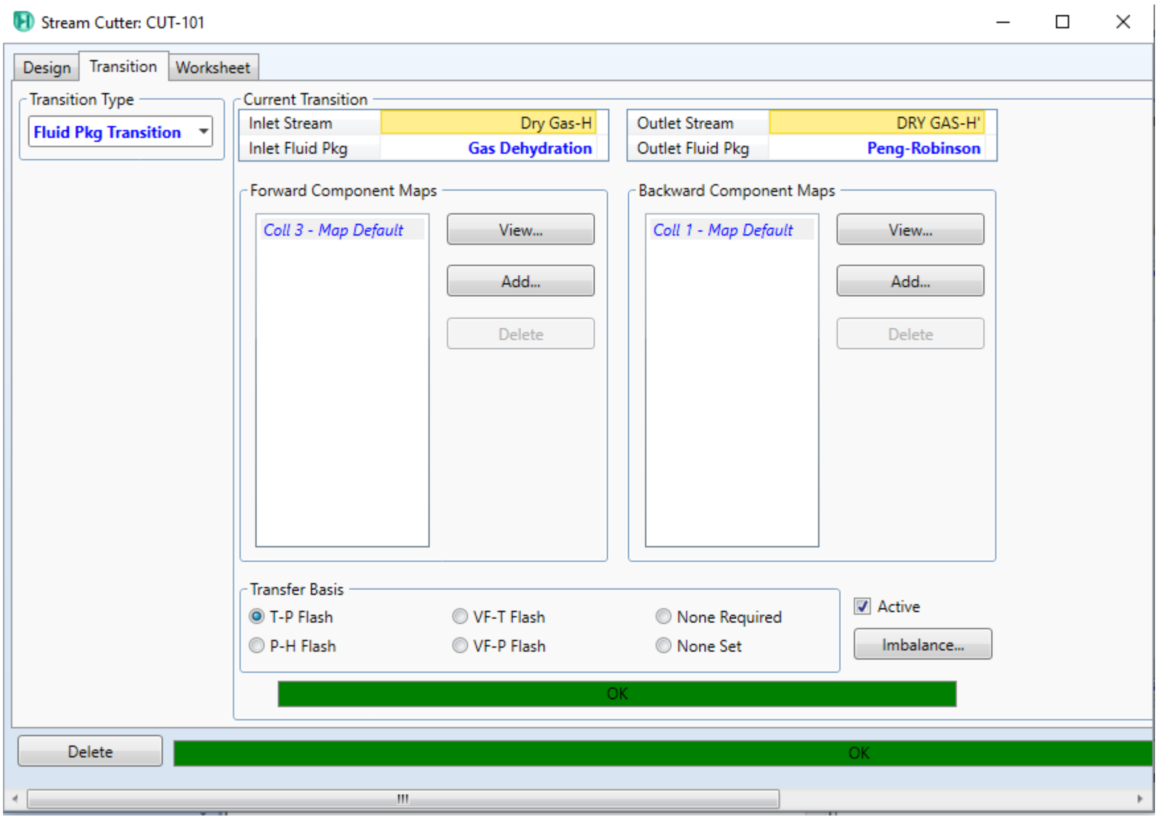
Where,

$$Z = \frac{pV}{RT} \qquad (3.5)$$

$$A = \frac{ap}{(RT)^2} \qquad (3.6)$$

$$B = \frac{bp}{RT} \qquad (3.7)$$

Similar to what happened earlier, the transition of the stream fluid package from the glycol package to Peng-Robinson occurs in a stream cutter set, the configuration of which can be seen in Figure 3.9.



Stream Cutter: CUT-101

Transition Type: Fluid Pkg Transition

Current Transition
Inlet Stream: Dry Gas-H
Inlet Fluid Pkg: Gas Dehydration
Outlet Stream: DRY GAS-H'
Outlet Fluid Pkg: Peng-Robinson

Forward Component Maps: Coll 3 - Map Default

Backward Component Maps: Coll 1 - Map Default

Transfer Basis
● T-P Flash
○ P-H Flash
○ VF-T Flash
○ VF-P Flash
○ None Required
○ None Set

☑ Active

Figure 3.9: Stream Cutter- Fluid package Transition (Glycol to Peng-Robinson)

The dry gas is then sent to a refrigeration unit employing a combined turbo-expansion and mechanical refrigeration strategy. The major objective of this unit is to enable the gas to meet standard technical specifications required for appropriate pipeline transportation. Generally, this includes maintaining the dew point of gas at low temperatures to prevent hydrate formation in the pipeline. For this, the dry gas, which is initially at a temperature of 32.22 °C is made to pass through a series of three Heat exchangers where it loses heat to reach a temperature of -33 °C. Even at this temperature, the Hydrocarbon Dew point is at 12.74 °C. Although these conditions are ideal for further processing and recovery of Natural Gas Liquids (NGLs), to bring sales gas within pipeline transportation requirements, we need to reduce the dew point further. For this purpose, a separator (V-100) is installed downstream of the three heat exchangers which serves to strip the gas from the liquid that has been generated due to cooling. The liquid stream mostly contains NGLs, while the overhead gas stream contains 95% methane and is suitable for transportation as sales gas. Before this, the last stage includes reducing the temperature to -40.38 °C, which also lowers the dew point to -88.95 °C. Then, the sales gas stream is compressed to 61.04 bar by means of 02 centrifugal compressors, operating with an adiabatic efficiency of 75%. Figure 3.10 gives the PFD of the refrigeration section of the natural gas processing.

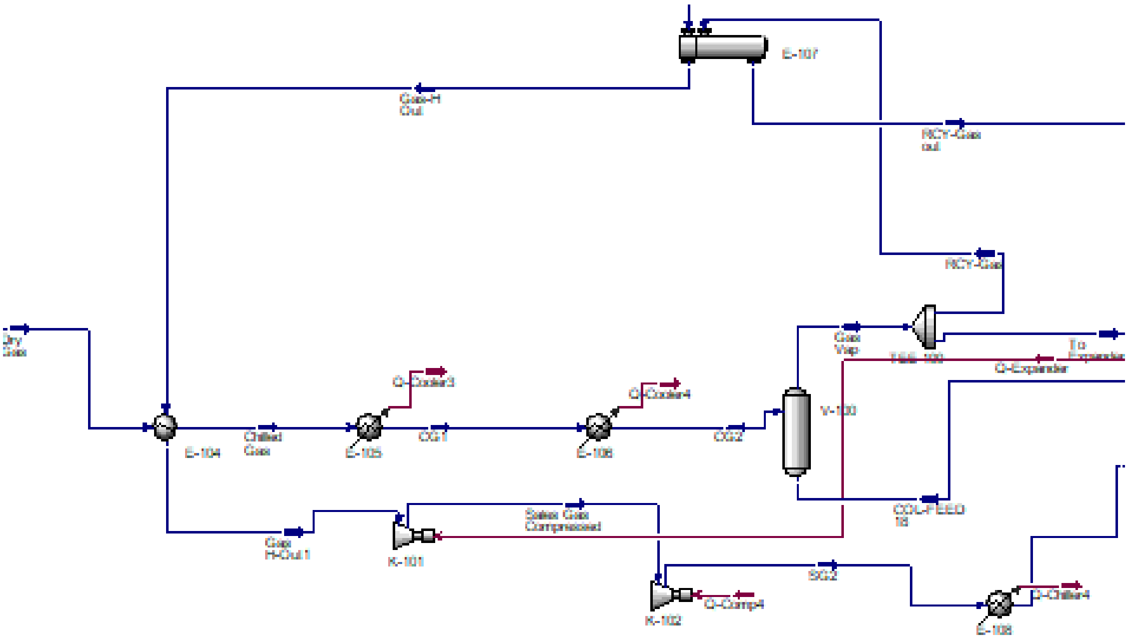


Figure 3.10: Process Flow Diagram of NG Refrigeration Loop in ASPEN HYSYS

The separation of natural gas (CH_4) is completely achieved when the overhead vapor product of the separator V-100 is sent into an NGL recovery column. The recovery column is simulated as a 30-stage reboiled absorber, where the feed enters at four stages (stages 1,5,10, 18). The boil up ratio is set at 0.8 to converge the column with success. The reboiler stage temperature is 4.44 °C. The overhead vapor product of this column is 97% methane gas, coming out at -88.44 °C and 24.29 bar. This will processed as discussed previously and exited from the system as sales gas. The NGL, which comes out as the bottom product, has a composition that contains 75% ethane and propane. Hence, this stream will proceed further in the NGL Fractionation train and enter the De-ethanizer and De propanizer columns, respectively, for the extraction of LPG. The De-ethanizer is represented by a 14-stage full reflux distillation column where the NGL enters at the 6th stage. Ethane is extracted as the overhead product with an 85.67% purity and is sent to storage. The bottom stream of the De-ethanizer is in a liquid state containing 71.25% propane and sent to a de-propanizer for LPG Extraction. The de-propanizer is a distillation column where NGL Enters at 64.21 °C and 15.89 bar. The column is modeled as a 24-stage total condensing distillation column. The NGL Feed is entering from the 11th Inlet stage. The overhead condenser produces a liquid stream only since it is a total condensing unit. A reflux is sent back to the column with a reflux ratio of 0.007. The remaining liquid is extracted as LPG. The Full Process Flow diagram of the Natural gas processing plant is shown in Figure 3.12 and Figure 3.13.

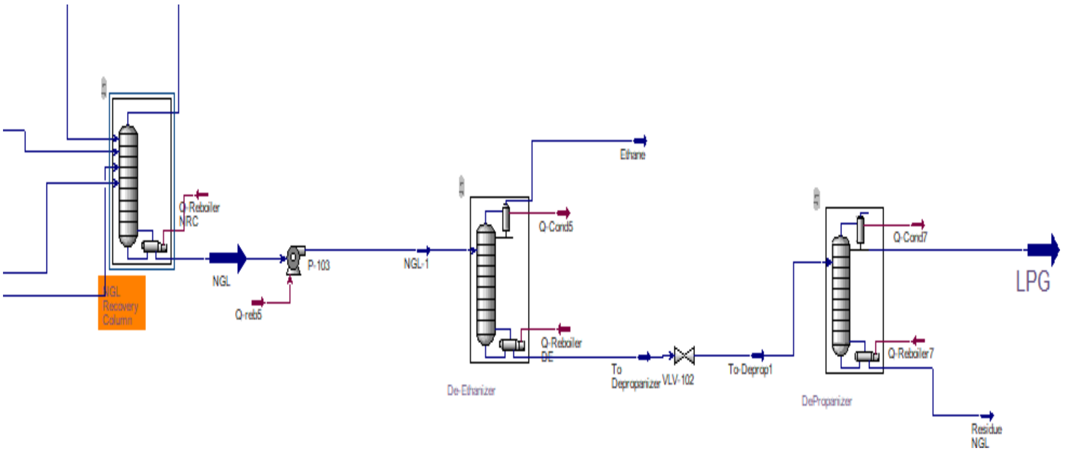


Figure 3.11: Process Flow Diagram of NGL Fractionation Train in ASPEN HYSYS

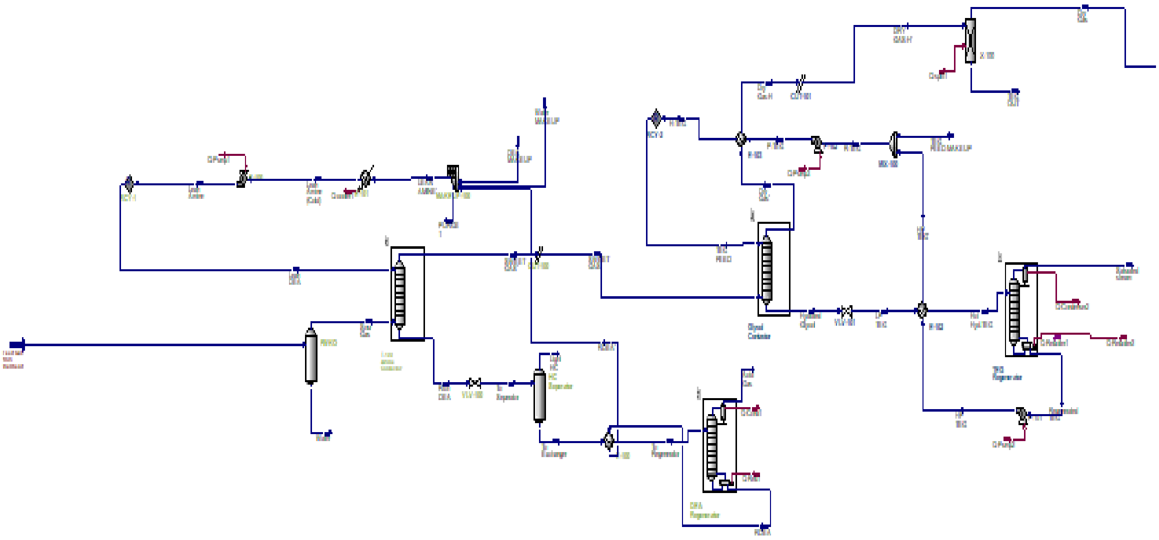


Figure 3.12: Process Flow Diagram of NGPP ASPEN HYSYS

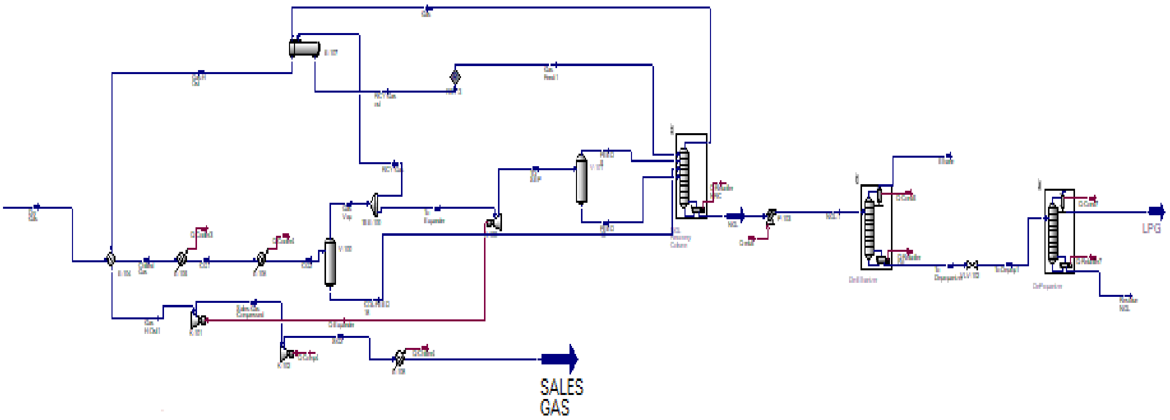


Figure 3.13: Process Flow Diagram of NGPP ASPEN HYSYS

3.3.1 Simulation Validation:

The data obtained from the simulation is validated against actual plant data to verify that the model developed in ASPEN HYSYS is precise. The products' LPG and sales gas streams were compared for both the simulation and actual mass flow rates. The results indicate very little difference percentages which means we can proceed with performing further analysis on the Natural gas processing plant simulation developed in ASPEN HYSYS. The actual data was gathered from the literature for the khurmala gas plant in Iraq. The difference percentage was calculated using the following expression.

$$Difference\ \% = \frac{Simulation\ Value - Actual\ Value}{Actual\ Value}\ \text{x}\ 100 \tag{3.8}$$

Table 3.3: Simulated Model Validation Results

Product	Actual (kg/hr)	HYSYS Results (kg/hr)	Difference %
LPG	40,456.74	40,458.53	0.0044
Sales Gas	159,908.16	159,914.91	0.0042
Residue NGL	9482.18	9505.27	0.2435

3.4 Economic Analysis

ASPEN Process Economic Analyzer (APEA) was utilized to perform Cost analysis of the developed base and optimized cases. This costing module allows to integrate economy with the developed simulation by estimating capital and operating costs of the designed setup. APEA is based on Icarus technology, which does not base its calculations on the capacity-factored curves for equipment sizing or bare equipment costs. Instead, it estimates cost by following industry-standard design codes and manufacturing procedures to represent the cost of equipment.

The main steps involved in the integrated economic analysis are listed below: -

1- Activation of the cost engine
2- Providing the required Material and utility Streams’ cost estimates.
3- Mapping unit operations to equipment as per best industrial practices.
4- Sizing of the equipment.
5- Running the economic evaluation and analyzing results.

After activating the cost engine, as the data from the ASPEN HYSYS simulation is transferred to the Economic Analyzer, the module develops estimates based on industry-standard and geography basis to estimate material, utilities, and equipment costs. The most crucial step in Economic analysis is the mapping of the unit operations. Although APEA offers the convenience of automatically mapping and sizing the equipment, it is strongly recommended that the equipment be mapped manually, keeping in view the operational requirements of the process. For example, the heat exchangers can be selected as shell and tube type plate type or spiral exchangers depending upon the process requirement. After mapping has been completed successfully, auto-sizing can be selected, which will trigger the APEA sizing module to set equipment dimensions in accordance with the process requirements and selected mapping. Any sizing errors or missing values will be indicated for which appropriate revised values can be input in the module. Triggering the evaluation gives an estimate of both capital and operating costs along with the estimation of certain economic performance parameters, such as the payback period and total product sales. The economic evaluation, in its calculation, incorporates both direct and indirect costs, such as: -

- Material and labor costs for Equipment and setting, piping, civil, structural steel, instrument and controls, Electric equipment, paint, and insulations etc.
- Indirect non-field costs, for example, freight, taxes, Basic and detailed Engineering, Material acquisition and other contingencies.

The cost of utilities required in the process is estimated by the module based on the utilities selected by the user or as recommended by the sizing expert.

3.5 Energy Analysis

Aspen Energy Analyzer (AEA):

AEA uses pinch embedded technology in ASPEN HYSYS for heat integration and optimization of energy consumption. It identifies targets to achieve Maximum Energy recovery and provides energy saving potential and also suggests solutions for minimum consumption using the pinch technology. These solutions may propose addition of heat exchangers to minimize overall utility requirements or imply design alternatives for improved performance. In general, the AEA allows to input thermal data for streams (Inlet, Outlet Temperatures, and enthalpy) and then estimates the pinch considering minimum energy requirements. However, another beneficial utilization of AEA is to predict potential energy saving potential for the existing network by analyzing the simulated utilities data. To perform this analysis, we will follow the steps as listed: -

- Before activating Energy analysis, the utilities that apply to the simulation will be edited in the utility manager. If no utilities are selected, the energy analyzer will self-estimate the utilities’ requirements and choose the best matches.
- Input the “process type” while configuring the AEA. Inserting the correct process type is crucial since it will impact the accuracy of the results. In the case of natural gas

processing, a “low-temperature process” will be selected because of the refrigeration loop and generally low temperatures occurring through the process.

- The scope of the energy analysis will be defined.
- The energy analyzer will be toggled to generate energy savings potential and savings summary.

Pinch Analysis:

Pinch analysis is an efficient and commonly practiced method for heat integration and energy optimization. It determines a pinch point, where the temperature difference between the hot streams and cold streams is the lowest. The analysis produces energy targets for all unit operations and suggests retrofit or alternate designs for energy-efficient performance.

Exergy Analysis:

Energy Analysis based on thermodynamics describes the efficiency, performance, and behavior of the systems by analyzing the conversion of energy from one form to another. Conventional energy analysis is constructed on the principles of the first law of thermodynamics (conservation of energy). A first law-based energy analysis is essentially identifying and accounting for the energies entering and exiting the system. Although simple, their exist inaccuracy and inefficiencies in employing conventional energy analysis due to the thermodynamic losses that are not assessed (mainly in wastes and products). Also, in utilizing the first-law analysis, the system energy balance is not calculated to determine locations of system inefficiencies. In simpler terms, locations where maximum energy can be saved are not identified. Therefore, a meticulous approach is to use exergy analysis to determine thermodynamic performance of the system. The exergy analysis integrates the first and second law of thermodynamics to determine the system’s thermodynamic efficiency. A major objective of exergy analysis is to identify and characterize the exergy destructions in the process to improve overall efficiency (Figure 3.14.). Exergy can be attributed to the maximum amount of useful work generated by a stream or a system when it is brought to equilibrium with a reference. It essentially gives the measure of usefulness or quality of energy and acts as a stringent indicator of thermodynamic performance of the system. For real processes, exergy is always destroyed due to the irreversibility’s that are coupled with the process. Exergy is mathematically expressed as: -

$$E = E_{PH} + E_{CH} + E_{KN} + E_{PT} \qquad (3.9)$$

Since there are no significant chemical reactions occurring during the process, and natural gas processing is essentially a separation and heat exchange process, the chemical, kinetic and potential exergies are omitted from the expression. Physical exergy is termed as the maximum work that can be extracted from a stream when its pressure and temperature is brought to ambient conditions. It is represented as: -

$$Ex_{ph} = m[(H - H_0 - T_0(S - S)] \qquad (3.10)$$

Where "m" is the mass flow rate of the stream, "h" is the enthalpy, "T" represents the temperature and "S" is the entropy of the stream.

To quantify exergy analysis, we will make use of performance parameters such as exergy efficiency, exergy destruction and exergy improvement potential.

Exergy Efficiency: -

Exergy efficiency, or in other words, also termed as second-law efficiency or rational efficiency is the ratio of thermal efficiency of an actual system to an ideal (reversible) system. It is the measure of how effectively a process consumes energy. In terms of exergy, it can be expressed as the ratio of output exergy to input exergy: -

$$\eta = \frac{E_{out}}{E_{in}} \times 100 \qquad (3.11)$$

The higher the exergy efficiency, the higher the energy savings in the system, as energy losses and irreversibility will be minimized. A higher exergy efficiency will indicate a sustainable system that produces more useful work.

Exergy Destruction: -

Exergy destruction is the measure of irreversibility and energy degradation of the system. It helps identify the systems within a process that are the most thermally underperforming or where the most resource degradation occurs.

$$I = Ex_{destroyed} = \sum Ex_{in} - \sum Ex_{out} \qquad (3.12)$$

Exergetic Improvement Potential: -

Improvement potential relates directly to the loss of useful work or the irreversibility encountered during the process. It is suggested as a performance indicator quantifying the loss which can be minimized for maximized improvement and optimal thermal performance.

$$IP = (1 - \eta)(E_{in} - E_{out}) \tag{3.13}$$

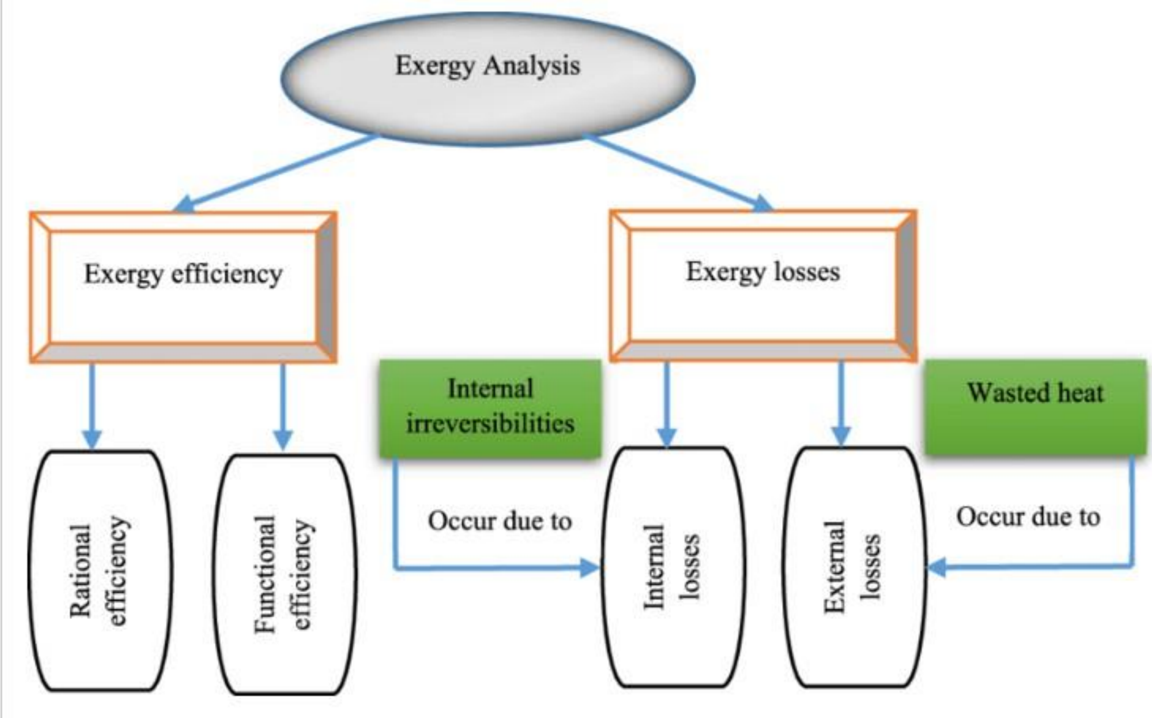


Figure 3.14: Exergy Analysis Performance Parameters

CHAPTER 4: RESULTS AND DISCUSSION

This chapter begins with a discussion of steady-state results for the base case. Section 4.1 discusses the energy analysis and economic analysis of the developed simulation. The economic and energy savings potentials are also highlighted. After that, section 4.2 discusses the development of an optimized design for the enhancement of the thermo-economic performance of the plant. Finally, the last section, 4.3, discusses the findings and optimized case results post-analysis. Also, the limitations and future recommendations are discussed in this section.

4.1 Base Case Simulation Results

4.1.1 Production efficiency:

After the steady-state simulation is developed in ASPEN HYSYS and validated against the actual data, steady-state analysis is conducted for the base model. To measure the production efficiency of the process, product recovery rates for methane and propane are calculated. Since these components are the valuable products recovered and marketed. The product recovery rate (R) is expressed as

$$Product\ Recovery\ Rate\ (R) = \frac{Component\ Molar\ Flow\ in\ Product\ Stream}{Component\ Moalr\ Flow\ in\ Feed\ Stream} \text{ x } 100 \qquad (4.1)$$

The stream conditions and composition for LPG (Natural Gas) are presented in Tables 4.1 and 4.2.

Table 4.1: LPG Product Stream composition

Component	Mole Fraction	Liquid Phase
Nitrogen	0.0000	0.0000
CO2	0.0000	0.0000
H2S	0.0000	0.0000
Methane	0.0000	0.0000
Ethane	0.0083	0.0053
Propane	0.7633	0.7788
i-Butane	0.0994	0.1240
n-Butane	0.0506	0.0631
i-Pentane	0.0123	0.0191
n-Pentane	0.0052	0.0080
n-Hexane	0.0009	0.0017

n-Heptane	0.0001	0.0001
TEGlycol	0.0000	0.0000
H2O	0.0000	0.0000
DEAmine	0.0000	0.0000

Table 4.2: LPG Product Stream conditions

Stream Name	LPG
Temperature [C]	51.82
Pressure [bar_g]	14.84
Molar Flow [kgmole/h]	867.90
Mass Flow [kg/h]	40458.53

The purity of LPG achieved from the simulated Natural gas processing plant is 82.33 % propane, which is acceptable for being marketed as a commercial and domestic fuel. The recovery rate for LPG is calculated using the molar flow of propane in the LPG stream and the molar flow in the feed stream.

$$LPG\ Recovery\ rate\ (R) = \frac{714.16\ kgmole/hr}{812.80\ kgmole/hr} \text{x}\ 100 = 87.86\% \qquad (4.2)$$

So, this gives us a recovery rate of approximately 87% which indicates a comprehensive development of the simulation. Similarly, the recovery rate of other marketable product is calculated, that is, the sales gas. The stream conditions and composition are presented in Tables 4.3 and 4.4, followed by the calculation of the sales gas recovery rate.

Table 4.3: Sales Gas Product Stream compositions

Component	Mole Fraction	Vapor Phase
Nitrogen	0.0013	0.0013
CO2	0.0000	0.0000
H2S	0.0000	0.0000
Methane	0.9731	0.9731
Ethane	0.0247	0.0247
Propane	0.0009	0.0009
i-Butane	0.0000	0.0000
n-Butane	0.0000	0.0000
i-Pentane	0.0000	0.0000
n-Pentane	0.0000	0.0000
n-Hexane	0.0000	0.0000
n-Heptane	0.0000	0.0000

TEGlycol	0.0000	0.0000
H2O	0.0000	0.0000
DEAmine	0.0000	0.0000

Table 4.4: Sales Gas Product Stream conditions

Stream Name	**LPG**
Temperature [C]	51.82
Pressure [bar_g]	14.84
Molar Flow [kgmole/h]	867.90
Mass Flow [kg/h]	40458.53

The sales gas purity achieved from this process is approximately 97 %, and the recovery rate of sales gas (methane) is found similarly as was found for the LPG stream.

$$Sales\ Gas\ Recovery\ rate\ (R) = \frac{9470.17\ kgmole/hr}{9550.40\ kgmole/hr} \text{x}\ 100 = 99.15 \tag{4.3}$$

The sales gas recovery rate is 99.15%, which is extremely efficient since the major economic output of the natural gas processing plant is the treated natural gas. The process is designed to keep the efficiency of natural gas production at an optimum point.

4.1.2 Economic Analysis:

After performing the analysis to evaluate production performance parameters, an economic analysis is carried out to assess the viability of the process. ASPEN Process Economic Analyzer (APEA) is utilized for this. First, the stream prices are determined and given as input to the evaluator. Along with this, the utility stream type is specified in the flowsheet. However, APEA possesses the capability to self-determine the appropriate utilities required for the process from a built-in database. Stream prices are entered for feed and product streams to estimate the cost. The stream prices are referred from the US database for energy costs in 2024 and are presented in Table 4.5. The stream basis for stream cost estimation is set as a mass flow basis. The operating life of the plant is assumed to be 25 years, and the length of plant startup is set at 0.5 years to set up the costing options. The operational year is kept at a 24/7 processing cycle, which means a total of 8760 hours.

Table 4.5: Market Stream Price for Economic Evaluation

FEED STREAM PRICE		
Sr #	**Stream Name**	**Stream Price (USD/kg)**
1	Feed Gas from the Wellhead	0.1503869
2	DEA MAKEUP	0.0909
3	TEG FEED-MAKEUP	0.1089
PRODUCT STREAM PRICE		
Sr #	**Stream Name**	**Stream Price (USD/kg)**
1	LIGHT HC	0.5037
2	SALES GAS	0.2879
3	Ethane	0.0286392
4	LPG	0.1973
5	Residue NGL	39.62580785

For Economic evaluation, all the heat exchangers and condensers/reboilers associated with columns were mapped as TEMA Shell and Tube Heat exchangers. Compressors were mapped as horizontal centrifugal compressors pumps as single or multi-staged centrifugal pumps depending upon the process requirements. All towers were mapped as multi-diameter trayed towers. After successful mapping, auto-sizing of the equipment was selected. HP steam was selected as a hot utility for the reboilers of TEG and DEA Regenerator to satisfy process conditions to eliminate sizing errors. Refrigerant-1 was selected as the cold utility for the condenser of De-ethanizer and Exchanger E-105, while a very low-temperature cold utility type was selected for E-106 to complete successful sizing. Toggling the evaluation button gave the cost estimation, a summary of which is presented in Table 4.6.

Table 4.6: Base Case Economic Analysis Summary

Summary	
Total Capital Cost (USD)	**34,110,600**
Total Operating Cost (USD)	403,468,000
Total Raw Materials Cost(USD)	358,172,000
Total Product Sales (USD)	4,002,970,000
Total Utilities Cost (USD)	12,714,100
Desired Rate of Return	20
P.O. Period	1.10369
Equipment Cost (USD)	11,736,000
Total Installed Cost (USD)	18,760,800

The desired payback period stands at 1.1 years, which demonstrates a lucrative economic opportunity for the Natural gas processing Unit.

4.1.3 Exergy Analysis:

Steady state exergy analysis for the Natural Gas Processing Plant was performed. The Overall exergy efficiency of the process was determined using the expression:

$$\eta = \frac{Output\ Stream\ Exergy}{(Inlet\ Stream\ Exergy + Energy)} \times 100 \qquad (4.4)$$

For calculation of Output and Inlet stream exergy, only physical exergy is considered. Physical exergy, in simpler terms is represented as

$$Physical\ Exergy = Mass\ Exergy * Mass\ Flow\ Rate$$

and,

$$Exergy\ of\ Energy\ Streams = Q\ (1 - \frac{T_o}{T}) \qquad (4.5)$$

Where T_o and P_o are the reference environment temperature and pressure. The reference temperature and pressure are specified as 25^oC and 101.325 kPa. Individual Exergy destruction and efficiencies of the four subsystems which constitute the natural gas processing plant, namely, Gas Sweetening Unit, Gas Dehydration Unit, Natural gas refrigeration and NGL Fractionation Unit are also calculated to determine thermal inefficiencies existing within the plant. The exergy data is presented in Table 4.7 below.

Table 4.7: NGPP Exergy Analysis Summary

OVERALL EXERGY ANALYSIS			
FEED STREAMS (Inlet Stream Exergy)			
STREAM NAME	**MASS FLOW RATE (kg/hr)**	**MASS EXERGY (KJ/kg)**	**PHYSICAL EXERGY (KJ/hr)**
Feed Gas from Well-head	271900	445.4552129	121111031.2
Water Makeup	3849	21.97955389	84594.92118
TEG Feed-Makeup	1.356	0.168373778	0.228264791
Energy Streams Exergy			
Utility Stream	**Power (KW)**	**Heat Flow (KJ/hr)**	**Energy Exergy**
Q-Pump 1	943.29	3395850.43	1740910.66

Q-Cooler 1	23122.46	83240845.89	42674104.54
Q-cond1	4679.31	16845507.81	8635988.18
Q-Reboiler 1	31262.97	112546704.77	57697994.22
Q-Pump 2	0.71	2543.92	1304.16
Q-Pump 3	38.87	139938.32	71740.53
Q-condenser 2	6.84	24617.09	12620.15
Q-Reboiler 2	2062.46	7424872.95	3806422.21
Q-cooler 3	5255.84	18921028.55	9700020.96
Q-Cooler 4	7756.55	27923581.95	14315254.04
Q-Expander	1195.22	4302792.64	2205862.05
Q-Comp 4	6867.70	24723734.62	12674825.98
Q-Chiller 4	5695.25	20502907.25	10510984.10
Q-Reboiler NRC	6748.75	24295496.44	12455286.15
Q-reb 5	16.53	59507.54	30507.03
Q-Cond 5	7638.57	27498857.78	14097515.70
Q-Reboiler DE	12509.75	45035084.72	23087606.73
Q-cond7	3677.10	13237543.21	6786335.44
Q-Reboiler 7	2508.68	9031235.45	4629937.16

PRODUCT STREAMS (Outlet Stream Exergy)

STREAM NAME	MASS FLOW RATE (kg/hr)	MASS EXERGY (KJ/kg)	PHYSICAL EXERGY (KJ/hr)
Water	4095.294136	7.015811625	28731.81221
light HC	596.9330932	192.2215844	114743.4249
Acid gas	26483.14878	77.63382616	2055988.168
purge 1	5.645969361	27.96001645	157.8613962
Sat stream	482.4672445	185.7064735	89597.29056
TEG out	18.31740676	8.173500544	149.7173341
Sales Gas	159911.7981	601.301729	96155240.68
Ethane	34208.00532	249.9892103	8551632.237
LPG	40458.53361	106.2157633	4297334.03
Residue NGL	9505.879079	48.73025919	463223.9514

Then, making use of Eq. 4.4 the overall exergy efficiency of the process was calculated to be 32.26 %. The exergy destruction caused due to the irreversibility of the system amounts to 65,159 KW, and the exergy improvement potential is 44,133 KW. An exergy analysis was also carried out for the four sub-systems to assess the location of the most exergy destruction to analyze the results further. The results are presented in Table 4.8.

Table 4.8: NGPP Sub-system Exergetic Performance

SUB-SYSTEMS EXERGY SUMMARY				
PARAMETER	**Gas Sweetening Unit**	**Gas Dehydration Unit**	**Refrigeration /Compression Loop**	**NGL Fractionation Train**
Inlet stream exergy (kj/hr)	121195626.16	114093938.01	193110728.71	120393089.65
Energy flow exergy (kj/hr)	110748997.60	3892087.06	49406947.12	61087188.20
Output stream exergy (kj/hr)	116024721.41	114680297.41	216548330.33	13312190.22
Exergy efficiency (%)	50.02	97.20	89.29	7.34
Exergy destruction (kw)	32229.09	918.26	7213.71	46713.36
Exergy improvement potential (kw)	16092.71	25.73	772.46	43286.78

From the results, it can be deduced that the Gas dehydration unit is the most thermally efficient, with 97% followed by 89% efficiency of the Natural gas Refrigeration loop. This is due to the fact that the least physical interactions have occurred in these sub-systems, causing minimum irreversibility in the system, which gives a superior exergetic performance. Surprisingly, about half of the exergy destruction for the whole process is caused in the NGL Fractionation Unit, where the exergy efficiency is only a meager 7.34% and the improvement potential is 43,286 KW. This is due to the fact that most exergy destruction is caused by the condensers and reboilers associated with the columns in the NGL Fractionation train. Other factors contributing to the exergy losses can be attributed to the high differences between the feed and outlet temperatures of these columns. This gives rise to a temperature gradient within the column stages, which, coupled with the interaction between the components of feed (due to composition differences), adds to significant exergy losses. Also, the pressure drop within each column is a major contributor to irreversibility and causes low exergy efficiency.

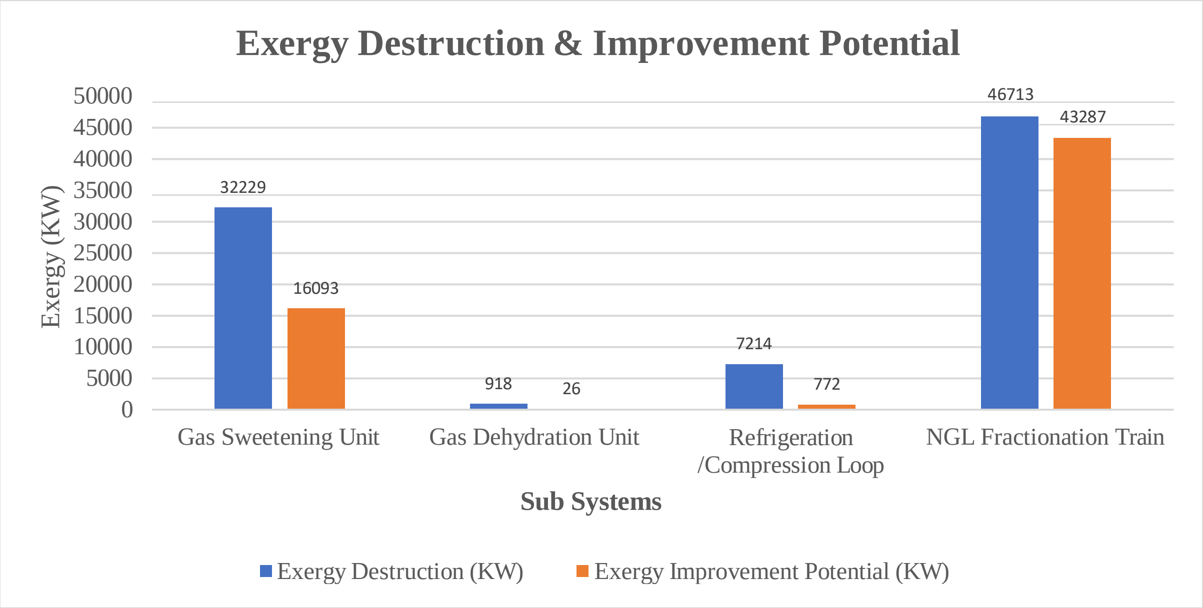


Figure 4.1: Unit classified Exergy Destruction & Improvement Potential of NGPP

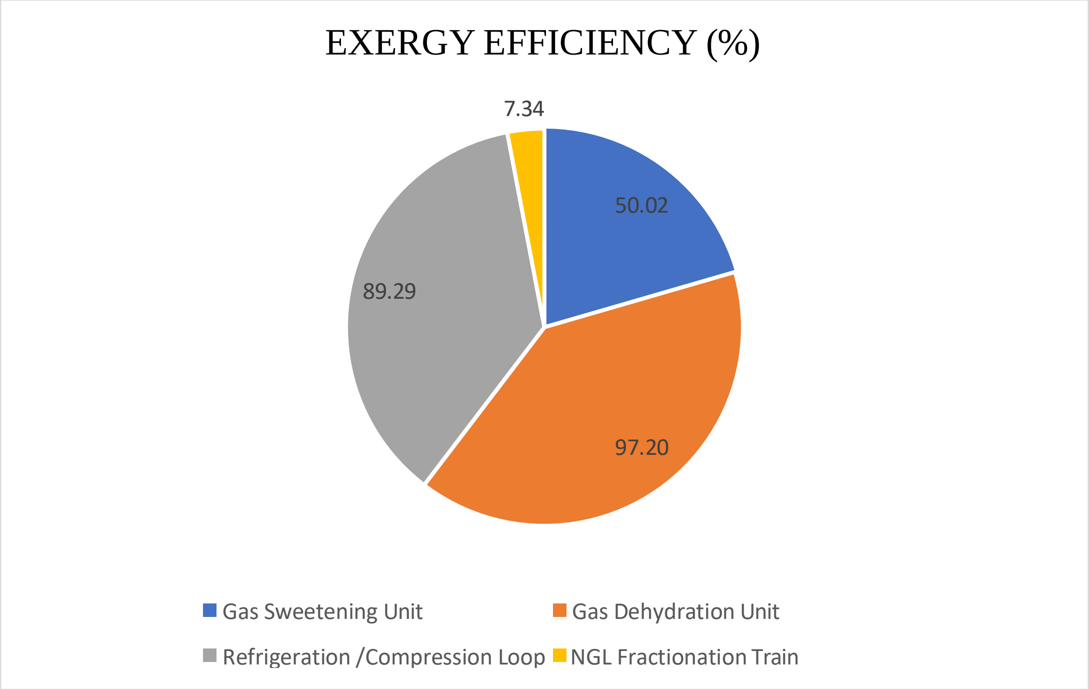


Figure 4.2: Unit classified Exergy Efficiency of NGPP

4.1.4 Activated Energy Analysis:

Since most of the exergy inefficiency is associated with the Natural Gas Fractionation Train, we will perform the activated energy analysis to assess the energy-saving potentials and performance improvement in this unit only. Foremost, from the utility manager in the ASPEN HYSYS Data base, the redundant utilities that are not present in the model are deleted to enable the energy analyzer to select appropriate utilities to satisfy process conditions. Then, the energy analysis case is configured by defining the process as a “low temperature” process since Natural gas processing operates at generally low temperatures. Then, the scope is defined so as to only limit within the fractionation unit. The results of the energy analysis are presented in Table 4.9 and shown in Figure 4.3. The Energy Targets for cold and hot utility streams are calculated using the pinch analysis. The Minimum Energy Requirement (MER), or the quantity of maximum energy that can be recovered, is calculated during the analysis and presented as available savings.

Table 4.9: NGL Fractionation Unit Energy Savings Potential

Property	Actual	Target	Available Savings	% Of Actual
Total Utilities [kW]	63,111.11	50,638.89	12,477.78	19.74
Heating Utilities [kW]	52,861.11	46,611.11	6,250.00	11.78
Cooling Utilities [kW]	10,250	4,022.22	6,227.78	60.75
Carbon Emissions [kg/h]	12,690	10,190	2,502.00	19.71

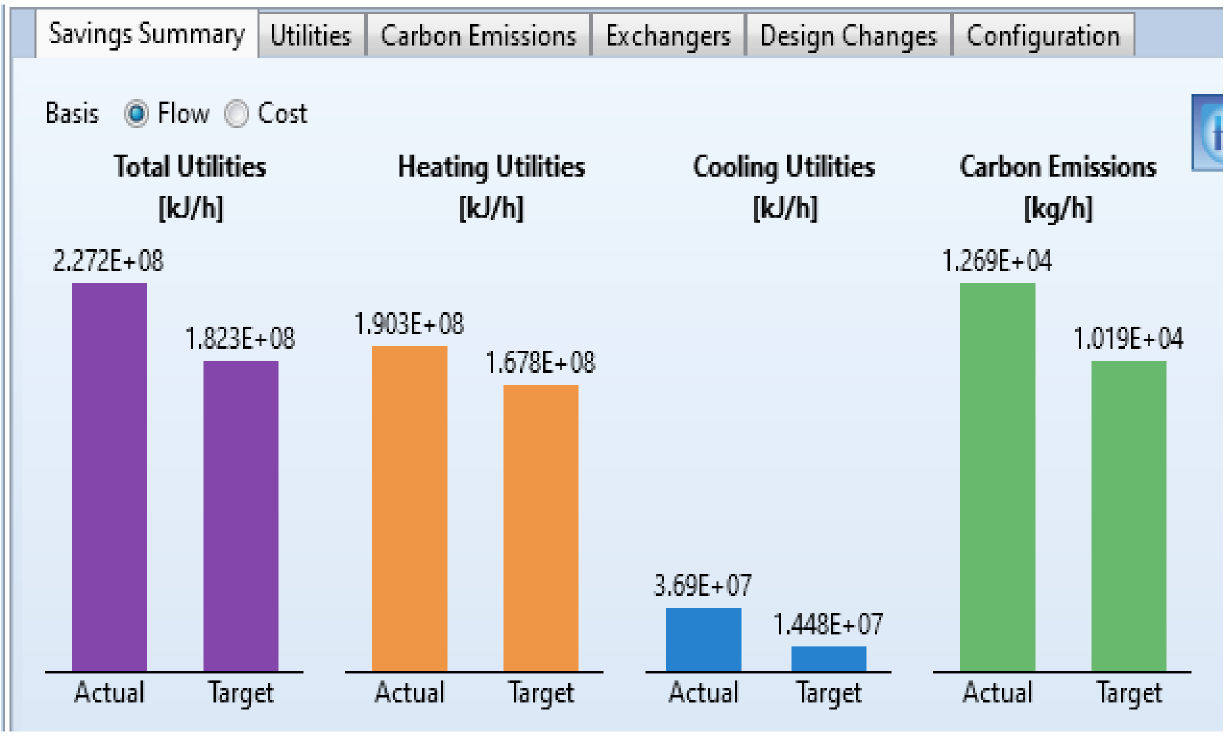

Basis ● Flow ○ Cost

Figure 4.3: NGL Fractionation Train Energy Savings

The energy usage details for each utility can be seen in the Table 4.10.

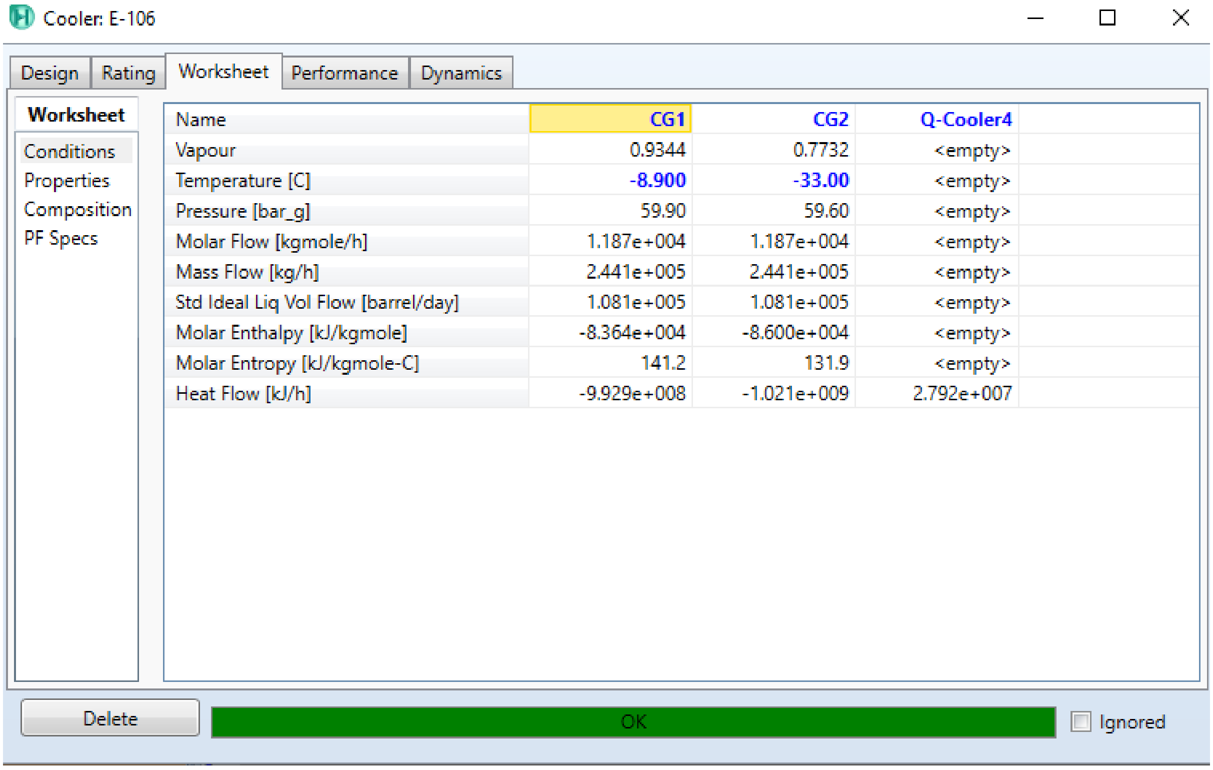
Table 4.10: NGL Fractionation Unit Utility Targets

Utility Type	**Current**	**Target**	**Saving Potential**	**Energy Cost Savings**	**Energy Cost Savings**
	[kW]	**[kW]**	**[kJ/h]**	**[Cost/Yr]**	**[%]**
LP Steam	52861.11	46611.11	6227.78	373,099	11.78
Total Hot Utilities	52861.11	46611.11	6227.78	373,099	11.78
Refrigerant 1	6536.11	2629.72	3905.56	337,433	59.77
Cooling Water	3713.89	1393.33	2320.00	15,544	62.48
Total Cold Utilities	10250.00	4022.22	6227.78	352,977	59.88

4.2 Optimized Design Analysis

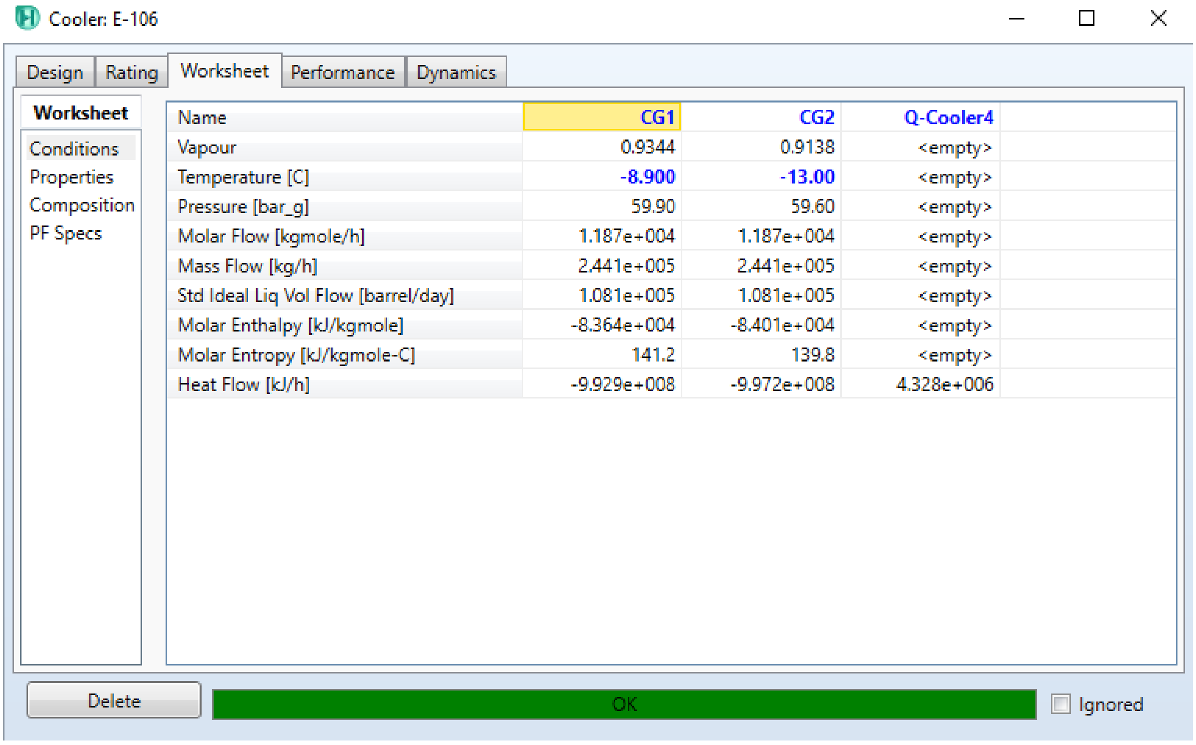
Two retrofit solutions are proposed after the analysis. Since the main NGL Recovery column (COL-5), which separates the Natural gas and NGLs, has the least thermal efficiency due to an extremely high-temperature gradient in its stages, the energy-saving potential is quite high. Similarly, the de-propanizer column, due to the high-temperature difference, also provides an opportunity for energy recovery. Another solution is to preheat the streams entering the NGL Fractionation Unit using heat exchangers or by simply modifying the process conditions upstream of the unit. Due to its simplicity and the limitations of using the ASPEN I/O Algorithm in the study, modification of the process conditions was utilized for the optimized case.

The Cooler E-106 was previously used to cool down the dry gas stream to a temperature of -33°C. As a part of an effort to reduce energy consumption and improve exergy efficiency, the outlet temperature for this cooler was set at -13°C. This ultimately led to the reduction of reboiler and condenser duties throughout the NGL Fractionation Train. The only significant difference this causes is the decrease in the liquid fraction of the product stream of E-106. This reduction consequently impacts the NGL recovery. However, the effect is slight. Regardless, the sales gas recovery rate is improved, which increases the overall economy of the process. The vapor fraction of the cooler E-106 was 0.7732 which consequently changed to 0.9138 when the temperature was increased, as shown in Figure 4.5.



Name	CG1	CG2	Q-Cooler4
Vapour	0.9344	0.7732	<empty>
Temperature [C]	-8.900	-33.00	<empty>
Pressure [bar_g]	59.90	59.60	<empty>
Molar Flow [kgmole/h]	1.187e+004	1.187e+004	<empty>
Mass Flow [kg/h]	2.441e+005	2.441e+005	<empty>
Std Ideal Liq Vol Flow [barrel/day]	1.081e+005	1.081e+005	<empty>
Molar Enthalpy [kJ/kgmole]	-8.364e+004	-8.600e+004	<empty>
Molar Entropy [kJ/kgmole-C]	141.2	131.9	<empty>
Heat Flow [kJ/h]	-9.929e+008	-1.021e+009	2.792e+007

Figure 4.4: E-106 Cooler Base Case Conditions



Name	CG1	CG2	Q-Cooler4
Vapour	0.9344	0.9138	<empty>
Temperature [C]	-8.900	-13.00	<empty>
Pressure [bar_g]	59.90	59.60	<empty>
Molar Flow [kgmole/h]	1.187e+004	1.187e+004	<empty>
Mass Flow [kg/h]	2.441e+005	2.441e+005	<empty>
Std Ideal Liq Vol Flow [barrel/day]	1.081e+005	1.081e+005	<empty>
Molar Enthalpy [kJ/kgmole]	-8.364e+004	-8.401e+004	<empty>
Molar Entropy [kJ/kgmole-C]	141.2	139.8	<empty>
Heat Flow [kJ/h]	-9.929e+008	-9.972e+008	4.328e+006

Figure 4.5: E-106 Cooler Optimized Case Conditions

4.2.1 *Comparison of the base case and optimized designs:*

The increase in temperature causes the purity of methane in natural gas to be affected. This is due to the fact that NGLs, especially ethane, boil off along with sales gas due to the increased temperature. The purity is still within range of sales gas specifications (>70%). The recovery rate is found to be slightly improved as calculated: -

$$Sales\ Gas\ Recovery\ rate\ (R) = \frac{9485.25\ kgmole/hr}{9550.40\ kgmole/hr} \text{x}\ 100 = 99.31\% \qquad (4.6)$$

Quite the opposite, the purity of LPG is increased to 84.14% however, the recovery rate is reduced to 78.02%. This can be attributed to the fact that the slightly higher temperatures of the optimized process have caused heavier hydrocarbons, such as propane, to vaporize and escape with the sales gas and ethane streams. The purity, as a result, increases, but the recovery rate is decreased.

$$LPG\ Recovery\ rate\ (R) = \frac{634.16\ kgmole/hr}{812.80\ kgmole/hr} \text{x}\ 100 = 78.02\% \qquad (4.7)$$

4.2.2 *Exergy Analysis of Optimized Case:*

The same performance indicators will be utilized as were used earlier to evaluate the thermal performance. The results can be reviewed in the Table 4.11.

Table 4.11: Optimized Case Exergy Analysis Summary

NGL FRACTIONATION TRAIN EXERGY SUMMARY			
PROPERTY	**BASE CASE**	**OPTIMIZED CASE**	**% Change**
INLET STREAM EXERGY (KW)	33442.52	11595.54	-65.32%
ENERGY FLOW EXERGY (KW)	16968.66	8599.88	-49.31%
OUTPUT STREAM EXERGY (KW)	3697.83	2502.35	-32.32%
EXERGY DESTRUCTION (KW)	46713.36	17693.06	-62.12%
EXERGY EFFICIENCY (%)	**7.34**	**12.39**	**+68.80%**
EXERGY IMPROVEMENT POTENTIAL	43287	15501	-64.19%

The results demonstrate a significant improvement in the thermal performance of the NGL Fractionation Unit with the manipulation of a single variable. The exergy efficiency is improved by 68%, and the exergy destruction is reduced by more than half, as was in the base case. This is majorly due to the fact that condenser and reboiler duties for all towers have significantly reduced, along with the fact that temperature and pressure gradients within the column stages are not unusually high.

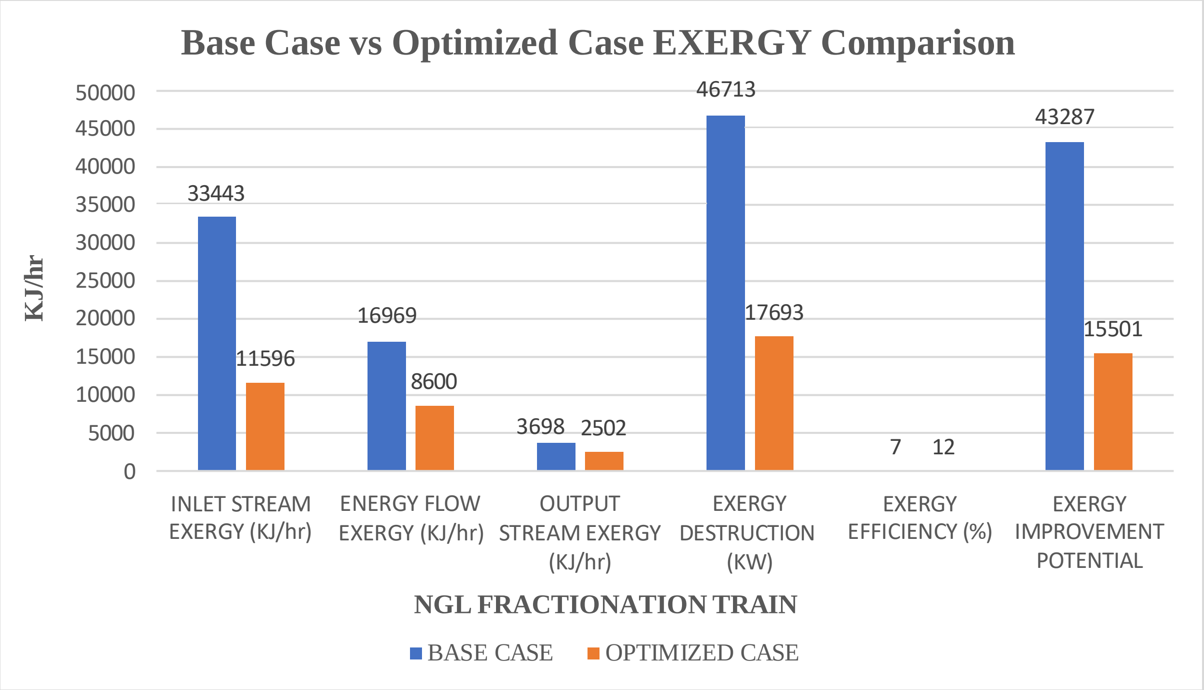


Figure 4.6: Exergy base comparison-Base vs Optimized Case

4.2.3 Optimized Case Energy Analysis:

The Analysis run on AEA gives revised energy targets based on Minimum Energy Requirement and maximum Energy recovery. When the configuration settings and utilities' selection are kept the same as in the base case, results reveal that the actual consumption and target requirement for the total utilities has significantly reduced. This is due to a major decrease in hot utility requirements since the temperature is enhanced. The actual consumption of cold utility has decreased. However, the target calculated has seen growth as well. This is due to the increased energy requirement for cooling water utility streams.

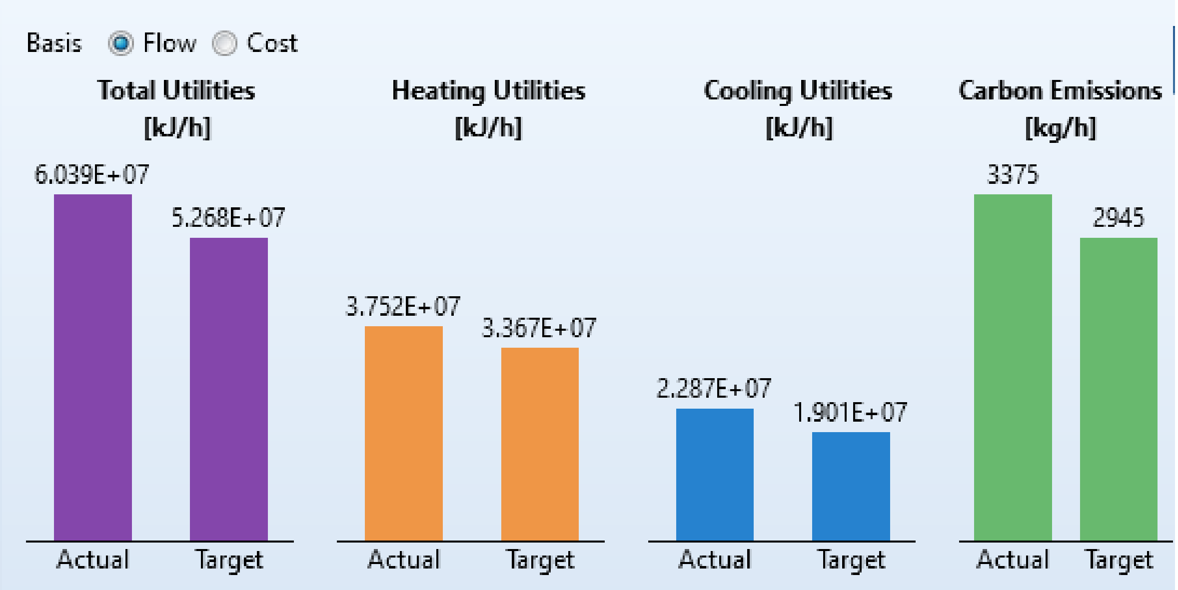


Figure 4.7: Optimized Case AEA Energy Targets Summary

A comprehensive comparison between the energy consumption of actual utilities is provided in Table 4.12. The Heating utilities' energy requirement has decreased significantly due to the fact that the temperatures of streams have increased and are now closer to a Stable reference Environment (SRE). The overall decrease in energy required for utilities to satisfy process conditions is close to 73%.

Table 4.12: Optimized Case Energy Consumption

Comparison of Utility Consumption					
Property	**Base Case**		**Optimized Case**		**% Decrease in Actual Consumption**
	Actual	**Target**	**Actual**	**Target**	
Total Utilities [kW]	63,111.11	50,638.89	16,775.00	14,633.33	73.41%
Heating Utilities [kW]	52,861.11	46,611.11	10,422.22	9,352.78	80.28%
Cooling Utilities [kW]	10,250	4,022.22	6,352.78	5,280.56	38.02%
Carbon Emissions [kg/h]	12,690	10,190	3,375.00	2,945.00	73.40%

4.2.4 Optimized Case Economic Analysis

The economic analysis is performed to evaluate the total annual cost, capital cost, utility cost, operating cost and the payback period of the process. The cost analysis also presents the total product cost of the optimized natural gas process. For this analysis, similar costing options as in the base case were selected. However, equipment mapping and sizing were selected to tailor the optimized process conditions. Aspen Process Economic Analyzer (APEA) was utilized to perform the analysis.

Table 4.13: Optimized Case Economic Performance

Summary	Base Case	Optimized Case	% Change
Total Capital Cost [USD]	**34,110,600**	**25,480,800**	**-25.29**
Total Operating Cost [USD/Year]	403,468,000	411,210,000	**1.92**
Total Raw Materials Cost [USD/Year]	358,172,000	358,418,000	**0.07**
Total Product Sales [USD/Year]	4,002,970,000	4,845,650,000	**21.05**
Total Utilities Cost [USD/Year]	12,714,100	14,751,400	**16.02**
Desired Rate of Return [Percent/'Year]	20	20	**0.00**
P.O. Period [Year]	1.10369	1.08595	**-1.61**
Equipment Cost [USD]	11,736,000	7,181,800	**-38.81**
Total Installed Cost [USD]	18,760,800	12,311,000	**-34.38**

There is a slight increase in the utilities cost per annum for the optimized case since cold utility minimum energy requirements have increased, causing an increase in external energy consumption. This increase subsequently results in the escalation of operational costs/yr. However, there is a significant decrease in capital cost and the enhanced profitability of the process due to an increase in product sales by 21.05%. The results demonstrate a strong economic performance strengthening the optimization strategy.

4.2.5 Discussion of the results

The maximum irreversibility or energy loss was found to be in the NGL Fractionation train of the natural gas processing unit. The total exergy destruction for the process was 65,159 KW, of which 72% of the total exergy destroyed was in the Fractionation Unit. The overall exergy efficiency of the NGPP was 32.26%, where the Dehydration section had the highest thermal efficiency of 97% and the NGL Recovery section operated with the lowest efficiency of 7%. Energy analysis performed for the base case revealed that the energy-saving potential for this section was 19%. This high energy consumption and low efficiency can be attributed to the presence of reboilers and condensers operating to maintain high-temperature gradients within the towers. The option to reduce reboiler and condenser duties was simulated by reducing the heat duty of coolers in the preceding refrigeration section. This will raise the temperature of the stream entering the NGL Recovery section, replicating a pre-heating scenario. The change in operating conditions resulted in a 68% increase in the exergy efficiency, reducing the exergy destruction by 62% down to 17,693 KW. The operating cost was slightly increased; however, the profits of the plant were significantly increased due to enhanced recovery rates of the product stream. The payback period was reduced by 1.6%, bringing it down to 1.08 years, making a strong economic case for building strategic decisions.

4.3 Limitations of the study

The study aims to enhance the thermo-economic performance of the process of LPG Production by Natural gas treatment and processing. Exergy and Pinch-based Energy analyses are performed to assess the performance of the developed model. Production efficiency is measured on the foundation of performance indicators like recovery rates, purity and production profit.

CHAPTER 5: CONCLUSIONS

Summary of the Study

As the environmental concerns around the globe become graver, regulations and transition strategies are being developed meticulously. Natural gas and its associated Liquids have become an economic and feasible option to act as a transition fuel till the world shifts completely towards green energy. Since energy costs have been on a consistent rise, the development of energy-efficient processes is of significant interest. A thermo techno-economic analysis, in this regard, can be employed to conduct a detailed energy analysis coupled with the economic examination to assess optimum operating conditions for the process ASPEN HYSYS is utilized to develop and validate the simulation of a Natural Gas processing Plant. The economic analysis, enabled by APEA, indicates that the economic performance of the current process is strong, giving a payback period of 1.103 years. The exergy analysis conducted for this case reveals that the process is operating at an efficiency of 32.26%. The exergy destruction and exergy improvement potential for the overall process are 65,159 KW and 44,143 KW. The highest exergy destruction is in the NGL Fractionation train 46,173 KW and the lowest contribution is by Gas Dehydration Unit, which is 918.26 KW.

Further energy analysis is conducted to evaluate energy saving potential utilizing the AEA. 19% potential energy consumption for the total utilities can be reduced by maximizing energy savings. The actual energy consumption is 63,111 KW, while the target calculated by pinch analysis is 50,638 KW. Pre-heating the feed stream entering the NGL Fractionation unit improves the exergy efficiency to 12.39%, reducing the exergy destruction and exergy improvement potential by 62.12% and 64.19%, respectively. This results in a 73% reduction in actual utility consumption for the NGL recovery process. The design modification results in a significant decrease in capital costs by 25%. Although the total operating cost slightly increases by 1.92%, the payback period is reduced to 1.08 years due to a 21% increase generated by product sales. The results demonstrate the crucial nature and impact of refrigeration temperatures on the economy of the process and suggest that introducing meticulously designed and measured changes in operating conditions results in significant energy savings, thermal performance improvement and economic viability of the process.

Future Work

The results and recommendations are generated based on analysis conducted by introducing changes in temperature conditions of the streams only. However, other variables such as pressure, feed composition and mass flow rate can be varied by introducing uncertainty to simulate a real-case scenario. Optimization using mathematical approaches and AI must be studied to reach better thermal performance and investigate further optimum conditions.

REFERENCES

[1] "https://www.man-es.com/discover/decarbonization-glossary---man-energy-solutions/liquefied-petroleum-gas."

[2] S. Kartohardjono and A. Eviany, "The Impact of Gas Production Decline to LPG Recovery and Optimization Strategy to Prevent LPG Facility From Shutdown NOx reduction through hollow fiber membrane modules View project CO2 absorption though superhydrophobic membrane contactor View project The Impact of Gas Production Decline to LPG Recovery and Optimization Strategy to Prevent LPG Facility From Shutdown," *International Journal of Advanced Science and Technology*, vol. 29, no. 7s, pp. 3296–3304, 2020, [Online]. Available: https://www.researchgate.net/publication/342168844

[3] R. K. Abdulrahman and I. M. Sebastine, "Natural gas sweetening process simulation and optimization: A case study of Khurmala field in Iraqi Kurdistan region," *J Nat Gas Sci Eng*, vol. 14, pp. 116–120, Sep. 2013, doi: 10.1016/j.jngse.2013.06.005.

[4] F. Chouaibi, J. Belghaieb, and M. R. Jeday, "Evaluation of the Energy Performances of an LPG Production Unit Using Pinch Analysis and Exergy Balance."

[5] "techno commercial analysis of LPG recovery from natural gas using ASPEN HYSYS".

[6] E. A. Torres and W. L. R. Gallo, "EXERGETIC EVALUATION OF A COGENERATION SYSTEM IN A PETROCHEMICAL COMPLEX."

[7] S. V. de A. Franco, D. da Cunha Ribeiro, and A. P. Meneguelo, "A comprehensive approach to evaluate feed stream composition effect on natural gas processing unit energy consumption," *J Nat Gas Sci Eng*, vol. 83, Nov. 2020, doi: 10.1016/j.jngse.2020.103607.

[8] I. B. Tuychiyev and A. B. Kamolov, "INTERNATIONAL JOURNAL ON ORANGE TECHNOLOGY Modeling and Simulation of the Process of Natural Gas Fractionation," 2021, [Online]. Available: https://journals.researchparks.org/index.php/IJOT

[9] M. Mazumder and Q. Xu, "Modeling and Optimization for a Comprehensive Gas Processing Plant with Sensitivity Analysis and Economic Evaluation," *Chem Eng Technol*, vol. 43, no. 11, pp. 2198–2207, Nov. 2020, doi: 10.1002/ceat.202000216.

[10] H. Hu, H. Jiang, J. Jing, H. Pu, J. Tan, and N. Leng, "Optimization and Exergy Analysis of Natural Gas Liquid Recovery Processes for the Maximization of Plant Profits," *Chem Eng Technol*, vol. 42, no. 1, pp. 182–195, Jan. 2019, doi: 10.1002/ceat.201800238.

[11] T. Saadi, M. R. Jeday, and J. N. Jaubert, "Exergetic analysis of an LPG production plant using HYSYS software," in *Energy Procedia*, Elsevier Ltd, 2019, pp. 1385–1390. doi: 10.1016/j.egypro.2018.11.303.

[12] A. E. R. Sayed, I. Ashour, and M. Gadalla, "Integrated process development for an optimum gas processing plant," *Chemical Engineering Research and Design*, vol. 124, pp. 114–123, 2017, doi: 10.1016/j.cherd.2017.05.031.

[13] S. D. Salas, L. Contreras-Salas, P. Rubio-Dueñas, J. Chebeir, and J. A. Romagnoli, "A multi-objective evolutionary optimization framework for a natural gas liquids recovery unit," *Comput Chem Eng*, vol. 151, Aug. 2021, doi: 10.1016/j.compchemeng.2021.107363.

[14] M. Zaman Shahruddin *et al.*, "Natural gas liquid (NGL) distillation process using driving force and thermal pinch analysis methods: Energy and economic assessment," *Malaysian Journal of Chemical Engineering & Technology Journal Homepage*, vol. 3, no. 2, pp. 18–24, 2020, doi: 10.24191/mjcet.v3i2.XXXX.

[15] B. Ghorbani, G. R. Salehi, H. Ghaemmaleki, M. Amidpour, and M. H. Hamedi, "Simulation and optimization of refrigeration cycle in NGL recovery plants with exergy-pinch analysis," *J Nat Gas Sci Eng*, vol. 7, pp. 35–43, 2012, doi: 10.1016/j.jngse.2012.03.003.