MAXIMIZATION OF NATURAL GAS LIQUIDS RECOVERY OF THE GHANA

NATIONAL GAS PLANT- ATUABO

BY

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A PROJECT SUBMITTED TO THE SCHOOL OF GRADUATE STUDIES, KWAME NKRUMAH UNIVERSITY OF

SCIENCE AND TECHNOLOGY, KUMASI IN PARTIAL FULFILMENT OF THE REQUIREMENT FOR THE DEGREE

OF

PROFESSIONAL MASTERS IN ENGINEERING WITH MANAGEMENT, OIL AND GAS

DEPARTMENT OF CHEMICAL ENGINEERING

COLLEGE OF ENGINEERING

©NOVEMBER 2015

DECLARATION

I hereby declare that, except for reference to other peoples work, which I have acknowledged, this Project submitted to the School of Graduate Studies, Kwame Nkrumah University of Science and Technology, Kumasi is the result of my own work and has not been presented for any other degree elsewhere.

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ACKNOWLEDGEMENT

ABSTRACT

Ghana National Gas Company was set up in 2011 to process natural gas produced from the Jubilee fields. The natural gas produced from the reservoir is piped to the surface as a dense phase and taken through the well head treatment to produce a liquid and gaseous phase. The two phases are separated with a liquid separator unit after which the sweet lean gas obtained is taken through dew point control.

After dew point control, Natural gas liquids are recovered from the lean gas stream by expansion refrigeration with the sales gas being forwarded to Aboadze thermal plant and the liquids, fractioned into LPGs, pentanes and condensates sold separately for domestic consumption, power production and refinery feed respectively.

Ghana gas has installed a Joule Thompson valve for this purpose in phase 1 to recover $70m^3$/hr. of natural gas liquids at an efficiency of 60% as per design. There is still a high amount of Natural gas liquids which will be left in the lean gas stream to be forwarded to the Aboadze thermal plant

The use of an alternative expansion refrigeration method, Turbo expansion is likely to recover more liquids as its 90% plus efficient. This project looks at how much liquids can be generated as per the forecasted flows of both methods and the yearly revenue to be accrued justifying the need to install a Turbo Expander in the next phase.
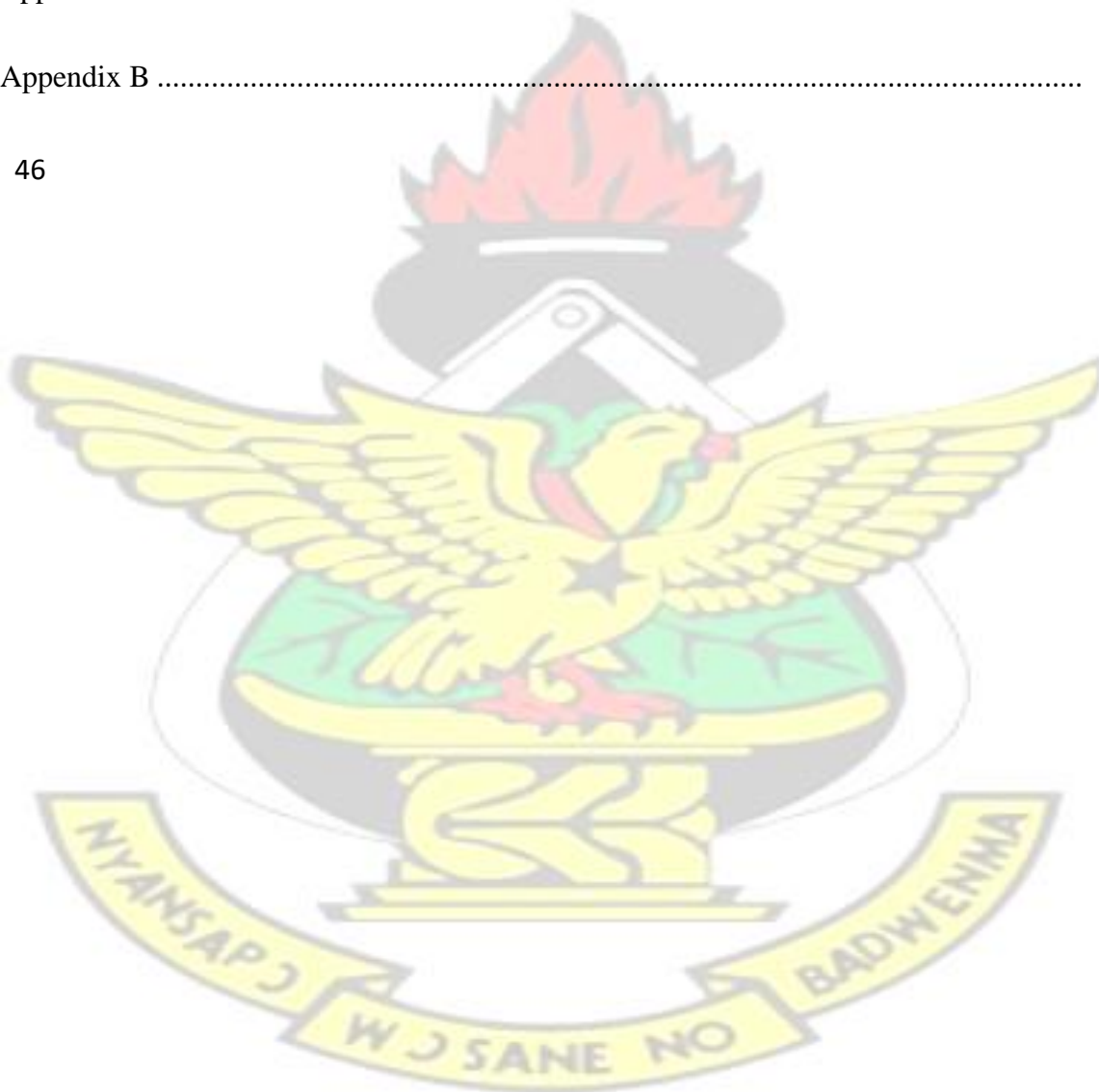
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ABBREVIATIONS

Abbreviation	Description
BBL/D	Barrel per day
BTU	British Thermal Units
FPSO	Floating, Production, Storage and Offloading
GNGC	Ghana National Gas Company
GNPC	Ghana National Petroleum Company
GDP	Gross Domestic Product
JT	Joule Thompson
Km^2	Kilometer square
LPG	Liquified Petroleum Gas
LT	Low Temperature
MMSCFD	Million standard cubic feet per day
m^3/hr	Meter cube per hour
MW	Mega Watt
NGL	Natural gas liquids
VRA	Volta River Authority
^0C	Degree Celsius
W/W	Weight per weight

CHAPTER ONE

1.0.Introduction

Ghana is a West African country, with 25 million estimated populations [Ghana statistical service, 2010] and covering an area of 238,500 km^2 .Ghana, is former British colony called Gold Coast and was the first country to attain independence in sub-Saharan Africa from the British colonial masters in 1957.

It is bounded by Republic of Cote d'Ivoire on west, Republic of Togo to the east, Republic of Burkina Faso on the northern border and the Sea (Gulf of Guinea) on the southern border. It is one of the most stable multi- party states democracies on the continent, and has held several elections with peaceful and smooth political transitions in a civil conflict ridden West African region.

Ghana has an agricultural dominated economy, with the agric sector employing almost 40% of the working population and contributing 21.3% of the country's gross domestic product (GDP) and it's a leading exporter of cocoa in the world. It has a very strong mining sector producing commodities such as diamond, manganese, bauxite, gold, and recently crude oil. Crude oil has become a very important subsector of our industrial sector contributing 6.1% of GDP in 2013 [Ghana Statistical service, 2013]

1.1.0 Hydrocarbon exploration in Ghana

Crude oil exploration begun in the country in 1896, with exploration wells drilled around Half-Asini when oil seeps were found in the on-shore Tano Basin, located in the Western Region. In 1970, the Salt pond by a Signal Amoco well off shore Salt-pond about 100 km on the west of Accra towards Cape coast. This field produced about 3.47 million barrels of

oil and 14 billion cubic feet of gas which was flared. The platform, Mr. Louie, was used for the production. Operation was restarted in 2000 with a current production of about 600 barrels of oil per day. [Tullow Oil, 2010]

In t h e y e a r 2004, the government of Ghana gave out licenses for off-shore oil and gas exploration and production blocks to several multinational companies. This led to the increase of exploration activities for crude oil in the country with companies like; Tullow oil, ENI, Vitol, Kosmos, Hess Corporation, and others intensifying their search. Figure 1.1 below illustrates the oil blocks of Ghana.
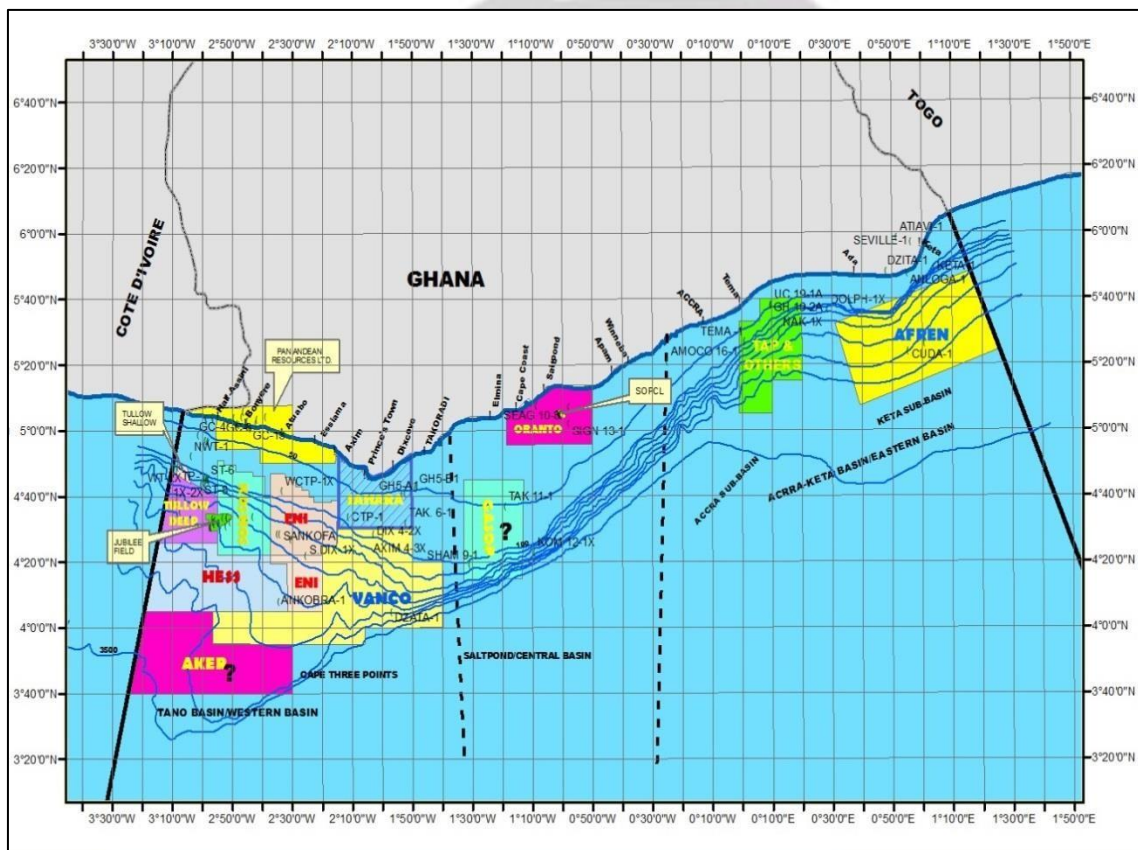
Figure 1.1. Oil blocks of Ghana (Source; Tullow oil, 2010)

1.2.0 Oil discovery in Ghana

In 2007, oil with associated gas was discovered in economic quantities in Ghana. The area located in the Western Region was named as "Jubilee Fields" (figure 1.2) by the Jubilee Partners; Tullow oil, Anadarko energy, Kosmos energy, EO Group, Sabre Oil & Gas, and Ghana National Petroleum (GNPC). The Jubilee field block is about 60km offshore between Deep water Tano block and Western Cape three Points block. The reserves are spread out in the Deep water Tano and 'West Cape Three Points blocks', with the wells at a water depth between 1,100m and 1,300 m and at a total depth between 3,400m and 4,200m. [Offshore-Technology.com, 2011] The field has an estimated recoverable reserve of more than 370 million barrels; with an upside potential of 1.8-1.9 billion barrels with 5trillion cubic feet of associated natural gas. A total of 17 wells have been drilled consisting of; 9 oil production wells, 2 gas injection wells and 6 water injection wells. [Tullow Oil, June 2010]. Figure 1.3 below shows the Drill Rig Eirik Raude which made the discovery

Figure 1.2 location of the Jubilee fields (Source; Tullow oil, 2010)

Figure 1. 3 Eirik Raude Platform (Source; Tullow oil, 2010)

The partners of the Deep water Tano block were Tullow oil, 49.95%, Kosmos Energy with 18%, Anadarko Petroleum Corporation, 18%, Sabre Oil & Gas, 4.05%, and Ghana National Petroleum Corporation (GNPC) with 10%.

West Cape Three Points block is held by Tullow oil Ghana, 22.9%, Kosmos Energy, 30.88%, Anadarko Petroleum, 30.88%, Sabre Oil & Gas, 1.85%, Ghana National Petroleum (GNPC), 10%, and EO Group, 3.5%. EO group offloaded its shares to Tullow oil in 2012.

1.3.0 Oil Production in Ghana

Development of the production facility was carried out within a record three and half years, oil production begun in December 2010 after the installation of the production vessel FPSO Kwame Nkrumah (figure 1.4). The FPSO Kwame Nkrumah named after

Ghana's first president, was designed to process 230,000 Barrels of sea water per day for water injection, 120,000 Barrels/day of oil, and 160mmscfd of natural gas [Tullow Oil, June 2010]

Figure1.4.0 FPSO Kwame Nkrumah (Source; Tullow oil, 2010)

1.4.0 Gas Production in Ghana

The jubilee field operators have since 2011 been re-injecting almost 80% of the natural gas from the field back into the reservoir, and utilizing the remaining 20% as fuel for the FPSO, and flaring a small amount of the 20%, it was estimated that continuous reinjection of gas into the jubilee reservoir could threaten the safety of the reservoir [Ministry of Energy Ghana, 2012].

The gas can be used to fire thermal plants which are currently being fired with gas imported from Nigeria via the West African Pipe line and crude oil. The gas from the West African

Pipeline is very erratic and the use of crude oil is very expensive as compared to natural gas.

In recognition of the immense benefit of the natural gas, the President, Professor John Evans Atta Mills in 2011 commissioned a "National Gas Development Task Force" to investigate and make appropriate recommendations for the rapid development of a National gas commercialization infrastructure system. The Task Force recommended the treatment of associated gas from the Jubilee Field for commercial use. [Ghana Gas, 2011]

Acting on the recommendation of the Task force, President John Atta Mills established an indigenous company, Ghana National Gas Company to undertake the project.

Ghana National Gas Company was formed in July 2011 to build, own and operate the relevant infrastructure required for the gathering, processing, transporting and marketing of processed Natural Gas. [Ghana Gas, 2011]

1.4.1 Western Corridor Gas Infrastructure Development Project

Ghana Gas in November 2011 contracted Sinopec International Petroleum Service Corporation to undertake the Western Corridor Gas Infrastructure Development Project also known as the "Early Phase Gas Infrastructure Development Project" which includes; 45km Offshore Pipeline, 110km onshore pipeline from Atuabo to Aboadze Thermal Plant, 75km lateral lean gas pipeline to Prestea, 150mmscfd capacity Natural gas processing Plant, NGLs Export System, and Offices. A $1billion loan was secured from the Chinese Development Bank for this purpose. [Ghana National Gas Company, 2013].

Figure 1.5 illustrates the western corridor infrastructure supply chain.

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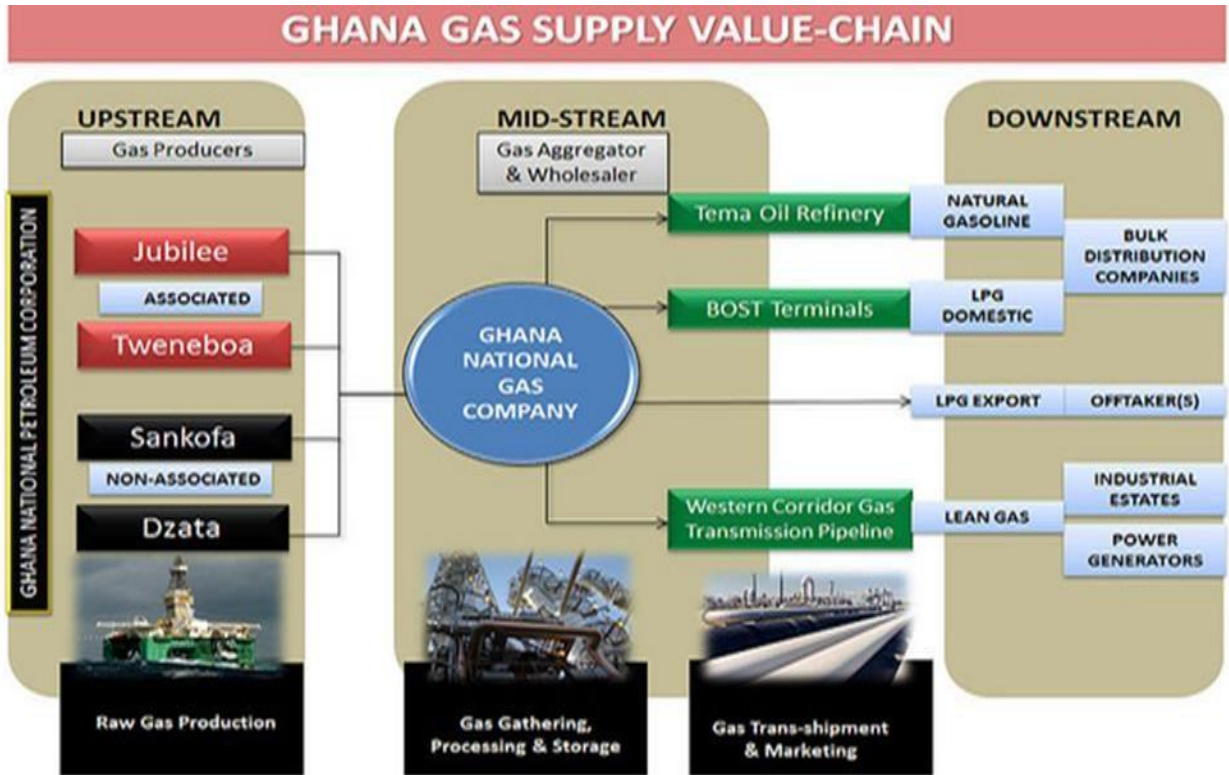
Figure1.5 .Western corridor Gas Infrastructure Project- Supply chain (Source; Ghana Gas, 2010)

1.5.0 Problem Statement

As part of the Western corridor Natural gas infrastructure project, and in line with its mandate, the Ghana National Gas company is constructing a 150mmscfd phase one plant in which it has installed a Joule Thompson valve with the aim of recovering natural gas liquids from the raw gas being processed.

The sales gas produced will be sold to VRA separate from the other products, LPG, npentane, and condensates which are generated as 'by products' from the raw gas treatment.

The LPG, n-pentane and heavy condensates are obtained from the fractionation of natural gas liquids which are also generated after dew pointing, and their relative amounts are dependent on the amount of NGL's recovered. The LPG is mainly used for domestic purpose in Ghana, and it's in high demand with the government's policy in place to reduce the use of wood as fuel domestically in order to control de-forestation.

The n-pentane produced will be used to fire a 50 MW thermal plant to be built by Cenet Power whiles the heavy condensates will be sold to Tema oil refinery.

It is very important therefore to recover as much NGL as possible from the lean gas stream to maximize the recovery of LPG's, n-pentane and heavy condensates and to prevent;

✓ Loss of revenue as VRA will not pay a premium for the rich-lean gas with a higher heating value than that specified in the sales contract as a result of NGL constituents in the sales gas stream.

✓ Risk of hydrate formation in pipelines causing blockage and reduction in gas flow

✓ Liquid slagging may occur, causing damage to equipment downstream.

The installed JT valve will recover 70m3/hr. of NGL from the lean gas stream with an efficiency of 60%. This project therefore looks an alternative for increasing the recovery of natural gas liquids by the installation of a Turbo expander as the flows to the plant are increased from 120mmscfd in 2014 to700mmscfd in 2030

1.6.0 Objectives

The objectives of this project are to;

✓ Estimate the amount of natural gas liquids to be generated with the forecasted flows from 2014 to 2030 when the JT valve is used and when a Turbo expander is used and to quantify the difference in NGL recovery

✓ Estimate the amount of revenue to be generated from the use of the JT valve and the Turbo Expander over the project life span and evaluate the difference.

CHAPTER TWO

Literature review

2.1.0 Gas Processing

Natural gas is normally obtained from underground wells/reservoirs and has a complex combination of hydrocarbons and non-hydrocarbons. It is formed as a result of the decomposition of plant and animal remains buried deep down the surface of the earth and is composed mainly of methane, with heavier hydrocarbons such as Ethane, Propane, Butane and Pentane in addition to non-hydrocarbons such as Carbon dioxide, Nitrogen and Hydrogen Sulphide. It is mostly saturated with water and may contain varying amount of compounds such as Helium, Carbonyl Sulphide, Carbon di-Sulphide and Mercaptans [Ben Asante, 2013]

Prior to natural gas usage as a fuel or feed stock to petrochemical industry, it is treated to produce what is referred to as "pipeline quality" (Dry natural gas). Natural Gas is processed for two main purposes which are:

• Removal impurities from the gas

• To increase liquids recovery from the raw gas stream

2.2.0. Natural Gas Processing Schematic and Process Description

The processing of raw gas involves the three major steps;

➤ Gas-oil separation usually carried out at the well head

➤ Gas treating or conditioning which involves the removal of impurities such as hydrogen Sulphide gas and Carbon dioxide (gas sweetening) and gas-water separation

➤ Gas processing consists of two main operations which are Natural Gas Liquids recovery and separation from the gas stream and its subsequent fractionation into the required economic components products.

Figure 2.1. below shows a typical schematic of the natural gas processing.
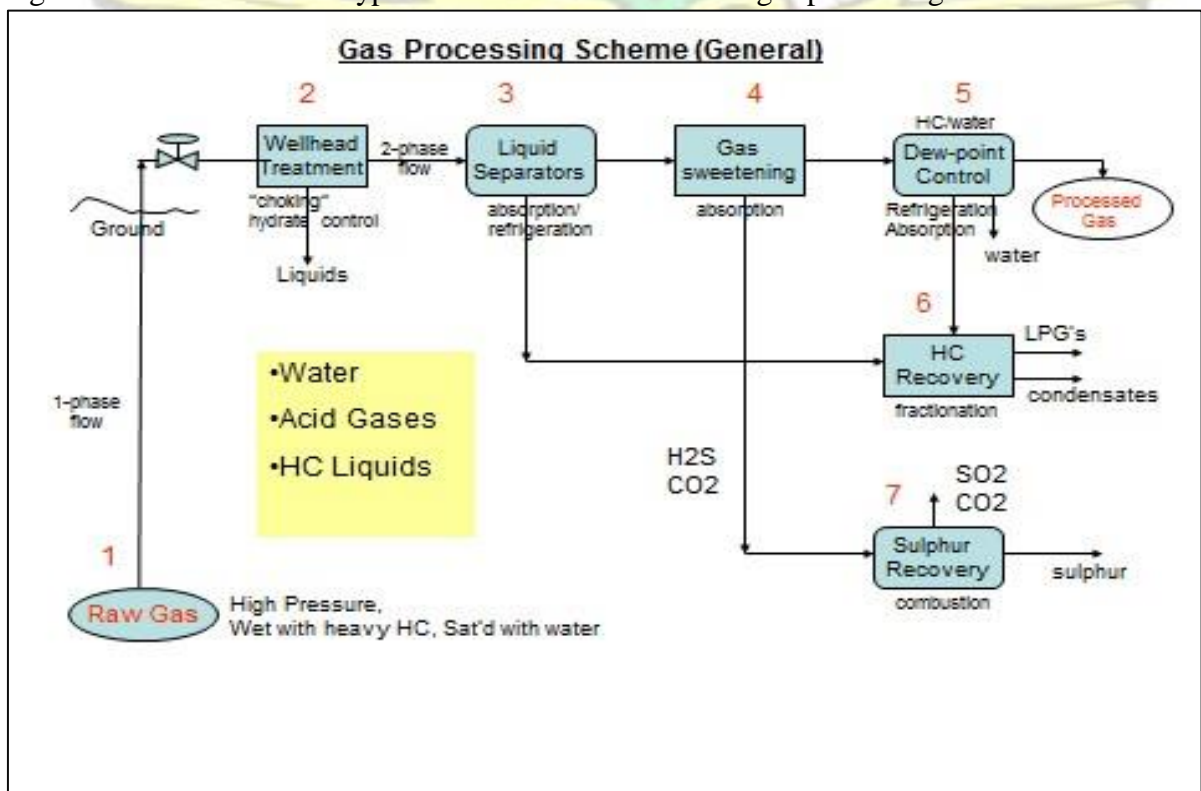
Figure 2.1 Gas processing Schematic (Source; Ben Asante, 2013)

2.3.0. Natural Gas Liquids Recovery

Natural Gas Liquids (NGLs) are liquids recovered from the raw natural gas stream including ethane, Liquefied Petroleum Gas (LPG) and heavy hydrocarbons to meet specific contractual and operating requirements [Ben Asante, 2013]

Natural Gas Liquids recovery is based on the principle of phase changes, thus to recover and separate NGL's, a new phase has to be created from the gas stream for the separation to occur. A phase change for NGL recovery and separation most often involves the regulation of one or more of the following operating parameters: pressure, temperature, and composition/concentration. [Abdel-al, 2003]

There are two ways of achieving the change in phase which are;

➢ By the use of a mass separation Agent (MSA)

➢ By the use of an energy separation Agent (ESA)

2.4.0 Mass Separation Agent

Separation by mass separation agent is achieved by controlling the composition or concentration of the hydrocarbons to be recovered by using;

➢ Adsorption or

➢ Absorption

2.4.1 Adsorption (Solid Bed Adsorption)

This method involves the use of solid adsorbents which have the ability to adsorb heavy hydrocarbons from the natural gas stream. The adsorbent mostly used are Silica gel or activated charcoal. The adsorption process provides a new surface area, on the surface of the solid material, to ''adsorb'' the components to be recovered and separated as NGL. The desired liquid components are entrained on the surface of the solid adsorbent, and are regenerated off in a high concentration; hence, their condensation efficiency is enhanced [Abdel Aal et al, 2003]. Figure 2.2 below shows an example of a solid bed adsorption schematic.
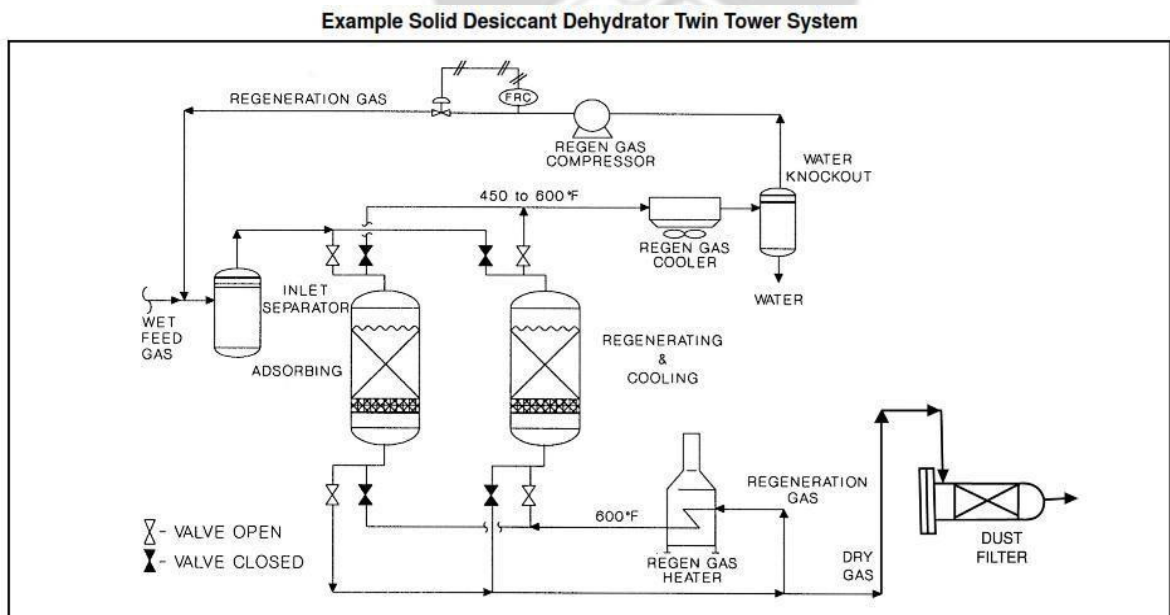
Figure 2.2 Solid Desiccant Dehydrator Twin Tower System (Source GPSA, 2011)

2.4.2 Absorption (Lean Oil Absorption)

Absorption is comparatively the oldest and least efficient method of recovering liquids from raw gas streams. The gas being processed is contacted in an absorption column normally kept at ambient temperature and a pressure close to the sales gas pressure with an absorption oil (lean oil), which absorbs the most heavy hydrocarbons (C_3-C_7) from the natural gas stream preferentially [Abdel Aal et al, 2003]. The process is as shown in figure 2.3 below.
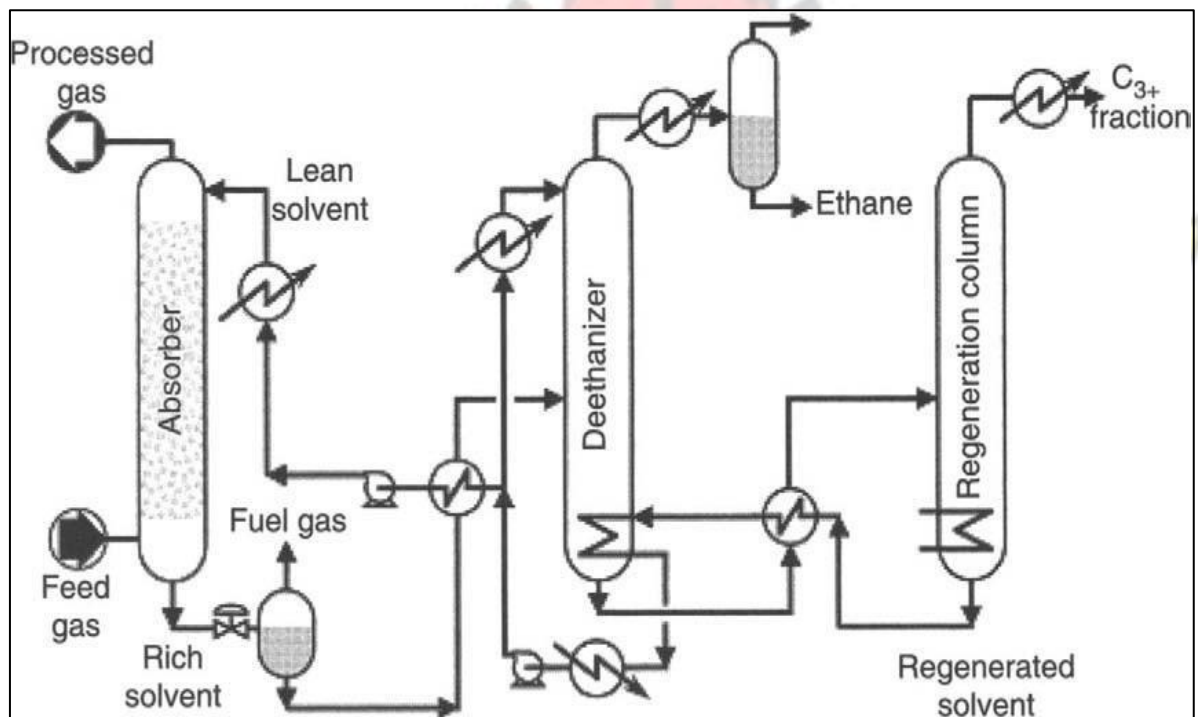
Figure 2.3 Schematic of a lean absorption system (Source GPSA, 2011)

2.5.0 Energy Separation Agent

This method is achieved by removing or applying mostly heat energy from the gas stream, to obtain a phase change. To achieve NGL recovery, pressure is kept constant whilst temperature is decreased by refrigeration using one of the following techniques;

- ➢ Mechanical Refrigeration
- ➢ Expansion Refrigeration

2.5.1 Mechanical Refrigeration

A refrigeration system is used to lower the temperature of a system below that normally achievable with air or water. A mechanical refrigeration system is essentially a heat pump where heat is removed from the process fluid and 'pumped' to the environment. Work is required to pump heat and the amount of work required depends on the temperature differential. Compression refrigeration is the most common example of mechanical refrigeration systems [Ben Asante, 2013]. Figure 2.4.0 below, shows the schematic of a mechanical refrigeration system.
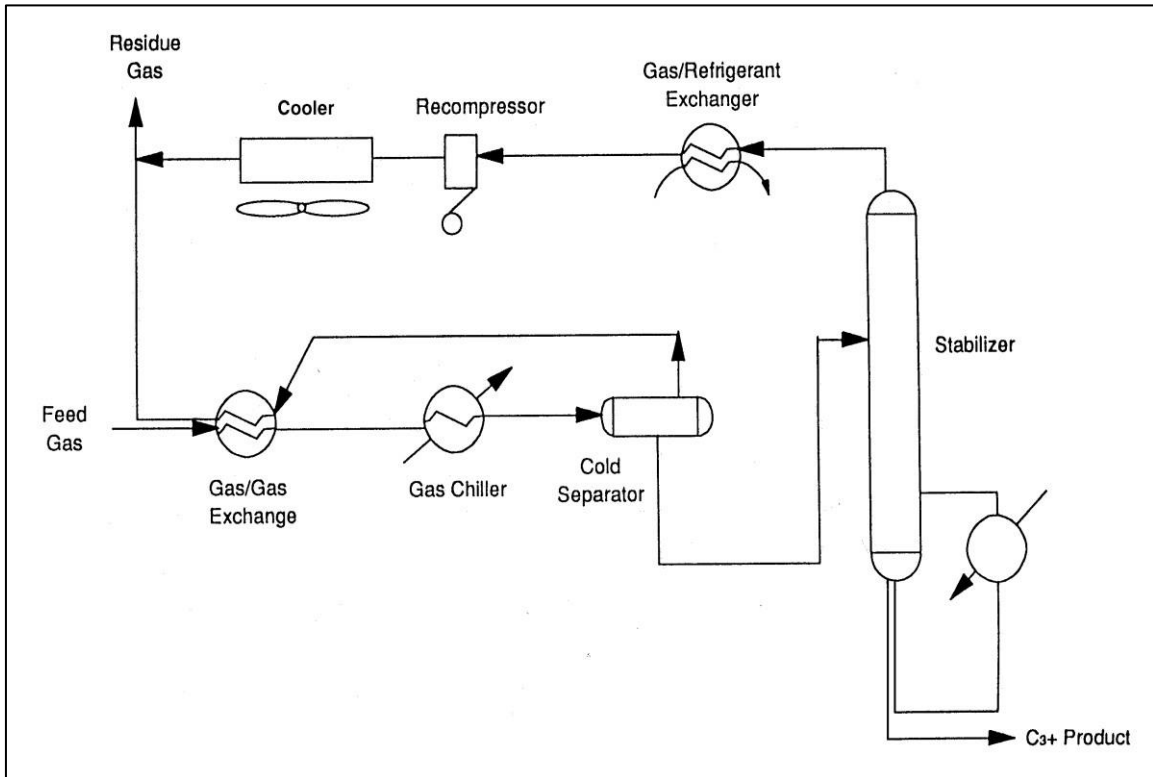
Figure 2.4.0 Schematic of a mechanical refrigeration system (Source; GPSA 11th edition, 2011)

2.5.2 Expansion refrigeration (Cryogenic Refrigeration)

This method is based on Joule-Thompson effect which states that, when a non-ideal gas is expanded by reducing the pressure form a high pressure region to a low pressure, there is a corresponding change in temperature. In general, the Gas processing industry considers cryogenic processing to be processes which are carried out below –50°F [GPSA]. To achieve these temperatures, a combination of pressure expansion and

chilling is utilized.

In the gas processing industry, two major types of expansion refrigeration are employed, and they are;

➢ Valve Expansion (Joule Thompson valve)

➢ Turbine Expansion (Turbo Expander)

2.5.2.1 Valve Expansion

Valve expansion technology has been used for over 60 years. It is relatively simple but less efficient (40-60% liquid recovery) as compared to turbine expansion but, very effective for dew point control. The general concept is to chill the gas stream by expanding the gas across the J-T valve. With the appropriate heat exchange mechanism and a substantial pressure differential normally about 100 bar minimum across the J-T valve; cryogenic temperatures can be achieved resulting in high extraction efficiencies. The expansion valve is normally placed in series with a Low Temperature Separator or a Low Temperature Exchanger. Valve expansion is thermodynamically isenthalpic and no work is done on or by the process fluid. The liquid-rich gas is normally heated in a gasgas exchanger where it exchanges heat with the sales gas, and hydrate control is achieved by using methanol or glycol. NGLs and water-rich inhibitor are removed from the process stream in the cold separator. Figure.2.5.0 below shows the schematic of a JT valve in series with a condensate stabilizer.
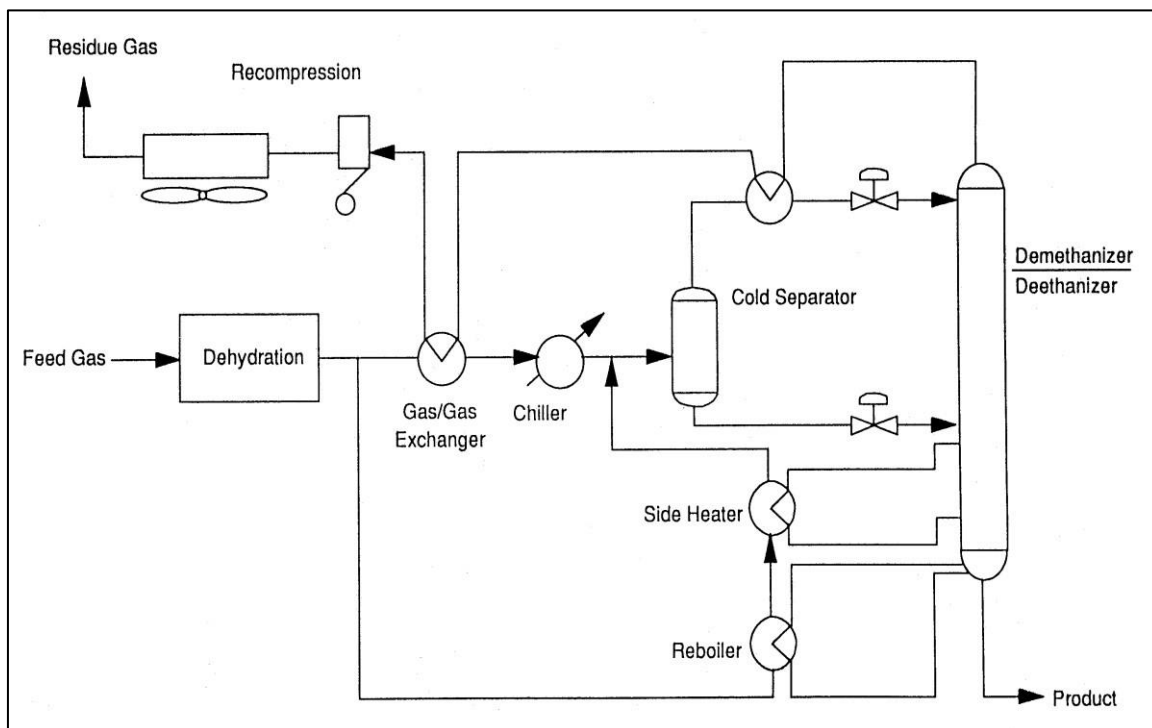
Figure 2.5 Schematic of a JT Valve in series with a condensate stabilizer (Source; GPSA 11th edition, 2011)

2.5.2.2 Turbo Expansion

This is a process whereby the feed gas pressure is used to produce refrigeration by expanding it across a turbine. The turbo expander expands and recovers useful work from the gas expansion and the useful work used to drive a centrifugal compressor to recompress the residual gas from the process. Because the expansion process is near isentropic, the turbo-expander reduces the gas temperature significantly and more than the use of expansion across a J-T valve. The process as was originally conceived, utilized a top feed, and non-refluxed de-methanizer. As the desire increases for higher recovery levels, different designs have been developed. The aim of these new designs is to produce reflux for the de-methanizer to achieve lower overhead temperatures and higher ethane recovery. The turbo expander process has been used on a wide range of process conditions and, in

addition to Ethane recovery projects; it is used often as a process for high Propane recovery. The process can be adapted to switch from recovering ethane to ethane rejection operation with very small changes in operation. Figure 2.6 shows the schematic of a conventional turbo expansion process
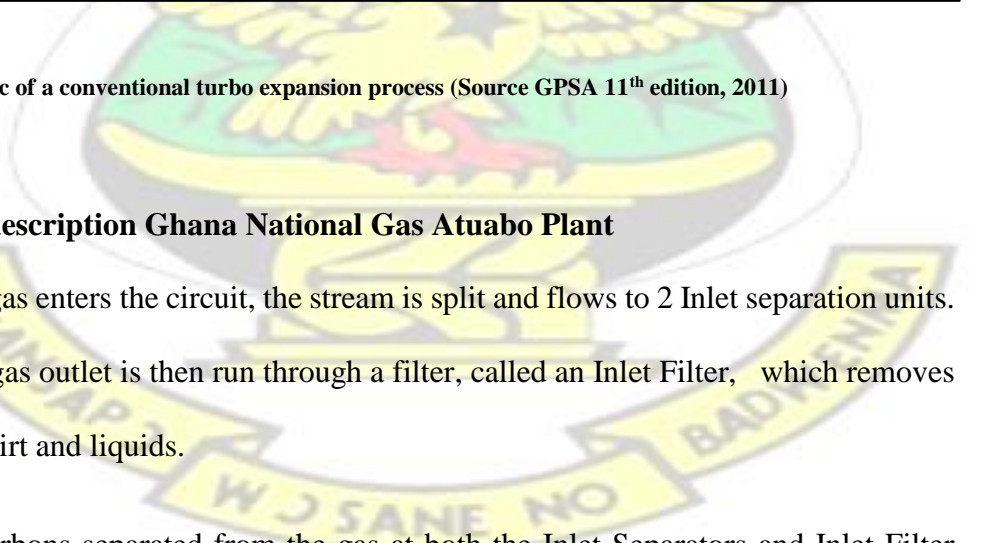
Figure 2.6 Schematic of a conventional turbo expansion process (Source GPSA 11th edition, 2011)

2.6.0 Process description Ghana National Gas Atuabo Plant

When the raw gas enters the circuit, the stream is split and flows to 2 Inlet separation units. The separated gas outlet is then run through a filter, called an Inlet Filter, which removes any entrained dirt and liquids.

Liquid hydrocarbons separated from the gas at both the Inlet Separators and Inlet Filter flow to the De-ethanizer. The flow is controlled by two level control valves, which dictate the hydrocarbon liquid level in the Inlet Separator and filter. In the Inlet Separator, any

water accompanied with the raw gas settles at the boot. After the water attains a high level, it is discharge into the oily water sump for treatment. Upon realizing a high water content in the raw gas, Methanol is dozed and injected at the upstream of the Inlet Separator. The methanol can be directed to any point of hydrate formation.

From the Inlet filter, the gas splits into two and goes through the Gas/Gas Exchanger and Gas/Liquid Exchanger. The two streams then commingle, after exiting both heat exchangers, at a temperature of -14.5 ºC and further flow through a JT valve, where it goes through isenthalpic expansion to 51 bar (g) and -37 ºC. Hydrates are prevented from forming by the injection of ethylene glycol solution onto the Gas/Gas exchanger and Gas/liquid exchanger tube sheet faces, and upstream of the J-T choke valves, with an injection quill in the piping. The gas stream moves to a 3-Phase separator known as Low Temperature Separator, which further separates gaseous stream from condensed liquids. The condensed liquids are further sent to the shell of the Gas/Liquids Exchanger, then to the De-ethanizer on hydrocarbon level control.

The Low Temperature Separator is equipped with an aqueous phase recovery boot, where rich ethylene glycol hydrate inhibitor is separated and sent to regeneration.

The rich Ethylene Glycol solution received from the Low Temperature separator is forwarded to the Ethylene Glycol regeneration unit where the Glycol is pre-heated, filtered, and flashed for any residual hydrocarbon. It is then reboiled in the glycol regeneration reboiler to 70% W/W, cooled, and then pumped by a positive displacement pump at the pressure of the Inlet filter. Prior to the reboiling, the low temperature rich Ethylene Glycol from the Low Temperature separator undergoes two stages of preheating before it is flashed

at 5 bar(g) pressure. The first preheating occurs at the regenerator reflux coil and the second preheating takes place as it passes through the surge tank coils. The flushing occurs in the Glycol flash drum at the above mentioned pressure. Entrained hydrocarbons are flushed in this drum and further sent to the flare. From the Glycol Flash drum, the glycol is passes through the Glycol Particulate Filter and Glycol Charcoal Filter respectively before it ends into the Glycol Regenerator. The regenerator consists of the reboiler, a bath-type exchanger that boils excess water out of the glycol. The water-glycol vapor mixture is treated in the still column by condensing the entrained glycol in the vapor and further dropping the condensed glycol to the reboiler shell. This condensation is provided by the cold rich glycol that flows through the reflux coil in the top of the still column. The produced water vapor is vented from the top of the still column. In the reboiler the glycol gets into the surge tank by overflowing its stand pipe. The produced lean glycol is further cooled by the rich glycol present in the surge tank coil. Re-boiling heat is provided with the Glycol Regenerator Tube Bundle, which uses heat media, flowing through a two-pass multi-tube bundle. Lean glycol exits the regenerator and is further cooled with finned pipe. The lean glycol is then pumped to a high pressure, as the pressure from the inlet filter, with the Glycol pumps, and sent to the lean glycol header tank for injection into the gas upstream of the gas/gas and gas/liquids heat exchangers.

The gas from the LTS goes back to the gas/gas exchanger and heated to 13 °C before exporting it to a gas metering station, and the plant's battery limit. Gas pressure will be 50 bar (gauge) at the plant battery limit.

The Deethanizer is a top feed, non-refluxed trayed tower. The Deethanizer liquids are reboiled with the Deethanizer Reboiler, a heat exchanger using heat medium from the plant hot oil supply. The bottoms product purity is controlled by reboiler outlet temperature and the Deethanizer is held at a constant pressure using a pressure control valve on the outlet gas stream. The Deethanizer operates at 35 bar (gauge). The C_3+ bottoms product at 165 ℃ is fed to a Debutanizer.

The deethanizer overheads flow to the deethanizer over-heads scrubber, which removes any entrained liquids prior to compression. The gas after existing the scrubber goes to the suction compartment of the deethanizer overheads compressor at a pressure of 32.5barg, where compression is done to attain sales gas pressure of 50barg.

The deethanizer bottoms feed the debutanizer unit which fractionates the C_3^+ to produce the specified LPG product in the overhead. The bottoms of the debutanizer product contains too much pentane to meet the Residual Vapor Pressure specification for stable condensate, so the Debutanizer bottoms product is fed to a pentane stripper column, which strips C_5 from the column feed such that, the bottom of the stripper product will be at a stable condensate residual vapor pressure of 0.65kg/cm2. Figure 2.7.0 below shows the schematic of the Ghana Gas Atuabo Plant.

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2.7.0 Ghana Gas Atuabo Product specification

Table 1.0.0 below shows Ghana National Gas Atuabo plant Product specifications the primary objective is to produce sales gas for Aboadze thermal plant.

Table 1 Atuabo plant final product specifications

SPECIFICATION	VALUE
Sales Gas Specification	
Water dewpoint	< 4LB/MMSCF
Hydrocarbon Dewpoint	< -26.6 C @ 50 barg
Hydrogen sulphide	< 4 ppm vol
Total Sulphur	< 28 ppm
CO2	< 2.0 mol%
Nitrogen	< 1.0 mol%
Wobbe Index	46-52 MJ/m3
Higher Heating Valve	<1250 Btu/SCF
Sales Gas Temperature	< 50 C
Sales Gas Pressure	50 barg minimum
LPG Specification	
TVP @ 37.8 C	< 9.5 kg/cm2
C2/C3 content	< 1.5 vol% expected
C5+ content	< 2.0 vol%
95% vol evaporated point	-2 C min, +2 C max.
Total Sulphur	< 200 ppm max.
Hydrogen sulphide	< 20 ppm max
Mercaptans	< 20 ppm max
CONDENSATE Specification	
RVP	< 0.65 kg/cm2
INITIAL BPT	70 C max
FINAL BPT	110 C max

(Source: Ghana National Gas Company Technical Directorate, 2014)

CHAPTER 3

Methodology

3.0. Composition

The composition of the raw gas stream to be fed to the Atuabo Gas plant from the Jubilee fields used in this study is as shown in Table 2.0.0. For the purpose of this project, it was assumed that there is not much variation in the feed composition to the plant. The feed composition was conveniently kept constant because the method of Natural gas liquids recovery employed is by energy separation that is Joule Thompson and Turbo Expansion.

Table 2 Raw gas composition, Ghana gas Atuabo plant feed

Composition	Raw gas mol%	Composition	Raw gas mol%
C1	78.127	CO_2	1.442
C2	8.943	H_2O	0.00422
C3	7.241	Benzene	0
C4	0.892	Toluene	0
C4	2.188	E-Benzene	0
C5	0.453	P-Xylene	0
C5	0.4	2-Mhexane	0
C6	0.073	M cyclopentane	0
C7	0	Cyclohexane	0
N2	0.241	M cyclohexane	0

(Source Ghana: National Gas Company Technical directorate 2014)

3.1.0 Flowrates

The Flowrates were varied according to projected yearly flows to estimate recoveries from the turbo expander instead and JT valve. Table 3.0 below shows projected flow rates for the entire life span of the project used for this study

Table 3 Projected Natural gas Flowrates to Atuabo plant

YEAR	FLOW FORECAST
2014	120MMSCFD
2015	150MMSCFD
2016	300MMSCFD
2017	300MMSCFD
2018	450MMSCFD
2019	450MMSCFD
2020	560 MMSCFD
2025	640MMSCFD
2030	700MMSCFD

(Source: Ghana National Gas Company Technical directorate, 2014)

3.2.0. Stripping calculations

Stripping calculations were done using an Excel based program, 'StrippingCalc' provided by the Ghana Gas Technical directorate. This software mainly requires inputs for the composition and flow rate to determine the yield of each component and subsequently the total natural gas liquids yield. Figure 3.1 below illustrates the stripping calculations

JUBILEE FIELD LIQUIDS EXTRACTION

OUTPUT STREAM

Sales Gas   103.71
mmcfd

c5+   Condensates   1569.55
bpd

6185.51   C3   6185.51
c3   bpd

RICH GAS
INPUT

120 mmcfd

2510.68
ic4   C4   3806.44
bpd

Expander

Total LPGs   9991.9
Total Liquids   11561.5 bpd

Fractionating Tower
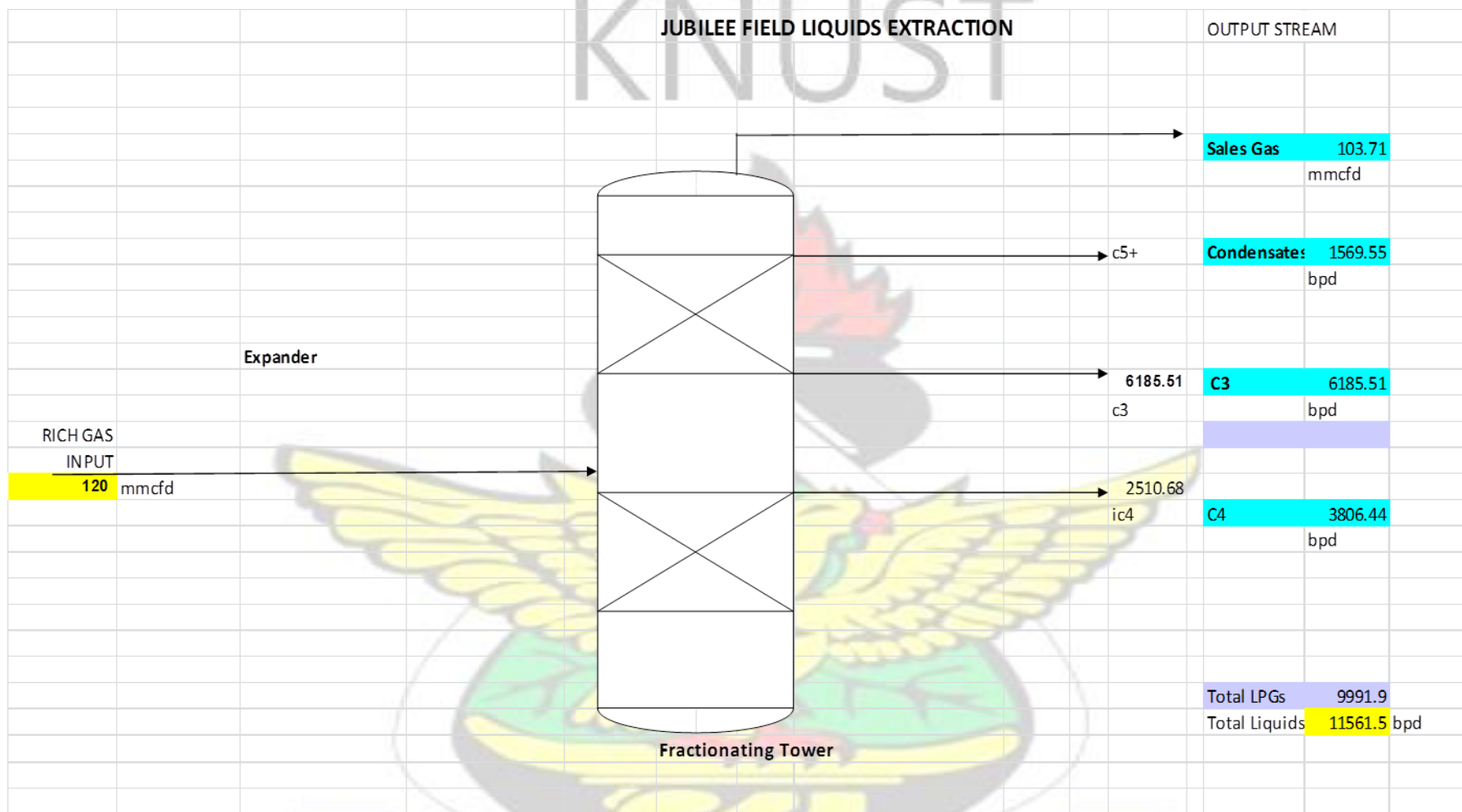
Figure 3.1 Schematic of stripping calculations

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CHAPTER FOUR

4.0 RESULTS AND DISCUSSION

This chapter discusses the results obtained from running the stripping calculations for the various flows forecasted for the life of the plant and an economic analysis.

On running strip calculations, a flow of 120mmscfd resulted in a total liquid recovery of 7054.01bbl/day on the JT valve whilst the turbo expander produced 9167.83bbl/day. On increasing the flow to 150mmscfd, the recovery rose to 8817.51bbl/day on the JT valve and 11459.78bbl/day on the Turbo expander.

A flow of 300mmscfd resulted in a total of 17635.02bbl/day of natural gas liquids being recovered with the JT valve and 22919.57bbl/day with the Turbo expander. On running 450mmscfd, we obtained 26452.54bbl/day of liquids from the JT valve and 34379.35bbl/day from the Turbo expander. 560mmscfd flow resulted in 32918.71bbl/day with the JT valve and 42783.19bbl/day on the Turbo expander.

Finally a flow of 640mmscfd resulted in 37621.39bbl/day and 48895.08bbl/day on the JT valve and Turbo expander respectively, with 700mmscfd producing 41148.39bbl/day and 53478.99bbl/day on the JT valve and Turbo expander respectively

The difference in recovery between the JT valve and the Turbo expander, running at 120mmscfd and 150mmscfd was 2,113.82bbl/d. and 2,642.27bbl/d respectively. When the flows were increased to 300mmscfd and 450mmscfd, the difference in recovery were

5,284.55bbl/d and 7,926.81bbl/d respectively. Running at 560 and 640mmscfd, the difference recorded was 9,864.48bbl/d and 11,273.69bbl/d respectively. 700mmscfd produced a difference of 12,330.60bbl/d of natural gas liquids. In all there is a 30% increase in recovery when the Turbo expander is used instead of the JT valve. Table 4.0.0 Figure 4.1 and below, shows the results obtained from stripping calculations

Table 4 Results of the stripping calculations

FLOWS (MMSCFD)	JT VALVE(bbl/d)	Turbo Expander (bbl/d)	Variance (bbl/d)
120	7,054.01	9,167.83	2,113.82
150	8,817.51	11,459.78	2,642.27
300	17,635.02	22,919.57	5,284.55
300	17,635.02	22,919.57	5,284.55
450	26,452.54	34,379.35	7,926.81
450	26,452.54	34,379.35	7,926.81
560	32,918.71	42,783.19	9,864.48

640	37,621.39	48,895.08	11,273.69
700	41,148.39	53,478.99	12,330.60
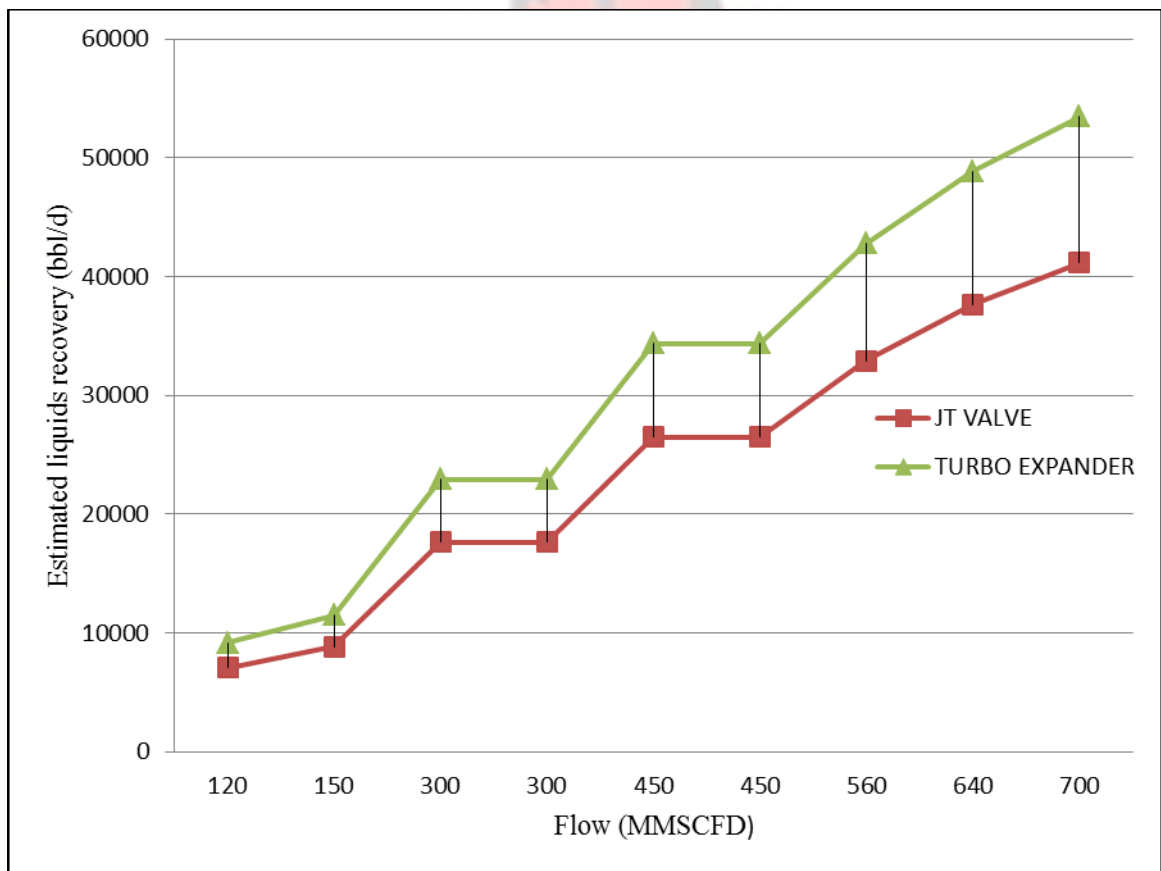
Figure 4.1 A graph showing estimated liquids recovery against projected flow rates

4.1.0 Economic Analysis

The Natural gas liquids recovered are sold separately from the lean gas. VRA is only interested in the lean gas but there is already a vast market for the LPG's which are obtained from the NGL's. Earlier this year, there were shortages of this product which is used domestically and now being adopted for commercial vehicles.

In using a JT valve, there are massive loses in NGL revenue as a substantial amount of the NGL's remain in the Lean gas stream creating a rich-lean gas but sold at the price of a normal lean gas as per contract with VR.A. Table 5.0 below shows the amount of revenue to be generated from the NGL, plant was assumed to have a 90% availability and price of NGL on the market $ 39 per barrel.

The CAPEX for the expander process as per the quote received from contactors are as follows;

Gas dehydration Unit (Mol sieve)   = USD$14.6 million

Turbo Expander unit   = USD$6.85million

Total cost   =USD$21.45million

Table 5 Comparison of quantity of NGL produced and revenue generated with the JT valve and Turbo Expander

Flows (MMSCFD)	JT Valve(bbl/d)	Total Yearly Revenue($)	Turbo Expander (bbl/d)	Total Yearly Revenue($)	Variance (bbl/d)	Total Yearly revenue ($)	% Increment in Recovery
120	7,054.01	$ 90,372,449.12	9,167.83	$ 117,453,654.05	2,113.82	$ 27,081,204.93	29.9662178
150	8,817.51	$ 112,965,529.37	11,459.78	$ 146,816,971.47	2,642.27	$ 33,851,442.11	29.96616959
300	17,635.02	$ 225,931,058.73	22,919.57	$ 293,634,071.06	5,284.55	$ 67,703,012.33	29.96622629
300	17,635.02	$ 225,931,058.73	22,919.57	$ 293,634,071.06	5,284.55	$ 67,703,012.33	29.96622629
450	26,452.54	$ 338,896,716.21	34,379.35	$ 440,451,042.53	7,926.81	$ 101,554,326.32	29.96615826
450	26,452.54	$ 338,896,716.21	34,379.35	$ 440,451,042.53	7,926.81	$ 101,554,326.32	29.96615826
560	32,918.71	$ 421,738,053.17	42,783.19	$ 548,116,838.69	9,864.48	$ 126,378,785.52	29.96618033
640	37,621.39	$ 481,986,437.99	48,895.08	$ 626,419,317.42	11,273.69	$ 144,432,879.44	29.96617084
700	41,148.39	$ 527,172,598.49	53,478.99	$ 685,146,080.39	12,330.60	$ 157,973,481.90	29.96617851
		$ 2,763,890,618.00		$ 3,592,123,089.17		$ 828,232,471.17	

At a flow of 120mmscfd, the total revenue to be generated from the NGL's is $90,372,449.12 with the JT valve running and $117,453,654.05 with the Turbo expander. The total loss of revenue will be $27,081,204.93 per anum.

A flow of 150mmscfd generates yearly revenue of $112,965,529.37 with the JT valve and $146,816,971.47 with the Turbo expander, resulting in an annual revenue difference of $33,851,442.11.between the JT valve and the Turbo Expander.

Pushing the flow up to 300mmscfd leads to a total annual revenue of $225931058.73 and $293634071.06 for the JT valve and Turbo Expander respectively, with a difference of $67703012.33.

450mmscfd will generate $338896716.21 from the JT valve and $440451042.53 from the Turbo Expander giving a difference of $101554326.32 in revenue between the two for the year. A flow of 560mmscfd also generates revenue of $421,738,053.17 and $548116838.69 from the JT valve and Turbo Expander respectively, with a difference of $126378785.52 annually.

At 640mmscfd, the expected revenue is $481986437.99 and 626419317.42 for the JT valve and Turbo expander respectively with a difference of $114432879.44.

Finally when we attain a flow of 700mmscfd, the revenue expected form selling natural gas liquids generated from the JT valve is $527172598.49 and $685146080.39 from the Turbo Expander, a difference of $157973481.90 annually.

4.2.0 Conclusion

Gas production from the Jubilee fields is projected to increase from 120mmscfd in 2014 to 700mmscfd in 2030. The results show that by running a JT valve throughout the lifetime of the project will lead to a massive loss of $828,232,471.17 in revenue, using the current natural gas liquid market price of $39 and plant availability of 90%. The stripping calculations has shown that there will be about 30% overall improvement in the recovery of NGL's when a turbo expander is used instead of a JT valve. It will therefore make a lot of economic sense to add the expander process at a cost of USD$21.45 million with one year expected payback.

4.3.0 Recommendations

I would recommend that, the plant installs a turbo expander after the following;

1) Further study to be conducted taking into consideration temperature and pressure variances over the years using HYSYS

2) Feasibility of installing the Turbo expander in series or parallel with the existing setup

3) A comparison of the operating and maintenance cost between the JT valve and the Turbo expander.

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Appendix A

Stripping calculations for the JT valve

Table A-1

	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio, scf/gal	Volume Stripped barrels/day strpped		
Flow, mmcfd	120	Initial Comp.				Separation Residual Comp.			Resid Flow Rate	110.14	MMcfd
										9.86	MMcfd
N	0.241	0.002	0	0.2410	0.000	0.241	0.263	91.13	0.00		
co2	1.442	0.014	0	1.4420	0.000	1.442	1.571	59.1	0.00		
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127	0.781	0	78.1270	0.000	78.127	85.117	59.14	0.00		
C2	8.943	0.089	0	8.9430	0.000	8.943	9.743	37.5	0.00		
C3	7.241	0.072	60	7.2410	4.345	2.896	3.156	36.4	3409.83		
iC4	0.892	0.009	85	0.8920	0.758	0.134	0.146	30.64	706.92		
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	1966.23		
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	472.54		
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	412.94		
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	85.55		
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004	1.000		100.004	8.217	91.787	100.000		7054.01	bbls/D(Total)	

Table A-2 1

Flow, mmcfd		150 Initial Comp.				Separation Residual Comp.		**Resid Flow Rate**	137.68 MMcfd
	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio, Volume Stripped	12.33 MMcfd
								scf/gal barrels/day strpped	
N	0.241	0.002	0	0.2410	0.000	0.241	0.263 91.13	0.00	
co2	1.442	0.014	0	1.4420	0.000	1.442	1.571 59.1	0.00	
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005 175.6	0.00	
C1	78.127	0.781	0	78.1270	0.000	78.127	85.117 59.14	0.00	
C2	8.943	0.089	0	8.9430	0.000	8.943	9.743 37.5	0.00	
C3	7.241	0.072	60	7.2410	4.345	2.896	3.156 36.4	4262.29	
iC4	0.892	0.009	85	0.8920	0.758	0.134	0.146 30.64	883.65	
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000 31.79	2457.79	
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000 27.39	590.67	
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000 27.68	516.18	
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000 24.38	106.94	
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000 24.16	0.00	
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000 24.56	0.00	
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000 28.33	0.00	
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000 35.83	0.00	
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000 29.45	0.00	
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000 21.57	0.00	
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000 21.84	0.00	
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000 24.94	0.00	
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000 29.94	0.00	
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000 25.98	0.00	
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000 25.8	0.00	
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000 26.36	0.00	
H	0.000	0.000	0	0.0000	0.000	0.000	0.000 111.2	0.00	
He	0.000	0.000	0	0.0000	0.000	0.000	0.000 98.69	0.00	
O	0.000	0.000	0	0.0000	0.000	0.000	0.000 112.9	0.00	
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000 74.51	0.00	
Total	100.004	1.000		100.004	8.217	91.787	100.000	8817.51 bbls/D(Total)	

Table A-3

	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio, scf/gal	Volume Stripped barrels/day strpped	
Flow, mmcfd	300	Initial Comp.					Separation Residual Comp.		Resid Flow Rate	275.36 MMcfd
										24.65 MMcfd
N	0.241	0.002	0	0.2410	0.000	0.241	0.263	91.13	0.00	
co2	1.442	0.014	0	1.4420	0.000	1.442	1.571	59.1	0.00	
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00	
C1	78.127	0.781	0	78.1270	0.000	78.127	85.117	59.14	0.00	
C2	8.943	0.089	0	8.9430	0.000	8.943	9.743	37.5	0.00	
C3	7.241	0.072	60	7.2410	4.345	2.896	3.156	36.4	8524.57	
iC4	0.892	0.009	85	0.8920	0.758	0.134	0.146	30.64	1767.30	
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	4915.57	
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	1181.35	
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	1032.35	
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	213.88	
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00	
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00	
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00	
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00	
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00	
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00	
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00	
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00	
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00	
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00	
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00	
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00	
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00	
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00	
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00	
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00	
Total	100.004	1.000		100.004	8.217	91.787	100.000		17635.02 bbls/D(Total)	

Table A-4

	Flow, mmcfd	450	Initial Comp.				Separation	Residual Comp.		**Resid Flow Rate**	413.04	MMcfd
	Feed %		x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped	36.98	MMcfd
									scf/gal	barrels/day strpped		
N	0.241		0.002	0	0.2410	0.000	0.241	0.263	91.13	0.00		
co2	1.442		0.014	0	1.4420	0.000	1.442	1.571	59.1	0.00		
H2O	0.00422		0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127		0.781	0	78.1270	0.000	78.127	85.117	59.14	0.00		
C2	8.943		0.089	0	8.9430	0.000	8.943	9.743	37.5	0.00		
C3	7.241		0.072	60	7.2410	4.345	2.896	3.156	36.4	12786.86		
iC4	0.892		0.009	85	0.8920	0.758	0.134	0.146	30.64	2650.95		
nC4	2.188		0.022	100	2.1880	2.188	0.000	0.000	31.79	7373.36		
iC5	0.453		0.005	100	0.4530	0.453	0.000	0.000	27.39	1772.02		
nC5	0.4		0.004	100	0.4000	0.400	0.000	0.000	27.68	1548.53		
c6	0.073		0.001	100	0.0730	0.073	0.000	0.000	24.38	320.81		
2m-C5	0		0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0		0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0		0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0		0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0		0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0		0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0		0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0		0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0		0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0		0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0		0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0		0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000		0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000		0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000		0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000		0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004		1.000		100.004	8.217	91.787	100.000		26452.54 bbls/D(Total)		

Table A

5

	Feed %	x1	%Removal	Initial moles	Moles rem	moles after	x2	Ratio, scf/gal	Volume Stripped, barrels/day strpped
Flow, mmcfd 560	Initial Comp.					Separation Residual Comp.		**Resid Flow Rate** 514.01 MMcfd	46.01 MMcfd
N	0.241	0.002	0	0.2410	0.000	0.241	**0.263**	91.13	0.00
co2	1.442	0.014	0	1.4420	0.000	1.442	**1.571**	59.1	0.00
H2O	0.00422	0.000	0	0.0042	0.000	0.004	**0.005**		0.00
C1	78.127	0.781	0	78.1270	0.000	78.127	**85.117**	175.6	0.00
C2	8.943	0.089	0	8.9430	0.000	8.943	**9.743**	59.14	0.00
								37.5	
C3	7.241	0.072	60	7.2410	4.345	2.896	**3.156**	36.4	15912.54
iC4	0.892	0.009	85	0.8920	0.758	0.134	**0.146**	30.64	3298.96
nC4	2.188	0.022	100	2.1880	2.188	0.000	**0.000**	31.79	9175.74
iC5	0.453	0.005	100	0.4530	0.453	0.000	**0.000**	27.39	2205.18
nC5	0.4	0.004	100	0.4000	0.400	0.000	**0.000**	27.68	1927.06
c6	0.073	0.001	100	0.0730	0.073	0.000	**0.000**	24.38	399.23
2m-C5	0	0.000	100	0.0000	0.000	0.000	**0.000**	24.16	0.00
			100	0.0000	0.000	0.000	**0.000**	24.56	0.00
3m-C5	0	0.000	100	0.0000	0.000	0.000	**0.000**		0.00
			100	0.0000	0.000	0.000	**0.000**		0.00
m-c-C5	0	0.000	100	0.0000	0.000	0.000	**0.000**	28.33	0.00
			100	0.0000	0.000	0.000	**0.000**		0.00
Benzene	0	0.000	100	0.0000	0.000	0.000	**0.000**	35.83	0.00
			0	0.0000	0.000	0.000	**0.000**		0.00

Table A-

c-C6	**0**	0.000	**0**	0.0000	0.000	0.000	**0.000**	29.45	0.00		
			0	0.0000	0.000	0.000	**0.000**		0.00		
			0	0.0000	0.000	0.000	**0.000**		0.00		
2-m-C6	**0**	0.000		0.0000	0.000	0.000	**0.000**	21.57			
3-m-C6	**0**	0.000						21.84			
m-c-C6	**0**	0.000						24.94			
Toluene	**0**	0.000						29.94			
et-Benzene	**0**	0.000						25.98			
p-Xylene	**0**	0.000						25.8			
o-Xylene	**0**	0.000						26.36			
								111.2			
H	**0.000**	0.000						98.69			
								112.9			
He	**0.000**	0.000						74.51			
O	**0.000**	0.000									
H2s	**0.000**	0.000									
Total		100.004	1.000		100.004	8.217	91.787	100.000		**32918.71** bbls/D(Total)	

Table A

-6 1

	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio, scf/gal	Volume Stripped barrels/day strpped		
Flow, mmcfd	640	Initial Comp.				Separation	Residual Comp.		Resid Flow Rate	587.44	MMcfd
										52.59	MMcfd
N	0.241	0.002	0	0.2410	0.000	0.241	0.263	91.13	0.00		
co2	1.442	0.014	0	1.4420	0.000	1.442	1.571	59.1	0.00		
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127	0.781	0	78.1270	0.000	78.127	85.117	59.14	0.00		
C2	8.943	0.089	0	8.9430	0.000	8.943	9.743	37.5	0.00		
C3	7.241	0.072	60	7.2410	4.345	2.896	3.156	36.4	18185.76		
iC4	0.892	0.009	85	0.8920	0.758	0.134	0.146	30.64	3770.24		
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	10486.55		
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	2520.21		
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	2202.36		
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	456.27		
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004	1.000		100.004	8.217	91.787	100.000		37621.39	bbls/D	(Total)

Table A-

KNUST

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	Flow, mmcfd	700 Initial Comp.				Separation Residual		Comp.	Resid Flow Rate	642.51	MMcfd
	Feed %	x1	%Removal	nitial moles	Moles rem	omoles after	x2	Ratio,	Volume Stripped	57.52	MMcfd
								scf/gal	barrels/day strpped		
N	0.241	0.002	0	0.2410	0.000	0.241	0.241	91.13	0.00		
			0				0.263 1.4420	59.1	0.00		
co2	1.442	0.014	0		0.000	1.442			0.00		
			0			1.571 0.0042			0.00		
			0		0.000	0.004			0.00		
H2O	0.00422	0.000				0.005		175.6			
				78.1270	0.000	78.127 85.117		59.14			
C1	78.127	0.781		8.9430	0.000	8.943 9.743		37.5			
C2	8.943	0.089									
C3	7.241	0.072	60	7.2410	4.345	2.896		36.4	19890.67		
			85			3.156 0.8920			4123.70		
iC4	0.892	0.009	100		0.758	0.134		30.64	11469.67		
			100			0.146 2.1880			2756.48		
nC4	2.188	0.022			2.188	0.000		31.79			
						0.000					
iC5	0.453	0.005		0.4530	0.453	0.000 0.000		27.39			
nC5	0.4	0.004	100	0.4000	0.400	0.000		27.68	2408.83		
			100			0.000 0.0730			499.04		
c6	0.073	0.001	100		0.073	0.000		24.38	0.00		
			100			0.000 0.0000		24.16	0.00		
			100		0.000	0.000		24.56	0.00		
2m-C5	0	0.000	100			0.000 0.0000			0.00		
			100		0.000	0.000			0.00		
3m-C5	0	0.000	100			0.000 0.0000			0.00		
			100		0.000	0.000			0.00		
			100						0.00		
m-c-C5	0	0.000	100			0.000 0.0000		28.33	0.00		
			100		0.000	0.000			0.00		
Benzene	0	0.000	100			0.000 0.0000		35.83	0.00		
					0.000	0.000			0.00		

Table A

KNUST

c-C6	0	0.000	100 0 0 0 0	0.000	0.0000 0.000	29.45	0.00 0.00
2-m-C6	0	0.000		0.000 0.000	0.0000	21.57	0.00 0.00
3-m-C6	0	0.000		0.000 0.000	0.0000	21.84	
m-c-C6	0	0.000		0.000 0.000	0.0000	24.94	
Toluene	0	0.000		0.000 0.000	0.0000	29.94	
et-Benzene	0	0.000		0.000 0.000	0.0000	25.98	
p-Xylene	0	0.000		0.000 0.000	0.0000	25.8	
o-Xylene	0	0.000		0.000 0.000	0.0000	26.36 111.2 98.69	
H	0.000	0.000		0.000 0.000	0.0000	112.9 74.51	
He	0.000	0.000	0.0000	0.000 0.000	0.000		
O	0.000	0.000					
H2s	0.000	0.000					
Total	100.004	1.000	100.004	8.217 91.787	100.000		**41148.39** bbls/D(Total)

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Table A-

KNUST

Appendix B

Stripping calculations for the Turbo Expander

Table B-1 1

Flow, mmcfd	120	Initial Comp.				Separation	Residual Comp.		Resid Flow Rate	106.94	MMcfd
	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped	13.06	MMcfd
									scf/gal	barrels/day strpped		
N	0.241	0.002	0	0.2410	0.000	0.241	0.270	91.13	0.00		
co2	1.442	0.014	0	1.4420	0.000	1.442	1.618	59.1	0.00		
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127	0.781	0	78.1270	0.000	78.127	87.666	59.14	0.00		
C2	8.943	0.089	0	8.9430	0.000	8.943	10.035	37.5	0.00		
C3	7.241	0.072	95	7.2410	6.879	0.362	0.406	36.4	5398.90		
iC4	0.892	0.009	100	0.8920	0.892	0.000	0.000	30.64	831.67		
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	1966.23		
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	472.54		
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	412.94		
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	85.55		
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004	1.000		100.004	10.885	89.119	100.000		9167.83	bbls/D(Total)	

2

Table B-1

Flow, mmcfd	150 Initial Comp.					Separation	Residual Comp.		Resid Flow Rate	133.68 MMcfd
	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped	16.33 MMcfd
								scf/gal	barrels/day strpped	
N	0.241	0.002	0	0.2410	0.000	0.241	0.270	91.13	0.00	
co2	1.442	0.014	0	1.4420	0.000	1.442	1.618	59.1	0.00	
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00	
C1	78.127	0.781	0	78.1270	0.000	78.127	87.666	59.14	0.00	
C2	8.943	0.089	0	8.9430	0.000	8.943	10.035	37.5	0.00	
C3	7.241	0.072	95	7.2410	6.879	0.362	0.406	36.4	6748.62	
iC4	0.892	0.009	100	0.8920	0.892	0.000	0.000	30.64	1039.59	
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	2457.79	
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	590.67	
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	516.18	
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	106.94	
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00	
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00	
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00	
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00	
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00	
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00	
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00	
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00	
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00	
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00	
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00	
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00	
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00	
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00	
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00	
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00	
Total	100.004	1.000		100.004	10.885	89.119	100.000		11459.78 bbls/D(Total)	

3

Table B-1

Flow, mmcfd	300	Initial Comp.				Separation	Residual Comp.	Resid Flow Rate		267.36	MMcfd
Feed %		x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped	32.65	MMcfd
								scf/gal	barrels/day strpped		
N	0.241	0.002	0	0.2410	0.000	0.241	0.270	91.13	0.00		
co2	1.442	0.014	0	1.4420	0.000	1.442	1.618	59.1	0.00		
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127	0.781	0	78.1270	0.000	78.127	87.666	59.14	0.00		
C2	8.943	0.089	0	8.9430	0.000	8.943	10.035	37.5	0.00		
C3	7.241	0.072	95	7.2410	6.879	0.362	0.406	36.4	13497.24		
iC4	0.892	0.009	100	0.8920	0.892	0.000	0.000	30.64	2079.18		
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	4915.57		
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	1181.35		
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	1032.35		
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	213.88		
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004	1.000		100.004	10.885	89.119	100.000		22919.57	bbls/D(Total)	

4

50

Table B-1

Flow, mmcfd	450	Initial Comp.				Separation	Residual Comp.	Resid Flow Rate		401.04	MMcfd
	Feed %	x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped	48.98	MMcfd
									scf/gal barrels/day strpped		
N	0.241	0.002	0	0.2410	0.000	0.241	0.270	91.13	0.00		
co2	1.442	0.014	0	1.4420	0.000	1.442	1.618	59.1	0.00		
H2O	0.00422	0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00		
C1	78.127	0.781	0	78.1270	0.000	78.127	87.666	59.14	0.00		
C2	8.943	0.089	0	8.9430	0.000	8.943	10.035	37.5	0.00		
C3	7.241	0.072	95	7.2410	6.879	0.362	0.406	36.4	20245.86		
iC4	0.892	0.009	100	0.8920	0.892	0.000	0.000	30.64	3118.76		
nC4	2.188	0.022	100	2.1880	2.188	0.000	0.000	31.79	7373.36		
iC5	0.453	0.005	100	0.4530	0.453	0.000	0.000	27.39	1772.02		
nC5	0.4	0.004	100	0.4000	0.400	0.000	0.000	27.68	1548.53		
c6	0.073	0.001	100	0.0730	0.073	0.000	0.000	24.38	320.81		
2m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00		
3m-C5	0	0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00		
m-c-C5	0	0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00		
Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00		
c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00		
2-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00		
3-m-C6	0	0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00		
m-c-C6	0	0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00		
Toluene	0	0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00		
et-Benzene	0	0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00		
p-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00		
o-Xylene	0	0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00		
H	0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00		
He	0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00		
O	0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00		
H2s	0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00		
Total	100.004	1.000		100.004	10.885	89.119	100.000		34379.35 bbls/D(Total)		

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Table B

	Flow, mmcfd	560	Initial Comp.				Separation Residual		Comp.	Resid Flow Rate	499.07	MMcfd
	Feed %		x1	**%Removal**	nitial moles	Moles rem	omoles after	x2	Ratio,	Volume Stripped	60.96	MMcfd
									scf/gal	**barrels/day strpped**		

Component	Feed %	x1	%Removal	Initial moles / Moles rem / moles after	x2	Ratio scf/gal	Volume Stripped barrels/day
N	0.241	0.002	0	0.2410 0.000 0.241	0.270 1.4420	91.13	0.00
			0	0.000 1.442 1.618 0.0042 0.000		59.1	0.00
co2	1.442	0.014	0	0.004 0.005 78.1270 0.000 78.127			0.00
			0	87.666			0.00
			0	8.9430 0.000 8.943	10.035		0.00
H2O	0.00422	0.000				175.6	
						59.14	
						37.5	
C1	78.127	0.781					
C2	8.943	0.089					
C3	7.241	0.072	95	7.2410 6.879 0.362		36.4	25194.85
			100	0.406 0.8920			3881.13
iC4	0.892	0.009	100	0.892 0.000		30.64	9175.74
			100	0.000 2.1880			2205.18
nC4	2.188	0.022		2.188 0.000		31.79	
				0.000			
iC5	0.453	0.005		0.4530 0.453 0.000 0.000		27.39	
nC5	0.4	0.004	100	0.4000 0.400 0.000		27.68	1927.06
			100	0.000 0.0730			399.23
c6	0.073	0.001	100	0.073 0.000		24.38	0.00
			100	0.000 0.0000		24.16	0.00
			100	0.000 0.000		24.56	0.00
2m-C5	0	0.000	100	0.000 0.0000			0.00
			100	0.000 0.000			0.00
3m-C5	0	0.000	100	0.000 0.0000			0.00
			100	0.000 0.000			0.00
m-c-C5	0	0.000	100	0.000 0.0000		28.33	0.00
			100	0.000 0.000			0.00
Benzene	0	0.000	100	0.000 0.0000		35.83	0.00
			100	0.000 0.000			0.00
c-C6	0	0.000	0	0.000 0.0000		29.45	0.00
			0	0.000 0.000			0.00
2-m-C6	0	0.000	0	0.000 0.0000		21.57	0.00
				0.000 0.000			
3-m-C6	0	0.000		0.000 0.0000		21.84	
				0.000 0.000			
m-c-C6	0	0.000		0.000 0.0000		24.94	
				0.000 0.000			
Toluene	0	0.000		0.000 0.0000		29.94	
				0.000 0.000			

Table B-1

et-Benzene	0	0.000			**0.000** 0.000	0.0000 0.000	25.98		6
p-Xylene	0	0.000			**0.000** 0.000	0.0000 0.000	25.8		
o-Xylene	0	0.000			**0.000** 0.000	0.0000 0.000	26.36		
H	0.000	0.000			**0.000** 0.000	0.0000 0.000	111.2 98.69		
He	0.000	0.000		0.0000 0.000	0.000 **0.000**		112.9 74.51		
O	0.000	0.000							
H2s	0.000	0.000							
Total	100.004	1.000		100.004 10.885	89.119 100.000			**42783.19** bbls/D(Total)	
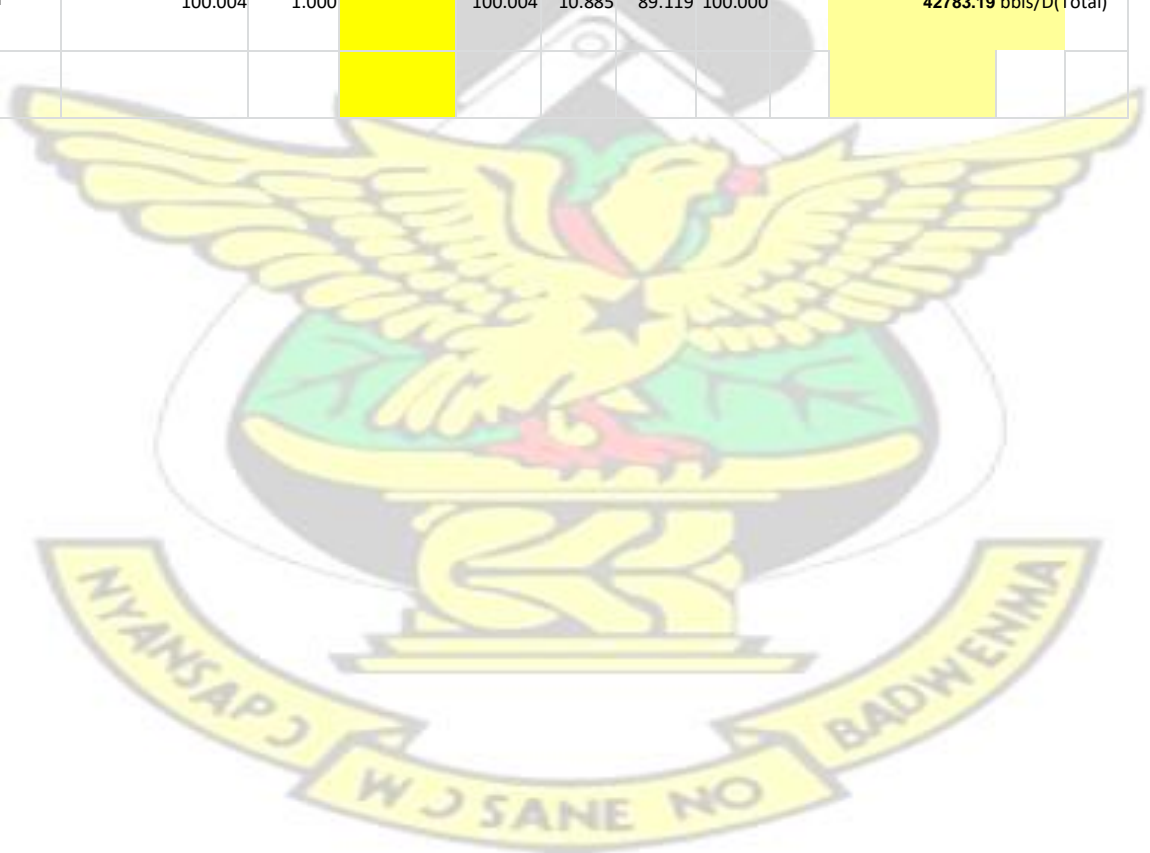
	Flow, mmcfd	640	Initial Comp.				Separation Residual Comp.		Resid Flow Rate		570.36 MMcfd
	Feed %	x1	**%Removal**	initial moles	Moles remo	moles after	x2	Ratio,	Volume Stripped		69.66 MMcfd
								scf/gal	barrels/day strpped		

Component	Feed %	x1	%Removal	Initial moles	Moles removed	moles after	x2	scf/gal	barrels/day strpped
N	0.241	0.002	0 / 0 / 0 / 0 / 0	0.2410	0.000	0.241	**0.270** 1.4420 0.000	91.13	0.00
co2	1.442	0.014		1.442 **1.618** 0.0042	0.000 0.004	**0.005**		59.1	0.00
				78.1270 0.000 78.127 **87.666**					0.00
				8.9430	0.000	8.943	**10.035**		0.00 / 0.00
H2O	0.00422	0.000						175.6	
								59.14	
C1	78.127	0.781						37.5	
C2	8.943	0.089							
C3	7.241	0.072	95 / 100 / 100 / 100	7.2410	6.879 **0.406** 0.8920 0.000	0.362 0.892 0.000 2.1880	**0.892**	36.4	28794.11 / 4435.58 / 10486.55 / 2520.21
iC4	0.892	0.009			2.188	0.000 **0.000**		30.64	
nC4	2.188	0.022		0.4530	0.453	0.000	**0.000**	31.79	
iC5	0.453	0.005						27.39	
nC5	0.4	0.004	100 / 100 / 100	0.4000	0.400 **0.000** 0.0730 0.000	0.000 0.073 **0.000** 0.0000	**0.000**	27.68	2202.36 / 456.27 / 0.00
c6	0.073	0.001	100 / 100 / 100 / 100 / 100	0.000 0.000	**0.000** 0.000 **0.000**	0.0000 0.000 0.0000		24.38 / 24.16 / 24.56	0.00
2m-C5	0	0.000	100 / 100 / 100	0.000	**0.000** 0.000	0.0000 0.000			0.00 / 0.00 / 0.00
3m-C5	0	0.000	100 / 100 / 100	0.000	**0.000** 0.000	0.0000 0.000			0.00 / 0.00 / 0.00
m-c-C5	0	0.000	100 / 100 / 100	0.000	**0.000** 0.000	0.0000 0.000		28.33	0.00 / 0.00 / 0.00
Benzene	0	0.000	0 / 0 / 0	0.000	**0.000** 0.000	0.0000 0.000		35.83	0.00 / 0.00 / 0.00
c-C6	0	0.000	0	0.000	**0.000** 0.000	0.0000 0.000		29.45	0.00
2-m-C6	0	0.000		0.000	**0.000** 0.000	0.0000 0.000		21.57	
3-m-C6	0	0.000		0.000	**0.000** 0.000	0.0000 0.000		21.84	
m-c-C6	0	0.000		0.000	0.000 **0.000**	0.000		24.94	
Toluene	0	0.000		0.0000	0.000	0.000	**0.000**	29.94	

Table B-

Table B-1

et-Benzene	**0**	0.000						25.98		**7**
p-Xylene	**0**	0.000						25.8		
o-Xylene	**0**	0.000						26.36		
H	**0.000**	0.000						111.2 98.69 112.9 74.51		
He	**0.000**	0.000								
O	**0.000**	0.000								
H2s	**0.000**	0.000								
Total	100.004	1.000		100.004	10.885	89.119	100.000		**48895.08** bbls/D(Total)	

Table B-1

	Flow, mmcfd	700	Initial Comp.				Separation	Residual Comp.		Resid Flow Rate		623.83	MMcfd
	Feed %		x1	%Removal	nitial moles	Moles rem	moles after	x2	Ratio,	Volume Stripped		76.19	MMcfd
									scf/gal	barrels/day strpped			
N	0.241		0.002	0	0.2410	0.000	0.241	0.270	91.13	0.00			
co2	1.442		0.014	0	1.4420	0.000	1.442	1.618	59.1	0.00			
H2O	0.00422		0.000	0	0.0042	0.000	0.004	0.005	175.6	0.00			
C1	78.127		0.781	0	78.1270	0.000	78.127	87.666	59.14	0.00			
C2	8.943		0.089	0	8.9430	0.000	8.943	10.035	37.5	0.00			
C3	7.241		0.072	95	7.2410	6.879	0.362	0.406	36.4	31493.56			
iC4	0.892		0.009	100	0.8920	0.892	0.000	0.000	30.64	4851.41			
nC4	2.188		0.022	100	2.1880	2.188	0.000	0.000	31.79	11469.67			
iC5	0.453		0.005	100	0.4530	0.453	0.000	0.000	27.39	2756.48			
nC5	0.4		0.004	100	0.4000	0.400	0.000	0.000	27.68	2408.83			
c6	0.073		0.001	100	0.0730	0.073	0.000	0.000	24.38	499.04			
2m-C5	0		0.000	100	0.0000	0.000	0.000	0.000	24.16	0.00			
3m-C5	0		0.000	100	0.0000	0.000	0.000	0.000	24.56	0.00			
m-c-C5	0		0.000	100	0.0000	0.000	0.000	0.000	28.33	0.00			
Benzene	0		0.000	100	0.0000	0.000	0.000	0.000	35.83	0.00			
c-C6	0		0.000	100	0.0000	0.000	0.000	0.000	29.45	0.00			
2-m-C6	0		0.000	100	0.0000	0.000	0.000	0.000	21.57	0.00			
3-m-C6	0		0.000	100	0.0000	0.000	0.000	0.000	21.84	0.00			
m-c-C6	0		0.000	100	0.0000	0.000	0.000	0.000	24.94	0.00			
Toluene	0		0.000	100	0.0000	0.000	0.000	0.000	29.94	0.00			
et-Benzene	0		0.000	100	0.0000	0.000	0.000	0.000	25.98	0.00			
p-Xylene	0		0.000	100	0.0000	0.000	0.000	0.000	25.8	0.00			
o-Xylene	0		0.000	100	0.0000	0.000	0.000	0.000	26.36	0.00			
H		0.000	0.000	0	0.0000	0.000	0.000	0.000	111.2	0.00			
He		0.000	0.000	0	0.0000	0.000	0.000	0.000	98.69	0.00			
O		0.000	0.000	0	0.0000	0.000	0.000	0.000	112.9	0.00			
H2s		0.000	0.000	0	0.0000	0.000	0.000	0.000	74.51	0.00			
Total	100.004		1.000		100.004	10.885	89.119	100.000		53478.99	bbls/D(Total)		