ANALYSIS AND OPTIMIZATION OF THE EFFECT OF FEED GAS TEMPERATURE ON ACID GAS REMOVAL UNIT PERFORMANCE

Iwan Febrianto[1], Nelson Saksono[2]

[1,2]Chemical Engineering / Faculty of Engineering, Universitas Indonesia, Depok

Author's Email: iwan.febrianto@ui.ac.id

Abstract. *The Gas Gathering Station (SPG) in field X processes gas from 16 (sixteen) wells before being sent as sales gas to consumers. The sixteen wells have decreased in well pressure since 2011, thus affecting the performance of the Acid Gas Removal Unit (AGRU). The AGRU facility in field X is designed to reduce the acid gas content of CO_2 by 21 mol% with a feed gas capacity of 85 MMSCFD. The condition of the existing AGRU is operating at 40% of the design capacity. A decrease in reservoir pressure causes an increase in the feed gas temperature and an increase in the water content of the well. Based on the reconstruction of the design conditions into the simulation model, the amine composition consists of MDEA 0.3618 and MEA 0.088 wt fraction. The increase in feed gas temperature to 146 F causes foaming due to condensation of hydrocarbon heavy fraction, so it is necessary to modify it with the addition of a chiller to cool the feed gas to 60 F so that, based on the simulation of the flow rate of gas entering AGRU, it can reach 83.7 MMSCFD. There was an increase in gas production of 38.1 MMSCFD and condensate of 1,376 BPD. Economically, the chiller addition modification project is feasible with the economic parameters of NPV of US$ 132,000,000, IRR of 348.19%, POT of 0.31 year, and PV ratio of 19.06.*

Keywords: *AGRU, reconstruction, foaming, chiller, economy*

1. INTRODUCTION

The Gas Gathering Station (SPG) facility in field X processes gas from 16 (sixteen) production wells with a feed gas capacity of 85 MMSCFD at a gas temperature of 105 F separator and a pressure of 800 psig with 21% mol CO_2 gas content. The main process facilities consist of a Production / Test Manifold, a Separation Unit, an Acid Gas Removal Unit (AGRU), and a Dehydration Unit (DHU). The fluid from the wells goes to the manifold and the liquid is separated at the HP Separator, which operates at a pressure of 775 psig and a temperature of 91 F, then goes to the HP Scrubber before entering AGRU. The Sweet Gas from AGRU will be sent to DHU to reach a moisture content of 7 lb / MMSCFD. The liquid from the HP Separator goes to the MP Separator, which operates at a pressure of 200 psig, and then goes to the LP Separator, which operates at a pressure of 75 psig. The resulting liquid is then sent to the Condensate Tank. The gas from MP Separator and LP Separator is used for fuel gas. (Process Description, 2005).

The Acid Gas Removal Unit (AGRU) uses BASF's aMDEA licensor solution to reduce the CO_2 content to 5% mol. The aMDEA solution is a property of the licensor that is not found in the UNISIM software, so it is necessary to simulate the reconstruction according to the design using a mixture of MDEA and MEA activator. The current condition, the reservoir will cause a decrease in reservoir/tubing pressure and an increase in water production, which will cause an increase in the temperature of the feed gas as shown in Figure 1. and changes in the fluid composition of the reservoir so that the existing AGRU only operates at 40% of the design capacity of the feed gas at 85 MMSCFD due to foaming. The condensation of the heavy fraction of the feed gas has a temperature hotter than the incoming amine solution, which can cause foaming which will result in amine losses and plant shutdown. The increase in absorber temperature will also reduce the effectiveness of CO_2 and H_2S absorption by the amine solution.

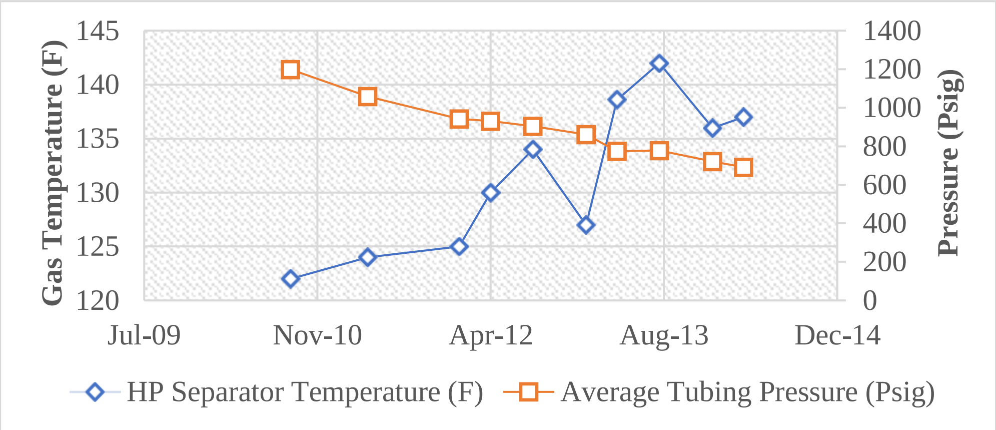

Figure 1. The Effect of Decreased Reservoir Pressure on Gas Temperature

In this paper, a reconstruction simulation model using UNISIM software according to the design to get a mixture of MDEA and MEA activator, then performs an AGRU simulation with changes in pressure and temperature feed gas, then calculates the project's economy to modify existing facilities so that AGRU can operate optimally.

2. LITERATURE REVIEW

The technology for removing CO_2 and sulfur content from acid gas in the survey consists of solvent absorption, solid absorption, direct conversion, and membrane. Solvent absorption technology uses generic amines, namely primary (MEA, DGA), secondary (DEA, DIPA), and tertiary (TEA, MDEA). The MDEA solution is used at a concentration of 30-50% by weight in the air. Unlimited acid gas loading is typically 0.7 - 0.8 moles of acid per mole of amines. So that it can reduce the amount of circulating energy (pump energy). MDEA is not easily degraded either by heat or chemical and heat reaction with low H_2S. Maddox et al. (1998) suggested that MDEA could be used as a non-selective solution to remove H_2S and CO_2, or it could be used as a selective solvent that removes H_2S and CO_2. Maddox et al. (1998) stated that the MDEA selectivity was raised by:

1. Temperature: The lower temperature will increase the selectivity.
2. Pressure: lower pressure will increase selectivity.
3. CO_2 / H_2S ratio: the higher it will support selectivity.

MDEA is a superior amine because the corrosion rate and low degradation rate result in the ability to use high solution concentrations (Polasek and Bullin, 2006). The theoretical loading of MDEA is 1 mole of acid gas/mol of amine (Kohl, 1999). This makes it more attractive for CO_2 removal if it can overcome MDEA's low reaction rate with CO_2. The process of CO_2 absorption by MDEA must be increased by adding an activator (Arkema Co., 2000). Amines such as methanolamine (MEA), diethanolamine (DEA), and piperazine (PZ) have been used as activators for MDEA blends to increase reaction rates. The Pz / MDEA mixture absorbs CO2 faster than monoethanolamine (MEA) or diethanolamine (DEA) mixed with MDEA at the same concentration (Bishnoi, 2002). The following is shown in Table 1. The reaction constants in the amine solution.

Table 1. Amine Reaction Constants on CO_2 Absorption
(Optimized Treating Inc., 2008)

Amine	Reaction Constants (L/mol.s)
MEA	6000
DGA	4500
DEA	1300
DIPA	100
Piperazine	59000
MMDEA	7100
MDEA	4

3. METHODOLOGY

The research was conducted using a process simulation following the algorithmic stages as shown in Figure 2.

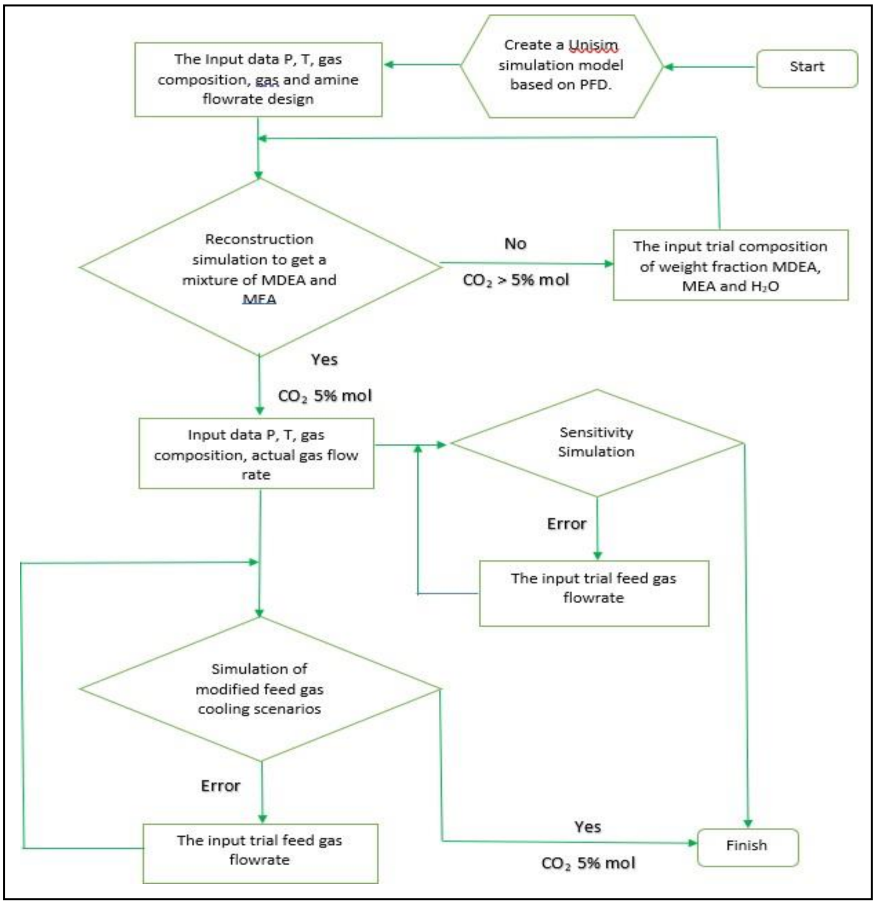

Figure 1. The Algorithmic Process Simulation

4. RESULT AND DISCUSSION

4.1. Reconstruction Simulation

Design reconstruction simulation modeling using feed gas data in AGRU with operating parameters based on the design data. Then, after obtaining the reconstruction model, it is continued with a simulation according to the current conditions. The differences in design and currentl data as input parameters in the simulation are shown in Table 2.

Table 2. Design & Current Data Input

Parameter	**Desain**	**Aktual**
Name	**FG**	**FG**
Vapour Fraction	1.0	1.0
Temperature [F]	83.2	140.6
Pressure [psig]	650.0	650.0
Molar Flow [MMSCFD]	85.0	86.0
Mass Flow [lb/hr]	221049.5	228105.0
Comp Mole Frac (CO_2)	0.2100	0.2105
Comp Mole Frac (H_2S)	0.00E+00	1.80E-05

Modeling on the UNISIM software follows the Process Flow Diagram (PFD) in the design conditions shown in Figure 3.

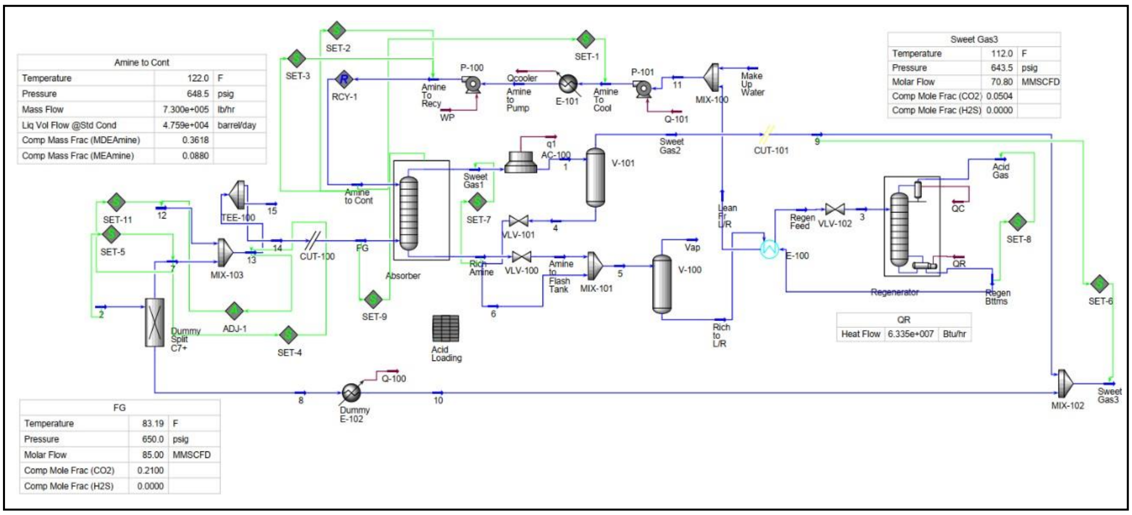

Figure 2. The Modelling PFD Design

Based on the process simulation reconstruction, the results of a mixture of amine, MDEA 0.3618 with MEA activator 0.088 by weight fraction to get the CO_2 content in the sales gas of 5% mol with a lean amine flow rate according to the design. From the reconstruction model, it continued with the simulation of the current conditions with an increase in temperature, an increase in flow rate, and a change in the composition of CO_2 and H_2S in the feed gas. It is found that the CO_2 content in the sales gas has increased to 6.43 mol% compared to the design conditions. Under these conditions, the sales gas does not meet the specifications of $CO_2 \leq 5\%$ mol.

4.2. Sensitivity Feed Gas Pressure

The sensitivity simulation was carried out at a feed gas flow rate of 70 MMSCFD because it was related to the convergence of the simulator in this case study. Changes in feed gas pressure resulting in changes in CO_2 and H_2S in gas sales are shown in Figure 4.

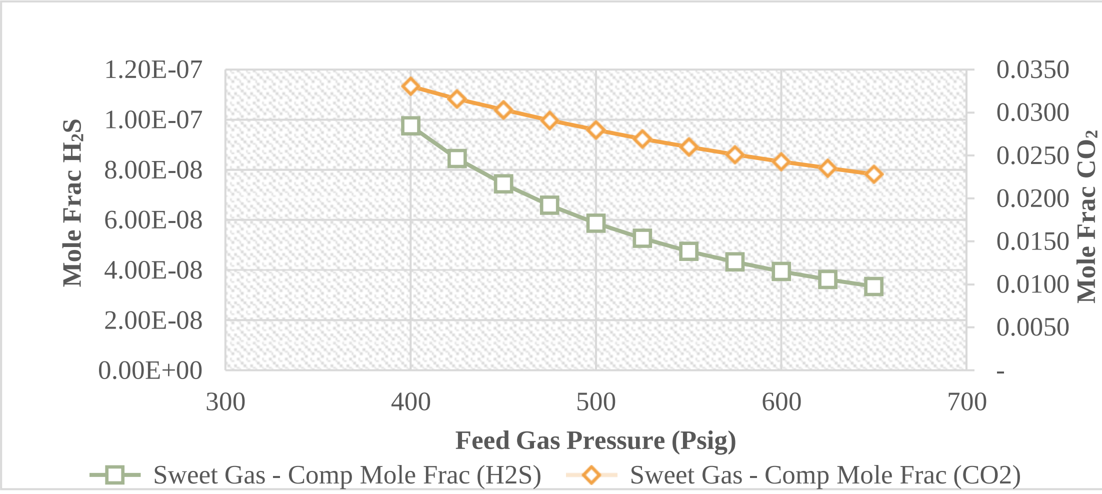

Figure 3. Sensitivity Feed Gas Pressure

The effect of a greater absorber pressure will increase absorption efficiency so that the acid content in the sales gas (CO_2 and H_2S) will be lower as shown in Figure 4. Reducing the absorber pressure will increase the acid content due to an increase in component volatility (J. Park, S. Yoon, S.-Y Oh et al., 2020). The reduction in pressure contributes to reducing the partial pressure of CO_2 in the feed gas and consequently decreases the rate of reaction with amines, which decreases the efficiency of CO_2 removal (A.Y. Ibrahim et al., 2014).

4.3. Sensitivity Feed Gas Temperature

Changes in feed gas temperature resulting in changes in CO_2 and H_2S in gas sales are shown in Figure 5.

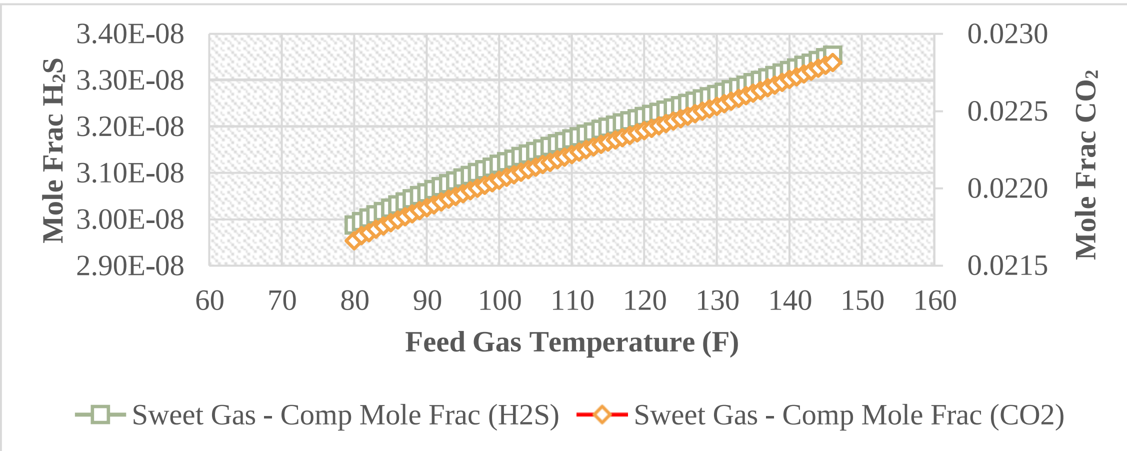

Figure 4. Sensitivity Feed Gas Temperature

Based on Figure 5, the higher the feed gas temperature will produce sales gas with higher acid content. Hydrocarbon condensation in the absorber can be avoided by keeping the lean amine temperature at least 5 °C above the hydrocarbon dew point (GPSA-1998). The feed gas temperature will affect the condensation of the heavy hydrocarbon fraction in the AGRU operation, so foaming will occur. A heavy fraction of hydrocarbons carried by the feed gas due to high temperatures will dissolve in the amine and circulate through the stripper, which will disrupt the lean amine pump so that a plant shutdown can occur. The lean amine temperature that enters the absorber is 122 F so that the maximum feed gas temperature is 113 F to avoid foaming.

4.4. Equipment Modification

With the increase in feed gas temperature, the current condition of the plant can only operate at a feed gas flow rate of 40 MMSCFD from a design capacity of 85 MMSCFD due to foaming at the AGRU unit. Foaming occurs due to the presence of a heavy fraction of hydrocarbons dissolved in the amine circulation. Based on the current simulation of 40 MMSCFD feed gas, the mole fraction of CO_2 in the sales gas is 0.0018 and 931 BPD d of liquid condensate with 13.7 psia of RVP. There are 4 (four) alternative modifications to reduce the temperature of the feed gas which are shown in Table 3.

Table 3. Current Condition and Modification Alternative

Parameter	Current	*Chiller*	*Air Cooler*	*Cooling Water*	*JT Valve*
Feed Gas *Outlet Separator* (MMSCFD)	40	84.7	79.8	80.1	83.7
Feed Gas Temp Outlet Separator (F)	146	146	146	146	146
Feed Gas *Inlet AGRU* (MMSCFD)	39.99	83.7	79.56	79.71	83.45
Feed Gas Temp *Outlet* Modif. (F)	140.6	60	113	95	103.7
Feed Gas Temp *Inlet AGRU* (F)	140.6	81	113	95	113
% Mol CO_2 Inlet AGRU	21.05	21.21	21.11	21.14	21.11
% Mol CO_2 Outlet AGRU	0.18	5	5.11	5.42	4.99
Sales Gas (MMSCFD)	31.35	69.43	65.95	66.37	69.09
Condensate (BPD)	931	2307	1856	1909	1946
RVP Condensate (Psia)	13.7	15	13.5	13.6	13.5

Based on Table 3. It can be seen in the simulation of current conditions and alternative modifications to reduce the temperature of the feed gas. The current condition can only operate with 40 MMSCFD of feed gas due to an increase in temperature from the well to 146 F, which causes foaming. The most optimal cooling alternative uses a chiller so that the feed gas temperature can be cooled down to 60 F and enters the absorber at the design temperature of 81 F, which results in the highest increase in gas and condensate production among other alternatives. The temperature delta between the amines is maintained above 5 C of feed gas (GPSA-1998) so that heavy fraction condensation does not occur, which causes foaming.

4.5. Modification Economics

The CAPEX calculation based on the previous year's contract price which is converted using the Chemical Engineering Capital Index (CEPCI) approach to 2021 is shown in Table 4.

Table 4. Main Equipment Cost

No	Equipment	Specification	Volume (Ea)	Unit cost (US$)	Cost (US$)
1	Gas/Gas Exchanger	2.25 MMBTU/Hr	1	308,464.56	308,464.56
2	Chiller Package	7.55E6 MMBTU/Hr	1	2,091,756.25	2,091,756.25
	Total				2,400,220.81

Based on the price of the main equipment, the total CAPEX calculation is carried out based on the Plant Cost Factor (Chemical Engineering Economics, 2013), which is shown in Table 5.

Table 5. Total CAPEX

No	Component	*Plant Cost Factor*	Cost US$
1	**Project Management, Engineering & Construction:**		
	Construction & Engineering	0.3	720,066
	Contractor Fee	0.1	240,022
	Contigency	0.15	360,033
2	**Preparation, Civil and Building:**		
	Land Preparation	0.05	120,011
	Environment	0.1	240,022
	Building	0.05	120,011
	Fondation	0.07	168,015
3	**Procurement:**		
	Main Equipement	1	2,400,221
	Piping	0.15	360,033
	Electrical	0.1	240,022
	Instrument	0.1	240,022
	Utilities	0.3	720,066
	Insulation	0.02	48,004
	Painting, fire proofing and safety	0.02	48,004
	Total Plant Cost		**6,024,554**
4	**Commisioning and Performance Test :**		
	Plant Start Up	0.05	301,228
	Working Capital	0.1	602,455
	Total CAPEX		**6,928,237**

Economic calculations are carried out using the KKKS (Cooperation Contract Contractor) Cost Recovery scheme model. The revenue calculation uses the sales gas price of 4 US \$/MMBTU, and the condensate price is 63.5 US \$/bbl. OPEX calculation of gas operating costs of 1.7 US \$/MSCF and oil 17.5 US \$/bbl. The economic results with a production age of 10 years are shown in Table 6. as follows:

Table 6. The Economic Result

Parameter	Unit	Result
IRR	%	348.19%
NPV@10%	US$ (000)	132,020
POT discounted	Year	0.31
PV Ratio		19.06

Based on Table 6, it shows that the modification of the addition of chiller equipment will result in additional gas and condensate production as gross revenue. By using the economic calculation of the KKKS (Cooperation Contract Contractor) cost recovery scheme model, the IRR value is 348.19%. With a factor discount rate of 10%, the NPV is US $132.020,000, the POT investment return rate is 0.31 years, and a PV ratio of 19.06. Thus, the economic indicators of NPV are positive, the IRR > of bank interest rates and the fast POT and PV ratio of more than 1. So it can be concluded that the modification of the addition chiller is feasible to increase company revenue.

5. CONCLUSION

The research using UNISIM software has succeeded in simulating aMDEA reconstruction modeling. The results of the design reconstruction simulation obtained the composition of the amine mixture (MDEA and MEA) with a weight fraction of MDEA 0.3618 and MEA 0.088 to obtain the percentage of CO_2 in the 5% mol of sales gas. The effect of the lower feed gas pressure is the higher the CO_2 and H_2S content in the sales gas product because the amine will be more volatile and the decrease in the partial pressure of the acid in the feed gas consequently decreases the rate of reaction with amines. While the effect of the feed gas temperature is higher, the higher the CO_2 and H_2S content in the sales gas product will also result in foaming due to condensation of heavy fractions. Reducing the feed gas temperature to 60 F using a chiller by considering the temperature of the hydrate formation so that AGRU can operate with a feed gas flow rate of 83.7 MMSCFD from the original 40 MMSCFD. There was an increase in sales gas production of 38.1 MMSCFD and condensate of 1,376 BPD. Modification of the addition of a chiller requires a CAPEX cost of US $ 6,928,237 and a feasible economic calculation result with NPV parameters of US $ 132,000,000, IRR of 348.19%, POT of 0.31 years, and PV Ratio of 19.06.

6. REFERENCES

Abdi, M.A. (2008) *Design and operations of natural gas sweetening facilities*. Course prepared for Iranian Gas Company. Faculty of Engineering and Applied Science, Memorial University of Newfoundland (MUN).

Aliabad, H. Zare, Mirzaei, S. (2009) *Removal of CO_2 and H_2S using aqueous alkanolamine solutions*. International Journal of Chemical and Biological Engineering 2:2.

Arkema Co. (2000) MDEA *proven technology for gas treating systems.*

Bishnoi, S., Rochelle, J.T. (2002) *Thermodynamics of piperazine / methyl diethanolamine/ water / carbon dioxide.* The University of Texas at Austin, USA.

Bullin, J.A., Polasek, J.C. (1990) *The use of mdea and mixtures of amines for bulk CO_2 removal*. Bryan Research & Engineering, Inc., Bryan, Texas.

Chung-Sung, T. (2006) *Absorption of carbon dioxide with piperazine and its mixtures in a rotating packed bed.* Separat. Purif. Technol. Taiwan.

Garrett, Donald E. (2013) *Chemical engineering economics*. New York: Van Nostrand Reinhold, 16th Edition.

Gas Processor Supplier Association. (1998) *Engineering data book*. Tulsa, Oklahoma, Eleventh Edition.

Ibrahim, A.Y, et al. (2014) *Effect of piperazine on carbon dioxide removal from natural gas using aqueous methyl diethanol amine*. Journal of Natural Gas Science and Engineering.

KKKS. (2005) *Process description dan desain basis stasiun pengumpul gas X*.

Kohl, A.L., Nielson, R.B. (1997) *Gas purification*. Houston: Gulf Publishing Company, Fifth ed.

Kolmetz, Karl. (2012) *Natural gas sweetening systems selection, sizing and troubleshooting*. Malaysia: KLM Technology Group.

Maddox, R. and Morgan, D.J. *Gas Conditioning and Sulfur Recovery*. Amine based Processes and Amine Process Calculation, 1998

Optimized Treating, Inc. (2008) *Piperazine, why it's used and how it works.* Vol. 2. Issue 4.

Pandey, M. (2005) *Process optimization in gas sweetening unit-a case study.* In: Proceedings of the International Petroleum Technology Conference, Doha, Qatar.

Polasek, John, Bullin, Jerry A. (2006) *Selecting amines for sweetening units.* Bryan, Texas: Research & Engineering, Inc.