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Escuela de Ciencias Químicas e Ingeniería

Operational Improvements in the Shushufindi Gas Plant as an Alternative to Increasing the Availability of Liquefied Petroleum Gas (LPG) in the Ecuadorian Market

Trabajo de integración curricular presentado como requisito para la obtención del título de Petroquímica

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For my parents Jose Echeverria and Lupe Recalde, for being the people who have accompanied me in my triumphs and failures throughout my life and for teaching me to fight for my dreams despite the circumstances.

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Maritza Ruby Echeverria Recalde

RESUMEN

El gas licuado de petróleo (GLP) es una mezcla de hidrocarburos ligeros, compuesta principalmente de propano y butanos. Se obtiene mediante el procesamiento del gas natural y/o en procesos de refinación de crudo de petróleo. En la actualidad, el GLP tiene aplicaciones muy variadas, en la industria, el transporte, la agricultura, o para cocinar. En Ecuador, el consumo de GLP ha tenido un comportamiento creciente a lo largo de los años y se centra principalmente en los sectores doméstico, industrial, automotriz y agrícola. La producción de GLP se focaliza en las refinerías de Esmeraldas y La Libertad, y en la Planta de Gas Shushufindi (PGSh), que en conjunto producen aproximadamente 1,92 MMbl/año. Sin embargo, para satisfacer el mercado de GLP en el país, el gobierno realiza importaciones que representan aproximadamente el 80% de la demanda nacional. Una alternativa para incrementar la oferta de GLP en el país es valorizar el gas natural asociado a la producción de hidrocarburos en la región Amazónica. La PGSh, se encuentra ubicada en la provincia de Sucumbíos, y es el principal centro de procesamiento de hidrocarburos para la producción de GLP. El presente trabajo tiene como objetivo llevar a cabo un estudio técnico y de factibilidad económica de mejoras operacionales que puedan implementarse en la PGSh con la finalidad de incrementar la oferta de GLP en el mercado nacional. El estudio consistió primero en identificar el proceso industrial de producción de GLP en la PGSh, donde se reconoció las operaciones unitarias existentes, las condiciones de operación de los equipos y las diferentes secciones de la planta en base a información pública y disponible. Segundo, se realizó un análisis detallado de la operación de la planta empleando herramientas de simulación y datos reales. Tercero, se propuso mejoras operacionales que podrían implementarse en la PGSh, considerando modificaciones en condiciones de operación, posible instalación de válvulas de expansión e incorporación de una nueva unidad de enfriamiento. Finalmente, se realizó un análisis técnico y de factibilidad económica de las mejoras operacionales propuestas. Los resultados mostraron que es posible incrementar la producción actual de GLP de la PGSh en un 30% mediante la incorporación de una nueva unidad de enfriamiento con etano. El costo estimado de la unidad de enfriamiento es de 1.5 MMUSD. Sin embargo, se requiere de un estudio económico detallado para definir el costo final asociado a la implementación y puesta en marcha de la nueva unidad de enfriamiento.

Palabras Claves: gas licuado de petróleo, Planta de Gas Shushufindi, mejoras operacionales, oferta, demanda, simulación de procesos.

ABSTRACT

Liquefied petroleum gas (LPG) is a mixture of light hydrocarbons, composed mainly of propane and butanes. It is obtained through the processing of natural gas and/or in crude oil refining processes. Currently, LPG has very varied applications, in industry, transport, agriculture, or for cooking. In Ecuador, the LPG consumption has had a growing behavior over the years and is mainly focused on the domestic, industrial, automotive, and agricultural sectors. LPG production focuses on the Esmeraldas and La Libertad refineries, and the Shushufindi Gas Plant (ShGP), which all together produce approximately 1.92 MMbl/yr. However, to satisfy the country's LPG market, the government carries out imports that represent approximately 80% of the national demand. An alternative to increase the LPG supply in Ecuador is to value the natural gas associated with the production of hydrocarbons in the Amazon region. The ShGP is located in the Sucumbios province and is a main hydrocarbon processing center for the LPG production. This work aims to carry out a technical and economic feasibility study of operational improvements that can be implemented in the ShGP to increase the supply of LPG in the national market. The study consisted of first, identifying the industrial process for the LPG production in the ShGP, where the existing unit operations, the equipment operating conditions, and the different sections of the plant were recognized based on public and available information. Second, a detailed analysis of the plant's operation was carried out using simulation tools and real data. Third, operational improvements were proposed that could be implemented in the ShGP, considering modifications in operating conditions, possible installation of expansion valves, and incorporation of a new cooling unit. Finally, a technical and economic feasibility analysis of the proposed operational improvements was carried out. The results showed that it is possible to increase the current LPG production of ShGP by 30% by incorporating a new cooling unit using ethane-based refrigerants. The estimated cost of the cooling unit is 1.5 MMUSD. However, a detailed economic study is required to define the final cost associated with the implementation and start-up of the new cooling unit.

Keywords**:** liquefied petroleum gas, Shushufindi gas plant, operational improvements, supply, demand, process simulation.

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

ABBREVIATIONS

AC	Actual case
ASTM	American Society for Testing and Materials
A	Heat exchange area (ft^2)
BC	Base Case
C_1	Methane
C_2	Ethane
C_3	Propane
C_4's	Butanes (n-butane and i-butane)
C_5^+	Natural gasoline
CO_2	Carbon dioxide
C_{fob}	Adjusted base cost
C_b	Unit base cost in 1968
C_{md}	Adjusted module cost
Cp	Specific heat
Fd	Module that represents the type of exchanger
Fp	Module that represents the working pressure
Fm	Module that represents the equipment material
gpm	Gallons per minute
GPM	Liquid hydrocarbon content expressed in gallons that can be obtained for every 1000 cubic feet of natural gas at standard conditions
H_2S	Hydrogen sulfide
INEN	Servicio Ecuatoriano de Normalización
LPG	Liquefied Petroleum Gas
LPG Cyl-15Kg/d	15 Kg LPG Cylinders per day
LMTD	Mean logarithmic difference of temperature (°F)
MMBl/yr	Million barrels per year
MMUSD	Million dollars per year
MMscfd/d	Millions of standard cubic feet per day
$\dot{m}$	Mass flow (lb/h)

NG	Natural Gas
NGL	Natural Gas Liquids
N_2	Nitrogen
NTE	Norma Técnica Ecuatoriana
ppmv	Parts per million by volume
Q	Power exchanged by the fluids (Btu/h)
ShGP	Shushufindi Gas Plant
Tm/d	Metric tons per day
ΔT	Inlet temperature minus outlet temperature of the stream (ºF)
Te_1	Inlet temperature of cold side (ºF)
Ts_1	Outlet temperature of cold side (ºF)
Te_2	Inlet temperature of hot side (ºF)
Ts_2	Outlet temperature of hot side (ºF)
U	Global coefficient of thermal transfer (Btu/h*ft^2*ºF)

CHAPTER I

1. INTRODUCTION

Natural gas (NG) supplies comprise approximately 23% of the world's energy sources, becoming the third source of energy, after oil and coal [1]. Furthermore, NG is considered one of the safest, cleanest, and most efficient energies of all energy sources [2]. Products such as liquefied petroleum gas (LPG) are produced at natural gas processing facilities, which currently has a high impact on the domestic, industrial, automotive, and agricultural sectors [3].

Liquefied petroleum gas is a mixture of light hydrocarbons, mainly composed of propane (C_3) and butanes (C_4's) [4]. Besides, LPG is obtained through the processing of associated natural gas and/or in crude oil refining processes [5]. Nowadays, LPG has some applications; it is used as fuel in industrial processes, for cooking food, in heating appliances, and vehicles [6]. Furthermore, the growing demand for this fuel worldwide, due to the multiple uses it has, makes the oil and gas industries focus on finding more effective production and processing methods [7].

Preliminary studies have shown that it is possible to increase LPG production and profitability through operational modifications of existing industrial facilities [8]. Currently, the oil and gas industries use process simulation tools to achieve higher production yields, reduce environmental impacts, and increase economic profitability [9]. The simulation techniques allow analyzing, contrasting, and finding improvements or operational changes that can be implemented in existing industrial facilities [10].

In Ecuador, LPG production is obtained through the processing of associated natural gas and in refining processes. This production focuses on the Esmeraldas Refinery, the Libertad Refinery, and the Shushufindi Gas Plant (ShGP) [11]. In 2019, the national production was 1.92 million barrels (1.92 MMbl), which represents 13.69% of the national demand. To satisfy the LPG market in the country, the government carries out imports that represent 86.31% of national demand and an investment of 385.32 million dollars per year (385.32 MMUSD/yr) [12].

Therefore, the objective of this work is to carry out a technical and economic feasibility study of operational improvements that can be implemented in the ShGP to increase the availability of LPG in the Ecuadorian market. In this study, public and available information from the ShGP and a commercial simulation software (PRO/II) were used to assess the impact of the proposed operational improvements.

This study focuses on the Shushufindi Gas Plant since it is a main hydrocarbon processing center for LPG production in Ecuador. The Shushufindi Gas Plant is part of the Shushufindi industrial complex and it is located in the Amazon region, in Sucumbios province [13]. It was designed to process 25 million standard cubic feet per day (25 MMscfd) of associated natural gas and 150 gallons per minute (150 gpm) of condensates to obtain approximately 500 metric tons per day (500 Tm/d) of LPG [14]. Currently, the plant processes 14.14 MMscfd of natural gas and 78.83 gpm of condensates to produce approximately 249.73 Tm/d of LPG [15].

The methodology proposed in this study consisted first of identifying the industrial process of LPG production in the ShGP, where the existing unit operations, the equipment operating conditions, and the different sections of the plant were recognized. Second, a detailed analysis of the plant operation was carried out using process simulation tools and real data. Third, operational improvements applicable to the ShGP were proposed, considering modifications in operating conditions, possible installation of expansion valves, and incorporation of a new cooling unit. Finally, a technical and economic feasibility analysis of the proposed operational improvements was carried out.

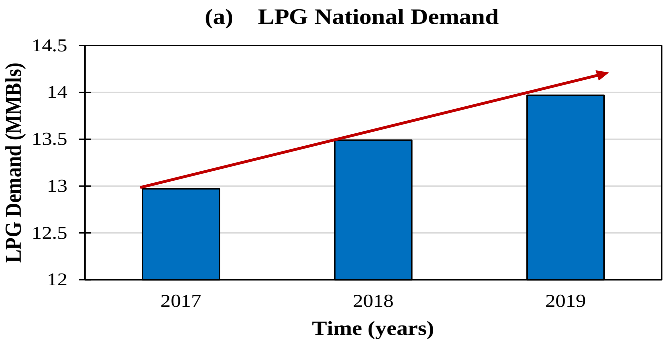
1.1. Problem Approach

In recent years, LPG consumption in Ecuador has shown an increasing behavior. According to statistical data from the EP Petroecuador company, the national demand for LPG in the last three years has increased by approximately 0.5 MMbl [16]. LPG consumption in the country is mainly focused on the domestic, industrial, agricultural, and automotive sectors, where LPG for domestic use is the one with the highest demand nationwide. In the last year, the national demand was 13.96 MMbl of LPG, where 88.64% was for domestic use [12,17].

Currently, the sale price of one Kg of LPG for domestic use is 0.11 USD [18]; however, this value includes a subsidy of approximately 70% of the import price. This subsidy is paid by the government, thus generating expenses for the state. Furthermore, the subsidy does not apply to the industrial sector since it is considered that natural or legal persons who wish to obtain LPG in large quantities may be able to assume a higher price [19].

LPG production in Ecuador focuses on the Esmeraldas Refinery, the Libertad Refinery, and the Shushufindi Gas Plant [11]. Last year, national production reached 1.92 MMbl of LPG, representing 13.69% of national consumption. To satisfy the country's LPG market, the government has to import around 12.12 MMbl of LPG each year, which represents 86.31% of the national demand and a cost of 385.32 MMUSD/yr [12].

Taking into account the growing demand, the low production and the high percentage imports of LPG in the Ecuadorian market (Figure 1), this study focuses on searching operational improvements that can be implemented at the Shushufindi Gas Plant to increase the availability of this fuel in the national market. This study focuses on the ShGP since it is a main hydrocarbon processing center for LPG production.



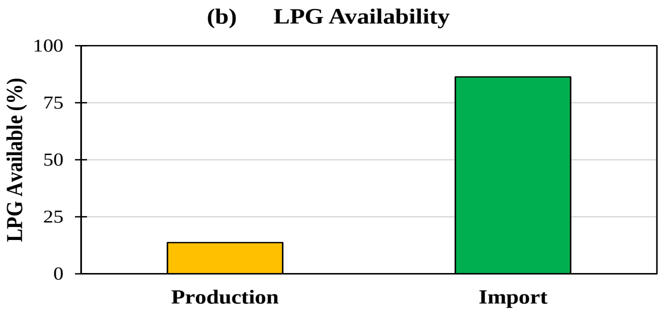


Figure 1. *LPG in the Ecuadorian market: (a) National demand, (b) Availability*

1.2. Objectives

1.2.1. General Objective

To carry out a technical and economic feasibility study of operational improvements applicable to the Shushufindi Gas Plant to increase LPG availability in the national market.

1.2.2. Specific Objectives

- To identify the existing unit operations, the equipment operating conditions, and the different sections that comprise the ShGP, based on public and available information.

- To carry out an analysis of the ShGP operation, employing simulation tools and using real data (flow rates and composition of streams, operating conditions, and equipment specifications).

- To propose applicable operational improvements in the ShGP considering modification in operating conditions and/or new process unit incorporations to increase LPG availability in the national market.

- To perform a technical and economic feasibility analysis of the proposed operational improvements.

CHAPTER II

2. BACKGROUND AND LITERATURE REVIEW

2.1. Natural Gas

Natural gas is a mixture of low molecular weight gaseous hydrocarbons mainly made up of methane (C_1), and of significant proportions of ethane (C_2), propane, heavy hydrocarbons (C_4^+), and some nonhydrocarbons gases, such as nitrogen (N_2), carbon dioxide (CO_2) and hydrogen sulfide (H_2S) [20]. In addition, NG constitutes the third source of energy, after oil and coal and it is considered one of the safest, cleanest, and most efficient energies of all energy sources [2].

Natural gas is found in nature in deep deposits of porous rocks, either associated with crude oil (associated gas) or in deposits containing little or no crude oil (non-associated gas) [21]. Natural gas is classified according to its liquid content (as either lean or rich gases), and according to H_2S and CO_2 content (as either sweet or sour gases) [22]. To quantify the liquid content present in a natural gas mixture, the industry uses GPM, or gallons of liquids recoverable per 1000 standard cubic feet (Mscf) of gas [23]. Therefore, a lean gas contains $GPM < 1$, whereas a rich gas contains $GPM \geq 3$. A sweet gas contains negligible amounts of CO_2 and H_2S, whereas a sour gas has unacceptable quantities of H_2S and CO_2, which when interacting with water cause corrosion in the processing facilities [22].

2.1.1. Natural Gas Processing

Natural gas from the well contains hydrocarbons, CO_2, H_2S, and water along with many other impurities. Therefore, crude NG must be processed to: remove materials that inhibit the use of gas as an industrial or residential fuel, separate components that have a higher value such as petrochemical feedstocks, stand-alone fuels (e.g., propane), or industrial gases (e.g., ethane), and fractioning to increase the energy density of the gas for storage or transportation [24]. Part of the processing can be accomplished at or near the wellhead (field processing). The complete processing of natural gas is carried out in a processing plant, usually located in a natural gas production region. The NG processing philosophy depends on the composition and conditions of the feed gas, and the desired product streams (product specifications) [22]. In Figure 2, a simplified natural gas processing scheme is shown.

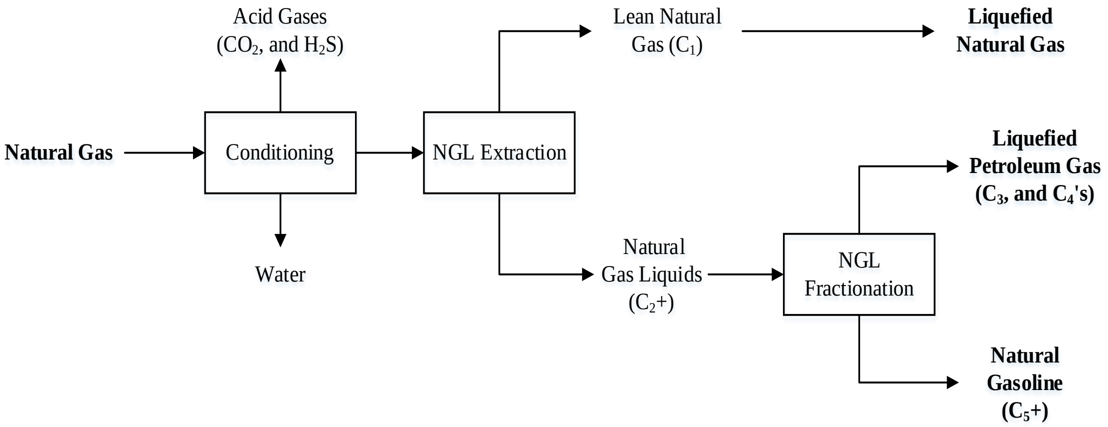


Figure 2. *Natural gas processing steps*

- **Conditioning**

Natural gas conditioning primarily involves the removal of water and acid gases such as CO_2 and H_2S. The acid gases removal comprises the reduction of CO_2 and H_2S, along with other sulfur species, to levels low enough to meet contractual specifications or allow processing without corrosion and clogging problems. A quality gas is defined as one that contains $CO_2 < 2$ vol.%, $N_2 < 4$ vol.%, or $H_2S < 4$ ppmv [22]. To remove these impurities, most plants use water-based absorbents; however, other solvents and processes are used, such as amine absorption, molecular sieve adsorption, membranes, etc.

The CO_2, as a by-product, if quantities are large, can be used as an injection fluid in EOR (Enhanced Oil Recovery) projects. If CO_2 amounts are low, it is vented as long as it complies with impurities' environmental regulations. In the case of H_2S, it can be incinerated and ventilated as long as environmental regulations are observed. Also, H_2S can be converted to elemental sulfur by the Claus process or a similar process [22].

Liquid and gas streams can be saturated with water after amine treatment or from underground storage. Therefore, dehydration or water removal is important to reduce corrosion of pipelines and equipment, prevent hydrate formation, and meet product specifications. The most common processes for dehydrating natural gas are physical absorption and adsorption. Water levels in natural gas can be reduced to the 10 ppmv range in a physical absorption process in which the gas comes into contact with a liquid that absorbs water vapor [22]. The most commonly used absorbents are glycols, ethylene glycol (EG), diethylene glycol (DEG), triethylene glycol (TEG), tetraethylene glycol (TREG) and propylene glycol. NG can also be dehydrated through the physical adsorption process, where solids are used as the adsorbent medium. The solids commonly used in this physical adsorption process are synthetic zeolites or molecular sieves that have an extremely high surface-to-volume ratio [22].

- **Natural gas liquids extraction**

Recovery of natural gas liquids (NGL) generally involves cold separation processes to recover ethane and heavier hydrocarbons. In other words, the condensation of the less volatile fraction is obtained by a decrease in temperature at low temperatures [22]. At present, there are several methods and process configurations to recover NGL. The best configuration depends on the desired products, gas volumes, gas composition, inlet and outlet pressure, and associated costs. The temperature drop can commonly be achieved by three main methods: Joule-Thompson expansion, turbo expansion or cryogenic expansion, and external or mechanical cooling [25,26].

Joule – Thomson expansion: gas cooling can be achieved by expanding the gas from high to low pressure; this is achieved through the throttle or expansion valve. The expansion is given in an isenthalpic process since when the pressure drops, the temperature also decreases, and the heavy hydrocarbons found in the gas condense. Furthermore, the Joule-Thomson expansion (J-T expansion) is one of the most widely used and practical methods for extracting NGL as it is an easy process to operate and has low capital and maintenance costs [26].

Turbo expander: this process was created for the high recovery of liquids (heavy components of hydrocarbons) through refrigeration. This method is based on a machine that has two functions; it expands from a radial turbine and compresses with the centrifugal compressor. The expansion turbine causes the gas to expand isotropically eliminating the enthalpy of the gas stream, which causes cooling; the extracted energy is also used to roll the shaft and operate the compressor, which recompresses the waste gas stream. The turbo expander is a technology where up to 90% of C_3 and 100% of C_4^+ can be recovered. However, it has a high capital investment cost [25].

Mechanical refrigeration: it is the simplest and most direct process for NGL recovery at large-scale. This method is provided by a refrigeration cycle, which normally uses propane as the refrigerant [25]. The refrigeration cycle consists of four stages (Figure 3) [22]:

1. Vapor compression of saturated refrigerant where centrifugal or reciprocating compressors are used to move refrigerants from the lowest to the highest range operating pressure conditions.
2. Condensation of the refrigerant by heat exchange with a refrigerant fluid, generally air.
3. Expansion through a valve (J-T expansion) to cool and condense the refrigerant.
4. Heat exchange with the fluid (hydrocarbons) to be cooled by evaporation of the refrigerant back to point 1.

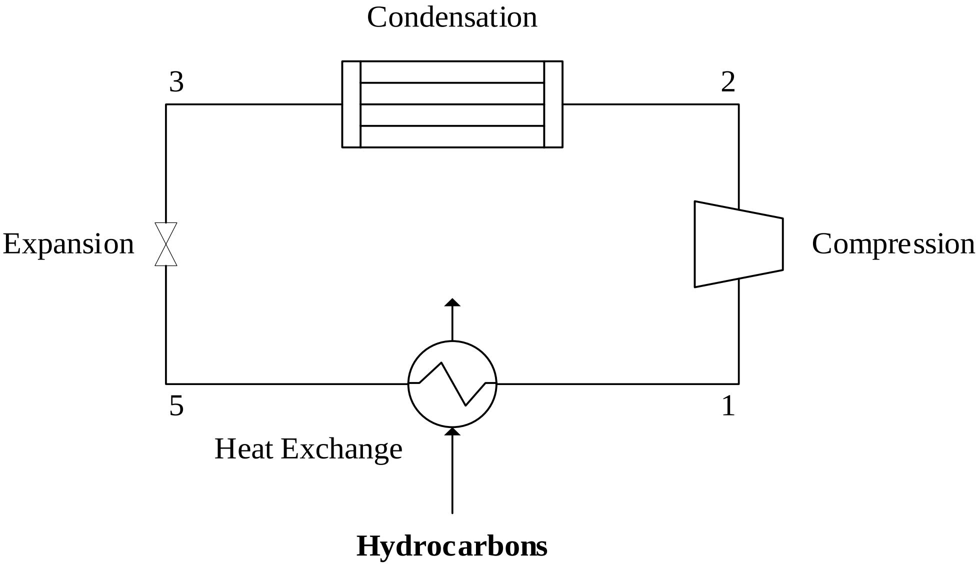


Figure 3. *Refrigeration cycle stages*

- **Natural gas liquids fractionation**

Natural gas liquids are fractionated through a sequence of distillation columns. The fractionation train depends on the current economy and customer requirements. The NGL fractionation is done based on the different boiling points of each hydrocarbon [27]. Figure 4 shows a possible NGL fractionation configuration, where propane is extracted using a depropanizer tower, butanes using a debutanizer tower, and finally, a butane separator to separate isobutane (i-C_4) from normal butane (n-C_4) [10].

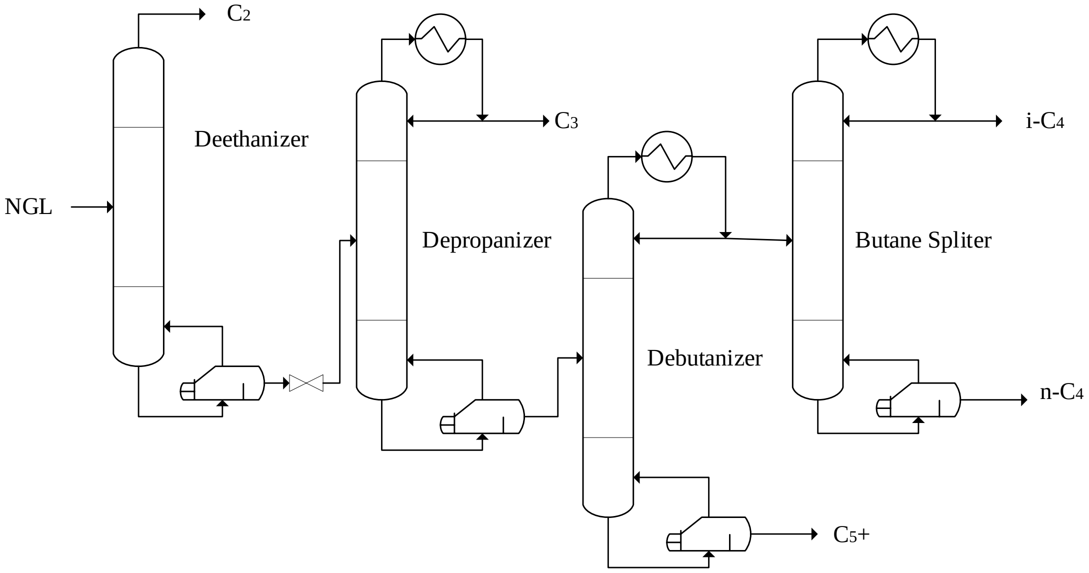


Figure 4. *NGL fractionation configuration*

2.2. Liquefied Petroleum Gas

Liquefied Petroleum Gas is a mixture of light hydrocarbons composed mainly of propane and butanes and in a smaller proportion of ethane, pentane, and others [4]. At normal pressure and temperature conditions (14.69 psi and 68 ºF), LPG is in a gaseous state. However, under moderate pressure conditions and at room temperature, LPG can be stored and handled in the liquid phase [28]. According to its volatility LPG is classified into three types: commercial propane (high volatility), commercial butane (low volatility), and commercial propane-butane mixtures (intermediate volatility). LPG is called commercial propane when the hydrocarbon mixture has a minimum of 80 vol.% propane. Commercial butane is the hydrocarbon mixture consisting essentially of butanes, and which can have a maximum of 20 vol.% propane [29]. For GLP commercialization, an odorizing agent (ethyl mercaptan) is added to give it a characteristic odor and to detect its presence. Also, Table A.1 (Appendix A) presents the specifications to be met by LPG, according to the Gas Processors Association (GPA) [30].

Worldwide, liquefied petroleum gas originates in two ways: 60% of LPG production comes from natural gas processing, and the remaining 40% is the product of the crude oil refining process [5]. In an oil refinery, LPG is produced in several stages: atmospheric distillation, reforming, cracking, and others. The LPG obtained from natural gas processing is based on the fractionation, where LPG (C_3 and C_4's) is separated from natural gasoline (C_5^+).

LPG due to its characteristics and calorific value (11739 Kcal/Kg) [31] is very useful for different uses in the daily life of people, shops and industries. In the domestic or residential sector, the fuel is used for cooking food, heating water and heating. In the industrial sector, it is used as fuel or raw material. The agricultural sector uses this fuel to promote grain conservation, as well as for heating in greenhouses. The automotive sector uses LPG as fuel [6].

2.2.1. LPG Recovery Technology

Various technologies have been developed to extract LPG from NG. The LPG recovery processes can be classified mainly into two groups: conventional and advanced processes.

In the conventional process, fractionation columns are used to separate the different hydrocarbons based on the boiling points of each compound [27]. Figure 4 shows a conventional fractionation process for the LPG recovery [7]. In the first stage, the extraction of LPG from NG is carried out in a deethanizer column. In this column, C_1 and C_2 will separate at the top of the column as vapor phase, and heavier hydrocarbons (C_3^+) flow through the bottom in the liquid phase and then enter the debutanizer column. In this column, C_3 and C_4 (LPG) are separated at the top of the column, while natural gasoline (C_5^+) flows through the bottom. To obtain pure propane, the C_3 and C_4 mixture is separated in a depropanizer column. Finally, the n-butane and isobutane are separated in a butane separator (butane splitter).

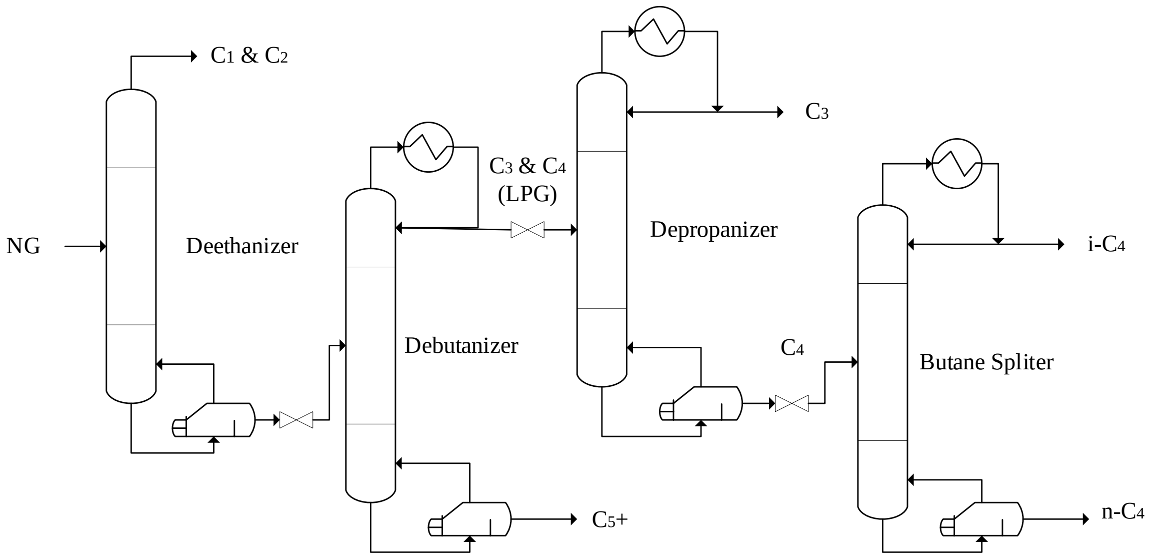


Figure 5. *Conventional fractionation process for the LPG recovery*

Advanced processes are based on more sophisticated process configurations involving different NGL recovery methods. The efficiency of these processes depends on the configuration of the industrial process [22]. Advanced NGL or LPG recovery technology includes gas subcooling process (GSP), overhead recycling process (ORH), single-column top recycling process (SCORE), and a few others [32]. One of these processes is described below.

Gas sub-cooled process (GSP): this gas sub-cooled process is shown in Figure 6. In this process, a portion of the feed gas that is after vaporization is condensed and sub-cooled flashed down to the tower operating pressure, and supplied to the tower as its top feed. The remainder of the feed gas is also expanded to lower pressure by using turbo expander for vapor streams and thereafter fed to the tower at one or more intermediate feed points. The cold liquids supplied to the top of the tower act as reflux, contacting and rectifying the vapor leaving the expander by absorbing the ethane-plus components for recovery in the bottom product [7].

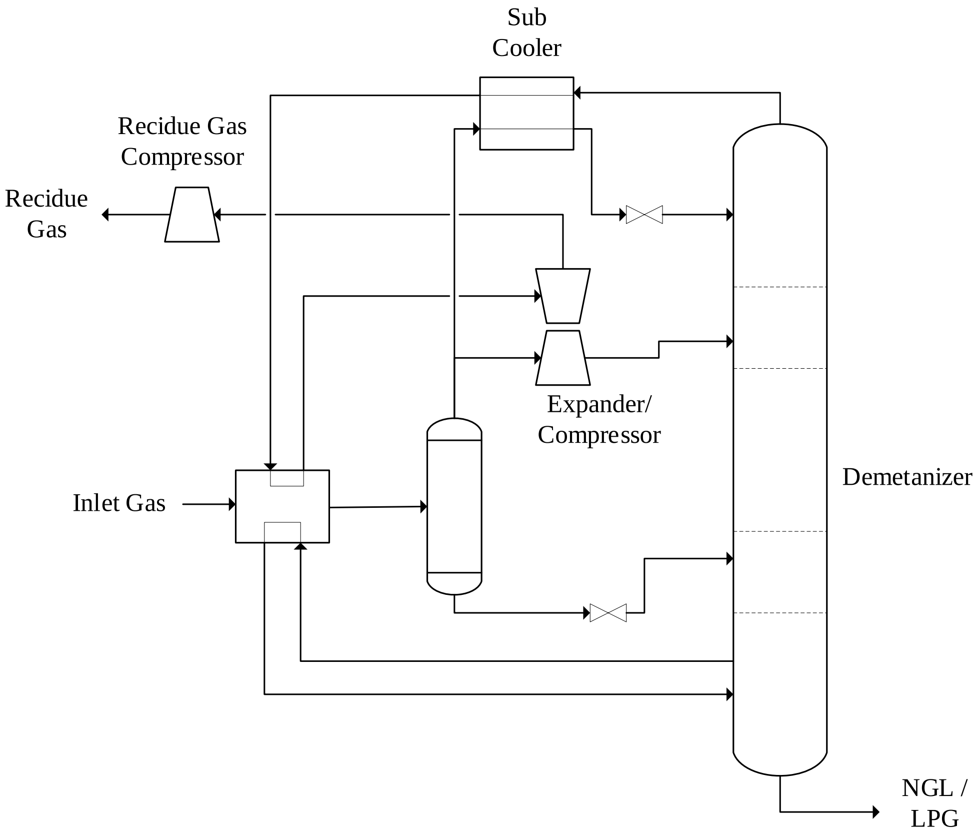


Figure 6. *Gas sub-cooled process (GSP)*

2.3. Liquefied Petroleum Gas in Ecuador

Ecuadorian LPG is a mixture of propane and butanes and in a smaller proportion of ethane, pentane, and others. Its composition varies depending on its source of production; LPG from crude oil refining and LPG from natural gas processing [13]. In addition, it is important to point out that the Ecuadorian LPG composition is not available for public access. However, according to Ecuadorian regulations, LPG can contain a minimum of 60% of C_3 and a maximum of 40% of C_4 (Table 1) [33,35]. Also, Ecuadorian LPG must comply with the requirements established by the Ecuadorian Standardization Service (INEN) (Table A.2) [35].

Table 1. *Quality of the Ecuadorian LPG*

Component	**Content in LPG (vol.%)**	
	min	**max**
Propane (C_3)	60	100
Butanes (C_4's)	0	40
Pentane and heavies (C_5^+)	-	2

2.3.1. Demand

LPG demand in Ecuador is mainly focused on the domestic, industrial, agricultural, and automotive sectors [16]. In 2019, EP Petroecuador shipped 13.96 MMbl of Liquefied Petroleum Gas, to meet national demand [12]. LPG for domestic use was the one with the highest demand since a dispatch of 12.38 MMbl was counted, representing 88.64% of national consumption. In contrast, LPG for industrial use reached 1.21 MMbl of LPG, with a demand of 8.7%. LPG for the agricultural sector reached a dispatch of 0.30 MMbl, representing 2.15% of consumption throughout the country. To a lesser extent, LPG for vehicular use obtained commercialization of 0.07 MMbl, representing 0.54% [17]. Figure 7 shows the LPG demand in each sector. Besides, according to their nominal capacity, 5, 10, 15 and 45 Kg cylinders are sold in Ecuador [36]. 15 Kg cylinders are intended for domestic use and 45 Kg cylinders for industrial use.

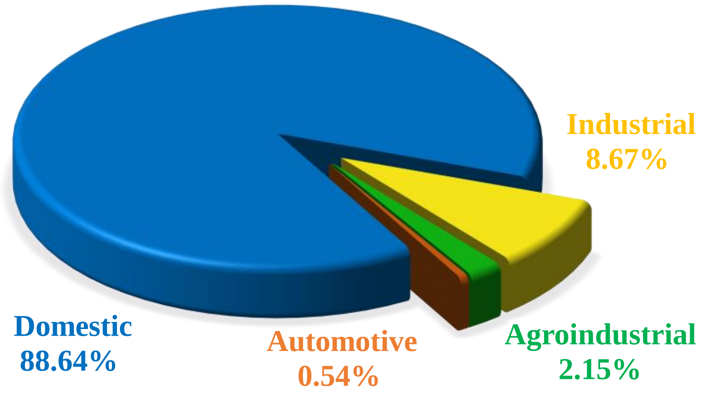


Figure 7. *LPG demand in Ecuador by sector*

2.3.2. Production

LPG production in Ecuador is obtained by processing the associated natural gas from the oil production fields and in refining processes. This production focuses on the Esmeraldas Refinery, the Libertad Refinery, and the Shushufindi Gas Plant [11]. For the previous year, national production reached 1.92 MMbl, representing 13.69% of national consumption (Figure 8). The Esmeraldas refinery process 0.84 MMbl/yr, the Libertad Refinery 0.02 MMbl/yr, and the Shushufindi Gas Plant 1.06 MMbl/yr [12].

2.3.3. Import

To meet national demand, the Ecuadorian state imports more than 80% of the LPG consumed in this country (Figure 8). In 2019, Ecuador imported 12.12 MMbl corresponding to a cost of 385.331 MMUSD [12].

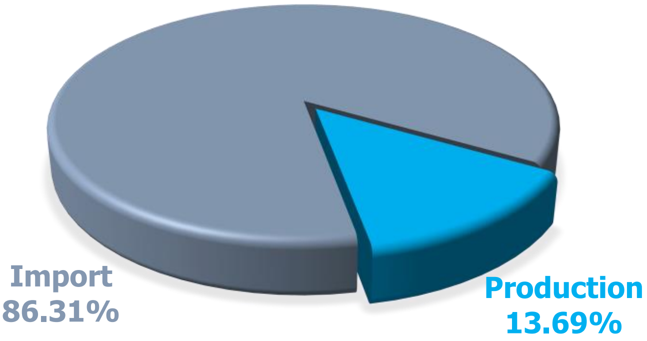


Figure 8. *LPG availability - production and import*

CHAPTER III

3. SHUSHUFINDI GAS PLANT

The Shushufindi Gas Plant is part of the Shushufindi Industrial Complex. It is located in the Amazon region, in the Sucumbios province, in the Shushufindi canton (Figure 9). Besides, ShGP is the main hydrocarbon processing center for LPG production in Ecuador [13].

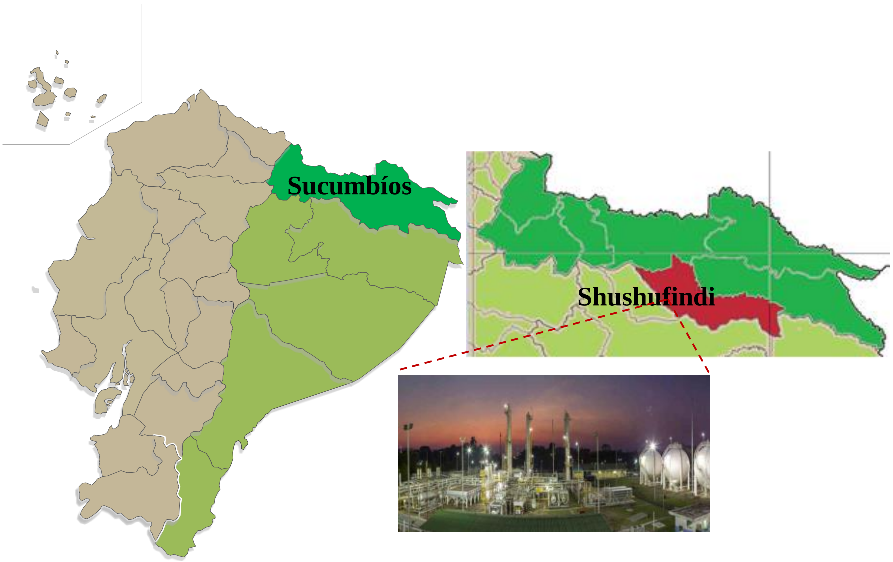


Figure 9. *Shushufindi Gas Plant geographical location*

The main objective of the ShGP is to produce liquefied petroleum gas, and natural gasoline and residual gas as by-products. Natural gasoline is sent to the Shushufindi Refinery for the production of gasoline. The residual gas is used for internal consumption in furnaces, turbines, boilers, and generators at the Shushufindi Industrial Complex. Also, the residual gas excess is sent to EP Petroamazonas to the Central, Sur, and Limoncocha stations [37].

The ShGP started its operations in 1984; it began processing 180 Tm/d of LPG with the feed of 14 MMscfd of natural gas from the Shushufindi and Aguarico oil fields [14,37]. Also, the plant underwent two extensions.

- The first stage included the assembly of high-power compressors at the Secoya station and the construction of pipelines, over an area of 42 Km, to collect and transport the gas and condensates produced in Secoya field and send them to the ShGP for processing. This stage came into operation in July 1990, and the LPG production reached 220 Tm/d [14,37].

- The second stage began operations in March 1992 and contemplated the expansion of the gas plant to process 500 Tm/d of LPG, using natural gas from the Libertador, Secoya, and Shushufindi oil fields as raw material [14,37].

3.1. ShGP Feed Sources

The ShGP has the capacity to process 25 MMscfd of associated natural gas and 150 gpm of condensates, at an inlet pressure of 565 psia and 120 ºF, to obtain a production of approximately 500 Tm/d of LPG [13,14,37]. The inlet gas is received from the Central, South, and North stations, and the inlet condensates are received from the Central, South, North and Secoya stations [14,37]. At the present, the ShGP processes 14.14 MMscfd of natural gas and 78.83 gpm of condensates to produce 249.73 Tm/d of LPG [15]. Table 2 shows the feed streams data.

Table 2. *ShGP feed stream data*

Component	**Formula**	**Natural Gas (mol %)**	**Condensates (mol %)**	**Refrigerant-Propane (mol %)**
Nitrogen	N_2	3.74	17.42	0.4
Carbon Dioxide	CO_2	12.60	1.95	-
Methane	C_1	45.04	2.52	-
Ethane	C_2	7.39	3.19	0.76
Propane	C_3	22.53	40.66	96.44
i-Butane	IC_4	2.68	8.44	1.29
n-Butane	NC_4	5.50	19.67	0.96
i-Pentane	IC_5	-	6.16	-
n-Pentane	NC_5	-	-	-
Hexane plus	C_6^+	0.53	-	0.15
Flow rate		25 MMscfd	150 gpm	Varies in each unit
Temperature		120 ºF	120 ºF	
Pressure		565 psia	565 psia	

Ref.: [13,14,37,38]

3.2.ShGP Block Diagram

The natural gas processing facilities in the ShGP can be divided as follows (Figure 10): inlet and conditioning section, cooling section, and fractionation section [13,14,37,39].

In the reception and conditioning section, the most water amount contained in the feed is condensed for later disposal in the inlet separator. Besides, gas and liquid dehydrators absorb water vapor from process streams. In the cooling section, it is operated at low temperatures, which gives way to the condensation of hydrocarbons (NGL). Finally, in the fractionation section, the LPG is obtained. The unit operations that make up each section are described in 3.3 and 3.4 headings.

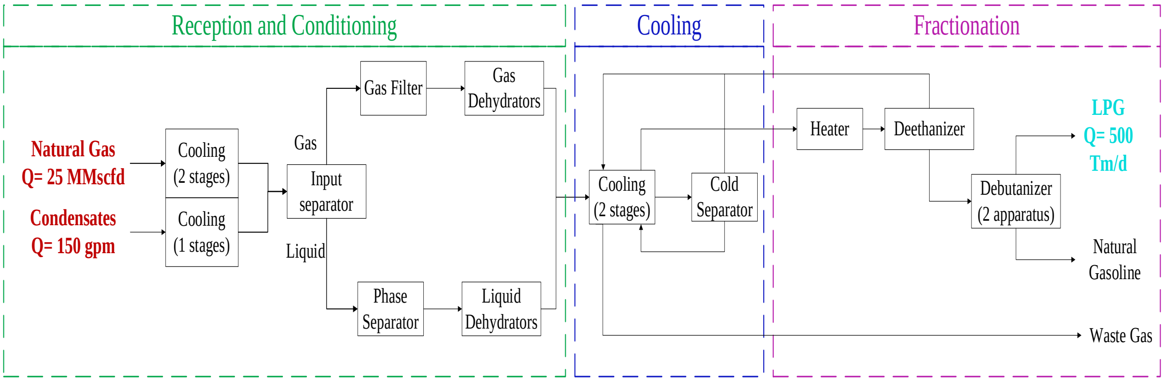


Figure 10. *Shushufindi Gas Plant: block diagram*

3.3. Detailed Process Description

The process of LPG production in the ShGP is described below, and it can be better understood using the process flow diagram (PFD) of the ShGP from Figure 11 [13,14].

3.3.1. Reception and Conditioning Section

The plant was designed to process 25 MMscfd of associated natural gas and 150 gpm of condensates, at an inlet pressure of 565 psia and 120 ºF. The inlet gas is received from the Central, South, and North stations, and the inlet condensates are received from the Central, South, North and Secoya stations [14,37]. The inlet natural gas is initially cooled from 120 ºF to 100.4 ºF in the GE-00.01 exchanger, where water is used as the cooling medium. The water enters at a temperature of 89.6 °F and returns at 100.4 °F. Subsequently, the gas stream complements its cooling from 100.4 ºF to 80 ºF in GE-00.02 exchanger. In this unit, the gas charge is cooled with propane. The inlet condensates are cooled in the GE-00.03 A/B two-shell exchanger using waste gas from the GE-15.01 exchanger. The condensates charge cools from 120 ºF to 80 ºF, while the waste gas heats from 59 ºF to 105 ºF.

Gas and condensates enter the inlet separator GV-16.01, where gas, liquid hydrocarbons, and condensed water are separated. The water is separated from the liquid hydrocarbons, and it is sent to a burning pit, where the small hydrocarbon traces are burned, and the water is treated. The liquid hydrocarbons are driven by the condensate pumps GP-17.01/02/13/14 through the phase separators GV-16.22/23. In phase separators, the water entrained after the separation is removed. Gas circulates through a separator filter GL-21.01, where the water is separated from the gas. Then, gas and condensates enter into dehydrators, where the greatest amount of moisture present in these streams is removed.

There are two dehydrating units for gas GV-16.02/03 and four for liquids GV-16.05/06/26/27. While one unit is dehydrating, the other is regenerating (lead/lag configuration). These dehydrators use the same principle of dehydration bases on molecular sieves formed by synthetic zeolite, which adsorbs water through a physical process. Regeneration of the zeolite is carried out using residual gas and preheated to 500 ºF for 6 hours. At the dehydrators' outlet, the gas and condensates pass through the dust filters GL-21.02/03/04/05/06/07, the purpose of which is to avoid any drag of dust from the sieves.

3.3.2. Cooling Section

Gas and condensates after dehydration go to the cooling train; 32% of condensates combine with 32% of gas to pass through the gas-gas exchanger GE-15.01, and 68% of condensates combine with 68% of gas pass through the gas-liquid exchanger GE-15.02. The streams' temperature drops to -22 ºF; this abrupt temperature change allows the liquefaction of the C_3, C_4's, and C_5^+ fractions. Then, the streams are combined to pass through the propane-cooled exchanger GE-15.03. In this exchanger, the temperature of the hydrocarbons stream drops from -22 ºF to -40 ºF. Next, the stream circulates through the cold separator GV-16.07, where the gas phase is separated from the liquid phase.

The cold separator function is to separate the uncondensed residual gas from the condensed liquids. The residual gas, which flows through the upper part of the separator, combines with the gas stream from the reflux accumulator of the deethanizer. The combined gas stream cools the process stream of the exchanger GE-15.01. Also, it cools the condensates charge entering the exchanger GE-00.03A/B. The residual gas temperature goes from -32 ºF to 59 ºF in the GE-15.01 and from 59 ºF to 105 ºF in the GE-00.03A/B. Liquid from the cold separator is sent through the GE-15.02 exchanger as a cooling agent, where the liquid temperature increases from -40ºF to 60.26 ºF.

3.3.3. Fractionation Section

The hydrocarbons from the exchanger GE-15.02 passes to the exchanger GE-15.11 for its final preheating and subsequent feeding to the deethanizer GV-16.09. The GE-15.11 preheats the charge from 60.26 ºF to 90 ºF with hot propane. The hydrocarbon stream from exchanger GE-15.11 is fed to the deethanizer column through plate 19 according to the design bases. The function of the deethanizer is to separate methane and ethane from propane and the heavier components contained in the column feed. Some propane comes out of the head of the column, but this amount should be minimal. The gases at the top of the deethanizer are partially condensed by propane cooling in the condensers GE-15.04A/B/C and sent to the reflux accumulator GV-16.10. The gases from the reflux accumulator go to the residual gas system, and the liquid is pumped back to the deethanizer. The reflux to the deethanizer is 159.4 gpm. The liquid from the bottom of the column is extracted, sent to the reboilers GE-15.05/A, and then passed to the debutanizers GV-16.24/11.

According to the design bases, half the liquid charge from the reboilers of the deethanizer flow through each debutanizer. Both columns have the same working principle, and the feed flow enters through plate 19 in each column. The function of the debutanizer is to separate propane and butane from the heavier components contained in the column feed. The gases at the top of the debutanizer are fully condensed in a cooling system with air in GA-19.07/03, and then with water in GE-15.09/10 condensers. Then, the condensate enters into reflux accumulator GV-16.21/12. A part of the accumulator liquids is returned to the debutanizer as reflux (137.6 gpm). The rest of the condensate (LPG) is cooled in the exchanger GE-19.09/08, and stored in the LPG spheres. The debutanizer tails are sent to the reboiler GE-15.08/06, then pass through a cooler with air GA-19.05/04 and, finally, enter into natural gasoline storage tank.

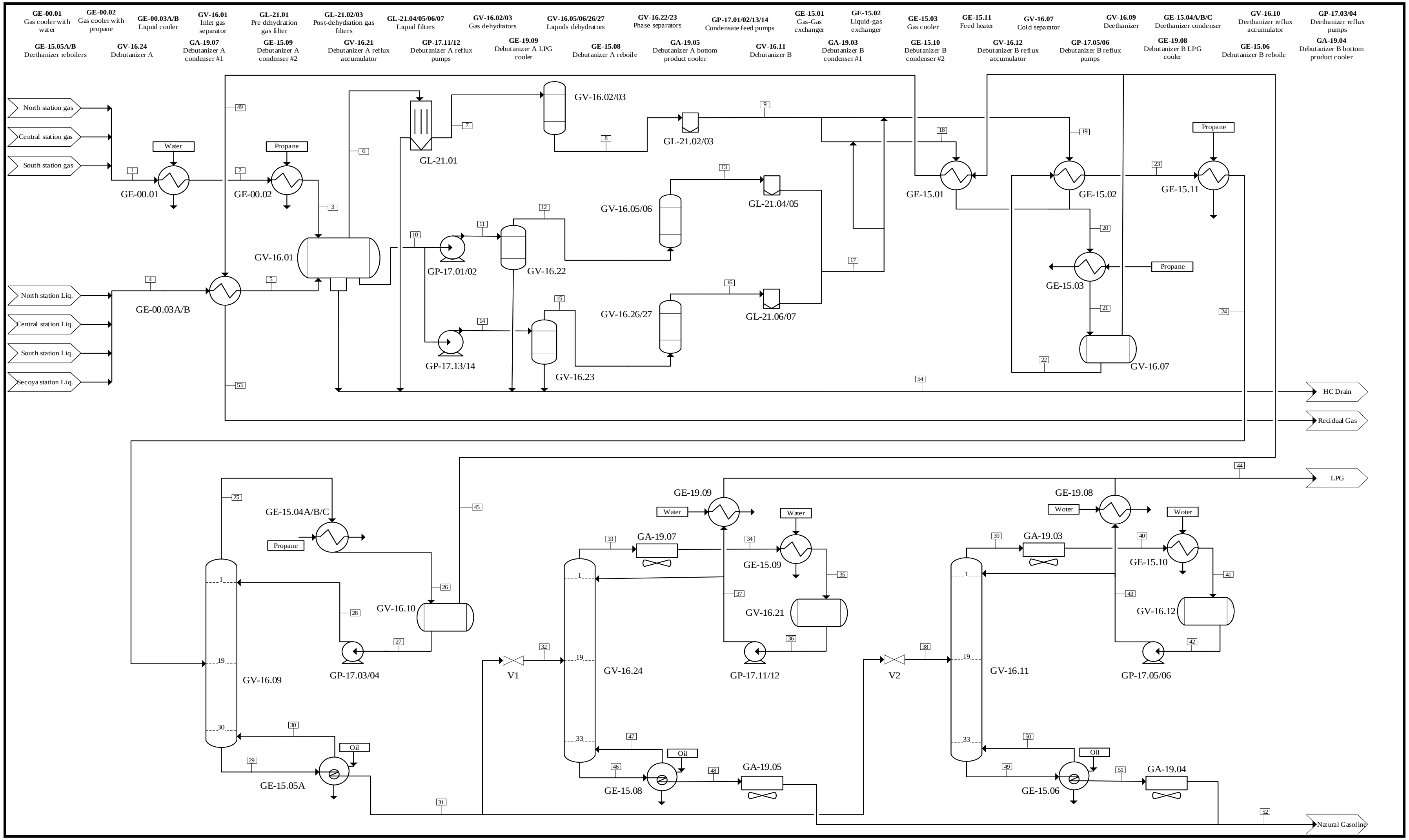


Figure 11. *ShGP: process flow diagram (PFD)*

3.4.Unit Operations and Equipment Operating Conditions

Tables 3, 4 and 5 show the unit operations necessary to LPG production in each section. Besides, the operational conditions and equipment specifications are presented.

Table 3. *Unit operations and equipment data-Reception and conditioning section*

RECEPTION AND CONDITIONING SECTION			
Unit	**Description**	**Data**	
GE-00.01	Gas pre-cooler with water	Hot side: inlet gas	
		Outlet Temperature	100.4 ºF
		Pressure drop	*2.5 psi
		Cold side: water	
		Inlet temperature	89.6 ºF
		Outlet temperature	100.4 ºF
		Inlet pressure	*73.48 psia
		Pressure drop	*2.5 psia
GE-00.02	Gas pre-cooler with propane	Hot side: inlet gas	
		Outlet Temperature	80 ºF
		Pressure drop	*2.5 psi
		Cold side: propane	
		Inlet temperature	34.7 ºF
		Outlet temperature	36.14 ºF
		Inlet pressure	73.96 psia
		Pressure drop	*2.5 psi
GE-00.03 A/B	Liquid pre-cooler	Hot side: inlet liquids	
		Outlet Temperature	80 ºF
		Pressure drop	*2.5 psi
		Cold side: residual gas	
		Pressure drop	*2.5 psi
GV-16.01	Inlet gas separator	Pressure Drop	0
		Adiabatic	
GL-21.01	Pre dehydration gas filter	Output water/Input water	*0.0001
GP-17.01/02/13/14	Condensate feed pumps	Pressure rise	35.5584 psi
GV-16.22/23	Phase separators	Pressure Drop	0
		Adiabatic	
GV-16.02/03/05/06/26/27	Gas and liquids dehydrators	Molecular sieve: zeolite	

*Assumed data (from engineering best practices)

Table 4. Unit operations and equipment data-Cooling section

COOLING SECTION			
Unit	**Description**	**Data**	
GE-15.01	Gas-Gas exchanger	Hot side: inlet liquid&gas	
		Outlet Temperature	-17 ºF
		Pressure drop	*2.5 psi
		Cold side: residual gas	
		Pressure drop	*2.5 psi
GE-15.02	Liquid-gas exchanger	Hot side: inlet liquid&gas	
		Outlet Temperature	-22 ºF
		Pressure drop	*2.5 psi
		Cold side: condensed	
		Pressure drop	*2.5 psi
GE-15.03	Gas cooler	Hot side: inlet liquid&gas	
		Outlet Temperature	-40 ºF
		Pressure drop	*2.5 psi
		Cold side: propane	
		Inlet temperature	-40 ºF
		Outlet temperature	-32.8 ºF
		Inlet pressure	88.99 psia
		Pressure drop	*2.5 psi
GV-16.07	Cold separator	Pressure Drop	0
		Adiabatic	

*Assumed data (from engineering best practices)

Table 5. *Unit operations and equipment data-Fractionation section*

FRACTIONATION SECTION			
Unit	**Description**	**Data**	
GE-15.11	Feed heater	Hot side: Propane	
		Inlet temperature	126.14 ºF
		Outlet temperature	97.88 ºF
		Inlet pressure	166.98 psia
		Pressure drop	*2.5 psi
		Cold side: condensed	
		Outlet temperature	90 ºF
		Pressure drop	*2.5 psi
GV-16.09	Deethanizer	Actual number of trays	30
		Overall tray efficiency	*55%
		Feed location	Tray 10
		Top tray pressure	*Defined to inlet stream -
		Column pressure drop	0.5
		Partial condenser	*5 psi
		Initial Estimates	
		Top tray temperature	-2 ºF
		Reboiler temperature	235 ºF
		Specifications	
		Reflux rate	159.4 gpm
		Overhead product rate	10.82 MMscfd
		Bottom product rate	322 gpm
GV-16.24/11	Debutanizer	Actual number of trays	30
		Overall tray efficiency	*55%
		Feed location	Tray 10
		Top tray pressure	*Defined to inlet stream -
		Column pressure drop	0.5
		Initial Estimates	*5 psi
		Specifications	
		Reflux rate	137.6 gpm
		Overhead product rate	91.4 gpm
		Bottom product rate	33 gpm
GE-19.08/09	GLP cooler	Hot side: GLP	
		Outlet Temperature	100 ºF
		Pressure drop	*2.5 psi
		Cold side: water	
GA-19.04/05	Bottom cooler	Hot side: natural gasoline	
		Outlet Temperature	110 ºF
		Pressure drop	*2.5 psi
		Cold side: utility air	
V1/2	Valves	Output pressure	215 psia

*Assumed data (from engineering best practices)

CHAPTER IV

4. METHODOLOGY

To implement operational improvements in the ShGP, as an alternative to increasing the availability of LPG in the Ecuadorian market consisted of first identifying the industrial process of LPG production. At this point, the existing unit operations, the equipment operating conditions, and the different sections of the ShGP were recognized, based on public and available information. Second, an analysis of the ShGP operation was carried out using simulation tools and real data (stream flows and composition, operating conditions, and equipment specifications). Third, operational improvements were proposed that could be implemented in the ShGP, considering modifications in operating conditions, possible installation of expansion valves, and incorporating a new cooling unit. Finally, a technical and economic feasibility analysis of the proposed operational improvements was carried out. Figure 12 shows a schematic representation of the methodology developed in this study.

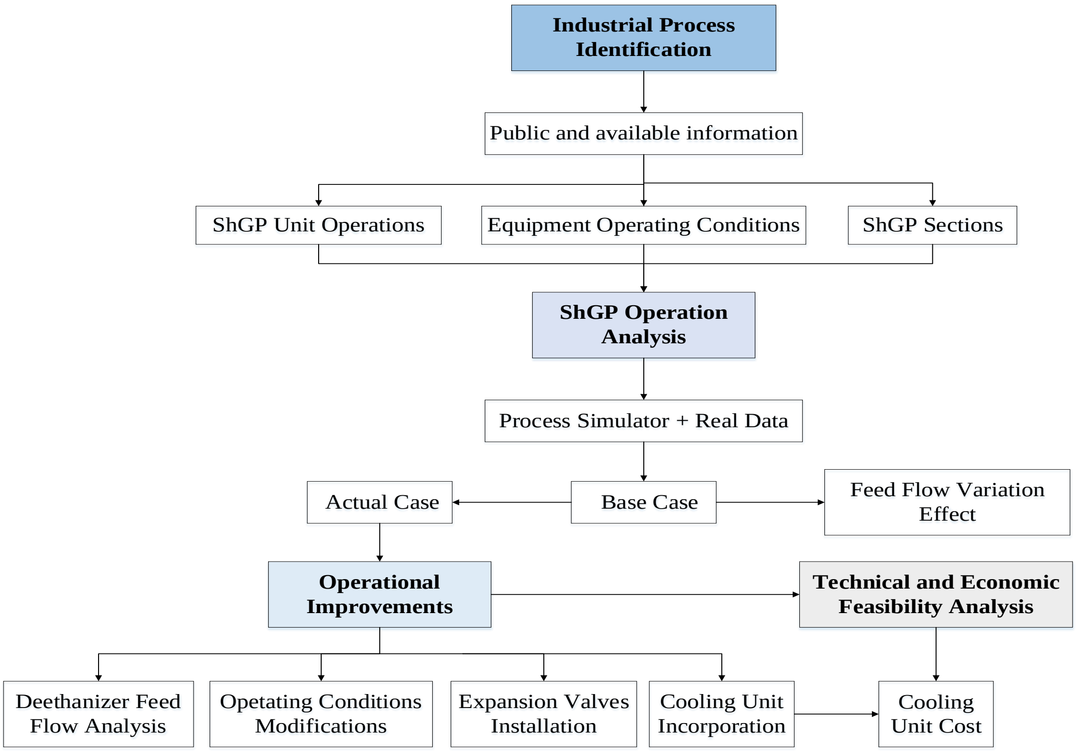


Figure 12. *Methodology diagram*

4.1. Industrial Process Identification

The first step to implementing operational improvements in the ShGP was to identify existing unit operations, equipment operating conditions, and the different sections of the Shushufindi Gas Plant. For this, ten degree-thesis related to this study (dated from 2006-2018), documents and reports of the EP Petroecuador company for the last three years, and documents of the Ecuadorian Standardization Service (INEN) and the American Society for Testing and Materials (ASTM) were reviewed. Besides, various process flow diagrams of the plant were reviewed and analyzed.

The description of the ShGP, the installed equipment, the operating conditions, and the process for obtaining LPG were detailed in Chapter III.

4.2. ShGP Operation Analysis

4.2.1. Base Case

The base case (BC) was established under design bases for the plant's feed, where the ShGP processes 25 MMscfd of natural gas and 150 gpm of condensates, to produce approximately 500 Tm/d of LPG. To build the base case, a series of steps were followed that reproduced a model that will fit the ShGP’s optimal operation. For this, a commercial process simulation software (PRO/II), real data from the ShGP (flow and composition data, operating conditions, and equipment specifications), and assumptions based on engineering best practices were used. In Appendix C, one can find the keyword file (.inp) of the base case.

Using the process simulator, the base case construction consisted on:

1. To build the process diagram in parts (Figure 13), considering the simulation strategies described in section 4.2.3.

2. To establish the unit system (English System).

3. To define the components of the streams.

4. To select the Peng Robinson (PRO1) and Glycol (GLYC01) thermodynamic methods for calculating the physical properties of the components and fluids handled.

5. To define the feed streams (pressure, temperature, flow, and composition) using the data in Table 2 (Chapter III).

6. To enter the equipment operating conditions using tables 3, 4 and 5 of chapter III.

7. To run the simulation, and analyze the results.

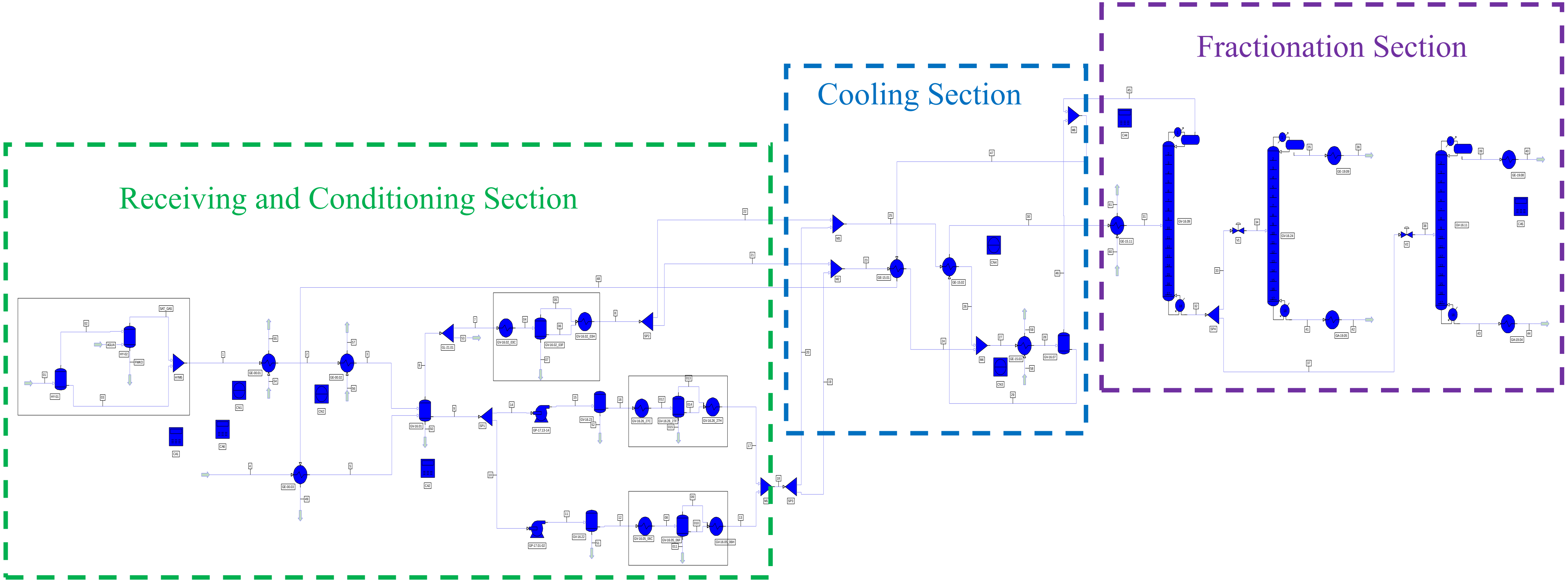


Figure 13. *Process flow diagram*

4.2.2. Actual Case

The actual case (AC) starts from the base case, and it is constituted in the current feeding conditions of the ShGP. At the present, the ShGP operates at approximately 50% of its capacity and processes 14.14 MMscfd of natural gas and 78.83 gpm of condensates, to produce 249.73 Tm/d of LPG according to data reported by EP Petroecuador [15].

4.2.3. Simulation Strategies

In the industrial process of LPG production in the ShGP, there are stream conditions and some equipment that the PRO/II process simulator does not handle. Therefore, it was necessary to define some simulation strategies that reproduced a model that would fit the ShGP's current operation. These strategies are described below.

- **Water saturation system**

The chromatographic analysis of the natural gas processed in the ShGP does not show the water content in the composition since, in practice, it is assumed that the feed gas is saturated with water at the plant inlet conditions. However, the operation of the ShGP involves units responsible for removing water from the process streams. Therefore, in this study, a water saturation system was established for the natural gas stream to estimate the water content in the ShGP feed.

This saturation system was established at ShGP input operating conditions (120 ºF and 565 psia) and is mainly made up of two separators (SA-01 and SA-02), as shown in Figure 14. In SA-01, after the natural gas stream enters, the gas is separated from the liquid hydrocarbons. Next, the gas stream enters SA-02 separator, where a flow of water also comes. In the SA-02 separator, free water (FWKO) is eliminated at the bottom and saturated gas (Sat_Gas) at the top. Finally, the Sat_Gas stream and the liquid hydrocarbons from the first separator are brought together to form the inlet gas stream to the ShGP. To verify the water content present in the gas stream (160 lb. water/MMscf of wet gas to 120 ºF and 565 psia), a water content diagram of the hydrocarbon gases was used as a function of temperature and pressure [26].

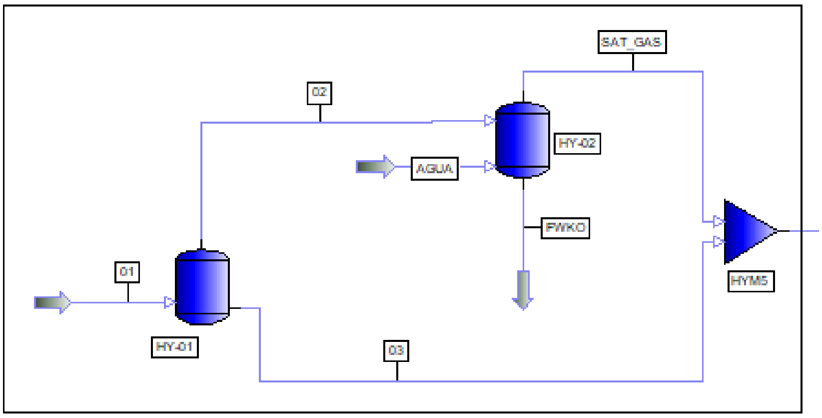


Figure 14. *Simulation strategies: water saturation system*

- **Dehydration System**

In the ShGP, the dehydration of gas and condensates is carried out in molecular sieves; however, the PRO/II process simulator does not handle this equipment. Therefore, in this study, a dehydration system was established for the gas line and the condensate line, with the same operating principle. This system is composed of a cooler (GV-C) that promotes the condensation of water, a separator (GV-F) where water is eliminated, and a heater (GV-H) that allows returning to the line operational conditions. Figure 15 illustrates the condensates dehydration system.

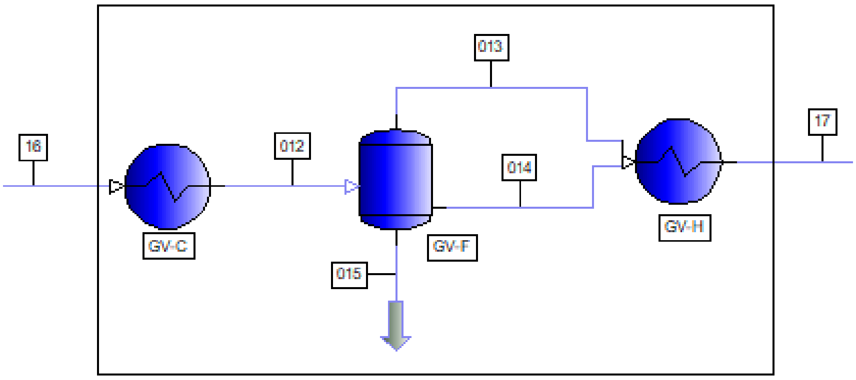


Figure 15. *Simulation strategies: condensates dehydration system*

4.2.4. Feed Flows Variation Effect

This analysis of feed flows variation effect starts from the base case and studies the behavior of LPG production (Tm/d) and quality (C_3 content in LPG) when varying the feed flow rates of ShGP. For this, three cases were proposed that have the purpose of knowing which are the minimum conditions that guarantee the production and quality of LPG in the ShGP.

- **Feed Flow Percentage Variation**

The first case consisted of varying the natural gas and condensates flows in percentage until reaching the base case feeding conditions (10, 30, 50, 70, 90%, and CB). Also, these variations were made considering the natural gas and condensates flows fluctuations from the different stations of Shushufindi [14,37]. Table 6 shows the percentage variation in feed rates.

Table 6. *Feed flows percentage variation*

Variation (%)	**10**	**30**	**50**	**70**	**90**	**BC**
Natural Gas (MMscfd)	2.50	7.50	12.50	17.50	22.50	25
Condensates (gpm)	15	45	75	105	135	150

- **Natural Gas Variation**

In the second case, the natural gas flow was varied until reaching the base case feeding conditions (5, 10, 15, 20, and 25 MMscfd). Furthermore, the condensate flow was kept constant. These variations were made considering the natural gas flows fluctuations from the North, South and Central stations of Shushufindi field [14,37]. Table 7 shows the corresponding variations.

Table 7. *Natural gas variation*

Variation Nº	**1**	**2**	**3**	**4**	**BC**
Natural Gas (MMscfd)	5	10	15	20	25
Condensates (gpm)	150	150	150	150	150

- **Condensates Variation**

In the third case, the condensate flow was varied until reaching the base case feeding conditions (25, 50, 75, 100, and 150 gpm). Also, the natural gas flow was kept constant. These variations were made considering the condensates flows fluctuations from the North, South, Central and Secoya stations of Shushufindi field [14,37]. Table 8 presents the corresponding variations.

Table 8. *Condensates variation*

Variation Nº	**1**	**2**	**3**	**4**	**BC**
Natural Gas (MMscfd)	25	25	25	25	25
Condensates (gpm)	25	50	75	100	150

4.3. Operational Improvements

4.3.1. Deethanizer Feed Flow

The analysis started from the actual case and consisted of varying the deethanizer feed flow until reaching the deethanizer feed flow of case base without modifying the fractionation columns' operating conditions. This analysis aims to determine the feasibility of making operational modifications in the previous sections to the fractionation to increase the deethanizer feed flow and the LPG supply in the Ecuadorian market. After this analysis, below are described as possible operational improvements to increase the deethanizer feed flow. These alternatives are described from least to greatest economic impact and in infrastructure.

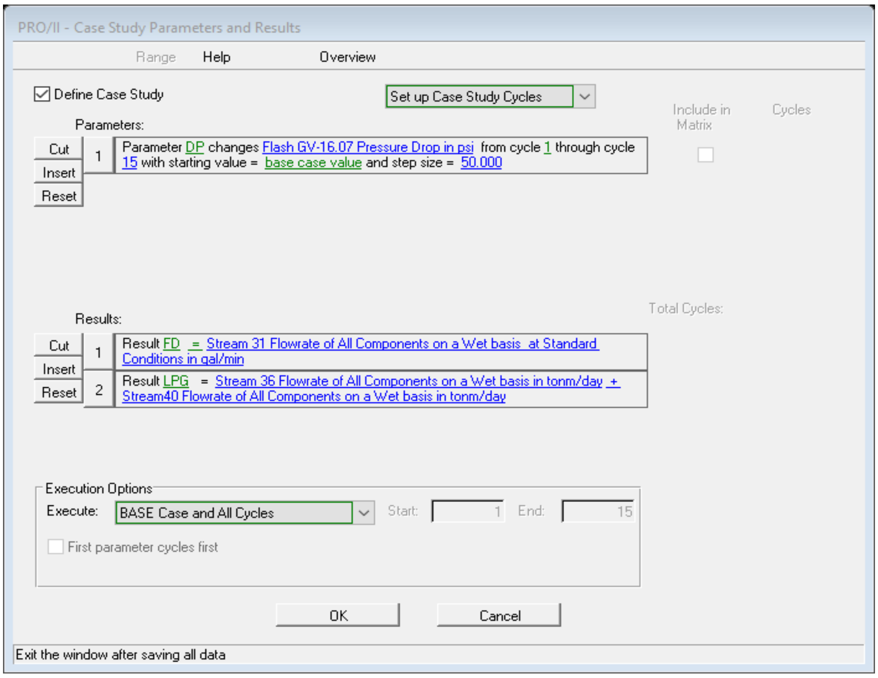
4.3.2. Operating Conditions Modifications

Natural gas liquids recovery generally involves the condensation of hydrocarbons by lowering the temperature [22]. Also, one of the possible alternatives to increase the deethanizer feed flow, with low economic and infrastructure impact, is the modification of the operating conditions. Therefore, this analysis started from the actual case and consisted of examining the effect of the deethanizer feed flow and the LPG production by varying the equipment operating conditions previous to the fractionation section described in Table 9.

Table 9. *Operating conditions modifications: equipment analysis*

Unit	Description	Modificable operating conditions
GE-00.01	Gas pre-cooler with water	Hot side pressure drop and Hot product temperature
GE-00.02	Gas pre-cooler with propane	Hot side pressure drop and Hot product temperature
GE-00.03 A/B	Liquid pre-cooler	Hot side pressure drop and Hot product temperature
GV-16.01	Inlet gas separator	Pressure drop and Duty
GV-16.22/23	Phase separators	Pressure drop and Duty
GE-15.01	Gas-Gas exchanger	Hot side pressure drop and Hot product temperature
GE-15.02	Liquid-gas exchanger	Hot side pressure drop and Hot product temperature
GE-15.03	Gas cooler	Hot side pressure drop and Hot product temperature
GV-16.07	Cold separator	Pressure drop and Duty

In this analysis, the Pro/II Study Case tool was used. As an example, Figure 16 shows the case study for the GV-16.07 cold separator, where the variable to be manipulated is located in the parameter section, and the variables to be observed are located in the results section.



☑ Define Case Study

Set up Case Study Cycles

Include in Matrix ☐

Cycles

Parameters:

1 Parameter DP changes Flash GV-16.07 Pressure Drop in psi from cycle 1 through cycle 15 with starting value = base case value and step size = 50.000

Total Cycles:

Results:

1 Result FD = Stream 31 Flowrate of All Components on a Wet basis at Standard Conditions in gal/min

2 Result LPG = Stream 36 Flowrate of All Components on a Wet basis in tonm/day + Stream40 Flowrate of All Components on a Wet basis in tonm/day

Execution Options

Execute: BASE Case and All Cycles Start: 1 End: 15

☐ First parameter cycles first

Exit the window after saving all data

Figure 16. *PRO/II-CASE study parameters and results dialog box to the cold separator*

4.3.3. Expansion Valves Installation

One of the commonly used methods to achieve high levels of liquids recovery from natural gas is through the Joule-Thomson expansion, where refrigeration of the gas is achieved by expanding it, from high to low pressure [25]. Besides, valve installation is one of the lowest investments that can be made to achieve certain goals. Therefore, in this study, the possible installation of valves in streams before separation was analyzed, hoping to increase the deethanizer feed flow and LPG production. The analysis consisted of incorporating valves before the GV-16.01, GV-16.22, GV-16.23, and GV-16.07 separators, and examining the effect of the deethanizer feed flow by varying the pressure drop of these valves. Figure 17 shows the process diagram of the ShGP with the valves incorporated.

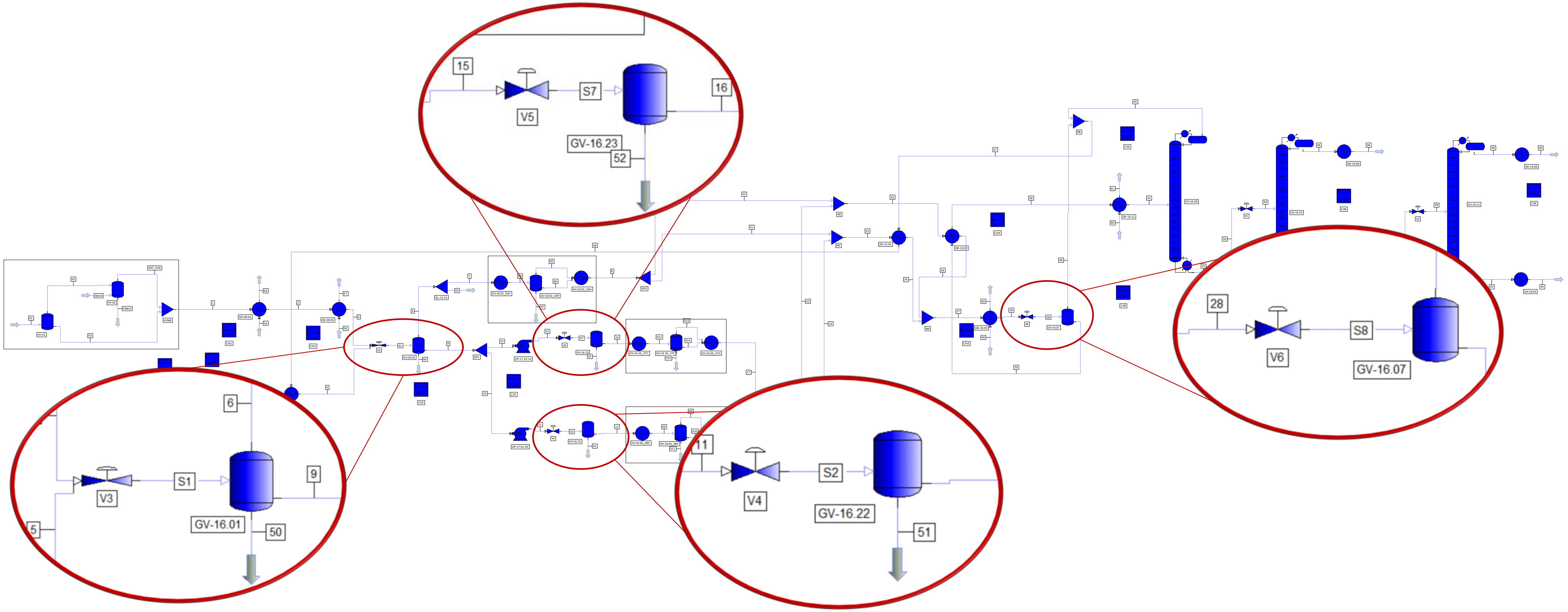


Figure 17. *ShGP process diagram with J-T expansion valves (capture)*

4.3.4. Cooling Unit Incorporation

The third alternative to increase the deethanizer feed flow is a possible installation of new heat exchange equipment in the cooling section, which must cool the process stream below -40 ºF. In this case, an ethane-base refrigerant was considered since it has a boiling temperature of -127.48 ºF. This alternative is the one that has the most significant economic and infrastructure impact since it requires the acquisition and installation of a new refrigeration system, as well as a change in operational philosophy.

The new ethane-cooled heat exchanger GE-C_2 (Figure 18), which will be incorporated after the Gas cooler GE-15.03, aims to lower the temperature of the process stream to increase the feed flow of the deethanizer. In this exchanger, the process stream cools from -40 to -100 ºF, while the ethane stream from -101 to -100 ºF.

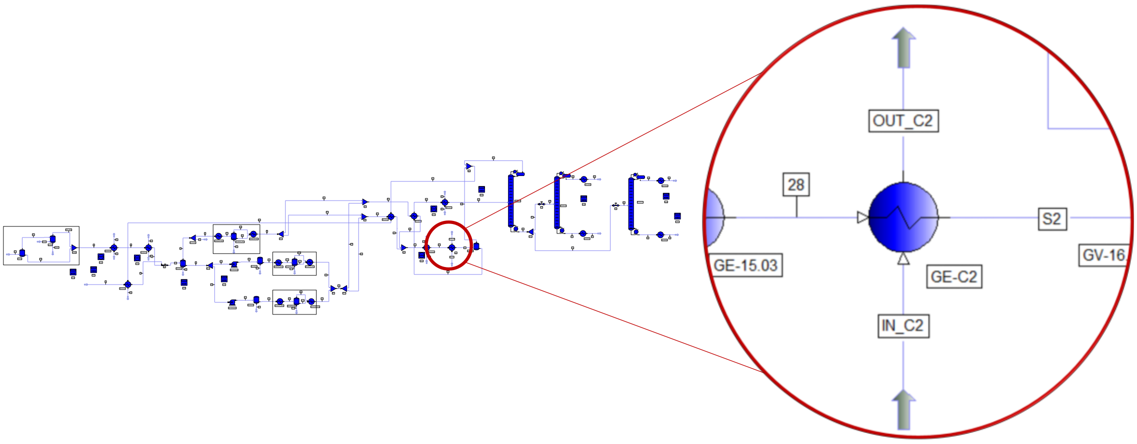


Figure 18. *ShGP process diagram with GE-C_2 heat exchanger (capture)*

4.4. Economic Feasibility Study

The economic feasibility study of this work is mainly focused on the cost estimating of the GE-C2 ethane cooled exchanger since it is the main equipment to achieve our objectives. However, a detailed economic study is required to define the final cost associated with the implementation and start-up of the new cooling unit.

4.4.1. Heat Exchanger Cost Estimate

The cost of the GE-C2 exchanger is estimated using the equipment module cost technique introduced by Guthrie (1968) [41]. This method consists of calculating the module adjusted cost in 1968 (C_{md1968}) using the following expression.

$$C_{md1968} = C_{md} + \left(C_{fob} - C_b\right) \tag{1}$$

where:

C_{md}: adjusted module cost.

C_{fob}: adjusted base cost.

C_b: unit base cost in 1968.

The base cost (C_b) (reference cost in 1968) is estimated based on the heat exchange area (A) of the GE-C2 cooling unit using Figure 19.

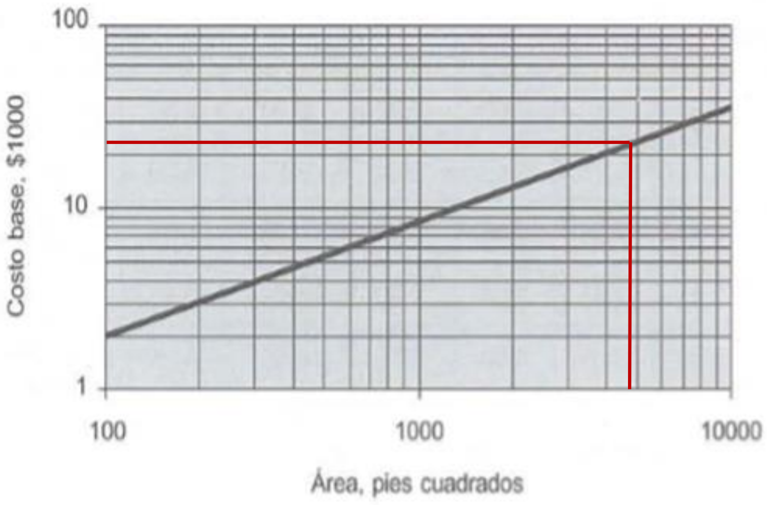


Figure 19. *Base cost of a heat exchanger (reference year: 1968)*

To obtain the adjusted modulus cost (C_{md}), expression two, and the modulus factor for heat exchangers from Table B.1 (Appendix B) are used.

$$C_{md} = C_b * Modulus\ factor \tag{2}$$

To calculate the adjusted base cost (C_{fob}), the following expression, and the adjustment factors from Table B.2 are used.

$$C_{fob} = (C_b * (Fd + Fp) * Fm) \tag{3}$$

where:

Fd: module that represents the type of exchanger.

Fp: module that represents the working pressure.

Fm: module that represents the equipment material.

To carry the equipment cost to the most recent year, the expression four and the Chemical Engineering Plant Cost Index (CPECI) of 1968 (113.7) and the most recent year published (2019 = 607.5) are used.

$$Cost\ 2019 = \frac{Index\ in\ actual\ year}{Index\ in\ 1968} * C_{md1968} \tag{4}$$

Finally, Guthrie recommends using a factor of 15%. Besides, to estimate the cost of this equipment in the Ecuadorian market, the cost obtained was increased by 50%. This increase refers to transport and insurance costs and customs duties and taxes.

- Heat Exchange Area

The following expression was used to calculate the heat exchange area which is used to estimate the GE-C2 cost.

$$A = \frac{Q}{U*LMTD} = 4840.41\ ft^2 \tag{5}$$

where:

A: heat exchange area (ft^2).

Q: power exchanged by the fluids (Btu/h).

U: global coefficient of thermal transfer (Btu/ft^2*ºF).

$LMTD$: mean logarithmic difference of temperature (ºF).

The power exchanged by the fluids is determined by:

$$Q = \dot{m} * C_P * \Delta T \quad (6)$$

where:

$\dot{m}$: mass flow (lb/h).

Cp: specific heat (Btu/lb*ºF).

ΔT: inlet temperature minus outlet temperature of the stream (ºF).

The average logarithmic temperature difference is calculated as follows:

$$LMTD = \frac{(Te1-Te2)-(Ts1-Te2)}{\ln\left(\frac{Te1-Ts2}{Ts1-Te2}\right)} \quad (7)$$

where:

$Te_{1:}$ inlet temperature of cold side (ºF).

$Ts_{1:}$ outlet temperature of cold side (ºF).

$Te_{2:}$ inlet temperature of hot side (ºF).

$Ts_{2:}$ outlet temperature of hot side (ºF).

The value of the global coefficient of thermal transfer (Btu/h*ft^2*ºF) was estimated based on data reported by Coulson & Richardson's [42].

CHAPTER V

5. RESULTS

5.1. ShGP operation analysis

5.1.1. Base Case

Upon reaching the convergence of the base case in the process simulator, the GPM or liquid hydrocarbon content expressed in gallons that can be obtained for every 1000 cubic feet of natural gas at standard conditions was first quantified [23]. The GPM for ShGP natural gas is 12.16, which means it is a rich gas.

Table 10 shows the ShGP feed flows at design bases, the LPG production and quality according to the data reported by EP Petroecuador and the Ecuadorian regulatory agencies, and the data obtained from the simulation. According to the design bases, the ShGP has the capacity to process 25 MMscfd of natural gas and 150 gpm of condensates to produce approximately 500 Tm/d of LPG. Although, for the base case construction, public and available information were used, assumptions based on the engineering best practices were made, and simulation strategies were implemented, which are described in the previous chapter; the LPG recovery in the process simulator was 499.81 Tm/d, with an error of 0.04% concerning the data reported.

According to the LPG quality, it is important to point out that the LPG composition of the simulation could not be compared with reported data since these are not available for public access. However, according to Ecuadorian regulation, Ecuadorian LPG can contain a minimum of 60% of C_3 and a maximum of 40% of C_4. Therefore, the LPG recovered in the simulation is within the quality specifications.

Table 10. *LPG production and quality-Base Case*

Feed Flow		LPG Production		LPG Quality	
		Data	**Simulation**	**Normative**	**Simulation**
Design flow	Natural Gas: 25 MMscfd Condensates: 150 gpm	500 Tm/d	499.81 Tm/d	Min C_3 60% Max C_4 40%	C_3 99% C_4 1%

5.1.2. Actual Case

Table 11 shows the results for the actual case. At present, the ShGP processes 14.14 MMscfd of natural gas and 78.83 gpm of condensates. When these feed rates are entered in the base case simulation, the LPG production is 258.52 Tm/d (optimal production for these feed rates). However, the current LPG production reported for ShGP is 249.73 Tm/d; 8.79 Tm/d less than the expected optimal production. Furthermore, if optimal LPG production is achieved in the ShGP, imports would be reduced by 0.31%. All this shows us the need to implement operational improvements in the ShGP to reach its maximum performance and, therefore, increase LPG availability in the Ecuadorian market. Concerning quality, the GLP of the simulation is within the regulations' parameters since the C_3 content is 99.73%.

Table 11. *LPG production and quality-Actual Case*

Feed Flow		LPG Production		LPG Quality	
		Data	**Simulation**	**Standards**	**Simulation**
Actual Flow	Natural Gas: 14.14 MMscfd Condensates: 78.83 gpm	249.73 Tm/d	258.52 Tm/d	Min C_3 60% Max C_4 40%	C_3 99.73% C_4 0.27%

5.1.3. Feed Flow Percentage Variation

Figure 20 shows the LPG production (Tm/d) and quality (C_3 content) versus the feed flows percentage variation. Where, as the percentage variation in feed flows increases, LPG production increases linearly until reaching the feed established in the base case. Furthermore, LPG quality complies with the parameters established by Ecuadorian regulatory agencies, and remains constant for all variations. According to the design bases, each debutanizer has the capacity to process 250 Tm/d of LPG. Therefore, under 50% variation in feed flows, the plant can produce LPG only using a debutanizer since, in this percentage variation, the plant produces 250 Tm/d of LPG. According to EP Petroecuador reports, the ShGP is currently working at 50% of its capacity, so that the plant could generate its production in one of the debutanizer plants.

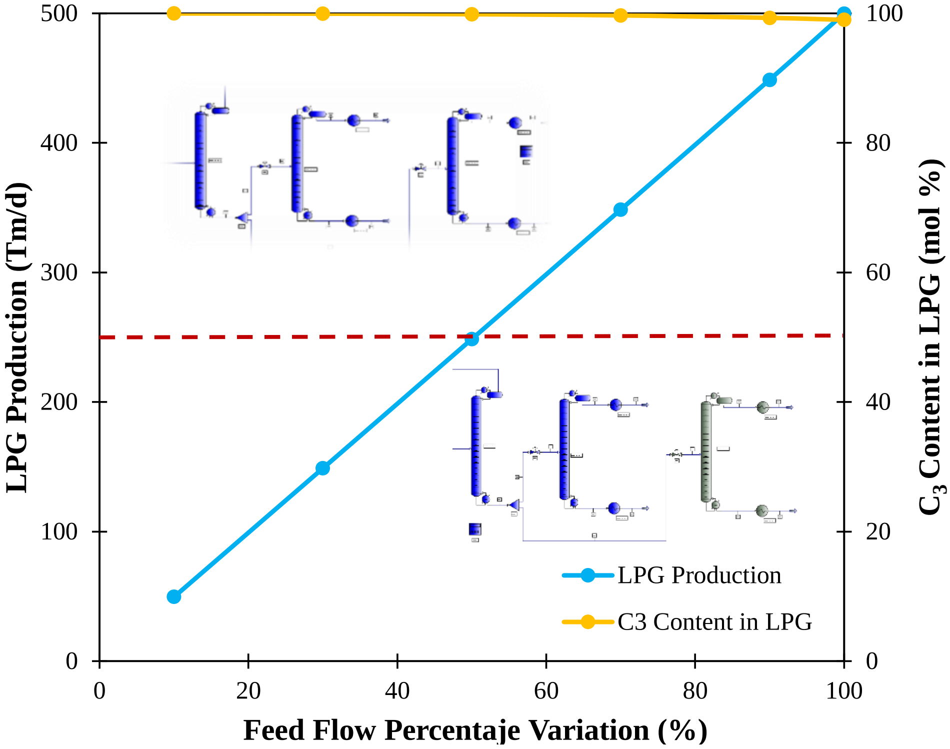


Figure 20. *LPG production and quality vs. Feed flows percentage variation*

Furthermore, taking into account the sector with the highest demand for this fuel (domestic sector), and the 15 Kg cylinders used in this sector, Figure 21 represents the LPG production in 15 Kg cylinders for variations considered. Currently, the plant produces daily (249.73 Tm/d) 16,648 15 Kg LPG cylinders (LPG Cyl-15 Kg/d). However, if the plant operates at 50% of its capacity, an approximate production of 17,000 LPG Cyl-15 Kg/d would be expected. This means that approximately 35,000 daily LPG cylinders could be marketed in the national market if the plant had optimal production.

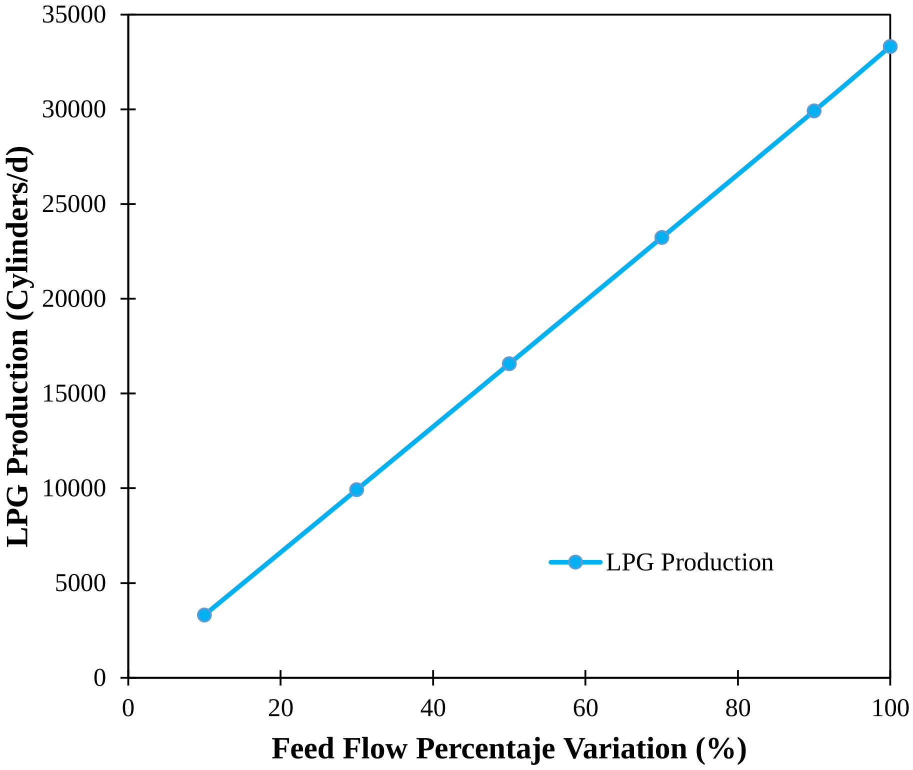


Figure 21. *LPG production in 15 Kg cylinders vs. Feed flows percentage variation*

5.1.4. Natural Gas Variation

Figure 22 represents the LPG production and quality versus the natural gas variation. Where, as the natural gas flow increases, the LPG production, and quality increase. Also, from 9 MMscfd of natural gas, the LPG meets the quality specifications since the C_3 content in the LPG is $\geq 60\%$. These variations were made considering the natural gas flow fluctuations from the Central, South, and North stations of Shushufindi [14,37].

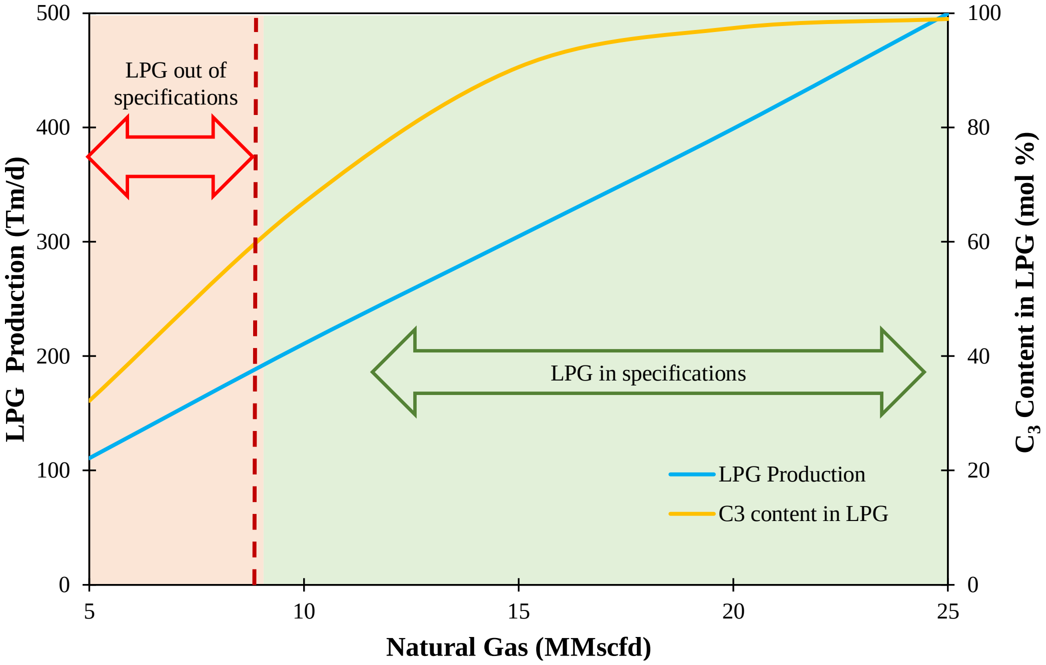


Figure 22. *LPG production and quality vs. Natural gas variation*

5.1.5. Condensates Variation

Figure 23 shows the LPG production and quality versus the condensates feed variation. Where, as the condensate flow increases, the LPG production and quality increase. However, below 40 gpm of condensate, LPG does not meet quality specifications. These variations were made considering the condensates flow fluctuations from the Central, South, North, and Secoya stations of Shushufindi [14,37].

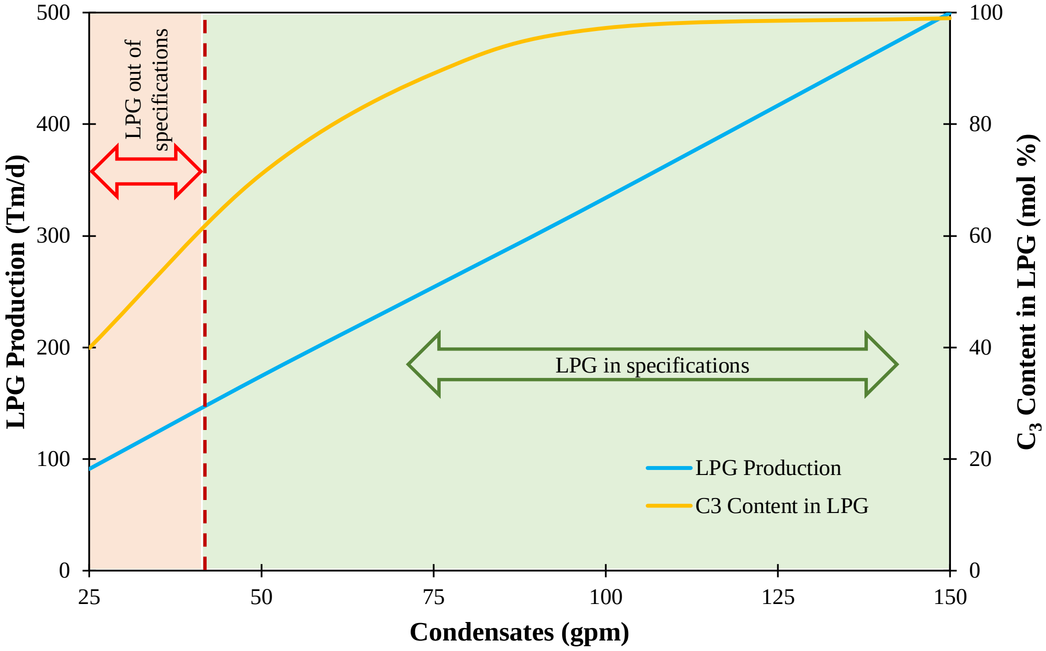


Figure 23. *LPG production and quality vs. Condensates variation*

Variations in feed flows can generate changes in the feed's richness and, therefore, affect the product quality. Therefore, when natural gas or condensates are varied (Figures 22 and 23), the feed richness changes, and consequently, the LPG quality is affected. However, when natural gas and condensates are varied at the same time (Figure 20), the feed richness is maintained, and therefore the LPG quality remains constant.

5.2. Operational Improvements

5.2.1. Deethanizer Feed Flow

When evaluating the variation of the deethanizer feed flow, it was found that it is feasible to make operational modifications in the sections of the plant previous to the fractionation section to increase the deethanizer feed flow and, therefore, the LPG supply in the national market. According to the data reported by the simulation and Figure 24, the feed flow of the deethanizer at current conditions is 233.37 gpm. Furthermore, it could be observed that as the deethanizer feed flow increases, the LPG production has an increasing behavior, and the quality remains constant and within specifications. Upon reaching the deethanizer feed flow at design bases (422.29 gpm), 467.78 Tm/d of LPG were obtained; 32.22 Tm/d of LPG less than the amount established for these conditions.

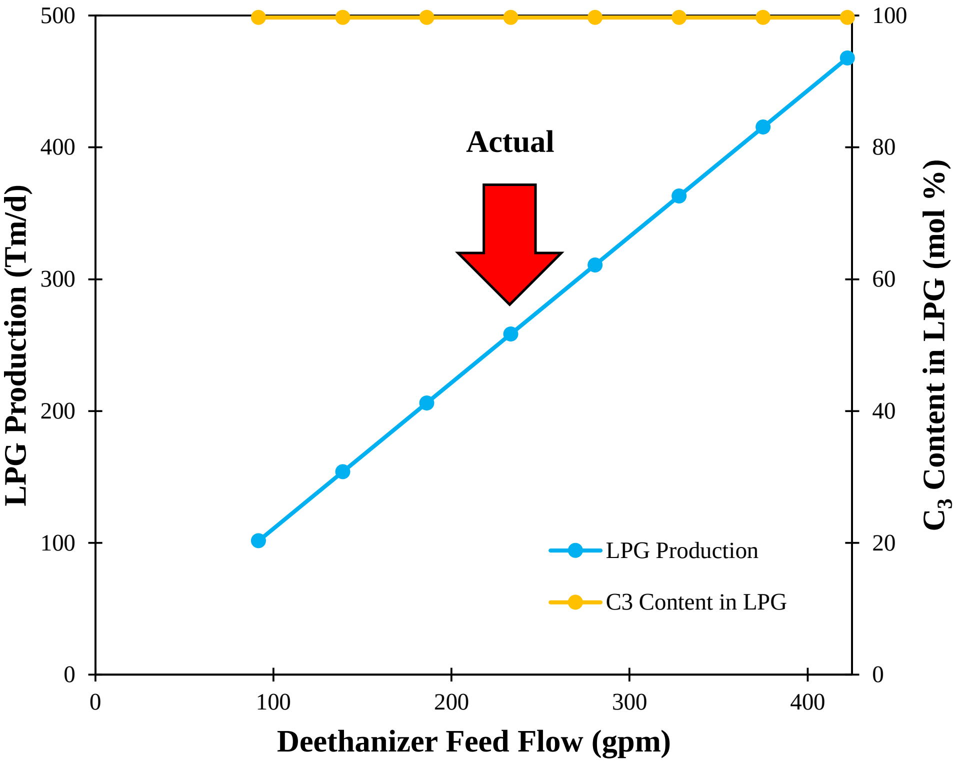


Figure 24. *LPG production and quality vs. Deethanizer feed flow*

5.2.2. Operating Conditions Modifications

Natural gas liquids recovery generally involves the condensation of hydrocarbons by decreasing the temperature [22]. Therefore, in this study, the effect of the deethanizer feed flow, and the LPG production by varying the operating conditions of various equipment (Table 10) were analyzed. One of the analyzed equipment was the cold separator GV-16.07, which operates adiabatically (pressure drop = 0; duty = 0). In this equipment, the effect of pressure drop and duty on the deethanizer feed flow and LPG production was studied. Figure 25 shows that as the pressure drop increases, the deethanizer feed flow and LPG production decrease in parallel. Also, Figure 26 shows a brief increase in the interest flows as the heat decreases (from right to left); however, from -6*10^6 BTU/h, the flows remain constant.

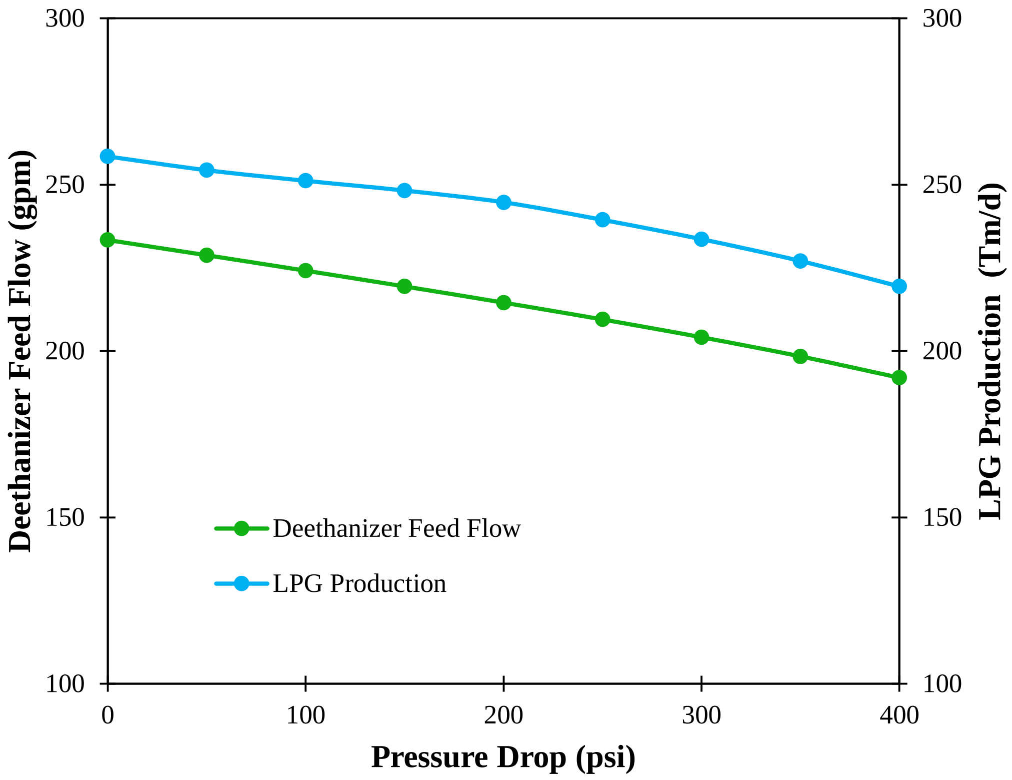


Figure 25. *Deethanizer feed flow and LPG production vs. Pressure drop - Separator GV-16.07*

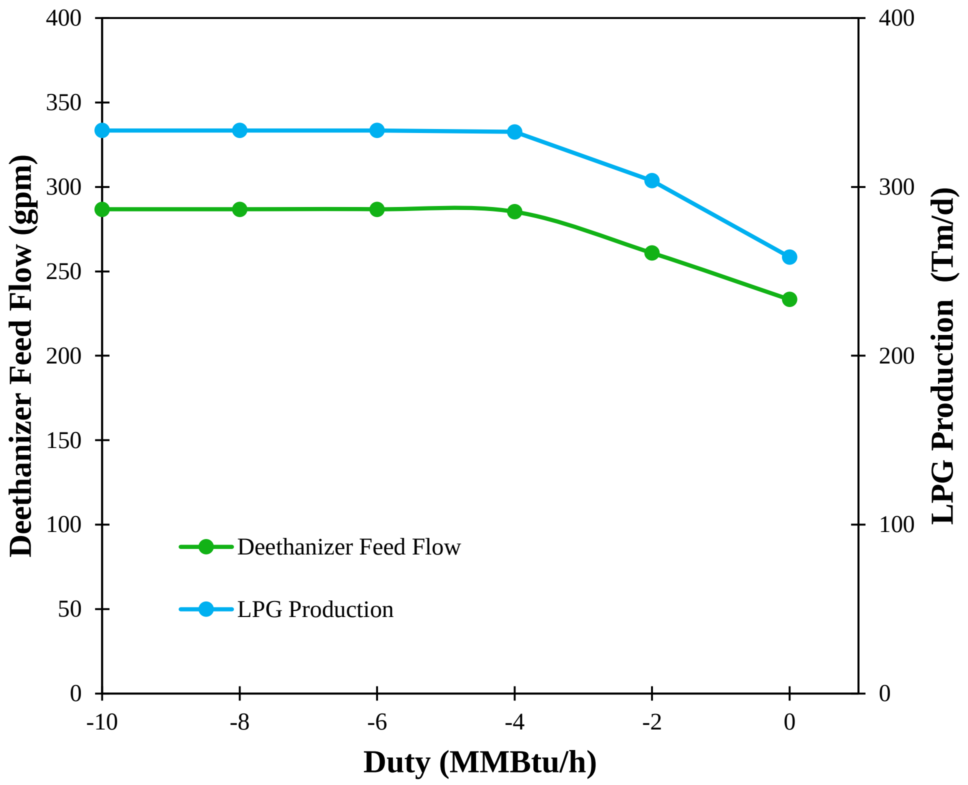


Figure 26. *Deethanizer feed flow and LPG production vs. Duty - Separator GV-16.07*

Another piece of equipment examined was the GE-15.03 gas cooler, which operates at a pressure drop = 2.5 and the specified outlet temperature is -40 ºF. In this equipment, the pressure drop and the hot side outlet temperature were evaluated. Figure 27 shows that the pressure drop's variation does not have a significant effect on the interest flow rates; they remain constant. Besides, Figure 28 shows a significant increase (from right to left) in the deethanizer feed and LPG production as the temperature decreases. Furthermore, it can be seen that from -130ºF to the left, the flow rates become constant. However, this exchanger works with C_3 as a refrigerant with a boiling temperature of -43.67 ºF, which does not allow to decrease the temperature beyond that value. Therefore, to work below -40 ºF, it is necessary to evaluate the refrigeration system and/or use another refrigerant, considering the costs associated with the installation of a new refrigeration system. The possible installation of a new cooling unit that works with ethane as a refrigerant (Boiling temperature = -127.48 ºF) is discussed below.

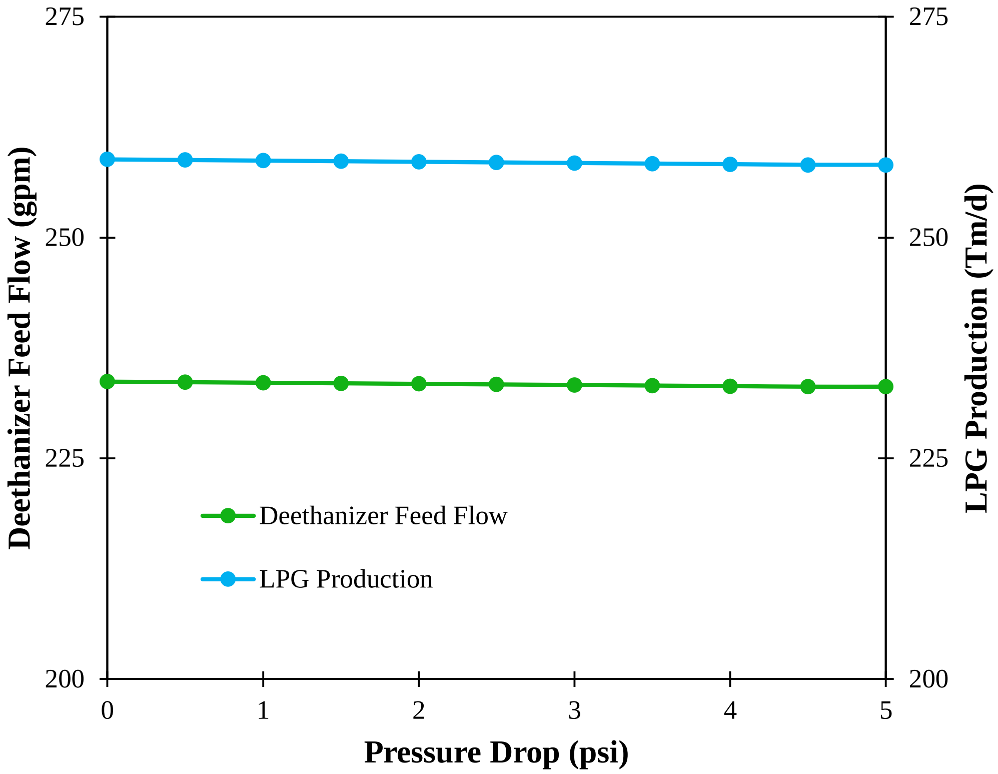


Figure 27. *Deethanizer feed flow and LPG production vs. Pressure drop of Gas Cooler GE-15.03*

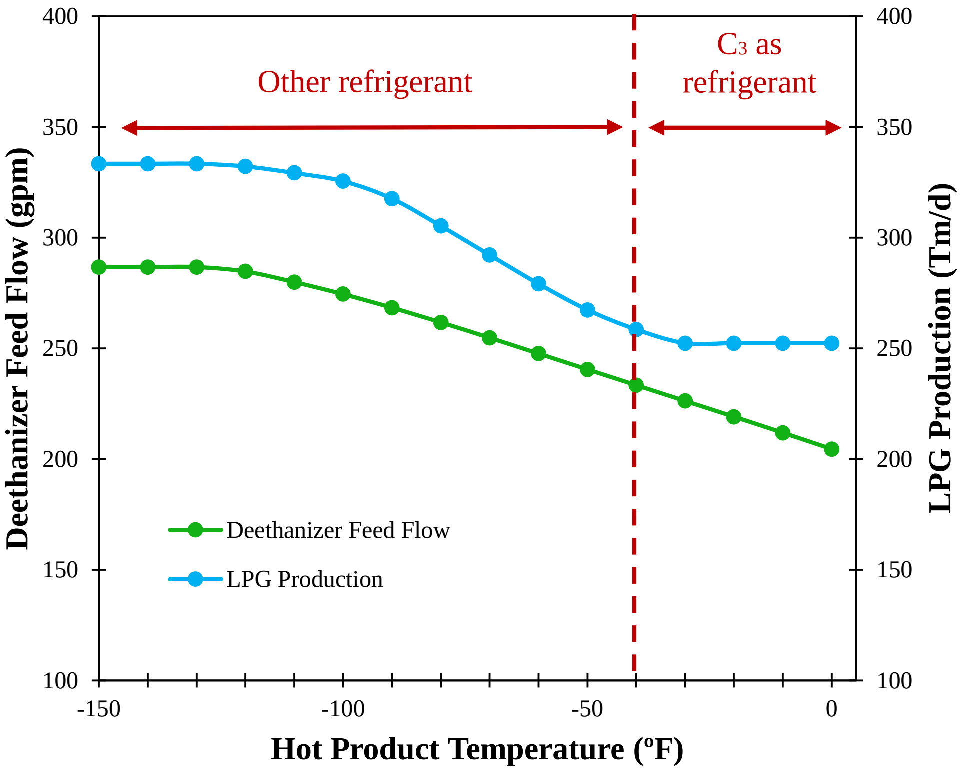


Figure 28. *Deethanizer feed flow and LPG production vs. Hot side out temperature of Gas Cooler GE-15.03*

After examining the effect of the deethanizer feed flow when varying the equipment's operating conditions before the fractionation section, results were obtained that did not significantly favor the increase in LPG production. This could be due to the plant's current configuration not allowing us to appreciate the effect of the decrease in temperature on the amount of final product.

5.2.3. Expansion Valves Installation

One of the commonly used methods to achieve high levels of liquid recovery from natural gas is through the Joule-Thomson expansion, where gas is cooled by expanding from high to low pressure. Therefore, in this study, the possible installation of valves in streams before the separators GV-16.01, GV-16.22, GV-16.23, and GV-16.07 was analyzed. Figure 29 shows the possible valve installation in the stream before the GV-16.07 separator, where it can be observed that as the pressure drop increases, the deethanizer feed flow and the LPG production decrease.

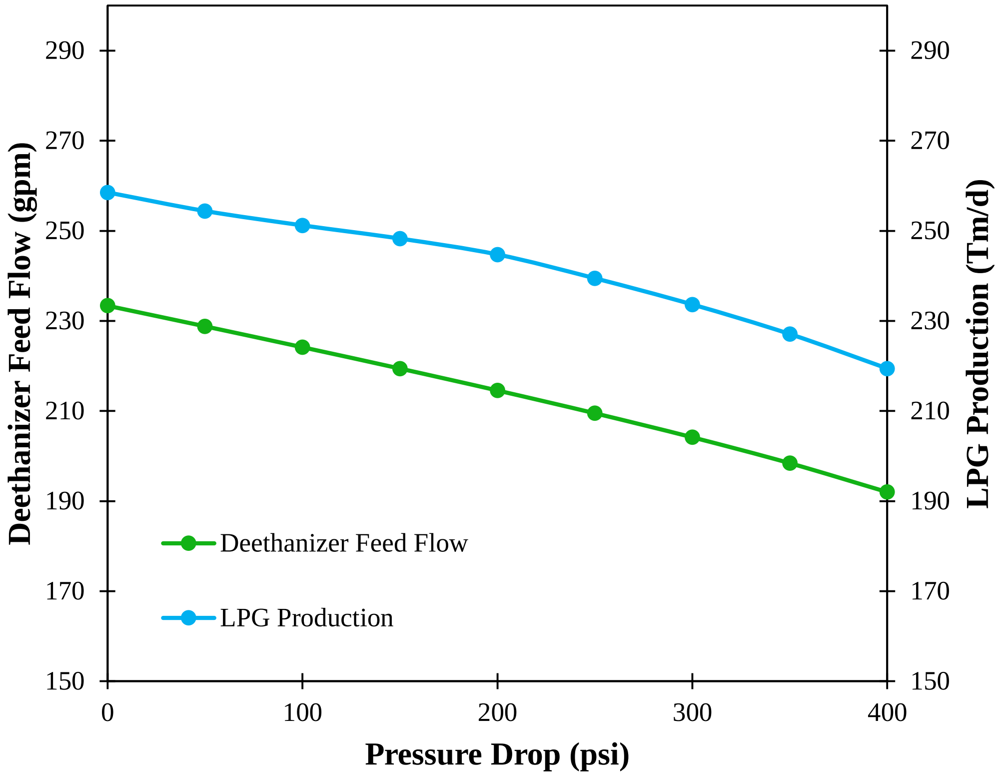


Figure 29. *Deethanizer feed flow and LPG production vs. Pressure drop of valve V6*

The possible installation of expansion valves in pre-separation streams showed an unfavorable result for increased the deethanizer feed flow and LPG production. For all cases where a valve was installed, LPG production always had a decreasing behavior. This may be due to the dynamics of the plant and the operational conditions it currently manages. Therefore, the possible installation of expansion valves is not an alternative to achieve our objectives.

5.2.4. Cooling Unit Incorporation

The incorporation of the GE-C_2 Ethane Cooled Exchanger cools the process stream from -40 ºF to -100 ºF. This temperature change generates an increase in the deethanizer feed flow and, therefore, in LPG production. According to the data obtained in the simulation, the deethanizer feed flow at current conditions (actual feed flows) is 233.37 gpm. However, with the incorporation of the GE-C_2 exchanger, this flow increases to 274.51 gpm.

According to EP Petroecuador, the current LPG production in the ShGP is 249.73 Tm/d. However, at current conditions (actual feed flows), and with the incorporation of the GE-C_2 exchanger, LPG production increases to 325.55 Tm/d. Figure 30 shows the current LPG production in 15 Kg cylinders (16,648 LPG Cyl-15 Kg/d) and the LPG production after incorporating the GE-C_2 heat exchanger (21,703 LPG Cyl-15 Kg/d). This increase represents approximately 30% of ShGP current production. However, this alternative is the one that has the greatest economic and infrastructure impact, since the costs associated with the acquisition and installation of a new cooling unit must be considered, as well as a change in the operating philosophy.

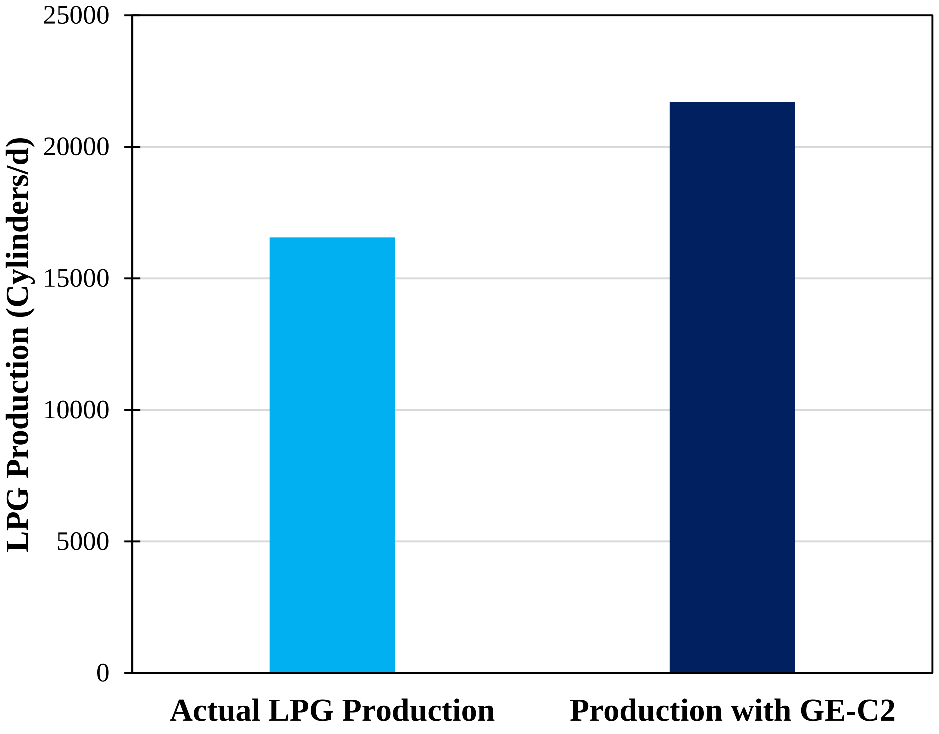


Figure 30. *Current LPG production and LPG production with the incorporation of GE-C_2 exchanger*

5.3. Economic Feasibility Study

The economic feasibility study of this work is mainly focused on estimating the cost of the GE-C2 ethane cooled exchanger since it is the main equipment to achieve our objectives. The estimated cost for the GE-C_2 exchanger is approximately 1.5 MMUSD in the Ecuadorian market, according to the calculation carried out, taking into account the methodology proposed by Guthrie [41]. The current reported production of the ShGP is 249.73 Tm/d of LPG. By incorporating the GE-C_2 exchanger, this production increases to 325.55 Tm/d (75.82 Tm/d more than current production). When analyzed from the point of view of 15Kg cylinders, there is an increase from 16,648 to 21,703 Cyl 15Kg/d, (5,055 LPG Cyl -15Kg/d more than the current production). This increase represents approximately 30% of ShGP current production. However, a detailed economic study is required to define the final cost associated with the implementation and start-up of the new cooling unit.

CONCLUSIONS AND RECOMMENDATIONS

- The base case's LPG production rate was very close to the data reported by EP Petroecuador, obtaining an error of 0.04%, even though public and available data from the ShGP were used, and several assumptions had to be made based on engineering best practices. Furthermore, the quality of LPG is within the parameters established by Ecuadorian regulations.

- According to the actual case analysis, the LPG production obtained in the simulator differs from the current production reported by 8.79 Tm/d of LPG. This difference shows us the need to implement operational improvements in the ShGP to obtain the plant's maximum performance.

- At design bases, each debutanizer generates 250 Tm/d of LPG. Therefore, if the ShGP operates below 50% of its capacity, it could only be considered a debutanizer for LPG production.

- Variations in feed flows can generate changes in the feed's richness and, therefore, affect the product quality. Therefore, when natural gas or condensates are varied, the feed richness changes, and consequently, the LPG quality is affected. However, when natural gas and condensates are varied at the same time, the feed richness is maintained, and therefore the LPG quality remains constant.

- Through the analysis of the deethanizer feed flow, it was possible to know that it is possible to increase the LPG supply to the current feeding conditions through operational modifications in the plant sections before fractionation.

- Among the operational improvement alternatives that were proposed to increase the LPG availability in the national market are the modification of operating conditions, possible incorporation of expansion valves, and integration of a heat exchanger with a refrigerant that works below -40 ºF. The improved operation that showed greater efficiency was incorporating a heat exchanger that works with ethane as a refrigerant (GE-C_2).

- The incorporation of the GE-C_2 exchanger showed an LPG increase of 75.82 Tm/d more than the current production. However, this alternative is the one that has the most significant economic and infrastructure impact since it requires the acquisition and installation of new equipment, as well as a change in operational philosophy.

- The estimated cost for the GE-C_2 exchanger is approximately 1.5 MMUSD in the Ecuadorian market, according to the calculation carried out, taking into account the methodology proposed by Guthrie (1968).

- It is recommended to deepen the cooling system's analysis for the GE-C_2 exchanger since this work focused only on the leading equipment.

- A detailed economic study is recommended to define the final cost associated with the implementation and start-up of the new cooling unit.

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APPENDIX

APPENDIX A:

Liquefied Petroleum Gas Specifications and Requirements

Table A. 1. *Liquefied petroleum gas specifications* [33]

Product Characteristics	**Commercial Propane**	**Commercial Butane**	**Commercial B-P Mixture**	**Propane HD-5**	**Test Method**
Composition	Predominantly propane and/or butylenes	Predominantly butanes and/or butylenes	Predominantly mixture of butanes and/or butylenes with propane and/or propylene	Not less than 90 liquid volume percent propane; not more than 5 liquid volume percent propylene	ASTM D-2163-91
Vapor pressure at 100 ºF, psig, max.	208	70	208	208	ASTM D-1267-95
at 37.8 ºC , KPa, max.	1434	483	1434	1434	
Volatile residue:					ASTM D-1837-94
Temperature at 95% evaporation,					ASTM D-2163-91
Deg. F, max.	-37	36	36	37	ASTM D-2163-91
Deg. C, max.	-38.3	2.2	2.2	-38.3	
Or butane and heavier, liquid volume percent max.	2.5	-	-	2.5	
Pentane and heavier, liquid volume percent max.	-	2.0	2.0	-	
Residual matter:					ASTM D-2158-92
Residue on evaporation of 100 ml, max.	0.05 ml	-	-	0.05 ml	ASTM D-2158-92
Oil stain observation	Pass (1)	-	-	Pass (1)	
Corrosion, copper strip, max.	No. 1	No. 1	No. 1	No. 1	ASTM D-1838-91 (Note A)
Total sulfur, ppmw	185	140	140	123	ASTM D-2784-92 (Note B)
Moisture content	pass	-	-	pass	GPA Propane Dryness Test (Cobalt Bromide) or D-2713-91
Free water content	-	none	none	-	-

(1) An acceptable product shall not yield a persistent oil ring when 0.3 ml of solvent residue mixture is added to a filter paper in 0.1 increments and examined in daylight after 2 minutes as described in ASTM D-2158.

NOTE A: This method may not accurately determine the corrosiveness of the liquefied petroleum gas if the sample contains corrosion inhibitors or other chemicals which diminish the corrosiveness of the same to the copper strip. Therefore, the addition of such compounds for the sole purpose of biasing the test is prohibited.

NOTE B: Be advised that the total sulfur limits in these specifications do include sulfur compounds used for stanching purposes.

Table A. 2. *Liquefied petroleum gas requirements* [35]

Requirement	**Unit**	**Commercial Propane**	**Commercial Butane**	**Commercial P-B Mixture**	**Test Method**
Relative density 15,6 °C/15,6 °C		Report	Report	Report	ASTM D1657 ASTM D2598
Hydrogen sulfide		Negative	Negative	Negative	ASTM D2163 IP 272
Vapor pressure 37,8 °C, max.	kPa	1 434	483	Report [a]	ASTM D1267 ASTM D6897 ASTM D2598
Sulfur content[b], max.	mg/kg	185	185	185	ASTM D6667 IP 272
Observed oil stain		Negative	Negative	Negative	ASTM D2158
Residue by evaporation of 100 mL, max.	mL	0,05	0,05	0,05	ASTM D2158
Evaporation temperature of 95% volatile residue, max.	°C	-38,3	2,2	2,2	ASTM D1837
Butane (C_4) and heavy, max.	%[c]	2,5			ASTM D2163
Pentano (C_5) and heavy, max.	%[c]		2,0	2,0	ASTM D2163
Corrosion to copper foil		No. 1	No. 1	No. 1	ASTM D1838
Moisture content		Negative			ASTM D2713

[a] The vapor pressure must not exceed 1430 kPa, nor must it exceed the value calculated between the observed vapor pressure and the observed relative density using the following relationship: Vapor pressure = 1 167-1 880 x (relative density at 15, 6 °C/15.6 °C).

[b] The sulfur content includes the sulfur compounds of the odorant added to the liquefied petroleum gas.

[c] % corresponds to volume fraction expressed as a percentage.

APPENDIX B:

Cost estimation

Table B. 1. *Setting factors for heat exchanger*

Design type	Fd	Design pressure (psi)	Fp
Reboiler, Kettle	1.35	Less than 150	0.00
Floating heat	1.00	300	0.10
Tube U	0.85	400	0.25
Fixed Tubes	0.80	800	0.52
		1000	0.55

Table B. 2. *Shell/tube material, Fm*

Area, ft^2	AC/AC	AC/Brass	AC/Mo	AC/Al	Al/Al	AC/Monel	Monel/Monel	AC/Ti	Ti/Ti
Low. 100	1.00	1.05	1.60	1.52	2.50	2.00	3.20	4.10	10.28
100 to 500	1.00	1.10	1.75	1.78	3.10	2.30	3.50	5.20	10.60
500 to 1000	1.00	1.15	1.82	2.25	3.26	2.50	3.65	6.15	10.75
1000 to 5000	1.00	1.30	2.15	2.81	3.75	3.10	4.25	8.95	13.05
5000 to 10000	1.00	1.52	2.50	3.52	4.50	3.75	4.95	11.1	16.60

Table B. 3. *Modulus factors*

Unit	Modulus Factor
Process ovens	2,30
Direct fire heaters	2.30
Heat exchangers	3.39
Air coolers	2.54
Vertical containers	4.34
Horizontal containers	3.29
Pumps	3.49
Compressors	3.21

APPENDIX C:

Keyword File (.INP)

Base Case

```
TITLE
  TOLERANCE STREAM =0.001,-0.01,0.001,0.01, TEMPERATURE=-0.001, &
         PRESSURE=0.001, DUTY=0.0001, MISCELLANEOUS=0.001, &
         FLASH=3e-006, MBAL=1
  DIMENSION ENGLISH, TEMP=F, PRES=PSIA, WT=LB, TIME=HR, LENGTH=FT, &
         FLENGTH=IN, LIQVOL=FT3, VAPVOL=FT3, LDENSITY=LB/FT3, &
         VDENSITY=LB/FT3, XDENSITY=API, SPVOL=FT3/LB-MOL, &
         SPVVOL=FT3/LB-MOL, ENERGY=BTU, WORK=HP, DUTY=BTU/HR, &
         CONDUCT=BTU/HR, HTCOEF=BTU/HR, FOUL=HFF/B, VISCOSITY=CP, &
         KVIS=CST, SURFACE=D/CM, STDTEMP(F)=60, &
         STDPRES(PSIA)=14.69594300599, PBASIS(PSIA)=14.69594300599
  SEQUENCE SIMSCI
  CALCULATION TRIALS=50, RECYCLE=ALL, TVPBASIS=100, RVPBASIS=APIN, &
         COMPCHECK=CALC, MAXOPS=1000000, CDATA=FIX, FLASH=DEFAULT, &
         DVARIABLE=ON, PHASE=SIMSCI, TMAX=OLDLIMIT, TMIN=OLDLIMIT
COMPONENT DATA
  LIBID 1,N2/2,CO2/3,C1/4,C2/5,C3/6,IC4/7,NC4/8,IC5/9,NC5/ &
        11,H2O, BANK=SIMSCI,PROCESS
  PETRO 10,C6PLUS,,73,210
  ASSAY FIT=ALTERNATE, CHARACTERIZE=TWU, MW=TWU, CONVERSION=API94, &
         GRAVITY=WATSONK, TBPIP=1, TBPEP=98, NBP=LV, &
         CURVEFIT=CURRENT, KVRECONCILE=TAILS, FORMATION=VER91
THERMODYNAMIC DATA
  METHOD SYSTEM=PR, KVAL(LLE)=SRKK, TRANSPORT=PETR, SET=PR01, &
         DEFAULT
    WATER PROPERTY=SATURATED
  METHOD SYSTEM=GLYC, TRANSPORT=PETR, SET=GLYC01
STREAM DATA
  PROPERTY STREAM=4, TEMPERATURE=120, PRESSURE=565, PHASE=M, &
         RATE(LV)=1203.12, COMPOSITION(M)=1,17.417/2,2.518/3,1.953/ &
        4,3.194/5,40.657/6,8.435/7,19.67/8,6.156, NORMALIZE
  PROPERTY STREAM=54, TEMPERATURE=89.6, PRESSURE=73.48, PHASE=M,  &
        RATE(M)=8193.31, COMPOSITION(M)=11,1, NORMALIZE
  PROPERTY STREAM=56, TEMPERATURE=34.7, PRESSURE=73.961, PHASE=M, &
         RATE(M)=456.08, COMPOSITION(M)=1,0.4/4,0.76/5,96.44/6,1.29/ &
        7,0.96/10,0.15, NORMALIZE
  PROPERTY STREAM=58, TEMPERATURE=-40, PRESSURE=88.99594300599, &
         PHASE=M, RATE(M)=11320.4, COMPOSITION(M)=1,0.4/4,0.76/ &
        5,96.44/6,1.29/7,0.96/10,0.15, NORMALIZE
  PROPERTY STREAM=30, TEMPERATURE=135.0107703234, &
         PRESSURE=286.6086019559, PHASE=M, RATE(M)=2296.663303477, &
         COMPOSITION(M)=3,0.1499241128076/4,0.08498208024197/ &
        5,0.3496290066869/6,0.07102336883796/7,0.2311406010018/ &
        8,0.01398583411617/2,0.08867592778664/ &
        1,0.003701170172265, NORMALIZE
  PROPERTY STREAM=01, TEMPERATURE=120, PRESSURE=565, PHASE=M,  &
        RATE(GV)=1041666.666667, COMPOSITION(M)=1,3.74/2,12.6/ &
        3,45.04/4,7.39/5,22.53/6,2.68/7,5.5/10,0.53, NORMALIZE
  PROPERTY STREAM=AGUA, TEMPERATURE=120, PRESSURE=565, PHASE=M,  &
        RATE(M)=5000, COMPOSITION(M)=11,1, NORMALIZE
  PROPERTY STREAM=60, TEMPERATURE=126.14, PRESSURE=181.677993006, &
         PHASE=M, RATE(M)=1871.67, COMPOSITION(M)=1,0.4/4,0.76/ &
        5,96.44/6,1.29/7,0.96/10,0.15, NORMALIZE
UNIT OPERATIONS
  FLASH UID=HY-01
      FEED 01
      PRODUCT  L=03, V=02
      ADIABATIC
      METHOD SET=PR01
  FLASH UID=HY-02
      FEED AGUA,02
      PRODUCT  V=SAT_GAS, W=FWKO
      ADIABATIC
      METHOD SET=PR01
  MIXER UID=HYM5
      FEED SAT_GAS,03
      PRODUCT  M=1
  CALCULATOR UID=CA1
      DEFINE P(1) AS  STREAM=1, RATE(WT,LB/H), COMP=11,WET
      DEFINE P(2) AS  STREAM=1, RATE(GV,MMFT/H),TOTAL,WET
      DEFINE P(3) AS  STREAM=1, RATE(LBM/H), COMP=11,WET
```

```
      DEFINE P(4) AS  STREAM=SAT_GAS, RATE(GV,MMFT/H),TOTAL,WET
      PROCEDURE
R1=P1/P2
R2=P3
RETURN
  HX   UID=GE-00.01
      HOT  FEED=1, M=2, DP=2.5, METH=PR01
      COLD FEED=54, M=55, DP=2.5, METH=PR01
      CONFIGURE COUNTER
      OPER HTEMP=100.4
  CONTROLLER UID=CN1
      SPEC STREAM=55, TEMPERATURE(F), VALUE=100.4
      VARY STREAM=54, RATE(LBM/H)
      CPARAMETER IPRINT, SOLVE
  HX   UID=GE-00.02
      HOT  FEED=2, M=3, DP=2.5, METH=PR01
      COLD FEED=56, M=57, DP=2.5, METH=PR01
      CONFIGURE COUNTER
      OPER HTEMP=80
  CONTROLLER UID=CN2
      SPEC STREAM=57, TEMPERATURE(F), VALUE=36.14
      VARY STREAM=56, RATE(LBM/H)
      CPARAMETER IPRINT, SOLVE
  HX   UID=GE-15.03
      HOT  FEED=27, M=28, DP=2.5
      COLD FEED=58, M=59, DP=2.5
      CONFIGURE COUNTER
      OPER HTEMP=-40
  CONTROLLER UID=CN3
      SPEC STREAM=59, TEMPERATURE(F), VALUE=-32.8
      VARY STREAM=58, RATE(LBM/H)
      CPARAMETER IPRINT, CONTINUE
  FLASH UID=GV-16.07
      FEED 28
      PRODUCT  V=46, L=29
      ADIABATIC
  FLASH UID=GV-16.01
      FEED 5,3
      PRODUCT  V=6, L=9, W=50
      ADIABATIC
      METHOD SET=PR01
  SPLITTER UID=GL-21.01
      FEED 6
      PRODUCT  M=53, M=7
      OPERATION OPTION=FILL
      SPEC STREAM=53, RATE(LBM/H), COMP=11,WET, DIVIDE, STREAM=6, &
                 RATE(LBM/H), COMP=11,WET, VALUE=0.0001
      METHOD SET=PR01
  HX   UID=GV-16.02_03C
      HOT  FEED=7, M=04
      OPER HTEMP=-50
  FLASH UID=GV-16.02_03F
      FEED 04
      PRODUCT  W=07, V=05, L=06
      ADIABATIC
  HX   UID=GV-16.02_03H
      COLD FEED=05,06, M=8
      DEFINE CTEM(F) AS  STREAM=7, TEMPERATURE(F)
  SPLITTER UID=SP2
      FEED 8
      PRODUCT  M=21, M=22
      OPERATION OPTION=FILL
      SPEC STREAM=21, RATE(LBM/H),TOTAL,WET, DIVIDE, STREAM=8, &
                 RATE(LBM/H),TOTAL,WET, VALUE=0.32
  SPLITTER UID=SP1
      FEED 9
      PRODUCT  M=10, M=14
      OPERATION OPTION=FILL
      SPEC STREAM=10, RATE(LV,GAL/M),TOTAL,WET, VALUE=50
  PUMP UID=GP-17.01-02
      FEED 10
      PRODUCT  M=11
      OPERATION DP=35.5584
  FLASH UID=GV-16.22
      FEED 11
      PRODUCT  L=12, W=51
      ADIABATIC
```

```
    METHOD SET=PR01
HX   UID=GV-16.05_06C
    HOT  FEED=12, M=08
    OPER HTEMP=-50
FLASH UID=GV-16.05_06F
    FEED 08
    PRODUCT  W=011, L=010, V=09
    ADIABATIC
HX   UID=GV-16.05_06H
    COLD FEED=010,09, M=13
    DEFINE CTEM(F) AS  STREAM=12, TEMPERATURE(F)
PUMP UID=GP-17.13-14
    FEED 14
    PRODUCT  M=15
    OPERATION DP=35.5584
FLASH UID=GV-16.23
    FEED 15
    PRODUCT  L=16, W=52
    ADIABATIC
HX   UID=GV-16.26_27C
    HOT  FEED=16, M=012
    OPER HTEMP=-50
FLASH UID=GV-16.26_27F
    FEED 012
    PRODUCT  W=015, L=014, V=013
    ADIABATIC
HX   UID=GV-16.26_27H
    COLD FEED=014,013, M=17
    DEFINE CTEM(F) AS  STREAM=16, TEMPERATURE(F)
MIXER UID=M1
    FEED 13,17
    PRODUCT  M=18
SPLITTER UID=SP3
    FEED 18
    PRODUCT  M=19, M=20
    OPERATION OPTION=FILL
    SPEC STREAM=19, RATE(LBM/H),TOTAL,WET, DIVIDE, STREAM=18, &
               RATE(LBM/H),TOTAL,WET, VALUE=0.32
MIXER UID=M2
    FEED 21,19
    PRODUCT  M=23
MIXER UID=M3
    FEED 22,20
    PRODUCT  M=25
HX   UID=GE-15.02
    HOT  FEED=25, M=26, DP=2.5
    COLD FEED=29, M=30, DP=2.5
    CONFIGURE COUNTER
    OPER HIHO=105.6749
HX   UID=GE-15.11
    HOT  FEED=60, M=61, DP=2.5
    COLD FEED=30, M=31, DP=2.5
    CONFIGURE COUNTER
    OPER CTEMP=90
CONTROLLER UID=CN4
    SPEC STREAM=61, TEMPERATURE(F), VALUE=97.88
    VARY STREAM=60, RATE(LBM/H)
    CPARAMETER IPRINT, CONTINUE
COLUMN UID=GV-16.09
    PARAMETER TRAY=18,IO
    FEED 31,10,TNOTSEPARATE, NOTSEPARATE
    PRODUCT  BTMS(M)=32, OVHD(M)=45, SUPERSEDE=ON
    CONDENSER TYPE=PART
    DUTY  1,1,,CONDENSER
    DUTY  2,18,,REBOILER
    PSPEC DPCOLUMN=5
    PRINT PROPTABLE=PART, DIAGRAM
    ESTIMATE MODEL=CHEM, TTEMP=-2, RTEMP=235
    TEMPERATURE 2,-2/18,235
    SPEC ID=COL1SPEC2, RRATIO, VALUE=1.4696
    SPEC ID=COL1SPEC1, STREAM=32, RATE(LV,GAL/M),TOTAL,WET, DIVIDE, &
               STREAM=31, RATE(LV,GAL/M),TOTAL,WET, VALUE=0.7608
    DEFINE PRODUCT(LBM/H)=45 AS  STREAM=31, RATE(LBM/H), COMP=1,4, &
              WET
    DEFINE PTOP(PSIA) AS  STREAM=31, PRESSURE(PSIA), MINUS,0.5
    VARY DNAME=REBOILER,CONDENSER
    REBOILER TYPE=KETTLE
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      METHOD SET=GLYC01
  MIXER UID=M6
      FEED 46,45
      PRODUCT  M=47
  HX   UID=GE-15.01
      HOT  FEED=23, M=24, DP=2.5
      COLD FEED=47, M=48, DP=2.5
      CONFIGURE COUNTER
      OPER HIHO=100.6749
  HX   UID=GE-00.03
      HOT  FEED=4, M=5, DP=2.5, METH=PR01
      COLD FEED=48, M=49, DP=2.5, METH=PR01
      CONFIGURE COUNTER
      OPER HIHO=40
  MIXER UID=M4
      FEED 24,26
      PRODUCT  M=27
  SPLITTER UID=SP4
      FEED 32
      PRODUCT  M=33, M=37
      OPERATION OPTION=FILL
      SPEC STREAM=33, RATE(LBM/H),TOTAL,WET, DIVIDE, STREAM=32, &
                 RATE(LBM/H),TOTAL,WET, VALUE=0.5
  VALVE UID=V2
      FEED 37
      PRODUCT  M=38
      OPERATION PRESSURE=215
  COLUMN UID=GV-16.11
      PARAMETER TRAY=18,IO
      FEED 38,10,TNOTSEPARATE, NOTSEPARATE
      PRODUCT  OVHD(M)=39, BTMS(M)=43, SUPERSEDE=ON
      CONDENSER TYPE=BUBB
      DUTY  1,1,,CONDENSER
      DUTY  2,18,,REBOILER
      PSPEC DPCOLUMN=5
      PRINT PROPTABLE=PART
      ESTIMATE MODEL=CHEM
      SPEC ID=GV-16.SPEC1, RRATIO, VALUE=1.5055
      SPEC ID=GV-16.SPEC2, STREAM=39, RATE(LV,GAL/M),TOTAL,WET, &
                 DIVIDE, STREAM=38, RATE(LV,GAL/M),TOTAL,WET, &
                 VALUE=0.5677
      DEFINE PRODUCT(LBM/H)=39 AS  STREAM=38, RATE(LBM/H), COMP=1,5, &
                WET
      DEFINE PTOP(PSIA) AS  STREAM=38, PRESSURE(PSIA), MINUS,0.5
      VARY DNAME=CONDENSER,REBOILER
      REBOILER TYPE=KETTLE
      METHOD SET=GLYC01
  VALVE UID=V1
      FEED 33
      PRODUCT  M=34
      OPERATION PRESSURE=215
  COLUMN UID=GV-16.24
      PARAMETER TRAY=18,IO
      FEED 34,10,TNOTSEPARATE, NOTSEPARATE
      PRODUCT  OVHD(M)=35, BTMS(M)=41, SUPERSEDE=ON
      CONDENSER TYPE=BUBB
      DUTY  1,1,,CONDENSER
      DUTY  2,18,,REBOILER
      PSPEC DPCOLUMN=5
      PRINT PROPTABLE=PART
      ESTIMATE MODEL=CHEM
      SPEC ID=COL2SPEC1, RRATIO, VALUE=1.5055
      SPEC ID=COL2SPEC2, STREAM=35, RATE(LV,GAL/M),TOTAL,WET, DIVIDE, &
                 STREAM=34, RATE(LV,GAL/M),TOTAL,WET, VALUE=0.5677
      DEFINE PRODUCT(LBM/H)=35 AS  STREAM=34, RATE(LBM/H), COMP=1,5, &
                WET
      DEFINE PTOP(PSIA) AS  STREAM=34, PRESSURE(PSIA), MINUS,0.5
      VARY DNAME=CONDENSER,REBOILER
      REBOILER TYPE=KETTLE
      METHOD SET=GLYC01
  CALCULATOR UID=CA2
      DEFINE P(1) AS  STREAM=6, RATE(WT,LB/H), COMP=11,WET
      DEFINE P(2) AS  STREAM=6, RATE(GV,MMFT/H),TOTAL,WET
      DEFINE P(3) AS  STREAM=8, RATE(WT,LB/H), COMP=11,WET
      DEFINE P(4) AS  STREAM=8, RATE(GV,MMFT/H),TOTAL,WET
      PROCEDURE
R1=P1/P2
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R2=P3/P4
RETURN
  CALCULATOR UID=CA8
      DEFINE P(1) AS  STREAM=35, RATE(LV,GAL/M),TOTAL,WET, DIVIDE, &
                 STREAM=34, RATE(LV,GAL/M),TOTAL,WET
      DEFINE P(2) AS  COLUMN=GV-16.24, RRATIO
      PROCEDURE
R1=P1
R2=P2
RETURN
  CALCULATOR UID=CA4
      DEFINE P(1) AS  STREAM=46, RATE(GV,MMFT/D),TOTAL,WET, PLUS, &
                 STREAM=45, RATE(GV,MMFT/D),TOTAL,WET
      DEFINE P(2) AS  STREAM=45, RATE(GV,MMFT/D),TOTAL,WET
      DEFINE P(3) AS  STREAM=46, RATE(GV,MMFT/D),TOTAL,WET
      PROCEDURE
R1=P1
R2=P2
R3=P3
RETURN
  CALCULATOR UID=CA3
      DEFINE P(1) AS  STREAM=12, RATE(LBM/H), COMP=11,WET, MINUS, &
                 STREAM=13, RATE(LBM/H), COMP=11,WET
      DEFINE P(2) AS  STREAM=12, RATE(LBM/H), COMP=11,WET
      DEFINE P(3) AS  STREAM=011, RATE(LBM/H), COMP=3,10,WET, DIVIDE, &
                 STREAM=12, RATE(LBM/H), COMP=3,10,WET
      PROCEDURE
R1=P1/P2*100
R2=P3*100
RETURN
  HX   UID=GE-19.09
      HOT  FEED=35, M=36, DP=2.5
      UTILITY WATER, TIN=89.6, TEMPERATURE=100.4
      CONFIGURE COUNTER
      OPER HTEMP=100
  HX   UID=GE-19.08
      HOT  FEED=39, M=40, DP=2.5
      UTILITY WATER, TIN=89.6, TEMPERATURE=100.4
      CONFIGURE COUNTER
      OPER HTEMP=100
  HX   UID=GA-19.05
      HOT  FEED=41, M=42, DP=2.5
      UTILITY AIR, TIN=71.6, TEMPERATURE=100, CPMETHOD=IDEA
      CONFIGURE COUNTER
      OPER HTEMP=110
  CALCULATOR UID=CA9
      DEFINE P(1) AS  HX=GE-15.02, HIHO(F)
      PROCEDURE
R1=P1
RETURN
  HX   UID=GA-19.04
      HOT  FEED=43, M=44, DP=2.5
      UTILITY AIR, TIN=71.6, TEMPERATURE=100, CPMETHOD=IDEA
      CONFIGURE COUNTER
      OPER HTEMP=110
  CALCULATOR UID=CA5
      DEFINE P(1) AS  STREAM=36, RATE(WT,TM/D),TOTAL,WET, PLUS, &
                 STREAM=40, RATE(WT,TM/D),TOTAL,WET
      DEFINE P(2) AS  STREAM=42, RATE(LV,GAL/M),TOTAL,WET, PLUS, &
                 STREAM=44, RATE(LV,GAL/M),TOTAL,WET
      DEFINE P(3) AS  STREAM=40, SPGR,WET
      DEFINE P(4) AS  STREAM=40, RVP
      DEFINE P(5) AS  STREAM=40,PCT(LV), COMP=6,10,WET
      DEFINE P(6) AS  STREAM=40,PCT(LV), COMP=8,10,WET
      DEFINE P(7) AS  STREAM=1, RATE(LBM/D),TOTAL,WET
      DEFINE P(8) AS  STREAM=4, RATE(LBM/D),TOTAL,WET
      PROCEDURE
R1=P1
R2=P2
R3=P3
R4=P4*6.894757
R5=P5
R6=P6
R7=P7*P1/(P7+P8)
R8=P8*P1/(P8+P7)
RETURN
  CALCULATOR UID=CA6
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      DEFINE P(1) AS  STREAM=1, RATE(LV,GAL/H), COMP=5,10,WET, PLUS, &
                 STREAM=4, RATE(LV,GAL/H), COMP=5,10,WET
      DEFINE P(2) AS  STREAM=1, RATE(GV,MFT3/H),TOTAL,WET, PLUS, &
                 STREAM=4, RATE(GV,MFT3/H),TOTAL,WET
      DEFINE P(3) AS  STREAM=36, RATE(LV,GAL/D),TOTAL,WET, PLUS, &
                 STREAM=40, RATE(LV,GAL/D),TOTAL,WET
      DEFINE P(4) AS  STREAM=36, RATE(WT,TM/D),TOTAL,WET, PLUS, &
                 STREAM=40, RATE(WT,TM/D),TOTAL,WET
      PROCEDURE
R1=P1/P2
R2=P3/P1*100
R3=P4
R4=P1
RETURN
  CALCULATOR UID=CA10
      DEFINE P(1) AS  STREAM=28, CP(BTU/LB-F)
      DEFINE P(2) AS  STREAM=28, RATE(WT,LB/H),TOTAL,WET
      DEFINE P(3) AS  STREAM=28, RATE(LV,GAL/S),TOTAL,WET
      PROCEDURE
R1=P1
R2=P2
R3=P3
RETURN
  CALCULATOR UID=CA7
      DEFINE P(1) AS  STREAM=10, RATE(LV,GAL/M),TOTAL,WET, DIVIDE, &
                 STREAM=9, RATE(LV,GAL/M),TOTAL,WET
      PROCEDURE
R1=P1
RETURN
END
```