Improved Performance of the Natural-Gas-Sweetening Benfield-HiPure Process Using Process Simulation

R. Ochieng*[a], A. S. Berrouk[a], J. Slagle[b], L.G. Lyddon[b], P. E. Krouskop[b]

[a]Department of Chemical Engineering, Petroleum Institute, Abu Dhabi-UAE

[b]Bryan Research & Engineering, Inc., Bryan, Texas-USA

Email: ochienrich@gmail.com

ABSTRACT

Natural Gas processing plants are an essential part of the energy industry, providing clean burning fuels and valuable chemical feedstock. The importance and complexity of gas processing plants have increased over the years, leading to improvements in energy efficiency and integration with petrochemical plants. These improvements are aided by the use of computer simulation models as tools for designing, troubleshooting, and optimizing gas treating plants.

This work discusses the major optimization techniques based on the Benfield HiPure process at Abu Dhabi Gas Liquefaction Company Limited (ADGAS) and the use of a process simulation tool, ProMax®. At ADGAS' Train 3 plant in Das Island, high pressure natural gas containing 6 to 7 mole % acid gas first comes into contact with a 30 weight % Potassium Carbonate (K_2CO_3) solution promoted with 3 weight % Diethanolamine (DEA). The gas is then contacted with a 20 weight % DEA solution downstream.

The results from the simulations show a close match with the plant operating data. The simulation model was then used to explore the effect of changes in process parameters on ADGAS' plant performance.

Keywords: Process Simulation, Benfield HiPure, Parametric Study, Gas Sweetening, LNG

Abu Dhabi Gas Liquefaction Ltd (ADGAS) Plant

ADGAS, a part of ADNOC (Abu Dhabi National Oil Company) group, is known for the production of Liquefied Natural Gas (LNG) since 1977. ADGAS operates three LNG Trains. The first two trains, (Trains 1 & 2) have been in operation since 1977, each with a capacity of 180 tons per hour of LNG. The third train (Train 3) was commissioned in 1994 and is capable of producing 380 tons per hour of LNG[1].

The Train 3 gas sweetening plant is a "Benfield HiPure" design supplied by UOP, and is a hybrid arrangement of the basic Benfield and Amine units[5].

Carbon dioxide and hydrogen sulfide removal from natural gas is a key step in the liquefied natural gas (LNG) process, in particular for sour gas streams containing significant concentrations of these acid gases. This plant is seeking sweet gas which contains no more than 5 ppmv and 50 ppmv of H_2S and CO_2, respectively[1]. Higher acid gas concentrations will directly affect the quality of LNG product and/or pose serious operational problems to the cryogenic columns. Failure to remove carbon dioxide can cause freeze-out on surfaces inside heat exchangers or plug lines which may lead to safety hazards and/or reduced operation efficiency. In presence of water, CO_2 and H_2S also form acids which cause corrosion of process equipment[2-4]. Therefore, removal of these contaminants is an operational necessity in any LNG producing plant.

The Benfield HiPure Process of ADGAS

The Benfield HiPure design was described in 1974 by Benson and Parrish[5]. It uses two independent but compatible circulating solutions to remove acid gases (H_2S & CO_2) from natural gas. In the first stage, the bulk of the acid gas is removed in a carbonate absorption system, where hot potassium carbonate promoted with diethanolamine (DEA) is employed as the solvent. In the second stage, the remaining acid gases are removed in an amine absorption system using DEA as the solvent. The DEA system provides the final trim removal of the acid gases to achieve the required sweet gas specification of less than 5 ppmv H_2S and 50 ppmv CO_2. The integrated schematic of the Benfield HiPure process is shown in **Figure 1**.

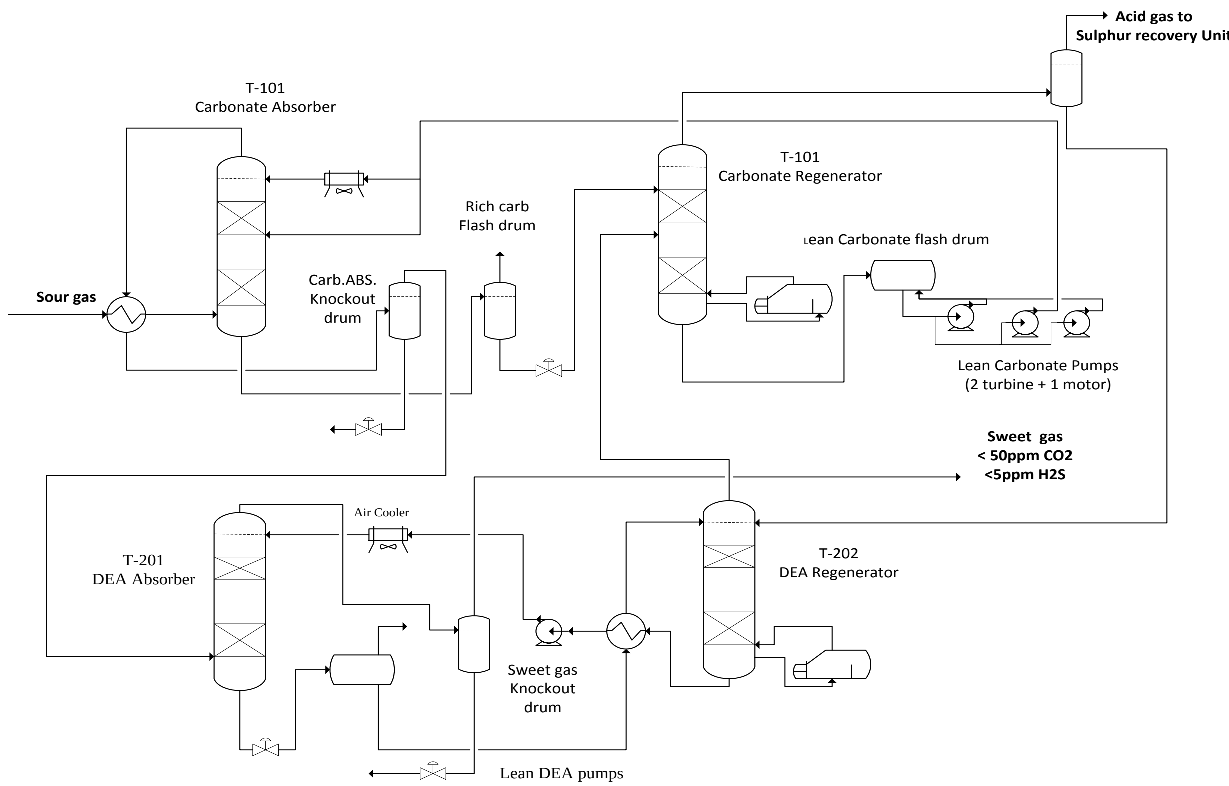

Figure 1: ADGAS' Gas Sweetening unit –Process Schematic

The hot potassium carbonate absorption system is comprised of a split flow absorber and a regenerator with no side draws. The carbonate absorber and regenerator are both tall vertical packed bed columns. The treated gas from the carbonate absorber is fed directly into the amine absorber.

The DEA amine system is comprised of an absorber and a stripper, both tall columns using a packed bed arrangement. After absorbing the acid gases, the rich solution from the absorber is pumped to the DEA regenerator. The regenerator has no condenser, and the overhead gas is fed to the middle of the carbonate regenerator, which does have a condenser. Liquid from the carbonate regenerator condenser is fed to the top of the DEA regenerator as reflux. The exit gas from the DEA absorber (sweet gas) is passed on for further processing to produce LNG. The stripped acid gases (H_2S and CO_2) from both the carbonate and DEA regenerators proceed to a sulfur recovery unit (SRU), where the acid gases are processed to produce molten liquid sulfur.

The feed gas to Train 3 is high pressure gas of about 52 bar (g) with an average H_2S and CO_2 content of about 2.2% and 4.7%, respectively. The sweetened gas produced by this plant is about 0.4 ppmv H_2S and 19 ppmv CO_2, which meets the required design specifications [1].

Both the hot potassium carbonate and DEA absorbers operate at a pressure of about 50 bar(g) and the regenerators operate at lower pressures of about 0.8 bar(g). The necessary heat load for the regeneration is supplied through reboilers associated with each of the regenerators. Operating data and column internals are shown in **Table 1** and **Table 2**.

Table 1: Typical Operating Data for ADGAS' Train 3 Plant

Parameter		Value
Feed Gas Flow Rate (MMSCFD)		476.93
Feed Gas Temperature (^oC)		25.03
Feed Gas Pressure (barg)		52.08
CO_2 Feed Gas Composition (%)		4.67
H_2S Feed Gas Composition (%)		2.11
Hot Potassium Carbonate Unit		
Circulation Rate (m^3/hr)	Main:	343.50
	Split:	1292.20
Lean Solvent Temperature (^oC)	Main:	81.84
	Split:	117
Lean Solvent Pressure (barg)		51.4
K_2CO_3 Concentration (wt %)		30
Promoter Concentration (DEA) (wt %)		3
Amine Unit		
Circulation Rate (m^3/hr)		109.8
Lean Solvent Temperature (^oC)		49.94
Lean Solvent Pressure (barg)		53.71
DEA Concentration (wt %)		20

Table 2: Absorber Configurations for ADGAS' Train 3 Plant

Hot Potassium Carbonate Absorber		
Top section		
	Column Diameter (m)	3.581
	Packing Height (m)	9.144
	Bed 1 Packing Type	#2.5 S.S Mini Rings
Bottom Section		
	Column Diameter (m)	4.724
	Packing Height (m)	9.144
	Bed 2 Packing Type	#3 S.S Mini Rings
Amine Absorber		
	Column Diameter (m)	2.972
	Packing Height (m)	15.24
	Packing Type	#3 S.S Mini Rings

Process Optimization through Process Simulation

Process optimization is the ultimate goal of process simulation.

Simulation models help illuminate the bottlenecks in the processes and identify changes to help optimize plant performance. Process simulation can be described as a logical model for a chemical process that can be used to evaluate the process response for a given set of inputs. In a typical engineering process, process simulation provides the capability for the designer to understand the consequences of new design before the actual implementation of the process. This greatly minimizes the risks associated with implementation of less than optimum designs. Simulations also enable prediction of process responses to proposed changes in process parameters for proposed improvement projects[6, 7].

Much work has been published on the modeling and simulation of both the Hot Potassium Carbonate [3, 8] and Amine systems [2, 3, 4, 10, 11], however, limited studies have been available for the Benfield HiPure process [5]. This work focuses on the simulation of the ADGAS Train 3 plant using ProMax® modeling software due to its capabilities in modeling gas sweetening and other electrolytic processes [9]. ProMax was used to perform a parametric study to provide guidance for new operating conditions that can improve performance of the gas sweetening unit.

ADGAS Train#3 Simulation Model

The plant model is set up in ProMax using current operating conditions to set a benchmark, or base case, for the case studies. Since the solvents are strong electrolyte solutions, the "Electrolytic ELR-PR" property package supplied with ProMax is used to predict the liquid phase thermodynamic properties. In the model, the potassium carbonate solution is specified as a quantified mixture of KOH, CO_2 and water. The KOH and CO_2, along with other components, are treated as ionic species in solution when using the Electrolytic Property Package. The TSWEET Kinetics model predicts the CO_2-amine or CO_2-carbonate kinetic reactions taking place in acid gas absorbers. TSWEET kinetics accounts for the relatively slow reaction of CO_2 with amine or carbonate solutions. This allows ProMax to model the rate-based reactions taking place in the absorber.

The two units of the Benfield HiPure process are modeled in ProMax on two separate flow sheets for clarity. The pressure drops in the columns are assumed to be 0.2 bar, and 0.1 - 0.2 bar for the heat exchangers. The packed columns are modeled assuming a Height Equivalent of a Theoretical Plate (HETP) of 1.5 meters (5 ft)[13]. **Figures 2** and **3** represent the Benfield/hot Potassium Carbonate and DEA models in ProMax, respectively.

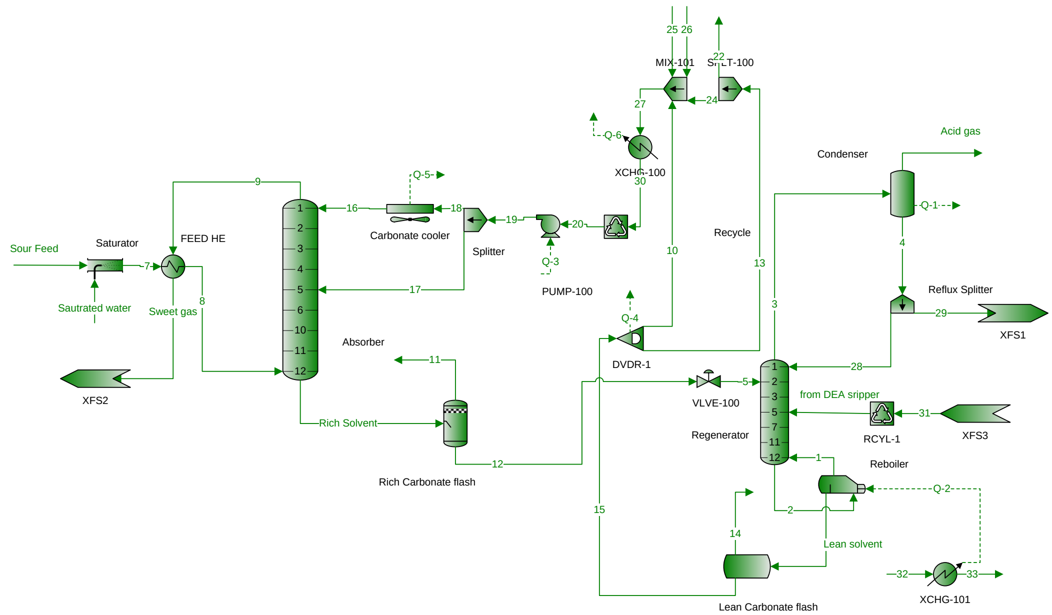

Figure 2: Schematic of Hot Potassium Carbonate Unit of the Benfield HiPure Process

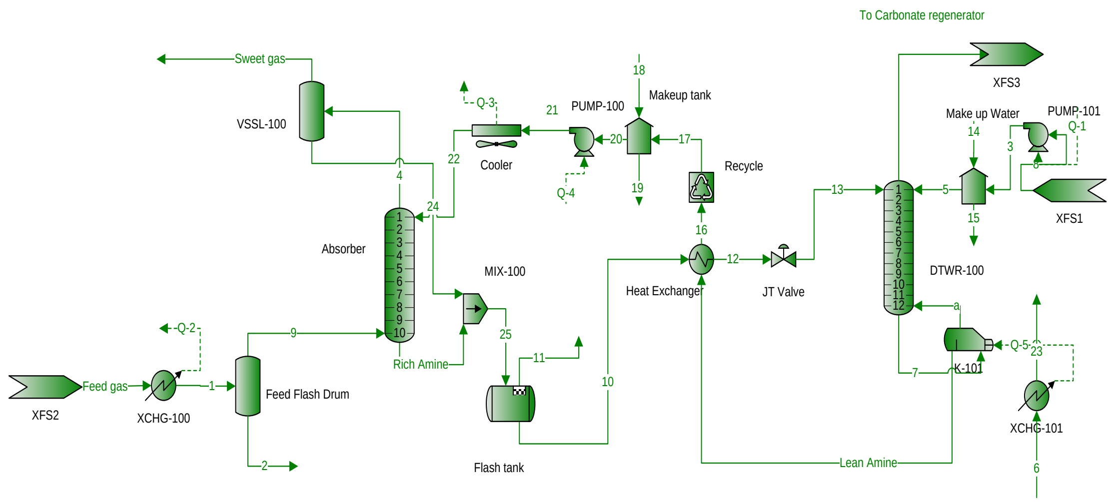

Figure 3: Schematic of the DEA Amine Unit of the Benfield HiPure Process

Comparison of Operating Data and Simulation Results

Tables 3 and **4** summarize a comparison of ADGAS operating data to the results obtained from ProMax. **Table 3** represents the Hot Potassium Carbonate unit and **Table 4** represents the DEA unit. The simulation results match the operating plant data very closely and should provide a solid basis for parametric studies.

Table 3: Comparison of Plant Operating Data to the Simulation Results for the Hot Potassium Carbonate Unit

Components	Sour Gas	Carbonate Absorber Overhead Product	
		Plant Data	Simulation
CO_2 (ppmv)	47000	574.1	570
H_2S (ppmv)	21000	707.4	683
Nitrogen (v/v%)	2.1		2.28
Methane (v/v%)	81.41		86.14
Ethane (v/v%)	5.61		5.94
Propane (v/v%)	2.71		2.87
i-Butane (v/v%)	0.36		0.39
n-Butane (v/v%)	0.60		0.64
i-Pentane (v/v%)	0.13		0.14
n-Pentane (v/v%)	0.24		0.25

Table 4: Comparison of Plant Operating Data to the Simulation Results for the DEA Unit

Components	Sweet Gas	
	Plant Data	Simulation
CO_2 (ppmv)	19	25
H_2S (ppmv)	0.41	0.40
Nitrogen (v/v%)		2.31
Methane (v/v%)		87.11
Ethane (v/v%)		6.00
Propane(v/v%)		2.90
i-Butane (v/v%)		0.39
n-Butane (v/v%)		0.64
i-Pentane (v/v%)		0.14
n-Pentane (v/v%)		0.25

Parameter Sensitivity Analysis

A parametric study is carried out to determine operating improvements to increase the LNG production capacity and efficiency. Most of the problems encountered in gas sweetening plants can be avoided through creating an envelope of operation on process parameters using a simulation tool[6,7]. Prior knowledge of anticipated conditions, operating ranges and their effects on plant performance are helpful in setting up control points and appropriately reacting to changes in the plant. A proactive approach to plant operation is always better than a reactive one.

Change in Feed Gas Flow Rate

The primary objective is to evaluate the columns in their ability to sustain an increase in gas flow. Of course, as the gas flow increases, the solvent circulation rates also need to increase. Therefore, each case is adjusted in order to maintain the current plant's sweet gas composition. The gas flow is increased from 350 MMSCFD to 750 MSCFD. The fraction flooding and corresponding solvent flow rates are reported in **Figures 4, 5 and 6**.

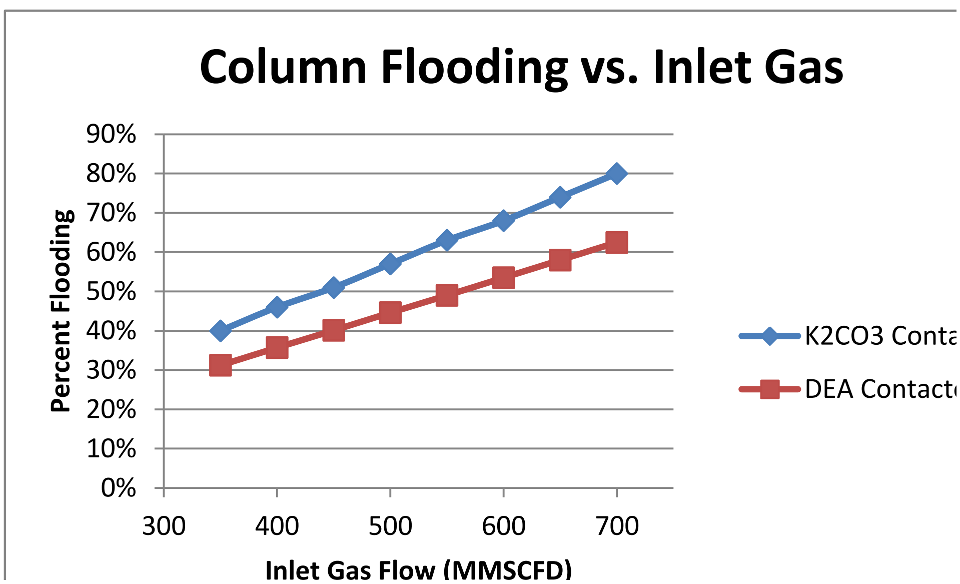

Figure 4: Effect of Inlet Gas Flow on the Column Flooding

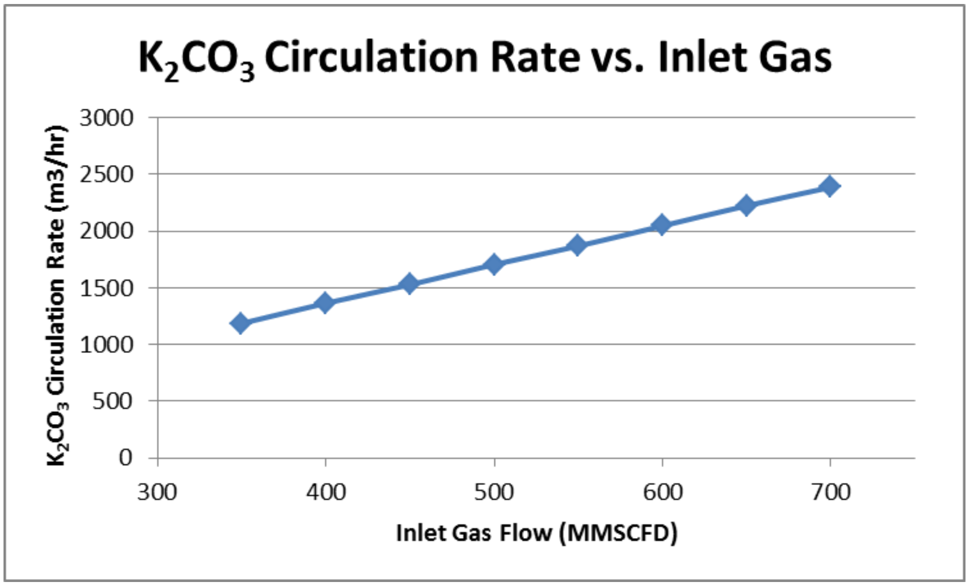

Figure 5: Effect of Inlet Gas Flow on the K_2CO_3 Circulation

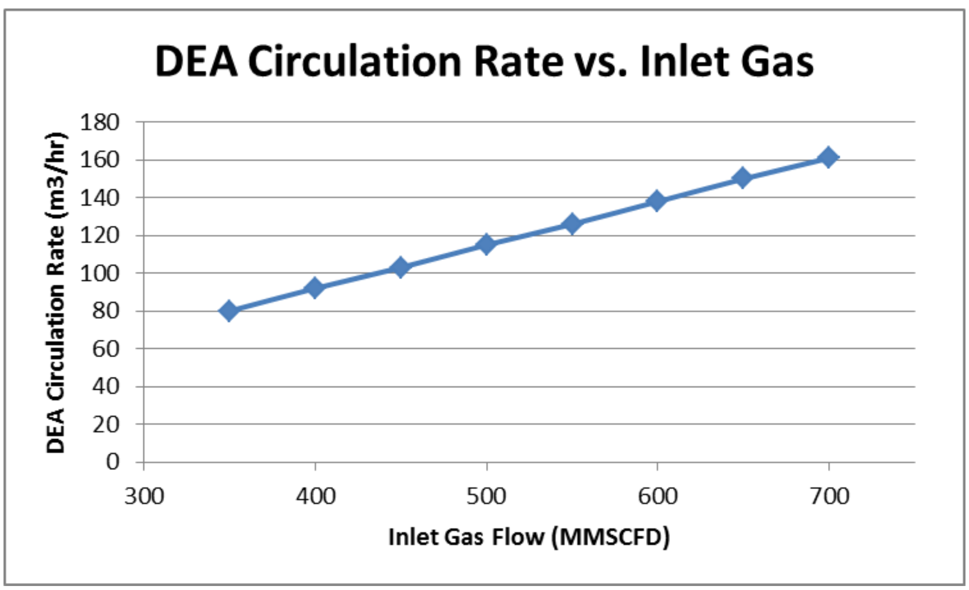

Figure 6: Effect of Inlet Gas Flow on the DEA Circulation

The plant's absorption columns have plenty of capacity to increase the inlet gas flow rate. The next step would be to evaluate the heat exchangers and pumps. Assuming only minor equipment replacement or modification, the sweetening plant may treat up to 650 MMSCFD of sour gas. At 650 MMSCFD, the plant will require 2,225 m^3/hr of K_2CO_3 and 150 m^3/hr of DEA which will result in 74%flooding in the K_2CO_3 absorber and the same sweet gas composition as current operations. For this study we are limiting flooding to 75% as calculated by the column rating utility in ProMax.

Effect of Changes in K_2CO_3 Concentration on Total Reboiler Duty

To investigate the effects of changes in solvent composition on the reboiler duty from the K_2CO_3 unit, the potassium carbonate solution concentration is varied in the range of 25-35 weight % K_2CO_3. Since CO_2 is the limiting species in this plant, the K_2CO_3 reboiler steam rate is varied in order to achieve the same CO_2 content in the treated gas as the base case. CO_2 is the limiting species since changes in operating parameters would cause the CO_2 to be off-spec more easily than the H_2S. The K_2CO_3 solvent flow rate is kept constant along with all parameters in the DEA section, including reboiler duty. The reboiler duty of the K_2CO_3 section is reported in **Figure 7**.

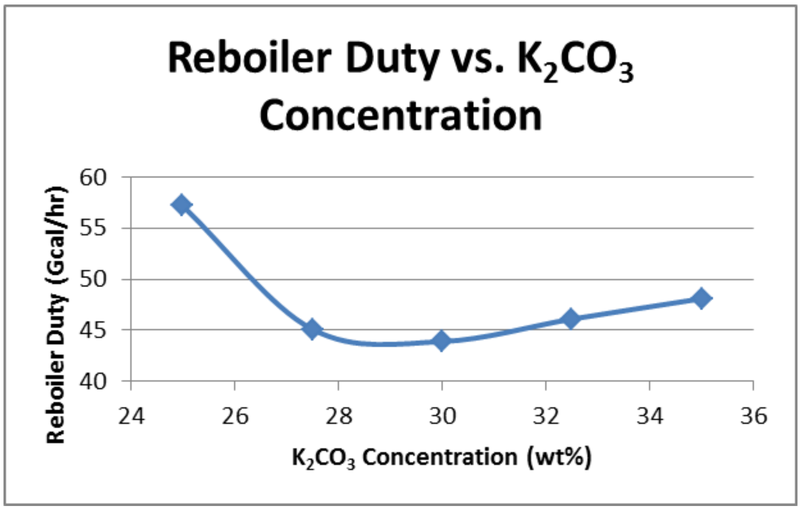

Figure 7: Effect of the K_2CO_3 Concentration on the Reboiler Duty

An optimum K_2CO_3 concentration is observed at about 29 to 30 wt% K_2CO_3. In this case, increasing or decreasing the K_2CO_3 concentration would decrease the unit's efficiency. At concentrations less than 29 wt% the reboiler duty increases due to less solvent available to absorb CO_2, requiring more duty to strip the K_2CO_3 solution to lower bicarbonate concentrations. At concentrations greater than 30 wt%, the reboiler duty increases due to more CO_2 absorbed, requiring increased energy for stripping. It is also important to avoid increasing the K_2CO_3 concentration excessively, in order to prevent salt precipitation [12].

A similar study may be performed on the DEA promoter concentration to find the optimum DEA wt%, or on the DEA concentration in the DEA unit.

Changes in Solvent Flow Rates

To investigate the effect of changes in solvent flow rate on the performance of the Benfield HiPure process, the K_2CO_3 solvent flow rate is varied between 1400 and 1700 m^3/hr. The corresponding DEA unit solvent rate is adjusted to achieve the H_2S and CO_2 treated gas compositions observed in the base case simulation. The reboiler duties are listed in **Table 5**.

Table 5: Effect of Change in Solvent Flow Rates on Reboiler Duty

K_2CO_3 Solvent Flow Rate (m^3/hr)	DEA Flow Rate (m^3/hr)	K_2CO_3 Reboiler Duty (Gcal/hr)	DEA Reboiler Duty (Gcal/hr)	Total Duty (Gcal/hr)
1700	106	48.5	15.0	63.5
1635.7 (current operation)	109.8 (current operation)	46.4	15.1	61.5
1600	110	45.6	15.5	61.1
1500	120	42.8	16.4	59.2
1450	140	41.4	19.1	60.5
1400	190	39.8	25.9	65.7

As shown in **Table 5**, there is an optimum Potassium Carbonate solution flow rate which results in the minimum overall plant duty. This occurs at about 1500 m^3/hr K_2CO_3 solvent rate and 120 m^3/hr DEA solvent rate. This would result in a 3.8% decrease in overall plant reboiler duty.

Optimized Region of Operation

In a separate study using the process simulator, the solvent concentration, circulation rates and reboiler duties are varied to obtain the best region of operation as given in **Table 6**. Whereas the previous studies were based on achieving the same H_2S and CO_2 concentrations in the treated gas as the current operating conditions, this study allows the concentrations to increase while staying within the specifications.

Table 6: Optimized ADGAS' operation variables

Optimized Variables	Current Operation	Proposed Changes
K_2CO_3/DEA Flow Rate (m^3/hr)	1635.7	1457.5
DEA Flow Rate (m^3/hr)	109.8	108.6
K_2CO_3/DEA (Wt %)	30/3	30/3
DEA (Wt %)	20	20
Carbonate Section Duty (Gcal/hr)	46.4	45.2
Amine Section Duty (Gcal/hr)	15.1	10.2
Total Reboiler Duty (Gcal/hr)	61.5	55.4
Output		
Sweet Gas Flow Rate (kg/hr)	413238	413468
H_2S Composition (ppmv)	0.4	0.98
CO_2 Composition (ppmv)	25	41.9

As seen from the table, a 10 percent decrease in solvent circulation rate and total reboiler duty is observed. ADGAS will see a slight increase in the sweet gas flow rate and a decrease in the total reboiler duty. Although the acid gas content will increase slightly as compared to the original case, the treated gas specification for LNG production is still achieved. Generally, careful observations should be considered when involving solvent concentrations in the global parametric optimization.

Conclusions

The current Benfield unit has various optimization steps it may take to improve operations and increase revenue. While the current unit is operating very close to the optimum K_2CO_3 concentration, optimization of solvent circulation rates and reboiler duties at a fixed solvent concentration can save up to 10% of the energy used in the original case.

The simulation study also shows the potential to increase the sour gas flow rate to the plant, should existing pumps and exchangers support it. When the gas flow is increased to 650

MMSCFD, the columns have a maximum flooding of 74%. If the gas is available, this can lead to a significant increase in revenue. For example, assuming the LNG is sold for $8/MMBTU and gas quality of 1124 BTU/SCF, ADGAS has the potential to increase their sales by nearly $1.6 million per day or $500 million per year. While detailed engineering will need to be performed to increase capacity by this magnitude, a 10% reduction in energy cost may be realized immediately.

Additional studies might be performed to help further optimize the plant. Such studies might include examining the effect of reducing cooler temperature on the overall plant efficiency, or finding the optimum promoter concentration, or even the effects of a K_2CO_3 promoter other than DEA.

Acknowledgement

We acknowledge GASCO and ADGAS operating companies of Abu Dhabi National Oil Company (ADNOC) for their technical and financial support.

References

[1] AlBishr, T., "Train-3 Acid Gas Removal Unit Performance," in *SOGAT Proceedings of 8th International Conference*, Abu Dhabi , U.A.E, 2012, pp. 511-532.

[2] Maddox , R. N., *Gas and Liquid Sweetening*, 2nd ed., Campbell Petroleum Series, John M. Campbell, Norman, OK, 1974.

[3] Kohl, A. L. and Riesenfield, F. C., *Gas Purification*, 3rd ed., Gulf Publishing, Houston, TX, 1979.

[4] Kohl, A. and Nielsen, R., *Gas Purification*, 5th ed., Gulf Publishing Co., Houston, TX, 1997.

[5] Benson, H. E. and Parrish, R. W., "HiPure Process Removes CO_2/H_2S", *Hydrocarbon Processing*, April 1974, pp. 81-82.

[6] Mohammad, R., Schneiders, L., and Neiderer, J., "Carbon Dioxide Capture from Power Plants: Part I : A Parametric Study of the Technical Performance Based on Monoethanolamine," *International Journal of Greenhouse Gas Control*, Vol. 1, pp. 37-46, 2007.

[7] Hassan, B. and Shamim, T., "Parametric and Exergetic Analysis of a Power Plant with CO_2 and Capture using Chemical Looping Combustion," in *International Conference on Clean and Green Energy*, Singapore, 2012.

[8] Field, J. H., Johnson, G. E., Benson, H. E., and J. S. Tosh, "Removing Hydrogen Sulfide by Hot Potassium Carbonate Absorption," *Bureau of Mines Report of Investigations*, 1960, p. 5660.

[9] ProMax Version 3.2, Bryan Research and Engineering, Inc., 2012.

[10] Abry, R. G. and Dupart, R. S., "Amine Plant Troubleshooting and Optimization," *Hydrocarbon Processing*, April 1995.

[11] Bullin, J. A., Polasek, J. C., and Holmes, J. W., "Optimization of New and Existing Amine Gas Sweetening Plants Using Computer Simulation " in *Proceedings of the 60th Annual GPA Convention*, Tulsa, Oklahoma, Gas Processors Association, 1981, pp. 142-148.

[12] Bocard, J. P. and Mayland, B. J., "New Charts for Hot Carbonate Process," *Hydrocarbon Processing & Petroleum REFINER*, Vol. 41, No. 4, April 1962.

[13] Shiveler, G., Solis, G. S., Gonzalez, L. H. P., and Bueno, M. L., "Retrofit of a H_2S Selective Amine Absorber Using MellapakPlus Structured Packing," in *Proceedings of the 2005 Spring AIChE Meeting*, 2005.