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AMINE PROCESS CALCULATIONS

Even though the removal of acid gas constituents by ethanolamine solutions is in wide-spread use surprisingly little information of a quantitative nature is available for use in process/design calculations. The reaction rate data necessary for design of the system using a reaction rate approach are so few and scattered as to be practically useless. Though equilibrium data are available for some systems, in most cases they are limited in scope and probably would not cover optimum concentration and temperature ranges for the particular process unit of interest. However, the equilibrium data provide the best avenue currently available for estimating the various parameters for an ethanolamine treating system. Techniques for using these equilibrium data will be discussed in the pages that follow.

Shortcut Calculations

Rapid estimates of order of magnitude values for the primary parameters in sweetening units can be obtained quickly. Sisson (4.1) has presented nomographs that can be used to quickly estimate the MEA circulation rate required for removal of CO_2 from a gas stream. The MEA solution concentration, the allowable loading at the bottom of the contactor (moles CO_2/mole MEA), the temperature of the MEA solution leaving the bottom of the contactor and the volume and concentration of the gas to be processed must be fixed. With this information the density of the flowing solution can be determined from Figure 4.1. Using this information and the concentration of the MEA solution in weight percent the point on pivot line #1 of Figure 4.2 can be fixed. From this point a straight line is drawn through the solution loading to pivot line #2. From the point on pivot line #2 a straight line is drawn to the volume of gas to determine the point on pivot line #3. A straight line from the gas composition through pivot point #3 extended to the right hand axis will determine the circulation rate in gallons per minute.

Example Problem

Consider the case of 20 MMscfd of gas containing 15 mole % CO_2 that is to be sweetened using 19 weight % MEA solution. The MEA solution is estimated to leave the bottom of the contactor at 150°F. What is the required MEA circulation rate?

On Figure 4.1 connect the 150°F. temperature and the 19 weight % MEA solution with a straight line. The density of the MEA solution is determined to be 8.2175 lbs per gallon..

Connecting 8.2175 lbs per gallon with 19 weight % MEA on Figure 4.2 and then following stepwise the procedure for using Figure 4.2 yields an MEA circulation rate requirement of 477 GPM.
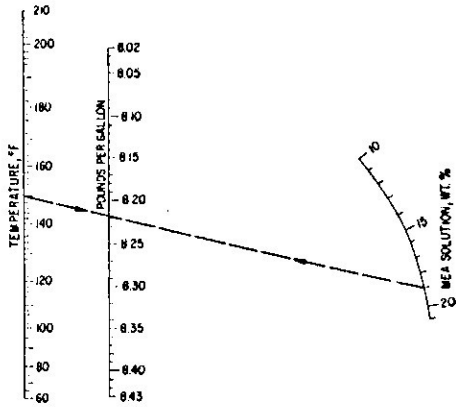
Figure 4.1
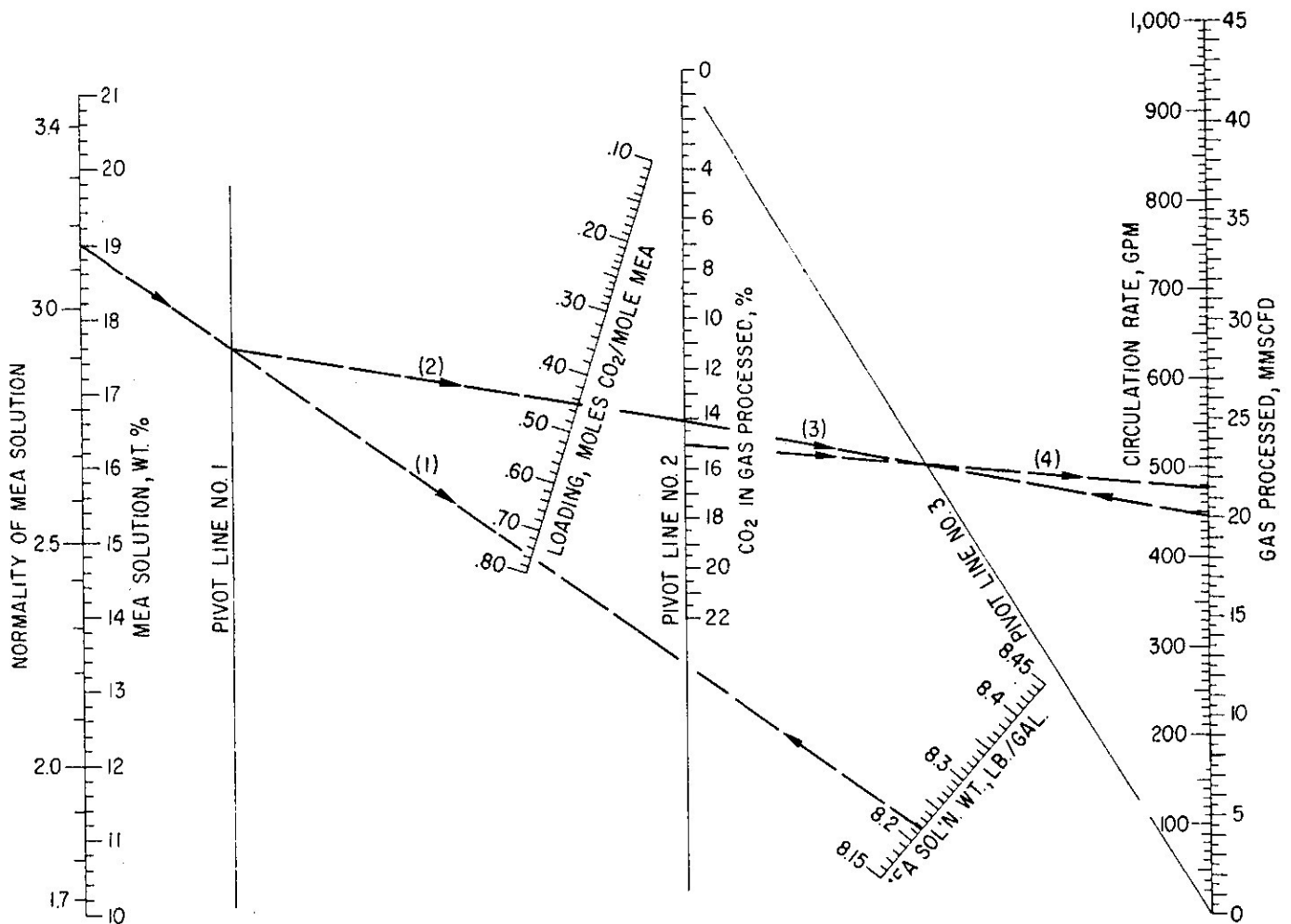Density of MEA solutions as a function of temperature and concentration

Figure 4.2   Nomograph for estimating MEA circulation rate

Jones (4.2) presented a simplified procedure for estimating circulation rates for MEA and DEA when both H_2S and CO_2 are present. The equations suggested by Jones are:

$$\text{For MEA:} \quad \text{GPM} = 41.0 \times QX/Z \tag{4.1}$$

$$\text{For DEA:} \quad \text{GPM} = 45.0 \times QX/Z \quad \text{(conventional)} \tag{4.2}$$

$$\text{GPM} = 32.0 \times QX/Z \quad \text{(high-loading)} \tag{4.3}$$

where Q = Gas to be processed, MMscfd

$\quad$ X = Acid gas content, volume percent

$\quad\quad = \text{Mol } \% \ CO_2 + \dfrac{\text{Grains } H_2S}{632}$

$\quad$ Z = Amine concentration, wt. %

The circulation rates, according to Jones, are based on loadings of 0.33 mole acid gas/mole MEA, 0.5 mole acid gas per mole DEA for conventional plants and 0.7 mole acid gas/mole DEA for high loading DEA plants. Figure 4.3 shows a graphical solution of Equations 4.1 and 4.3.

Jones suggests using data from Table 4.1 to estimate heat exchange requirements and Table 4.2 to estimate top horse power requirements. The contactor size can be estimated from Figure 4.4 and the other vessels in the plant from Table 4.3.

Figure 4.5 shows a graph taken from Jones' paper which provides an estimate of the investment required for ethanolamine sweetening units. The investment in Figure 4.5 assumes skid mounting for the separators, reflux accumulators, heat exchangers, etc., and does not include transportation from fabrication shop to the plant site or foundations and piping required at the location.

Table 4.4 shows a breakdown of the annual costs for an ethanolamine treating unit. Table 4.4 clearly shows that sweetening of natural gas is an expensive proposition.

TABLE 4.1

HEAT EXCHANGE REQUIREMENTS FOR AMINE PLANTS

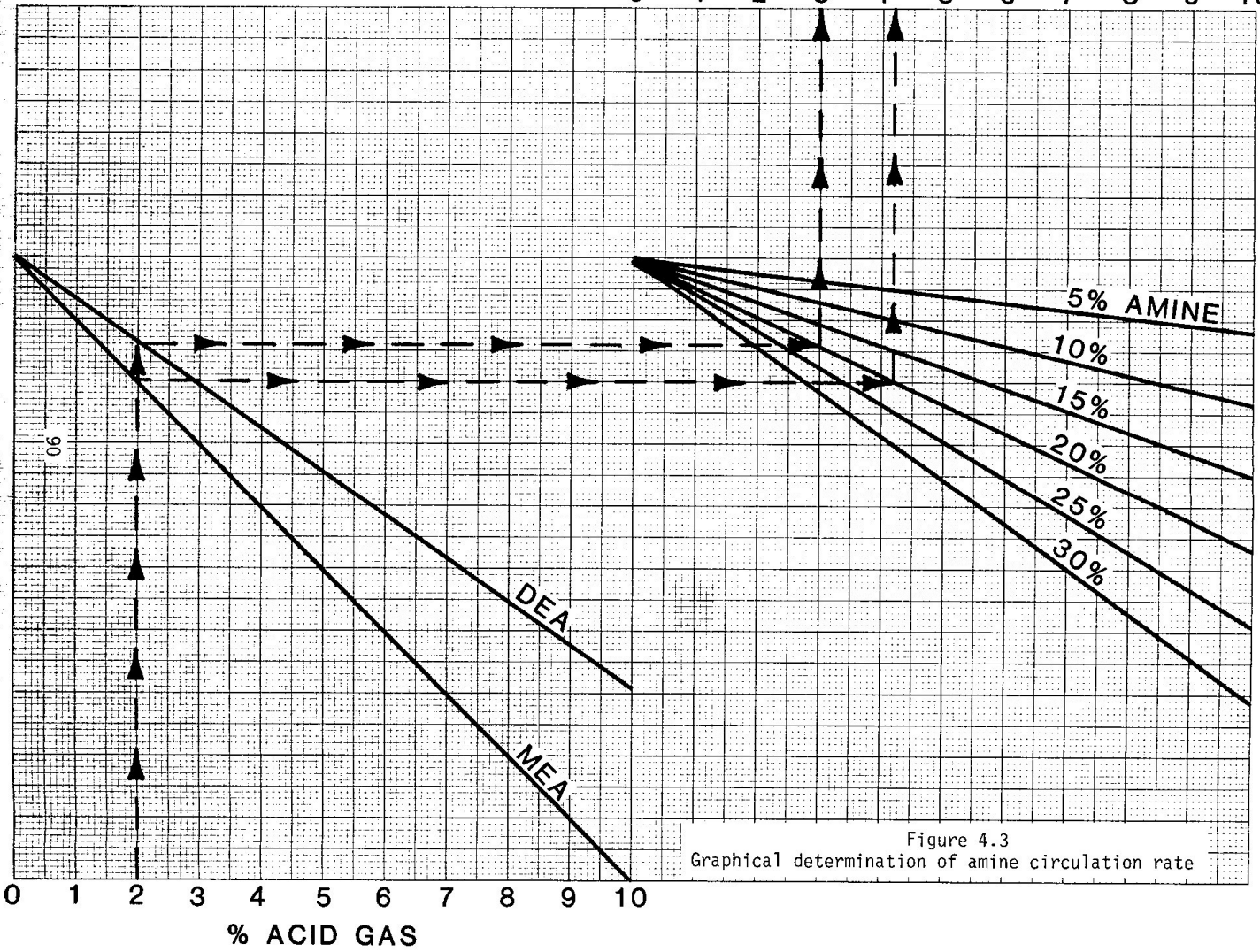|	Duty-BTU/Hr.	Area-Sq. Ft.
Reboiler (Direct Fired)	72,000 x GPM	11.30 x GPM
Solution Exchangers	45,000 x GPM	11.25 x GPM
Solution Coolers (Air Cooled)	15,000 x GPM	10.20 x GPM
Reflux Condenser (Air Cooled)	30,000 x GPM	5.20 x GPM

GPM CIRCULATION REQUIRED PER MMCFD

5% AMINE
10%
15%
20%
25%
30%

DEA

MEA

% ACID GAS

Figure 4.3
Graphical determination of amine circulation rate

TABLE 4.2
Pump Horsepower Requirements For Amine Plants

Solution Pumps	GPM x PSIG x 0.00065 = HP
Booster Pumps	GPM x 0.06 = HP
Reflux Pumps	GPM x 0.36 = HP
Aerial Cooler	GPM x 0.36 = HP

TABLE 4.3
Regeneration Vessel Sizes
(Inches)

Solution Circl. Rate GPM	Still Diameter	Surge Tank Diam.	Length	Reflux Accum. Diam.	Length	Flash Tank Diam.	Length	Carbon Filter Diam.	Length
10	16	24	72	16	36	24	72	16	84
25	24	42	96	24	48	42	96	24	84
50	30	48	144	30	96	48	144	36	96
100	42	60	192	42	96	60	192	48	96
200	60	84	288	60	96	84	288	60	96
300	72	84	384	72	96	84	384	72	96
400	84	96	384	84	96	96	384	84	96

TABLE 4.4
Amine Unit Operating Expenses

Basis: Treating plant with 200 Gpm circulation; treating 20 MMcfd gas with 8% acid gas: January, 1976, cost index

	¢/Mcf	$/Year
Operating Labor (5 @ 15,000)	1.03	75,000
Supervision (1 @ 18,000)	0.25	18,000
Employee Benefits @ 35% Payroll.	0.44	32,550
Utilities	0.74	54,000
Chemicals & Supplies	0.33	24,000
Repair Materials & Labor @ 3% Investment	0.41	30,000
Direct Overhead @ 5% Investment.	0.68	50,000
Corporate Overhead @ 3.5% Investment . .	0.48	35,000
Depreciation (10 Year Straight Line) . .	1.37	100,000
Interest (@ 12%, first year)	1.64	120,000
Insurance & Taxes @ 2% Investment. . . .	0.27	20,000
	7.64	558,550
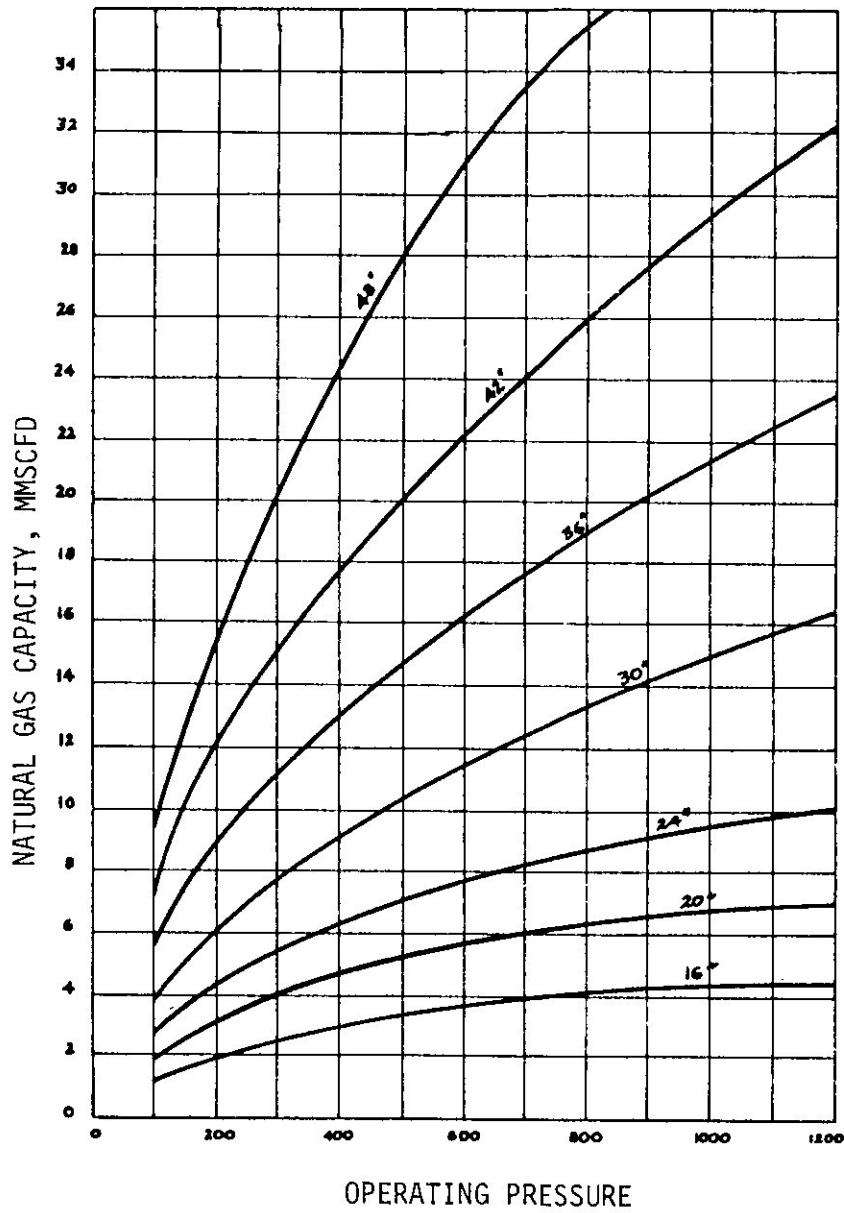
Figure 4.4   Amine Contactor Size Required

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TABLE 4.5
Effect Of Hydrogen Sulfide and Carbon Dioxide Loading On Equilibrium
Partial Pressures Over A 15% Solution Of MEA

(Calculated by SGP Program)

$\dfrac{\text{Mol } H_2S}{\text{Mol MEA}}$	$\dfrac{\text{Mol } CO_2}{\text{Mol MEA}}$	Part. Press. H_2S, mm Hg	Part. Press. CO_2, mm Hg
0.1	0.000	0.382	0.000
0.1	0.001	0.387	0.000
0.1	0.01	0.431	0.001
0.1	0.05	0.648	0.006
0.1	0.10	0.985	0.020
0.05	0.10	0.348	0.013
0.01	0.10	0.049	0.009
0.001	0.10	0.004	0.008
0.000	0.10	0.000	0.008

Figure 4.5
Amine Plant Cost (1977 dollars)
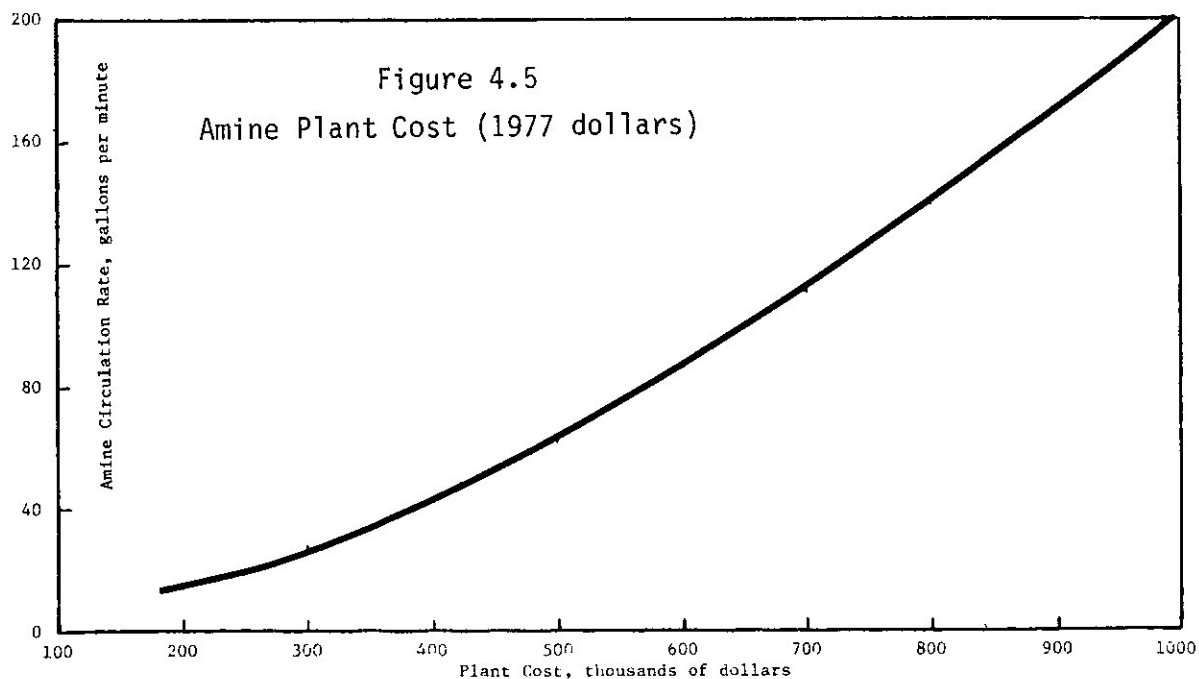
MONOETHANOLAMINE

The first short cut method discussed above is restricted in application to those gases containing only carbon dioxide. The second procedure treats hydrogen sulfide and carbon dioxide as lumped "acid gas". In the general case both hydrogen sulfide and carbon dioxide will be present in the gas stream. As the data presented in Table 4.5 clearly show, there is an interaction between H_2S and CO_2 reacted with monoethanolamine. The data shown in Table 4.5 are typical of the behavior displayed by MEA, DEA, DGA, and DIPA, in the presence of gases containing both H_2S and CO_2. The presence of even what could be considered trace amount of one of the acid gas constituents can cause significant differences in equilibrium solution loading and/or equilibrium partial pressures for the other acid gas constituent. This is displayed also in Fig. 4.6 through 4.13 for MEA.

Jones et al., (4.3) made a systematic investigation of the equilibrium concentrations of gas and liquid for the H_2S — CO_2 — MEA system. The concentration of the MEA was 15.3% by weight. They smoothed the experimental data and presented it in tabular form. Using the data in that form required much interpolation and cross plotting. McCoy and Maddox (4.4) plotted the Jones et al. data in nomograph form. The McCoy and Maddox charts are shown in Fig. 4.6 to 4.13. Interestingly, one of the primary variables in determining MEA — acid gas equilibrium is the ratio of H_2S to CO_2 concentrations in the sour gas.

Figure 4.6   Equilibrium data for H_2S and 15.3% MEA at 40°C with CO_2 present

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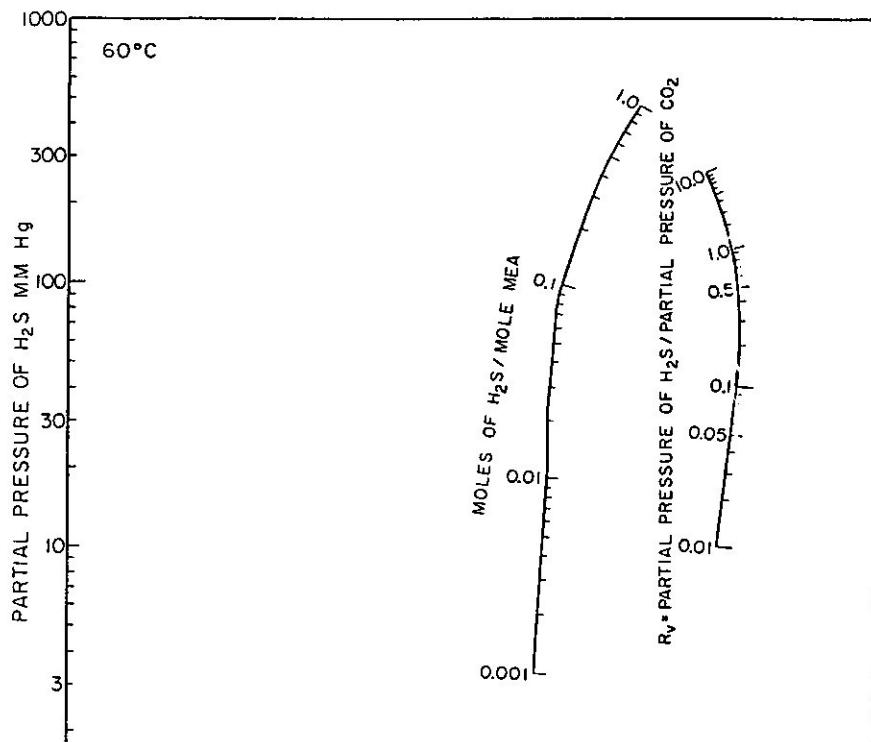
Figure 4.7 Equilibrium data for H_2S and 15.3% MEA at 60°C with CO_2 present. (Ref. 4.9 Courtesy Chemical Rubber Co.)
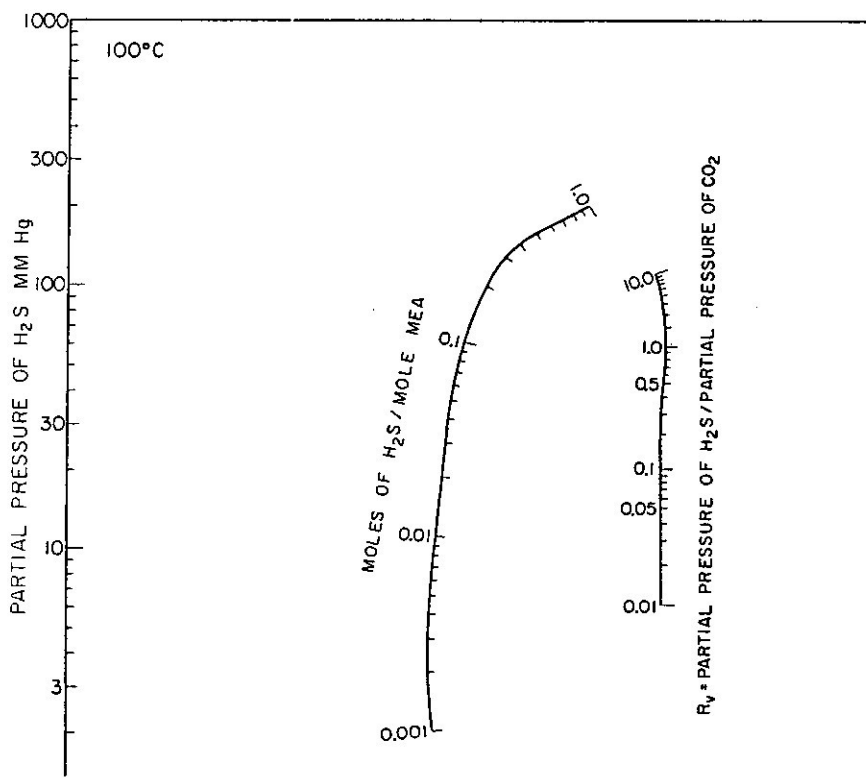
Figure 4.8 Equilibrium data for H_2S and 15.3% MEA at 100°C with CO_2 present

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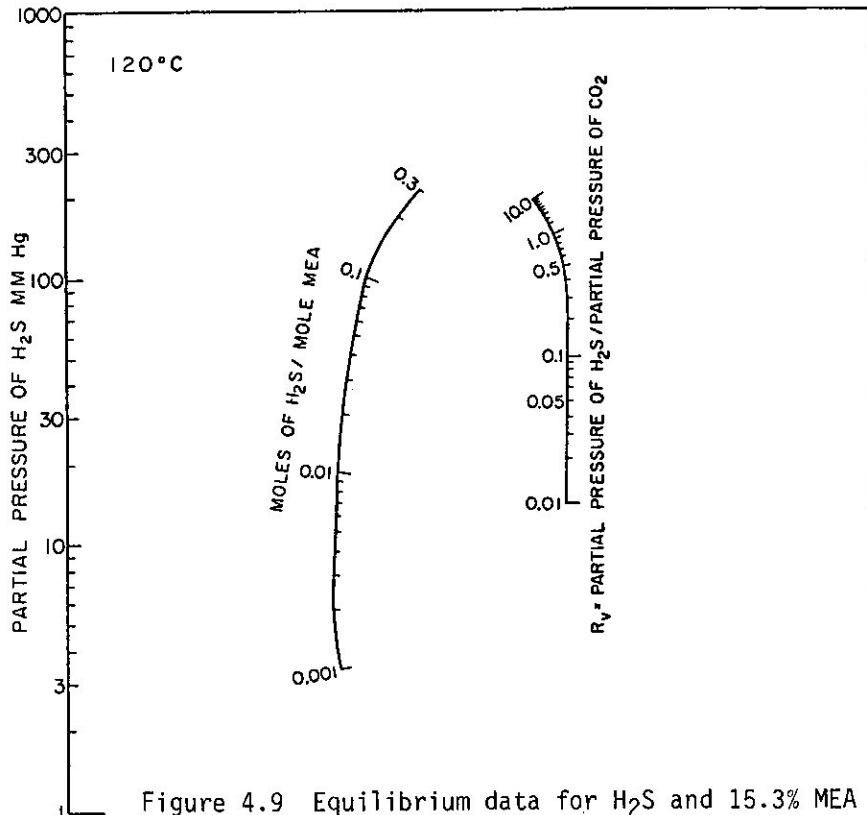
Figure 4.9   Equilibrium data for H$_2$S and 15.3% MEA at 120°C with CO$_2$ present
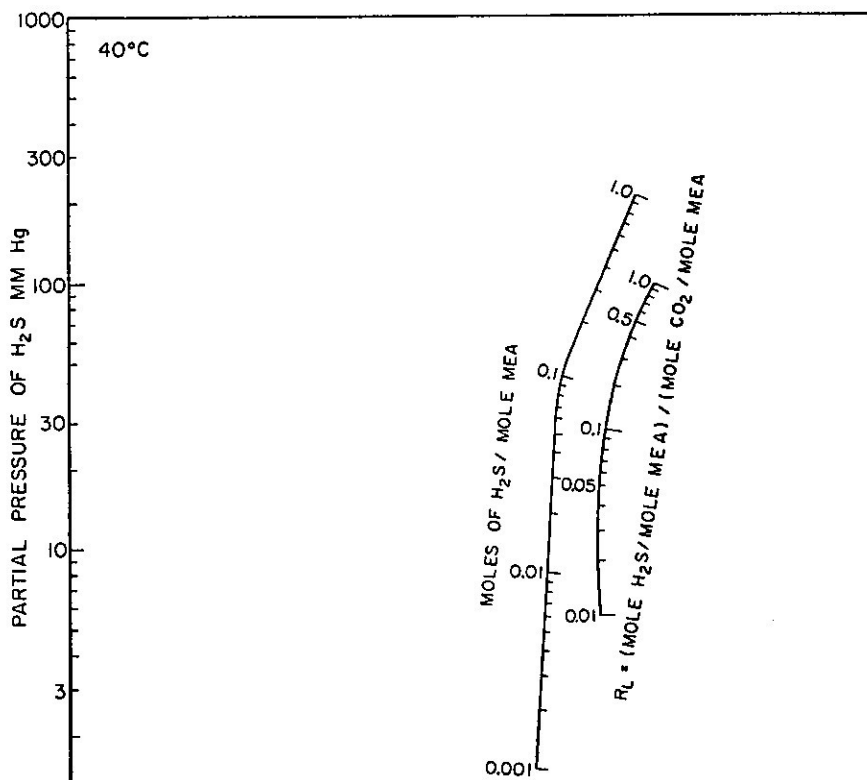
Figure 4.10   Equilibrium data for CO$_2$ and 15.3% MEA at 40°C with H$_2$S present
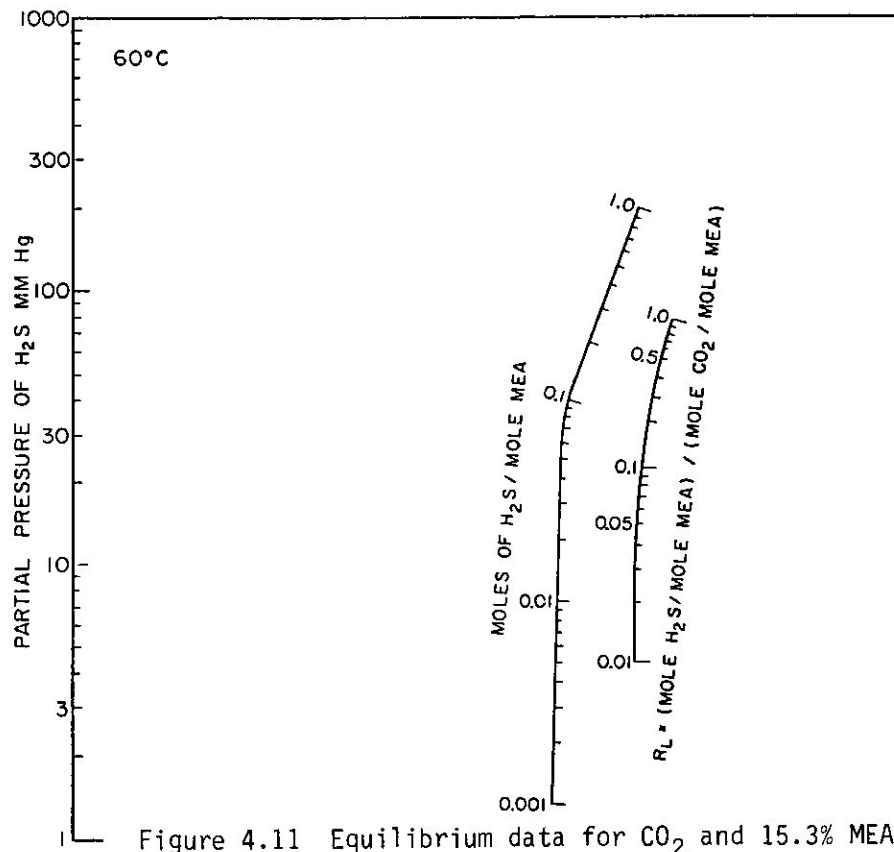
Figure 4.11 Equilibrium data for CO_2 and 15.3% MEA at 60°C with H_2S present
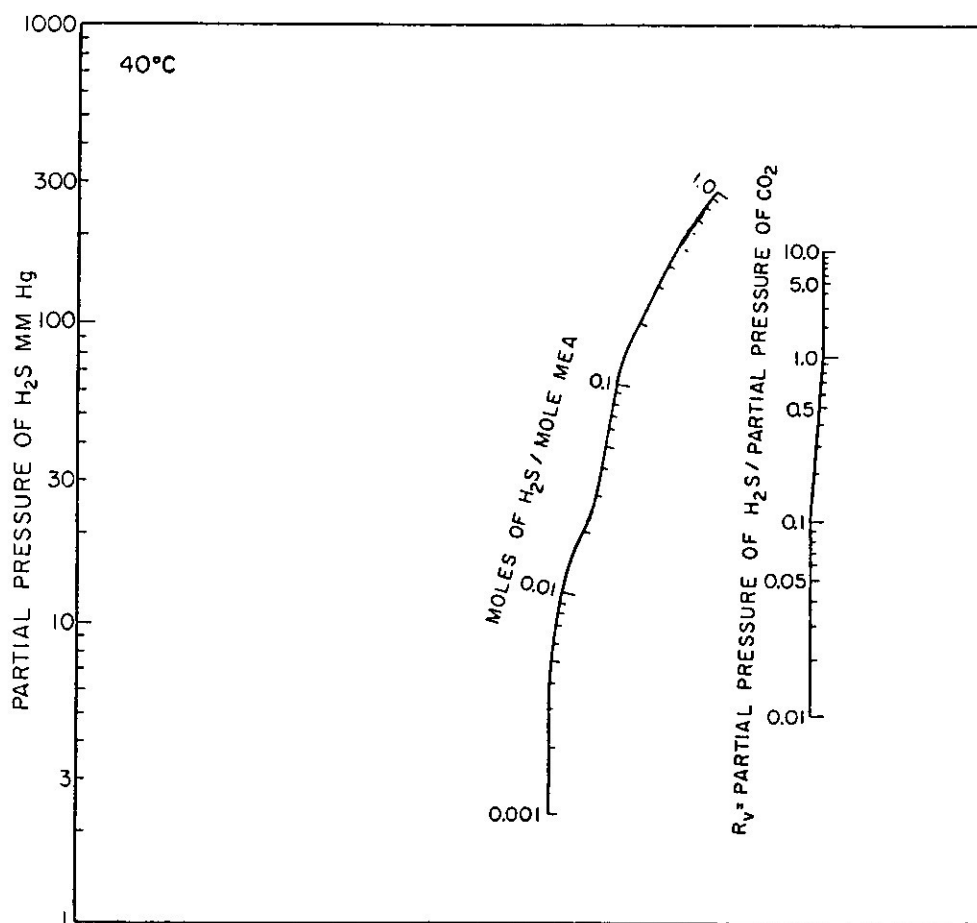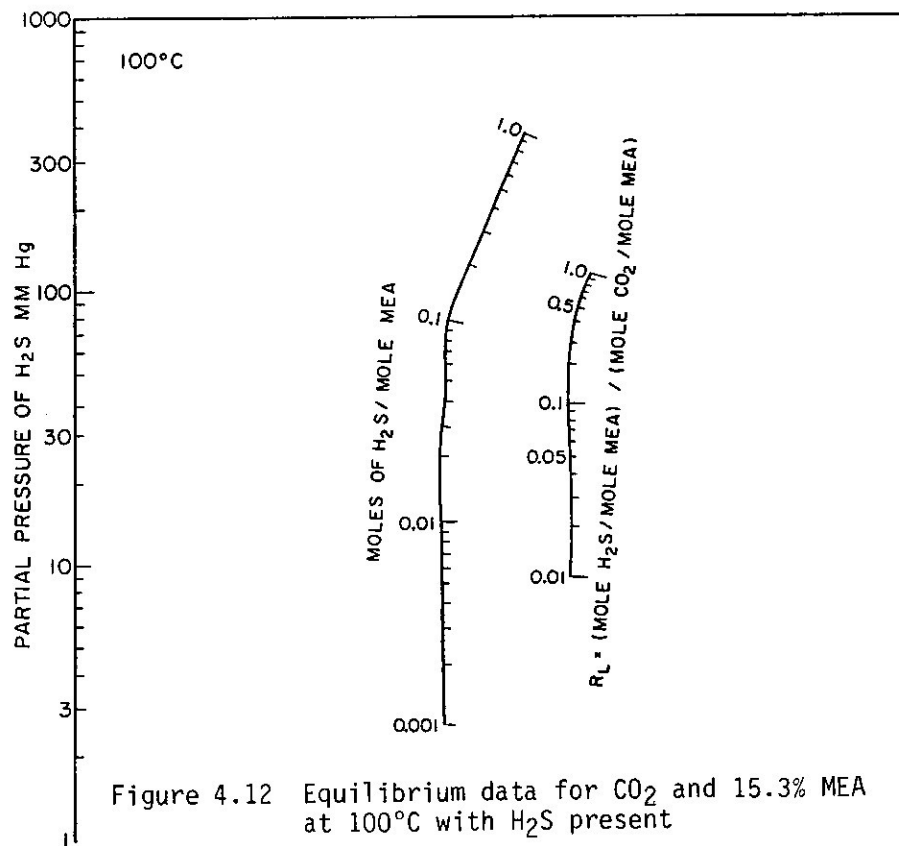

Figure 4.12 Equilibrium data for CO_2 and 15.3% MEA at 100°C with H_2S present

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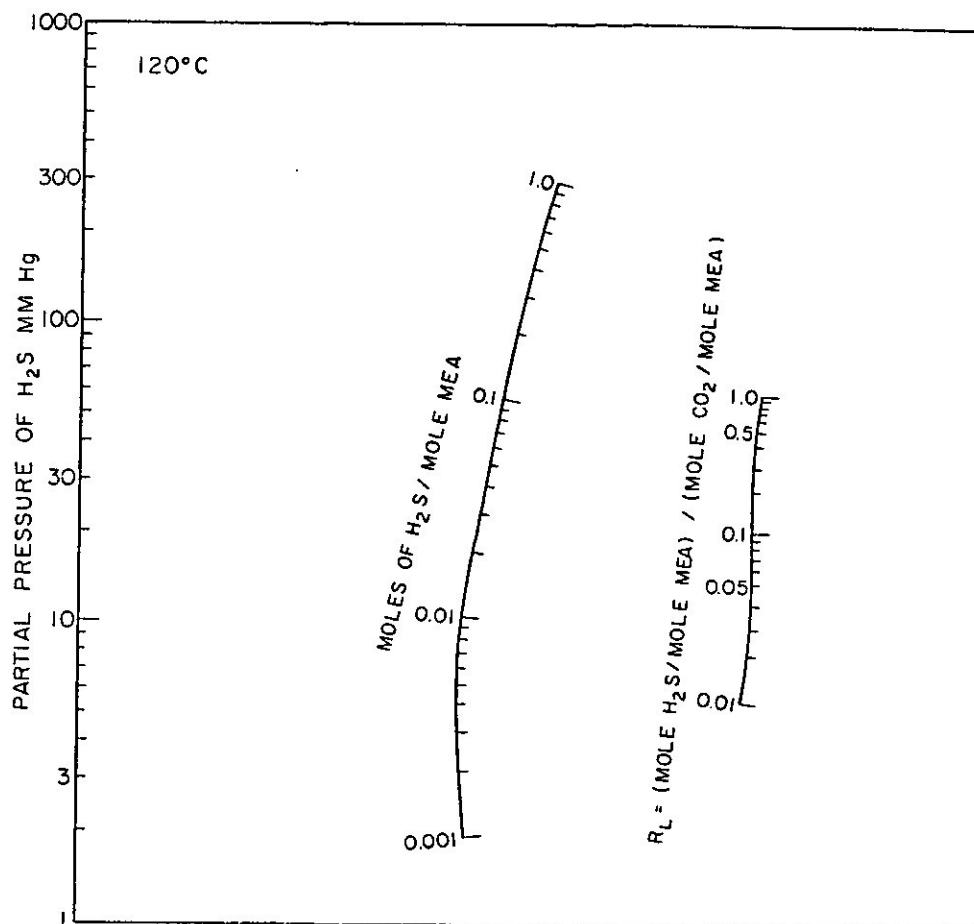
Figure 4.13   Equilibrium data for CO_2 and 15.3% MEA
at 120°C with H_2S present

Fitzgerald and Richardson (4.5, 4.6) made a study of several amine processing units operating on a variety of natural gas feed streams. Their data are plotted in Figs. 4.14 and 4.15. These data clearly show that high values of the H_2S/CO_2 ratio in the sour gas result in higher residual H_2S content in the amine solution. They also show that, regardless of the reboiler steam rate, there is for each H_2S/CO_2 ratio an asymptotic value of residual acid gas concentration below which the H_2S content cannot be reduced.

From Fig. 4.14 the residual H_2S content of the lean amine can be determined directly, once the feed gas composition is known and the steam rate to the amine stripper column has been fixed. Also shown on Fig. 4.14 is a dashed line indicating the limit of stripper operation to produce pipeline specification natural gas, with contact taking place at 900 psig and 110° F. This dashed line can be used as a general guide for other operating conditions.

Fitzgerald and Richardson also studied the effect of steam stripping rate and H_2S to CO_2 ratio in the sour gas on the retention of CO_2 in the stripped amine solution. Fig. 4.15 shows the correlation they derived from their studies on operating plants. Fig. 4.15, when used in combination with Fig. 4.14 and Figures 4.6 to 4.13, provides vital information to be used in finalized process design for MEA treating units.
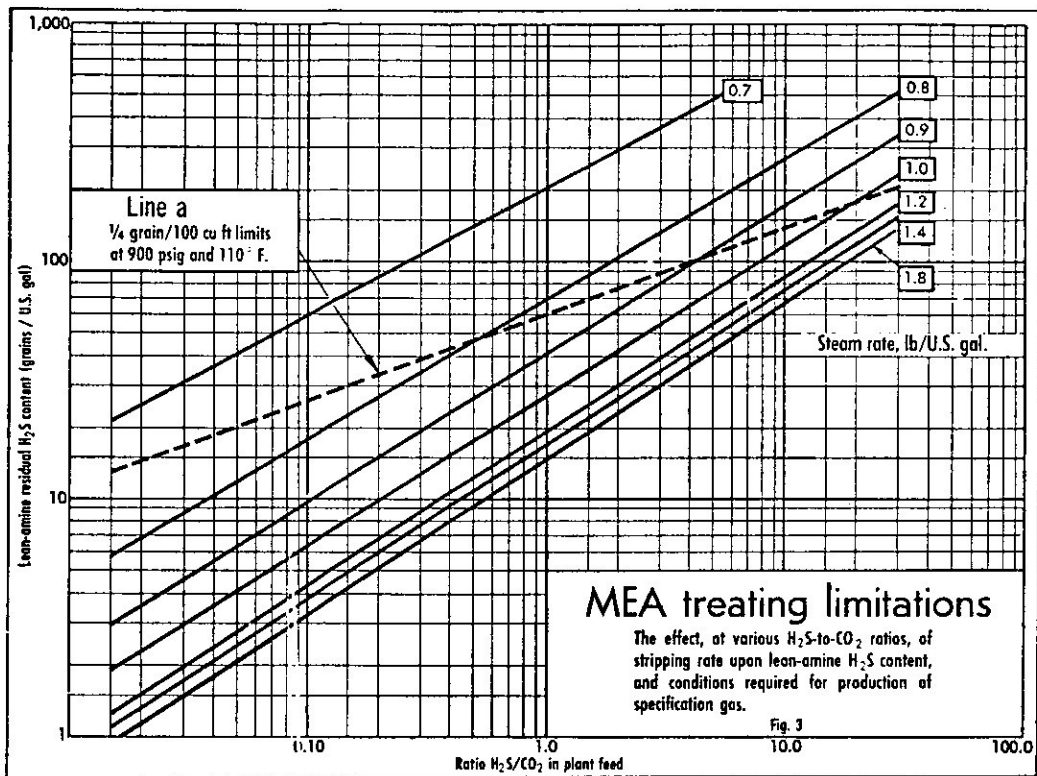
Figure 4.14   (Ref. 4.5   Courtesy Oil Gas J.)
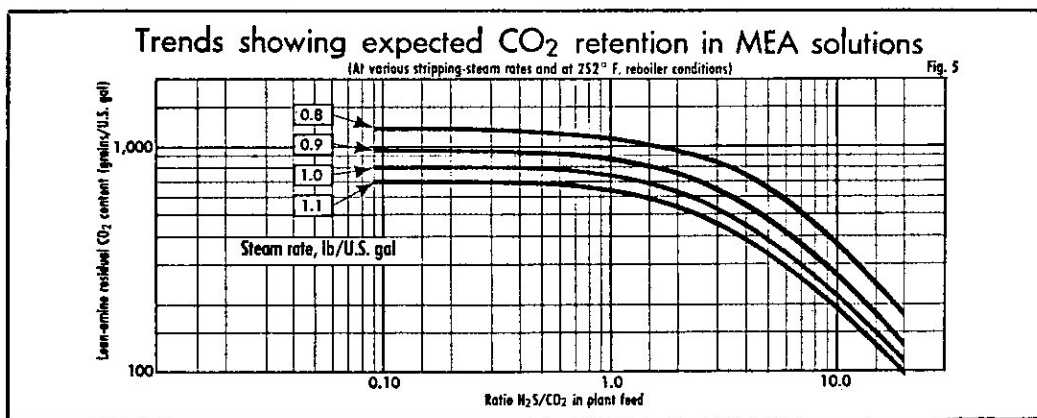
Figure 4.15   (Ref. 4.5   Courtesy Oil Gas J.)

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Fitzgerald and Richardson also show a typical temperature profile for an MEA absorber. This is reproduced in Fig. 4.16. In general, conditions at the top of the contactor will control outlet gas composition, and conditions at the bottom of the tower will control the holding capacity of the rich amine solution for acid gas. Tower top conditions seem to very closely approximate the temperature of the lean amine. Temperatures at the bottom of the tower are controlled by the heat of reaction of the H_2S and CO_2 in the amine. Kohl and Riesenfeld (4.7) also present amine contactor temperature profiles. Their profile for a contactor handling a high acid gas content gas is similar to that shown in Fig. 4.16. Their profile for a contactor treating a lean natural gas (less than 0.5 per cent total acid gas) shows reverse conditions with the amine solution temperature being highest at the top of the tower. Their profiles, however, were made on towers handling a mixture of glycol and amine and may or may not be exactly typical of an MEA contactor.
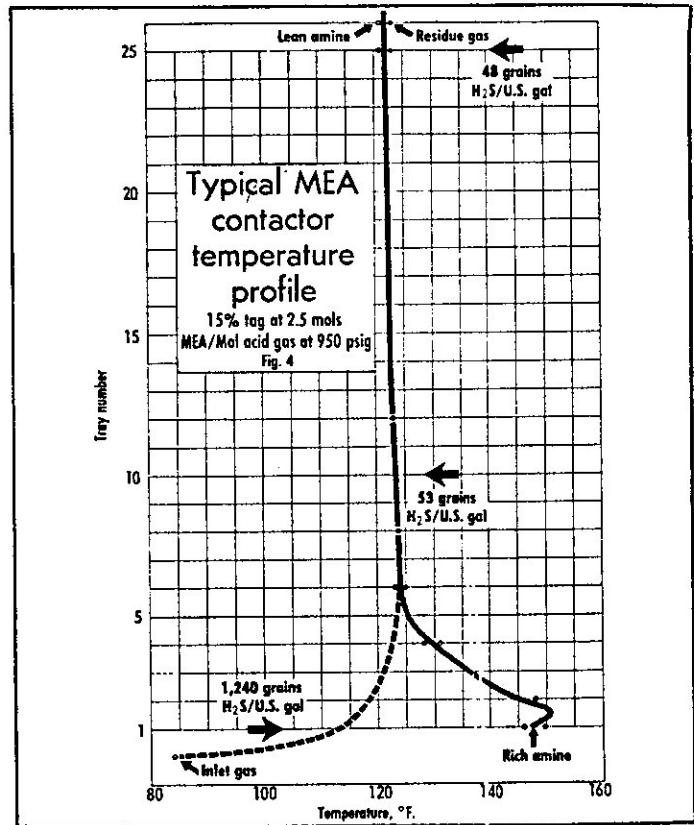
Figure 4.16   (Ref. 4.5   Courtesy Oil Gas J.)

A DESIGN PROCEDURE

The application of the foregoing data is illustrated by the example problem which follows.

Example Problem

A natural gas stream of 50 MMscfd at 900 psig is to be sweetened. The sour gas contains 0.50% H_2S and 2.0% CO_2. The gas is to be sweetened to pipeline specifications of 0.25 grains H_2S per 100 cu ft of gas. Assume that the gas is available at $90^\circ F$. and that lean amine solution can be cooled to $110^\circ F$.

The first step in the design of the process is to determine the solution circulation rate. This involves estimating the unstripped acid gas content of the lean amine in order to estimate conditions at the top of the column, and also determining the total acid gas pickup allowed in order to determine rich amine solution concentration at the bottom of the column. Zapffe(4.8) recommends a 65% approach to the equilibrium concentration at the bottom of the tower as the maximum that should be assumed for design purposes. Kohl and Riesenfeld(4.7) recommend that a 75% approach to the equilibrium concentration be used. Take a middle road between these two recommendations and assume that the rich amine solution reaches 70% of its equilibrium concentration at the bottom of the absorber.

Fig. 4.14 and 4.15 can be used to estimate the composition of the lean amine leaving the bottom of the stripper. Assume a stripping steam rate of 1 lb of steam per U.S. gallon of MEA solution. Then for an H_2S to CO_2 ratio of 0.25 (the existing ratio in the sour gas) from Fig.4.15, find a residual CO_2 concentration in the lean amine of approximately 800 grains of CO_2 per U.S. gallon.

Fig. 4.14 determines the residual H_2S concentration of the lean amine. Using the assumed steam stripping rate and the specified acid gas ratio in the feed gas, find a residual concentration of 12 grains of hydrogen sulfide per U.S. gallon of amine solution.

The residual concentrations of acid gas on a molar basis are:

$$\frac{800 \text{ gr } CO_2}{7000 \text{ gr/lb}} = 0.114 \text{ lb } CO_2 \qquad \frac{0.114 \text{ lb } CO_2}{44 \text{ lb/mol}} = 0.0026 \text{ mol } CO_2/\text{gal}$$

$$12 \text{ gr } H_2S = 0.00171 \text{ lb } H_2S \qquad \frac{0.00171 \text{ lb } H_2S}{34 \text{ lb/mol}} = 0.0000504 \text{ mol } H_2S/\text{gal}$$

$$(0.0026 + 0.0000504) = 0.00265 \text{ mols total acid gas per gallon of solution}$$

Taking the density of the 15.3 per cent by weight amine solution to be 8.3 lb per U.S. gallon, there will be

0.15(% by wt) x (0.997 x 8.33)lb/gal = 1.246 lb MEA per gallon

$$\frac{1.246 \text{ lb MEA}}{61.1 \text{ lb/mol}} = 0.0204 \text{ mols of MEA/gal}$$

$$\frac{0.0026 \text{ mol } CO_2/\text{gal}}{0.0204 \text{ mol MEA/gal}} = 0.1275 \text{ mol unstripped } CO_2/\text{mol MEA}$$

$$\frac{0.0000504 \text{ mol } H_2S/\text{gal}}{0.0204 \text{ mol MEA/gal}} = 0.0025 \text{ mol unstripped } H_2S/\text{mol MEA}$$

Assuming ideal gases, the partial pressure of H_2S in the sour gas will be

$$0.005 \text{ x } 915 \text{ x } \frac{760}{14.7} = 237 \text{ mm Hg}$$

The partial pressure of CO_2 is $0.02 \text{ x } 915 \text{ x } \frac{760}{14.7} = 948 \text{ mm Hg}$

The ratio of acid gas partial pressures is $R_v = \frac{237}{948} = 0.25$

Before the equilibrium composition of the MEA leaving the bottom of the contactor can be determined the temperature of the rich amine stream must be known. As a first assumption, assume that the rich amine leaves the contactor at 140°F. (60°C.). The equilibrium composition of the H_2S in the amine can then be read from the known value of R_v (0.25) and the H_2S partial pressure of 237 millimeters of mercury. From Fig. 4.7, read the equilibrium concentration of H_2S to be 0.096 moles of H_2S per mole of MEA. From Fig. 4.11, using the 237 millimeters of H_2S partial pressure in the sour gas and the equilibrium concentration of 0.096 moles of H_2S per mole of MEA, read an R_L value of 0.17.

$$R_L = \frac{\text{mols } H_2S/\text{mole MEA}}{\text{mols } CO_2/\text{mole MEA}} \qquad \text{Mols } CO_2/\text{mole MEA} = \frac{0.096}{0.17} = 0.565$$

The equilibrium acid gas concentration of the MEA solution at the estimated temperature leaving the absorber of 140°F. will be:

0.096 mol H_2S

0.565 mol CO_2

Using a 70% approach to the equilibrium concentration:

$0.096 \times 0.7 = 0.0672$ mol H_2S/mol MEA

$0.565 \times 0.7 = 0.3955$ mol CO_2/mol MEA

This is the concentration of the H_2S and CO_2 in the rich MEA solution leaving the bottom of the contactor. The net pickup in the contactor will be the difference between the concentration leaving the bottom of the contactor and the unstripped acid gas remaining in the lean amine solution as it enters the top of the contactor.

$0.0672 - 0.0025 = 0.0647$ mol H_2S pickup/mol MEA

$0.3955 - 0.1275 = 0.2680$ mol CO_2 pickup/mol MEA

The total amount of acid gas to be removed can be determined from the total gas flow rate and the percentage of acid gas in the sour gas stream.

50 mmscfd = 34,700 scfm

Since the gas contains 2% of CO_2, there will be

$0.02 \times 34,700 = 694$ scfm CO_2 $\dfrac{694 \text{ scfm}}{380 \text{ scf/mol}} = 1.826$ mol CO_2/min.

0.5% H_2S = 174 scfm H_2S = 0.458 mol H_2S/minute

For the sake of simplicity in material balance calculations, assume that all acid gas carried in with the sour gas stream will be removed by the amine solution. This is not quite correct because there will be 0.25 grains of H_2S remaining per 100 scf of gas. However, this is only approximately 4.0×10^{-6} mol-fraction. There will be about the same amount or slightly more CO_2, but for material balance purposes consider these concentrations negligible.

$\dfrac{0.458}{0.0647} = 7.079$ mol MEA/min needed for H_2S

$\dfrac{1.826}{0.2680} = 6.813$ mol MEA/min for CO_2

$\dfrac{7.079}{0.0204} = 347$ gpm required circulation rate of MEA solution

The design circulation rate of 15% MEA solution would be 350 gpm.

The calculated acid gas pickup is 0.0323 moles acid gas per mole of MEA, and very close to the maximum recommended. A rule-of-thumb for acid gas pickup is not more than 0.35 moles pickup per mole of MEA, or, on another basis, no more than 2.5 standard cu ft of total acid gas pickup per gallon of MEA solution.

The reader should not infer that an acid gas concentration in the sour gas of 2.5% will always give proper amine solution loading. The ratio of H_2S to CO_2 in the acid gas is important. As an illustration, in the case at hand, with an H_2S partial pressure of 237 mm of mercury, the equilibrium loading of H_2S in the rich amine solution would be approximately 0.72 moles of H_2S per mole of MEA (from Fig. 4.7). This shows very clearly that CO_2 in the gas has an effect on H_2S. The reverse is true also. The effect of one acid gas constituent on the other is not simple, but can be predicted through the use of the Jones data as plotted by McCoy and Maddox in Fig. 4.6 to 4.13.

The next step would be to check the temperature of the rich amine solution leaving the bottom of the contactor. This can be done by making an over-all heat balance on the absorber to determine the rich amine temperature. There are

$$\frac{694 \text{ cu ft/min}}{380 \text{ cu ft/mol}} = 1.826 \text{ mol } CO_2/min$$

$$1.826 \text{ mol } CO_2/min \times 44 \text{ lb/mol} = 80.3 \text{ lb } CO_2/min$$

Kohl and Riesenfeld(4.7) give a heat of reaction between CO_2 and MEA of 825 Btu per pound, and for H_2S and MEA a heat of reaction of 820 Btu per pound.

$$80.3 \text{ lb } CO_2/min \times 825 \text{ Btu/lb} = 6.63 \times 10^4 \text{ Btu/min}$$

Likewise, there are 15.6 lb of H_2S reacting per minute.

$$15.6 \text{ lb } H_2S/min \times 820 \text{ Btu/lb} = 1.28 \times 10^4 \text{ Btu/min}$$

The total heat release from the reaction of the sour gas and the amine is

$$(6.63 + 1.28) \times 10^4 = 7.91 \times 10^4 \text{ Btu/min}$$

The heat pickup by the sweet gas can be determined using the enthalpy charts in the NGPSA Engineering Data Book(4.10). Taking the sweet gas to be pure methane (molecular weight = 16) and estimating the temperature of the sweet gas leaving the contactor to be the same as the entering lean amine (based on Fig. 4.16), the heat pickup by the sweet gas in passing through the contactor will be

enthalpy at 110°F.   258 Btu/lb
enthalpy at 90°F.   247 Btu/lb
                       11 Btu/lb pickup

The mass of sweet gas per minute will be

$$(\frac{34,700}{380} \times .975) \times 16 = 1,425 \text{ lb/min}$$

The total heat pickup by the gas will be

$$1.425 \times 10^3 \times 11 = 1.56 \times 10^4 \text{ Btu/min}$$

The temperature rise of the amine solution will be

$$\frac{(7.91 \times 10^4 - 1.56 \times 10^4) \frac{\text{Btu}}{\text{min}}}{(8.3 \frac{\text{lb}}{\text{gal}}) (350 \frac{\text{gal}}{\text{min}}) (0.915 \frac{\text{Btu}}{\text{lb}^\circ\text{F}})} = 23.9^\circ\text{F}.$$

This gives a strong amine solution temperature of 134°F. This is close enough to the assumed value of 140°F. that there is no need to repeat the lengthy foregoing calculations.

Contact Performance

Fixing the number of trays, either theoretical or actual, in the absorber is not a simple task. As a matter of fact, many authors feel that computation of the number of theoretical trays required is an exercise in futility (4.14). The calculation of rates and efficiencies in a simple absorption process is complicated enough. In the case of absorption followed by chemical reaction, the calculations become exceedingly difficult.

Limited amounts of experimental work have been reported on the dynamic absorption (reaction) of pure H_2S and CO_2 in water solutions of ethanolamines. However, these data are insufficient to provide a sound basis for rational plant/process design. Essentially no data are available on the simultaneous dynamic absorption/reaction of H_2S and CO_2 into any of the ethanolamines in water solution.

DIETHANOLAMINE

Solubility of CO_2 and H_2S in DEA

Lee et al. (4.11, 4.12) have taken mutual solubility data for CO_2 and H_2S in DEA solutions. The solution strengths were 3.5 normal and 2.0 normal and the temperatures were 50 and 100° C. The 2.0 normal solution is 20% DEA by weight and the 3.5 normal solution is 35% DEA by weight. The temperatures correspond approximately to temperatures at the bottom of the contactor and the bottom of the regenerator. The Lee et al. data are shown in Fig. 4.17 to 4.24.

For temperatures other than those for which data are taken, Lee et al. recommend interpolating by making use of the fact that log vapor pressure theoretically is a straight line function of reciprocal absolute temperature. To illustrate this point, assume that the equilibrium loading of a 2.0N DEA solution is 0.6 moles of CO_2 per mole of DEA at 70° C. No H_2S is present.

Lee et al. also report heat of solution values for CO_2 and H_2S in DEA. The values for CO_2 are shown in Table 4.6. The heat of solution of CO_2 in DEA is dependent on both the strength of the DEA solution and the amount of the DEA that has reacted with the CO_2. The heats of solution for H_2S in DEA are shown in Table 4.7. The H_2S heat values are dependent only on the amount of DEA that has reacted with H_2S and are largely independent of the DEA solution strength.

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The effect of acid gas loading on MEA solution pH is interesting. This may explain part of the reason for corrosion problems in MEA plants. Two trends are definitely clear from Fig. 4.27 — increases in acid gas loading lower the pH of the solution as does an increase in temperature. Both of these factors may be at work in some areas of severe corrosion such as heat exchangers and the top of the stripping column.

The prediction of tray efficiencies and/or mass transfer rates is not exact. In general, 20 actual trays (or an equivalent height of packed column) will be provided in the absorber. In most instances, these 20 trays will provide sufficient contact for treating of sour gas to pipeline specifications, providing that the solution is properly regenerated and that gas loadings in the rich amine solution are not excessive.

STRIPPER DESIGN

The primary problem in designing the stripper is estimating the heat duty on the reboiler. This heat load includes:

1. *The sensible heat required to raise the temperature of the feed amine solution to that of the regenerated solution leaving the reboiler.*

2. *The heat of reaction required for dissolution of the acid gases from the amine.*

3. *The stripper column condenser duty which is essentially the heat of vaporization of the water leaving the stripping section of the column that is condensed and returned to the stripper as reflux.*

Properly estimating the condition and enthalpy of the feed to the stripper is a very complicated problem. As the feed to the stripper passes through the amine-to-amine heat exchanger, its temperature will rise. As the temperature increases, a portion of the absorbed acid gases will be vaporized. The portion of the acid gases that vaporizes will require a *heat of dissolution.* Estimating the amount and composition of acid gases regenerated is a difficult, trial-and-error procedure.

A simple practical approach is to assume that no vaporization occurs in the amine-to-amine heat exchanger and that the feed solution simply takes up the heat released by the lean amine solution; that all of this heat goes to raise the temperature of the feed to the stripper. This increase in feed temperature does tend to decrease the amount of heat required in the reboiler. However, since no acid gases are assumed stripped from the solution in the amine-to-amine exchanger, this tends to increase the reboiler duty with the two opposite effects essentially counteracting each other.

In the general case, however, the amine-to-amine heat exchanger will not transfer the maximum amount of heat. As a rule-of-thumb, the hot, rich amine solution leaving the amine exchanger should be at about 180° to 200° F. for proper feed temperature to the stripper. In general, this will require less heat than is available for transfer from the hot lean solution leaving the bottom of the stripper. Also, in most cases, the two amine streams will exchange heat on essentially an equal temperature difference basis — a 10° lowering in the lean amine solution will result in a 10° rise in the rich amine stream. As a result, the engineer is usually safe in arbitrarily fixing the temperature of the feed stream to the stripper column and proceeding as follows to calculate the stripper, based primarily on heat balance considerations.

Example Problem

For the purposes of this example problem, assume that the rich amine
feed to the stripper is entering at 190°F. and that the reboiler is
operating at 240°F. The original estimate of acid gas content in the
lean amine was based on a steam rate to the reboiler of 1 lb per U.S.
gallon. Assume the steam enters the reboiler saturated at 250°F.
(latent heat equal 945 Btu/lb).

A heat balance around the stripper is:

$$H_{RA} + Q_R = H_{AG} + Q_C + H_{LA}$$

where: H_{RA} = heat entering with rich amine solution Btu/unit
of time

H_{AG} = heat leaving with acid gas from top of stripper,
Btu/unit of time

H_{LA} = heat leaving stripper in lean amine solution
Btu/unit of time

Q_R = reboiler duty, Btu/unit of time

Q_C = condenser duty, Btu/unit of time

The problem now is to properly evaluate each of these terms. There is
sufficient information to evaluate the left-hand side of the equation
and two of the three terms on the right-hand side of the heat balance
equation, with the condenser loading being unknown.

With the feed entering the column at 190°F. there are:

350 gal/min of 15% MEA solution containing 0.476 moles of
H_2S and 2.735 moles of CO_2

The lean solution leaving the reboiler at 240°F. will be 350 gal/min
of 15% MEA containing 12 grains of H_2S and 800 grains of CO_2 per gallon,
equivalent to a total of 0.0176 moles of H_2S and 0.909 moles of CO_2 per
minute.

Acid gas overhead will be 0.458 moles of H_2S and 1.826 moles of CO_2 per
minute. With the acid gas overhead will be sufficient water vapor for
reflux to the stripper. The amount of this water vapor will be deter-
mined from the condenser loading as shown below.

The term (H_{LA} - H_{RA}) can be evaluated as the difference in heat required
for the lean amine constituents to be heated from the feed temperature
of 190°F. to the lean amine temperature of 240°F. This sensible heat
requirement is:

2947 lb/min x 0.98 Btu/lb-°F. x (240 - 190)°F. = 144,380 Btu/min

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The acid gas portion of the rich amine solution is changed from a reacted gas in the liquid phase to acid gases in the overhead. The amount of heat required for this conversion is essentially the heat of reaction of the acid gases with the amine. This has previously been calculated as 8.09×10^4 Btu/min.

The heat released from the steam in the reboiler is:

$$350 \text{ gal/min} \times 945 \text{ Btu/lb} = 330,750 \text{ Btu/min}$$

This is based on the assumption that the heat release in the reboiler is the heat of condensation of the saturated steam entering the reboiler.

Rearranging the heat balance equation gives:

$$Q_C = (H_{RA} - H_{LA}) - H_{AG} + Q_R$$

Substituting

$$Q_C = -(144,380 + 80,900) + 330,750 = 105,470 \text{ Btu/min}$$

The amount of water condensed in the condenser can be calculated by assuming that the condenser operates at 20 psia, where water had a heat of vaporization of 970 Btu/lb. This gives:

$$\frac{105,470 \text{ Btu/min}}{18 \text{ lb/mole} \times 970 \text{ Btu/lb}} = 6.04 \text{ moles water condensed/min}$$

Recommendations for the amount of water returned to the column as reflux vary from 2-to-1 to 3-to-1 on the basis of the total acid gas going overhead in the stripper.

$$\frac{6.04 \text{ moles of water/min}}{2.34 \text{ moles of acid gas/min}} = 2.58 \text{ moles water/mole acid gas}$$

This falls within the recommended range and should give satisfactory stripping operation.

Modelling Ethanolamine — Acid Gas Equilibrium

Over the years many efforts have been made to systematize the equilibrium between ethanolamines and acid gases in such a way that information could be made useful for process/design calculations. One approach which has met with reasonable success is to describe the reactions taking place between the ethanolamines and acid gas constituents. With appropriate pseudo equilibrium constants written for each reaction the equilibrium composition of all species in solution can be calculated. Through use of appropriate relationships to equilibrate vapor (gas phase) and liquid composition the entire ethanolamine — water — natural gas — acid gas system can be modeled at equilibrium. Kent and Eisenberg (4.15, 4.16) proposed a set of equations as follows to model the CO_2/H_2S reactions with ethanolamines: