Simulation of the Production of Liquefied Petroleum Gas (Lpg) from Flare Gas System

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Abstract

Gas flaring is associated with economic loss, health and environmental challenges, resulting from waste of valuable products, blood abnormalities and depletion of ozone layer, respectively. Therefore, there is a need for the recovery of these valuable products from gas flaring system. One of such products is the Liquefied Petroleum Gas [LPG] produced from the liquefaction of the natural gas which could be used as a source of energy. In this work, the maximization of LPG production from flare gas generated from a pretreated natural gas stream was carried out using HYSYS V8.6. Other products obtained from the process include C1, C2 and natural gasoline [C5+]. The results obtained show that the plant can recover 98.5 % LPG from the component of gas feed. This is a huge improvement on the 92 % recovery in a typical conventional plant. The result of the economic analysis show that the plant is economical and profitable. The LPG recovered can be used as a source of energy, and the process could help to mitigate climate change, minimize economic loss and health challenges.

Keywords: Liquefied Petroleum Gas; Flare gas; Simulation; Production Rate, Natural Gas

Introduction

Natural gas is a flammable mixture of hydrocarbon gases comprising predominantly of methane and other gases such as ethane, propane, butane and pentane. The gas is the cleanest, safest and the most important energy sources. It exists as associated [seen in oil fields] or non-associated [isolated in natural gas fields] gas [1,2]. This gas can be flared for safety reasons [relieve pressure] or in order to evacuate it from the well due to lack of gas processing and transportation infrastructures [3]. Gas flaring is currently perceived as a major challenging energy and ecological problem causing economic loss and hazardous effect to the atmosphere as flares may emit methane and other unstable organic compounds, which have shown to aggravate asthma. The flared gas is also carcinogenic and poisonous resulting to blood abnormalities, and the emitted particulate matter may cause cancer and heart attack [4]. In 2018, World Bank reported that approximately US$20 billion was lost to gas flaring globally with Nigeria losing N233 billion based on the estimation by PricewaterhouseCoopers. The quantity of gas flaring in Nigeria has been on decline since 2012 and was 10% in 2018. This decline is probably due to increase in reinjection, utilization of the gas and global policy on gas flaring. The volume of gas flared in Nigeria in 2018 was 7.4 billion cubic feet placing the country on the 10th position among the gas flaring countries in the world, those in the first five position are Russia, Iraq, Iran, United States and Algeria [5].

Nevertheless, environmental and economic conditions have increased the need to mitigate the quantity of gas being flared. The recovery of flared gas reduces noise, thermal radiation, operation and maintenance cost, reduction in air contamination and gas emission. Several steps are employed today worldwide to reduce flared gas loss, such as adequate operation and maintenance of flare gas systems, changing start-up and shut-down, methods [6]. Eliminating leaking valves, adequate use of hydrocarbon gases needed to properly maintain process of flaring and enhanced management of stream to get smokeless burning are also employed in reduction of flared gas losses. The reduction or recovery of flared gas can be carried out by collection, compression and injection/reinjection; collection and delivery to a nearby gas-gathering systems; use as an on-site fuel source; use as feedstock for other petrochemical production; generating electricity, and conversion of gas to liquid [Liquefied Petroleum Gas].

However, the conversion of flared gas to LPG is preferred because, it is safe, cost effective, provides opportunity to increase the quantity of gas transported and stored, produces various byproducts such as methane, ethane and natural gasoline which are important feedstocks in the petrochemical industry [7]. Various researchers have worked on recovering LPG from flare gas and notably among them are simulated the LPG recovery system us-

ing four separate columns resulting to products such as C3, i-C4, n-C4, CH3, C2 and natural gasoline. The demerit of this configuration was a relative high capital cost. While simulated the LPG process using three columns including deethanizer, depropanizer and debutanizer, and pretreated natural gas as the feed stream given rise to reduction in the capital cost. The current work is focused on the maximizing the LPG production from flare gas using HYSYS to simulate the process with a pretreated natural gas stream as the feedstock. The Deethanizer and Demethanizer were merged to reduce the capital cost and increase the quantity of the products produced [8].

Materials and methods

Aspen HYSYS Version 8.6 was used to simulate the LPG recovery process, carry out the material and energy balances, equipment sizing and rating, cost estimation [fixed cost and working capital cost [assumed to be 15% of the fixed capital cost]. The feed was a pretreated flare gas obtained from Port Harcourt Refinery Eleme and the compositions are presented in Table 1.

Table 1 Composition of Feed Gas

Property	Feed
Temperature (°C)	25
Pressure (bar)	22.75
Mass Flow(kmol/hr)	5177
Composition (Mole Fraction)	
Methane (C_2H_4)	0.7252
Ethane (C_2H_6)	0.1176
Propane (C_3H_8)	0.0750
Iso-butane (C_4H_{10})	0.0204
Normal-butane (C_4H_{10})	0.0197
Iso-pentane (C_5H_{12})	0.0147
Normal-pentane (C_5H_{12})	0.0102
Normal-hexane (C_6H_{14})	0.0037
Normal-heptane (C_7H_{16})	0.0047
NormalOctane (C_8H_{18})	0.0027
Carbondioxide (CO_2)	0.0018
Nitrogen (N_2)	0.0043

Process description

The feed gas was first cooled from 25 °C to -65°C using a cooler to enable effective separation of vapour and liquid in the column. It was then charged to the de(M)ethanizer column at 2270 kpa, where methane and ethane were separated out as the top product at -66.34 °C, NGLs as base product at 94.87°C. The NGLs from the De(M)ethanizer were first charged to a pressure control valve to decrease the pressure from 2270 kpa to 1770 kpa. This was done to enable the light components from the NGLs to partially vapourize. It was then passed through a cooler to decrease the temperature from 83.64 °C to a design temperature of 65 °C before it was channeled to the Depropanizer, where C3 (LPG) having a mass flow of 694.9 kg/h was sep-

arated out as top product at 49.80 °C and 1700 kpa operating pressure, while butane flows through the bottom at 122.4°C and 1769 kpa operating pressure. The C4 from the Depropanizer is charged to a pressure control valve to decrease pressure from 1769 kpa to 1600 kpa, before charging it to a Debutanizer. The Debutanizer having an operating pressure of 1600 kpa and 117.7 °C temperature separates C4 and C5 as top and bottom products, respectively. Butane separates out through the top at 89.32 °C and 1585 kpa, while pentane flows through the bottom at 154.9 °C and 1600 kpa. The C4 from the debutanizer was therefore charged to the butane Splitter, where butane was separated to iso-butane at -4.588 °C and 150 kpa and normal-butane at 50.58 °C and 500kpa (See Fig. 1)
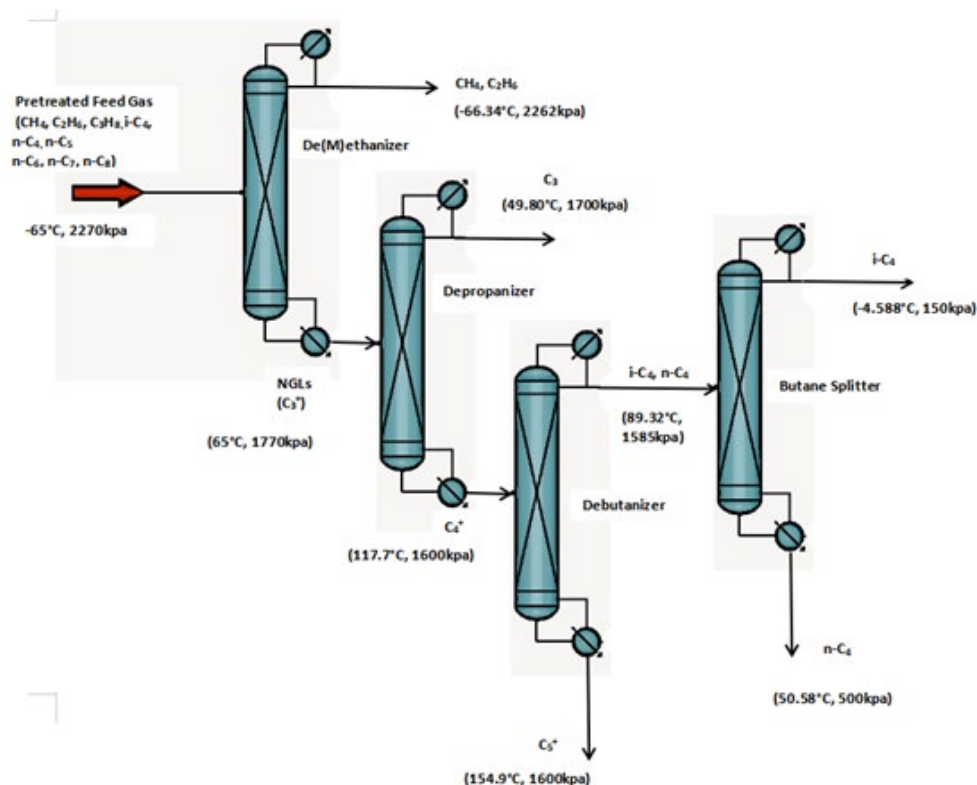
Figure1: Simulation of LPG from flare gas

Results and Discussion
Simulation of the LPG from flare gas
This was carried out using Aspen HYSYS Version 8.6 as described in section 2.0. The process parameters temperature and pressure, mass flowrate, compositions and heat flow of the various streams obtained are presented in Table 1. As can been see from Table 1, the feed was supplied to De[M]ethanizer at a flow rate of 5177 kg/hr, temperature of -90 °C and pressure of 2270 kpa. Due to difference in boiling point of the feed component, ethane and methane were separated as the overhead product at

Table1. Stream Parameters

Units		Feed	De(M)eth Ovhd	De(M)eth Btm	Deprop Feed	Deprop Ovhd	Deprop Btm	Debut Feed	Debut Ovhd	Debut Btm	Splitter Feed	Splitter Ovhd	Splitter Btm
Temp	°C	-90	-66.34	94.87	65	49.80	122.4	117.7	89.32	154.9	44.2	-4.588	50.58
Major Comp		C1	C1, C2	C3+	C3+	C3	C4	C4	C4	C5+	C4	i-C4	n-C4
Pressure	Kpa	2270	2262	2270	1770	1700	1769	1600	1585	1600	695.0	150.0	500
Mass flow	kg/h	5177	3329	1848	1848	694.9	1153	1153	485.8	666.9	485.8	249.4	171.1
Heat flow	kJ/h	-20280000	-14870000	-439200	-4558000	-1839000	-2558000	-2558000	-1181000	-1360000	-1244000	-6754000	-4251000
C1		0.7252	0.8543	0	0	0	0	0	0	0	0	0	0
C2		0.1176	0.1384	0.0006	0.0006	0.0012	0	0	0	0	0	0	0
C3		0.0750	0.0001	0.4958	0.4958	0.9850	0.0449	0.0449	0.0898	0	0.0898	0.1018	0
i-C4		0.0204	0	0.1350	0.1350	0.0123	0.2481	0.2481	0.4767	0.0202	0.4767	0.6927	0.0920
n-C4		0.0197	0	0.1304	0.1304	0.0015	0.2492	0.2492	0.4137	0.0850	0.4137	0.2054	0.8500
i-C5		0.0147	0	0.0973	0.0973	0	0.1870	0.1870	0.0159	0.3576	0.0159	0	0.4654
n-C5		0.0102	0	0.0675	0.0675	0	0.1297	0.1297	0.0039	0.2552	0.0039	0	0.0115
De(M)eth - De(M)ethanizer; Deprop - Depropanizer; Debut - Debuthanizer; Ovhd - Overhead; Btm - Bottom													

-66.34°C and 2262 kpa operating pressure. 85.43% of C1 and 13.84% of C2 were recovered while 49.58 % C3, 13.50 % i-butane, 13.04 % n-butane, 9.73% i-pentane and 6.75% n-pantane were recovered at the bottom. Small amount of C3+ was produced and there was no production of C4+ in the De[M]ethanizer overhead. It was observed that a higher percentage of the lighter component was recovered at the top, while the heavier components were recovered at the bottom. [See Fig. 2].
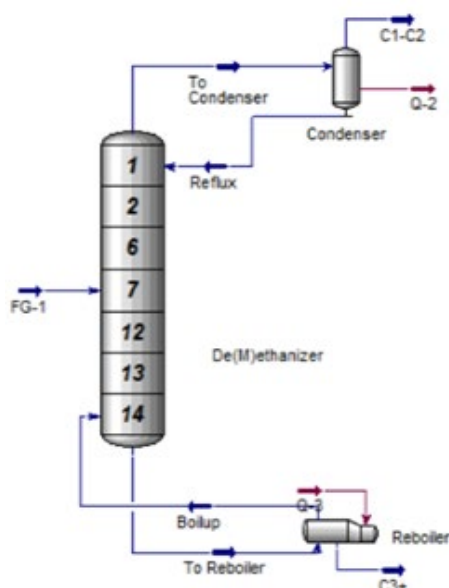
Figure:2 HYSYS simulation – De[M]ethanizer

Figure:3 HYSYS simulation – Depropanizer

In the depropanizer, 1848 kg/hr of C3+ was fed to the column at 65 °C and 1770 kpa. LPG [propane] was purely recovered as the overhead product at 49.80 °C and 1700 kpa, having a molar composition of 98.50 %. 1.23% of i-butane was also recovered at the overhead with small production of ethane as well. C4+ was collected as bottoms product at 122.4 °C and 1769 kpa having a composition of 4.49 % propane, 24.81 % i-butane, 24.92 % n-butane, 18.70% i-pentane and 12.97% n-pentane. It was observed that some LPG [propane] was still recovered underneath the depropanizer and there was no production of C1 and C2 at the bottom of the depropanizer [See Fig. 3]. Butane [C4+]
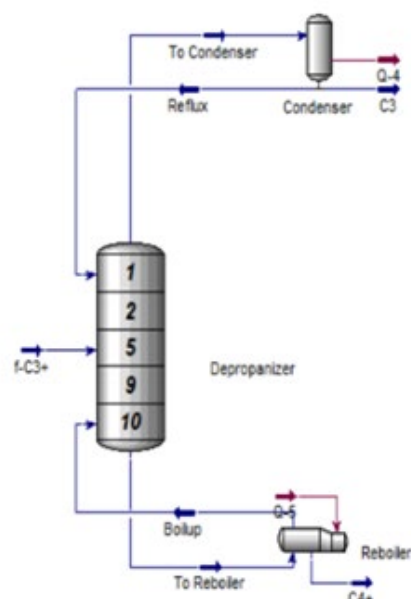
was fed into the debutanizer at 117.7 °C and 1600 kpa, where i-C4 and n-C4 were recovered as the overhead product at 89.32 °C and 1585 kpa having a molar composition of 47.67 % i-C4 and 41.37 % n-C4 with some traces of propane with a molar composition of 8.98 %. There was no production of ethane and methane at the overhead of the debutanizer. At the bottom of the Debutanizer, C5+ was separated out at 154.9 °C and 1600 kpa having a composition of 2.02 % i-C4, 8.50 % n-C4, 35.76% i-C5 and 25.52% n-C5 [see Fig. 3]. There was no production of propane at the bottom.
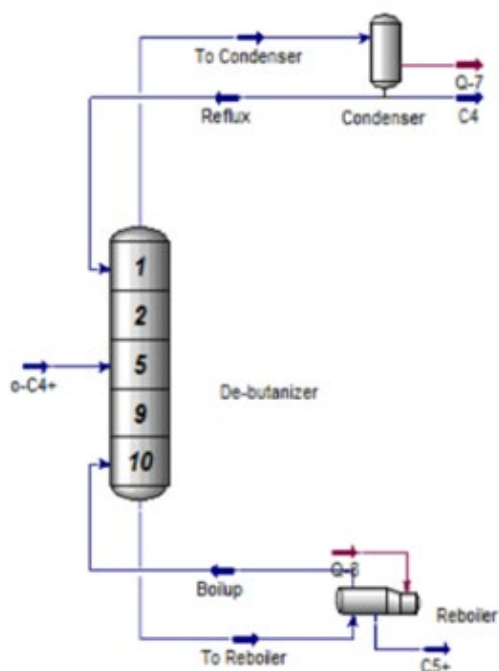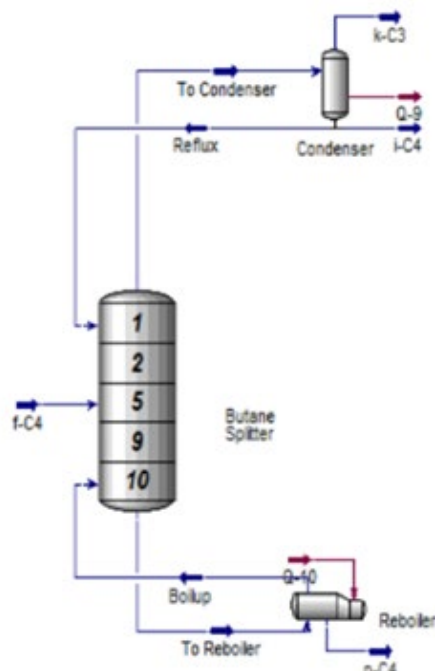

Figure:4 HYSYS simulation-Debutanizer

Figure:5 HYSYS simulation-Butane Splitter

Butane was fed into the butane Splitter at 44.2 °C and 695 kpa where the entrained propane [LPG] from the debutanizer overhead having a composition of 10.18 % and i-C4 with a composition of 69.27 % were separated at the top with operating conditions -4.588 °C and 150 kpa. It was observed that 20.54 % n-C4 was also recovered at the top, but sent back to the column to be separated at the base. Underneath the butane Splitter, 85 % n-C4, 46.54 % i-C5, 9.20 % i-C4 and 1.15% n-C4, were separated out at 50.58 °C and 500 kpa. There was no production of methane, ethane and propane at the bottoms of the butane Splitter [Fig. 5].

Equipment Sizing

The sizing of the equipment in the plant was done using HYSYS and the results obtained are presented in Table 2. From the table, the number of stages for the De(M)ethanizer was 14, and Depropanizer, Debutanizer and butane Splitter had 10, respectively. Other parameters of the each column include diameter [1.5 m], tray spacing [0.55 m], tray volume [0.9719 m3], hold up [8.836E-02 m3] and weeping factor [1.0] [see Table 2].

Table 2. Equipment Sizing of columns

Columns	Diameter (m)	No of Stages	Tray Spacing (m)	Tray volume (m3)	Hold up (m3)	Weeping Factor
De(M)ethanizer	1.5	14	0.55	0.9719	8.836E-02	1.0
Depropanizer	1.5	10	0.55	0.9719	8.836E-02	1.0
Debutanizer	1.5	10	0.55	0.9719	8.836E-02	1.0
Butane Splitter	1.5	10	0.55	0.9719	8.836E-002	1.0

The number of reboilers and condensers required in the simulation process was four respectively. The dimensions of the reboilers and condensers were the same with each possessing a diameter [1.193 m], length [1.789 m] and volume [2 m3].

Equipment Rating

The rating of the coolers and pressure control valves was conducting using HYSYS and the results obtained are presented in Tables 3 and 4. Three coolers were used and there was significant change in temperature, pressure and heat duty along the coolers. In the first cooler, it was observed that feed [pretreated flare gas] temperature was reduced from 25°C to -65 °C, resulting to a -90 °C change in temperature. A variation in pressure from 2275 kpa to 2700 kpa resulting to a 5.00 kpa pressure drop was observed. The quantity of heats supplied was 1.787E+007 KJ/h. In the second cooler,

Table 4. Equipment rating of coolers

Parameters	Cooler 1	Cooler 2	Cooler 3
Change in Temp (°C)	-90	-18.64	-9.331
Pressure drop (kpa)	5	0	5
Duty (KJ/h)	1.787E+007	1.669E+005	6.295E+004

feed (C_3^+) temperature was reduced from 83.64 °C to 65 °C resulting to a reduction in temperature of -18.64 °C. There was no change in pressure, and quantity of heat supplied was 1.669E+005 KJ/h. In the third cooler, feed (C_4) temperature changed from 53.53 °C to 44.20 °C, leading to a decrease in temperature of -9.331 °C.There was also variation in pressure from 700 kpa to 695 kpa resulting to a 5 kpa pressure drop. Quantity of heat 6.295E+004 KJ/h (see table 3).

Similarly, three pressure control valves (PVC) were used in the simulation to decrease the pressure of the components to meet the product specification. The first PCV reduced the feed (C_3^+) pressure from the de(M)ethanizer at 2270 kpa to 1770 kpa, resulting to a pressure drop of 500 kpa. The mass flow rate of C_3^+ was 32.72 kgmole/hr, and the valve opening was 50 %. The second PCV from the depropanizer reduced the feed (C_4^+) pressure from 1769 kpa to 1600 kpa resulting to a pressure drop of 169 kpa. The mass flow rate of (C_4^+) was recorded as 17.03 kgmole/hr and the valve opening was 50 %. The third PCV from the debutanizer reduced the feed (C_4) pressure from 1585 kpa to 700 kpa resulting to a pressure drop of 88.5kpa. Mass flow rate was 8.502kgmole/hr and valve opening of 50 % were obtained (see Table 4).

Table 5. Equipment rating for pressure control valves (PCV)

Parameters	PCV-1	PCV-2	PCV-3
Pressure Drop (kpa)	500.00	169.00	885.00
Mass flow (kgmole/hr)	32.72	17.03	8.502
Valve opening (%)	50	50	50

Flowsheet Calculation

The mass balance on the plant was carried out using HYSYS and the results obtained are presented in Table 6. The total mass flow rate of the flare gas into the system was 5177 kg/h and the mass flowrate out of the system from six different streams; De[M]ethanizer overhead, Depropanizer overhead, Debutanizer bottom, butane splitter distillate was 5176.608188 kg/h. Comparing the inlet and outlet mass flowrate, 0.391812 kg/h was obtained resulting to a percentage error of 0.007% which is not appreciable.

Table 6. Quantity of material flow through the system

Mass in		Mass out	
Stream	Mass flow rate (kg/h)	Stream	Mass flow (kg/h)
Flare gas	5177	De(M)eth Ovhd	3328.937267
		Deprop Ovhd	694.9360189
		Debut bottom	666.8892814
		n-C4	171.1047307
		i-C4	249.4159022
		Butane splitter dis	65.32498802
Total	5177	Total	5176.608188
Difference = Mass In - Mass Out = 0.391812			
% Error = (0.391812/5177)*100 = 0.007%			
The percentage error is smaller than 1% and thus acceptable			

Energy Balance

The overall energy flow into the system from the flare gas and the reboilers was 10692276.45 kJ/h while the energy out of the system from the condensers and the coolers was 10679313.82 kJ/h. The difference in energy was calculated and resulted to 12962.63 KJ/h, having a percentage error of 0.12%, which is smaller than one and thus acceptable [see Tables 4.8].

Table 4.8: Heat balance for the inlet and outlet streams

Inlet Streams	Heat Flow (kJ/h)	Outlet Streams	Heat Flow (kJ/h)
Flare gas	-18492832.72	Cooler 1Q	1787008.636
De(M)eth RebQ	3191605.4	De(M)eth CondQ	2173986.489
Deprop RebQ	18188658.32	C1, C2	-14870930.1
Debut RebQ	2028290.357	Deprop CondQ	18011927.02
But Split RebQ	5776555.087	C3	-1838849.686
		Cooler 2Q	166915.3779
		Debut CondQ	2013978.211
		C5+	-1359969.878
		But Split CondQ	5786821.476
		n-C4	-425089.4767
		i-C4	-675385.8889
		k-C3	-154044.2307
		Cooler 3Q	62945.87672
Total	10692276.45	Total	10679313.82

Cost Estimation

The economic analysis of the plant in terms of the fixed capital, working capital, start up cost, annual revenue and profit was conducted to determine if the LGP production is profitable. The results show that the fixed capital, working capital, start up cost and profit were $5,205,200.00, $5,205,200.00, $780780.00, $5,985,980.00 and $95,182,350.60, respectively. With an annual profit of $95,182,350.60, the plant is considered profitable. When compared with result obtained by, the current plant is cheaper and more cost effective [8,9-22].

Conclusion

The simulation of the production of LPG from flared gas system has been demonstrated using HYSYS. The plant has the capacity to effectively recover 98.5 % LPG content leading to high production of LPG, while giving flexibility to product blending by separating propane, iso-butane and normal-butane product streams. Based on the economic analysis, the plant is considered to be profitable and economical. The recovery of LPG from flared gas system is a value addition and could help to reduce the quantity of gas being flared, thermal radiation, operation and maintenance cost, and air contamination.

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