Troubleshooting of ADGAS' Benfield - HiPure Plant of Natural Gas Sweetening Using Process Simulation

R. Ochieng*, A. S. Berrouk, C. Peters
Department of Chemical Engineering, Petroleum Institute, Abu Dhabi-UAE
J. Slagle, L.G. Lyddon, P. E. Krouskop
Bryan Research & Engineering, Inc., Bryan, Texas-USA
Email: ochienrich@gmail.com

ABSTRACT

Natural Gas processing plants, an essential part of the energy industry, provide one of the cleanest-burning fuels and valuable chemical feedstock. The importance and complexity of gas processing plants have increased over the years to increase their energy efficiency and their integration with petrochemical plants. The advantages of computer simulation models as tools for designing and troubleshooting gas treating plants is increasingly obvious.

This paper will discuss the major problems faced in operation of the Benfield HiPure process at Abu Dhabi Gas Liquefaction Company Limited (ADGAS) and the use of a process simulation tool, ProMax®, to investigate and suggest ways of overcoming some of these problems. 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 also used to provide physical insight on how the changes in process parameters can affect ADGAS' plant performance, so as to suggest ways of improving or avoiding some these changes.

Keywords: Simulation, Benfield HiPure, parametric study, ProMax, Gas Sweetening, LNG

Troubleshooting of ADGAS' Benfield - HiPure Plant of Natural Gas Sweetening Using Process Simulation

INTRODUCTION

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 (1 & 2) have been in operation since 1977, each producing a capacity of 180 tons per hour of LNG. The third one (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 of a "Benfield HiPure" design supplied by UOP, and is a hybrid arrangement of the basic Benfield and Amine units. In this paper, an outline of some of the problems faced in the operation of ADGAS' Train 3 gas treating plant will be provided, and possible measures that can be taken to solve some of these problems through process simulation are suggested.

The Benfield HiPure Process of ADGAS

The Benfield HiPure design was described in 1974 by Benson and Parrish [2]. 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 does the final trim removal of the acid gases to produce 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.

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 then 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 30 ppmv CO_2 which meets its required design specifications [1].

Both the Carbonate and DEA absorbers operate at a pressure of about 50 bar(g) and the regenerators operate at lower pressures of about 0.8bar(g). The necessary heat load for the regeneration is supplied through reboilers associated with each of the regenerators.
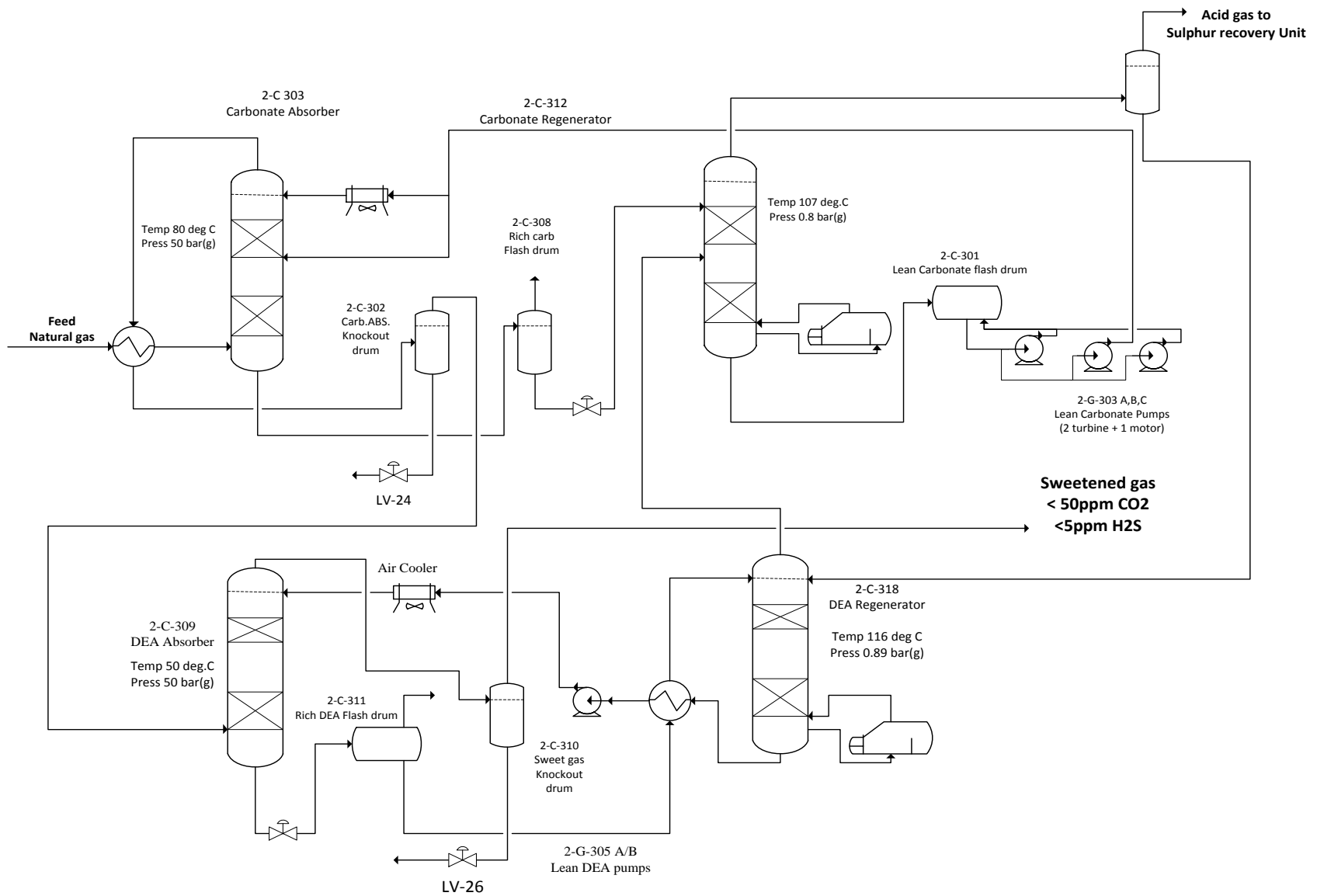
2-C 303
Carbonate Absorber

2-C-312
Carbonate Regenerator

Acid gas to
Sulphur recovery Unit

Temp 80 deg C
Press 50 bar(g)

Temp 107 deg.C
Press 0.8 bar(g)

2-C-308
Rich carb
Flash drum

2-C-301
Lean Carbonate flash drum

Feed
Natural gas

2-C-302
Carb.ABS.
Knockout
drum

2-G-303 A,B,C
Lean Carbonate Pumps
(2 turbine + 1 motor)

LV-24

Sweetened gas
< 50ppm CO2
<5ppm H2S

Air Cooler

2-C-318
DEA Regenerator

2-C-309
DEA Absorber

Temp 50 deg.C
Press 50 bar(g)

Temp 116 deg C
Press 0.89 bar(g)

2-C-311
Rich DEA Flash drum

2-C-310
Sweet gas
Knockout
drum

2-G-305 A/B
Lean DEA pumps

LV-26

Figure 1: ADGAS –Gas Sweetening Plant –Process Schematic

Operation Issues at ADGAS Plant

Issues with operation of the acid gas sweetening unit may occur whenever the feed gas source changes, when liquid hydrocarbon entrainment from such actions as pigging operations upstream occurs, or as a result of process upsets within the plant. One of the major issues encountered in the acid gas sweetening unit is foaming in the packed bed absorbers. This foaming causes large pressure drops and poor gas/liquid contact which results in high slippage of the acid gases into the overhead product.

Experience has shown that the DEA absorber is highly sensitive to foaming issues and requires vigilant observation. When there is a foaming issue in the DEA absorber, the pressure drop across the packed beds is high and fluctuates frequently. In cases of severe foaming, there could be liquid carryover in the exiting gas from the top bed. These effects may require a reduction in the feed gas rate to the absorber to correct the situation. The reduction in rate is undesirable since it ultimately reduces the LNG production. Recent reports from ADGAS also have shown that a change in process parameters during operation is another cause of these problems. As reported by Tamadher [1], one of the major causes of higher pressure drops and foaming in the absorber is a high flow rate of solvent. In this case critical process monitoring was an important issue so the effects of change in the sensitive parameters such as solvent flow rate, steam flow rate, feed gas quality, and solvent concentrations is noted.

Table A shows the sample operating data and Table B shows the absorber configurations of ADGAS' Train 3 plant. From this data a normal operation of the plant may be expected; however, due to unavoidable circumstances such as change in feed conditions, aging of the plant, etc., the actual operation tends to deviate from normal requiring immediate attention to stabilize production.

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

Parameter		Value
Feed Gas Flow Rate (MMSCFD)		476.93
Feed Gas Temperature (^{o}C)		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 (^{o}C)	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 (^{o}C)		49.94
Lean Solvent Pressure (barg)		53.71
DEA Concentration (wt %)		20

Table B: 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

To illustrate the changes in the quality of feed gas to the Train 3 plant, the temperature, pressure, and composition of H_2S and CO_2 in the feed gas were plotted against time for a period of 192 hours as shown in Figures 2 and 3, respectively. Figure 2 shows that the temperature of the feed gas is highly unstable. Figure 3 shows that the H_2S feed composition is more stable than that of CO_2. The figures also show that there was a large disturbance in feed gas pressure and composition immediately after the 92nd hour.
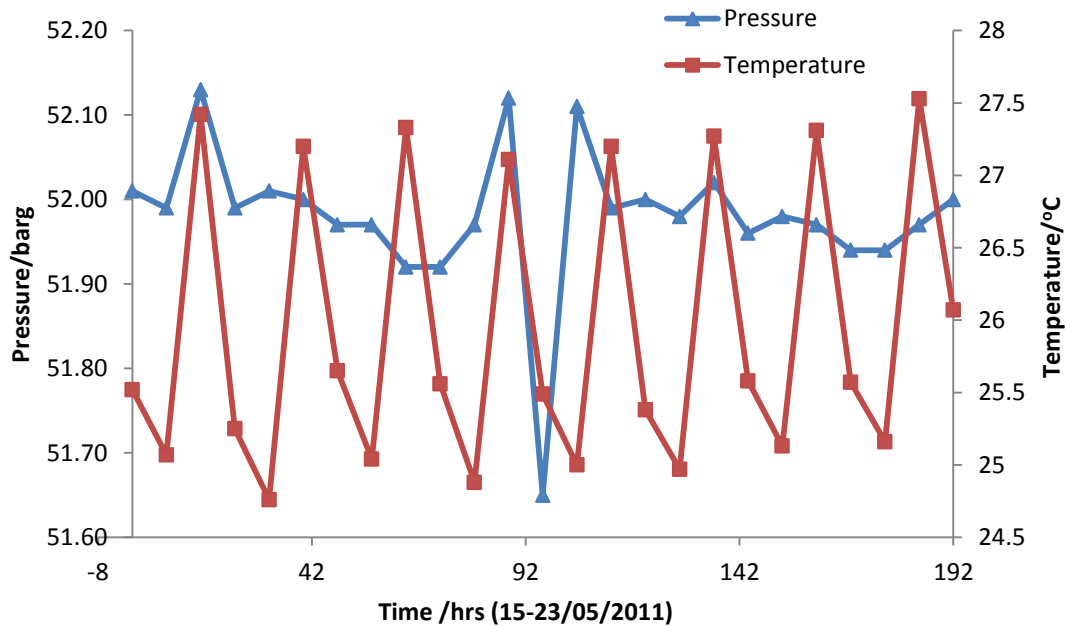

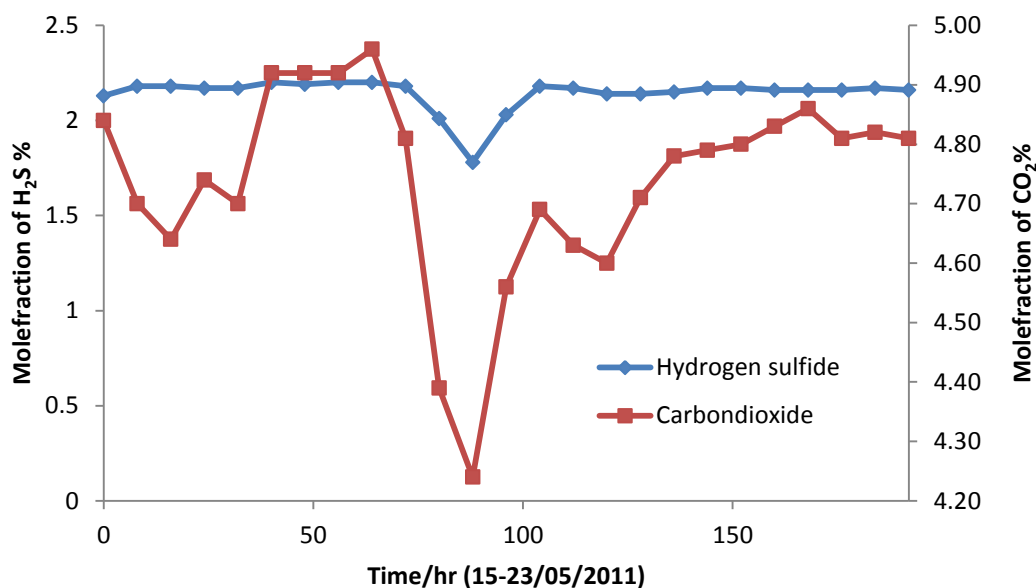**Figure 2: Pressure and Temperature Changes in the Feed Gas for a Period of 192 Hours**

Figure 3: Sour gas Composition Change in the Feed Gas for a Period of 192 Hours

The changes in the feed gas quality reported in Figures 2 and 3 are large enough to destabilize the absorbers and cause a reduction in production capacity. The changes in pressure and temperature can cause condensation of hydrocarbons and lead to foaming problems. The changes in composition (Figure 3) may lead to changes in the H_2S/CO_2 ratio which is likely to affect the overhead product. These fluctuations in feed gas quality may be attributed to changes in sour gas supplier, changes in reservoir conditions, or receiving gas from multiple wells.

BENEFITS OF COMPUTER MODELING FOR PROCESS OPTIMIZATION

In chemical processes, the optimization of processes depends on how well the process can be controlled in response to the changes in the operating environment. The control depends on whether the appropriate parameters are measured, which ultimately depends upon how well a process is understood. Many computer based process modeling techniques have been developed, which along with advances in networking and communication, facilitate capturing of sufficient data and thorough analysis to respond dynamically to the fluctuations in the process environment [3].

Simulation models help illuminate the bottlenecks in the processes and dependence on critical resources required for deriving optimum performance and improving complex cross functional processes. 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 [4, 5].

Much work has been published on the modeling and simulation of both the Hot Potassium Carbonate [6, 7, 8] and Amine systems [4, 9, 10]; however, limited studies have been available for the Benfield HiPure process [11]. This work focuses on the simulation of the ADGAS Train 3 plant using ProMax® modeling software due to its

capabilities in modeling gas sweetening processes [12]. ProMax will be used to perform a parametric study to provide guidance for new operating conditions that will help stabilize performance of the gas sweetening unit.

ADGAS PLANT SIMULATION MODEL

The plant model was set up in ProMax using current operating conditions to set a benchmark for the case studies. Since the solvents are strong electrolyte solutions, the "Electrolytic ELR-PR" property package supplied with ProMax was 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 are treated as ionic species in solution when using the Electrolytic Property Package. The TSWEET Kinetics model was used for predicting the CO_2-amine or CO_2-carbonate kinetic reactions taking place in acid gas absorbers. TSWEET kinetics accounts for the relatively slow absorption of CO_2 by amine or carbonate solutions.

The two units of the Benfield HiPure process were modeled in ProMax on two separate flow sheets for clarity. The pressure drops in the columns were assumed to be 0.2 bar, and 0.1 - 0.2 bar for the heat exchangers. The packed columns were modeled assuming an HETP of 1.5 meters (5 ft) [15]. Figures 4 and 5 represent the Benfield/hot Potassium Carbonate and DEA models in ProMax, respectively.
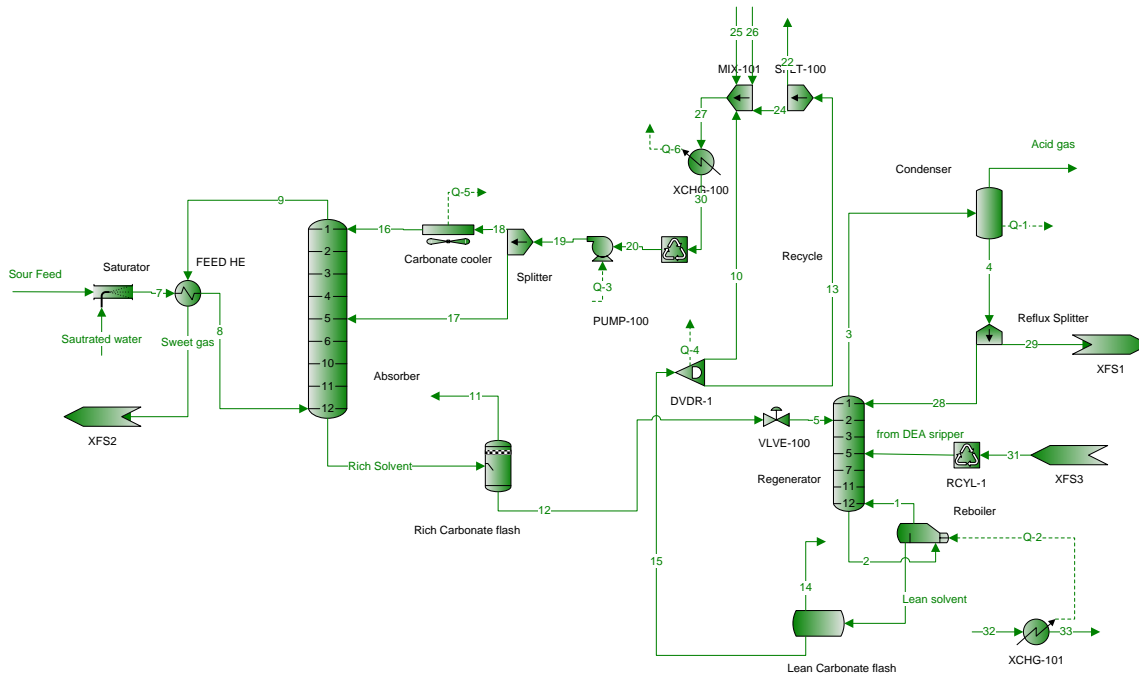

Figure 4: Schematic of Hot Potassium Carbonate Unit of the Benfield HiPure Process

Figure 5: Schematic of the DEA Amine Unit of the Benfield HiPure Process

Comparison of Operating Data and Simulation Results

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

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

Components	Sour Gas Feed (mol %)	Carbonate Absorber Overhead Product (ppmv)		
		Design Data	Plant Data	Simulation
CO_2	4.7	1232.88	574.1	570
H_2S	2.1	759	707.4	683
Nitrogen	2.1	14345.68		22818.46
Methane	81.406	853935.92		861351.56
Ethane	5.611	68661.45		59383.61
Propane	2.707	30510.97		28655.37
i-Butane	0.364	560.4		3855.63
n-Butane	0.601	10241		6356.60
i-Pentane	0.129	2584.19		1365.06
n-Pentane	0.236	2584.19		2500.10
COS	0.005	0.4		49.82
Methyl Mercaptan	0.003	12.1		25.84
Ethyl Mercaptan	0.005	35		47.70
Isopropyl Mercaptan	0.003	24.7		25.95
n-Propyl Mercaptan	0.001	12.4		13.16
Butyl Mercaptan	0.001	10		9.95

Table D: Comparison of Plant Data to the Simulation Results for the DEA Amine Unit

Components	Sweet Gas (ppmv)		
	Design Data	Plant Data	Simulation
CO_2	49.83	19	25
H_2S	0.4983	0.41	0.40
Nitrogen	14436.96541		23080.82
Methane	859281.22		871149.96
Ethane	69087.96		60057.92
Propane	30702.1		28983.20
i-Butane	5628.93		3900.03
n-Butane	10306.84		6429.53
i-Pentane	2601		1380.82
n-Pentane	2601		2528.93
COS	0.2		50.30
Methyl Mercaptan	10		25.38
Ethyl Mercaptan	30.8		47.31
Isopropyl Mercaptan	22.3		25.93
n-Propyl Mercaptan	11.3		13.15
Butyl Mercaptan	9.8		10.02

PROCESS PARAMETRIC STUDY

Due to the problems faced by ADGAS plant operation, a parametric study was carried out to determine possible solutions to the plant instabilities and to also suggest improvements to increase the LNG production capacity. 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 [5]. 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.

In this work, the following process parameters are considered for study of the ADGAS plant: changes in feed flow rate, feed gas H_2S/CO_2 ratio, HiPure solvent composition, and solvent volumetric flow rates. The sensitivity analysis was carried out on each of these process parameters while keeping others constant except as noted. This study will help ADGAS operators understand the cause-and-effect relationship between process parameters which will guide in monitoring of process operations.

Changes in Feed Gas Flow Rate

To determine the effect of changes in feed gas flow rate on the process performance, the concentrations of H_2S and CO_2 in the treated gas were observed as the feed gas flow rate was varied over a range of 200 to 500 million standard cubic feet per day (MMSCFD). Figure 6 below illustrates the increases in H_2S and CO_2 content in the potassium carbonate absorber overhead stream as the feed gas rate increases.
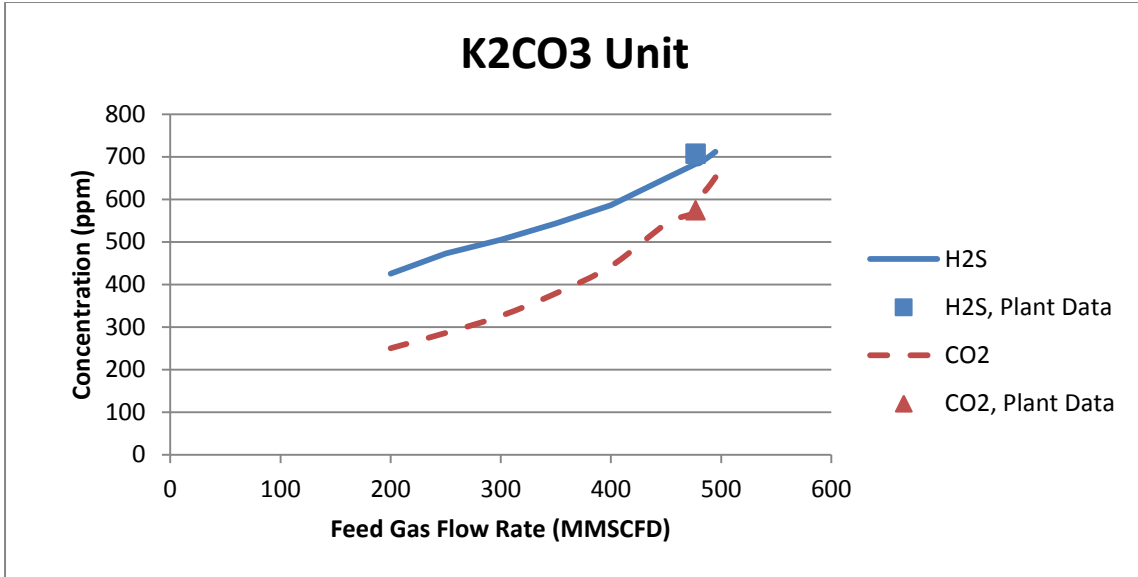
Figure 6 : Effect of Increasing Feed Gas Flow Rate on the Potassium Carbonate Absorber Acid Gas

Figure 7 below illustrates the H_2S and CO_2 content in the DEA absorber overhead stream as the feed gas rate increases. The CO_2 increases as expected with increasing feed gas rate, but only from 2.1 to 31.8 ppmv since the DEA absorber sour feed gas CO_2 content is relatively low with the bulk of the acid gas having been removed in the Potassium Carbonate unit. The H_2S is about constant, changing by only 0.1 ppm between 200 and 500 MMSCFD. Unlike the CO_2, the H_2S actually decreases slightly with increasing feed gas rate. At lower feed gas rates there is less CO_2 per amount of solvent since the solvent rate is held constant. This results in more CO_2 being absorbed at the lower flow rates which displaces the H_2S.
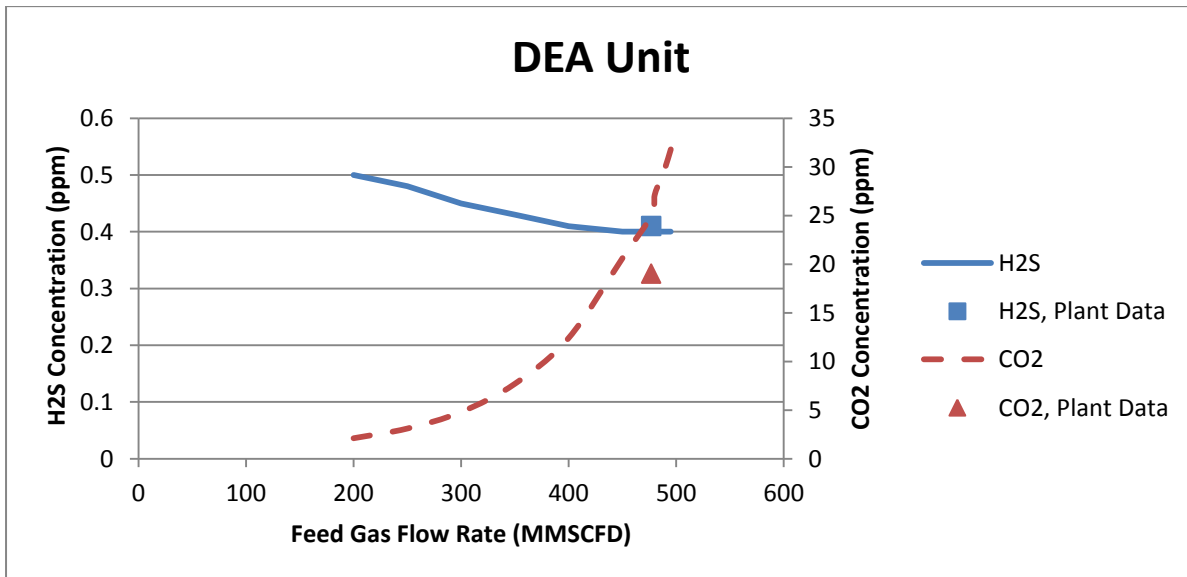
Figure 7: Effect of Increasing Feed Gas Flow Rate on the DEA Absorber Acid Gas

The plots indicate that the feed gas rate should not be increased much above the current inlet rate of 477 MMSCFD at current conditions of composition, temperature and pressure to ensure the treated gas meets the H_2S and CO_2 specifications.

Changes in Feed Gas H_2S/CO_2 Ratio

To understand the effect of feed acid gas composition (excluding mercaptans) to the Benfield HiPure plant, the feed gas H_2S/CO_2 ratio was varied from 0.05 to 0.6. The effect of this ratio on acid gas removal is shown in Figures 8 and 9 below. Figure 8 shows that an increase in the feed H_2S/CO_2 ratio increases the amount of H_2S and CO_2 absorbed by the Potassium Carbonate unit. There is also an observed increase in acid gas absorption for the overall sweetening unit (Figure 9).

Potassium Carbonate Unit Only
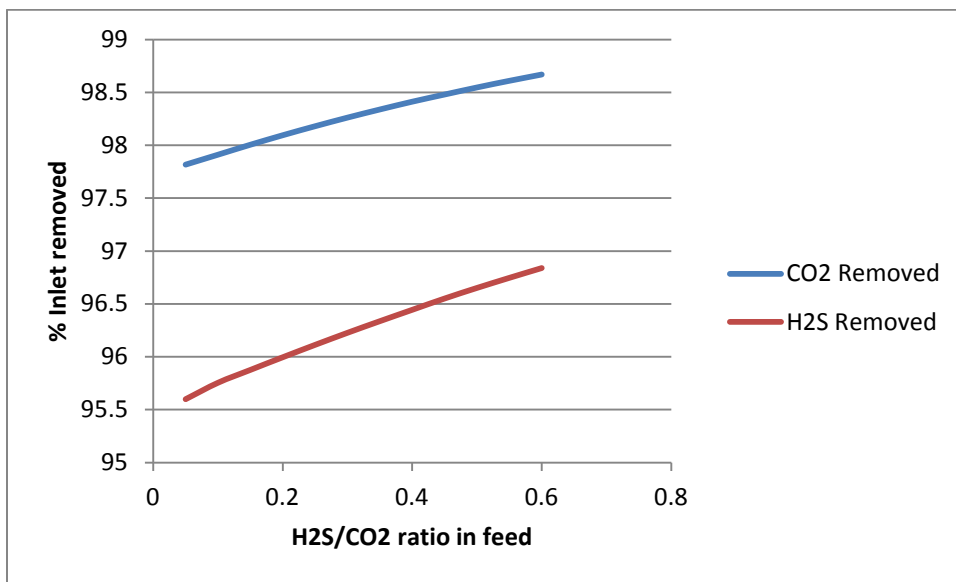
Figure 8: Effect of Increasing Feed Gas H_2S/CO_2 Ratio on Acid Gas Removal on the Carbonate Unit

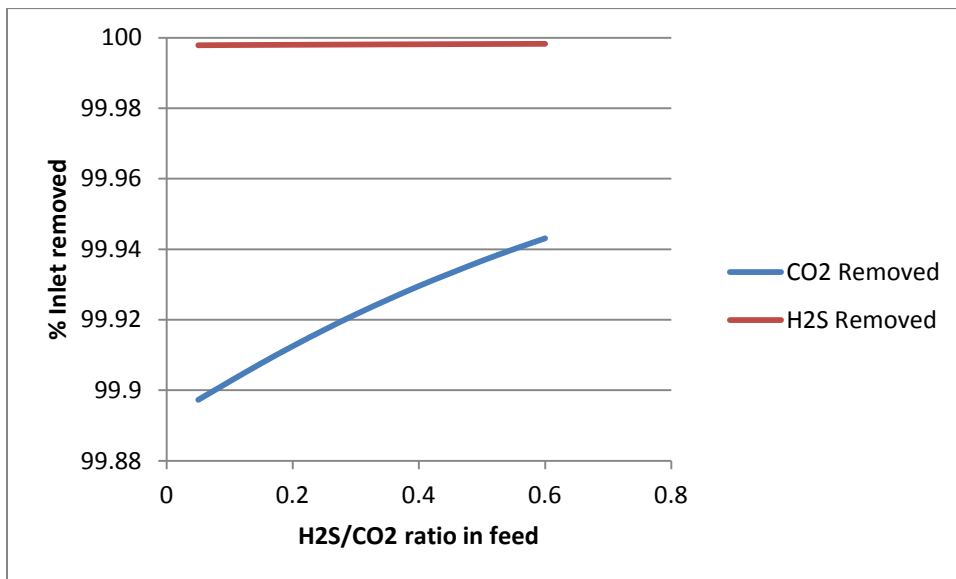**Potassium Carbonate Unit and DEA Amine Unit Together (Total Sweetening Plant)**

Figure 9: Effect of Increasing Feed Gas H_2S/CO_2 Ratio on Acid Gas Removal on the Overall Plant

These simulations were performed by keeping the Potassium Carbonate and DEA solution flow rates, reboiler steam rates, inlet gas flow rate, and inlet acid gas composition constant. The relative amount of CO_2 and H_2S is the only variation in the simulation cases.

The hot potassium carbonate process removes the bulk of the acid gas while the DEA amine unit of the Benfield HiPure process is considered the final polishing step in the removal of acid gases from the natural gas. Since the DEA section has limited contact with the acid gas, increasing the H_2S/CO_2 ratio will increase the rate of absorption of both acid gases.

Changes in K_2CO_3 and DEA Promoter Composition

To investigate the effect of changes in solvent concentration to the performance of ADGAS plant, the solvent concentrations of the Potassium Carbonate unit were varied simultaneously in the range of 20-33 weight % K_2CO_3 and 0-6 weight % DEA (which acts as a promoter). Changes to gas production and net reboiler duty were investigated.

Effect of Changes in K_2CO_3 and DEA Composition on Gas Production

Figure 10 shows that operating at a promoter concentration of 4-5 weight % DEA promoter while decreasing the hot potassium carbonate concentration to about 20-24 weight % will allow a slight 1.2 % increase in the production of gas meeting the required specifications (Gas Prodn). This represents a gain in the overall output of the plant. However; the reader should note that this happens only when the other process parameters are undisturbed.
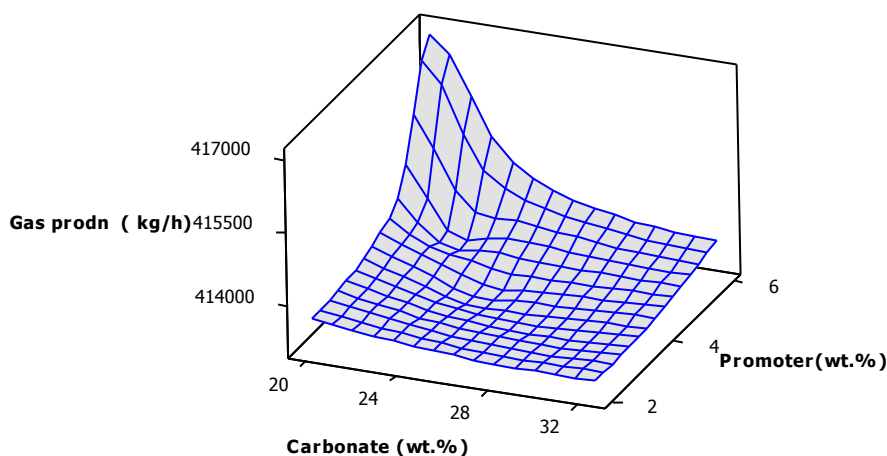

Figure 10: Effect of Change in Hot K_2CO_3 and Promoter Concentration on Gas Production

The increase in gas production with a decrease in potassium carbonate solution concentration can be attributed to the reduced absorption of hydrocarbons from the gas into the solution. Hydrocarbons are more soluble in higher concentration solvents [13], however, the solubility of hydrocarbons in Potassium Carbonate solution is relatively low, and the DEA concentration of the solution is low at 6 wt% or less. Although a higher concentration of the K_2CO_3 with DEA promoter is desired to meet gas specifications, an optimum may be determined to achieve minimum hydrocarbon losses.

Effect of Changes in K$_2$CO$_3$ and DEA Composition on Total Reboiler Duty

To investigate the effects of changes in solvent composition on the total reboiler duty from both units, the potassium carbonate solution concentration was varied in the range of 20-35 weight % K$_2$CO$_3$ and 0-6 weight % DEA promoter. The minimum net duty is obtained at a range of 20-26 weight % K$_2$CO$_3$ and 4-6 weight % DEA promoter concentrations as shown in Figure 11 below.
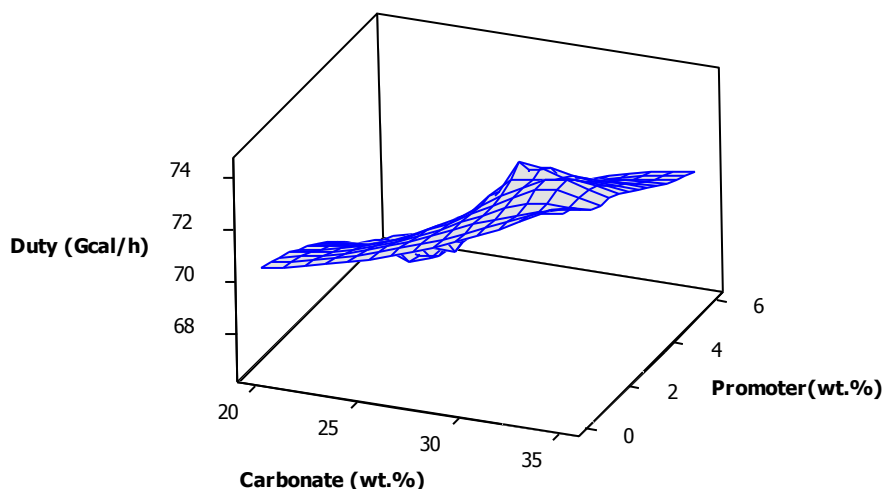

Figure 11: Effect of Change in Hot K$_2$CO$_3$ and Promoter Concentration on Net Reboiler Duty

The decrease in concentration of the solvent leads to a reduction in energy required to overcome the heat of reaction in the absorber, which later reduces the amount of energy required to regenerate the solvent. Though the concentration of the promoter increases by 33-100%, its overall amount is small and only introduces a slight increase in the required energy.

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$_2$CO$_3$ solvent flow rate was varied between 1400 and 1700 m^3/hr while the corresponding DEA unit solvent rate was found which would achieve the H$_2$S and CO$_2$ treated gas compositions currently observed in the simulated data set. Table E below lists the individual and total reboiler duties.

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

K$_2$CO$_3$ Solvent Flow Rate (m^3/hr)	DEA Flow Rate (m^3/hr)	K$_2$CO$_3$ Reboiler Duty (Gcal/hr)	DEA Reboiler Duty (Gcal/hr)	Total Duty (Gcal/hr)
1700	106	48.5	15.0	63.5
1625 (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 E there is an optimum Potassium Carbonate solution flow rate which results in the minimum overall plant duty.

CONCLUSIONS

The simulation results matched well with the plant data, providing a solid basis for parametric studies. Plant simulation indicates that an increase in feed gas supply from current operation of 477 MMSCFD to about 500 MMSCFD will cause the CO_2 content of the treated gas to approach the 50 ppmv specification with no significant effect on the H_2S content of the treated gas.

Feed gas fluctuations with increasing H_2S/CO_2 ratio might increase the acid gas removal in the Potassium Carbonate unit. Likewise, an increase in acid gas removal will be observed in the overall plant. Operating at a hot potassium carbonate concentration of about 20-24 weight % and 4-5 weight % DEA promoter could increase the gas production rate by 1.2% and reduce the reboiler duty by 5% of the current operation.

The simulations also show that there is a minimum overall plant reboiler duty which can be achieved by decreasing the hot potassium carbonate solution circulation rate and increasing the DEA solution circulation rate to a level that still allows the treated gas specifications to be achieved.

Other process parameters such as operating pressure and steam rate of both the Potassium Carbonate and DEA units could also be investigated to find the optimum operating conditions. Besides the parametric monitoring, other important observations should be utilized to stabilize plant operations. Such observations may include frequent monitoring of the accumulation of heavy hydrocarbons in the flash tanks to avoid solvent contamination. The recirculated solutions should also be closely monitored to avoid foaming caused by the contaminants.

ACKNOWLEDGEMENT

REFERENCES

1.   Tamadher A. B, "Train-3 Acid Gas Removal Unit Performance", SOGAT, *Proceedings of the 8th International Conference,* Abu Dhabi-U.A.E 2012, pp. 511-531.

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

3.   http://www.alagse.com/pm/p10.php, *Alagse Article.*

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

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

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

7. Yih, Siu-Ming, and Hung-Cherh Lai, "Simultaneous Absorption of Carbon Dioxide and Hydrogen Sulfide in a Packed Absorber-Stripper Unit", *Chem. Eng. Comm.*, Vol. 51, 1987, pp. 277-290.

8. Park, K. Y, and T. F. Edgar, "Simulation of the Hot Carbonate Process for Removal of CO_2 and H_2S from Medium Btu Gas," *Energy Progress*, Vol. 4, No. 3, 1984, pp. 174-180.

9. Kohl, A., and R. Nielsen, *Gas Purification,* 5th ed., Gulf Publishing Co., Houston, Texas, 1997.

10. Maddox, *Gas and Liquid Sweetening,* 2nd ed., Campbell Petroleum Series, John M. Campbell, Norman, OK, 1974.

11. Ochieng, R., A. S. Berrouk, C. J. Peters, P. E. Krouskop, L. G. Lyddon, and J. Slagle, "Simulation of the Benfield HiPure Process of Natural Gas Sweetening for LNG Production and Evaluation of Alternatives," *Proceedings of the SOGAT 8th International Conference ,* Abu Dhabi-U.A.E, 2012, pp. 63-77.

12. http://www.bre.com/; *Bryan Research and Engineering, Inc.*

13. Bullin, J. A., and W. J. Brown, "Hydrocarbons and BTEX Pickup and Control from Amine Systems," *Proceedings of the 83rd Annual GPA Convention*, Gas Processors Association, 2004.