ISSN: 2639-7536

Research Article

Petroleum and Chemical Industry International

Economic Analysis of Flare gas recovery system in a Refinery plant in Nigeria

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Submitted: 22 Apr 2022; **Accepted:** 25 Apr 2023; **Published:** 30 Apr 2023

Citation: Edeh, I., Olawale, Y. M. (2023). Economic Analysis of Flare gas recovery system in a Refinery plant in Nigeria. *Petro Chem Indus Intern, 6*(2), 142-150.

Abstract

The emission of greenhouse gases like methane and carbondioxide during gas flaring in the oil and gas refineries is of immense concern in mitigation of climate change. As a result of this, flare gas recovery system is encouraged through various policies to be installed in oil and gas refineries. The current work is on the economics analysis of flare gas recovery system in a refinery plant in Nigeria to determine the prospect and feasibility of installation and operation. The simulation of the system was carried out using Unisim Design 471 software, and the economic analysis conducted manually. The result of the cost estimation and profitability analysis show total purchased equipment cost ($52,327,866), total capital investment ($549,088,158), annual total production costs ($204,681,762), profit after tax ($2,007,068,515), rate of return (365.50 %) and payback time (3.28 months). Based on the result, the flare gas recovery system has a great prospect and feasibility, its installation and operation could help to minimize the greenhouse gases emission, energy consumption, create employment and increase the revenue generation of the refinery.

Keywords: Flare gas, recovery system, Refinery, economics, profitability

Introduction

Refinery gas flaring negative impact on environment extends to precipitate colossal health consequences. The greenhouse house effect stems from natural balance disruption caused by flaring emission gases. Examples of such gases include carbon dioxide, carbon monoxide, methane, water vapour, Nitrogen oxide family, soot and so on [1,2]. They are capable of causing health issues like cancer, lung damage, deformities in children, genetic mutation, asthma, bronchitis, pneumonia, neurological and reproductive disorder as well as environmental challenges which stall agricultural productivity, aquatic and wild lives [NOSDRA]. The environmental consequences result from climate change, acid rain and other forms of air pollution [3].

According to data released by National oil spill detection and response agency [NOSDRA], an arm of Nigeria Federal ministry of environment, through its satellite tracker, 1.8 billion Standard cubic feet [scf] per day of gas was flared in nine years, one that should attract about $3.6 billion in penalty, little of which was paid. The flared gas is valued at $6.3 billion and could generate 179.9 thousand GWh of electricity which represents a milestone in energy needs of Nigeria. The volume generated about 95.5 million tonnes of carbon dioxide emissions. In 2020, natural gas valued at $1.24 billion was burned by oil companies, one which could generate the annual electricity use of 804 million Nigerian citizens, according to the tracker [3].

The international dent caused by flaring in Nigeria is significantly monumental. Nigeria is ranked seventh in the global oil flaring index released in April 28, 2021 by world bank report tagged "Global gas Flaring Trackers" putting the country in hot spot. The six leading countries are Russia, Iraq, Iran, the United States, Algeria and Venezuela.

Forbes report of July 2021 showed that scope 1 and 2 emissions associated with refining operations account for around 5 % of the global total oil and gas emissions. The emissions result mainly from energy intensive processes of distillation and chemical conversion that are integral to refining along with the production of grey hydrogen which is the hydrogen produced from natural gas.

Refining profit margin is dipping and might not rebound soon due to large numbers of new refineries springing up in the middle east and Asia, emergence of biofuel and manufacture of electric vehicle. Wood Mackenzie's global composite profit margin averages US$1.8/bbl in 2021 but less than half of the US$4.25/bbl five-year average.

Nigeria, like other countries of the world, enacted a number of laws and regulations aimed at controlling discharges into the environment. They are in different forms and dimensions. One of such regulations is the National environmental protection [pollution abatement in industries and facilities generating wastes] regulations,1991. The Nigerian regulations, made under section 37 of the Federal Environmental Protection Agency Act, provide for control of discharge by industries in Nigeria. It reads “no industry or facility shall release hazardous or toxic substances into the air, water or land of Nigeria’s ecosystems beyond limits approved by the Federal Environmental Protection Agency”. Discharge, including, solid, gaseous and liquid waste from any industry or facility shall be analyzed and reported to the nearest office of the agency. The provision falls short of placing the analyses report in public domain so as to reflect compliance, accountability and transparency not only to the host communities but citizenry in general [4].

In Petroleum Industry act 2021, section 104 subsection 1-4, provides that a licensee, lessee, or marginal field operator can only flare or vent natural gas in the case of emergency if and only it is an acceptable safety practice established under this regulation. Section 107 of the act has it that a licensee or lessee can flare where it is required for facility start-up or testing gas equipment or plant and fine as prescribed by the commission will be paid to the government by the defaulting oil companies in the same way as royalties if such exercise fails. Such fines are deployed for environmental remediation and relief to host communities. The fines are prescribed by the Flare gas [Prevention of Waste and Pollution] Regulations [4]. In Section 106 subsection 1 it is made mandatory for licensee or lessee to install a metering equipment in any facility where natural gas may be flared or vented before petroleum production can commence and non-compliance attracts fine prescribed by regulatory Authority. Another section of the Act, section 107 stipulates that licensee or lessee producing natural gas shall, within 12 months of commencement of operation submit a natural gas flare elimination and monetization plan to the concerned Authority in accordance with established regulations of the Act. The Nigerian Midstream and Downstream Petroleum Regulatory Authority is the one issuing permit to gas firms and regulates gas flaring in the sector [5].

It is obvious from the above regulations that serious efforts are being made to discourage gas flaring and reduce it to the barest minimum. In all of the regulations, all the attentions are on all the oil producing companies. None of regulations is specifically directed at the refineries known for about 5 % emission pool as obtainable in United states of America where specific regulatory provisions and projections in clear terms are impeccably detailed. Other producing companies that do not fall in the category of oil and gas companies or refineries need to be adequately captured in specific terms as they are major contributors to gas flaring. It is imperative that such loophole be addressed in the future amendment to the Act. The Act provisions are excellently drafted but implementation and enforcement are equally important in order to drive it to achieve the intended results. Also, milestones recorded with the Act need to be documented so that its progress can be effectively monitored and vacuum yet to be filled by the Act can be clearly spelt out and required effort/review can be intensified in this regard.

Furthermore, some researchers have investigated the feasibility of flare gas recovery systems through economic analysis. Among such researchers include, and they carried out a technical and economic assessment of flare gas recovery in a giant gas refinery. They assessed the feasibility of using liquefaction, LPG production and a three-stage compression unit in recovery flare gas. The result of the economic analysis showe that the rate of return (ROR) obtained for liquefaction and LPG production units, respectively were greater than 200 %. In the same vain, conducted a technical, economic, and environmental assessment of flare gas recovery system while considering FGRS methods including pressurizing and injecting flare gas into oil wells, producing electricity and injecting surplus flare gas into oil wells, and producing power via a combined heat and power system (CHP) and an internal combustion engine. The result show that the best method of recovering flare gas under the conditions investigated was by pressurizing and injection based on economic reason of having an internal investment rate of 171 % and a payback period of 1.02 years. Similarly, Carried out a thermo-economical assessment of producing liquefied natural gas natural gas liquids from flare gases using an auxiliary natural gas flow rate, and poly refrigerant intergrated cycle operation [PRICO]. The result of the economic analysis gave a payback period of approximately 1.6 years [6-8].

In our previous work, we investigated the recovery and purity of some important constituents of flare gas like methane, hydrogen, propane, ethane and debutanized products via the simulation of the flare gas recovery system using Unisim Design 471 software. The current work is geared towards evaluating the prospect, viability and feasibility of installing and operating a flare gas recovery system in a refiner. This was carried out by determining the total purchased equipment, total capital investment, annual total production costs, profit after tax, rate of return, and payback time.

Materials and Methods

The details of the simulation of the flare gas recovery system, and material and energy balances have been reported in our previous work [9].

Process Description

The flare gas streams from different sections of the refinery represented by Fluid Catalytic Cracking Unit [FCCU] flare gas from a header was routed either to Flare system or Flare gas recovery system.

The flare gas streams from the header of the Fluid Catalytic Cracking Unit [FCCU] is passed to either flare system or flare gas recovery system [FGRS]. The flare gas at 193.1 kPa and 36.67 oC routing to FGRS was passed through a Let-down valve where the pressure of the gas was reduced to almost vacuum of about 6.895 kPa [see Figure 1]. After which It entered the Liquid ring Compressor at vacuum suction pressure and was subjected to compression to a discharge pressure of about 1620 kPa. The compressor was modelled using a centrifugal hybrid with adiabatic efficiency of 50 %. Hot and compressed flared gas enter the trim cooler E-100 where its temperature drops to 57.22 oC at 1620 kPa, before passing it to a three-phase separator, SEP 1 where the partially condensed gas stream is flashed into vapour, liquid and water. The vapour fraction is cooled further to a cryogenic condition of -42.78 oC and 1615 kPa in another cooler E-101 before being passed to another phase of flashing in a second three-phase separator SEP 2 where the remaining water and liquid in the condensed gas stream are knocked off. The liquid and water from both separators are manifolded into MIX-100 and MIX-101, respectively for further cooling where necessary.

The combined liquid stream are used as feed and enters the lower feed tray [15th stage] to the Demethanizer. The cooled gas stream at -42.78 oC and 1615 kPa is divided into two fractions at ratio 4:1 in a splitter TEE-100. The bulk gas fraction, COL1 STREAM 1 goes for further cryogenic cooling in multi-exchanger LNG-100 in the same way as smaller gas fraction COL 1 STREAM 2 does to a much lesser extent. COL 1 STREAM 1 exits LNG-100 at -153 oC and 1600 kPa to be used as reflux to the 30-tray Demethanizer which is a reboiled absorber and it enters through the top stage [tray]. COL 1 STREAM 2 on the other hand at -95 oC and 1600 kPa enters the column at the 2nd stage to function as the top feed. The reboiled absorber operates at 1600 kPa and 1650 kPa at the top and bottom stage, respectively at corresponding temperature of -153 oC and 25 oC. Hydrogen, being the lightest component in the feed stream, is discharged as vapour from the top of the column to increase the purity of methane, which is the primary product of interest, is withdrawn as side product in vapour form from the 4th stage. Both methane and hydrogen overheads are passed through the multi-exchanger LNG-100 for cold heat recovery to the both reflux and top feed stream. Purity of the methane and hydrogen produced are 91 % and 48.6 %, respectively.

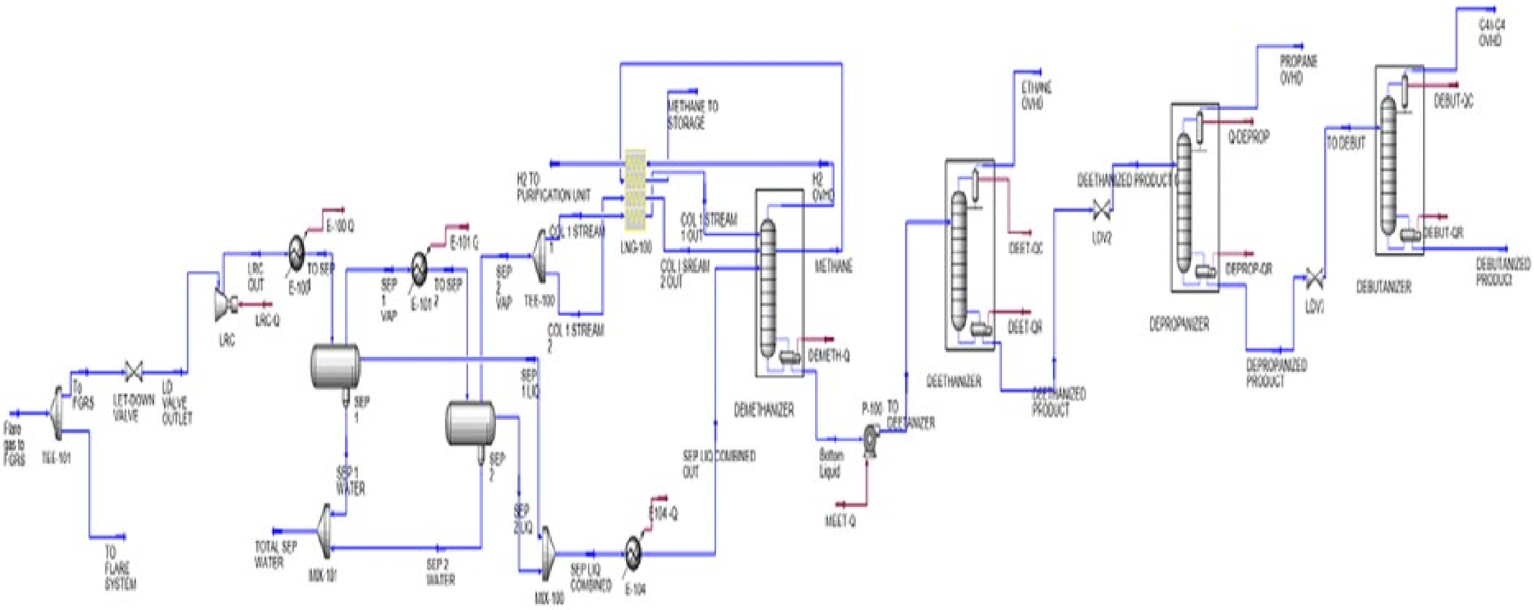

Figure 1: Simulated Flare Gas Recovery System for the Refinery under study

The Demethanized liquid is removed from the bottom of the column at 56.32 oC and 1650 kPa. The liquid are used as the feed to the Deethanizer entering through 13th stage after being pressurised by the pump, P-100 to 2600 kPa and heated to 54.72 oC. The Deethanizer is a 30-tray distillation column which operates at 2590 kPa, 1.21 oC at the top and 2700 kPa, 114.5 oC at the bottom. Ethane is discharged at 88.8 % purity as a vapour from the top of the column while Deethanized product removed from the bottom of the tower is the feed stream to the Depropanizer after pressure step-down to 1750 kPa in a let-down valve LDV2. The Deethanized product is separated into propane and Depropanized product in a distillation tower, Depropanizer which is also a thirty-tray column operating at 1700 kPa, 50.22 oC and 1750 kPa, 118.6 oC at the top and bottom end respectively. Propane of 98% purity is obtained at the top of the tower in form of vapour while Depropanized product obtained as liquid is the feed to the Debutanizer after pressure reduction to 550kPa in a let-down valve, LDV3. It enters the fractionator at the 14th stage. Depropanized liquid at 68.55 oC enters the 25-stage Debutanizer through the 12th stage where it fractionated into butane/isobutane vapour at the top end and Debutanized liquid at the bottom end. The column operates at 500 kPa, 45.55 oC at the top and 600 kPa, 104.7 oC at the bottom end. The Debutanized liquid is the gasoline fraction in crude stream [9].

Cost Estimation

Determining the Purchased Equipment Cost

This was carried out using Equation 1.

$C_e = a + bS^n$ (1)

Where; C_e is the purchased equipment cost on a Gulf Coast basis, Jan.2007 [CE index [CEPCI] = 509.7, NF refinery inflation index = 2059.1) [10]; a, b = constant [see, [10]; S is the size parameter, and n is the characteristic exponent for the equipment. The prices are for Carbon steel except where it is stated otherwise in the table.

In a situation where the size parameter falls outside the valid range, Equation 2 known as sixth-tenth rule was used to correct the limitation of Equation 1

$$C_2 = C_1 \left(\frac{S_2}{S_1}\right)^n \qquad (2)$$

C_2 is the cost of the equipment with Capacity S_2 and C_1 is the cost of the equipment with Capacity S_1. n is typically 0.6.

Equation 1 was used to determine the cost of the liquid ring compressor (LRC), demethanizer bottom liquid pump [P-100]. The Equation 2 was used to scale up the size parameter of the demethanizer bottom liquid pump (P-100). For the Trim coolers E-100, E-101, and E-104 using a U-tube shell and tube heat exchanger, the size parameter was the heat transfer area of the exhanger [m2] which was determined using Equation 3.

$$Q = UA\Delta T \text{ and } A = \frac{Q}{U\Delta T} \qquad (3)$$

$$A = \frac{Q}{U\Delta T}$$

where $\Delta T = LMTD = \frac{(T_{ih}-T_{oc})-(T_{oh}-T_{ic})}{ln\frac{(T_{ih}-T_{oc})}{(T_{oh}-T_{ic})}}$ for a counter-current flow; the approach temperature was assumed to be 7 oC and the inlet cooling fluid temperature was 25 OC

Vertical Three-Phase Separators

The cost of the separators (SEP 1 and SEP 2) was determined using Equation 1, and the size parameter was taken to be the thickness (t) and shell mas (SM). The thickness and shell mass were obtained using Equations 4 and 5

$$t = \frac{P_i D_i}{2SE - 1.2P_i} \qquad (4)$$

$$SM = \pi DLt\rho \qquad (5)$$

Where: t = wall thickness in m; D = Vessel diameter, m; S= allowable stress in N/mm2, E = weld efficiency, L = length of the vessel in m; ρ is the metal density in kg/m^3 (8000 kg/m^3), D = 3.4m, L=8.9m, take weld efficiency, E= 1 Allowable stress for Stainless steel at 100 ^{o}F is about 20ksi or 138N/mm^2 selected from allowable stress table B [10].

LNG-100

The LNG-100 was assumed to be a U-type of shell and tube heat exchanger. The cost of the LNG-100 was calculated using Equation 1, and the size parameter was the area of the heat exchanger (A), obtained using Eqaution 6.

$$A = \frac{U_A}{U} \qquad (6)$$

Where U_A = -133892.198046242 kJ/^{o}C-h and U = 3960kJ/h m^2 ^{o}C

Demethanizer

The demethanizer was estimated separately as a pressure vessel, tray and reboiler. The shell mass is the size parameter used for calculating the cost of a pressure vessel. The cost of the tray was obtained using Equation 1. For the pressure vessel, Equations 4 and 5 were used to calculate the shell mas before using Equation 1 to obtain the cost. The cost of the reboiler was estimated using area of the heat exchanger [Equation 3] as the size parameter and then Equation 1.

Deethanizer, Depropanizer and Debutanizer

The cost of deethanizer, depolarizer and debutanizer was estimated separately as a pressure vessel, tray, reboiler and condenser. The first three were obtained as in case of demethanizer, and the cost of the condenser was estimated as in the case of reboiler. The purchased cost of the equipment was summed in order to obtain the total purchased equipment cost. The calculations as mentioned earlier were based on the CEPCI index of 509.7 in Jan. 2007 [10]. The cost of the spare of each equipment was the same as the purchased equipment cost for all the equipment. Thus, the total purchased equipment cost was obtained by multiplying the purchased equipment cost by 2. In order to obtain the total purchased equipment cost as at October, 2022, the total purchased equipment cost was multiplied by 816.3 which was the CEPCI index in October, 2022 [10].

Estimating the Total Capital Investment [TCI].

The total capital investment [TCI] was obtained by summing the fixed capital investment and other outlays [11]. The fixed capital investment is the summation of the direct and indirect costs. The direct cost was calculated as the sum of the purchased equipment cost, offsite and physical plant cost, and indirect cost included the summation of design and engineering and contingencies.

The total physical plant cost [PPC] was calculated using the factors in Table 1 in Equation 7 [10]. The outlays included the working capital

PPC = PCE (1 + F1 + F2 + …….+F7 (7)

Table 1. Factors used in the determination of the total physical plant cost

Item factor		
F1	Equipment erection	0.3
F2	Piping	0.8
F3	Instrumentation	0.3
F4	Electrical	0.1
F5	Civil	0.2
F6	Structures and building	0.2
F7	Lagging and painting	0.1

The fixed capital cost (FC) was calculated using Equation 8 and factors in Table 2

PPC (1+F11) (1+F10+F12) (8)

The working capital cost (WC) was calculated as 20 % of FC G

The total capital investment (TCI) = FC + WC (9)

Table 2. Factors used in the determination of the fixed capital cost [10]

F10	Design and Engineering	0.30
F11	Offsites (OS)	0.30
F12	Contingencies	0.10

Estimating the Total Production Cost [TPC]

The total production cost was estimated by summing the direct production cost and the total general expenses. The direct production cost [DPC] was obtained by the summation of the total fixed cost [TFC] and the total variable cost [TVC]. The TFC was estimated by adding [S/N: 1-9], and TVC was calculated through the summation of [S/N: 10-12] in Table 3, respectively. The total general expenses was estimated by adding [S/N:13-15] in Table 3.

Table 3. Cost parameter asumptions used in the estimating the total production [TPC] [10]

S/N	Cost parameter	Range	Assumption
1.	Maintenance and repair cost (MTC)	5-10 % of FC	5 % of FC
2.	Operating labour cost (OLC)	6-20 % of TCI	15 % of TCI
3.	Laboratory charges	5-23 % of OLC	10% of OLC
4.	Supervision cost	10-20 % of OLC	15 % of OLC
5.	Plant overhead cost	5-15 % of OLC	10 % of OLC
6.	Capital charges	5-10 % of FC	6 % of FC
7.	Insurance	0-1 % of FC	1 % of FC
8.	Local texes	0-2 % of FC	1 % of FC
9.	Patent and Royalties	0-1 % of FC	1 % of FC
10.	Raw material cost (RMC)	10-50 % product cost	0
11.	Utilities	10-20 % of MTC	10 % of MTC
12.	Miscellaneous	10 - 20 % of MTC	10 % of MTC
13.	Sales expense		5 % of DPC
14.	Research and Development	2-4 % of Laboratory cost	4 % of Laboratory cost
15.	General overhead		5 % of DPC

Profitability Analysis

Anuual Revenue:- The annual revenue [income] was obtained by the summation of the earnings from the sales of the products such as methane, hydrogen, ethane, propane, and butane/i-butane. The profit before tax [PBT] and profit after tax [PAT] were calculated using Equations 10 and 11, respectively.

Profit before Tax (PBT) = Total income – Total production cost (10)

Profit after Tax (PAT) = PBT – Tax payable (11)

The rate of return (ROI), and the pay back period (PBP) were estimated using Equations 12 and 13, respectively.

$$\text{Rate of Return (ROI)} = \frac{PAT}{TCI} X\ 100 \quad (12)$$

$$\text{Pay Back Period (PBP)} = \frac{Total\ capital\ investment}{Profit\ after\ Tax} \quad (13)$$

Results and Discussion

Cost Estimation

Determining the Purchased Equipment Cost:- The purchased equipment cost helps in determining the overall cost of the recovery system. For each equipment, the purchased cost is presented in Table 4. The table shows that it would require a sum of two million, one hundred and sixty three thousand, nine hundred and thirty three dollars [$26,163,933] to purchase the equipment needed for the establishment of flare gas recovery system. This cost is doubled to fifty two million, three and twenty seven thousand, eight hundred and sixty six dollars ($52,327,866) if provision is to be made for spare parts of the equipment in order to ensure the availability and reliability of the system for optimum recovery. Considering the components of the flare gas recovery system which included compression, cooling, flashing and separation, the cost of compressor was 87.35 %. This means that the cost of compression is more expensive compared to the cost of other components of the system. This shows that cost of compression in the system is a controlling factor of the total purchased equipment cost.

Table 4. Purchased equipment cost

S/N	Equipment	Price ($)
1	Compressor	22854644
2	Pump	40897
3	Cooler E-100	629607
4	Cooler E-101	211875
6	Cooler E-104	46,945
7	Separator SEP 1	280,810
8	Separator SEP 2	112638
9	Cooler LNG-100	27,145
10	Demethanizer	107,116
11	Deethanizer	465,163
12	Depropanizer	722,525
13	Debutanizer	664,568
	Total	26163933

Estimating the Total Capital Investment

The result of the estimated total capital investment [TCI] of the flare gas recovery system is presented in Table 5. The table shows that to establish the kind of the system being proposed in this work, that it will require a TCI of five hundred and forty nine million and eighty eight thousand, one hundred and fifty eight dollars ($549,088,158) with about 83.3 % of the cost accruing from the fixed capital. The TCI is a determinant of both ROI and PBT because of dependence of their calculation on it.

Table 5. Estimated total capital investment [TCI]

S/N	Items	Cost ($)
1	Purchased equipment cost	52,327,866
2	Physical Plant Cost	251,413,992
3	Fixed Capital	457,573,465
4	Working Capital	91,514,693
5	Total Capital Investment	549,088,158

Estimating the Total Production Cost

As shown in Table 6, the total production cost of the flare gas recovery system is estimated to be two hundred and four million, six hundred and eighty one thousand, seven hundred and sixty two dollars [$204,681,762]. The table also shows that total variable cost is the least component.

Table 6. Estimated total production cost [TPC]

S/N	Items	Cost ($)
(A)	Fixed cost	
1	Maintenance and Repairs	22,878,673
2	Operating labor	82,363,224
3	Direct supervisory and Clerical labor	12,354,484
4	Laboratory Charges	8,236,322
5	Plant Overheads	8,236,322
6	Capital Charges	27,454,408
7	Insurance	4,575,735
8	Local Taxes	4,575,735
9	Patent & Royalties	4,575,735
	Total fixed operating cost (TFC)	175,250,638
(B)	VARIABLE COST	
1	Raw Material Cost	0
2	Utilities	8,236,322
3	Miscellaneous	2,287,867
	Total variable cost (TVC)	10,524,189
	Direct Production cost (DPC) = TFC +TVC	185,774,827
(C.)	GENERAL EXPENSES (GE)	

Sales Expenses	9,288,741
Research and Development	329,453
General Overheads	9,288,741
Total general expenses (TGE)	18,906,935
Annual total production cost (TPC) = DPC+TGE	204,681,762

Profitability Analysis

As shown in Table 7, the Profit After Tax (PAT), Rate of Return (ROI) and Pay Back Time (PBT) of the flare gas recovery system were determined to be $2,007,068,515, 365.50 % and 3.28 months, respectively. The huge profit after tax obtained shows that the investment would be lucrative. A very high ROI of over 360 % suggests that the investment would be of great benefit for any prospective investors. Comparing this with over 200 % obtained by shows that the current is more viable. The PBT of less than four months when compared to 1.02 years, and 1.6 years obtained by Respectively shows more viability, efficiency and enhanced liquidity status of the investment [6-8].

Table 7. Profitability analysis

Items	Amount/Value
Plant Operation	Continuous process
Plant Attainment Value for all sections	94.52 %
Annual Methane Production	530,747,460.14 litre
Annual Hydrogen Production	7,170,601.0114 kg
Annual Ethane Production	22,218,144 MMBtu
Annual Propane Production	1,538,931,180.8 Litre
Annual Butane/isobutane Production	2,124,624,522
Annual Debutanized product Production	437,898,796 litre
Market Selling Price of Methane	$1.31 per litre
Market Selling Price of Hydrogen	$16/per kg
Market Selling Price of Ethane	$1.36/MMBtu
Market Selling Price of Propane	$0.39 per litre

Conclusion

The economics analysis of a flare gas recovery system in a refinery in Nigeria has been conducted. The result shows that it would require a capital investment of about five hundred million dollars to establish such a venture. This cost is quite enormous, but, the proft margin after tax is greater with about one thousand five hundred million dollars owing to zero cost of the raw material [flare gas] as it is currently a waste in the refinery. The high rate of return of 365.50 % and low payback time of 3.28 months show that the investment is profitable, viable, feasible, bears less risk, and efficent. The low payback time also indicates improve liquidity position of the investment. Due to the profitability of this system, the refinery would benefit immensely from the installation and operation of the the flare gas recovery system through mitigation of the greenhouse gas emission, valorization of refinery waste [flare gas], and has health and enormous economic benefits.

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