Addition of Static Mixers Increases Treating Capacity in Central Texas Gas Plant

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ABSTRACT

Due to the addition of new wells, the feed to Ferguson Burleson County Gas Gathering System (FB) Anderson gas treating plant was scheduled to increase from 180 MMSCFD to a design capacity of 210 MMSCFD. The feed gas contained both CO2 (6.0 mol %) and H2S (25 ppm) at high pressure (980 psig). A feasibility study determined that a cost effective method to handle the additional gas volume was to switch from a single amine to an amine mixture and to add static mixers to treat a bypass stream for H2S. The primary amine contactors would perform bulk removal of the acid gases from the main stream. Although increased treating capacity was the ultimate goal, a secondary concern was that corrosion be kept to within acceptable limits. To be certain this concern was adequately addressed, all of the heat exchangers within the system were rated and heat fluxes were investigated for possible problem areas. Several different amine mixtures were also evaluated with an eye toward potential corrosion limitations. This paper discusses the modifications made to the system and the results of subsequent plant trials to determine the overall capacity increase.

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INTRODUCTION

The Ferguson Burleson County Gas Gathering System (FB) is a joint venture between a subsidiary of Mitchell Energy and Union Pacific Resources. FB operates a lean gas gathering system that spans Washington and Grimes counties in East Texas. The pipeline system gathers gas from wells and transports it to the FB Anderson treating plant where the gas is treated to remove acid gases and dehydrated for further distribution. In the fall of 1996, the FB Anderson plant found itself in the enviable position of having a gas surplus due to an increased drilling schedule. The existing plant capacity was approximately 180 MMSCFD, and the anticipated inlet gas flow rate for December 1996 was 210 MMSCFD. With the need for additional treating, a feasibility study was performed to determine the most cost effective method to achieve this goal. The end result of the study was to switch from a single amine to a mixed amine solution and install static mixers to treat bypass gas. Optimization of the existing regeneration system was necessary as well, since it was operating near full capacity.

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INLET GAS CONDITIONS

The inlet gas typically contains between 5.5 and 6.0 mol % CO_2 and 25 ppm of H_2S. The operating pressure is dependent upon the line pressure of the residue pipeline, but normally ranges from 950 to 1050 psia. The temperature of the gas is 60-90 ° F from the pipeline, but the feed to each of the amine contactors is controlled at approximately 102 ° F by a gas to gas exchanger equipped with a bypass. At the origination of the project, the plant was using a 50 wt.% MDEA solution and was able to treat the gas to a CO_2 outlet of around 2.7 mol % at a peak inlet flow rate of 180 MMSCFD. H_2S in the outlet gas was well below the 4 ppm specification.

PROCESS CONFIGURATION

Figure 1 is a process flow diagram of the FB Anderson plant prior to the addition of the static mixers. Two six foot diameter amine contactors were operated in parallel. The remainder of the system included the following: a single rich amine flash tank, carbon and sock filters, lean/rich exchangers, a seven foot diameter regenerator with associated reboiler and condenser, an amine surge tank, amine booster pumps, and amine coolers. Each piece of equipment was evaluated to determine where any process bottlenecks would occur.

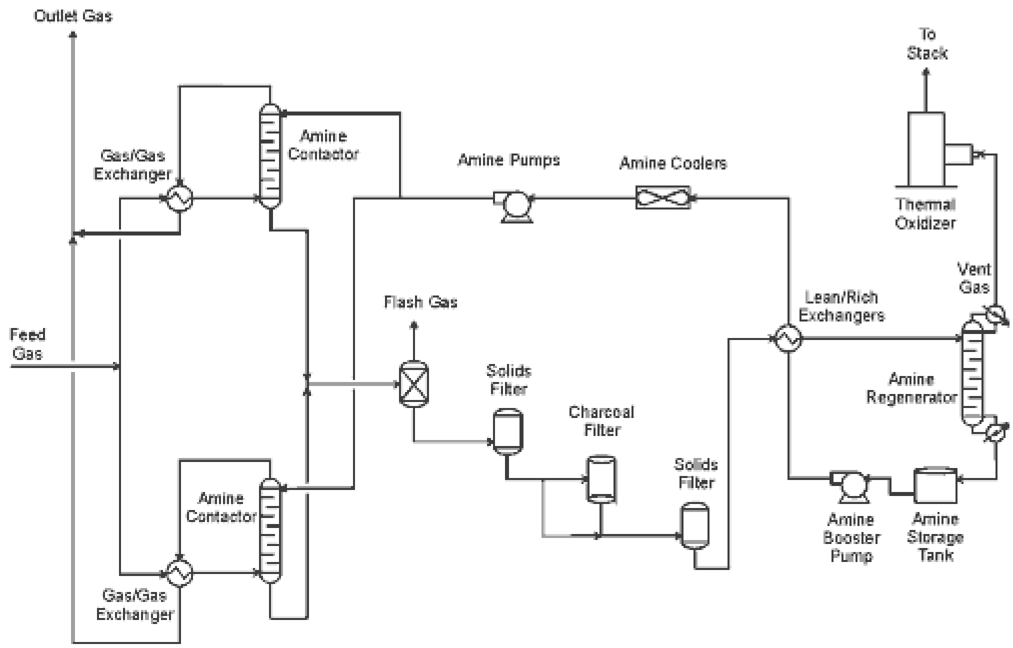

Figure 1. Anderson Plant Before Modification.

The amine contactors had a gas flow limit of 90 MMSCFD and a liquid handling capacity of 400 gpm of amine solution. Simulation results from the process simulator TSWEET® showed that even if the contactors were able to handle the additional gas (30-40 MMSCFD), 400 gpm of 50 wt.% amine solution would not be adequate to treat below the 3.0 % CO_2 specification. Rating of the lean/rich exchangers and regenerator showed that they were also running close to 100 % capacity.

PROPOSED CHANGES

Due to the capital expense of increasing the circulation rate of the regeneration system, the addition of a third contactor was eliminated as a possible answer. Instead, it was decided to maximize the existing system by the addition of DEA to the amine solution. To meet the required outlet CO_2 specification, a solution of 30 wt % MDEA and 20 wt % DEA was targeted. The mixed solvent would allow the facility to treat a 180 MMSCFD inlet stream

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down to approximately 2.2 mole % CO_2. The reduction of CO_2 in the outlet stream meant that approximately 40 MMSCFD of gas could be bypassed around the towers and then recombined for an overall outlet CO_2 concentration of 2.7-2.8 mol %. Although the CO_2 problem had been handled, simulation showed that the new design resulted in unacceptable levels of H_2S in the outlet stream.

To combat the high H_2S levels in the overall outlet gas stream, the bypass stream was treated for H_2S using a static mixer arrangement. Figure 2 shows the static mixers, designed and built by Koch Engineering Company, Inc., before (picture from Koch sales literature) and after installation (field photo). The static mixers were designed so that the H_2S would be absorbed while allowing the CO_2 to slip through unreacted. The design called for an amine flow rate of 40 gpm to be injected into the static mixers. Since the existing regeneration unit was already at maximum capacity, further study was needed to determine if the system could handle the additional 40 gpm. Figure 3 is a PFD of the amine system with the static mixers.

Figure 2. Static Mixer Before and After Installation.

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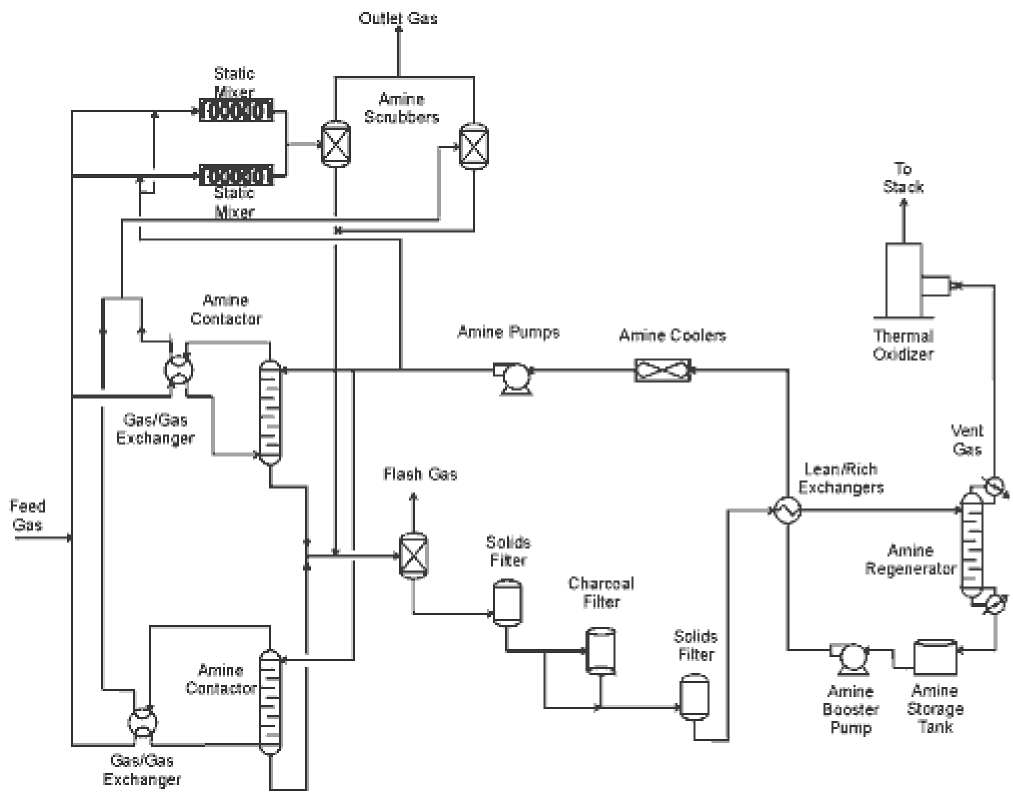

Figure 3. Anderson Plant After Modification.

POTENTIAL PROBLEMS

To determine the feasibility of the additional 40 gpm of circulation and the addition of a secondary amine (DEA), the entire regeneration system was thoroughly evaluated. A cumulative review and rating of the lean/rich exchangers and reboiler was performed. The analysis revealed that during normal operations (gas volume less than 180 MMSCFD) the lean/rich exchangers were only achieving 70% of their design capacity, based on design specifications and simulation results, and were already causing problems. Additional lean/rich exchangers were ordered and installed to help boost the capacity of the regeneration system. The additional lean/rich exchanger duty also allowed enough operating capacity to deal with the extra 40 gpm of amine and the addition of the DEA, which requires more heat to regenerate than MDEA. The reboiler was also rated and found to have sufficient area to accommodate the process changes. Insulation was added to the lean amine surge tank to help limit heat losses.

The stripper was reviewed for capacity limits with a secondary concern for potential corrosion problems. High CO_2 vapor phase concentrations in the reboiler and lower trays can lead to accelerated corrosion rates[1]. To reduce the CO_2 concentrations in the stripper, a specified minimum reflux rate of 8 gpm was established and maintained. Evaluation indicated that the stripper could handle the additional 40 gpm of amine; however, it is currently the limiting factor to any future upgrades because of flooding concerns.

STARTUP

Due to contractual commitments and limits on downtime for construction, the start-up procedure allowed for the steady addition of inlet gas into the plant. A plan was developed to permit step changes in the concentration of DEA and addition of gas flow through the static mixers. The changes were followed by simulations and analysis of field data to determine the next target amine concentration. Following this pattern, the maximum effective gas flow

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rate achieved by the plant was 210 MMSCFD. The specific steps used to make the changes are listed below. Table 1 shows the data collected during the testing period.

Flowrate Trial Procedure

1. Add DEA to MDEA solution and test for concentrations.
2. Check CO_2 and H_2S concentrations in the outlet gas.
3. Check for overall and individual amine concentrations.
4. Check loadings and run simulations for next step change.
5. Increase feed gas flowrate by 10 MMSCFD, bypassing all of it to the static mixers.
6. After the system moves to equilibrium, take additional data and perform simulations to determine the next DEA target value.
7. Repeat steps 1-6 until peak gas flow is achieved.

Note: The overall amine flow rate was kept constant throughout this procedure. 400 gpm of lean solvent was fed to each of the main contactors, and 20 gpm was fed to each of the static mixers.

Table 1. Flow Rate Test Trial Data

Sour Feed to Main Contactors	Sour Feed To Static Mixers	MDEA/DEA Conc.	Inlet CO_2 Conc.	Tower Outlet CO_2 Conc.	Overall Outlet CO_2 Conc.	Inlet H_2S Conc.	Overall Outlet H_2S Conc.
MMSCFD	MMSCFD	wt. %	mol %	mol %	mol %	ppm	ppm
180	0	50/0	6.0	2.7	2.7	25	1.0
180	10	37/14	6.0	2.4-2.5	2.7	25	1.8
180	20	23/20	6.0	2.3-2.4	2.6	25	1.6
180	30	20/25	6.0	2.2-2.3	2.6	25	2.0

CONCLUSIONS

Since the plant was started up after the retrofit, meeting the CO_2 and H_2S specifications was not a problem, even at the new design capacity. No increase in corrosion has been observed, nor any of the other typical amine unit problems such as foaming. Overall, the equipment changes allowed the plant to run at peak performance when necessary, and these changes were much less expensive than adding an entire amine train.

ACKNOWLEDGEMENTS

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REFERENCES

1. Gas Purification, Arthur Kohl and Richard Nielsen, pp. 195, 205, fifth edition, 1997.

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