Analysis of Various Flow Schemes for Sweetening with Amines

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ABSTRACT

There are many possible process variations for sweetening sour hydrocarbons with amines. Those to which we have given attention include the use of precontactors (static or jet eductor mixers), multiple absorber inlet nozzles, split flow units and pressure swing regeneration. Each of these variations is best suited to a certain set of operating conditions. Not all processes are appropriate for use with certain feed compositions or product requirements. This paper will discuss the application of the various flow scheme alternatives to a variety of different process conditions.

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INTRODUCTION

A standard amine unit is shown in Figure 1. The sour hydrocarbon is fed to the bottom of the absorber and is contacted by lean amine solution flowing downward through the column. The sweetened hydrocarbon exits the top of the absorber. Rich amine flows from the bottom of the absorber and typically enters a flash vessel where some of the dissolved hydrocarbons and acid gas are released. From there the rich amine enters the lean/rich exchanger where it is preheated before being fed to the regenerator. The regenerator produces acid gas overhead, and lean amine in the bottoms. The lean amine is cooled by the lean/rich exchanger, and continues on to the tank,booster pump, and lean amine cooler. Lean amine fed to the top of the absorber completes the circuit.

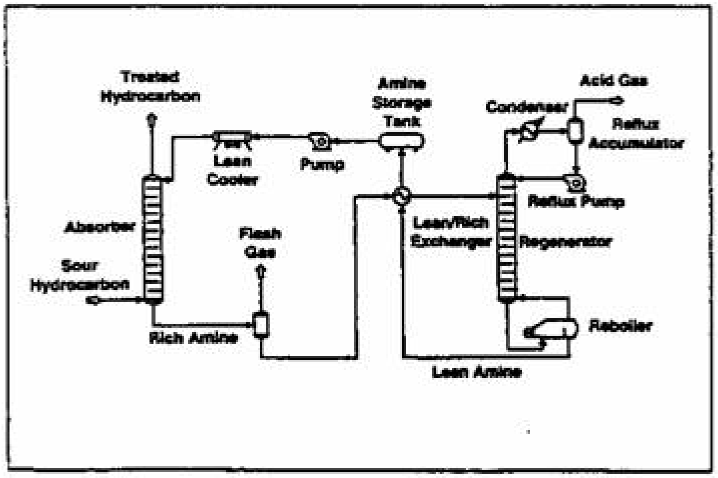

Figure 1. Standard amine sweetening plant

Variables which may be manipulated to meet gas or liquid hydrocarbon product specifications are the amine type and/or concentration and process configuration. Some examples of process configuration modifications are:

- Addition of a precontactor in the sour feed
- Use of multiple feed points in an absorber
- Split flow in which semi-lean amine is fed to some mid-point of the absorber while ultra-lean amine is fed at the top
- Pressure swing regeneration instead of rebelling

Table I briefly lists the applications, advantages and disadvantages of these four process configuration modifications.

Table I
Summary of process configuration modifications
Type of flow configuration

General purpose	Precontactor (static mixer)	Multiple Inlet nozzles	Spit flow	Pressure swing regeneration
	Bulk acid gas removal	Maximize CO_2 slip	Satisfy stringent H_2S specification	Bulk CO2 removal
Advantage	Increases plant capacity with low capital expenditure	Accommodates wide variations in feed acid gas	Lowers reboiler duty	Eliminates reboiler
Design or retrofit	Retrofit	Design or retrofit	Design or retrofit	Design
Amines used	Any	Tertiary	Any but usually primary or secondary	Tertiary
Disadvantages	Large pressure drop	Extra piping required	Extra equipment required	CO_2 pickup very limited

Special flow schemes for an amine sweetening unit may offer some advantages over the standard amine unit; however, for a design, the standard amine flow scheme should be considered first. Only when a simple flow scheme becomes limited should a more sophisticated flow scheme be pursued. For a retrofit situation, changing the amine type is usually the first option to consider before investigating a modified flow scheme.

PRECONTACTOR (JET EDUCTOR MIXER/STATIC MIXER)

The precontactor is typically a static mixer or jet eductor mixer for bulk acid gas removal and can be used to increase sour gas sweetening plant capacity. Static mixers are also commonly used in sour liquid hydrocarbon treating applications.[1] If a precontactor is effective in increasing plant capacity, it is almost certainly less costly than other methods of increasing plant capacity such as addition of an extra absorber, entire absorption train, or inlet gas compression. Any amine can be used depending on treated gas specifications, however, CO_2 pickup does not increase when tertiary amines are used due to kinetic effects. Addition of a static mixer may be thought of as the addition of an ideal stage (or portion of an ideal stage). The only potential problem associated with the use of static mixers would be associated with excessive pressure drop.[2]

Several circumstances could necessitate an increase in plant capacity.[2] One is increased sour gas flow due to an increase in the number of wells online. Another is a decrease in sour gas feed pressure resulting in increased gas volume. When the gas volume increases such that the absorber cannot handle the extra gas, the static mixer is placed in a bypass stream. In addition to an increase in sour gas volume, another difficulty which might be alleviated by the use of a precontactor would be an increase in acid gas content of the feed. Since the volume of gas does not increase, the precontactor may be placed in the feed to the absorber instead of in a bypass stream. A separator could be placed downstream of the static mixer to separate the rich amine. Alternatively, the bottom of the absorber could be used as a separator provided the gas/liquid feed into the bottom of the tower does not cause a problem with the inlet vapor distributor.[3] The static mixer could effect the bulk removal of a portion of the acid gas and allow meeting the sales gas specification without addition of another train or additional absorber. Isom and Rogers suggest as many as four alternative flow configurations for static mixers, some of which may not be beneficial.[3] Two other options investigated here are: static mixer placed in an absorber bypass stream, and static mixer placed in the absorber feed stream.

An actual case in which an anticipated increase in plant feed was compensated by the installation of static mixers in a bypass stream occurred at the Anderson Plant as described by Carter, et al.[4] The static mixers, used in combination with mixed amines, allowed a 17% increase in plant capacity. The plant feed contained about 6% CO_2 and 25 ppm H2S, with a sweet gas specification of 3% mol CO_2 and 4 ppm H_2S. The capacity of the plant was 180 MMscfd, which was expected to increase to 210 MMscfd. Since the gas could not be fed to the absorbers due to capacity limitations, the gas was bypassed around the absorbers with the static mixers placed in the bypass stream. Figure 2 shows a simplified flow configuration of the absorption section of the Anderson plant after installation of the static mixers. Although the static mixers did not cause an appreciable pickup of CO_2, the H_2S content was kept to within the specified 4 ppm by selective absorption. The Anderson plant originally circulated 50% wt MDEA. The addition of DEA after the static mixers were installed in the absorber bypass streams helped the CO_2 content of the treated gas stay below 3% mol.

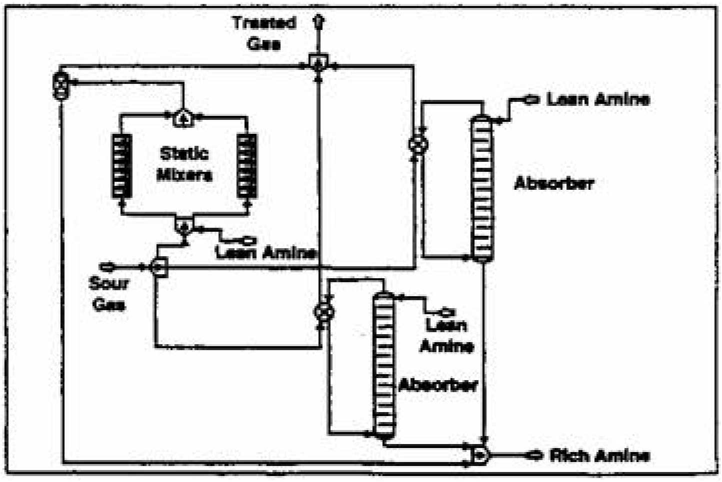

Figure 2. Simplified absorption section Anderson Gas Plant

To illustrate the effectiveness of the precontactor, the following example considers an MEA sweetening case in which the acid gas content of the feed gas increases. Case 1 uses the standard amine unit with a static mixer in the absorber feed, as shown in Figure 3. In this plant it is assumed that the absorber is at flood limit and cannot handle additional amine flow. The stripper however, can handle up to 10% additional amine flow. The conditions are given in Table II. Although the acid gas content, temperature, pressure and MEA concentration are taken from the Wildcat Hills gas plant described by Gregory, the inlet flow, circulation rate and reboiler steam rate were assumed since this information was not included in the data.[2] The static mixer in this example is assumed to represent 1 ideal stage.

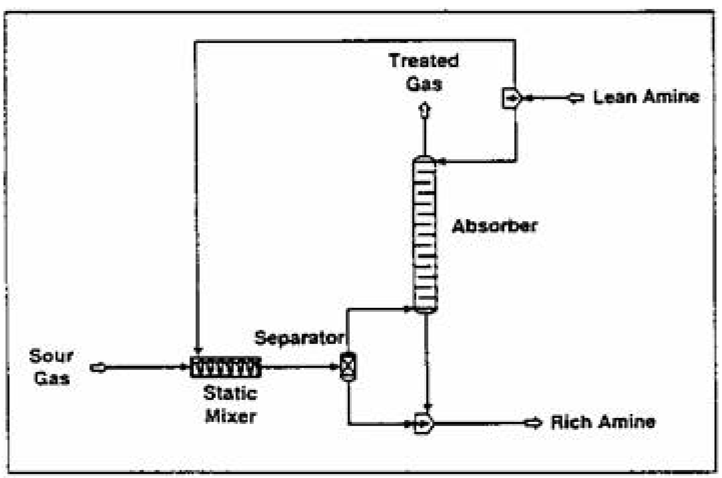

Figure 3. Case 1 - Static mixer in absorber feed stream.

Table II
Case 1: Example plant assumed operating conditions

Amine circulation rate	1250 gpm
Amine concentration	20% wt MEA
Sour feed flow rate	50 MMscfd
Sour feed temperature	100F
Sour feed pressure	85 psia
Sour feed CO2 content	7% mol
Sour feed H2S content	13% mol
Sweet gas CO2	0.105% mol

Sweet gas H2S	5.01 ppm
Reboiler steam rate	100/MMbtu/hr

If the acid gas in the feed were to increase to 16% H_2S, Figure 4 shows the CO_2 and H_2S content of the treated gas as a function of amine fed to the static mixer. This plot was generated using the process simulator TSWEET® to determine the amount of H_2S and CO_2 in the treated gas at varying amine circulation rates to the static mixer. The maximum for this example is 125 gpm (10% of 1250 gpm), and the plot shows that circulation rates higher than 125 gpm are ineffective regardless. In an actual plant, it would have to be determined whether or not the extra amine flow could be handled by the regenerator.

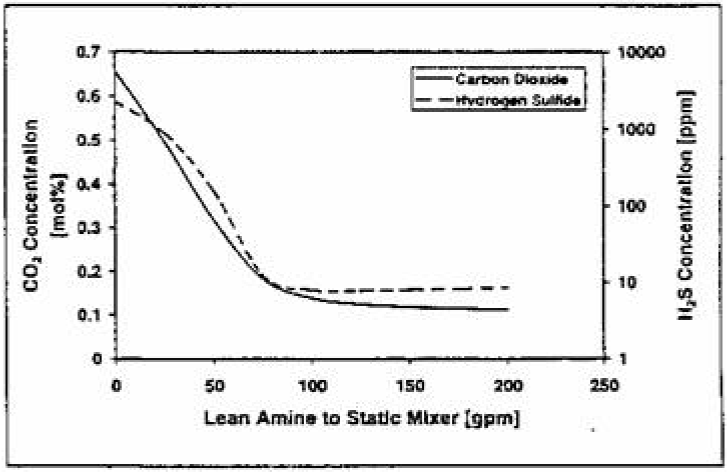

Figure 4. Case 1 - Effect of lean amine fed to static mixer on H_2s and CO_2 in treated gas.

MULTIPLE INLET NOZZLES

Use of multiple absorber inlet nozzles is applicable when maximum CO_2 slip is desired. Multiple lean amine feed locations are particularly useful if the sour feed flow rate varies and has a high CO_2 content By feeding lean amine to the absorber at different locations, the number of ideal stages may be varied. Fewer ideal stages allow more $C0_2$ rejection, and possibly increased H_2S absorption. Figure 5 shows an absorber with multiple amine feed locations.

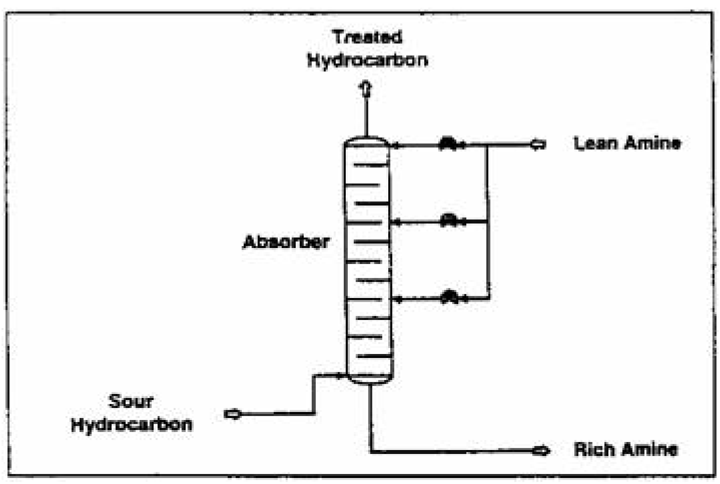

Figure 5. Absorber with multiple lean amine inlet.

The first example in this section, Case 2, illustrates the effect of decreasing the number of ideal stages by varying the single lean amine feed point to the absorber. This example uses a 3 ft diameter, 20 tray (7 ideal stage) absorber. The conditions are similar to those of the Signalta Plant as described by Mackenzie et al.[5] It is desired to maximize CO_2 rejection in the absorber to improve feed to the Claus Unit.

By adjusting the number of ideal stages, the CO_2 rejection can be maximized while still meeting the H_2S specification. If the $C0_2$ in the feed decreases, fewer ideal stages may be used, which allows more $C0_2$ slip in the absorber. Thus the CO_2 in the acid gas feed to the Claus unit is reduced. After validating TSWEET predictions, the conditions in the last column of Table III are used to generate the results listed in Table IV. For this particular example, 4 ideal stages (feed on stage 10) is the optimum number to reject the most possible $C0_2$ while still meeting the 4 ppm H_2S specification. Figures 6 and 7 are graphical representations of the effects of varying the number of ideal stages in the absorber.

Table III
Case 2: Data comparison and plant operating conditions

	Plant data	TSWEET predictions	Conditions used in example
Amine circulation rate	70 gpm (3 feed points)	70 gpm (3 feed points)	70 gpm (single feed point)
Amine concentration	32.3% mol MDEA	32.3% mol MDEA	35% mol MDEA
Sour feed flow rate	14.16 MMscfd	14.16 MMscfd	15 MMscfd
Sour feed temperature	60°F	60°F	80 °F
Sour feed pressure	390 psia	390 psia	400 psia
Sour feed CO_2 content	3.02% mol	3.02% mol	3% mol
Sour feed H_2S content	0.32% mol	0.32% mol	0.3% mol
Sweet gas CO_2 content	2. 13% mol	2.59% mol	—
Sweet gas H_2S content	3.2 ppm	3.6 ppm	< 4 ppm (spec.)
Reboiler steam rate	1.56 Ibfeal	1.561b/gal	1.51b/gal

Table IV
Case 2 - Effect of varying the number of ideal stages.

# Ideal stages	Approximate equivalent feed tray	H_2S Overhead (ppm)	CO_2 Overhead (%)	% $C0_2$ in claus feed (dry basis)
7	1	2.7	1.88	77.1
6	4	3.1	2.04	74.3
5	7	3.5	2.20	70.7
4	10	3.9	2.35	66.1
3	13	4.4	2.50	60.0
2	16	7.0	2.65	51.5

1	19	96.4	2.78	41.2

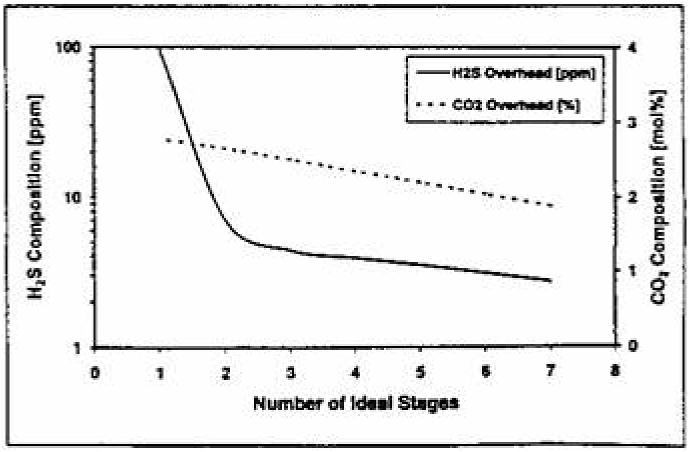

Figure 6. Case 2 – Effect of varying the number of ideal stages

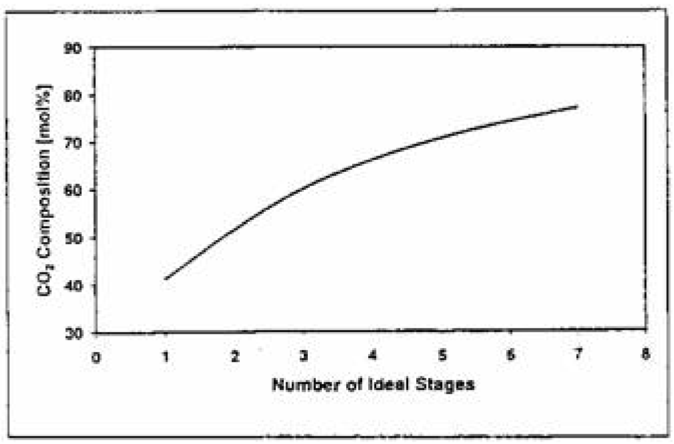

Figure 7. Case 2 - Improvement of Ciaus feed by varying the number of ideal stages in the absorber.

In some cases, the amount of H_2S in the treated gas may actually be reduced by decreasing the number of ideal stages and rejecting more CO_2.[6] In the following example (Case 3), the number of ideal stages was varied for a 20 tray contactor. The composition and conditions are from the Hungarian Gas Plant design conditions for the high pressure absorber described by Law.[6]

Table V
Case 3 - Plant operating conditions

	Plant data	TSWEET predictions	Conditions used in example
Amine circulation rate	—	30 gpm	30 gpm
Amine concentration	50 wt % MDEA	50% mol MDEA	50% mol MDEA
Sour feed flow rate	—	10 MMscfd	10 MMscfd
Sour feed temperature	95°F	95°F	80°F

Sour feed pressure	1289 psia	1289 psia	1300 psia
Sour feed CO_2 content	11.0% mol	11.0% mol	9.47% mol
Sour feed H_2S content	82.4 ppm	82.4 ppm	0.03% mol
Sweet gas CO_2 content	8.4% mol	8.4% mol	No Spec.
Sweet gas H_2S specification	3.3 ppm	1.5 ppm	<4ppm
Reboiler steam rate,	—	0.83 Ih/gal	1.0 Ib/gal

After comparing TSWEET predictions to the plant data, the number of ideal stages in the column was reduced to obtain the following results, also shown in Figure 8.

Table VI
Case 3 - Effect of varying the number of ideal stages

# Ideal stages	H_2S overhead (ppm)	CO_2 overhead %
7	10.5	5.9
6	7.2	6.0
5	2.2	6.2
4	0.8	7.2
3	0.7	8.1
2	0.8	8.6
1	5.3	9.0

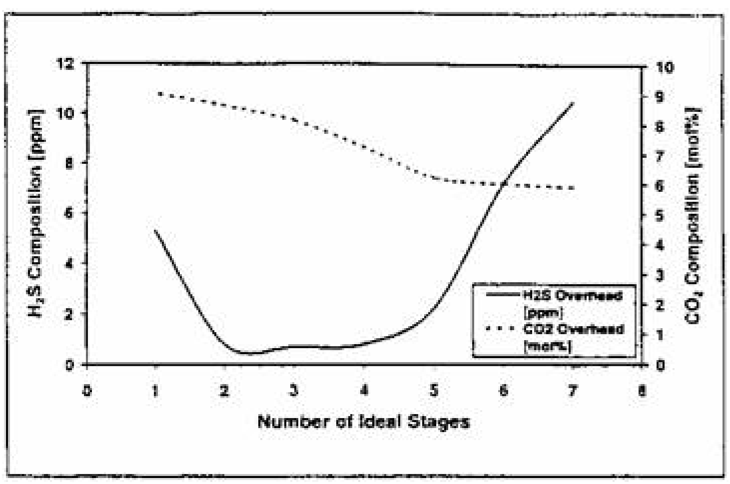

Figure 8. Case 3 - Effect of varying the number of ideal stages.

In this example, the H_2S actually decreases with decreasing number of ideal stages. One explanation forthis phenomenon is that when less CO_2 is absorbed, more amine is available to absorb H_2S. Another explanation offered by Law is that with less CO_2 pickup, the column internal temperature is lower, which favors H_2S pickup.[6] Most likely a combination of these two effects is causing the decrease in H_2S overhead concentration.

It should be noted that simply feeding lean amine to two or more locations simultaneously as described by Mackenzie etal. in the Signalta Plant may not show significant benefit, as shown for Case 4 in Table VII.[5] This table was generated using TSWEET and conditions similar to the Signalta Plant as shown in Table III. Absorber feeds are on trays 1 and 12 (assumed to be ideal stage 4).

Table VII
Case 4 - Effect of feeding lean amine to two absorber inlet locations

%to side feed	H_2S Overhead (ppm)	$C0_2$ Overhead (%)
0	2.7	1.88
10	3.0	1.92
20	3.2	1.95
30	3.6	1.99
40	4.1	2.04
50	4.7	2.08
60	5.5	2.13
70	6.6	2.17
80	8.9	2.22

The $C0_2$ slip improves only slightly because the tray residence time at the top is increased as a result of lower liquid flow on those trays. Thus, more CO_2 is absorbed due tokinetic effects. Another consideration is that at very low flow rates below design, the trays might not be operating properly with respect to hydraulics, and thus would be less efficient.

More benefit could be derived from two simultaneous feeds if the section above the middle feed were a smaller diameter than the section below.[7,8] In this case, the tray residence time above the feed would be shortened, allowing more $C0_2$ slip and thus more H_2S pickup in that section. The capital cost of the tower would be somewhat less, but at the expense of additional piping and a less flexible column. Lower column weir height above the feed location would also shorten the residence time, increasing CO_2 slip.

SPLIT FLOW UNIT

A split flow process is used when the H_2S specification is stringent. The traditional split flow configuration is generally used only with primary and secondary amines. In the split flow unit, a portion of semi-lean amine is drawn from the regenerator, cooled, and fed to the contactor at some mid-column feed point, as shown in Figure 9 for the Okotoks Plant.[9] The amine flow to the regenerator reboiler is lower because of the side draw, and a given amount of steam can strip the amine more thoroughly, resulting in an ultra-lean amine. In this way, the semi-lean amine picks up the bulk of the acid gas, leaving the ultra-lean amine fed to the top of the contactor to “polish" the gas to a level that could not be achieved by ordinary lean amine stripped at the same reboiler duty.[8] A split flow configuration is most advantageous in cases where the H_2S sweet gas specification is very low and a very lean amine is required to achieve the specification. It should also be noted that the treated gas specification can be achieved at lower reboiler duty, but at the expense of a more complicated plant. The stripper must be taller and more complex in addition to requiring an extra pump, heat exchanger, cooler and piping, not to mention the associated controls.[7,8]

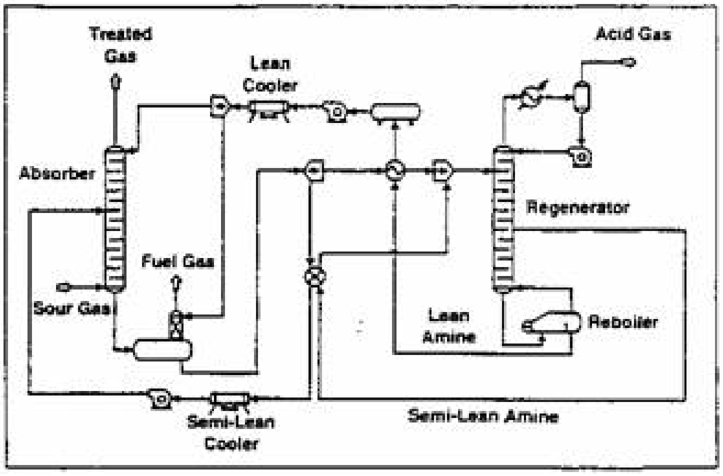

Figure 9. Case 5 - Okotoks plant.

The Okotoks plant described by Estep et al. (Figure 9), was an actual operating split flow MEA unit.[9] Table VIII lists the plant operating conditions for Case 5, and Table IX shows a comparison between the data and the process simulator TSWEET.

Table VIII
Case 5 - Okotoks plant operating conditions

Lean amine circulation rate	390 gpm
Semi-lean amine circulation rate	1620 gpm
Lean amine concentration	2 1.5% wt MEA
Semi-lean amine concentration	15% wt MEA
Sour feed flow rate	29.46 MMscfd
Sour feed temperature	58T
Sour feed pressure	565 psia
Sour feed CO_2 content	10.4% mol
Sour feed H_2S content	33.4% mol
Sweet gas H2S specification	0.05 gi^lOO cf

Table IX
Case 5 - Comparison between Okotoks plant data and process simulator

	Plant data	TSWEET
$C0_2$ out, ppm	—	27
H_2S out, giVlOO scf	0.01	0.009

Rich loading, mol/mol	0.65	0.68
Reboiler steam rate, lb/hr	113,600	113,583
Acid gas flow, MMscfd	13.80	13.57

After matching plant data, TSWEET was used to find the optimum amount of semi-lean amine for this plant. Figure 10 shows a plot of semi-lean flow vs. H_2S in gr/100 scf. For this plant the specification was 0.05 gr/100 scf, which could have been achieved by splitting only about 600 gpm as semi-lean. Because the plant was designed for the 1600 gpm split, an increase in the acid gas flow rate to the absorber or the H_2S concentration should not render the plant unable to meet the H_2S specification. Figure 11 shows the reboiler duty required to achieve the 0.05 gr/100 scf H_2S specification at increasing amounts of semi-lean for the same set of conditions. The split flow plant produces much cleaner gas for each lb of steam consumed, as stated previously, but at the expense of a more complicated plant.

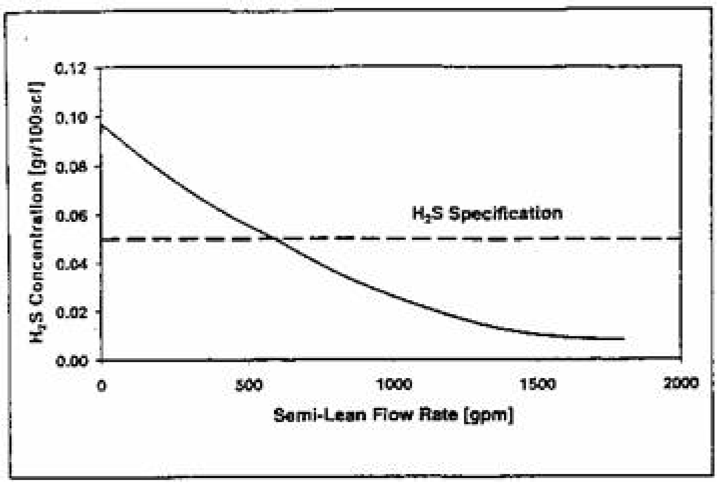

Figure 10. Case 5 - Split flow plot with 2000 gpm total circulation rate.

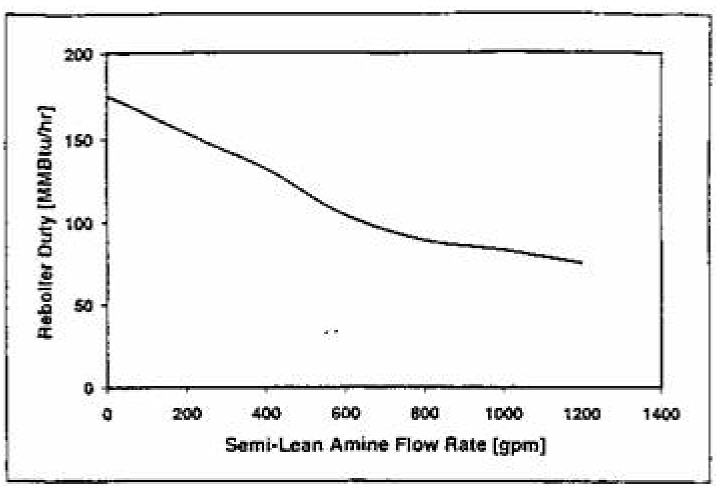

Figure 11. Case 5 - Energy required to meet specifications.

An even greater efficiency and energy savings can be achieved by modifying the conventional split flow unit by using recycles as described by Towler, et al.[10] In this modified flow scheme, shown in Figure 12, the condensate from the regenerator condenser is fed to and withdrawn from the regenerator above the rich feed, and subsequently fed to the reboiler to increase stripping efficiency. The semi-lean amine is reboiled to maintain the amine concentration. An even more extreme modification proposed by Towler et al. and shown in Figure 13.[10] This flow scheme employs a separate stripping column in which the semi-lean amine is partially stripped, with 20% of the bottoms fed to a second stripping column to be stripped to ultra-lean status. The stripped amine is fed to the middle of the absorber. This flow scheme is known as the double-loop absorber-stripper process.

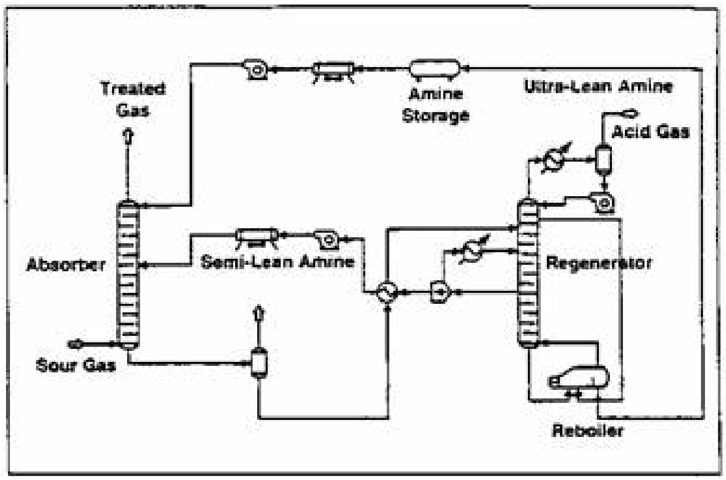

Figure 12. Thermodynamically efficient regeneration system.

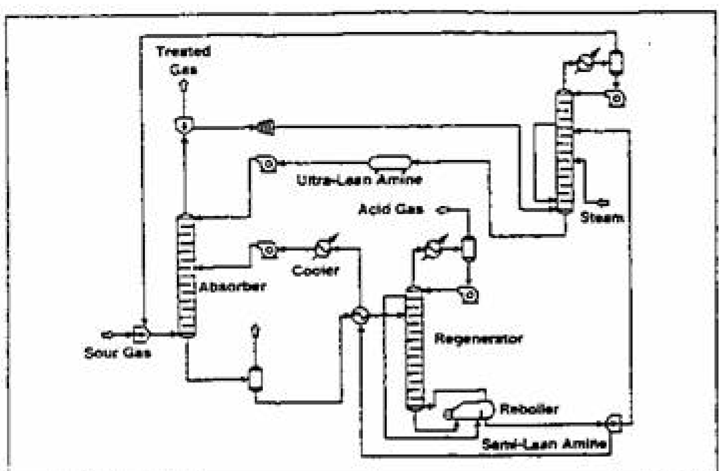

Figure 13. Double-loop absorber-stripper process.

A final alternative is split flow for removal of CO_2 from high pressure gas with promoted MDEA.[7] This process, shown in Figure 14, is a combination of split flow and pressure swing regeneration. (Pressure swing regeneration is discussed in the following section.)

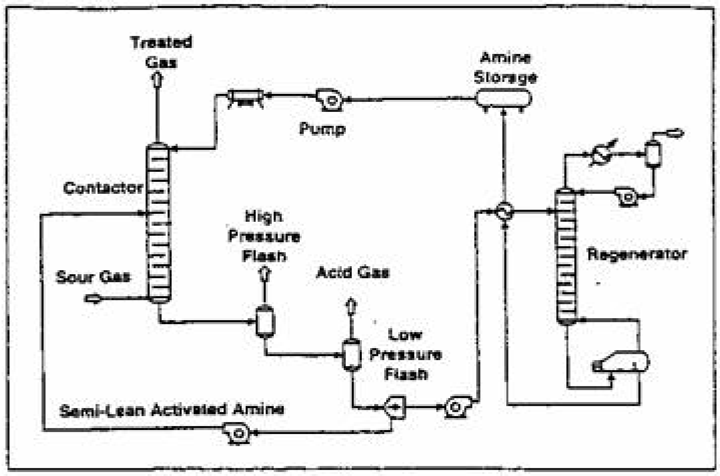

Figure 14. Activated MDEA split flow configuration.

To summarize, split flow may be used when other alternatives have been exhausted due to not meeting ^as specifications. Usually this occurs when the H_2S to CO_2 ratio is high and/or the H_2S specification is very stringent. In cases with H_2S only, it is possible that no amine or mixture of amines will allow the spec to be achieved, however, with a split flow setup, the H_2S specification is easily achieved.

PRESSURE SWING REGENERATION

A pressure swing type of regeneration is used only with tertiary amines (eg. MDEA or TEA) for the absorption of $C0_2$ only. The pressure swing uses the reduction of pressure to "regenerate" the amine. At lower pressures the absorbed $C0_2$ is released. For example, if the absorber pressure is 800 psia, a preliminary "high pressure" flash might be 100 psia, and the low pressure flash, which releases most of the CO^ might be 32 psia. The rich and lean loadings in a pressure swing regeneration plant are relatively high. Because the rich amine is not fully regenerated, the $C0_2$ pickup is relatively low. Pressure swing regeneration can only be used if bulk removal of only a portion of the CO_2 is acceptable. For this reason pressure swing regeneration for use with amines is not very common. Pressure swing regeneration is morecommon with physical solvent regeneration. The main benefit to this method is that no heat duty would be required for regeneration. Figure 15 shows a simplified pressure swing regeneration flow scheme.

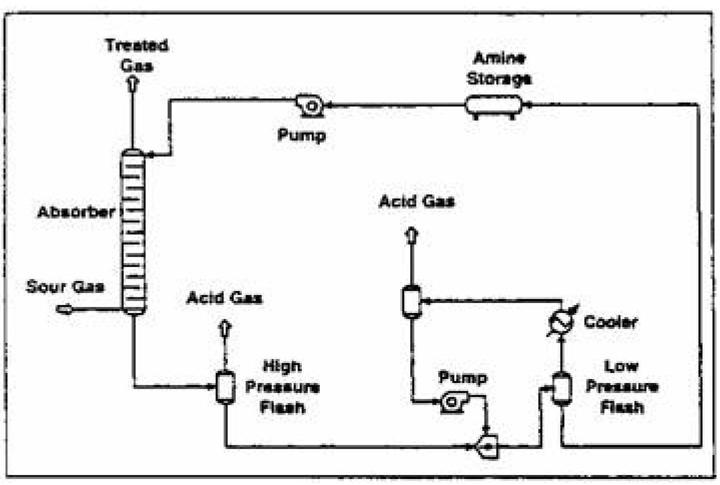

Figure 15. Simplified pressure swing regeneration.

If increased $C0_2$ pickup is desired, an alternative would be to heat the low pressure flash to drive off more CO_2.

TSWEET was used to model an example plant where the low pressure flash is heated to temperatures ranging from 140-240°F. The operating conditions for Case 6 are listed in the following table.

Table X
Case 6 - Pressure swing regeneration operating conditions

High pressure flash	100 psia
Low pressure flash	32 psia
Lean amine concentration	35% mol MDEA
Sour feed flow rate	2 MMscfd
Sour feed temperature	72 °F
Sour feed pressure	813 psia
Sour feed CO_2 content	50% mol

Figure 16 shows the decrease in $C0_2$ in the sweet gas as the low pressure flash temperature increases. The duty increases with increasing low pressure flash temperature, as expected, Figure 17.

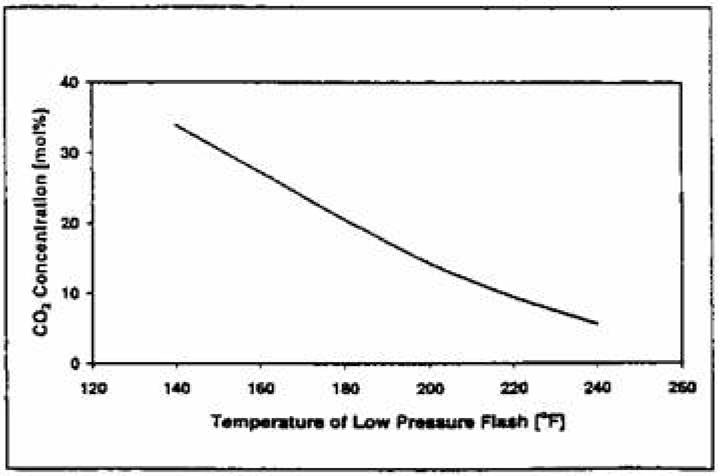

Figure 16. Case 6 - Effect of low pressure flash temperature on treated gas CO_2 content at 35 % wt MDEA.

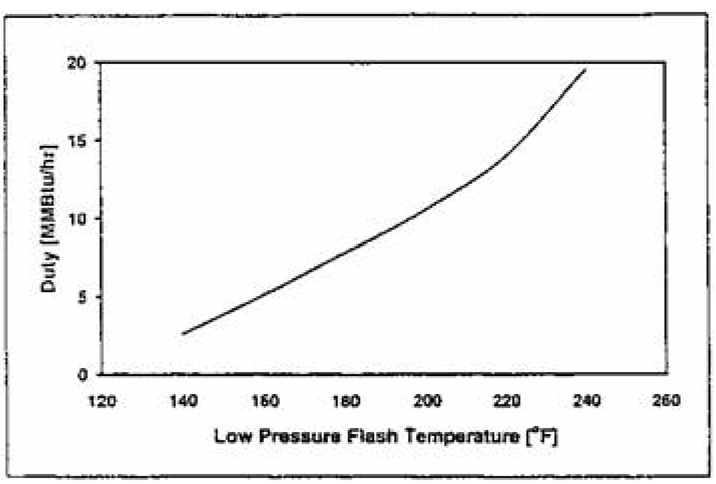

Figure17. Case 6 – Heat duty for 35% wt MDEA at varying low pressure flash temperatures.g low pressure flash temperatures.

When considering using a pressure swing or pressure swing with heated flash for regeneration, the $C0_2$ specification must be the governing factor. For the previous example feed and amine circulation, Table XI shows a comparison between pressure swing with no heating, pressure swing with low pressure flash heated to 180°F and a typical amine regenerator with reboiler and condenser. The cases with the heated flash and regenerator use a lean amine trim cooler in addition to the low pressure flash cooler.

Table XI
Case 7 -comparison between regeneration options pressure swing

	Pressure swing (no heating)	Heated low pressure flash (180°F)	Standard amine regenerator
$C0_2$ in treated gas, mol %	30.3	20.5	3.5
Heating duty, MMBtu/hr	0	7.8	13.4
Cooling duty, MMBtu/hr	0.08	7.8	13.2
Lean loading, mol/mol	0.654	0.384	0.006
Rich loading, mol/mol	0.831	0.611	0.297

SUMMARY

Modifications of the standard amine treating flowsheet include precontacting, multiple absorber inlet nozzles, split flow and pressure swing regeneration. The modifications to consider are primarily dependent on the objective of the process and circumstances resulting from process operation changes. Precontacting is useful for increasing plant capacity while multiple absorber amine feed nozzles are applicable for maximizing CO_2 slip. Split flow achieves very stringent H_2S specifications, and pressure swing regeneration accomplishes bulk $C0_2$ removal at little cost. These modifications should be explored in addition to other methods such as amine concentration, use of different amines, or mixtures of amines. Since these considerations for design or retrofit must be evaluated on a case by case basis, tools such as process simulation programs assist in the evaluation of the effectiveness of the modifications. In many cases, these and other plant modifications could have significant economic benefits.

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