Design Glycol Units for Maximum Efficiency

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ABSTRACT

In designing dehydration units for natural gas, several critical parameters exist which can be varied to achieve a specified dew point depression. This paper studies the effects of varying the glycol flow rate, number of stages in the contactor, reboiler temperature, and stripping gas rate on water content in glycol dehydration units. The effect of high carbon dioxide composition in the feed is also presented. Finally, the emissions of aromatic (BTEX) and other VOC's from the regenerator and flash will be considered.

INTRODUCTION

A common method to remove water from natural gas is glycol dehydration.[1] In this process, triethylene glycol (TEG) or diethylene glycol (DEG) is used to remove the presence of water in the gas stream. Water vapor can cause hydrate formation at low temperatures and high pressures or corrosion when it is in contact with hydrogen sulfide (H_2S) or carbon dioxide (CO_2), components regularly present in the gas stream.

Glycol dehydration units are typically represented by a contactor, a flash tank, heat exchangers, and a regenerator, as shown in Figure 1. The glycol, usually TEG, enters at the top of the contactor and absorbs water as it progresses toward the bottom of the column. A dry gas exits at the top of the contactor and may be used for cooling the incoming lean glycol.
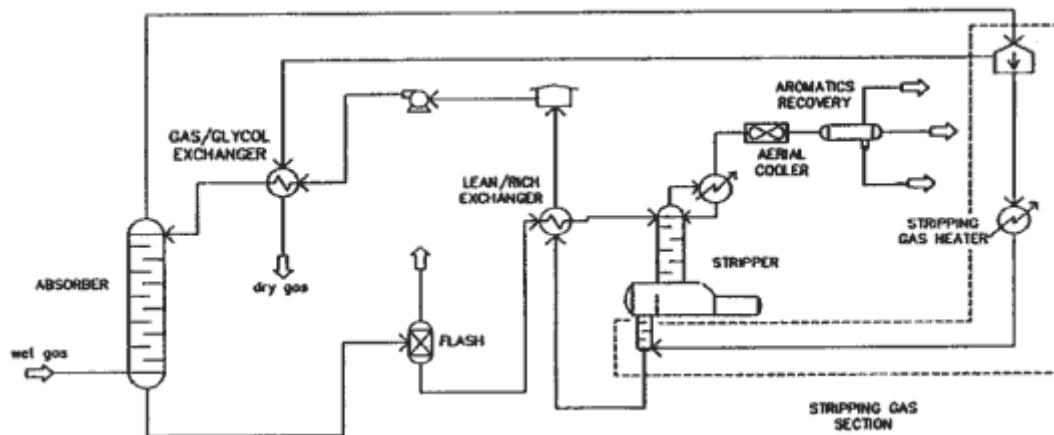
Figure 1. Typical glycol dehydration unit.

The rich stream flows to a separator or flash tank where gaseous hydrocarbons that were absorbed along with some of the water in the contactor are liberated and used as fuel. Finally, the glycol flows to the stripper where it is regenerated by boiling off the water and returned to the contactor. For processes requiring gas with very low water dew points, a stripping vapor will most likely be needed to aid the regeneration process. This technique is illustrated by the region enclosed in the dotted line in Figure 1. For maximum stripping, this vapor is normally injected into a short column at the bottom of the reboiler. However, the gas may also be introduced directly into the reboiler.

In the past years, glycol dehydration plants have been designed using rule-of-thumb procedures. While still sufficient for many applications, today more efficient designs are often required. In many cases the plant feed will contain small quantities of aromatic hydrocarbons that are quite soluble in the TEG. The aromatics are primarily comprised of benzene, toluene, ethylbenzene, or xylenes (BTEX). These aromatics are carried to the flash tank where a small fraction are released along with other volatile organic compounds (VOC's). The remaining VOC's and aromatics travel to the regenerator where the application of heat will remove virtually all volatile gases. Since the regenerator is usually vented to the atmosphere, the plant may have serious environmental impact, even in small plants where the total aromatic emissions can easily exceed 100 lb/day.[2]

Other problems have also caused special concern in glycol dehydrator design over the past few years. Enhanced oil recovery using CO_2 and dehydration of refinery gases have required glycol units that can properly dehydrate gases containing a few percent CO_2 to pure CO_2. In addition to the effect on dehydration, high quantities of CO_2 in the feed can accelerate corrosion in the regenerator. On the other hand, the CO_2 can act as a stripping vapor in the regenerator.[1]

This paper presents the progress of current work to modernize simulation software used in design and optimization of dehydration units. The results provide an analysis of the dehydration effectiveness at a variety of common operating variables for a typical dehydration facility. Next, the effect of carbon dioxide in the plant feed will be presented. Finally, emission calculations of aromatic and non-aromatic hydrocarbons absorbed in the glycol contactor are presented.

PROCESS DESCRIPTION

All calculations provided in this report are based upon calculations made with Bryan Research & Engineering's general purpose process simulator, PROSIM. The results are obtained in an effort to update and expand the capabilities of the BR&E dehydration program, DEHY. PROSIM allows the user to draw the process flow diagram on the computer screen and enter the operating parameters on "pop-up" forms. The program is flexible to handle almost any glycol dehydration scheme. Additionally, the program has heavy ends/crude characterization, complex heat exchanger, tray rating, and a variety of utility calculational operations. Dehydration results obtained by PROSIM have been compared to plant data obtained by Worley in the GPSA Engineering Data Book and other pertinent experimental vapor-liquid equilibrium data as noted throughout this report.[3]

In order to provide the results needed for this paper, a base set of operating parameters was selected as presented in Table I. Except when noted, these parameters are held constant throughout the analyses presented here. The process flow scheme is basically that presented in Figure 1 except for cases where stripping gas was not utilized. In all runs, the amount of water in the feed was determined by PROSIM so that the entering gas would be at the water dew point.

Table I

Base dehydration unit operation parameters used in this analysis.

Inlet gas temperature	90°F
Inlet gas pressure	500 psia
Inlet gas composition:	
Methane	85.1 mol%
Ethane	8.5 mol%
Propane	3.8 mol%
n-Butane	1.9 mol%

n-Pentane	0.7 mol%
Lean glycol temperature	90°F
Rich glycol flash pressure	65 psia
Regenerator pressure	1 atm
Equilibrium trays in contactor	2

DESIGN OF DEHYDRATION UNITS

When optimizing the design of dehydration facilities, the impact of the following parameters should normally be considered:

- Number of trays in the contactor
- Glycol circulation rate through the contactor
- Temperature of the reboiler in the regenerator
- Amount of stripping gas used, if any
- Operating pressure of the regenerator

Of the above parameters, only the first four are normally considered as variable parameters. The first two parameters affect the approach to equilibrium at the top of the absorber while the third and fourth dictate the value of the equilibrium water content by limiting the purity of the lean glycol to the absorber. The last parameter affects the lean glycol purity in a manner similar to reboiler temperature. However, the vast majority of units are vented to the atmosphere so this parameter is beyond control.

In addition to the design parameters listed above, several other factors influence the residual water content of the sales gas. However, often these factors are fixed and cannot normally be changed when optimizing a unit. First, the temperature of the inlet gas will dictate the total amount of water fed to the unit. Lower plant inlet gas temperatures will require less water to be removed by the glycol. Second, lean glycol temperature at the top of the contactor will affect the water partial pressure at the top stage. Consequently, high glycol temperatures will result in high water content in the overhead gas. However, this temperature is normally no cooler than 10°F above the inlet gas to prevent hydrocarbons in the feed from condensing in the solution. This limit is normally maintained by a gas/glycol exchanger that cools the lean glycol to approximately a 10 °F approach using the dry gas.

Other parameters in the plant have limited or no effect on the dry gas water content. The number of equilibrium stages in the regenerator has only a slight effect on the lean glycol purity. Equilibrium at the reboiler temperature and pressure is approached in the reboiler so that additional stages have no effect. Operating temperature of the lean/rich glycol exchanger only significantly impacts the reboiler heat duty.

ANALYSIS OF PERTINENT DESIGN PARAMETERS AND OTHER FACTORS

In order to investigate the important design variables, plots of the residual water content versus circulation rate are presented in Figures 2 to 6 for common values of the parameters discussed in the previous section.

Case 1. Effect of Number of Equilibrium Stages in Absorber

Figure 2 illustrates the effect of the number of equilibrium contact trays on residual water content using a 400 °F reboiler. Figure 3 presents a similar figure comparing dew point depressions instead of actual water content. Increasing the number of trays allows the gas to approach equilibrium with the lean glycol at a lower glycol circulation rate. Considering a typical glycol circulation rate of approximately 3 gal TEG/lb water removed, Figures 2 and 3 illustrate that a three equilibrium-stage contactor is virtually at equilibrium with the inlet glycol. In a two stage contactor, a circulation rate of 5 to 6 gal TEG/1b water would be required to approach equilibrium. Significantly higher flow rates would still be required when only one ideal stage is used.
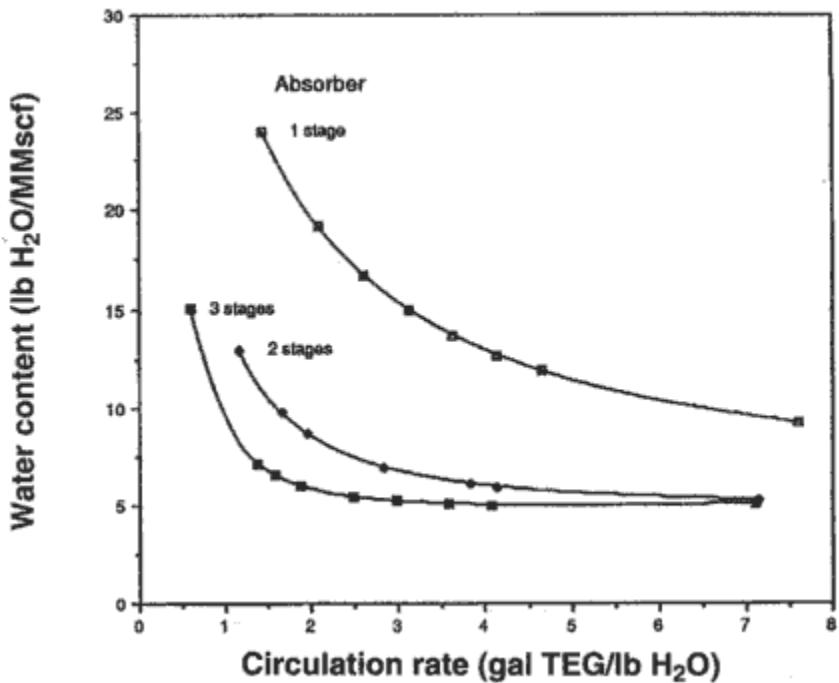
Figure 2. Effect of the number of equilibrium stages in the contactor on the water content of a stream
of sweet natural gas. Reboiler temperature 400°F.
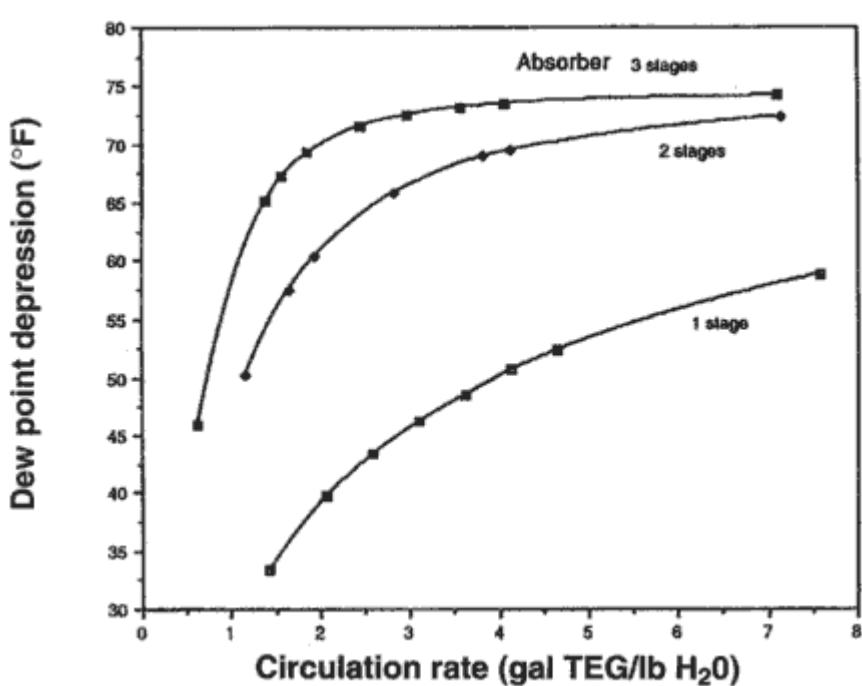
Figure 3. Effect of the number of equilibrium stages in the contactor on the dew point depression of
natural gas. Reboiler temperature 400°F.

Case 2. Effect of Reboiler Temperature

Figures 4 to 6 illustrate the overhead water content using a fixed number of equilibrium stages at reboiler
temperatures of 360, 380, and 400 °F. For two or three equilibrium stages, pipeline quality gas containing less
than 7 lb water/MMscf gas could be produced using either a 380 or 400 °F reboiler. At 380 °F, approximately 4.5

gal TEG/1b water circulation would be needed with two stages as opposed to approximately 2 gal TEG/lb with three stages. Similarly, at 400 ºF, approximately 3 gal TEG/1b water circulation would be needed with two stages as opposed to 1.5 gal TEG/lb with three stages. Note that these results are for the inlet gas temperature of 90 ºF. Higher temperatures would result in higher residual moisture at the same circulation rate. The reboiler temperature influences the overhead water content by changing the purity of the lean glycol. Glycol purities of 98.0, 98.5, and 98.8 wt % are obtained at 360, 380, and 400 ºF, respectively, at one atmosphere pressure.
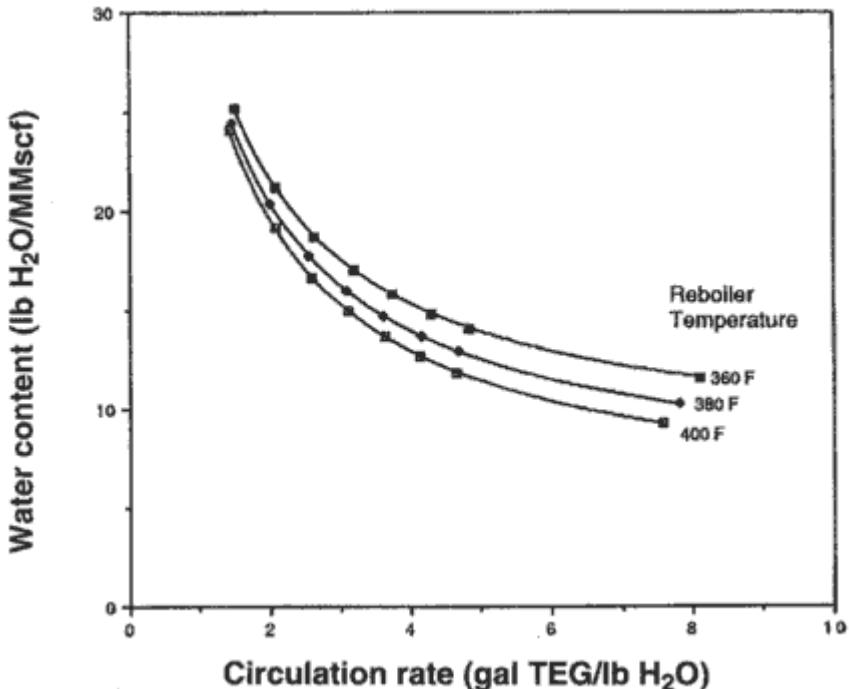
Figure 4. Water content for a natural gas stream treated by a one equilibrium-stage contactor.
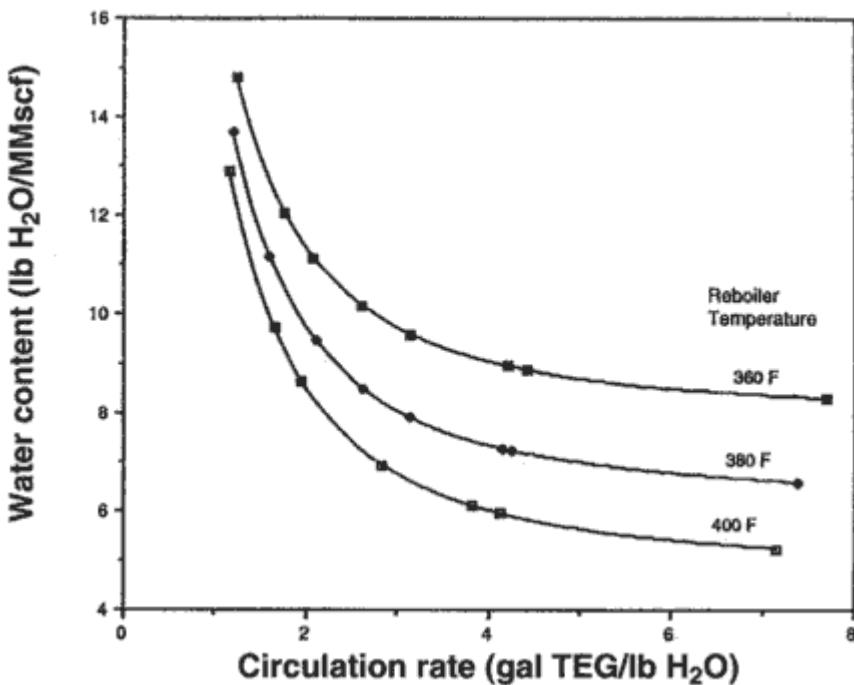
Figure 5. Water content for a natural gas stream treated by a two equilibrium-stage contactor.
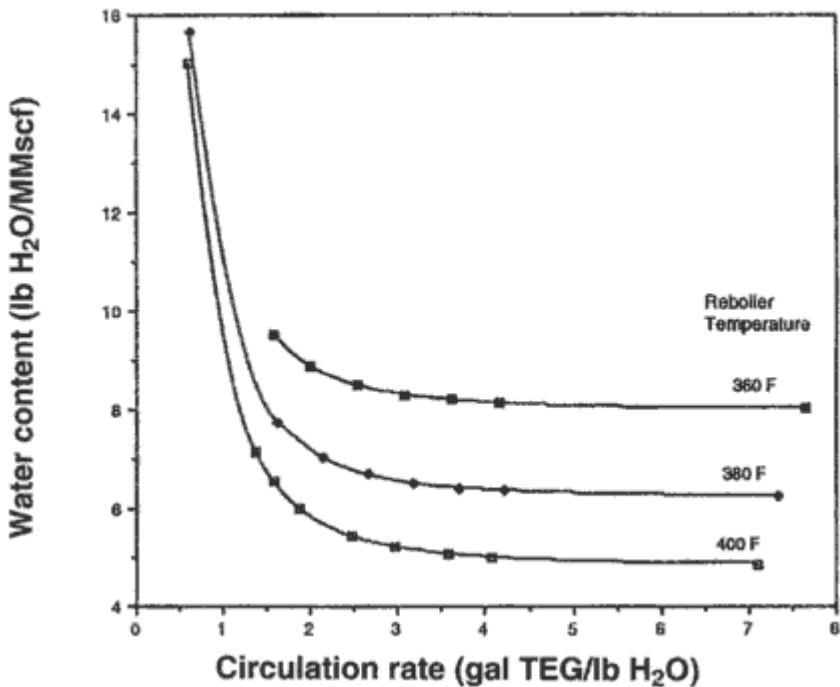
Figure 6. Water content for a natural gas stream treated by a three equilibrium-stage contactor.

Case 3. Effect of Stripping Gas

As stated earlier, applications requiring high dew point depressions will virtually always utilize stripping gas in the regenerator. Low dew points simply cannot be achieved using the maximum 98.8 wt % glycol obtainable with a 400 °F reboiler at atmospheric pressure. These low dew points will need up to 99.9 wt % glycol in the absorber. Increasing reboiler temperature is not an option due to the thermal decomposition temperature of 404 °F for TEG. Even a 400 °F reboiler can result in glycol decomposition due to higher film temperatures. Further, stripping gas has a much greater effect than increasing reboiler temperature. For maximum efficiency, stripping gas should be introduced in a short column after the hot glycol is removed from the reboiler. Stripping gas may be place directly in the reboiler, but the high water partial pressure in the vapor space limits the mass transfer driving force.

Figures 7 and 8 illustrate the effect of stripping gas on residual water content and dew point depression of the dry gas. As can be seen, even small stripping gas rates of 1 scf/gal circulated solution have a pronounced difference. With this stripping gas rate, the dry gas will contain about half the water of the same process without stripping gas. Increasing the stripping gas rate beyond 2 to 3 scf/gal will have little impact on dew point depression.
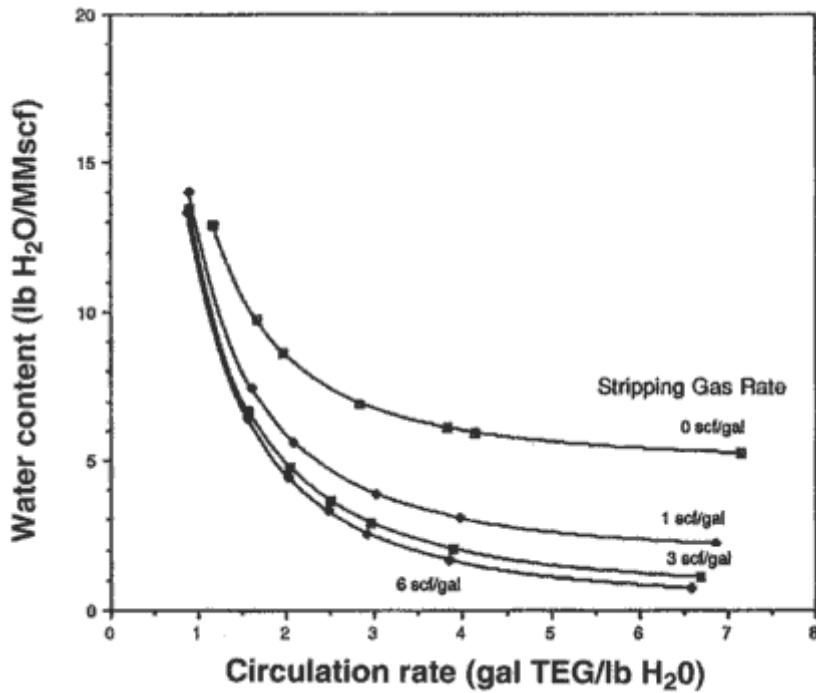
Figure 7. Effect of stripping gas rate on water content of a natural gas stream.
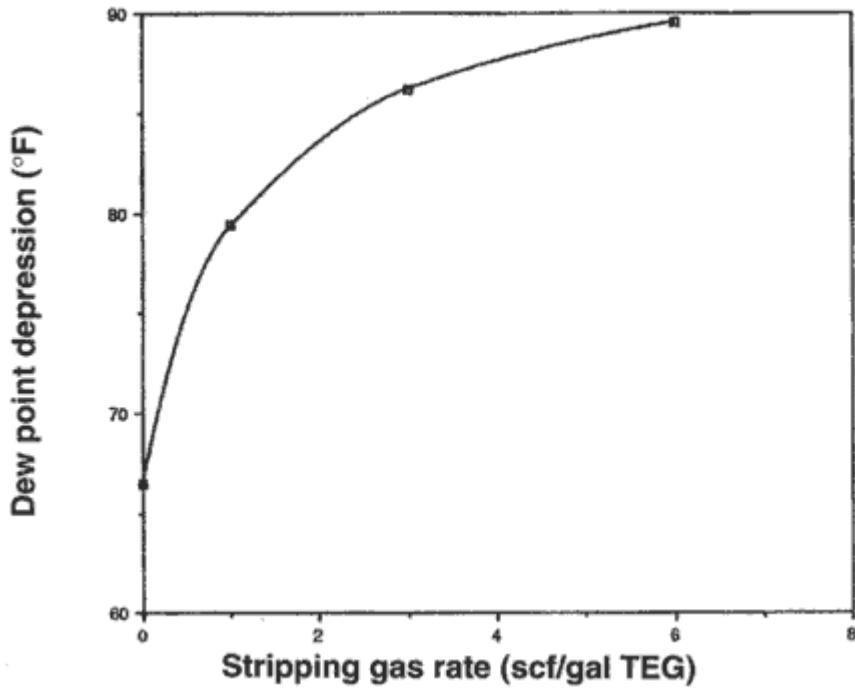
Figure 8. Effect of stripping gas rate on dew point depression, at three gal TEG/lb H_2O.

Case 4. Impact of Carbon Dioxide

In this phase of the work, a plant with significant CO_2 in the feed was modeled using PROSIM. The parameters in Table I were used as in the previous section but a total 50 mol % CO_2 was added to the plant feed. The amount of water in the feed was calculated by the program so that the gas remained at the dew point.

Figure 9 illustrates the effect of this CO_2 concentration on the dehydrated gas water content. The results indicate

that the addition of CO_2 slightly increased the water content about 1 lb water/MMscf gas at all glycol circulation rates. The most substantial data compilations for CO_2 - TEG - Water systems are that of Jou, Deshmukh, and Mather, and Takahashi and Kobayashi in GPA report TP-9.[4,5] In the GPA report, the water content of a natural gas mixture and of CO_2 is presented at various temperatures, pressures, and glycol concentrations. These data indicate that pure CO_2 water content above glycol solutions is slightly less than in natural gas at 865 psia and 565 psia. At 565 psia and 10 $^\circ$F, the difference in water content of CO_2 and natural gas above glycol solutions of less than 5 wt % water is insignificant. At 765 psia, the report indicates that the water content of CO_2 is greater than that of natural gas for temperatures greater than 75 to 100 $^\circ$F. For the most part, the difference in water content between pure CO_2 and natural gas is very small at conditions presented in the report. Therefore, the slight discrepancy between the program and the data is probably within the accuracy that can be expected for such complex systems. However, work is still being performed to investigate this difference.
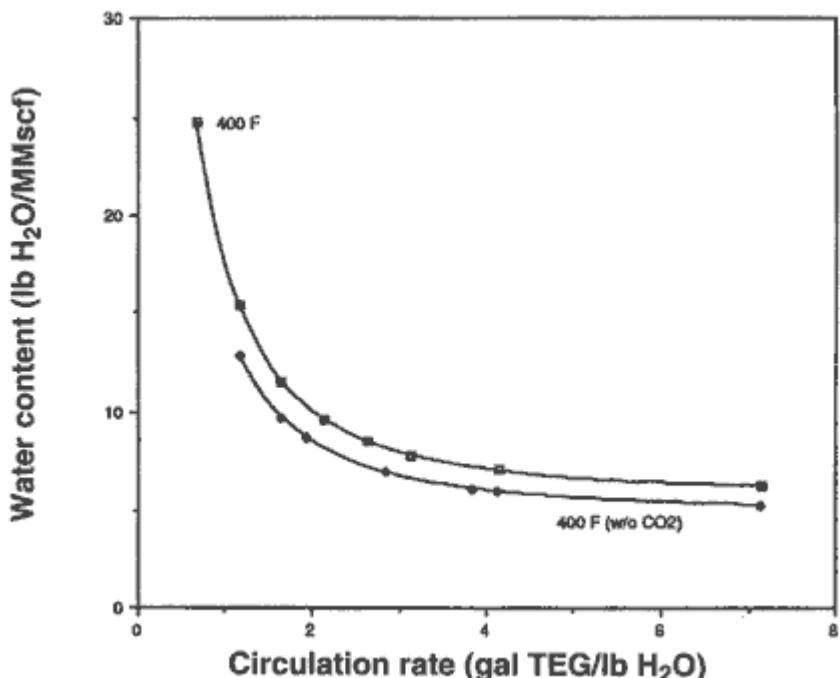
Figure 9. Water content of a natural gas stream with 50 mol % CO_2, treated by a two equilibrium-stage contactor.

Case 5. Aromatic and VOC Emissions in TEG Dehydration

In the last phase of the work, estimates of aromatic and VOC emissions from the regenerator and flash tank were calculated. According to other work, the total aromatic concentration in natural gas streams is normally well less than 0.05 to 0.1 mol %.[2] In order to give an approximation for the maximum amount of aromatics emitted per day, the parameters in Table I were used with the addition of 0.1 mol % total aromatic compounds. The actual ratios used for benzene:ethylbenzene, toluene:ethylbenzene, and o-xylene:ethylbenzene were 8:8, 13:3, and 12:6, respectively. These ratios were based on a feed analysis obtained for an operating plant. For this comparison, a feed flow rate of 20 MMscfd was taken.

The total non-aromatic and aromatic hydrocarbon emissions versus glycol circulation rate for the flash tank are presented in Figures 10 and 11, respectively, for contact pressures of 200, 500, and 900 psia. Figure 10 also includes the total non-aromatic emissions for the case with the same feed composition without aromatics at 500 psia to illustrate the slight increase in solubility in the presence of aromatics.
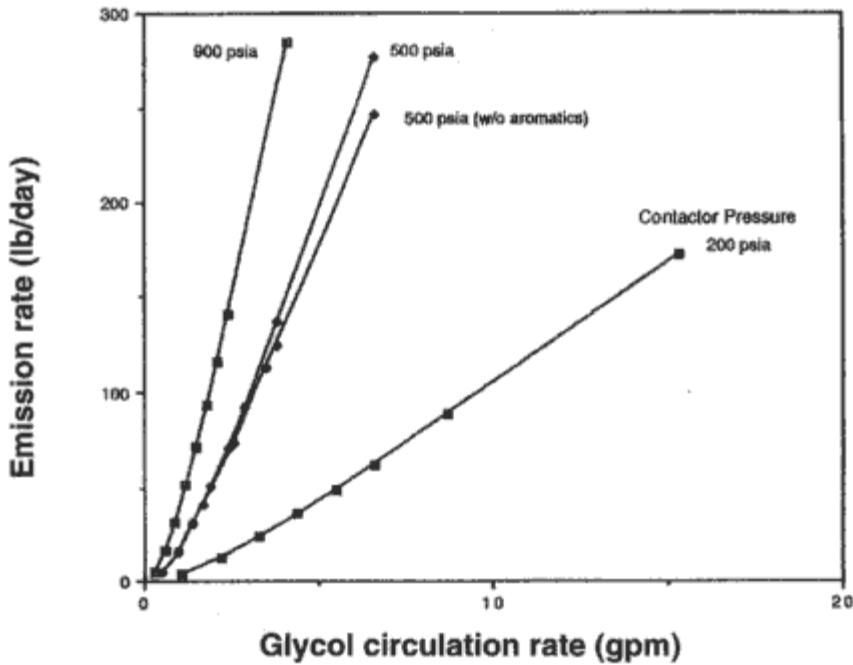
Figure 10. Total non-aromatic VOC emissions from flash tank. Contactor temperature 90°F.
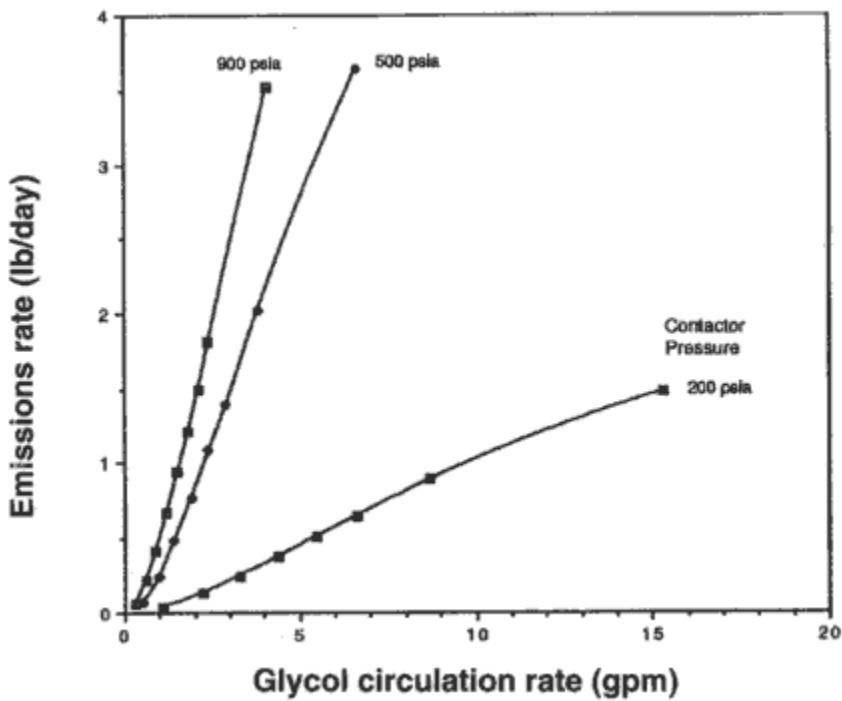
Figure 11. Total aromatic emissions at the flash tank. Contactor temperature 90°F.

The total non-aromatic emissions at the regenerator are presented in Figure 12. The emission rates for benzene, o-xylene, and total aromatics from the regenerator are given in Figures 13 to 15, respectively. Figure 13 indicates that the emission rate for benzene increases with glycol circulation rate and contact pressure. However, Figure 14 illustrates that the o-xylene emission rate at 900 psia and circulation rates greater than 3 gpm is smaller than at 500 psia for the same scenario. The only substantial data that exist on aromatic solubility in TEG are those of Robinson. In this compilation, the equilibrium K ratio (K=y/x) is shown to decrease with pressure for benzene and toluene as is the case with most volatile components. However, for ethylbenzene and o-xylene, the ratio is shown

to increase with pressure at certain temperature/glycol concentrations and to decrease with pressure at others. Most notable is that at temperatures in the range of the current scenario and relatively dry glycol concentrations where the K value increases with pressure. Since drier glycol would be present in higher circulation cases, the increase in K with pressure causes a decrease in solubility for ethylbenzene and o-xylene at high pressure. Since the aromatic portion of the feed is primarily o-xylene on a molar basis (and even higher on a mass basis), the total emission rate presented in Figure 15 will also exhibit this phenomenon at higher circulation rates.
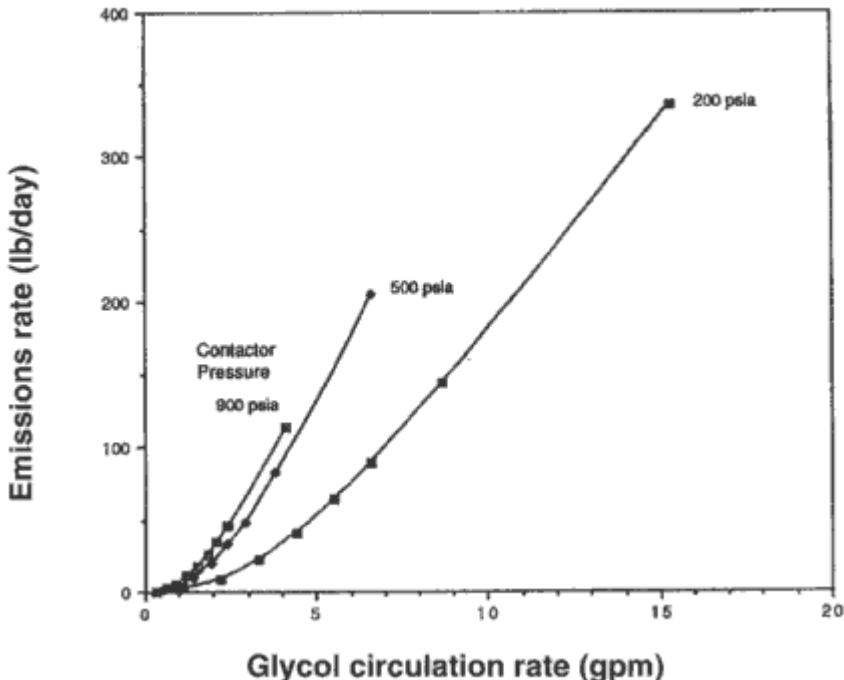
Figure 12. Total non-aromatic VOC emissions from regenerator. Contactor temperature 90°F.
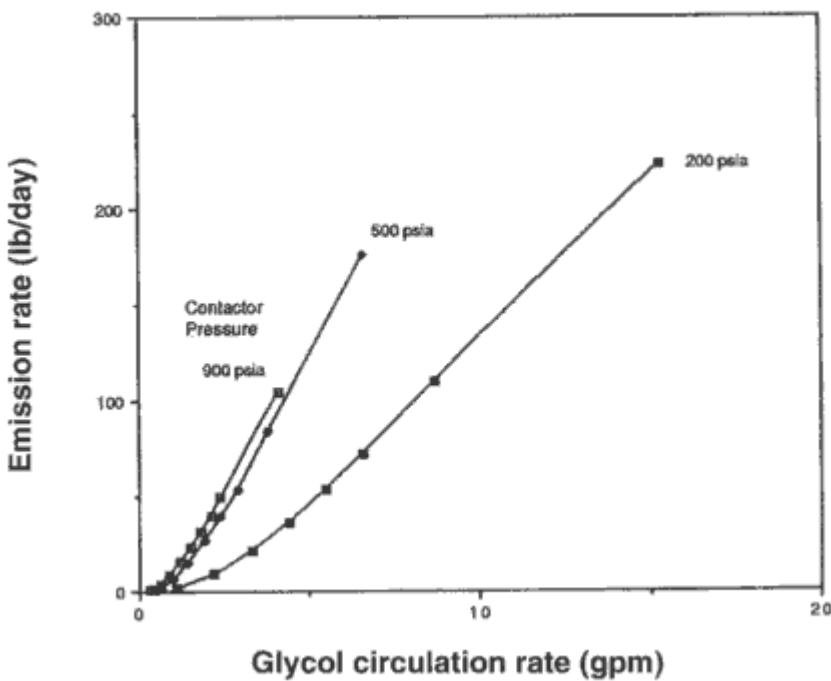
Figure 13. Benzene emissions at the regenerator. Contactor temperature 90°F.
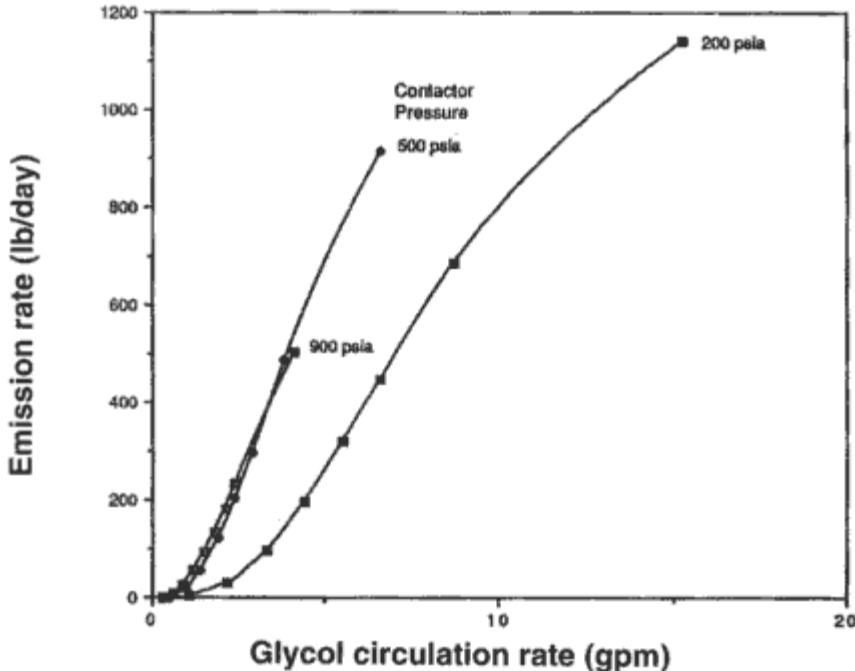
Figure 14. o-Xylene emissions at the regerator. Contactor temperature 90°F.
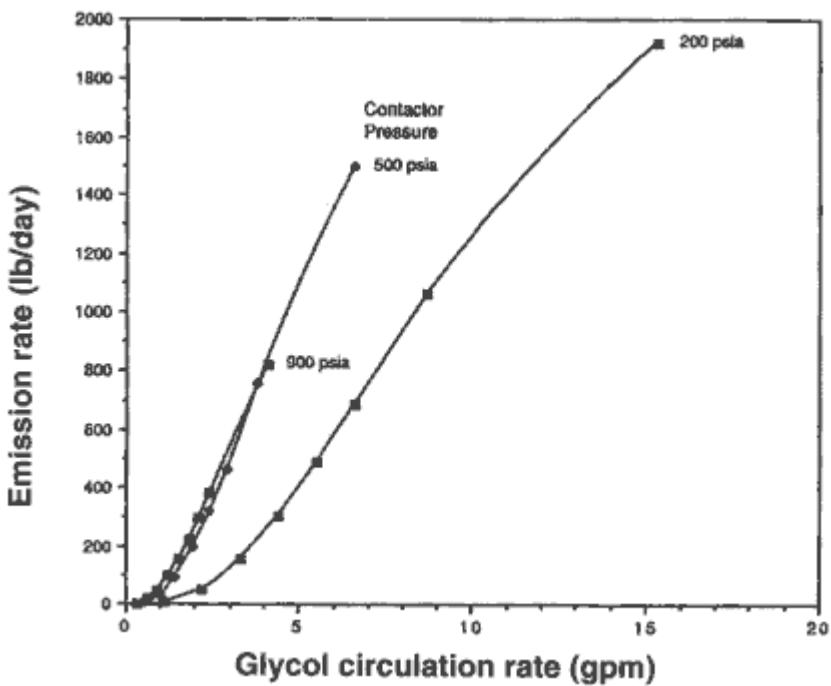
Figure 15. Total aromatic emissions at the regenerator. Contactor temperature 90°F.

These data also indicate that in systems with larger amounts of water in the glycol and higher temperatures (122 $^{\circ}$F), the K values for ethylbenzene and o-xylene decrease with pressure so that solubility would increase at the higher pressure range. Therefore, several cases at this temperature were modeled, and PROSIM indicated the solubility to increase with pressure except at high glycol circulation rates. Again, at these high circulation rates, the water content in the glycol would be much lower where the solubility decreases with pressure. Comparison of Figures 10, 11, 12, and 15, indicate that a significant fraction of the dissolved non-aromatic VOC's are released in the flash tank. However, only a small fraction of the aromatics will be liberated without the application of heat.

Special attention should be made to the fact that the majority of the aromatics released in the regenerator can be recovered using an aerial cooler operating about 120 °F and a three phase separator. Simple calculations yield that up to 97 % of the aromatics emitted at the high glycol circulation rates can be recovered. Often economics makes the return on this recovery system feasible.

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