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Simulation and Optimization of Natural Gas Sweetening Plant

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ABSTRACT

This study is pursuing to quantitatively simulate an existing natural gas sweetening plant with a commercial process simulator (Aspen Hysys version10), using Peng Robinson as the fluid package. In this dissertation, one of the typical processes present in the natural gas chain were modelled: via chemical absorption with methyldiethanol amine, which has absorption selectivity of H_2S preferentially to CO_2 but for purpose of this project, the process parameter was modified to increase CO_2 loading because the natural gas contain negligible amount of H_2S. As the plant is considered new, there is a requirement to review the current plant operation and investigate any possible optimization and modification that can lead to reduction in the flared waste gases without violating the sweet gas specifications. Reduction of the operating costs and increasing the environmental sustainability of the process are two critical concerns in Greenville oil and gas. The plant operation was optimized by carrying out sensitivity analysis of the operating parameters (amine circulation rate and concentration simultaneously) which showed a decrease in CO_2 loading in lean amine. Also, graph of mole fraction of CO_2 in the sweet gas was plotted against amine circulation rate and amine concentration which showed a reduction in CO_2 concentration in the sweet gas thereby meeting the liquefied natural Gas (LNG) specifications. It was found that increasing the absorption column temperature will increase the losses of water and amine to the sweet gas. There is a water loss of approximately 0.5 kmol h^{-1} for every 5°C increase in temperature. So, it is recommended that sensitive temperature difference between feed gas and the amine should be establish to eradicate excessive rich amine loss to the sweet gas at the top of the column. Also, it is recommended that, to avoid any condensation of light

hydrocarbons which can cause foaming. So, a flow rate difference of at least $7m^3/h$ should be establish between the feed gas entering the contactor and the amine in a counter current manner. The result of this work shows that at circulation rate of $22.7m^3/hrs$ and concentration of 45% yielded the optimum natural gas composition (CH_4 = 91.4, C_2H_6 = 3.3459, C_3H_8 = 2.7336, i-C_4H_{10} = 0.5784, n- C_4H_{10} = 0.8530etc) required for LNG production.

Key words: Simulation, Optimization, liquefied Natural Gas, Sweetening Plant, Aspen Hysys

1. INTRODUCTION

Natural gas sweetening, also known as amine gas treating, amine scrubbing, and acid gas removal, refers to a group of processes that used aqueous solutions of various alkylamines (commonly referred to simply as amine) to remove hydrogen sulphide (H_2S) and carbon dioxide (CO_2) from gases. It is a common unit used in refineries and is also used in petrochemical plants, natural gas processing plant and other industries. From economic point of view, dissolving CO_2 or H_2S in water contribute the production of acidic solution (Kent & Eisenberg, 1976) and consequently corrosion in pipeline and related equipment will occur during transportation of natural gas. Moreover, Natural gas that contains more than 4 ppmv of hydrogen sulphide (H_2S) is commonly referred to as "sour". This is because the odour of hydrogen sulphide gas in air at very low concentrations is similar to that of rotten eggs. Significant quantities of natural gas resources around the world are known to contain H_2S.

Natural gas is a prime source of energy, which can be used as an industrial and domestic fuel. To make natural gas suitable and environmentally safe to use, it is crucial to remove all contaminants that can affect its utilization and optimal energy capacity. These contaminants can also cause problems such as corrosion, freezing, plugging, erosion, health and environmental hazards if not removed. Ghanbarabadi & Khoshandam, (2015), studied simulation and comparison of sulfinol solvents performance with amine solvents in removing sulphur compounds and acid gases from natural sour gas. The acid gases of hydrogen sulphide (H_2S), carbon dioxide (CO_2) and water are impurities existing in natural gas brought up from wellhead (oil wells with associated gas or gas wells). The gas is considered sour if its H_2S content exceeds (4ppm volume-based). The process for removing acid gases from a natural gas stream is referred to as gas sweetening (Abkhiz*et al.*, 2014). With the increasingly strict environmental regulations on emissions from natural gas processing plants and also the market demand for high quality

natural gas, the gas sweetening process has become mandatory (Rezakazemi*et al*, 2011). The most widely used method for acid gases removal from natural gas, is amine gas sweetening process with more than 50% of the current acid gas removal technologies use aqueous solutions of alkanolamines. However, this gas sweetening process is energy intensive especially for amine regeneration (Wang *et al*., 2015). Therefore, optimizing the amine gas sweetening process could result in great energy savings and remarkable economic benefits for the existing gas sweetening plants. Depending on their molecular structure, the amines are characterized into three main groups: primary, secondary, and tertiary. Mixtures of amines are also used in industry. The waste acid gases from the process are normally incinerated or flared to the environment. In the incinerator, H_2S is converted to SO_2. The flaring of sour gases not only means that natural gas resources are wasted, but also results in the release of pollutants into the atmosphere. CO_2 is a major contributor to global warming and SO_2 causes acid rain; accelerated corrosion of buildings and reduced visibility. Recently, most of the publications on the pre- or post-treatment of natural gas is concentrating on CO_2 capture either using physical or chemical solvents (Budzianowski & Koziol (2015), Budzianowski (2011a), Budzianowski (2011b), Budzianowski (2015)). However, the major concern of industry is always safety requirement for the removal of H_2S, which is highly toxic, is still constraining the design of natural gas production and processing facilities (Duissenov, 2013). Therefore, optimizing the performance of existing processes or investigating possible process modifications should always take into consideration keeping H_2S concentration in treated natural gas under the allowable discharge concentration.

2. Materials And Methods

3.2.2 Design of Components (Equipment) of Treatment Process

Conservation of Mass

Rate of mass inflow = Rate of mass out flow+Rate of Accumulation+Rate of generation (1)

If there is no chemical reaction and no accumulation of mass within the system, so the steady state balance reduces to;

Rate of mass inflow = Rate of mass out flow (2)

3. Material balance of packed column

From the general material balance as shown in equation (1), we can deduce the material balance of the scrubber column as thus:

Flow rate in = flow rate out at steady-state, accumulate rate = 0 (1)

$GmY_1 + LmX_1 = GmY_2 + LmX_2$ (2)

But X_1 =0, assumed pure solvent

$GmY_1 = GmY_2 + Lm\ X_2$ (3)

$Gm\ (Y_1 – Y_2) = Lm\ X_2$ (4)

$$X_2 = \frac{G_m}{L_m}(Y_1 – Y_2) \quad (5)$$

Definition of terms:

Let:

Y = mole fraction of gas absorbed at the bottom column (exist)

X = mole fraction of the solvent at the exist

G = mass fraction of the gas at exist.

L = mass fraction of the solvent at the exist

$m = \frac{G_m}{L_m}$ can be obtained and rated to get, N_{0G}

The relationship between $\frac{mG_m}{L_m}$, and N_{OG}

Were:

N_{OG} = overall number of transfer unit

4. Column Diameter Design

The design models for obtaining the size of the column were adopted from (Sinnott&, Towler,2009).

$$G_m = \frac{Gi}{3600_s}(\text{Kmol/s}) \quad (6)$$

Were:

Gi = mass flow rate gotten from material balance.

$$L_m = \frac{L_i}{3600_s}(\text{Kmol/s}) \quad (7)$$

Were:

$$Li = 0.98\ \frac{Y_1 G_m}{X_2} \quad (8)$$

Select material type of packing i.e. 25mm Rashing rings ceramics

Packing factor Fp = 525 m^{-1}

$$\rho_g = \frac{M_{mix}}{22.4} \text{ x } \frac{T}{T_f} = \frac{M_{mix}}{22.4} \text{ x } \frac{273}{(30+273)} \quad (9)$$

ρ_g = density of the solvent known or gotten from data (literature)

$$F_{LV} = \frac{L_m}{G_m}\sqrt{\frac{\rho_g}{\rho_L}} \quad (10)$$

Taking 20mm H_2O per packing of pressure drop

K_4 is obtained from F_{LV} as above.

$$G_i^* = \left\{\frac{K_4\rho_g\,(\rho_L-\rho_g)}{13.1\,Fp\,(\mu L/\rho_g)^{0.1}}\right\}^{½} \text{ [kg/s.m}^2\text{]} \quad (11)$$

Thus, ac = column area is calculated as thus;

$$a_c = \frac{G_m}{Gi^*} \text{ [m}^2\text{]} \quad (12)$$

Hence, column diameter is given

$$D_c = \left(\frac{4a_c}{\pi}\right)^{½} \text{ [m]} \quad (13)$$

Estimation of H_{OG}, hence Height of column

(Sinnott&, Towler, 2009).Stated design models for the estimation of overall height of gas transfer unit which are given as:

Given: D_L (Diffusivity of liquid), [m^2/s]

D_v = Diffusivity of vapour, [m^2/s]

μv = Viscosity (vapour), [Ns/m^2]

$$\frac{a_w}{a} = 1 - \exp\left[-1.45\left(\frac{\sigma_c}{\sigma_L}\right)^{0.01}\left(\frac{L_i^*}{au_c}\right)^{0.01}\left(\frac{L_i^* a}{\rho_g^*}\right)^{-0.05}\left(\frac{L_i^*}{\rho_L\sigma_L a}\right)^{0.2}\right] \quad (14)$$

Where:

a = 194 m^2/m^3 (actual area of packing) per unit volume, see Chemical Engineering design Sinnott, 2009, Table 11.2, vol.6.

a_W= Effective interfacial area of packing per unit volume, m^2/m^3

σ_c = critical surface tension (61 x 10^{-3} N/m)

σ_L = surface tension of liquid, N/m

g = acceleration due to gravity, 9.81 m/s^2

5. Determination of mass transfer coefficients for both liquid and gas K_L and K_G.

$$K_L\left(\frac{\rho_C}{\mu_{lg}}\right)^{1/3} = 0.0051\left(\frac{L_i^*}{a_w u_L}\right)^{2/3}\left(\frac{\mu_L}{\rho_C D_C}\right)^{1/2}(adp)^{0.4} \quad (15)$$

$$K_G\left(\frac{RT}{aD_v}\right) = K_5\left(\frac{G_i^*}{aD_v}\right)^{0.7}\left(\frac{\mu_v}{\rho_g D_v}\right)^{1/3}\left(ad_p\right)^{-2.0} \quad (16)$$

Were:

K_5 = 5.23, for packing size above 15mm and for sizes below 15mm.

R = 0.08314 bar m^3/kmol.K

$$L_i^* = \frac{L_m}{G_c} \text{Kmol/m}^2\text{s} \quad (17)$$

$$G_i^* = \frac{G_m}{a_c}, \text{Kmol/m}^2\text{s} \quad (18)$$

P = given total pressure, atmospheric (101.325 kPa).

Determination of liquid and gas height of transfer unit, H_L and H_G

$$H_L = \frac{L_m}{K_L a_w C_t}, \text{ m} \quad (19)$$

Where:

$$C_t = \frac{\rho_L}{M_s} \quad (20)$$

And, Ms = molar mass of solvent.

$$H_G = \frac{G_m}{K_G a_w P} \quad (21)$$

$$H_{OG} = H_G + \frac{m G_m}{L_m} H_L \quad (22)$$

$$Z = N_{OG} H_{OG} \quad (23)$$

Where:

Z_T = Total height of the packed column

R= Ideal gas constant, bar m^3/kmol.K

m = Slope of the equilibrium line

C_t = Total concentration, kmol/m^3

dp = packing size, mm.

Process Simulation

The process simulation model developed using a commercial process simulator follows an "ideal-stage" approach, coupled with kinetic modelling for carbon dioxide. In an ideal-stage model, each stage is assumed to reach thermodynamic equilibrium, and a real column is modelled by determining the number of ideal stages that yields the same performance as the real column. The kinetics model accounts for different absorption rates of the relatively slow absorption of CO_2, which is kinetically limited. Simulation flow diagram are validated against plant operating data, as shown in Figure 1. The Greenville LNG amine gas sweetening plant was simulated by using Aspen Hysys V.10. The MDEA is utilized as an aqueous absorbent to absorb acid gases from sour gas stream. The first step of simulation

work was selection of suitable fluid package and provide the feed gas stream and amine compositions.

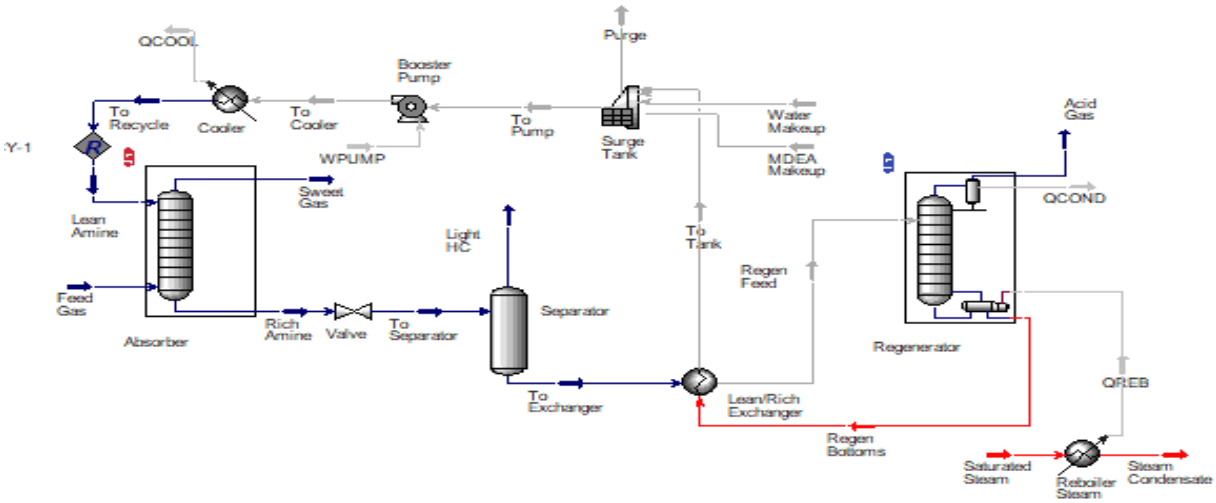

Figure 1: Aspen Hysys Process Simulation Model of the Plant

3. Results and Discussion

Table 1: Simulation Results

	CASES				
State	Amine Concentration %	Amine Flow Rate (m^3/h)	Regenerator Spec Value (KJ/h)	Sweet Gas CO_2 Composition (mol %)	Regenerator or Reboiler Bottom Acid Gas Loading
Case 1	40	18.17	5	0.071000	0.004762
Case 2	40	18.17	6	0.066000	0.003684
Case 3	40	18.17	7	0.064000	0.003143

Case 4	40	22.71	5	0.009694	0.006793
Case 5	40	22.71	6	0.009649	0.004863
Case 6	40	22.71	7	0.009626	0.00395
Case 7	40	27.26	5	0.009774	0.009925
Case 8	40	27.26	6	0.008718	0.006448
Case 9	40	27.26	7	0.008691	0.004948
Case 10	42.5	18.17	5	0.008986	0.004472
Case 11	42.5	18.17	6	0.007937	0.003388
Case 12	42.5	18.17	7	0.007912	0.002853
Case 13	42.5	22.71	5	0.006706	0.006447
Case 14	42.5	22.71	6	0.005666	0.004506
Case 15	42.5	22.71	7	0.005645	0.003597
Case 16	42.5	27.26	5	0.004645	0.003597
Case 17	42.5	27.26	6	0.004469	0.006041
Case 18	42.5	27.26	7	0.003747	0.004539
Case 19	45	18.17	5	0.003574	0.004233
Case	45	18.17	6	0.004521	0.003133

20					
Case 21	45	18.17	7	0.004494	0.002597
Case 22	45	22.71	5	0.002875	0.006152
Case 23	45	22.71	6	0.002840	0.0042
Case 24	45	22.71	7	0.001232	0.003288
Case 25	45	27.26	5	0.002925	0.009191
Case 26	45	27.26	6	0.003894	0.005688
Case 27	45	27.26	7	0.003878	0.004184

From Table 1 it was observed that keeping the temperature and pressure constant and increasing concentration and circulation rate of amine the amount of carbon dioxide (CO_2) removed increased. Also, it can be seen that at amine concentration of 40% and circulation rate of22.7m^3/hr the CO_2 content is 0.009626. When the amine concentration and circulation rate was increased to (42.5% and 27.3m^3/hr), there was a decreased in the CO_2content0.007645 in the sweet gas and when the amine concentration and circulation rate was increased to 45% and 22.7m^3/hr, the CO_2 content in the sweet gas was reduced to 0.001232. This showed that increasing amine concentration to 45% and circulation rate to 22.7m^3/hr will reduce the CO_2 content to 0.001232 as the optimal value

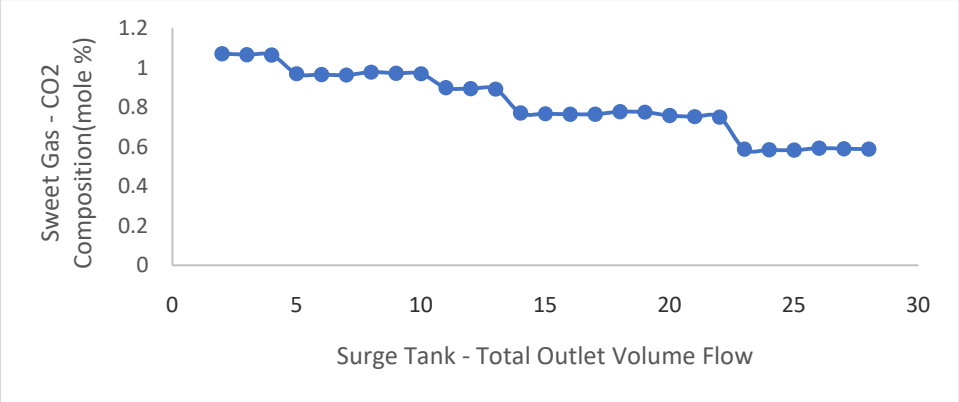

Figure 2: Graph of CO_2 Concentration in the Sweet Gas against Amine Circulation Rate

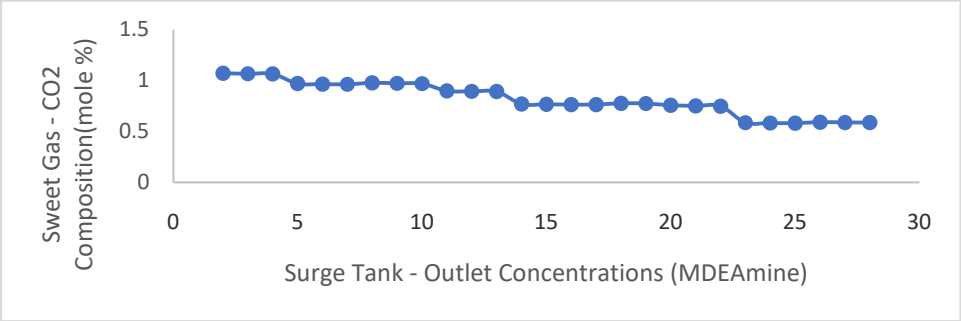

Figure 3: Graph of CO_2 Concentration in the Sweet Gas against Amine Concentration

A sensitivity analysis was carried out using Hysys version 10 (Figure 2 and Figure 3) to investigate the effect of the amine circulation rate and amine concentration on CO_2 in the sweet gas as well as the regeneration duty requirements, keeping the number of trays in the columns and other parameters fixed.

3.1 **Plant Data Validation**

Plant data validation before optimizing the actual gas sweetening process, the simulation results for the sweet gas stream specifications and required duties using the two Hysys packages are compared to the actual plant data from Greenville LNG. Table 4.1. Shows the plant data validation.

Table 4.5 Plant Data Validation

Components	**Feed Gas**	**Amine Plant Data**	**Sweet Gas Simulation Results**	**Amine Simulation Results**
CO_2 Content (%)	0.0286	0.000273	0.0012	0
Circulation Rate (m^3/h)	16	19.7	15.5	22.7
Concentration (%)	-	41	-	45

4. Conclusion

The plant model was set up in Hysys version 10, using current operating conditions in Greenville LNG RUMUJI, to set a benchmark, or base case. The plant data in Table 3.1 was used as simulation input data, then the flow sheet for conventional gas treatment system was developed, as shown in Figure3.1

It was found that the carbon dioxide in the rich amine increased with increasing amine circulation rates22.7m^3/h and concentration 45%. Regeneration reboiler duty (2969257kJ/h) increased due to increased flow rate (22.7m^3/h) entering the regeneration column (Table 4.3) which enhanced the stripping off CO_2 (0.003597) from the rich amine.

Also, CH_4 composition in the feed gas 87.1357% before simulation was increased to 91.4% in sweet gas. These resulted in enhanced acid gases removal efficiency in the absorption column thereby converting the raw gas to sweet gas that contains 91.4% CH_4which is within the LNG specification.

From the Data Validation in Table 4.1, this work has been able to discover how amine circulation rate and concentration can be tuned to increase high content CO_2 absorption by MDEA from the natural gas that meets CO_2 specification at the gas liquefaction unit.

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