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Investigating the Cause of Excessive Triethylene Glycol Loss during Dehydration of Natural Gas (A Case Study)

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ABSTRACT

The simulation model of Imo River AGG plant was done using Aspen Hysys process simulator. The operating data for the simulation was obtained from the above mentioned plant. This work established the optimum operating temperature and pressure that will give the least amount of TEG lose. Material and Energy Balance for the plant model was done using the principles of conservation of mass and energy. The plant was assumed to be operating at steady state condition. Sizing of major equipment such as contactor having a diameter and height of 0.6096m and 6.034m, was carriage out. Sensitivity analysis to determine the effect of process parameters such as temperature, pressure, density and molecular weight on number of trays in the contactor. Temperature was observed to be decreasing with the number of trays, pressure was constant because pressure of the inlet gas and inlet TEG were relatively close. Process optimization was carried out to determine the optimum temperature (35^0C) and the corresponding TEG loss (1.2705kg/hr), also the optimum pressure 6400kpa with a corresponding TEG loss (1.0782kg/hr). The simulation done assisted greatly in determining the optimum values, however, further research should be done using advanced process simulators such as Aspen Hysys Plus, Unisim, and Prosim.

1. INTRODUCTION

The Imo River Gas Gathering Plant has been designed and constructed as a part of the Oyigbo Node Associated Gas Gathering Project with a primary function of utilizing associated gas as a means of reducing flaring. The infrastructure provided by the project enables the associated gas from current and planned developments in the Oyigbo Node area to be incorporated into the system for utilization for this purpose. The compression systems provided enable the existing customers and the gas lift systems to be satisfied. In total, a maximum of 113 MMscf/d can be supplied to sales customers and approximately 45 MMscf/d to gas lift facilities (Ahmed *et al*., 2019).

The basic concept of the Oyigbo Node Associated Gas Gathering Project is to make gas available for gas lift and sales gas customers from a common pipeline. The three compression plants at Imo River, Agbada, and Oyigbo (located near existing gas lift plants) supply gas into the pipeline at 73.8 bars. The Imo River AG Plant is similar in design to the Oyigbo and Agbada AG Plants. The Imo River Plant consists of inlet facilities (pig receiver and slug catcher), compression facilities (three stage centrifugal compressor driven by a Solar Mars T-100 gas turbine), dehydration facilities (TEG contactor) and export facilities (flow and quality monitoring) (Alay *et al*., 2014).

The associated gas is gathered into the AG plants at two pressures, high pressure (HP @ approximately 13 bars) and low pressure (LP @ approximately 4 bars) corresponding to the operating pressures of the oil separators in the flow stations. Gas gathered from flow stations remote from the AG plants is transported at HP separator pressures, the LP gas being compressed in a local gas-engine driven reciprocating compressor. The compressor has two inlet streams, into both the first and second stages. The LP gas from the local flow station (Agbada II) is fed into the first stage. The HP gas from the local flow station (Nkali station) is fed into the second stage (Ahmed *et al*., 2019).

The compressor third stage discharge pressure is fed to TEG contactor. The TEG contactor outlet is controlled at a water dewpoint of minus 10 ^{0}C to prevent any hydration or corrosion problems in the downstream systems. Failure of the TEG regeneration system will cause the gas export from the plant to stop. The export metering facilities into the pipelines consist of field metering only based at Oyigbo. The liquids generated in the AG plants are metered and disposed of directly into the crude oil export pipelines downstream of the neighbouring flow station (Abdulrahma *et al*., 2014).

Gas to be sold is collected at the Oyigbo Sales Manifold Facility which is a stand-alone facility located adjacent to the Oyigbo AG Facility. Overall co-ordination for sales gas distribution is centred at Oyigbo North. The overall operational concept is based on the premise of individual AG plants operating at maximum throughput depending upon the gas availability at the flow stations. Each AG plant contains facilities to monitor and control its own operation in addition to the flow stations connected to it (Muhammad *et al*., 2020).

2. MATERIALS AND METHODS

2.1 General Material Balance Equation

In developing the model equations, the mass balance equations was derived using the principles of conservation of mass which states that matter is neither created nor destroyed but can only be converted from one form to another.

$$\begin{pmatrix} Rate of accumulation \\ of\ material\ within\ the \\ Unit\ opertation \end{pmatrix} = \begin{pmatrix} Rate\ of\ inflow \\ of material\ into\ the \\ Unit\ operation \end{pmatrix} - \begin{pmatrix} Rate\ of\ outflow \\ of\ material \\ from\ unit\ operation \end{pmatrix} \pm \begin{pmatrix} Rate\ of generation\ or \\ deplection\ of\ speciesi \\ by\ chemical\ reactor \end{pmatrix} \quad (1)$$

For a steady state process equation (1) is written as:

$$\begin{pmatrix} Rate\ of\ inflow \\ of\ material\ into\ the \\ Unit\ operation \end{pmatrix} = \begin{pmatrix} Rate\ of\ outflow \\ of\ material \\ from\ unit\ operation \end{pmatrix}$$

(2)

2.2 Material Balance over the Separation

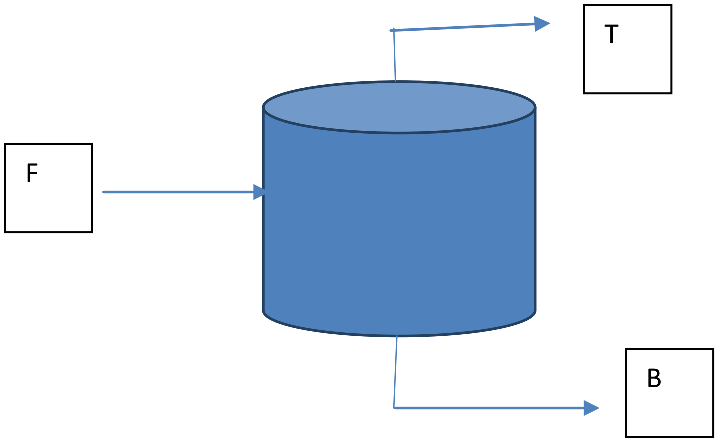

Figure 1: Material Balance over Separator

Taking Material Balance over the entire Separator

$$F = T + B \quad (3)$$

Taking Material Balance over each Component

$$F(X_F) = T(X_T) + B(X_B) \quad (4)$$

2.3 Material Balance over Contactor (Absorber)

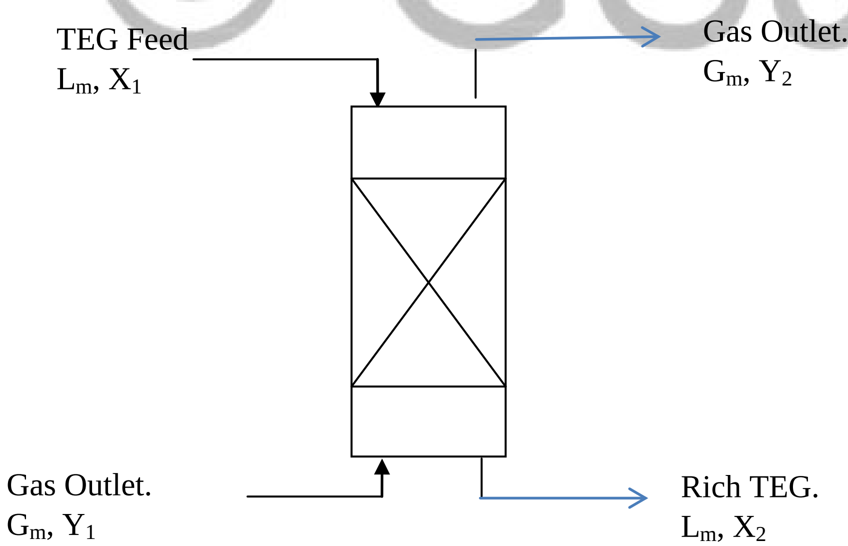

Figure 2: Material Balance over Contactor

Material Balance around the Absorber

$$G_m Y_1 + L_m X_2 = G_m Y_2 + L_m X_1 \quad (5)$$

$$G_m (Y_1 - Y_2) = L_m (X_1 - X_2) \quad (6)$$

2.6 Material Balance over Distillation Column

The regenerator is provided to recover the contaminated glycol for reuse in the glycol contactor. The rich glycol is removed from the contactor of level control, flows from the bottom of the contactor is routed to glycol flash drum and portion of the stream is directed through the still reflux condenser. The schematic diagram of regenerator is shown in the figure 3

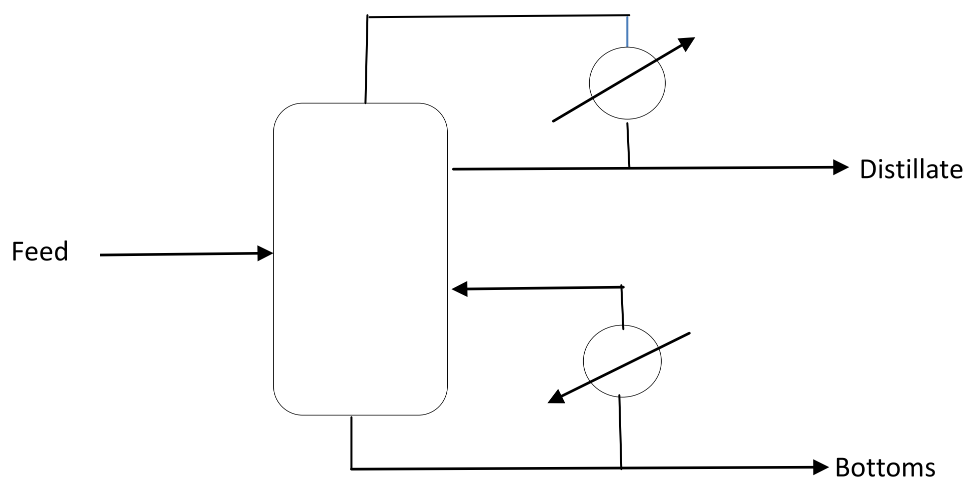

Figure 3: A Schematic Diagram of a Regenerator

Molar Flow Rate of Feed

The product from the decanter is given in terms of volumetric flow rate, so it will be converted to molar flow rate using the below formula:

Mass flow rate = density * volumetric flow rate

Molar flow rate = $\frac{Massflowrate}{molarmass}$ (7)

$F = \frac{\rho_{mix}}{M} V_o$ (8)

The volumetric flow rate of the feed to the Reactor = the volumetric flow rate of the feed of the product from the Reactor.

$\rho_{mix} = x_{fB}\rho_B + x_{fM}\rho_M + x_{fD}\rho_D$ (9)

Mass flow rate of the feed the Reactor = the mass flow rate of the feed of the product from the Reactor.

$M = x_{fB}M_B + x_{fM}M_M + x_{fD}M_D$ (10)

2.7 Determination of Distillate and Bottom Flow Rates

Substituting the assumptions for the distillation column into the general material balance equation as shown in equation 1, we obtained;

$$\begin{pmatrix} Rate\ of\ input \\ of component \\ i \end{pmatrix} = \begin{pmatrix} Rate\ of output \\ of\ component \\ i \end{pmatrix} \tag{11}$$

Taking overall balance of Figure 2

F = D + B (12)

From equation (12)

B = F –D (13)

Taking component balance with respect to benzene (B)

$$x_{fB}F = y_B D + x_{BB}B \tag{14}$$

Putting (13) into (14)

$$x_{fB}F = y_B D + x_{BB}(\mathrm{F} - \mathrm{D}) \tag{15}$$

Simplifying equation (15) and making D the subject of formula we have;

$$D = F(\frac{x_{fB} - x_{BB}}{y_B - x_{BB}}) \tag{16}$$

Taking balance around the condenser

$$V_n = L_n + D \tag{17}$$

But:

$$R = \frac{L_n}{D} \tag{18}$$

From equation (3.18)

$$L_n = R \times D \tag{19}$$

2.8 Determination of Upper and Lower Operating Line Equation

The upper operating line equation as proposed by Lewis – Sorel is given as;

$$Y_n = \frac{L_n}{L_n + D} x_n + \frac{D}{L_n + D} x_d \tag{20}$$

Similarly, the lower operating line equation is given by;

$$Y_m = \frac{L_m}{V_m} x_m + \frac{W}{V_m} x_{BB}$$

(21)

2.9 Process Description and Flow Diagram

The base case process flow diagram for Imo River ASS plant and Aspen Hysys are presented in figure 3.4 and figure 3.6

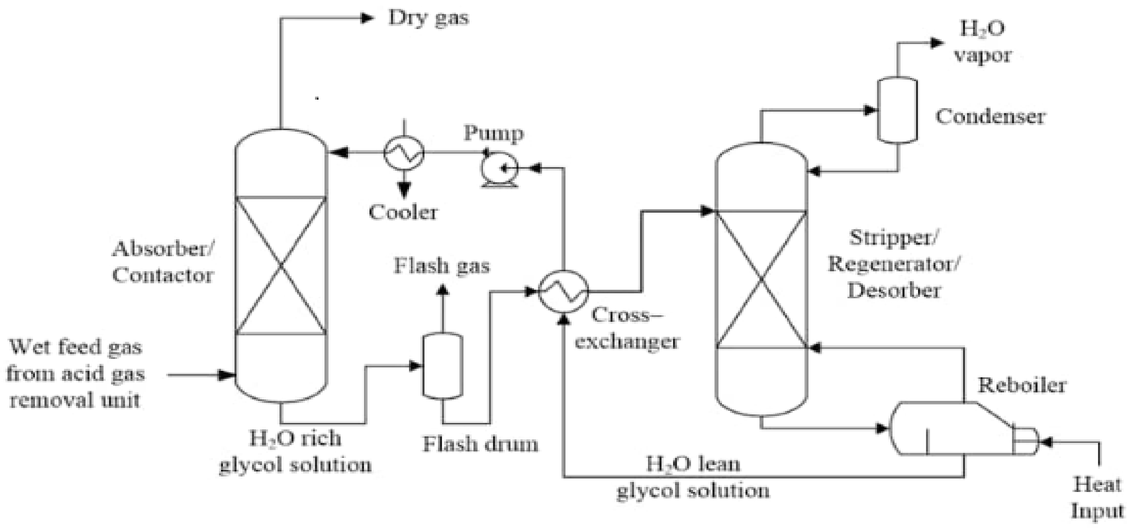

Figure 3.1: Base Case Process Flow Diagram for Imo River AGG Plant

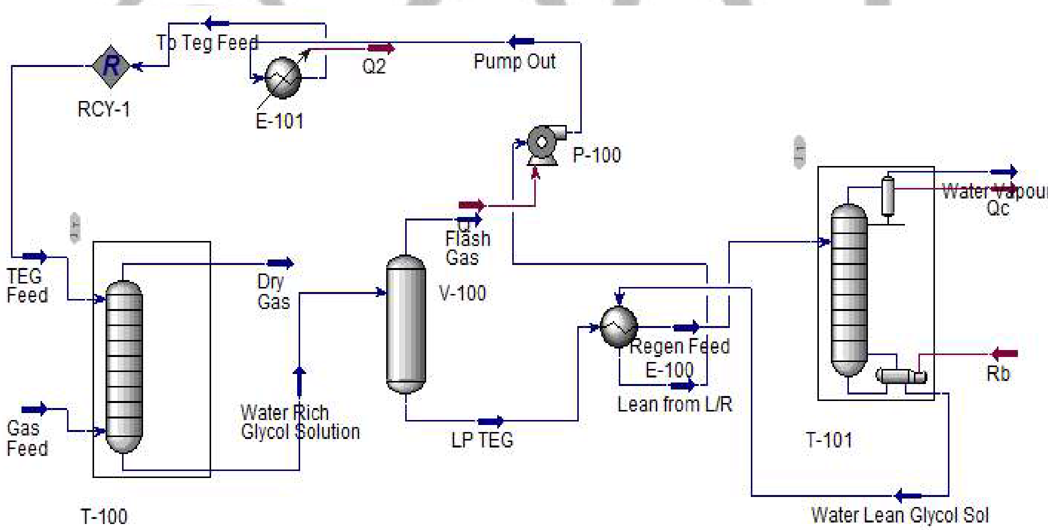

Figure 3.2: Aspen Hysys Process Flow Diagram

Lean, Water-free glycol (purity > 99%) is fed to the top of an absorber (also known as a "glycol contactor") where it is contacted with the wet natural gas stream. The glycol removes water from the natural gas by physical absorption and is carried out at the bottom of the column. Upon exiting the absorber the glycol stream is often referred to as "rich glycol'. The dry natural gas leaves the top of the absorption column and is fed either to a pipeline system or to a gas plant. Glycol absorbers can be either tray columns or packed columns.

After leaving the absorber, the rich glycol is fed to a flash vessel where hydrocarbon vapors are removed and any liquid hydrocarbons are skimmed from the glycol. This step is necessary as the absorber is typically operated at high pressure and the pressure must be reduced before the regeneration step. Due to the composition of the rich glycol, a vapour phase having high hydrocarbon content will form when the pressure is lowered.

After leaving the flash vessel, the rich glycol is heated in a cross-exchange and fed to the stripper (also known as a regenerator). The glycol stripper consists of a column, an overhead condenser, and a reboiler. The glycol is thermally regenerated to remove excess water and regain the high glycol purity. The rich glycols are used in heat transfers and cooling. It provides a variety of heat transfer characteristics; it also prevents the water from freezing at low temperatures within the piping system.

The hot, lean glycol is cooled by cross-exchange with rich glycol entering the stripper. It is then fed to a lean pump where its pressure is elevated to that of the glycol absorber. The lean solvent is cooled again with a trim cooler before being fed back into the absorber. This trim cooler can either be a cross exchanger with the dry gas leaving the absorber or an air-cooled exchanger.

3. RESULTS AND DISCUSSION

3.1 Material Balance from Hysys for Absorber

Table 1 shows the summary result of material balance composition for each component obtained from the result of Aspen Hysys simulation for Absorber. The feed data for the simulation was obtained from Imo River AGG plant and it consist of two inlet streams (NG and TEG Feed) and two outlet streams (Dry Gas and Rich TEG). The Wet feed gas Stream consist of mostly methane i.e 0.8980 mol% methane while the inlet TEG Feed contains mostly TEGlcol 0.999 mol % . The First outlet stream is Dry Gas and has 0.8981 mol % methane due to the dehydration of the natural gas by the TEG while the second outlet stream is Rich TEG which contains 0.906998 mol % of absorbed water.

Table 1: Result Summary of Material Balance from Hysys for Absorber

Equipment	Components	Inlet Composition		Outlet Composition	
		TEG Feed	Wet Feed Gas	Dry Gas	Water Rich Glycol Solution
Absorber	Nitrogen	0.0000	0.0009	0.001005	3.04E-06
	CO_2	0.0000	0.0037	0.02853	0.000861
	H_2S	0.0000	0.0005	0.015569	0.001981
	CH_4	0.0000	0.8234	0.89806195	0.000882
	Ethane	0.0000	0.0203	0.031144	2.24E-05
	Propane	0.0000	0.0062	0.014869	1.84E-05
	i-Butane	0.0000	0.0016	0.005927	6.66E-05
	n-Butane	0.0000	0.0020	0.003014	9.97E-06
	i-Pentane	0.0000	0.0009	0.001005	3.42E-05
	n-pentane	0.0000	0.0005	0.000502	8.16E-06
	TEGlcol	0.9990	0.0000	5.00E-08	0.089114
	H2O	0.0010	0.1400	0.000373	0.906998

Table 1 shows the composition of each component material balance obtained from the result of Aspen Hysys simulation for heat exchanger unit. The heat exchanger consist of two inlet streams: LP TEG which contains mostly water about 0.9 mol% and very little amount of other components, H_2O Lean Glycol Solution which contains mostly triethylene glycol(TEG) about 0.8 mol% and 0.19 mol% water as well as other components in minute concentrations. It also has two outlet streams which are: Regenerated feed which contains mostly water about 0.9 mol% and other components in trace amounts, Lean from LR which contains more of triethylene glycol about 0.8 mol% and 0.19 mol% water as well as other components in small amounts. A careful examination of Table 2 reveals that the composition of H_2O lean Glycol solution and Lean from LR are equal while that of LP TEG and Regen feed are also equal, hence, the results are in agreement with the principles of conservation of mass for a steady state process which states that total inflow of mass is equal to the total outflow of mass (Anyadiegwu *et al.*, 2014).

Table 2: Result summary of Material Balance from HysysFor Heat Exchanger

Equipment	Components	Inlet Compos		Outlet Composition	
		LP TEG	H2O LeanGlycol Solution	Regenerated Feed	Lean from LR
Heat Exchanger 1	Nitrogen	3.04E-06	1.51E-14	3.04E-06	1.51E-14
	CO_2	0.000861	5.65E-10	0.000860962	5.65E-10
	H_2S	0.001981	2.60E-08	0.001981474	2.60E-08
	CH_4	0.000882	6.49E-12	0.000882352	6.49E-12
	Ethane	2.24E-05	2.67E-13	2.24E-05	2.67E-13
	Propane	1.84E-05	8.17E-13	1.84E-05	8.17E-13
	i-Butane	6.66E-05	2.09E-11	6.66E-05	2.09E-11
	n-Butane	9.97E-06	1.93E-12	9.97E-06	1.93E-12
	i-Pentane	3.42E-05	4.54E-11	3.42E-05	4.54E-11
	n-pentane	8.16E-06	4.47E-12	8.16E-06	4.47E-12
	TEGlcol	0.089114	0.800005034	0.089114217	0.800005034
	H2O	0.906998	0.19999494	0.906998287	0.19999494

Table 2 shows the material balance composition of each components for TEG regenerator unit obtained from the result of Aspen Hysys simulation. The TEG Regeneration unit consist of one inlet stream and two outlet streams, the inlet stream is Regenerated feed and contains mostly water about 0.9 mol% and other components in trace amounts the results are in agreement with the principles of conservation of mass for a steady state process which states that total inflow of mass is equal to the total outflow of mass (Eric & Akademi, 2018).

3.2 Energy Balance Result

Table 3 shows the Energy balance summary of pump, Heat exchanger and Regenerator in terms of Duty (KJ/Hr) from hysys simulation, the pump duty has a value of 3837.6 KJ/hr but pump ratings are usually given in kilowatts hence this value is converted into 1.066 KW. From the results the Regenerator has the highest duty because it is attached to a reboiler. The pump is used to pump TEG into the contactor and from table 4.4, the amount of energy it needs to execute this is 3837.6 KJ/hr or 1.066 KW, Glycol flows from the pump discharge to the Lean Glycol Air Cooler (E-2647), where it is cooled to 50°C (122°F) by exchange with air. The cooled lean glycol then enters the Contactor (C-2635) above the structured packing. The first heat exchanger which is known as Glycol preflash exchanger is a multi-tube shell and tube type having its design temperature and pressure as 260^{o}C (500^{o}F) and 5.2 Barg (75.4 psig) respectively, its duty from the simulation result when converted to kilowatts gives 47.167 KW. The main function of the Glycol preflash exchanger is to raise the temperature of the glycol coming from the pump. The second heat exchanger which is known as Hot Lean/ Rich Glycol preflash exchanger is a multi-tube shell and tube type having its design temperature and pressure as 260^{o}C (500^{o}F) and 5.2 Barg (75.4 psig) respectively, its duty from the simulation result when converted to kilowatts gives 10 KW.

Table 3: Results of Energy Balance

Equipment	Heat Duty in KJ/Hr
Pump	3837.6
Heat Exchanger 1	1.698×10^5
Heat Exchanger 2	3.616×10^4
Regenerator	2.173×10^6

The main function of the Hot Lean/ Rich Glycol preflash exchanger is to drop the temperature of the glycol coming from the regenerator. The reflux rate is controlled by the amount of glycol allowed to pass through the Still Reflux Condenser. The glycol that flow through the condenser is limited by the level controller on the contactor, but can also be controlled with the manual bypass valve around the condenser. The reflux serves to condense any glycol in the vapors from the Still Column (E-2637) and promotes the fractionation of glycol and water in the packed section of the still. Water vapor exits the top of the still via a flame arrestor to the atmosphere while the condensed glycol is returned to the Glycol Regenerator (V-2642). The glycol stream leaving the still condenser at 56°C (133°F) is heated further by heat exchange with the lean glycol from the Rich Glycol Pre-flash Exchanger (E-2645). The stream, now heated to about 63°C (145°F), flows to the Glycol Flash Drum (V-2638), operating at a pressure of 3.45 Barg (50 psig). The regenerator has a duty of 2.173×10^6 KJ/hr and when converted to kilowatts gives 606.6KW (Finecountry *et al*., 2020).

3.3 Sizing Results

Table 5 shows the sizing result for the separator from hysys simulation, the diameter and height of the separator are given as 1.193m and 1.789m respectively. The separator was modeled as a Glycol flash drum. Glycol flows from the bottom of the Glycol Flash Drum (V-2638) on level control to the Glycol Cartridge Filters (S-2639 A/B). The filters are cartridge type filters, which remove solid particles from the glycol stream. From the cartridge filters, the glycol stream then flows to the Glycol Carbon Filter (S-2641). The filtering medium in this filter is activated carbon, which traps and absorbs various contaminants in the glycol such as dissolved hydrocarbons, oils, grease, and glycol degradation compounds. Removal of these materials is essential in order to prevent their accumulation, which causes fouling of equipment and foaming in the Reboiler (E-2643) and contactor (C-2635). The glycol stream then flows from the carbon filter through the tube side of the Lean/Rich Glycol Exchanger (E-2640), where it is heated by lean glycol from Glycol Surge Drum (V-2644) to about 163°C (325°F).

Table 5: Sizing for Separator

Parameter	**Value (m)**
Diameter	1.193
Height	1.789

Table 5 shows the sizing result of the absorber from hyssy simulation. The diameter and height of the absorber are 1.5m and 1.2m respectively.The gas ascends through the structured packing section where it is contacted by the descending glycol, which absorbs water vapor from the gas. Operating conditions of the Glycol Contactor (C-2635) are approximately 69 Barg at 41°C (1000 psig and 106°F). The Contactor (C-2635) is 1066 mm ID x 9144 S/S (42” ID x 30’-0” S/S) and is equipped with 4572 mm (180”) of structured packing. The glycol circulation rate is 2.61 m^3/hr (11.2 gpm). The dehydrated gas flows through a de-mister pad above the top liquid distributor and exits from the top of the tower. The gas leaving the tower is metered and continues on to the export facilities. Gas pressure in the Contactor and discharge piping is maintained by pressure controller (26PIC-351) and pressure valve (26PV-351). Prior to entering the export piping, some of the gas is diverted to the Fuel Gas system. Lean glycol from the regeneration module is supplied at the top of the structured packing through a liquid distributor. The liquid distributor ensures that glycol is distributed throughout the cross section of the Contactor (C-2635) column. Rich glycol is removed from the Contactor (C-2635) on level control from the chimney tray, which is located at the bottom of the structured packing section (Ibrahim *et al*., 2017).

Table 4.6: Absorber Sizing (Tray Section)

Parameter	**value**
Tray Diameter	1.5m
Weir Height	5×10^{-2}m
Weir Length	1.2m
Tray Spacing	0.5m
Tray Volume	0.8836m^3
Tray Hold up	8.836×10^{-2} m^3

Table 7 shows the sizing result from hysys simulation for a shell and tube heat exchanger.Heat exchanger 1 which is known as Glycol preflash exchanger is a multi-tube shell and tube type having its design temperature and pressure as 260°C (500°F) and 5.2 Barg (75.4 psig) respectively, its duty from the simulation result when converted to kilowatts gives 47.167 KW. The main function of the Glycol preflash exchanger is to raise the temperature of the glycol coming from the pump. Tube sizing are as follows: length of tube = 1m, tube outer diameter = 20mm, tube thickness = 2mm, tube pitch = 50mm, number of tubes per shell = 80. Shell side sizing are as follows: spacing = 800mm, diameter = 530.0259mm, and area = 5.03m^2 (Jokar *et al*., 2014).

3.4 Sensitivity Analysis

3.4.1 Temperature Variation with Number of Trays

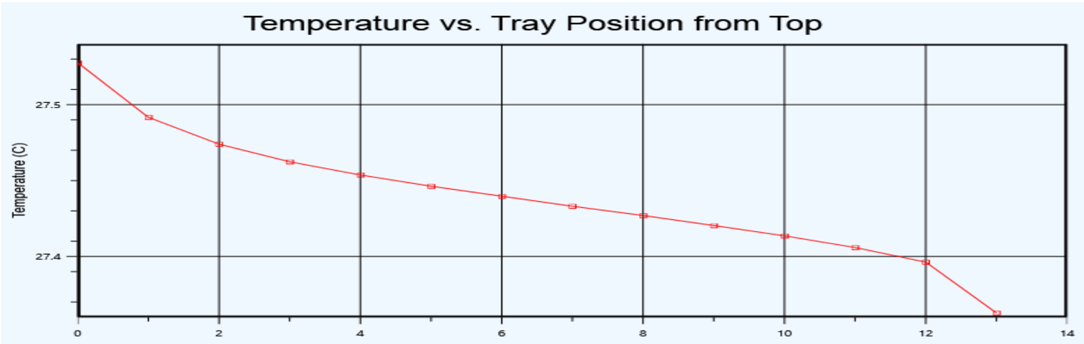

Figure 1: Temperature Variation with Number of Trays

Figure 1 shows the variation of pressure along the trays of the contactor, from the graph pressure is constant along the trays of the contactor, this is because the principles of operation of an absorber is relatively the same or close values of operating pressure for the inlet gas and the absorbent (TEG), and most times this value is operated at atmospheric conditions.

3.4.2 Effect of TEG Circulation Rate on Water Content of Natural Gas

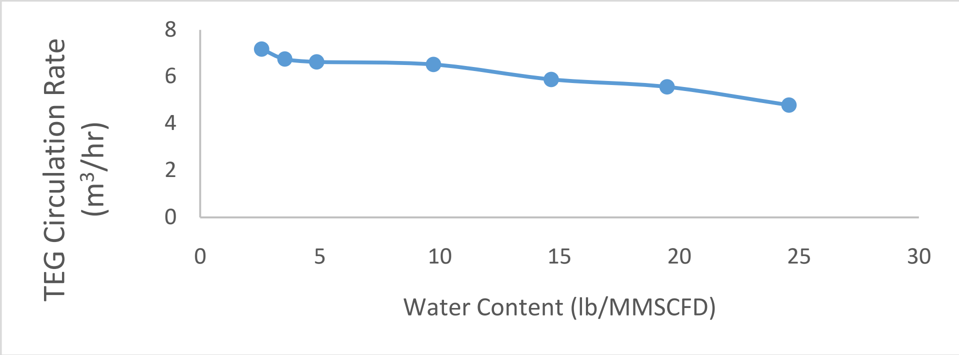

Figure 2: Variation of TEG Circulation rate on Water content

Figure 2 Shows how TEG flow rate affects the amount of water removed from natural gas known as water content usually below 7lb/MMSCFD before it can be accepted for pipeline transportation and to avoid corrosion of pipeline. From Figure 4.8 the more the TEG circulation rate the more the amount of water removed from the natural gas but using a TEG Circulation rate of 24.57m^3/h can achieve excellent drying results, however adoption of high TEG circulation rate may lead to increase in cost of operation and TEG losses due to accumulation of TEG in the contactor. Also using low TEG circulation rate could lead to hydrate formation as the water content in the Gas will be above the specified limit.

3.4.3 Effect of number of Equilibrium stages of the contactor on water content

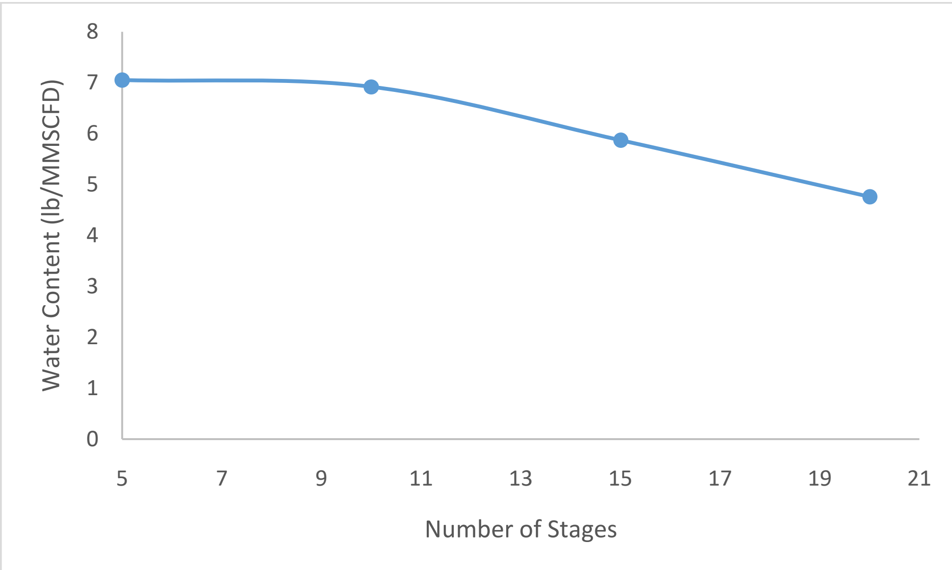

Figure 3: Variation of Number of Stages of Contactor on Water Content in Natural Gas

Figure 3 shows the variation of number of stages of contactor on water content in natural gas, increasing number of theoretical equilibrium stages the contactor results in reduced water content of the outlet gas regardless of the TEG flowrate, thus as expected, a greater number of contactor stages indicate a large surface area for effective transfer of water from the gas to the glycol. Also a greater number of trays indicates that the gas would have a higher contacting time in the vessel and would as such lead to more effective dehydration.

3.5 Process Optimization

The optimum conditions of the model were simulated to determine the points which give the least amount of TEG loss.

3.5.1 Determination of the Optimum Temperature that will give the Least TEG Loss

Regression analysis was used to fit a curve showing the relationship between TEG loss and temperature from the graph in Figure 4, the equation of the curve was determined using Microsoft excel as

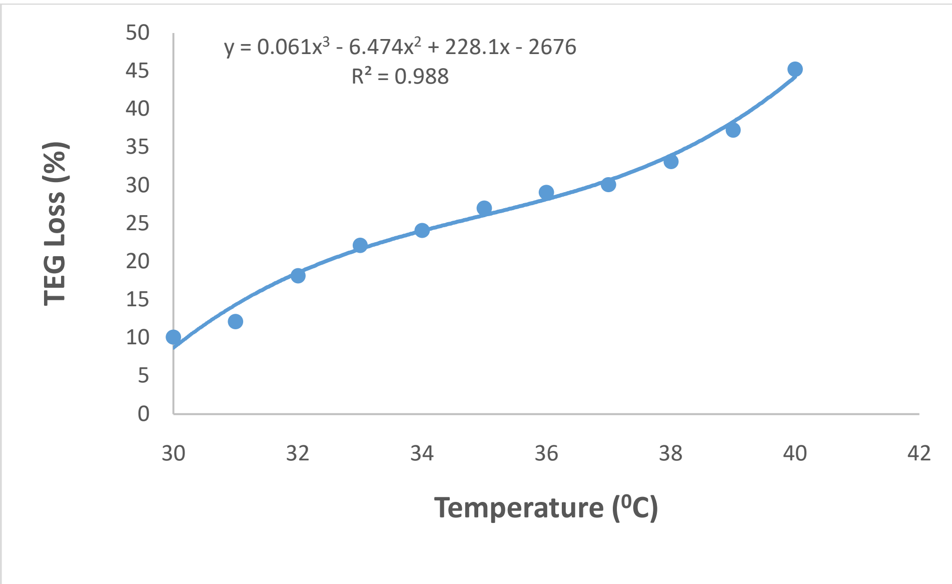

Figure 4: Regression Equation of TEG Loss in Temperature

$$y = 0.0618x^3 + 6.4744x^2 - 228.1x + 2676 \quad (22)$$

Where y = TEG loss in kg/hr

X = Temperature in ^{0}C

Rewriting the equation as

$$L = 0.0618\,T^3 + 6.4744T^2 - 228.1T + 2676 \quad (23)$$

Equation 4.1 was differentiated and equated to zero as shown in (4.2)

$$0.1854T^2+12.9488T-228.1 = 0 \quad (24)$$

Solving (24) as a quadratic equation gives two values T_1 = -84.42 and T_2 = 14.57, but then know that temperature cannot be negative so the only true value of T becomes T_2. To test the nature of this value was substituted in equation (25) after it was differentiated a second time in equation (24) to give to see if it gives a positive value then, it can be concluded that this value gives the minimum operating temperature.

$$0.3708T+12.9488 = 0 \qquad (25)$$

where $T = 14.57$, (4.3) gives $18.35 > 0$

Hence to minimize excessive loss of TEG at Imo River AGG plant, temperature should be kept at a value of 30^{o}C or lower than the operating temperature of 40^{0}C.

3.5.2 Determination of Optimum Pressure that will give the Least TEG Loss

Regress analysis was used to fit a curve showing the relationship between TEG loss and pressure from the graph in figure 5, the equation of the curve was determined to be

$$y = 9E - 8\,x^3 - 0.0017x^2 + 10.611x + 22408 \qquad (26)$$

Where y = TEG loss in kg/hr

X = Pressure in kpa

Rewriting the equation as

$$L = 9E - 8\,P^3 - 0.0017P^2 + 10.611P + 22408 \qquad (27)$$

Equation 4.5 was differentiated and equated to zero as shown in (4.6)

$$0.00000027P^2 - 0.034P + 10611 = 0 \qquad (28)$$

Solving (28) as a quadratic equation gives two values P_1 = 3202kpa and P_2 = 122,734kpa , but to find the minimum value of P we differentiate (27) a second time to give (29) and substitute the value of P_1 to see if it gives a positive value then we can conclude that this value gives the minimum operating Pressure.

$$0.00000054P - 0.034 = 0 \qquad (29)$$

Where P = 3202kpa, (4.7) gives $-0.0322 < 0$

The maximum pressure at which the plant should operate is 3202kpa but the operating pressure of the plant is 6800kpa which should be reduced to minimize excessive loss of TEG.

Substituting the value of the optimum pressure (3202Kpa) into equation (29) to find the corresponding TEG LOSS gave a loss of 4.19%.

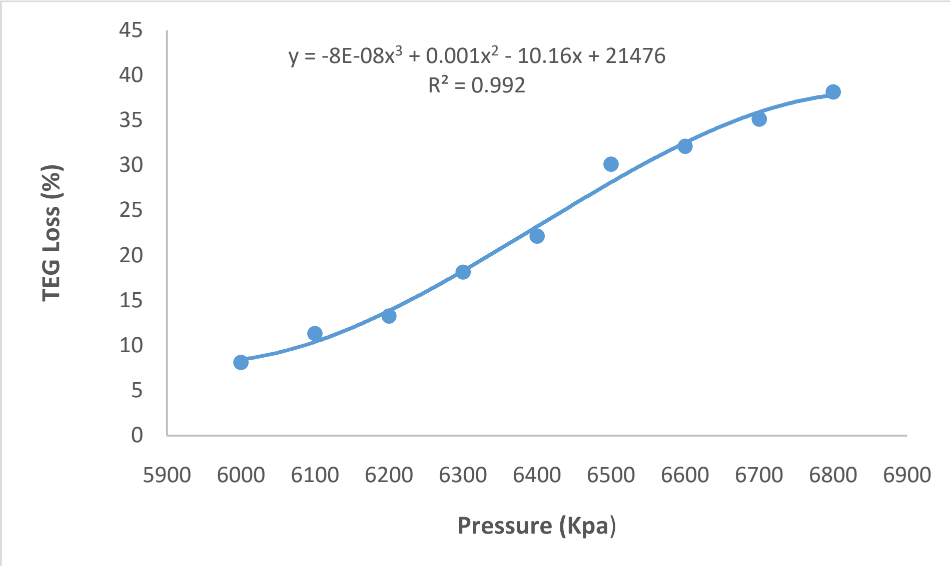

Figure 5: Regression Equation of TEG Loss in Pressure

4. CONCLUSION

This research studied the cause of excessive TEG loss in Imo River AGG Plant. The simulation was done using operating data from the plant. The material and energy balance of the plant was done using the principles of conservation of mass and energy. The plant sizing of the major equipment such as absorber (contactor), regenerator (distillation column) and heat exchanger were calculated. The diameter and height of the contactor was computed as 0.6096m and 6.034m respectively while the diameter and height of the regenerator was computed as 0.193m and 1.789m respectively and compared with base case specifications.
The major causes of TEG loss identified in this plant are:

(i) Operating at a high pressure of 78 bar (7800kpa) which should be reduced to 3202kpa to curb the effect of excessive TEG loss.
(ii) Operating at a temperature of 42°C which should be kept at an optimum value of 35°C to reduce TEG losses.
(iii) Using high TEG circulation rate of 24.57 m^3/hr which leads to accumulation of TEG in the contactor thereby causing loss of TEG, this should be reduced to about 15 m^3/h.

Sensitivity analysis was also carried out at it was absorbed that temperature was decreasing as pressure was constant throughout all the no of trays density and molecular weight decreased also along the trays and finally mole fraction of the food reduced along the trays.
Finally, operating pressure and temperature ranges that are close to the optimum temperature and pressure that gives the lowest FEG loss was recommended to be used as new operating conditions of the plant to minimize TEG losses. The new operating temperature range should be (30-32)°C and new operating pressure range should be (3200-6000) Kpa.

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