Optimization of New and Existing Amine Gas Sweetening Plants Using Computer Simulation

JERRY A BULLIN,
Chemical Engineering Dept., Texas A&M University,
College Station, Texas

JOHN C POLASEK, JOSEPH W HOLMES,
Bryan Research & Engineering, Inc.,
Bryan, Texas

ABSTRACT

Large volumes of natural gas are produced throughout this country and Canada for the purpose of heating homes, producing electricity and generating heat for a wide variety of industries. As it comes from the ground, much of the gas produced contains quantities of acid gases, notably H2S and CO2. The carbon dioxide is of little consequence for the most part, but H2S is quite toxic and virtually all of this gas must be removed before the gas can be sent to commerical pipeline.

Proceedings of the Sixtieth GPA Annual Convention. Tulsa, OK: Gas Processors Association, 1981: 142-8.

Bryan Research & Engineering, Inc.
Visit our Engineering Resources page for more articles.

INTRODUCTION

Large volumes of natural gas are produced throughout this country and Canada for the purpose of heating homes, producing electricity and generating heat for a wide variety of industries. As it comes from the ground, much of the gas produced contains quantities of acid gases, notably H_2S and CO_2. The carbon dioxide is of little consequence for the most part, but H_2S is quite toxic and virtually all of this gas must be removed before the gas can he sent to a commercial pipeline.

The specification in this country states that there must be less than .25 grain H_2S per 100 standard cubic feet of gas or approximately 4 ppm H_2S in the pipeline gas. I here are a large number of sweetening plants currently in operation and many more are constructed each year.

One of the most common methods of acid gas removal involves contacting the sour gus with,>n amine water solution. The general process diagram is shown in Figure 1.

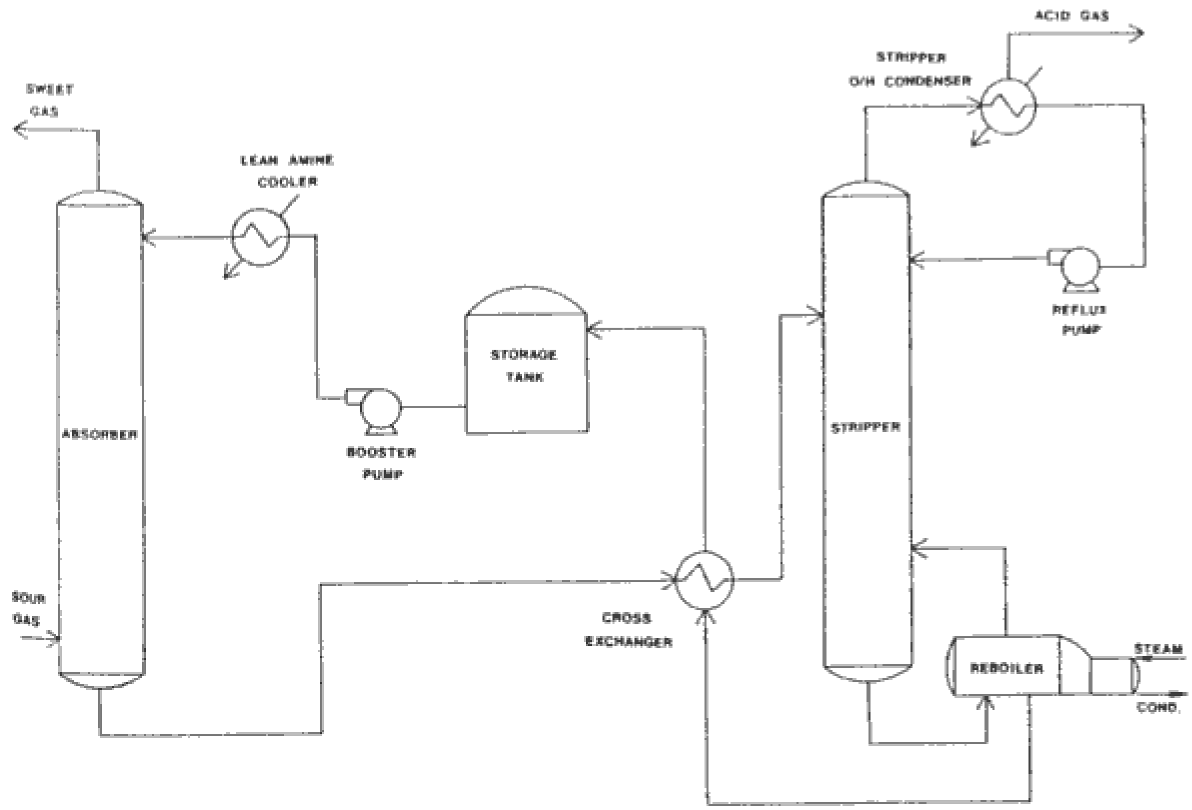

Figure 1. Process flow sheet for Example 1 plant.

The sour gas is contacted with the amine solution and the acid gases are absorbed. The solution is then heated and passed to a low pressure still column to regenerate the amine solution by driving the acid gases off. The solution is then recirculated to the absorber. The streams out of thc plant are a sweet gas stream which meets pipeline specifications and an extremely sour gas stream containing varying amounts of acid gases which is either flared or passed to a sulfur production facility.

Some of thc commonly used amines are monoethanolamine, diethanolamine, and diglycolamine. Each amine has a unique set of properties which make it desirable under certain conditions and undesirable under other conditions.

Designing an amine sweetening unit requires careful thought on the part of the design engineer. The list of parameters which must be decided is quite lengthy. The engineer must first decide what embellishments should be added to this basic flow sheet.

Precontactors to reduce absorber size, fuel gas flash drums and scrubbers, amine reclaimers, and multiple feed gases are all used at times. The designer must then decide the best amine for his particular feed gas and chosen flow sheet.

In the absorber, he must decide the maximum allowable acid gas concentration and the operating temperature and pressure of the stripper, the temperature of the overhead condenser, and the heat rate in the reboiler. All of these factors must he traded off to produce an economical and reliable plant.

To date, the design process for these plants has involved approximations based on existing plants and guesswork. With the advent of the energy crisis and the environmental crisis, guesswork and approximation are no longer good enough for practical design.

This paper describes the construction and use of a computer simulation program with predicts the performance of amine sweetening units based on reaction kinetics and tray by tray distillation calculations.

DESCRIPTION OF SIMULATION PROGRAM

The program utilizes the flexible flowsheet concept. This allows the user maximum flexibility in the layout of the projected unit. The basic flow sheet becomes the block diagram shown in Figure 2. Each block represents a given unit operation, with the user able to specify important parameters about each operation.

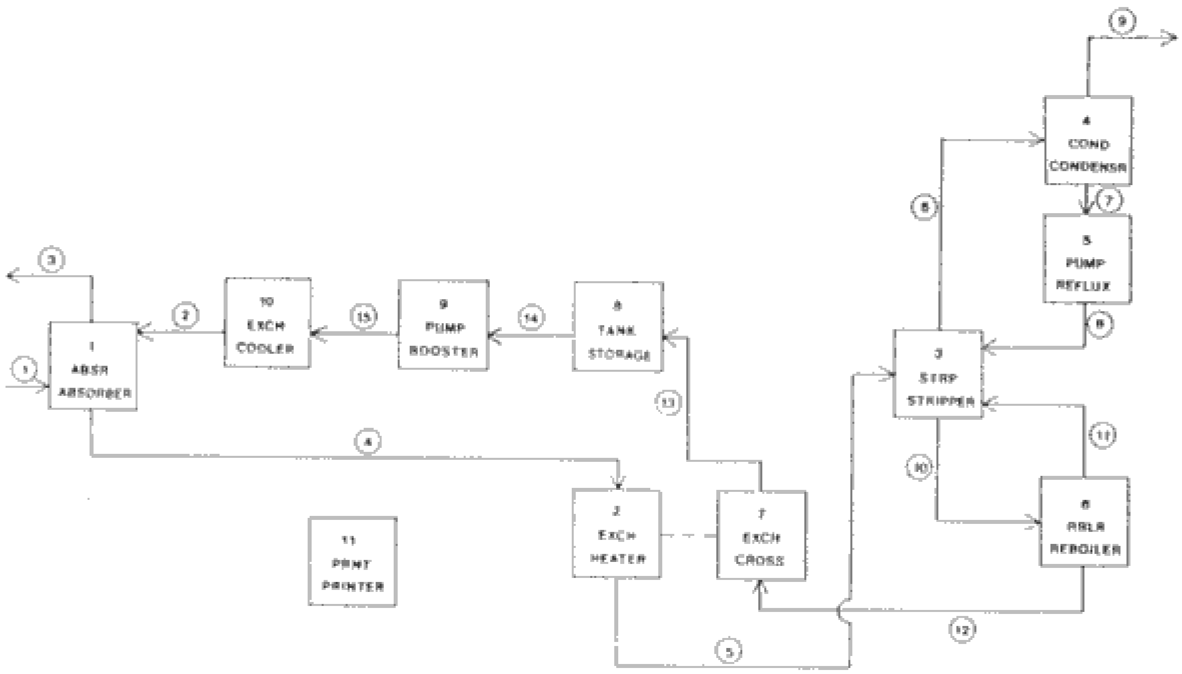

Figure 2. Process simulation diagram for Example 1 plant.

For the absorber block, the user must specify the amine type (MEA, DEA, or DGA), the weight percent amine, and either the desired acid gas loading or the percent of equilibrium acid gas loading. The user may optionally specify the pressure drop through the absorber, the pounds of liquid entrained overhead from the absorber, the tray spacing or packing factor, and the percent of flood.

The absorber block returns the amount of acid gases overhead and, if the optional parameters are specified, the size of the absorber.

Each side of the exchanger is a separate unit operation. The user must specify either the temperature out of the exchanger, the heat transferred, or the unit supplying the heat transferred information. The block returns the heat duty and outlet temperature as well as the log mean viscosity, log mean thermal conductivity, log mean density, and if the block is part of a cross exchanger, the log mean temperature difference.

The regenerator consists of three related blocks; a stripper, a reboiler, and a condenser. The parameters which must be specified in these blocks are the pressure overhead at the stripper, the steam rate to the reboiler, the pressure drop through the condenser and through the column, and the condenser temperature.

Optional parameters which can be specified are the tray spacing or packing factor in the column and the fraction liquid overhead from the reboiler. The blocks return the heat duties for the condenser and reboiler, the reflux ratio, the condenser knockout drum size, and the acid gas concentrations in the lean amine.

If the optional parameters are specified, the stripper is also sized.

The tank block is required to adequately close the recycle loop on the amine stream. It has an optional parameter of holdup time. The tank size necessary for the specified holdup is calculated by the block. The other parameters returned by the block are the circulation rate, the amine makeup, and water makeup necessary because of losses overhead from the absorber and regenerator.

The pump block is used to boost the lean amine solution pressure sufficiently to enter the absorber. The user must specify either the outlet pressure or the pressure rise through the block. The user may also specify the number of stages and efficiency of the pump.

The block returns the power required by the pump. Examples of the output from the program are shown in Figure 3.

Figure 3. Example program output.

BLOCK NUMBER 1.00 (ABSR) UNIT NAME: ABSORBER			
Amine Type	=	1.000	
Wt. Percent Amine	=	15.000	Percent
Amine Loading or -% Equil.	=	0.450	M/M amine or -%
Amine Cooler Block Num.	=	10.000	
Pressure Drop	=	10.000	psi
Entrained Liquid	=	1.000	lb/MM scf gas
Number of Ideal Stages	=	7.000	
Tray Spacng/-Packng Fct	=	2.000	Feet or dim'less
Percent of Flood	=	70.000	Percent
Absorber Diameter	=	4.250	Feet
Percent of Max Load (H2S)	=	68.296	Percent
H2S Overhead (Equilm)	=	0.049	Grains/100scf
CO2 Overhead (Equilm)	=	0.000	Percent
Actual Acid Gas Loading	=	0.450	Mole/mole amine
BLOCK NUMBER 2.00 (EXCH) UNIT NAME: HEATER			
Exchanger Type	=	1.000	
Target Parameter Value	=	220.000	
Pressure Drop	=	10.000	psi
Outlet Temperature	=	220.000	Degrees F.
Heat Duty	=	15129216.000	BTU/hr
Log Mean Viscosity	=	2.102	Centipoise
Log Mean Therm Cond	=	0.320	BTU/hr ft deg F.
Log Mean Density	=	60.976	lbs/cu ft
BLOCK NUMBER 3.00 (STRP) UNIT NAME: STRIPPER			
Stripper Pressure (Top)	=	12.000	psig
Steam Rate at Reboiler	=	1.000	lb/gal amine
Number of Ideal Stages	=	7.000	
Ideal Feed Stage	=	3.000	
Condenser Block Number	=	4.000	
Reboiler Block Number	=	6.000	
Reclaimer Block Number	=	0.000	
Tray Spacng/-Packng Fct	=	-84.000	Feet or dim'less
Percent of Flood	=	60.000	Percent
Stripper Diameter	=	6.531	Feet
H2S in Lean Amine	=	25.428	Grains/gal
CO2 in Lean Amine	=	0.101	CO2/amine (M/M)

STREAM INFORMATION FOR PLANT

Stream No.:	**1**	**2**	**3**	**4**
	lb-mole/hr	lb-mole/hr	lb-mole/hr	lb-mole/hr
MEA	0.000	405.394	0.075	405.318
DEA	0.000	0.000	0.000	0.000
DGA	0.000	0.000	0.000	0.000
H2	0.000	0.000	0.000	0.000

N2	0.000	0.000	0.000	0.000
O2	0.000	0.000	0.000	0.000
H2S	27.860	2.141	0.004	29.997
CS2	0.000	0.000	0.000	0.000
COS	0.000	0.000	0.000	0.000
SO2	0.000	0.000	0.000	0.000
NH3	0.000	0.000	0.000	0.000
HCN	0.000	0.000	0.000	0.000
CO	0.000	0.000	0.000	0.000
CO2	111.400	41.001	0.009	152.391
H2O	0.000	7787.887	8.612	7779.270
CH4	5014.000	0.000	4997.805	16.195
C2H6	557.100	0.000	555.375	1.726
C3H8	0.000	0.000	0.000	0.000
C4H10	0.000	0.000	0.000	0.000
C5H12	0.000	0.000	0.000	0.000
C6+	0.000	0.000	0.000	0.000
Press psig	900.000	949.500	890.000	900.000
Temp Deg F	90.000	110.000	110.000	129.677
Mol Frac Liq	0.000	1.000	0.000	1.000
Lbs/cu ft	2.801	62.058	2.584	61.721
Total mol/hr	5710.355	8236.418	5561.875	8384.891
Total lbs/hr	103028.625	166980.625	97025.125	172983.937
Flow gpm	0.000	335.453	0.000	349.408
Total ft3/hr	36788.305	2690.730	37545.902	2802.667
Visc centips	0.011	2.444	0.011	2.334
Tk BTU/hr-F	0.031	0.305	0.033	0.312
CP BTU/mol-F	9.005	19.457	9.180	19.200
From Block	Inlet	Cooler	Absorber	Absorber
To Block	Absorber	Absorber	Outlet	Heater

The program returns all of the input parameters and the calculated values for the other parameters. The figure shows the results section of a program run for the first two blocks of a plant. The run name is placed at the top of the sheet, then each block is labeled as to type and user specified name.

Each parameter is labeled as to name and units. The first block is the absorber, with the input specifications of amine type, weight percent amine, amine loading, and amine cooler block number. The amine cooler block number is necessary in order to properly evaluate the temperature of the inlet lean amine.

The user has also specified a pressure drop, a value for entrained liquid overhead in the absorber, a tray spacing, and a desired percent of flood. The program has returned an absorber diameter, a percent of maximum acid gas loading (in this case, the H_2S partial pressure becomes limiting first), the H_2S and the CO_2 in the product gas, and the actual acid gas loading.

This parameter is necessary if the user had specified a loading that was in excess of equilibrium or had specified a fraction of equilibrium. The second block is the first side of the cross exchanger.

The user has input an exchanger type of 1 (outlet temperature specified), the target parameter value (in this case 220^{o}F), and the pressure drop. The block returns the outlet temperature, the heat duty, the log mean viscosity, thermal conductivity, and density.

The program also returns the composition and physical properties of the streams flowing between the blocks. The rest of Figure 3 shows the first page of the stream information section of a typical program output. The lb moles per hour of each compound in the stream are listed first, followed by the physical properties of the stream.

The physical properties listed are the pressure, temperature, fraction liquid, density, total moles per hour, mass per hour, gallons per minute for liquid streams, cubic feet per hour for all streams, viscosity, thermal conductivity, and heat capacity. The block from which the stream originated and to which the stream goes are also identified to help the user.

Stream 1 (from block INLET) is an input specification which the user had to set. Stream 3 (to block OUTLET) is the sweet gas out of the plant. The stream specifications are enough to let the user completely identify the streams.

There are several other blocks included in the package for the use in the design of more complicated plants. These blocks include a precontactor block to reduce the flow rate through the absorber for high acid gas streams, a flash block used to generate low pressure fuel gas, a reclaimer block to remove a fraction of the amine if significant degradation is expected, and splitter and mixer blocks necessary to make these auxiliary blocks function properly.

These blocks greatly expand the use of the program, allowing the designer enough flexibility to mimic the flow sheet of almost any amine plant. However, the results of changing plant flowsheets will not be addressed at this time.

This paper will discuss the results of simulations employing the simple flowsheet shown above. Parameters which will be changed include number of trays in the stripper, steam rate in the stripper, and the H_2S/CO_2 feed ratio.

COMPARISON OF PROGRAM RESULTS TO EXPERIMENTAL DATA

The key parameter which concerns most users is the residual acid gas concentration leaving the bottom of the stripper. If the residual acid gas concentration, called the lean amine concentration, rises above a certain point, the H_2S concentration in the sweet gas will be too high to meet pipeline specifications.

In addition, each unit of acid gas recycled within the amine solution reduces the ability of the solution to pick up acid gas from the sour gas stream. This means that more solution must be circulated to remove the same mass of acid gases. There are two methods which can be used to reduce the residual acid gas concentration.

The most common method is to increase the stripping steam rate at the reboiler. This trades the expense of increased heat duties in both the reboiler and the condenser with the expense of increased amine circulation, which results in increased pumping costs, larger cross exchanger and trim cooler, and larger absorber.

In the design stage, a possibly cheaper method of reducing the residual acid gases is to increase the number of trays in the stripper. However, this option has not been pursued with much enthusiasm, because it was not possible to predict the incremental performance of an extra tray.

For this reason, most users have been content to continue using the 22 tray stripper, since this configuration has shown itself to be capable of effecting sufficient stripping at approximately 1.0 lb of steam per gallon of lean amine to produce pipeline specification gas for most sour gasses.

This program will show the incremental performance of adding or removing ideal stages to the stripper column, which according to the model can significantly reduce the stripping steam and circulation rate simultaneously with the equipment size.

Our test cases were taken from data collected and published in Oil and Gas Journal by Fitzgerald and Richardson (1966). The paper contains data on the residual H_2S concentrations for a number of plants using varying H_2S/CO_2 ratios, and varying steam rates. The paper also contains correlations for residual H_2S and CO_2 concentrations based on plant data. This data can thus be used to check the program.

Our first test case was chosen because it was reasonably smooth, had several data points, and was near the center of the residual H_2S chart. Assumed operating conditions which were not changed were:

1. Rich amine fed to the stripper at 220^{o}F.
2. Absorber overhead at 110^{o}F and 900 psig.

3. Stripper overhead pressure was 12 psig with a dP of 2.5 psi through the column. This made the reboiler pressure 14.5 psig and reboiler temperature 252°F.
4. The flow measurement for both the lbs steam per gallon lean amine and for the grains acid gas per gallon lean amine were based on the volume of amine solution at the reboiler bottom.
5. One pound of steam was regarded as 945 BTUs.

Assumption 4 was particularly troubling. Since the horizontal scale of the figure is so greatly expanded, changing the base volume from the reboiler bottoms to the absorber feed can move the program results from, for example, 0.80 lb/gallon to 0.85 lb/gallon. Accordingly, this gives additional uncertainty in the quoted data.

Figure 4 shows the residual acid gas concentrations from actual plant data, from the Fitzgerald and Richardson (1966) correlations based upon that data, and from the program. The lower collection of lines running from about 300 grains per gallon down to 30 grains per gallon are for H_2S and the higher set of lines are for CO_2.

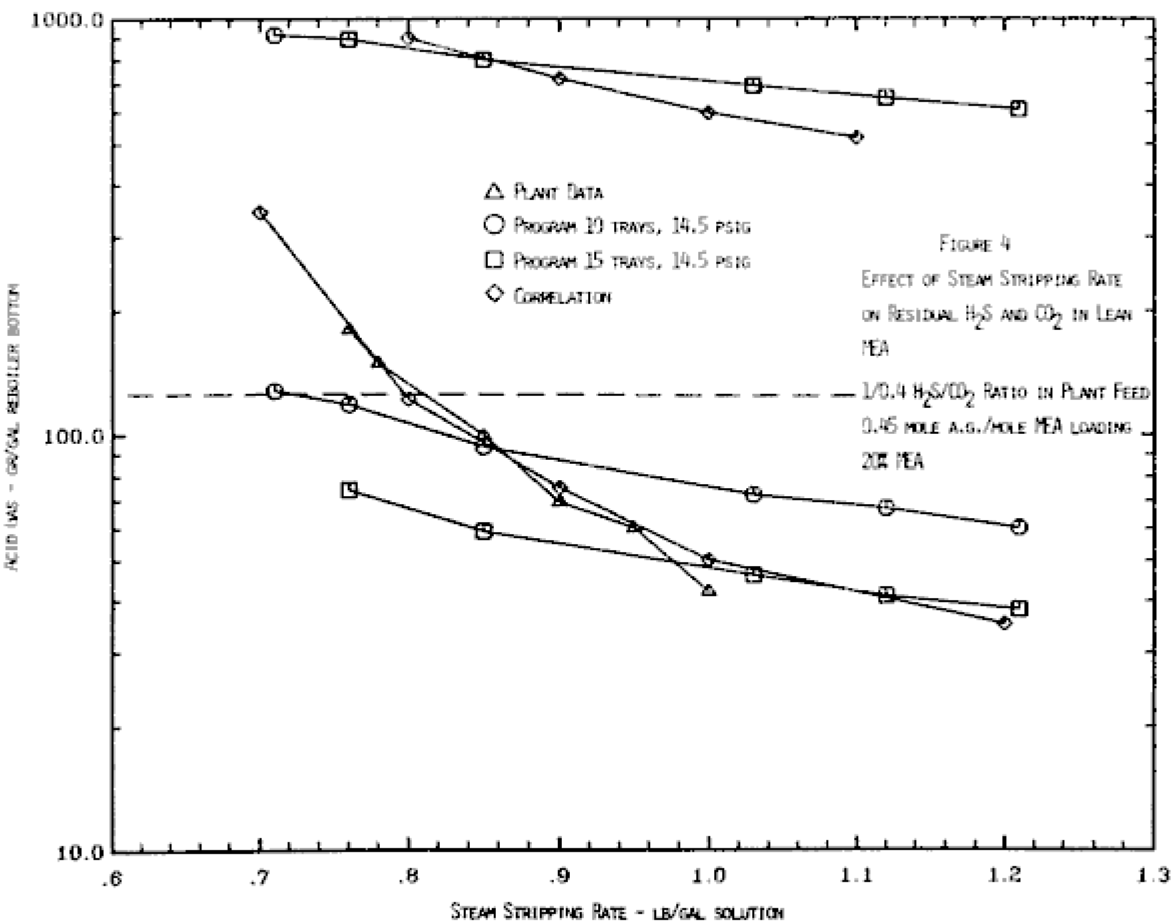

Figure 4. Effect of steam stripping rate on residual H2S anc CO2 in lean MEA.

The dotted line at 120 grains per gallon is the equilibrium point corresponding to .25 grains H_2S per 100 standard cubic feet of gas at 900 psig and 110°F. This equilibrium point is predicted both by Fitzgerald and Richardson and by the program. The correlation matches the plant data very closely as expected, since the correlation is based on this data.

Another obvious point is that the CO_2 residual concentration is almost independent of the number of stages in this plant. The points from a 10 ideal tray column fall on the same line as those from a 15 ideal tray column and both sets of points are nearly coincident with the correlation predictions.

No actual plant data was quoted for CO_2. The third item which should be noted from this figure is that the program predicts less slope on the residual lines than does the plant data. This was expected.

In an actual plant, the trays are designed for maximum efficiency at specific liquid and vapor flow rates. As the flows through the tray change from the design values, the efficiencies change, usually for the worse.

For instance, in the 15 tray example, the vapor rate into the bottom tray is 20% less for the case at .76 lb steam per gallon of lean amine than it is at 1.12 lb steam per gallon while the liquid rate into the tray decreases by 2%. Furthermore, the change in the vapor rate increases from tray to tray up the column.

Figure 5 shows the same information for a plant handling a very sour gas stream. The inlet gas to this plant contained 11 moles H_2S per mole of CO_2. As can be seen, the CO_2 residual concentration dropped while the H_2S residual concentration rose to roughly double the previous plant values.

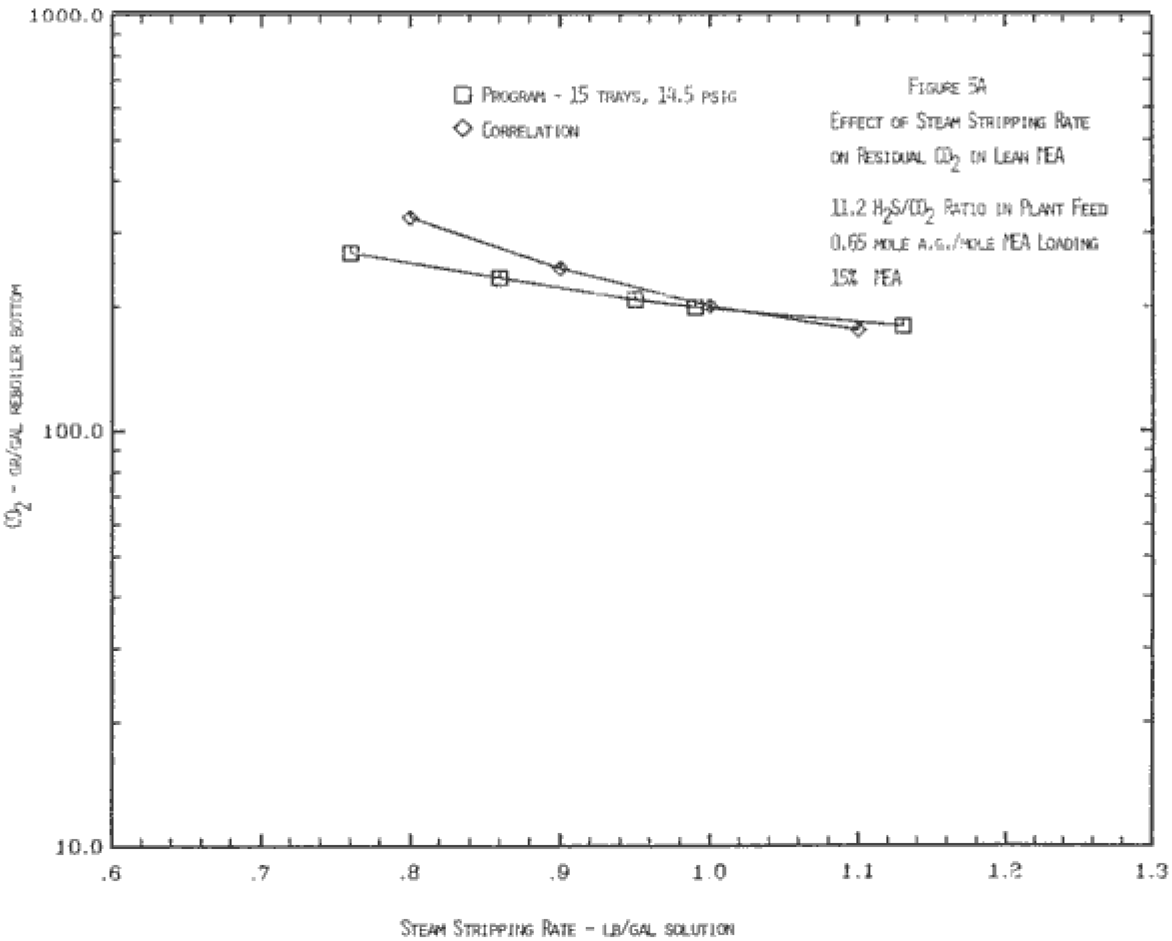

Figure 5a. Effect of steam stripping rate on residual CO2 in lean MEA.

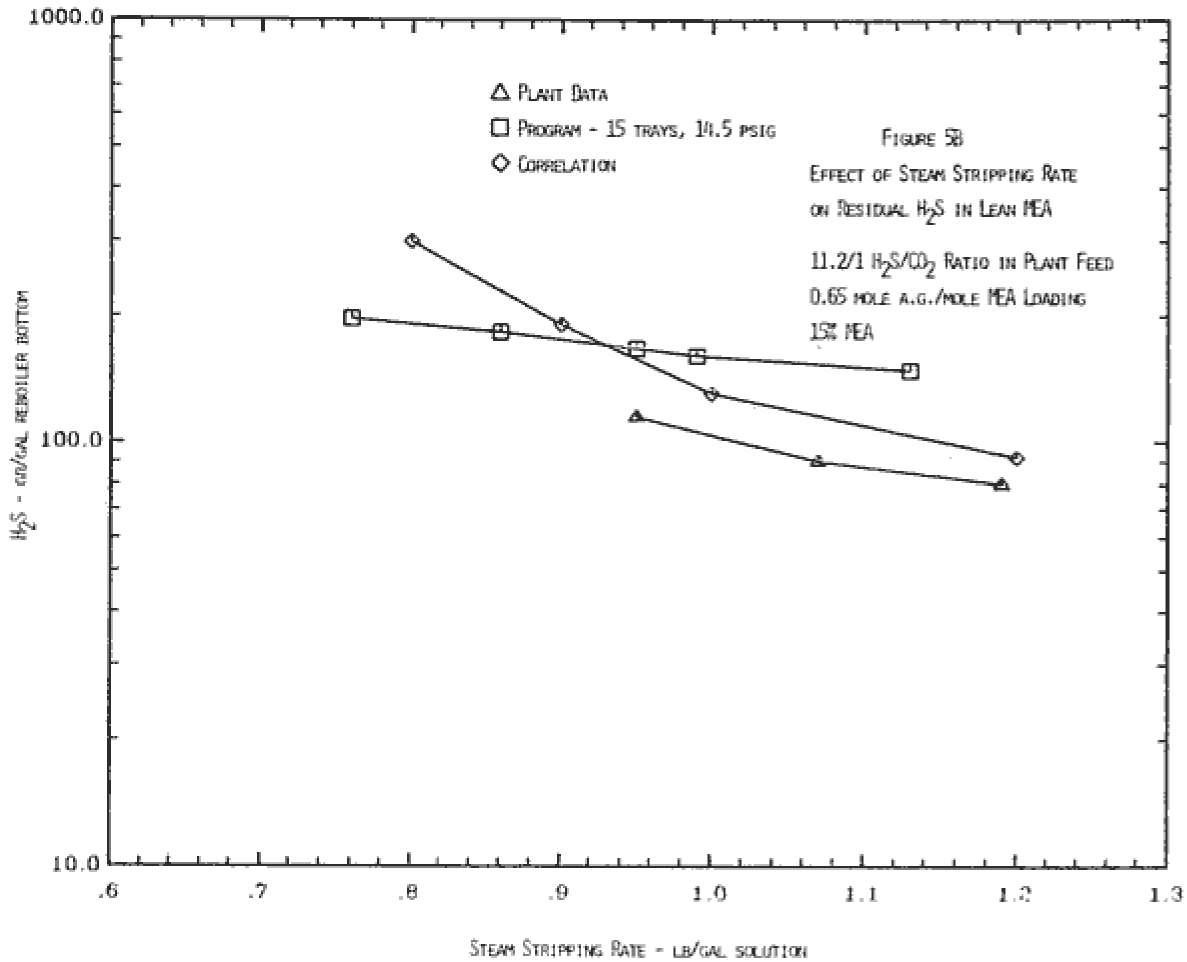

Figure 5b. Effect of steam stripping rate on residual H2S in lean MEA.

Again, the program results displayed flatter profiles than the actual plant data. Also note that, for 15 ideal trays, the predicted H_2S concentration was high, while the predicted CO_2 concentration was low. Also note that this plant performed better than the correlation predicted.

Figure 6 shows the program results for a plant handling exactly the opposite tvpe of stream. The feed gas to this plant contains 11 moles of CO_2 for each mole of H_2S. If this plant used the standard 22 tray configuration in the stripper, either the tray efficiencies were in the vicinity of 30% at best, or the analytical techniques used for H_2S were not as good as they could have been.

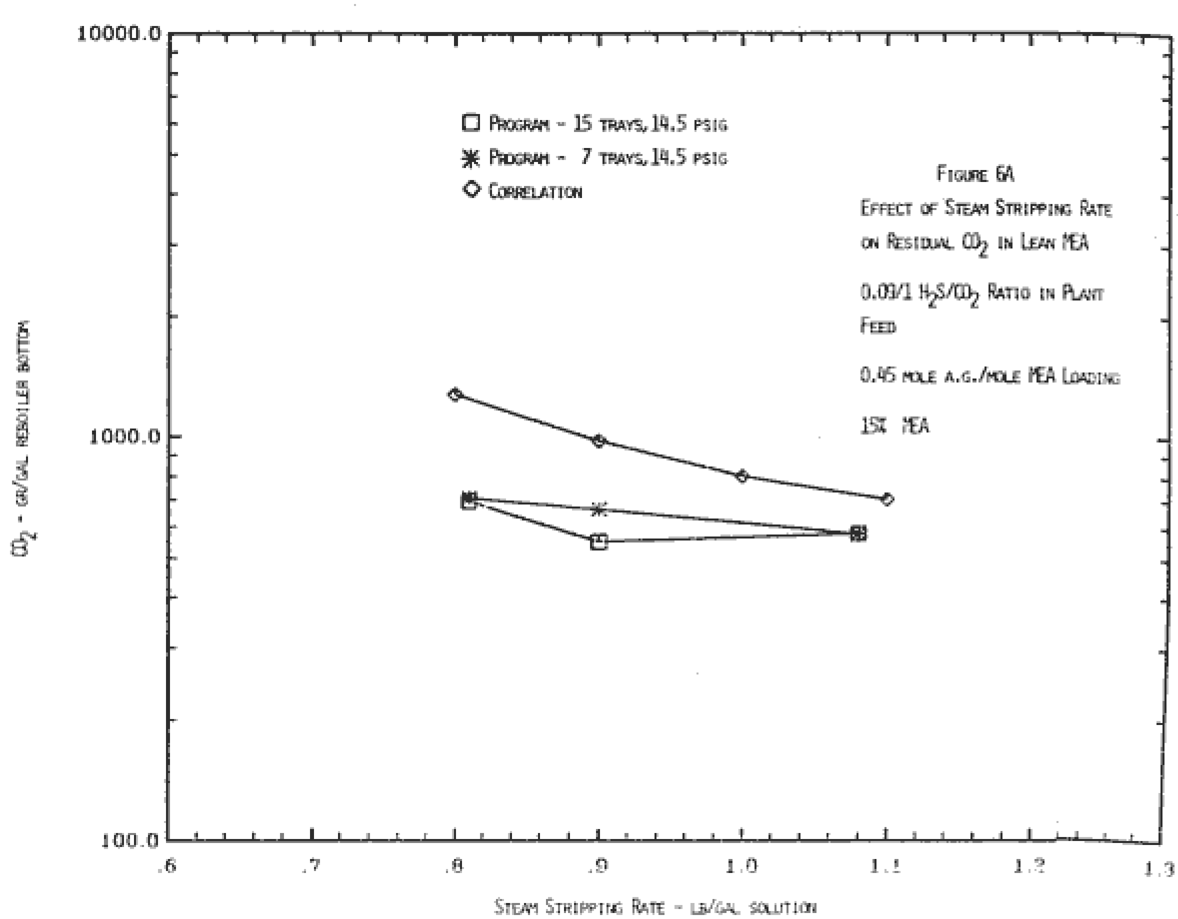

Figure 6a. Effect of steam stripping rate on residual CO2 in lean MEA.

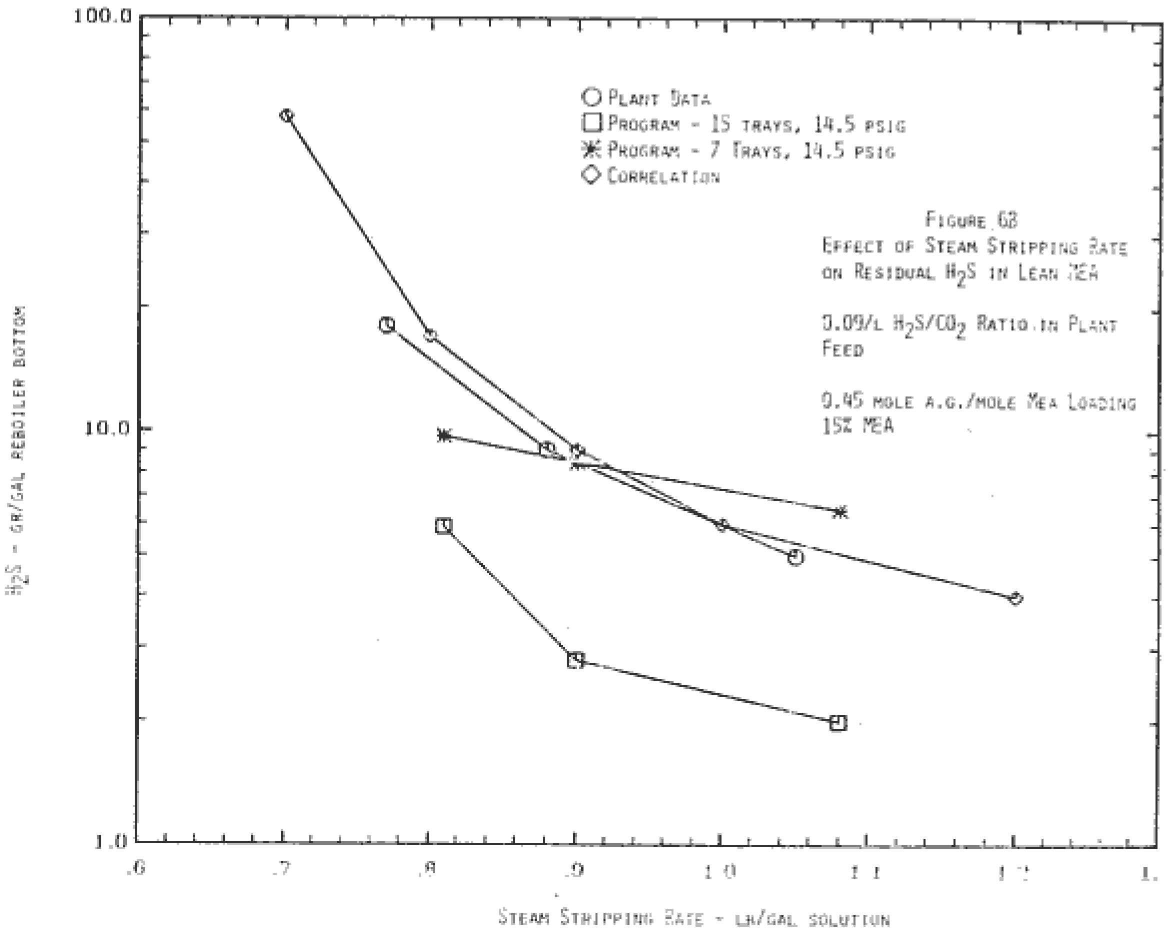

Figure 6b. Effect of steam stripping rate on residual H2S in lean MEA.

At 7 ideal stages, the program prediction line crosses the observed plant H_2S data at about .9 lb steam per gallon of lean amine, while the CO_2 prediction line is well below the correlation line.

SWEETENING PLANT OPTIMIZATION

The above example cases can also be examined to determine the optimum design. Tables 1 and 2 may be compared directly to see the savings which can be realized by increasing the number of trays in the stripper. Table 1 shows six different steam rates for the 10 ideal tray column, while Table 2 shows the same information for the 15 ideal tray column.

Table 1
2.5 H2S/CO2 Ratio; .45 Mole Acid Gas/Mole MEA; 20% MEA; 10 Trays

Steam Rate lb/gal	Absr. Dia. ft	Circ Rate gpm	Pump hp	Cross Exch. MMBTU/hr	Trim Cooler MMBTU/hr	Still Dia. ft	Reboiler MMBTU/hr	Condenser MMBTU/hr	Residual H2S gr/gal	Residual CO2 gr/gal	H2S Overhead gr/100scf
.71	4.19	324	268	14.5	6.6	5.83	13.1	5.3	128	915	.31
.76	4.18	316	261	14.1	6.5	5.88	13.6	5.8	119	895	.27
.85	4.14	300	249	13.3	6.3	5.95	14.5	6.9	94	800	.18
1.03	4.11	285	236	12.6	6.1	6.19	16.7	9.3	72	690	.11
1.12	4.09	280	232	12.3	6.1	6.31	17.9	10.5	67	645	.09
1.21	4.09	276	228	12.1	6.0	6.43	19.0	11.7	60	605	.08

Table 2
2.5 H2S/CO2 Ratio; .45 Mole Acid Gas/Mole MEA; 20% MEA; 15 Trays

Steam Rate lb/gal	Absr. Dia. ft	Circ Rate gpm	Pump hp	Cross Exch. MMBTU/hr	Trim Cooler MMBTU/hr	Still Dia. ft	Reboiler MMBTU/hr	Condenser MMBTU/hr	Residual H2S gr/gal	Residual CO2 gr/gal	H2S Overhead gr/100scf
.76	4.16	308	254	13.7	6.4	5.81	13.7	5.6	74	895	.16
.85	4.13	295	244	13.1	6.3	5.91	14.3	6.7	59	800	.11
1.03	4.10	282	233	12.4	6.1	6.15	16.5	9.1	46	690	.07
1.12	4.09	276	229	12.1	6.0	6.28	17.6	10.3	41	645	.06
1.21	4.08	273	226	12.0	6.0	6.41	18.8	11.5	38	605	.05

The item which limits the reduction in steam in this plant is the H_2S concentration in the sweet gas. It can be observed that the 0.76 lb steam/gallon case is well below the accepted limit. The optimum plant for the 10 tray column is the .85 lb/gallon plant.

The reboiler and condenser heat duties are minimized while the plant still produces pipeline specification gas and the absorber size is not much affected by steam rate. The trim cooler and cross exchanger are slightly larger, but unless some overriding consideration increases the prices of these items of equipment, the heat duties mentioned should be the determining factor in the cost of plant operation.

For the 15 tray column, the optimum operation point is the .76 Ib/gallon case. Comparing the two optimums, it can be seen that they are almost identical except for two items. The reboiler and condenser heat duties are markedly lower in the 15 tray case. The user would save some .80 million BTUs per hour in steam fed to the reboiler, and reduce the cooling load in the condenser by a like amount.

This means roughly a savings of 125 BTUs per pound of acid gas recovered. The cash savings represented by this are highly individual to each plant. If 50 million BTU per hour are being generated by a Claus sulfur plant on site, this steam decrease may be worthless. If, however, clean gas must be burned to generate the reboiler steam, the cost could be very important. Table 3 shows how the program can be used to determine the minimum steam rate required for a given number of trays. This plant is also limited by the necessity of meeting pipeline specifications at the top of the absorber. Looking to the far right column, it can be seen that a steam rate of at least .85 lb/gallon of steam is necessary.

Table 3
11.2 H2S/CO2 Ratio; .65 Mole Acid Gas/Mole MEA; 15% MEA; 15 Trays

Steam Rate lb/gal	Absr. Dia. ft	Circ Rate gpm	Pump hp	Cross Exch. MMBTU/hr	Trim Cooler MMBTU/hr	Still Dia. ft	Reboiler MMBTU/hr	Condenser MMBTU/hr	Residual H2S gr/gal	Residual CO2 gr/gal	H2S Overhead gr/100scf
.76	3.85	184.6	153	8.68	3.66	4.51	8.00	3.11	197	268	.253
.86	3.85	181.7	150	8.53	3.63	4.65	8.83	3.94	183	233	.209
.95	3.84	179.1	148	8.40	3.61	4.77	9.63	4.77	167	207	.171
.99	3.84	178.2	148	8.36	3.60	4.83	10.05	5.19	160	198	.157
1.13	3.84	176.4	147	8.26	3.58	5.02	11.31	6.41	148	179	.131

Little savings in size are realized by going upwards in steam stripping rate, and the heat duties and stripper diameter increase dramatically. Therefore, the optimum point to operate this plant is at .85 to .90 Ib steam per gallon of reboiler bottoms.

The last plant carrys a different limiting factor in its optimization. Because of the large amount of CO_2 in the stream, the H_2S carrying capacity of the amine is greatly enhanced. Pipeline quality gas can easily be generated by any reasonable steam rate.

The savings in this plant can be achieved by reducing the number of trays in the stripper to 3 or 4 ideal trays. Table 4 shows the effect of decreasing the number of ideal stages in the stripper at a steam rate of 1.08 lb/gallon reboiler bottoms. From this table it can be seen that as the number of stages decreases, the pump horsepower, equipment size, and heat duties slowly increase while the H_2S overhead in the absorber rises.

Table 4
0.9 H2S/CO2 Ratio; .45 Mole Acid Gas/Mole MEA; 15% MEA; 1.08 Steam/Gallon

No. of Ideal Trays	Absr. Dia. ft	Circ Rate gpm	Pump hp	Cross Exch. MMBTU/hr	Trim Cooler MMBTU/hr	Still Dia. ft	Reboiler MMBTU/hr	Condenser MMBTU/hr	Residual H2S gr/gal	Residual CO2 gr/gal	H2S Overhead gr/100scf
15	4.03	254.4	211	10.9	6.09	5.94	15.6	8.24	2.0	495	.003
10	4.04	258.5	215	11.2	6.16	5.98	15.8	8.45	3.9	529	.006
7	4.06	265.0	220	11.5	6.24	4.06	16.2	8.78	6.5	574	.010

However, even at 7 ideal stages, the overhead concentration is about 5% of the maximum allowed. This implies that equipment sizes will become economically prohibitive long before the plant becomes unable to meet pipeline specifications.

The point at which this occurs is a strong function of the individual plant location, ease of maintenance, availability of heat, discounts from equipment vendors, and other factors which the individual designer and user must decide upon.

The program can also be used to analyze existing plants. The user can input the actual operating conditions of the plant and see how close the required equipment sizes are to the actual field equipment. The user can then vary parameters to attempt to reduce or increase whatever parameter he wishes to optimize on.

In addition, if at actual operating conditions, a piece of equipment is grossly oversized or undersized, the user could possibly change the equipment to bring it more into line with the most efficient plant. Thus, the program has three main uses with regards to existing plants.

First, the program can be used to experiment with the plant without causing process upsets or damaging equipment. Second, the program can be used to isolate bottlenecks or massively oversized pieces of equipment, locating the optimization points in a marginal plant. Third, the program can evaluate the results of an amine or a feed gas change without actually making the change.

SUMMARY AND CONCLUSIONS

The simulation program can be a very useful tool in the design and analysis of amine sweetening plants. The plants can be optimized based on both the utility and equipment size requirements. In addition, process problems as well as plant optimization may be analyzed for existing plants. Especially, in cases where significant carbon dioxide is present in the sour gas, the strippers are oversized and the utility requirements are much larger than required. Thus the simulation program using tray to tray type calculations can significantly reduce both the capital and operating costs for amine plants.

