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Liquefied Petroleum and Compressed Natural Gas recovery option from Nigerian flare gas

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ABSTRACT.

This research developed a concept for selection of recovery plant option for the production liquefied petroleum gas (LPG) and compressed natural gas (CNG) from associated gas (flared gas) produced during exploration and production activities (oil and gas). The process was modelled and simulated using Aspen HYSYS version v8.0. The process flow involved dehydration, compression and cooling, separation before compressing the separated components into liquefied petroleum (LPG) gas and compressed natural gas (CNG). The concept was evaluated economically using conventional method of economic analysis. The concept was found to cost a net total investment of $78,903,030.80. The concept was found to be economically viable breaking even within the first two (2) years of operation.

Keyword: flare gas, recovery, LPG, CNG

Introduction.

The discovery and extraction of natural resources have brought different consequences to countries that are endowed with such resources. While some of these nations have become economically strong and self-sustaining, others have been drawn into serious economic hardships and conflicts. Proponents of the resource have it that the citizens of these countries suffer from abject poverty, environmental damages, pollutions, diseases, illiteracy and score very low on the United Nations Human Development Index. The Niger Delta region, is a region withhuge Oil and Gas resources (Ajuwo, 2013).

However, pollution from domestic and industry operations and activity, over-exploitation of Oil and Gas resources and poorly planned and managed communities and coastal developments are resulting in a rapid degradation of vulnerable land, coastal and offshore habitats and shared living marine resources of the region putting the economies and health of the populace at risk (Ayres,2008). The decline in water and air quality from gas flaring have been identified as a major Trans-boundary environmental problem to communities in the region.

Depleting gas reserves through flaring is a disaster that may likely affect unborn generation if not curbed. The worst aspect of gas flaring is the associated risk to the environment and it is a major contributor to global warming and climate change. Ajugwo (2013) reported that flaring of gas does only waste potentially valuable sources of energy but also adds significantly to carbon emission which harms the health of the people which is linked with terminal illness such as Cancer, asthma, chronic bronchitis, blood disorders and other diseases.

Flaring of gas also gives up additional revenue from the well site. Extracted methane gas could fuel onsite power generators, or be processed and sold as compressed natural gas (CNG) or liquefied natural gas (LNG). Flaring also eliminates potential income from extracted propane, butane, and other saleable NGL products.

Jinn (2015) in their studies of gas flaring impacts in the Niger Delta Region of Nigeria showed that of 1.78×10^{12} cubic meter of gas produced for a period of 51 years, 822.02×10^{9} meter cubic is utilised while 917.17×10^{9} cubic meters are flared. Globally, gas flaring is recognized as a colossal waste of natural resource of which around 150 billion cubic meters are flared annually (Kanshiro, et al., 2017).

Attempt has been made through policy and legislative statement to reduce gas flaring in Nigeria and the world in general. In 2016, the Nigerian government through the ministry of Petroleum resources developed a frame work tagged “7 big wins to grow Nigeria’s oil & gas industry’’ which focus on domestic utilization of oil and gas resources. The policy number three (3) of the big win focus on gas infrastructural development with the view at utilizing gas resources which have been wasted through flaring. The policy firmed view is on utilization of associated gas for the production of CNG and LPG and to boost power generation. Ayoola(2011) reported that, the Nigeria government in its effort adopted the 1973 amended petroleum decree which provides that oil and gas companies can utilise associated gas produced without paying royalty. In 1992, the Nigeria government introduced a fiscal incentives aimed at eliminating flaring of associated gas through the associated gas frame work agreement, these were targeted at improving associated gas utilisation during crude oil exploration (CGFR, 2013). Ajugwo(2013) reported on a need for gas field optimisation studies conducted by Exploration Company to include gas flaring plan.

Over the years Nigeria has been plagued with the problem of gas flaring, these gases release large amount of natural gas which has high global warming potential. Gas flaring has led to the following environmental and socio-political problems such as.

- Emission of Toxic substance
- Waste of natural resources and energy sources

The work aims to study the concept selection of recovery plant option of LPG and CNG production from flare gas.

The specific objectives of this work are to;

1. Review/evaluation of existing natural gas processing designs
2. Design a conceptual plan of optimal specification/performance
3. Compare the conceptual design with existing natural gas processing designs.

The significant of this study is to developed concept that will reduce associated gas flaring. The establishment of this plant is expected to contribute in the following ways to national GDP.

- Increase local productions of LPG and CNG
- It will boost federal government effort in power generation.
- Provide employment opportunities for both skilled and unskilled workforce and thereby solving unemployment challenge.
- It will solve climate change issues as control flaring which is the major contributor of greenhouse gas will be completely eradicated.

The study is focused specifically on associated gas utilisation for the production of LPG and CNG. Other feed stock used for production of LPG and CNG will not be considered. It is a small scale production concept that will be scaled up to larger scope

METHODOLOGY

A front end engineering design approach has been applied in this study and reservoir data obtained from different exploration and production reservoir data was simulated using HYSYS simulation tool.

PROCESS SIMULATION

Simulation Procedure

HYSYS version v8.8 was used for the steady state simulations. The Peng-Robinson Equation of State with default interaction parameters was applied throughout for the thermodynamic property prediction which is given as:

$$P = \frac{RT}{V_m - b} - \frac{a\propto}{V_m^2 + 2bV_m - b^2} \qquad \text{Eq (1)}$$

Feed Compositions

Reservoir fluid hydrocarbon compositions and pseudo-component properties were obtained from different fields within Nigeria. Well fluid compositions of recent samples taken in the field with Feed condition: Temperature: 40˚C (102˚F), Pressure: 3.89MPa (565 psia), (Table 1)

Table 1: Well stream Composition (Feed Composition)

Feed condition: Temperature: 40˚C (102˚F), Pressure: 3.89MPa (565 psia)			
S/NO.	COMPONENTS	MOLE FRACTIONS	MOL. WT.
1.	Methane (C1)	0.6950	16.04
2.	Ethane (C2)	0.0797	30.07
3	Propane (C3)	0.0677	44.10
4	i-Butane (i-C4)	0.0095	58.12
5	n-Butane (n-C4)	0.0213	58.12
6	i-Pentane (i-C5)	0.0054	72.15
7	n-Pentane (n-C5)	0.0065	72.15
8	Hexane (C6)	0.0020	86.18
9	Heptane (C7)	0.0016	100.20
10	Water	0.1113	18.02

PRODUCTION RATES

The design parameter considered for the topsides processing facility for the LPG and CNG production are assumed.

Table 2 Production Rates

Component	Total
Oil (bopd)	00.000
Flare Gas (kmol)	1,000

PROCESS DESCRIPTION

The associated gas considered for this design was obtained from different wells producing crude hydrocarbon. The gas obtained during oil production is assumed to be initially scrubbed before being passed through the dehydrator which completely removed any form of liquid such as water to give complete dry gas before being routed to the compressor. The dry gas is routed to the suction of the first stage compressor. The inlet condition of the gas was set at 40˚C and 565psia. The compressed gas from the first stage compressor is charged into the De-ethanizer column to separate methane and ethane from the dry gas, the propane and other hydrocarbon recovered at the bottom of the de-ethanizer were charged into de-butanizer where they were further separated into the propane and butane component. The mixture of propane and butane obtained at this stage were stored in a storage tank before being dispensed out.

EQUIPMENT DESCRIPTION AND CONSIDERATION

The liquefied petroleum gas and compressed natural gas plant concept were a skid mounted modular plant with the facility having the following: The equipment considered are in figure 1. and 2

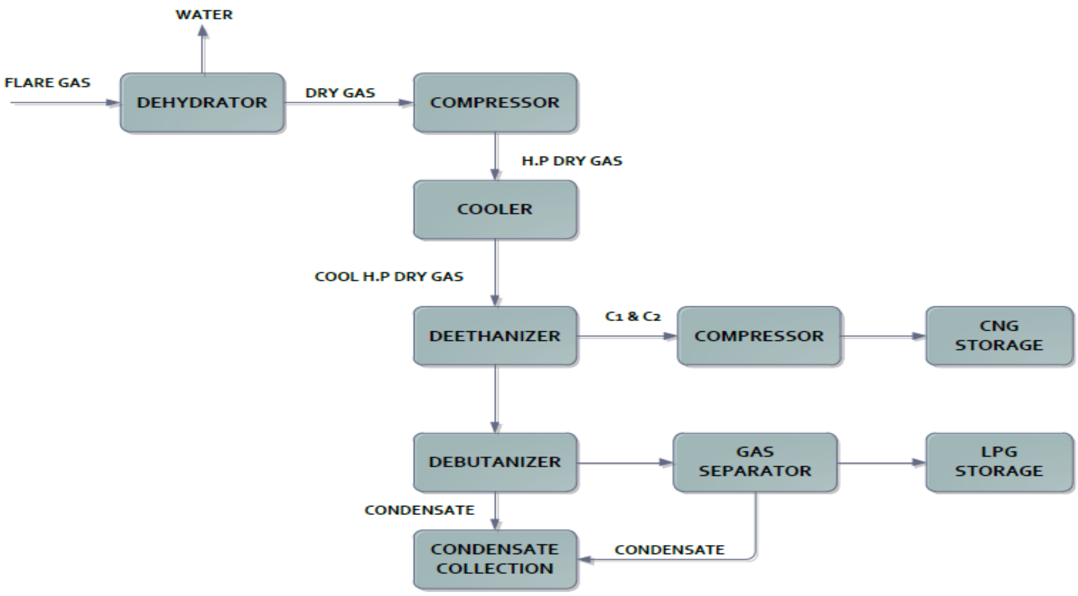

Figure 1: Block Flow Diagram

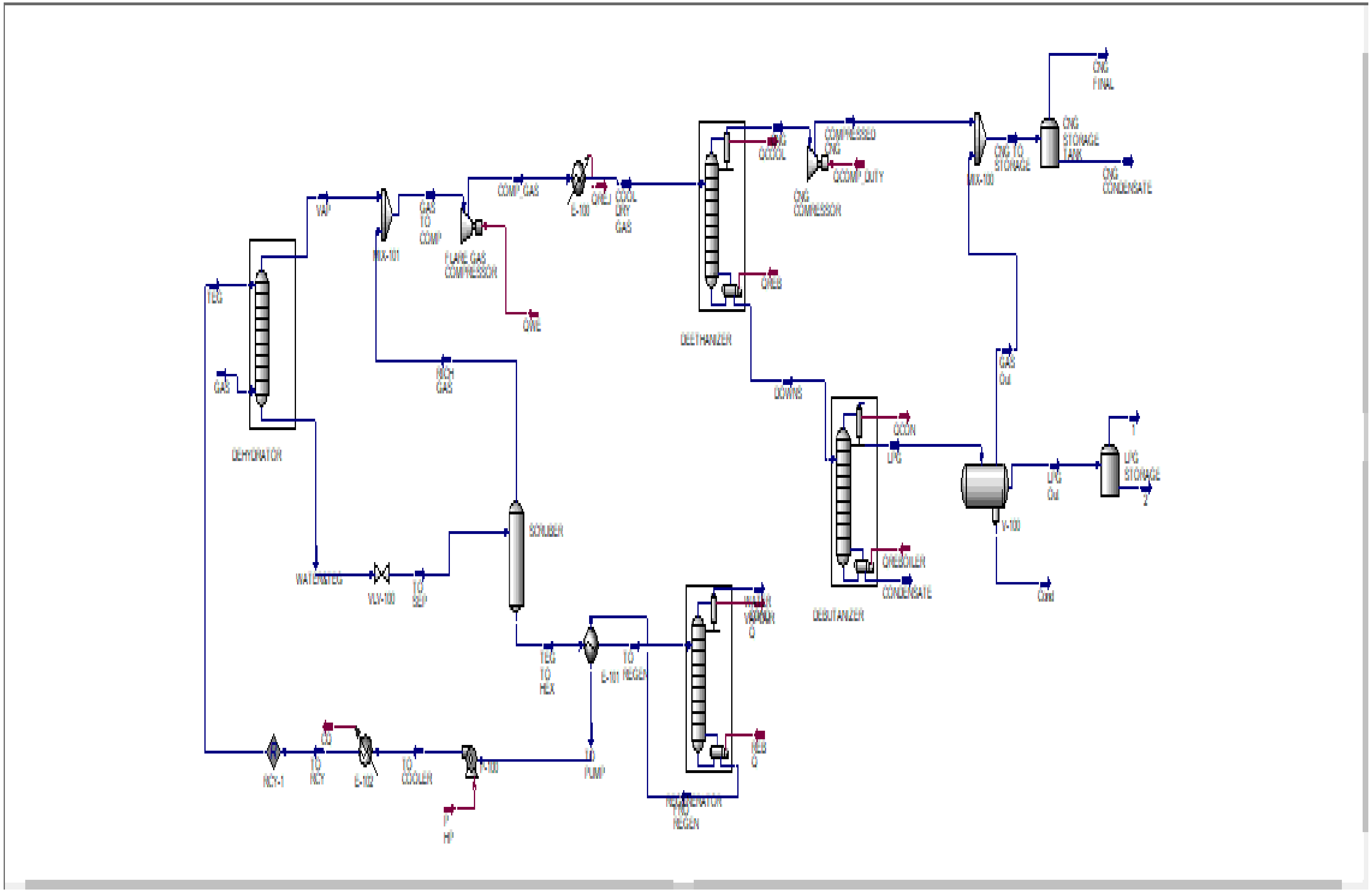

Figure 2:Hysys Simulation Process Flow Diagram

Material Balance around the system

Material balance around the system was done based on the fundamental laws of mass conservation with the assumption of steady state operation.

Input + Generation – Output – Consumption = Accumulation Eq2

Material Balance on dehydrator

Basis: 1000 kmol of wet flare gas (1 day of operation)

Assumption(s):

All water molecules are stripped from the flare gas i.e. flare gas is completely dry

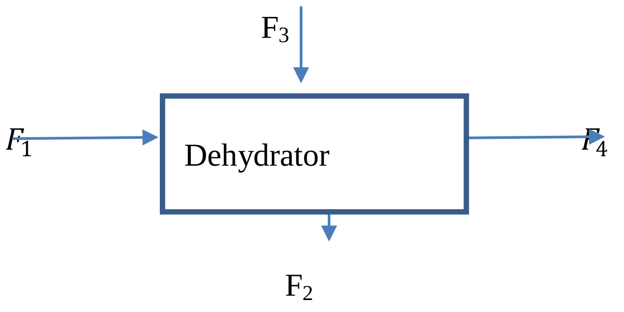

Figure 3: Block Diagram for Dehydrator Stream Balance

Overall material balance equation:

$F_1 + F_3 = F_2 + F_4$ Eq3

Using a feed ratio 0f 0.3

F_3=300 kmol

Water balance;

$0.1113F_1 = w2F_2$

Glycol Balance

$g_2 f_2 = f_3$

$0.1113f_1+0f_3 = w_2 f_2$

Gas Balance

$0.8887F_1=W_4F_4$

$F_2=F_1 + F_3 - F_4$

Material Balance on Deethanizer

Basis: 888.7 kmol of dry flare gas (1 day of operation)

Assumption(s):

- Overhead stream contains only methane and ethane
- 100% of methane and ethane is recovered from the flare gas at the overhead stream

Overall material balance equation:

$F_5 = F_6 + F_7$ Eq 4

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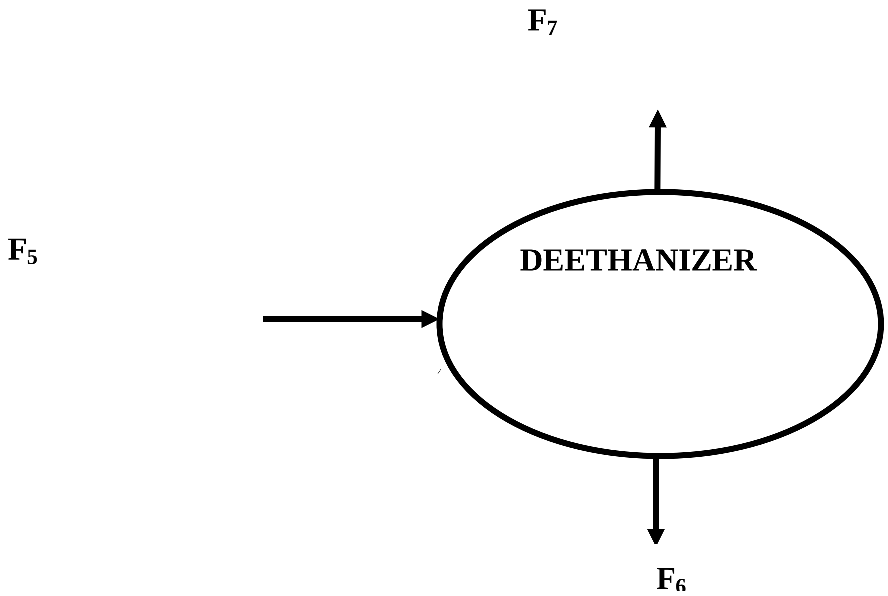

Figure 4: Material Stream for Deethanizer Stream

$F_7 = F_5 - F_6$ Eq5

Material Balance on Debutanizer
Basis: 114.02 kmol (1 day of operation)
Assumption(s):

- Overhead stream contains only propane, i-butane and n-butane
- 100% of propane, i-butane and n-butane are recovered from the flare gas at the overhead stream

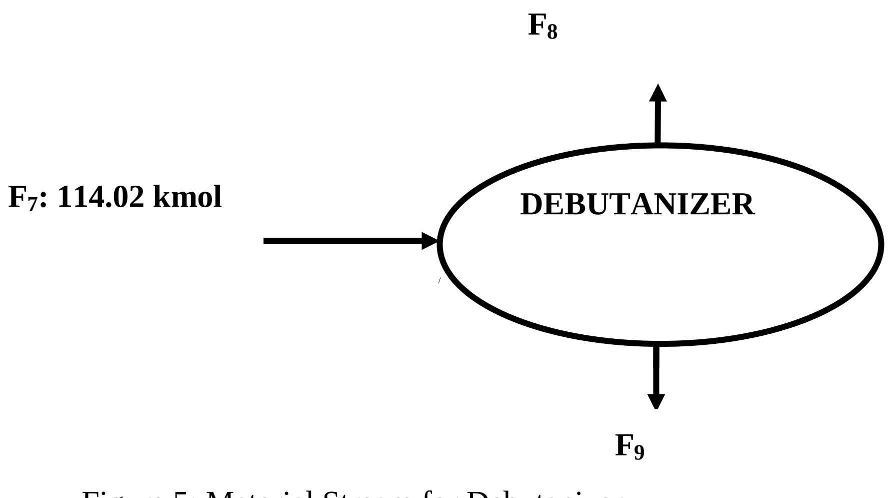

Figure 5: Material Stream for Debutanizer

Overall material balance equation:

$F_7 = F_8 + F_9$ Eq6

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Energy balance around the system

The energy balance around the system is carried out based on the fundamental laws of conversion of energy.

Energy balance on the flare compressor

Considering steady state for the compression process, the mass flow of flare gas through the compressor can be represented by:

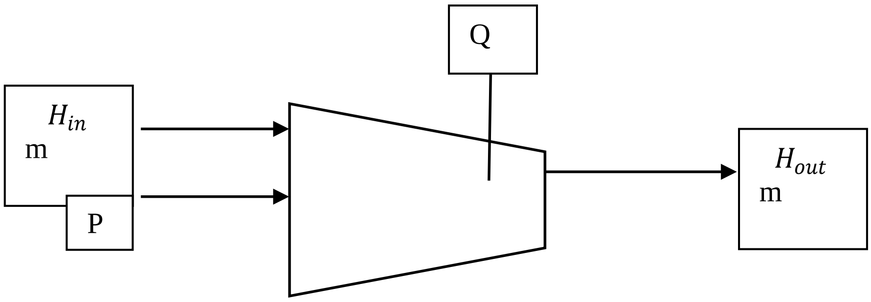

Figure 6: Energy Balance block Diagram

$$mH_{in} + P = mH_{out} + Q \qquad Eq\ 7$$

Where: m=mass rate of flow of the feed gas in and out of the system

H_{in} and $H_{out}$are the specific enthalpies into and out of the compressor

P is the mechanical power of the compressor

Q is the heat flow from the compressor

$W = H_4 - H_3$Eq8

$W = \dot{n} \times \widehat{W}, (ideal\ work)$Eq 9

Assuming a compressor efficiency of 75%,

$W_{actual} = \frac{W_{ideal}}{\eta}$Eq10

Assuming the compressor operates isentropically (reversible adiabatic) i.e. s_3=s_4

Energy balance on the gas coolers

The energy balance around the cool was carried out based on the fundamental laws of conservation of energy governed by the model equation:

$Q_{cooler} = \dot{n}(\widehat{H}_4 - \widehat{H}_5)$Eq 11

Energy balance on the de-ethanizer and debutanizer

The energy balance around the de-ethanizer and debutanizer was based on the fundamental laws of conservation of energy and they were carried out in line with the global model for energy balance for multi-stage distillation columns given as:

$\sum_{i=1}^{n} F_i H_i^F - \sum_{i=1}^{n} W_i H_i^v + U_i H_i^L + Q = 0$Eq 12

Where F is the feed stream of gas at stage i, $H_i^F$is the feed enthalpy, Q heat stream from stage I, U and W represents the distillates and bottom products from the process.

Cost Analysis

The cost analysis for the design project was done using HYSIS simulation software and the input information was obtained from the equipment sizing simulation and material balance carried out.

Capital Cost
Total Equipment Cost(PCE) = $\sum major\ equipment\ cost$.Eq 13
Total Physical Plant Cost(PPC) =PCE$\sum$ $(1+f_1+\ldots..f_9)$
Fixed capital =PPC$(1+f_{10}+f_{11}+f_{12})$
Working capital= 5% of the fixed capital
Total Investment Required= $\sum$ (Working Capital + Fixed Capital) Eq 14

Operating Cost
Utility cost = $\sum$ (Electricity + Cooling water +Refrigerant + Steam + Glycol)
Variable cost= $\sum$ (Cost of raw material +Utility cost)
Maintenance cost = (5% of fixed capital)
Plant overheads = 50% of operating labour
Laboratory = 30% of operating labour
Capital charges=10% of fixed capital
Insurance = 1% of fixed capital
Operating cost= $\sum$ (Utility cost + variable cost +Maintenance cost +plant overhead cost +Laboratory cost+ cost of Labour +capital Charges +insurance cost) Eq14

Revenue Estimation
Annual sales of CNG = Quantity of CNG* Cost of CNG
Annual sales of LPG = Quantity of LPG* Cost of LPG
Total annual sales = Annual sales of LPG + Annual sales of CNG
Annual profit= total annual sales – direct product cost
Annual profit after tax =Annual profit* tax rate Eq 15

Payback period
Pay-back period $=\frac{Net\ investment}{Profit\ \ after\ \ Tax+Depreciation}$ Eq16

RESULTS AND DISCUSSION

Results
Mass Balance and Energy Balance for CNG and LPG
Mass Balance on Dehydrator

Table 3: Result of mass balance on the dehydrator

	INPUT(kmol/day)		OUTPUT(kmol/day)			
STREAM	y1	F1	y2	F2	y3	F2
COMPONENTS						
C1	0.695	695	0	0	0.782	694.9634
C2	0.0797	79.7	0	0	0.0897	79.71639
C3	0.0677	67.7	0	0	0.0762	67.71894
i-C4	0.0095	9.5	0	0	0.0107	9.50909
n-C4	0.0213	21.3	0	0	0.024	21.3288
i-C5	0.0054	5.4	0	0	0.0061	5.42107
n-C5	0.0065	6.5	0	0	0.0073	6.48751
C6	0.002	2	0	0	0.0023	2.04401
C7	0.0016	1.6	0	0	0.0018	1.59966
Water	0.1113	111.3	1	111.3	0	0
TOTAL	1	1000		111.3		888.7

Mass balance on Deethanizer

Table 4: Result of mass balance on the deethanizer

	INPUT(kmol/day)		OUTPUT(kmol/day)			
STREAM	y3	F3	y6	F6	y7	F7
COMPONENTS						
C1	0.782	694.9634	0.8971	694.9654	0	0
C2	0.0897	79.71639	0.1029	79.71457	0	0
C3	0.0762	67.71894	0	0	0.5939	67.71648
i-C4	0.0107	9.50909	0	0	0.0834	9.509268
n-C4	0.024	21.3288	0	0	0.1871	21.33314
i-C5	0.0061	5.42107	0	0	0.0475	5.41595
n-C5	0.0073	6.48751	0	0	0.0569	6.487738
C6	0.0023	2.04401	0	0	0.0179	2.040958
C7	0.0018	1.59966	0	0	0.014	1.59628
Water	0	0	0	0	0	0
TOTAL		888.7		774.68		114.02

Mass balance on Debutanizer

Table 5: Mass balance on the debutanizer

	INPUT(KMOL/DAY)		OUTPUT(KMOL/DAY)			
STREAM	y7	F7	y8	F8	y9	F9
COMPONENTS						
C1	0	0	0	0	0	0
C2	0	0	0	0	0	0
C3	0.5939	67.71648	0.687	67.71072	0	0

i-C4	0.0834	9.509268	0.0965	9.51104	0	0
n-C4	0.1871	21.33314	0.2165	21.33824	0	0
i-C5	0.0475	5.41595		0	0.3486	5.389356
n-C5	0.0569	6.487738		0	0.4171	6.448366
C6	0.0179	2.040958		0	0.1314	2.031444
C7	0.014	1.59628		0	0.1029	1.590834
Water	0	0	0	0	0	0
TOTAL		114.02		98.56		15.46

Energy Balance Results

Table 5: Energy balance around the cooler

At inlet conditions: Temperature: 40˚C (102˚F); Pressure: 3.89MPa (565 psia)						
S/N	COMPONENTS	y_i	$\hat{H}_3$(kJ/mol)	$y_i\hat{H}_3$ (kJ/mol)	$\hat{S}_3$(kJ/mol.K)	$y_i\hat{S}_3$ (kJ/mol.K)
1.	C1	0.7820	14.5915	11.4106	0.07863	0.0615
2.	C2	0.0897	14.29432	1.2822	0.057755	0.0052
3	C3	0.0762	18.17398	1.3849	0.073606	0.0056
4	i-C4	0.0107	18.041	0.1930	0.077852	0.0008
5	n-C4	0.0240	17.42857	0.4183	0.0791667	0.0019
6	i-C5	0.0061	1.771397	0.0108	0.002818	0.0000
7	n-C5	0.0073	1.771397	0.0129	0.002818	0.0000
8	C6	0.0023	-5.90401	-0.0136	-0.0192	0.0000
9	C7	0.0018	-14.1764	-0.0255	-0.04279	-0.0255
	TOTAL			14.6736		0.0749
	Heat Duty (KJ/day)			1,371915.813		

Table 6: Energy balance around the cooler

Outlet conditions: Pressure= 5MPa , Temperature = 200 K				
S/NO	COMPONENTS	y_i	$\hat{H}_5$(kJ/mol)	$y_i\hat{H}_5$ (kJ/mol)
1.	C1	0.7820	8.7337	6.8298
2.	C2	0.0897	1.2902	0.1157
3	C3	0.0762	1.6459	0.1254
4	i-C4	0.0107	2.9387	0.0314
5	n-C4	0.0240	2.6947	0.0647
6	i-C5	0.0061	-16.632	-0.1015
7	n-C5	0.0073	-16.632	-0.1214
8	C6	0.0023	-25.975	-0.0597
9	C7	0.0018	-16.722	-0.0301
				6.8540

	Heat Duty (KJ/day)	7,978,215.38

Table 7: Energy balance around the de-ethanizer

Saturated properties of methane and ethane at 2MPa		
	H_{liquid} (kJ/mol)	H_{vapour} (kJ/mol)
Methane	3.3900	8.9006
Ethane	6.7294	16.5641

Equipment Design and Specification

The detail design (sizing and specification) for the concept developed for the production of LPG and CNG was done using process design software Aspen Hysys version v8.8 the equipment were sized using Aspen sizing tools.

Table 8: TEG Dehydration Units Sizing Specification Result

S/N	Process Data Dehydrator	parameter
1	Internal Tray type	sieve
2	Diameter	1.5m
3	Tray/Packed space	0.5m
4	Tray/Packed volume	0.88m^3
5	Hold up volume	0.088m^3
6	Weeping factor	1.00

Table 9: Regenerator Units Sizing

S/N	Process Parameter	
1	Internal tray type	Sieve Type
2	Diameter	1.5m
3	Hold up volume	0.08836 m^3
4	Tray volume	0.9719m^3
5	Weeping factors	1.0

Table 10: Vessel Sizing (Condenser & Reboiler Section)

S/N		Process Parameter	Data
1		Diameter	1.193m
2		Length	1.789m
3		Volume	2.00m^3
4		Hold up volume	1.00m^3
5		Pressure drop across the reboiler	11.11 KPa

Table 11: Heat Exchanger for the Regenerator

S/N	Process Parameter	Data
1	Number of shell passes	1

2	Number of shell in series	1
3	Number of shell in parallel	1
4	Tube passes per shell	2
5	Heat exchanger orientation	Horizontal counter current type
6	Overall heat transfer coeff (u)	516.3KJ/-m ^{2}c
7	Heat transfer areas	60.32m^2
8	Tube volume per shell	0.1930m^3
9	Shell Volume per shell	2.272m^2
10	LMTD	334.2 c
11	Heat duty	10410000 KJ/h

Table 12: Regenerator Pump

S/N	Process Parameter	Data
1	Capacity	10m^3/h
2	NPSH	46.44m
3	Pressure head	430.9m
4	Total power	117.3kW 157.3HP
5	Velocity head	1.575m

Table 13: Heat Exchanger for the Regenerator

S/N	Process Parameter	Data
1	Number of shell passes	1
2	Number of shell in series	1
3	Number of shell in parallel	1
4	Tube passes per shell	2
5	Heat exchanger orientation	Horizontal counter current type
6	Overall heat transfer coeff (u)	516.3KJ/-m ^{2}c
7	Heat transfer areas	60.32m^2
8	Tube volume per shell	0.1930m^3
9	Shell Volume per shell	2.272m^2
10	LMTD	334.2 c
	Heat duty	10410000 KJ/h

Table 14: Regenerator Pump

S/N	Process Parameter	Data
1	Capacity	10m^3/h
2	NPSH	46.44m
3	Pressure head	430.9m
4	Total power	117.3kW 157.3HP

5	Velocity head	1.575m

Table 15: Compressor Process Data Result

1	Adiabatic head	4615m
2	Polytropic head	4675m
3	Adiabatic fluid head	45.26kJ/Kg
4	Polytropic fluid head	45.85kJ/Kg
5	Adeadatic efficiency	75%
6	Polytopie efficiency	76%
7	Power consumed	429.1 HP
8	Flow rate	365.8 ACFM

Table 16: Heat Exchanger (Counter Current Orientation)

S/N	PROCESS PARAMETER	DATA
1	No of sheet	1
2	No of sheet in barrel	1
3	Tube passes per shell	2
4	Overall U(KJ/h)	514.3
5	UA(kJ/c-h)	31141
6	Heat transfer area per shell(m^2)	60.32m^2
7	Shell baffles	single with horizontal orientation
8	Baffle cut (%Area)	20
9	Baffels spacing(mm)	80
10	Tube OD(mm)	20
11	Tube ID(mm)	16
12	Tube thickness (mm)	2
13	Tube length(m)	6

Table 17: Debutanizer Sizing detailed

S/N	Tray Section /Packed Section Size	Data
1	Diameter	1.5m
2	Tray /packed space	0.55m
3	Tray/packed volume	0.9719m^3
4	Hold up	0.08835m^3
5	Vessel Sizing (Reboiler & Condenser)	
6	Diameter	1.193m
7	Length	1.789m
8	Volume	2.0m^3
9	Reflux ratio	2.5
10	Vent rate	765 Kg/h

Table 18: separator data sheet

S/N	PROCESS VARIABLES	DATA
1	volume	$15m^3$
2	diameter	2.3m
3	Length	3.5m
4	Vessel orientation	horizontal

Table 19: Scrubber Data sheet

S/N	PROCESS PARAMETER	DATA
1	Volume	$1000m^3$
2	Diameter	4.4 m
3	Height	6.6 m

Table 20: CNG Tank design data sheet

S/N	Process Variable	Data
1	Volume	$60m^3$
2	Diameter	3.7m
3	Height	5.5m

Table 21: LPG Tank data sheet

S/N	PROCESS VARIABLE	DATA
1	Volume	$20m^3$
2	Diameter	2.57m
3	Height	3.85m

Cost Analysis

Table 22: Capital Cost

S/N	Cost Parameter	Cost (USD)
1	Total Equipment cost (PCE)	4,713,800
2	Total Physical Plant cost (PPC)	11,313,120
3	Fixed Capital	15,838,368
4	Working Capital	4,751,510.40
5	Total Investment Required	20,589,878.40

Table 23: Operating Cost

S/N	Cost Parameter	Cost (USD)
1	Utility cost	8,858,950.56
2	Variable Cost	27,709,007.76
3	Maintenance cost	791,918.4
4	Plant Overheads	100,000
5	Labour	200,000
6	Laboratory cost	60,000
7	Capital Charges	1,583,836.8
8	Insurance	158,383.68
9	Fixed cost	2,894,138.88
	Operating cost	30,603,146.64

Table 24: Revenue Estimation

S/N	Cost Parameter	Cost (USD)
1	Annual Sales of CNG	233,909,674
2	Annual Sales of LPG	238,636,24.34
3	Total Annual Sales	257,773,298.34
4	Annual Profit	2271,701,51.70
5	Annual Profit After Tax	136,302,091

Discussion of Findings

DEHYDRATION DESIGN

The dehydration column is design to give complete removal of water and give 99% regeneration of the triethyl glycol. The contactor operates at the gas export pressure of 558.4 psia and 120 ^oF. The vessel is designed to remove water from the gas to the lowest level with a number of stage put at twenty-five. The TEG rate required per pound mole of water is 500kgmole/hr. The tray efficiency is 25% and the tray/packed volume is 0.88m^3 with sieve type of internal tray type and diameter of 1.5m. the regenerative section design to recover 100% of the tri-ethyl glycol used. The regenerative section comprises condenser and reboiler sized with a diameter of 1.193m with a hold up volume of 1.00m$^{3.}$

Compressor

The compressor is a single stage centrifugal compressor designed to run continuously when in operation. The compressor flowrate is design at 365.8 ACFM with a polytropic head of 4675m and adiabatic head of 4615m with adiabatic and polytropic efficiency of 75%. The compressed gas from the 1st stage compression station is routed to cooler to cool the gas 121.5 ^0F before transporting the gas to the Deethanizer column for further processing.

Heat Exchanger Design

The heat exchanger is design to cool the compressed gas before been routed into the dethethanizer. The cooler is design with two tubes per shell sized to heat transfer area per shell to 60.32m^2 withShell baffles of single with horizontal orientation with baffles cuts area of 20%.

De-Ethanizer

The Deethanizer tower is design to separate the component of the cooled compressed gas into the different component. The De-ethanizer separate the component of methane and ethane which is obtained at the upper section of the tower. The tower is sized to 1.5m diameter which have hold up volume of 0.08835m3 the distillation column is divided into tray section and vessel section which comprises of the Reboiler and Condenser. The vessel is sized with a diameter of 1.193m diameter and volume of 2.0m3.

DEBUTANIZER

The debutanizer is design to further separate the gas component of the C4 series which were not split by the dethanizer. The primary role of the debutaniser is to separate completely the component of the LPG which is further routed into three phase separator to split the component of the LPG, CNG and some condensate before storage. See the detail below for the sizing detailed.

LPG Separator

The LPG separator is a horizontal vessel designed to handle the bulk flow from THE debutanizer. The separator is designed as 3-phase separator. It is designed to handle the gas component downstream of the column.

The separator is provided with diverter to knock off the bulk flow of the liquid, a vane pack to minimise liquid carry-over, vortex breaker on the liquid line to avoid coagulation and a submersible weir.

Gas Scrubber Process Data

The mixture of water, Triethyl glycol and gas from dehydrator are routed to the 1st stage gas Scrubber to remove any carry-over from the dehydrator unit. The separated gas is routed to the suction of 1st stage gas Compressor. Pressure in the 1st stage Compressor is maintained nominally at 3350kpa. The Scrubber has a demister pad that removes water to minimum percentage. Any liquids removed normally from the demister pad flow into the bottom of the vessel which connects to the closed drain header back to regenerator reboiler for recycling. The liquid from the vessel is been discharged when the high liquid level is attained in the vessel. The cycle is repeated with the cross-over valve opened to allow upper compartment to drain into the lower compartment.

CNG Storage Tank

The CNG storage tank is a spherical shape type of tank to handle the high pressure associated with compressed gas. The tank design detail as seen in the table below.

LPG Storage Tank

The storage tank is design with a floating roof fitted with automatic bleeder and a pressure relief system. The sizing required for this concept are seen in the detail below.

Cost Analysis

The economic evaluation of the concept developed for the production of LPG and CNG from flare gas will be based all the equipment of dehydrator (Absorber Column), Dethanizer column, gas scrubber, separator, debutaniser column, heat exchangers of various types, pump and storage tanks in estimating the total capital investment.

Capital Cost Estimation

The cost required for the total investment cost required to make the concept achievable is estimated at $20,589,878.40 the total net investment required for initial start-up is put at $78,902,032.80.

OPERATING COST ESTIMATION

The operating cost is directly influence by the cost of electricity, cost of tri-ethyl glycol etc. the total production cost is estimated to be at $ 30,603,146.64.

REVENUE ESTIMATION

The revenue estimated is based on the cost of raw material used and the expected annual sales of LPG and CNG that is produced based on this concept, the revenue to be generated per annum from annual sales is placed at $257,773,298.34 which puts the profits after tax at $136,302,091.02

Payback Period

This payback period for the concept was developed based on the total net capital investment and the net profit and it’s evaluated to 1.75 year which shows that the breakeven period is quite small and investment will be recovered quickly.

Profitability Analysis

The profitability analysis was done using the entire plant developed to see how investor will recover their investment that will be vested into the concept developed. It was shown from the net annual profit and short pay-back period recorded that the process is deem profitable.

CONCLUSION

This research focus basically on the development of a concept that will produced CNG and LPG from associated gas (flare gas) from the production and exploration activities within the oil and gas sectorthat are ordinary been wasted and contribute immensely to greenhouse emission. This is important because the flare gas that is constantly waste through routine flaring and production associated flaring contain crucial compound that are used for the production of LPG and CNG and no empirical based research have been conducted on the possibility of utilizing the resources that are constantly been wasted.

Based on the simulation result obtained using HYSYS simulation tools to model the concepts, the following conclusion were drawn

1. Methane and ethane which constituted the major component of the flare gas are recovered completely at the dethanizer column and therefore compressed to give the CNG.
2. Propane and Butane which constituted LPG are recovered 100% from the debutanizer column which is further treated using separator to removed condensate and CNG carryover before been blended in a storage tank
3. The Economy Analysis shown that the concept is profitable and investor investment will be recover within two years of operation as shown in the economy evaluation

REFERENCES

GGFR Technology overview (2018). Utilization of small-scale associated gas. The World Bank technical report.

Jinn, G (2015). Gas Flaring in Nigeria, an overview justice in Nigeria now. Journal of Environment Pollution and Human Health, available www.justiceinnigerianow.ord

Kanshio, S, Agogo H.O and Joseph C. T. (2017). Techno-Economic Assessment of Mini-GTL Technology for flare gas Monetization in Nigeria, SPE Nigeria Annual International Conf and Exhibition, 31 July, 2017. SPE-189120-MS. Pp 1-11

Ajugwo A.O (2013). Negative effects of Gas Flaring: The Nigeria Experience, journal of Environment Pollution and human Health, 1 (1) pp 6-8

Arowolo A.A, and Adaja I.J, (2014). Trends of Natural Gas Exploitation in Nigeria and its Implications on the socio-economic stability and governance, 35th Nigeria Statistical Association Annual Conference.

AyoolaJ. T (2011), Gas flaring and its implications for Environmental Accounting in Nigeria, Journal of Sustainable Development. 4(5) 244-250.

Ayres R. U.(2008), sustainable Economics: where do we stand? Ecological economics, Elsevier, vol 67(2) pp 281-310