THE ECONOMICS OF NATURAL GAS LIQUIDS (NGLS)

FRACTIONATIONA CASE STUDY OF THE GAS PROCESSING PLANT OF

THE GHANA

NATIONAL GAS COMPANY.

By

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DECLARATION

I hereby declare that except for reference to other people's work which I have dully cited, this project submitted to the school of Graduate Studies, Kwame Nkrumah University of Science and Technology, Kumasi is the result of my own investigation, and have not been presented for any other degree elsewhere.

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ABSTRACT

This project work was done to ascertain the economic viability of separating NGLs into individual components as oppose to mix product such as LPG using the Ghana National Gas Company Processing Plant as a case study. The process flow diagram of the gas processing plant was reviewed to incorporate a depropanizer to recover propane and a butane splitter to recover isobutane and normal butane. The depropanizer and the butane splitter were sized using the procedure recommended by the Gas Producers and Suppliers Association and cost- benefit analysis conducted using Net Present Value approach the results of which showed that the project was viable. It was recommended that the material balance and the design parameters obtained should be confirmed by the use of a process modeling software and sensitivity analysis be conducted on the cash flow.

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LIST OF ABREVIATIONS

BSCFD   Billion Standard Cubic Feet Per Day

GNGC   Ghana National Gas Company

GPSA   Gas Processors & Suppliers Association

Lbs.   Pounds

LPG   Liquefied Petroleum Gas

MMSCFD   Million Standard Cubic Feet Per Day

NGLs   Natural Gas Liquids

OF

NPV   Net Present Value

Psia   Pounds force Square Inch Absolute

TCF   Trillion Standard Cubic Feet

CHAPTER 1

1.0   INTRODUCTION

Gas in a reservoir is normally at high pressure say 27579 kPa at 66 $^{\circ}$C. It is brought to the surface via tubing in the well to the well head. At the well head, the pressure is reduced ("choking") to be consistent with the pressure of the gathering pipeline (typically 9929 kPa or less). The drop in pressure from 27579 kPa to 9929 kPa is accompanied by a drop in temperature of the gas (Joule-Thompson effect). This cooling of the gas could result in condensation of some of the heavy hydrocarbon and water known as Hydrate. To prevent hydrate formation (hydrate control), the gas can be

a)   Heated above the hydrate formation temperature

b)   Injected with inhibitors (e.g. glycol or methanol) to lower the hydrate formation temperature

c)   dehydrated–water removal using for example Triethylene glycol (TEG) in a column

Gas from the well-head is normally transported as a 2-phase mixture to the separation plant. Most of the liquids are separated from the gas here. The acidic components of the gas stream are removed before further processing is done at this stage. The process of removal of the acidic component (specifically H_2S) of the gas is called sweetening. The acid component of the gas stream maybe removed with an amine column, using, for example, Monoethanol Amine (MEA) or Diethanolamine (DEA). The acid

gases (H_2S and CO_2) from the sweetening plant are sent to a sulphur recovery plant where the H_2S is reacted with O_2 to produce elemental sulphur. The residual SO_2 and CO_2 are rejected to the atmosphere depending on environmental regulations.

The gas from the sweetening plant may still contain some water and heavy hydrocarbons. These must be reduced such that the gas stays above both the dew-points of water and the hydrocarbons. This can be accomplished in several ways:

1. Refrigeration (e.g. by expansion) below the dew-point of the heavy hydrocarbon and water; glycol may be required to prevent hydrate formation

2. Water: dehydration / HC: absorption column

Hydrocarbon Liquids are recovered from both the separators and the dew-point controllers. The Liquids usually referred to as natural gas liquids (NGLs) may be sent to a fractionating column for separation of the individual components. This is essentially by distillation, utilizing the different boiling points of the components.

Natural Gas Liquids (NGLs) refers to liquids recovered from natural gas including ethane, Liquefied Petroleum Gas (LPG) and heavier hydrocarbons to meet specified contractual and operating requirements.

Natural gas liquids (NGLs) once separated from a natural gas stream are fractionated into individual component fractions. Fractionation involves separating components by relative volatility (α). The difficulty of a separation is directly related to the relative volatility of the components and the required purity of the product streams.

[Association of Gas Producer and Suppliers, 2004]

A gas processing plant requires at least one fractionating tower to produce liquids. The number of fractionators depends on the number of products required. The first is a deethanizer, which separates ethane from the NGL stream. Second, the depropanizer separates the propane, followed by a debutanizer which boils off the butane, and some cases a butane splitter separates normal butane and isobutene leaving natural gasoline.

The Ghana national Gas Company separates NGLs into ethane, Liquefied Petroleum Gas (LPG) thus a mixture of propane and butane, pentane and condensate normally referred to as the Y-grade. The company has plans of installing a turbo expander to increase its NGL recovery but analysis conducted by the company has shown that this will increase the amount of propane in the LPG beyond specification.

This means a depropanizer will have to be installed to recover the excess propane, therefore, the need to investigate the economics of separating the NGL into the pure component fractions of ethane, propane, normal butane and isobutene, pentane and heavier.

1.1   PROJECT OBJECTIVES

The goal of this project is to ascertain the economic viability of separating NGLs into the individual fractions of ethane, propane, normal butane, isobutene and pentane and heavier which will require the installation of a depropanizer and butane splitter as opposed to ethane, LPG (mixture of propane and butane), pentane and condensate.

Specific Objectives are:

1. Design a new process flow diagram of the Ghana national gas plant which will incorporate a depropanizer to recover propane and a butane splitter to recover isobutane and normal butane

2. Determine the sizes(columns diameter, height and number of trays) of the depropanizer and the butane splitter required to handle the NGL flow

3. Determine the operating pressure and temperature of the above fractionators required to meet the product specifications

4. Conduct a cost- benefit analysis to ascertain the economic viability of this project.

1.2   PROJECT OUTLINE

The first chapter is the introduction which gives a background and the problem to be investigated. Chapter two covers literature review of the oil and gas sector of Ghana. Chapter three explains procedures and approach used to achieve the project objectives. The details given explain how the design model and procedures used produced the results presented in Chapter four. Chapter five provides conclusions and recommendations.

CHAPTER 2

2.0   LITERATURE REVIEW

Ghana sold licenses for offshore oil exploration and production to different international companies in 2004 and by July 2007, Tullow Oil and Kosmos Energy discovered oil in commercial quantities in the Western Region of Ghana. The oil field was named the "Jubilee Field". Commercial production from the site started right in December 2010 by Tullow which is the main operator of the field.

2.1   GENERAL OVERVIEW OF JUBILEE OIL FIELD

The Jubilee Field is operated by Tullow oil plc. The recoverable reserves of the Jubilee oil field are around 600 million barrels with an upside potential of 1.8 billion barrels. In addition there are significant gas resources of approximately 800 BSCF (23 BSCM). Present production rate as at January 2014 is around 100 000 barrels of oil per day (15 900 m3/day) and 140 MSCF (4 MSCM) of natural gas per day.

2.2   THE GHANA NATIONAL GAS COMPANY

Ghana National Gas Company was incorporated as a limited liability company in July 2011 with the responsibility to build, own and operate infrastructure required for the gathering, processing, transporting and marketing of natural gas resources in the country.

The gas plant treats and processes the raw gas to meet the specifications for sales gas and to recover economic quantities of associated LPG and condensate. A schematic diagram of the plant is shown in figure 2.2. The plant capacity is 150 MMSCFD with a turn down ratio of 30% of nominal capacity and a life time of 25 years. A 12", 58 km long subsea pipeline transports the gas from the Jubilee FPSO to the reception facilities at the Gas Plant at Atuabo. The gas arriving at the plant is metered at the inlet facilities and

routed to a separation unit to separate any entrained water and condensed hydrocarbons from the gas stream, producing a water stream, a liquid hydrocarbon stream and a gas stream. The separated gas is filtered, and then chilled to condense the heavy hydrocarbons in the gas. The gas and liquefied heavy hydrocarbon streams are separated; gas is routed to the export gas header for gas export via the Atuabo-Takoradi pipeline. The liquid hydrocarbons are stripped of the remaining ethane and methane, and then fractionated into LPG, pentane and condensate streams. The pentane flows to the fuel gas system while the LPG and the condensate streams are cooled and stored.
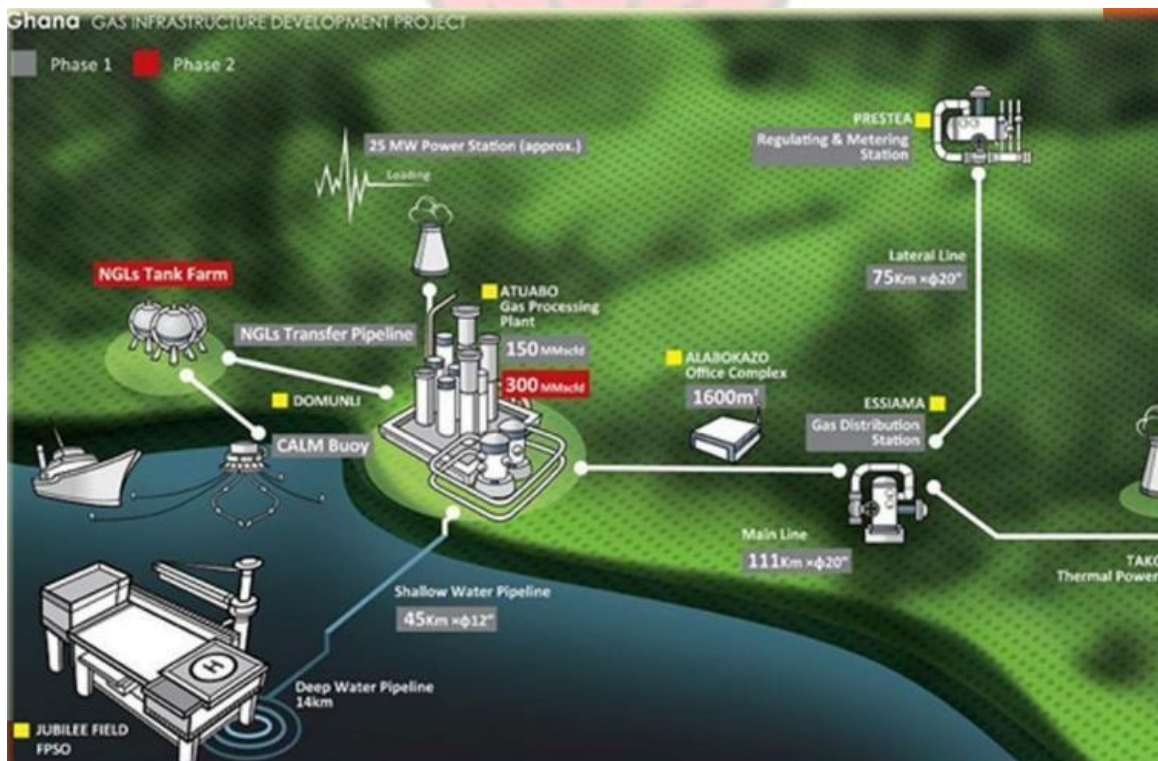
Figure 2.1   The Western corridor Gas Infrastructure Development project (Source: Ghana National Gas Company's brochure

CHAPTER 3

3.0   METHODS AND PROCEDURES

This chapter reviews the process flow diagram of the Ghana National Gas plant and incorporates a depropanizer and a butane splitter in order to separate the natural gas liquids into an individual component of ethane, propane, butane, isobutene, pentane and condensate. It also focuses on the detailed procedures and the information used to determine the size (diameter, height and number of trays), operating pressure, temperature and a cost-benefit analysis of the additional fractionating columns.

3.1   PROCESS FLOW OF THE GHANA NATIONAL GAS PLANT

The Ghana national gas company plant at Atuabo currently receives raw natural gas through a 14 km x 10'' (25.4 cm) deep water and 44 km x 12'' (30.48 cm) pipeline from the Jubilee oil field at a normial flow of 120 million standard cubic feet per day (3397990 m^3)) during the phase one of the gas infrastructure project. The second phase will process about 300 million standard cubic feet per day (8494974 m^3) of raw gas from Jubilee, Tweneboa and Odum oil fields combined. A process flow diagram of Ghana national gas company plant presented in figure 2.1 shows the major units and equipment in the process area.

Ghana Gas Processing Plant Process Flow Diagram

Figure 3.1 Process flow diagram of Ghana gas processing plant (Source; Ghana National gas company)

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The raw gas on arrival at the processing plant feeds two inlet separators to condense water vapour present in the gas and the water is collected in the boot of the separator and sent to the closed drain. The heavy hydrocarbon liquids called natural gas liquids (NGLs) are separated from the gas for fractionation.

The gas from the inlet separators is filtered and split through a Gas/Gas and a Gas/Liquid Exchangers respectively to recombine at a temperature of -10 ºC. The gas then flows through a Joule Thompson valve to drop the pressure 51 Barg (5201.325 kPa) and temperature to -32 ºC. Ethylene glycol solution is injected to prevent hydrate formation. The stream then flows to a Low Temperature Separator to remove hydrocarbon liquids which passes a gas/liquids exchanger and then to a Deethanizer. The gas stream from the low temperature separator exchanges heat with the incoming gas in the gas/gas exchanger and is warmed to 13 ºC before being sent to the Export gas header.

The NGLs recovered at the separators feed deethanizer to recover ethane as overhead product. The bottom product goes through another column called the debutanizer which recovers liquefied petroleum gas (LPG) as its overhead product mostly a mixture of propane and butane. The bottom product of the debutanizer which is composed mainly of pentane and heavier is directed to a pentane stripper to recover pentane from the stream. The heavier hydrocarbons remained are stored as condensates which have economic value as well.

2.2 DESIGNING THE FRACTIONATORS (DEPROPANIZER AND BUTANE SPLITTER)

The design of fractionators is based on the concept of equilibrium stages. Tray efficiency or stage efficiency is a correction factor for non-ideality. Generally, the separation is centered on two components: the light key and the heavy key. The phase behavior of this binary pair is important in understanding the science of fractionation. Thus, even though, we are dealing with a multicomponent mixture, the actual separation is between the light and heavy keys [GPSA 2004, section19].

In order to design the depropanizer and the butane splitter, the following method as recommended by GPSA was adopted.

1. Establish feed composition, flow rate, temperature and pressure; the composition of the raw gas at 120 – 150 million standard cubic feet per day (3397990 to 4247487 m^3) from the jubilee field was used to estimate the feed composition and flow rate for the depropanizer and the butane splitter by material balance approach.

2. Determine the product splits of the columns and establish condenser temperature and column pressure; the product split for both fractionators were determined based on the assumption that both columns are to obtain 90-95% of light key component (based on feed) in the overhead stream and 1% of the heavy key component (based on distillate) in the overhead stream. And also all (not true in practice) the components above the light key will go to overhead stream and all components below the heavy key will go to the bottom stream.

The condenser temperature thus the cooling medium temperature (raw water) for bubble point condition of the overhead product was assumed at 30-70 ^{0}c. The column pressure was

determined using the condenser temperature. The column pressure was used to calculate the

reboiler temperature at bubble point.

3. Minimum theoretical stages from Fenske equation as below;

$$S_m = \frac{\text{Log } S_f}{\text{Log}(\alpha avg)}$$

Equation 3.1

$$S_f = \left[\frac{X_D}{X_B}\right]_{LK} \cdot \left[\frac{X_B}{X_D}\right]_{HK}$$

Equation 3.2

$$\alpha_{avg} = \frac{\alpha_{top} + \alpha_{bottom}}{2}$$

Equation 3.3

$$\alpha = \frac{K_{LK}}{K_{HK}}$$

Equation 3.4

If volatility varies widely, a modified volatility β_{ij} is used where

$$\beta ij = \frac{K_{LK}}{((K_{HK})^b}$$

Equation 3.5

And Equation 2.1 becomes

$$S_m = \frac{\log\left[\left[\frac{X_D}{X_B}\right]_{LK} \cdot \left[\frac{X_B}{X_D}\right]_{HK} \left[\frac{B}{D}\right]^{1-b}\right]}{\log \beta_{ij}}$$

Equation 3.6

S_m = minimum theoretical stages

α_{avg} = average relative volatility of top and bottom products

S_f = separation factor

α = relative volatility of the key component

X_B = mole fraction of the bottom

Lk= light key component

Hk = heavy key component

X_D = mole fraction of the distillate

B = total moles of bottom D

= total moles of distillate

b = exponent from K-value plots

4. Minimum reflux ratio from the Underwood equations as shown below;

$$1 - q = \sum_{n=1}^{n} \left(\alpha_i \frac{X_{Fi}}{(\alpha_i - \emptyset)} \right)$$
Equation 3.7

$$(R_m + 1) = \sum_{n=1}^{n} \left(\alpha_i \frac{X_{Di}}{(\alpha_i - \emptyset)} \right)$$
Equation 3.8

Where q = thermal condition of feed (q=1 for boiling point feed, q=0 for dew point feed and 0< q < 1.0 for multiphase feed)

X_{Fi} =mole fraction of component i in the feed

X_{Di} = mole fraction of component i in the distillate

R_m = minimum reflux ratio $\alpha_{i\,=}$ relative volatility of component i

ϕ= constant obtained by trial and error

5. Obtain theoretical stages to operating reflux ratio from Erbar-Maddox Correlation of Stages vs. Reflux shown in Figure 2.2.
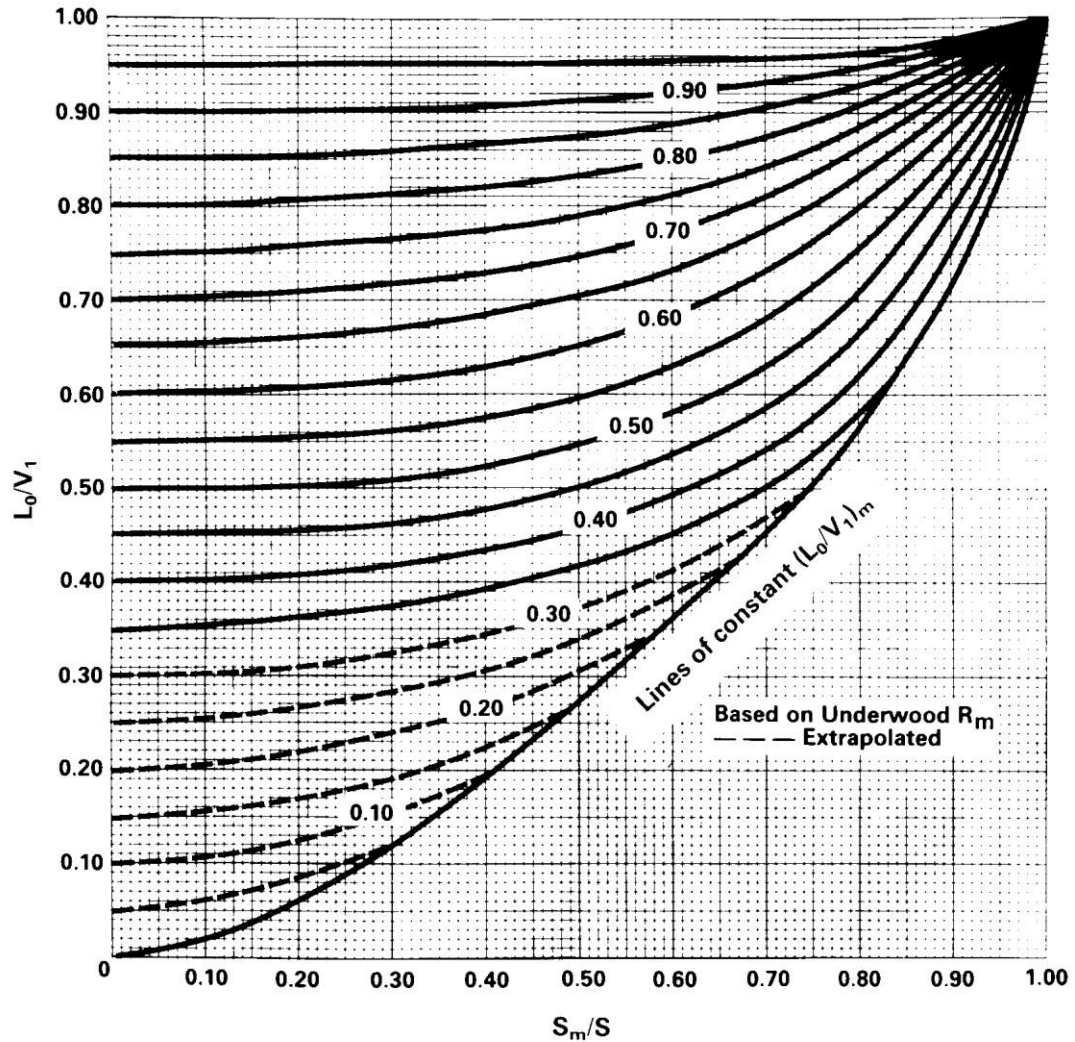
Figure 3.2 Erbar-Maddox Correlation of Stages vs. Reflux (GPSA, 2004)

This correlation relates the ratio of minimum stages to theoretical stages (S_m/S) to the minimum reflux ratio (Rm) and the operating reflux ratio (R). The actual number of trays (S) is then calculated from (S_m/S).

$$(L_o/V_i) = \frac{R}{R+1}$$ Equation 3.9

$$(L_o/V_i)_m = \frac{R_m}{R_m+1}$$
$$\text{Equation 3.10}$$

6. Determine the feed point location; this was estimated using the empirical equation given by Kirkbride (1944) shown below.

$$\log\left(\frac{N_r}{N_s}\right) = 0.206\log\left[\left(\frac{B}{D}\right)\left(\frac{X_{f,HK}}{X_{f,LK}}\right)\left(\frac{X_{b,LK}}{X_{d,HK}}\right)^2\right]$$
$$\text{Equation 3.11}$$

Where N_r = number stages above the feed, including any partial condenser

N_s = number stages below the feed, including the reboiler

B = molar flow of the bottom product

D = molar flow of the top product

X_f, HK= mole fraction of the heavy key component in the feed

X_f, LK = mole fraction of the light key component in the feed

X_d, HK= mole fraction of the heavy key component in the top product

X_b, LK = mole fraction of the light key component in the bottom product

7. Determine the column diameter; the column diameter required to handle the vapour load was found using Souders and Brown equations as shown below.

$$V_{max} = C\sqrt{\frac{\rho_l - \rho_v}{\rho_v}}$$ Equation
3.12 $$D = \sqrt{\frac{Q_v}{V_{max} \times 07854}}$$

Equation 3.13

V_{max} = maximum vapour velocity, feet per seconds (ft/s)

ρ_l = Liquid density, pounds per cubic feet (lbs/ft^3)

ρ_v = vapour density, pounds per cubic feet (lbs/ft^3)

C = constant

Q_v = vapour volumetric flow rate, cubic feet per seconds (ft^3/s)

D= diameter of column, feet (ft)

8. Height of the column; the tower height can be related to the number of trays in the column. Equation 3.14 assumes that a spacing of two feet between trays will be sufficient including additional five to ten feet at both ends of the tower. This includes a fifteen percent excess allowance of space (Douglas, 1988).

H $_{tower}$ = 2.3 N$_{actual}$ (ft.)   Equation 3.14

9. Determine the condenser and reboiler loads;

Assuming 15% heat loss by the reboilers:

a. Condenser duty

$$Q_c = D(R + 1)\Delta H_{vap}$$ Equation 3.15

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b. Cooling air flow rate for condenser;

$$m_{cw} = \frac{Q_c}{C_{pH_2o}(T_{out}-T_{in})}$$

Equation 3.16 c.

Reboiler duty

$$Q_c = (RD + F)[\Delta H_{vap} + Cp * \Delta T]$$

Equation 3.17

d. Hot oil flow rate for the reboiler;

$$m_{oil} = \frac{Q_R}{Cp\Delta T}$$

Equation 3.18

Where;

ΔH = heat of vaporization

V (vapour flow rate in the column) = (R + 1) D

L (liquid flow rate in top section = R*D

D = distillate flow

R = reflux ratio

10.   Thickness of the columns; the thickness of the columns were determined using the equations below.

$$t = t_h + \frac{1}{8}$$ For $t_h > t_a$ Equation 3.19

$$t = 0.5(t_h + t_a) + \frac{1}{8}$$ For $t_h \le t_a$ Equation 3.20

Where;

$$t_h = \frac{P.D}{2479-0.1P}$$ Equation 3.21

$$t_a = \frac{90X10^{-6}(H+S)^2}{D}$$ Equation 3.22

t_h = thickness required for hoop stress in inches

t_a = thickness required for wind velocity in inches

P = design pressure in psig

D = column diameter in feet

H = height of column in feet

S = height of skirt in feet

Insulation for column is assumed to be one inch thick

3.3   COST- BENEFIT ANALYSES

The economic tools used for the cost-benefit analysis were:

a) The Net Present value (NPV) approach which compared the present values of future cash flows expected from the additional product streams to the initial cash outlay for installing the additional fractionating columns. The NPV approach is based on the discounted cash flow formula.

$$PV = \frac{FV}{(1+i)^n}$$ Equation 3.23

PV = present value

FV = future value

i = discounted rate

n = year

b) The Net Cash flow approach which refers to the difference between the cash inflow and the cash outflow.

3.3.1 COST/EXPENSE

a. Capital cost of the equipment; a quick order of magnitude cost was arrived based on chemical engineering construction index 324.5 and quotes from vendors.

$$Vessels\ (\$) = 2(L + 2.2D)(100t + 45D + 5D)$$

Equation 3.24

$$Tray\ (\$/tray) = 170D(0.5 + \frac{10}{N})$$

Equation 3.25

$$Heat\ exchangers\ (\$) = 5000 + 23A$$

Equation 3.26

$$Pumps\ (\$) = 4000 + 50(GPM)^{1/2} + 3h$$

Equation 3.27

$$Motors\ (\$) = 55(Hp) + 220$$

Equation 3.28

$$Motors\ starters(\$) = 30(Hp) + 750$$

Equation 3.29

$$In\ plant\ pipelines\ (\$/ft) = 2d + 3$$

Equation 3.30

$$Control\ systems\ (\$/loop) = 5600 + CL + (1600CF)Dv$$

Equation 3.31

$$Vessel\ insu.\ (\$) = (24 + 8i)\left(D + \frac{i}{6}\right)H + (50 + 14i)\left(D + \frac{i}{6}\right)^2 + 57DS$$

$$Pipe\ insulations\ (\$/ft) = 6 + (0.35 + 0.725i)(d + 3)$$

Equation 3.32

Where:

L = length of vessel feet

T = thickness in inches

D = diameter of vessel in feet N

= number of trays

i = insulation thickness in inches

d = diameter of pipe in inches

P = pressure in psig = Density x gravity x head H

= height of column in feet

h = head of pump in feet

CL = level control CF

= flow control

dv = diameter of motor valve

HP= hydraulic power = head x flow x Specific gravity

A = area in square feet = $Q/ (U\Delta T_{lim})$

Q = duty of condenser/reboiler

U = overall heat transfer coefficient in Btu/ft^2 ^{0}F

ΔT_{lim} = log mean temperature

b.   Operating cost:

        i. Hot oil ii. Electricity - reflux and bottom pumps' power and energy for
        heating the hot oil iii. Instrument air (2% of operating cost)

c.   Cost of capital (using an average Treasury bill rate of 15% per annum)

d.   Depreciation (straight line, 25years useful life of the fractionators)

e.   Labour and maintenance cost of the fractionators: assume 5% of the equipment cost

f.   Contingency (10% of capital cost and 5% of operating cost)

3.3.2 EXPECTED CASH INFLOW/REVENUE

   a. Sales to be generated from the sale of propane, isobutene and normal butane.

   b. Salvage value of the equipment(15% of the initial investment cost)

CHAPTER 4

4.0   RESULTS AND DISCUSSIONS

This chapter presents and discusses the results of the various design stages and the
economic analysis for the depropanizer and butane splitter columns as outlined in chapter

two.

4.1 REVISED PROCESS FLOW DIAGRAM OF THE NGLS RECOVERY SECTION OF GHANA GAS PLANT

Simplified Process Flow Diagram of NGLs recovery section Ghana Gas plant incorporating a Depropanizer and a Butane splitter
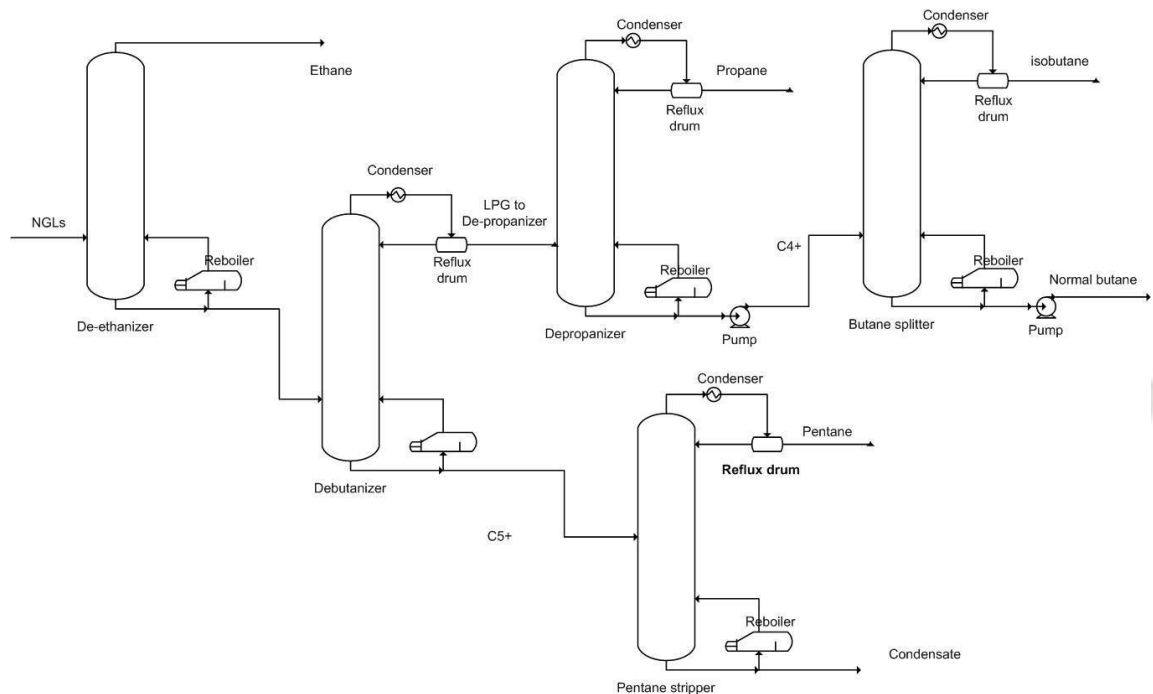
Figure 4.1   Process flow diagram of the NGLs recovery section of Ghana gas company plant incorporating a Depropanizer and a butane splitter

The overhead product of the debutanizer thus LPG is to feed a Depropanizer to separate it into propane as its overhead product and butane the bottom product. The butane is to be further separated into isobutene and normal butane by a butane splitter. All the other equipment will remain the same as described in section 3.1.

This new process flow will result in additional products of propane, normal butane and isobutene streams. These products will serve as feed stock for the petrochemical and allied industries. The normal butane can also be sold to homes for domestic purposes as substitute for LPG.

4.2   FEED COMPOSITION, FLOW RATE, COLUMN TEMPERATURE AND PRESSURE

4.2.1 FEED COMPOSITION

The feed composition to the debutanizer and volumes of natural gas liquids (NGLs) in barrels per day were derived from the raw gas compositions from the jubilee field at 150MMSCF (4247487 m^3) [Student Manual Intermediate Course - Main Process Systems, Ghana gas] by the use of a stripping calculation. The feed compositions and product splits of the debutanizer are presented in Table 3.1.

Table 4.1   Feed and product compositions of the debutanizer

Composition	Feed %	Moles of Distillate	%Distillate	Moles of Bottoms	% Bottoms
C$_3$	59.444	59.444	66.684	0.000	0.000
iC$_4$	9.900	9.900	11.106	0.000	0.000
nC$_4$	19.902	18.907	21.210	0.995	9.165
iC$_5$	4.947	0.891	1.000	4.055	37.347
nC$_5$	4.196	0.000	0.000	4.196	38.642
C$_6$	0.620	0.000	0.000	0.620	5.714
2m-C$_5$	0.000	0.000	0.000	0.000	0.000
3m-C$_5$	0.000	0.000	0.000	0.000	0.000

m-c-C_5	0.417	0.000	0.000	0.417	3.838
Benzene	0.337	0.000	0.000	0.337	3.103
c-C_6	0.000	0.000	0.000	0.000	0.000
2-m-C_6	0.051	0.000	0.000	0.051	0.472
3-m-C_6	0.000	0.000	0.000	0.000	0.000
m-c-C_6	0.137	0.000	0.000	0.137	1.266
Toluene	0.043	0.000	0.000	0.043	0.400
et-Benzene	0.001	0.000	0.000	0.001	0.007
p-Xylene	0.004	0.000	0.000	0.004	0.033
o-Xylene	0.001	0.000	0.000	0.001	0.013
He	0.000	0.000	0.000	0.000	0.000
H_2s	0.000	0.000	0.000	0.000	0.000
Total	100.00	89.142	100.000	10.858	100.000

The debutanizer distillate will feed the Depropanizer and the compositions of the various streams are shown in table 4.2.

Table 4.2   Feed and product splits of the Depropanizer

Composition	Feed %	Moles of Distillate	%Distillate	Moles of Bottoms	% Bottoms

C$_3$	66.684	63.350	99.000	3.334	9.259
iC$_4$	11.106	0.640	1.000	10.466	29.064
nC$_4$	21.210	0.000	0.000	21.210	58.900
iC$_5$	1.00	0.000	0.000	1.000	2.777
Total	100	63.99	100.00	36.01	100.00

The feed and product compositions of the butane splitter derived from the percent moles of the

Depropanizer distillate are presented in Table 4.3.

Table 4.3   Feed and product compositions of the butane splitter

Composition	Feed %	Moles of Distillate	%Distillate	Moles of Bottoms	% Bottoms
C$_3$	9.26	9.26	24.86	0.00	0.00
iC$_4$	29.06	27.61	74.14	1.45	2.36
nC$_4$	58.90	0.37	1.00	58.54	93.25
iC$_5$	2.78	0.000	0.00	2.78	4.43
Total	100	37.24	100.00	62.76	100.00

4.2.2 FLOW RATE

The total volume of NGLs stripped is 2370 m^3 per day as shown in Table 4.4. This is the

volume of feed to the debutanizer column. The various flow rates of streams to the

Depropanizer and Butane splitter columns derived from the feed to the debutanizer column are presented in Table 4.4.

Table 4.4   Flow rates of product streams of the various fractionators

	Debutanizer		Depropanizer		Butane splitter	
	Flow, m^3/day	Product	Flow, m^3/day	Product	Flow, m^3/day	Product
Feed	2370	NGLs	2113	LPG	761	Butane
Distillate	2113	LPG	1352	Propane	283	Isobutene
Bottom	257	Pentane+	761	Butane	477	N-butane

4.2.3 COLUMN TEMPERATURE AND PRESSURE

The column temperature and pressure for the Depropanizer and butane splitter obtained by bubble point calculation are presented in the tables 4.5 – 4.8 below.

Table 3.5 Bubble point pressure calculation of the Depropanizer at 80 ^{o}F (26.67 ^{0}C)

			try	200Psia		try	100 Psia	
Comp.	Feed %	xi	ki	xiki			ki	xiki
c3	99.000	0.990	0.800	0.792			1.400	1.386
ic4	1.000	0.010	0.320	0.003			0.540	0.005
nc4	0.000	0.000		0.000				0.000
ic5	0.000	0.000		0.000				0.000
Total	100	1.000		0.795	1.00			1.391
				0.795	1.000	1.391		
				200	166 Psia	100		

From Table 4.5 above, the bubble point pressure (column pressure) of the Depropanizer is 166 Psia (1145 kPa).

Table 4.6 Bubble point temperature calculation of the Depropanizer at 166 Psia (1145 kPa)

Comp.	Feed %	z_i	160°F (71°C) k_i	$z_i k_i$	180 °F (82 °C) k_i	$z_i k_i$
c3	9.259	0.093	2.100	0.194	2.400	0.222
ic4	29.064	0.291	0.950	0.276	1.200	0.349
nc4	58.900	0.589	0.710	0.418	0.920	0.542
ic5	2.777	0.028	0.365	0.010	0.490	0.014
Total	100	1.000		0.899	1.000	1.126
				0.899	1.00	1.126
				160	169°F	180

From table 4.6 above, the bubble point temperature (bottom temperature) of the Depropanizer is 169 ^{0}F (76 °C).

Table 4.7, Bubble point pressure calculation of the butane splitter at 80 °F (26.67 °C)

Comp.	Feed %	xi	ki	xiki		ki	xiki
			100 Psia (690 kPa)			70 Psia (483 kPa)	
c3	24.862	0.249	1.400	0.348		2.000	0.497
ic4	74.138	0.741	0.550	0.408		0.750	0.556
nc4	1.000	0.010	0.400	0.004		0.590	0.006
ic5	0.000	0.000		0.000			0.000
Total	100.000	1.000		0.760	1.000		1.059
				0.760	1.000	1.059	
				100	76	70	
					Psia		

From Table 4.7 above, the bubble point pressure i.e. column pressure of the butane splitter is

76 Psia (524 kPa)

Table 4.8 Bubble point temperature calculation of the butane splitter at 78 Psia (538 kPa)

Comp.	Feed %	z_i	k_i	$z_i k_i$	k_i	$z_i k_i$
			120 ºF (49ºC)		**160 ºF (71 ºC)**	
ic4	2.316	0.023	1.250	0.029	1.950	0.045
nc4	93.259	0.933	0.920	0.858	1.500	1.399
ic5	4.425	0.044	0.410	0.018	0.750	0.033
Total	100.000	1.000		0.905	1.00	1.477
				0.91	1.00	1.48
				120	127⁰F	160

From table 4.8 above, the bubble point temperature thus bottom temperature of the butane splitter is 127 ^{0}F (53 ºC).

4.3   MINIMUM NUMBER OF THEORETICAL STAGES FOR THE COLUMNS

The minimum stages of the columns were calculated from the (Fenske) equation 3.1 in chapter three.

4.3.1 MINIMUM STAGES OF THE DEPROPANIZER

Table 4.9 K-values and average volatilities of the Depropanizer products

Composition	Distillate			Bottom		
	K_1 values	x_1	$K_1.x_1$	K_2 values	x_2	$K_2.x_2$
C3	0.98	0.990	0.970	2.359	0.093	0.218
iC4	0.38	0.010	0.004	1.166	0.291	0.339
nC4				0.890	0.589	0.524
iC5				0.470	0.028	0.013
Total			0.974			1.095

From table 4.9, the average volatility αavg = 2.301 Using

equation 3.2, the separation factor is:

$$S_f = \left[\frac{99.00}{9.259}\right]_{LK} \cdot \left[\frac{29.064}{1.00}\right]_{HK} = 310.761$$

Using equation 3.1, the minimum stages:

$$S_m = \frac{\text{Log }(310.761)}{\text{Log}(2.301)} = 6.887 \text{ trays}$$

Correct for change in relative volatility by using equation 3.5:

$K_{LK} = 0.98 = \beta_{ij}\ (0.38)^b$ (condenser)

$K_{LK} = 2.359 = \beta_{ij}\ (1.166)^b$ (reboiler) Dividing

gives $2.4071 = (3.0684)^b$

$b = 0.7835;\ \beta_{ij} = 2.0915$

$$S_m = \frac{\log\left[\left[\frac{63.35}{3.334}\right]_{LK} \cdot \left[\frac{10.466}{0.640}\right]^{0.7835}{}_{HK}\left[\frac{36.01}{63.99}\right]^{1-0.7835}\right]}{\log\ (2.0915)}$$

$S_m = 6.789$ trays

4.3.2 MINIMUM STAGES OF THE BUTANE SPLITTER

Table 4.10 K-values and average volatilities of the butane splitter products

Composition	Distillate			Bottom		
	K_1 values	x_1	$K_1.x_1$	K_2 values	x_2	$K_2.x_2$
C3	1.800	0.249	0.448	-	0.000	0
iC4	0.700	0.741	0.519	1.373	0.023	0.032
nC4	0.500	0.010	0.005	1.023	0.933	0.954
iC5	-	-	-	0.470	0.044	0.021
Total			0.971			1.007

From table 4.10, the average volatility αavg = 1.371 Using

equation 3.2, the separation factor is:

$$S_f = \left[\frac{74.138}{2.316}\right]_{LK} \cdot \left[\frac{93.259}{1.00}\right]_{HK} = 2985.33$$

Using equation 2.1, the minimum stages:

$$S_m = \frac{\text{Log } (2985.33)}{\text{Log}(1.371)} = 25.358 \text{ trays}$$

Correct for change in relative volatility by using equation 3.5:

$K_{LK} = 0.7 = \beta ij \ (0.5)^b$ (condenser)

$K_{LK} = 1.373 = \beta ij \ (1.023)^b$ (reboiler) Dividing

gives $1.9614 = (2.046)^b$

$b = 0.941; \ \beta ij = 1.3439$

$$S_m = \frac{\log\left[\left[\frac{27.611}{3.334}\right]_{LK} \cdot \left[\frac{58.528}{0.372}\right]^{0.941}_{HK}\left[\frac{62.758}{37.242}\right]^{1-0.941}\right]}{\log\ (1.3439)}$$

$S_m = 23.36$ trays

4.4 MINIMUM REFLUX RATIO FROM THE UNDERWOOD EQUATIONS 4.4.1 MINIMUM REFLUX RATIO OF THE DEPROPANIZER

For boiling point feed thus q=1, the results of the Underwood equations are summarized in Table 4.11.

Table 4.11 Results of the Underwood equations for the Depropanizer

Depropanizer			αi.xf/(αi-φ)		αi.xD/(αi-φ)	
Composition	x_F	α_i(avg)	1.064	1.033	φ=1.033	x_D
C3	0.667	2.314	1.235	4.000	1.789	0.99
iC4	0.111	1.000	-1.735	-3.365	-0.303	0.01
nC4	0.212	0.761	-0.533	-0.594		
iC5	0.010	0.363	-0.005	-0.005		
Total	1.000		-1.039	0.0350	1.4855	

From Table 3.12, φ = 1.033 and (R_m +1) = 1.4855 therefore, R_m = 0.4855

4.4.2. MINIMUM REFLUX RATIO OF THE BUTANE SPLITTER

For boiling point feed thus q=1, the results of the Underwood equations are summarized in Table 4.12.

Table 4.12 Results of the Underwood equations for the Butane splitter

Butane Splitter			$\alpha_i \cdot x_f/(\alpha_i-\phi)$		$\alpha_i \cdot x_D/(\alpha_i-\phi)$	
Composition	Xf	α_i(avg)	1.1711	1.222	ϕ=1.222	x_D
C3	0.093	3.507	0.139	0.142	0.382	0.249
iC4	0.291	1.409	1.720	2.187	5.578	0.741
nC4	0.589	1.000	-3.442	-2.653	-0.045	0.010
iC5	0.028	1.336	0.225	0.326		
Total	1.000		-1.359	0.002	5.915	

From Table 4.13, ϕ = 1.222 and (R_m +1) = 5.915 therefore, R_m = 4.915

4.4. THEORETICAL STAGES TO OPERATING REFLUX RATIO FROM ERBAR-MADDOX CORRELATION

4.4.1 THEORETICAL STAGES TO OPERATING REFLUX RATIO FOR THE DEPROPANIZER

Assuming R = 1.3 R_m at optimum reflux:

R = 1.3 R_m = 1.3(0.4855)

R = 0.63115

$$\frac{L_0}{V_i} = \frac{R}{R+1} = \frac{0.63115}{0.6311+1} = 0.3869$$

$$\left(L_o / V_l\right)_m = \frac{R_m}{R_m+1} = \frac{0.4855}{0.4655+1} = 0.3268$$

From figure 3.3, the theoretical stages to operating reflux ratio $(S_m/S) = 0.55$

Therefore, $S = 6.789/0.55$

$S = 12.34$ trays (use 13trays including Reboiler)

4.4.2   Theoretical stages to operating reflux ratio for the Butane splitter Assuming

$R = 1.3\ R_m$ at optimum reflux:

$R = 1.3\ R_m = 1.3(4.915)$

$R = 6.3895$

$$\frac{L_0}{V_i} = \frac{R}{R+1} = \frac{6.3895}{6.3895+1} = 0.8647$$

$$\left(L_o / V_l\right)_m = \frac{R_m}{R_m+1} = \frac{4.915}{4.915+1} = 0.8311$$

From figure 2.3, the theoretical stages to operating reflux ratio for the butane splitter $(S_m/S) =$ 0.72

Therefore, $S = 23.36 /0.72$

$S = 32.44$ trays (use 33 trays including Reboiler)

4.5   FEED POINT LOCATION

Using Kirkbride equation 3.11, the feed point locations of the fractionators are presented below.

4.5.1 DEPROPANIZER FEED POINT LOCATION

$$\log\left(\frac{N_r}{N_s}\right) = 0.206 log\left[\left(\frac{36.01}{63.99}\right)\left(\frac{0.11106}{0.66684}\right)\left(\frac{0.0929}{0.01}\right)^2\right] = 0.1870$$

$$\left(\frac{N_r}{N_s}\right) = 1.5382$$

Also $N_r + N_s = 12$

Solving for N_r and N_s:

$N_s = 4.73 \approx 5$ and $N_r = 7.27 \approx 7$

4.5.2 BUTANE SPLITTER FEED POINT LOCATION

$$\log\left(\frac{N_r}{N_s}\right) = 0.206 log\left[\left(\frac{62.758}{37.242}\right)\left(\frac{0.589}{0.29064}\right)\left(\frac{0.02316}{0.01}\right)^2\right] = 0.26015$$

$$\left(\frac{N_r}{N_s}\right) = 1.8203$$

Also $N_r + N_s = 32$

Solving for N_r and N_s:

$N_s = 11.35 \approx 11$ and $N_r = 20.65 \approx 21$

Using Souders and Brown equations 3.12 and 3.13, the diameters of the columns are calculated as below.

4.6.1 DEPROPANIZER COLUMN DIAMETER

Table 4.13 Specific gravity of the liquids of the Depropanizer column

Comp.	Mole%	x	SG(lq)	xiSGi
c3	66.684	0.667	0.507	0.335
ic4	11.105	0.111	0.562	0.064
nc4	21.209	0.212	0.562	0.119
ic5	1	0.010	0.584	0.006
Total	100	1		0.526

The Liquid density (ρ_l) = (specific gravity of liquid) * (specific gravity of water) ρ_l

= 0.526 * 62.4 = 32.829 lbs. /ft^3

Vapour density

$$\rho_v = \frac{PMw}{zRT} = \frac{164*48.91}{0.9*10.73*584} = 1.42228$$

$$V_{max} = 0.150\sqrt{\frac{32.829-1.42228}{1.42228}} = 0.7049$$

Flashing the feed to the Depropanizer at the operating pressure 164 Psia (1131 kPa) and average

temperature 124 ^{0}F (51 ^{o}C), the vapour was 17 MMscfd (481382 m^3).

$$Q_{act} = \left(\frac{17*10^6*14.7}{86400*164}\right)\left(\frac{584}{520}\right) * 0.9 = 17.625 \; ft^3/s$$

$$D = \sqrt{\frac{17.625}{0.7049 X \; 0.7854}} = 5.64 \; ft$$

The column diameter is 5.64 ft. (1.719 m)

4.6.2   BUTANE SPLITTER COLUMN DIAMETER

Table 4.14, Specific gravity of the liquids in the butane splitter

Comp.	Mole%	x	SG(lq)	xiSGi
c3	9.259	0.093	0.507	0.047
ic4	29.064	0.291	0.564	0.164
nc4	58.899	0.589	0.563	0.332
ic5	2.777	0.028	0.584	0.016
Total	100	1		0.558

The liquid density (ρ_l) = 0.558 x 62.4 = 34.819 lbs. /ft³ (561 kg/m³) Vapour

density

$$\rho_v = \frac{PMw}{zRT} = \frac{78*57.214}{0.9*10.73*565} = 1.8179 \; \text{lbs. /ft}^3 \; (29 \; \text{kg/m}^3)$$

$$V_{max} = 0.150 \sqrt{\frac{34.819-0.8179}{0.8179}} = 0.9671 \; ft^2/s$$

Flashing the feed to the Depropanizer at the operating pressure 78psia and average temperature

105⁰F, the vapour was 6.017MMscfd.

$$Q_{act} = \left(\frac{6.017*10^6*14.7}{86400*78}\right)\left(\frac{565}{520}\right) * 0.9 = 12.798 \; ft^3/s$$

$$D = \sqrt{\frac{12.798}{0.9671X\ 0.7854}} = 4.105\ ft$$

The butane splitter column diameter is 4.106 ft (1.25 m)

4.7 HEIGHT OF THE COLUMN

Using equation 2.14, the height of the Depropanizer;

H = 2.3 * 12 = 27.6 ft. (8.41 m)

And height of the butane splitter is 73.6 ft. (22.43 m)

4.8. THICKNESS OF THE COLUMNS

The column thickness required for hoop's stress and wind velocity was calculated using equations 3.19-3.22.

4.8.1 THICKNESS OF THE DEPROPANIZER COLUMN

Thickness required for hoop's stress:

$$t_h = \frac{149.27X5.64}{2479-0.1(149.27)} = 0.342 \text{ inch (0.869 cm)}$$

Thickness required for wind velocity

$$t_a = \frac{90X10^{-6}(27.6)^2}{5.64} = 0.012 \text{ inch (0.031 cm)}$$

$$t = 0.342 + \frac{1}{8} = 0.467\ inch\ (1.186\ cm)$$

4.8.2 THICKNESS OF THE BUTANE SPLITTER COLUMN

$$t_h = \frac{61.27 X\ 4.105}{2479 - 0.1(61.27)} = 0.102 \text{ inch (0.256 cm)}$$

Thickness required for wind velocity

$$t_a = \frac{90 X 10^{-6}(73.6)^2}{4.105} = 0.119 \text{ inch (0.302 cm)}$$

$$t = 0.5(0.102 + 0.119) + \frac{1}{8} = 0.235\ inch \text{ (0.60 cm)}$$

4.9   CONDENSER AND REBOILER LOADS

The condenser and reboiler loads were calculated using Equation 2.15;

4.9.1 CONDENSER AND REBOILER LOADS FOR THE DEPROPANIZER

a. Depropanizer Condenser duty

$$Q_c = (47740.8 * 31.695)(0.63115 + 1)174.4 = 430448134 \text{ Btu/day } (454146823$$

$$\text{kJ/day)}$$

Assuming 15% heat loss, the Depropanizer condenser duty is 495015354Btu/day thus 21MMBtu/h (21761202 kJ/hour)

b. Cooling air Mass flow rate for the Depropanizer condenser;

Using Equation 3.16;

$$m_{air} = \frac{495015354}{0.998(568-535)} = 14925836\ lbs/D \text{ (6770246 kg/day)}$$

The Depropanizer condenser cooling air mass flow rate is 15MM lbs. /day (242094 kg/hour)

c. Depropanizer Reboiler duty

Using Equation 3.17;

$$Q_c = [(31.695 * 0.63115 * 47740.8) + (32.829 * 74606)][174.4 + (11 *$$

$$(168 - 124)] = 2241365853 \; Btu/day \quad \text{(2364766161 kJ)}$$

Assuming 15% heat loss, The Depropanizer Reboiler duty is 2578MMBtu/day

(113311712 kJ/hour).

d. Hot oil rate for the reboiler;

Using equation 3.18

$$m_{oil} = \frac{2577570731}{0.56*(987-960)} = 170474255 \; lbs/day \quad \text{(77325827 kg/day)}$$

The Depropanizer Reboiler hot oil mass rate is 170MMlbs. (3221909 kg/hour).

4.9.2 CONDENSER AND REBOILER LOADS FOR THE BUTANE SPLITTER

a. Butane splitter Condenser duty

Using Equation 3.15;

$$Q_c = (10005.19 * 34.264)(6.3895 + 1)159.8 = 404813727 \; Btu/day \quad \text{(427101092}$$

kJ/day)

Assuming 15% heat loss, the Butane splitter condenser duty is 466 MMBtu/day

(20465261 kJ/hour).

b. Cooling air mass flow rate for Butane splitter condenser;

Using Equation 3.16;

$$m_{air} = \frac{465535786}{1.005(568-540)} = 16543560 \; lbs/D \quad \text{(7504033 kg/day)}$$

The cooling air mass flow rate is 17MM lbs. /day (312668 kg/hour.)

c. Butane splitter Reboiler duty

Using Equation 3.17;
$$Q_r = [(34.264 * 6.3895 * 10005.19) + (34.844 * 26865.78))][159.8 + (11 *$$

$$(129 - 105)] = 1325030095 \; Btu/day \quad \text{(1397980757kJ/day)}$$

Assuming 15% heat loss, the Butane splitter Reboiler duty is 1524MMBtu/day

(66986578 kJ/hour)

d. Hot oil mass flow rate for the Butane splitter Reboiler;

Using equation 3.18

$$m_{oil} = \frac{1524784609}{0.56*(987-960)} = 100845543 \; lbs/day \; \text{(45742772 kg/day)}$$

The hot oil rate for the Butane splitter is 101 MMlbs/day (1905949 kg /hour)

Summary of the design parameters for both fractionators are presented in Table 4.15 below.

Table 4.15, Summary of design parameters for the Depropanizer and Butane splitter

		SIZE		
NO.	PARAMETER	DEPROPANIZER	BUTANE SPLITTER	UNIT
1	Column pressure	1145	524	kPa
2	Column temperature	76.11	52.78	^{O}C
3	Column diameter	1.75	1.25	m
4	Column height	8.41	22.43	m
5	Column thickness	1.19	0.61	cm
6	Insulation thickness	2.54	2.54	cm

7	Separation factor	310.76	2985.33	-
8	Theoretical trays	6.789	23.36	trays
9	Actual trays(plus reboiler)	13	33	trays
10	Minimum reflux ratio	0.49	4.92	-
11	Operating reflux ratio	0.63	6.39	-
12	Feed entry tray	5	11	tray
13	Condenser duty	21761202	427101092	kJ/h
14	Cooling air mass flow rate	242094	312668	kg/h
15	Reboiler duty	11331112	66986578	kJ/h
16	Hot oil flow rate	3221909	190549	Kg/h

4.10   COST- BENEFIT ANALYSES

The cost-benefit analysis compares all the estimated costs to be incurred with all the projected revenues to be generated.

4.10.1   CAPITAL COST OF EQUIPMENT:

The capital costs of the major equipment are estimated from equations 2.24 - 2.32:

4.10.1.1   CAPITAL COST OF DEPROPANIZER

1. $Vessels\ (\$) = 2\big((27.6 + (2.2 * 5.64)\big)\big((100 * 0.467) + (45 * 0.467 * 5.64) + (5 * 5.64)\big) = \$6942$

2. $Tray\ (\$/tray) = 170 * 5.64 \left(0.5 + \frac{10}{12}\right) = \$1278.4$

Therefore, cost of the 12 trays = $15341

3. $Heat\ exchangers_{(condenser)}\ (\$) = 5000 + 23 * 28144 = \$65230$

44

4. $Heat\ exchangers_{(reboiler)}\ (\$) = 5000 + 23 * 3989 = \$96747$

5. $Pumps_{bottoms}\ (\$) = 4000 + 50(116)^{1/2} + 3 * 11 = \$4572$

6. $Pumps_{reflux}\ (\$) = 4000 + 50(130)^{1/2} + 3 * 2 = \$4576$

7. $Motors_{bottoms}\ (\$) = 85(671) - 1000 = \$56050$

8. $Motors_{reflux}\ (\$) = 85(66) - 1000 = \$4613$

9. $Motors\ starters_{bottoms}(\$) = 30(671) + 750 = \$20880$

10. $Motors\ starters_{reflux}\ (\$) = 30(66) + 750 = \$2730$

11. $Control\ systems\ (\$/loop) = 5600 + 2600 = \$8200$

$Total\ for\ three\ loops = \$24600$

12. $Vessel\ insu.\ (\$) = (24 + 8)\left(5.64 + \frac{1}{6}\right)26.6 + (50 + 14)\left(5.64 + \frac{1}{6}\right)^2 +$

$57(5.64) = \$7422$

13. $Pipe\ insulations\ (\$/ft) = 6 + (0.35 + 0.725)(2 + 3) = \$11.375/ft$

$Total\ pipe\ insulations\ cost\ (\$) = 11.375*90 = \$1024$

14. In plant piping cost ($) = 0.20*6942=$1388

Therefore, cost of the Depropanizer =$1740910

Shipment, insurance, civil works and installation cost = 0.30*1740910 = $522273

Capital cost of the Depropanizer plus contingency =**$2,263,183.00**

4.10.1.2   CAPITAL COST OF BUTANE SPLITTER

1. $Vessels\ (\$) = 2\big((73.6 + (2.2 * 4.105)\big)\big((100 * 0.235) + (45 * 0.235 *$

$4.105) + (5 * 4.105)\big) = \$14450$

2. $Tray\ (\$/tray) = 170 * 4.105\left(0.5 + \frac{10}{32}\right) = \$567$

Therefore, cost of the 32 trays = $18144

3. $Heat\ exchangers_{(condens}$

4. $Heat\ exchangers_{(reboiler)}$

$Pumps_{bottoms}\ (\$) = 4000 + 50(73)^{1/2} + 3*11 = \$4460$ 5.

6. $Pumps_{reflux}\ (\$) = 4000 + 50(277)^{1/2} + 3*2 = \$4837$

$Motors_{bottoms}\ (\$) = 85(449) - 1000 = \$37155$ 7.

$Motors_{reflux}\ (\$) = 85(310) - 1000 = \$25323$ 8.   9.

$Motors\ starters_{bottoms}(\$) = 30(4$

10. $Motors\ starters_{reflux}\ (\$) = 30(310) + 750 = \$10050$

11. $Control\ systems\ (\$/loop) = 5600 + 2600 = \$8200$

$Total\ for\ three\ loops = \$24600$

12. $Vessel\ insu.\ (\$) = (24 + 8)\left(4.105 + \frac{1}{6}\right)73.6 + (50 + 14)\left(4.105 + \right.$

$\left.\frac{1}{6}\right)^2 + 57(4.105) = \$11462$

13. $Pipe\ insulations\ (\$/ft) = 6 + (0.35 + 0.725)(2 + 3) = \$11.375/ft$

$Total\ pipe\ insulations\ cost\ (\$) = 11.375*90 = \$1024$

14. In plant piping cost ($) = 0.20*14450=$2890

Therefore, cost of the butane splitter =$1691279

Shipment, insurance, civil works and installation cost = 0.30* 1691279= $507384

Capital cost of the Butane splitter plus contingency =**$2,198,663.00**

4.10.2   OTHER COSTS:

Used GH¢0.58 per kilowatt hour, exchange rate of US$1.0= GH¢3.15and US$3.46per

gallon of Petrotherm (hot oil), 1HP=0.746 and 98% plant availability and

utilization.

4.10.2.1   OTHER COSTS FOR THE DEPROPANIZER

1. Annual Operating Cost: I.   Hot oil:

$$\$ = 3.46 * 18350297 * 1.01 = \$64126948$$

II.   Electricity:

1. *Reflux pump* = 66*0.746*24*365*0.98*0.186 = $78392

2. *Bottoms transfer pump* = 671*0.746*24*365*0.98 *0.186= $799290

3. Hot   oil   heating   cost   =   2.937*10⁻⁴*2577570731*365*0.98*0.186= $50367039

Total annual operating cost for the Depropanizer = $115,371,669

2. Depreciation

$$d = \left(\frac{2263183 - 339477}{25}\right) = \underline{\$76,948}$$

3. labour and maintenance cost: =5% *2263183 = $113,159

4. contingency = 5% * 115602644 = $5,780,132

4.10.2.2   OTHER COSTS FOR THE BUTANE SPLITTER

1. Annual Operating Cost:

i.   Hot oil:

$$\$ = 3.46 * 14654627 * 1.01 = \$51,212,060$$

ii.   Electricity:

1. Reflux pump = 449*0.746*24*365*0.98*0.186 = $545,761

2. Bottoms transfer pump = 310*0.746*24*365*0.98 *0.186= $369,270

3. Hot oil heating cost = 2.937*10⁻⁴*1523784609*365*0.98*0.186= $29,775,524

Total annual operating cost for the butane splitter = $55,104,615

2. Depreciation

$$d = \left(\frac{2198663-329799}{25}\right) = \underline{\$74,755}$$

3. Labour and maintenance cost: =5% *2198663 = $109,933

4. Contingency= 5% * 95128785 = $4,756,439

4.10.3   EXPECTED CASH INFLOW/REVENUE

a. Sales

Using $1.04, $1.28 and $1.17 per gallons of propane, isobutene and normal butane respectively;

1. Propane = 357126*365*0.98*1.04 = $132,853,729

2. Isobutene = 126126*365*0.98*1.28 =$57,747,546

3. Normal butane =74844*365*0.98*1.17 = $31,322,888

b. Salvage values

1. Depropanizer=15% *2263183 = <u>$339,477</u>

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2. Butane splitter =15% *2198663 = <u>$329,799</u>

4.10.4 NET PRESENT VALUE (NPV)

Using a discounted rate of 15% and an inflation adjustment rate of 8% the cash outlays and inflows from year zero to year twenty-five and their net present value are presented in Table 4.16.

Table 4.16, results of net present value analysis from year zero to year twenty-five

YEAR	EXPENSES/COST ($)						INCOME ($)			NET CASH FLOW($)
	EQUIPMENT	LABOUR & MAINTENANCE	DEPRECIATION	UTILITIES	CONTINGENCY	TOTAL	SALES	SALVAGE VALUE	TOTAL	
0	4461846					4461846				-4461846
1		223092	151703	197274284	10536571	208185650	221924163		221924163	13738513
2		260215	176946	230100725	12289856	242827742	258852344		258852344	16024602
3		281032	191102	248508783	13273045	262253962	279560531		279560531	17306570
4		303514	206390	268389485	14334889	283234278	301925374		301925374	18691095
5		327795	222901	289860644	15481680	305893021	326079404		326079404	20186383
6		354019	240734	313049496	16720214	330364462	352165756		352165756	21801294
7		382340	259992	338093456	18057831	356793619	380339016		380339016	23545397
8		412928	280792	365140932	19502458	385337109	410766138		410766138	25429029
9		445962	303255	394352207	21062654	416164078	443627429		443627429	27463351
10		481639	327515	425900383	22747667	449457204	479117623		479117623	29660419
11		520170	353717	459972414	24567480	485413780	517447033		517447033	32033253

12		561784	382014	496770207	26532878	524246883	558842795		558842795	34595913
13		606726	412575	536511823	28655508	566186633	603550219		603550219	37363586
14		655264	445581	579432769	30947949	611481564	651834237		651834237	40352673
15		707686	481228	625787391	33423785	660400089	703980976		703980976	43580887
16		764300	519726	675850382	36097688	713232096	760299454		760299454	47067358
17		825444	561304	729918413	38985503	770290664	821123410		821123410	50832746
18		891480	606208	788311886	42104343	831913917	886813283		886813283	54899366
19		962798	654705	851376836	45472691	898467030	957758345		957758345	59291315
20		1039822	707081	919486983	49110506	970344393	1034379013		1034379013	64034620
21		1123008	763648	993045942	53039346	1047971944	1117129334		1117129334	69157390
22		1212849	824739	1072489617	57282494	1131809700	1206499681		1206499681	74689981
23		1309877	890719	1158288787	61865094	1222354476	1303019655		1303019655	80665180
24		1414667	961976	1250951890	66814301	1320142834	1407261228		1407261228	87118394
25		1527840	1038934	1351028041	72159445	1425754260	1519842126	669276	1520511402	94757142
TOTAL									NPV	$243,225,505

From Table 4.16, the net present value thus NPV is $243,225,505.00. This means the total revenue generated exceeds the total cost incurred by this investment

4.10.5 CASH FLOW ANALYSIS

An analysis of the cash outflow, the cash inflow and the net cash flow are presented in Table 4.17 below.

Table 4.17 Results of Cash flow Analysis

S/N	ITEM	DESCRIPTION	DEPROPANIZER	BUTANE SPLITTER	
		Cash flow Analysis		Value/$	
		Cash outflow			
1	Equipment cost	Fractionator Package	1,740,910.00	1,691,279	
		Shipment, insurance, civil works and installation cost	522,273.00	507,384.00	
		Sub-total	**2,263,183.00**	**2,198,663.00**	
2	Total equipment cost				**4,461,846.00**
3	Annual Operating Cost	Hot oil	64,126,948.00	51,212,060.00	
		Electricity	51,244,721.00	30,690,555.00	
		Depreciation (straight line)	76,948.00	74,755.00 labour	
		and maintenance cost	113,159.00	109,933.00	
		Contingency	5,780,132.00	4,756,439.00	
		Sub-total	**121,341,908.00**	**86,843,742.00**	
4	Total Annual Operating Cost				**208,185,650.00**

| 5 | Total cost per Equipment | 123,605,091.00 | 89,042,405.00 |
| 6 | Total cash outflow | | **212,647,496.00** |

Cash inflow		
7	Sales Propane	132,853,729.00	
	Iso-butane	57,747,546.00
	Normal butane	31,322,888.00
	Sub-total	**221,924,163.00**

8	Salvage values	Depropanizer	339,477.00	
		Butane splitter		329,799.00
		Sub-total		**669,276.00**
9	Total cash inflow	**222,593,439.00**		
10	Net cash flow	**9,945,943.00**		
11	Payback period (Year)	**1.0**		

CHAPTER 5

5.0   OBSERVATIONS, CONCLUSIONS AND RECOMMENDATIONS

This chapter presents conclusions drawn from and recommendations made from the results and discussions in chapter four (4).

5.1   OBSERVATIONS

From the results and discussions in chapter four, the following observations were made:

i.   The addition of Depropanizer and butane splitter to the process stream of Ghana national gas company processing plant will result in the separation of the NGL into its pure components.

ii.   2113 m^3/day of liquefied petroleum gas (LPG) will yield 1352 m^3 of propane, 283 m^3 isobutane and 477 m^3 of normal butane per day.

iii.   A 1.75 x 8.41 m, 12 actual trays Depropanizer and 1.25 x 22.43 m, 32 actual trays butane splitter are required to generate the barrels of products mentioned in point ii.

iv.   The minimum operating temperatures and pressures for the Depropanizer and butane splitter are 76.11^{0}C/1145 kPa and 52.78 ^{0}C/524 kPa respectively

v.   The two fractionators cost $4,461,846.00.

vi.   An additional $208,185,650.00 would be required as working capital for the first year of operation.

vii.   A positive net present value (NPV) of $243,225,505.00 and a net positive cash flow of $9,945,943.00 means the expected revenue exceeds the expected cost.

viii.   The investment has a payback period of one year.

5.2   CONCLUSION

Inferring from the observations outlined above, I can conclude that it is economically viable to separate the NGLs at the gas processing plant of the Ghana National Gas Company into individual fractions.

5.3   RECOMMENDATIONS

After a careful analysis of the results, observations and conclusion, I hereby recommend that:

i.   The material balance and the design parameters of the fractionators obtained by the approach used in this project should be confirmed by the use of a process modeling software such as Aspen hysis, Aspen plus etc.

ii.   Sensitivity analysis should be conducted to identify the factors that will pose as threats to this project.

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