

CAN STRIPPING REALLY OCCUR IN AMINE CONTACTORS?

*Laurance Reid Gas Conditions Conference
February 24-27, 2020 –Norman, Oklahoma USA*

Kyle Ross, PhD, PE
Bryan Research & Engineering, LLC
3131 Briarcrest Drive, Bryan, TX 77808
979.776.5220
k.ross@bre.com

Joel Cantrell, PhD
Bryan Research & Engineering, LLC
3131 Briarcrest Drive, Bryan, TX 77808
979.776.5220
j.cantrell@bre.com

ABSTRACT

Absorbers absorb. Strippers strip. It is what they do. But what happens if they don't? In discussions of amine contactors, the statement "Under the right conditions, stripping can occur" is sometimes heard. This has been said often enough that some people accept it as fact, while others dismiss the idea. We investigated this claim and did find that during co-absorption of both H₂S and CO₂, there can be movement of one of these two species from the amine into the vapor, and established conditions that cause it. As an example, we found that DEA can release H₂S in the when the feed gas also contains CO₂.

This paper will show when stripping cannot occur, explore potential causes of stripping in an absorber, and give examples of conditions that cause stripping.

Introduction

Amine absorbers use an amine and water solution to absorb unwanted acidic species such as H₂S and CO₂ from a vapor or liquid feed stream. The typical configuration has the acid-containing vapor or liquid stream flowing upward from the bottom of the absorber while the amine solution flows down countercurrently. With trayed absorbers, the countercurrently-flowing streams mix and then separate on each tray as the two streams pass by each other. With a packed absorber, mixing and separating occurs more or less continuously. In either case, a single mix and separate step is usually not sufficient to remove enough sour gas from the feed stream to meet product specifications, especially with a vapor feed stream. With a vapor feed, one will typically see somewhere around 20 trays or a depth of packing that gives similar performance.

Obviously, additional trays bring additional cost. If one could achieve the same sour gas removal with 19 trays versus 20 trays with all else being equal, then why have the additional tray? Implicit in that statement is that each additional tray gives additional benefit, namely additional sour gas removal or robustness to process upsets. More often than not in an absorber, most of the absorption occurs in the bottom trays, leaving the higher trays to remove the last small fraction necessary to meet the required product gas specifications.

In the typical scenario described above, the sour gases move in the expected direction on each tray, from the feed gas into the amine solution. If a sour gas moves in the opposite direction, i.e. the vapor strips sour gas from the amine, then the number of trays available for *absorption* of that sour gas decreases. If meeting the specification for that sour gas requires absorption on all the trays then stripping, or more generally, desorption, will cause out of specification product gas. For this reason, it is helpful to determine when desorption can occur, when it cannot, and what can be done to prevent desorption.

Desorption With Only One Acid Gas Species

There are situations where only a single acid gas species is present in the feed gas. This situation provides a useful starting point for analyzing the conditions that lead to desorption inside the column. So, the question at hand is,

Question: Can desorption occur when there is only one sour gas species?

One can contrive conditions where desorption could occur even with a single sour gas species. For instance, extremely warm or insufficiently regenerated semi-lean amine in a split amine feed arrangement could lead to desorption. The analysis below covers more typical conditions in a contactor. The assumptions behind the analysis allow for a wide range of conditions and provide the opportunity to determine whether desorption can occur when one might not expect it.

First, a few terms are defined. The column is numbered top to bottom from 1 to n.

\hat{f}_i^V = fugacity of the single sour gas in the vapor phase leaving stage i

T_i^L = liquid temperature on stage i

x_i^L = mole fraction of the single sour gas in the liquid phase on stage i

ϕ_i = vapor-phase fugacity coefficient on stage i

Next, the assumptions behind the analysis are,

Assumptions for 1 sour gas species

1. $\hat{f}_i^V = \phi_i y_i^V P$ and ϕ_i does not vary with temperature over the temperature range in the column. (y_i^V is the vapor mole fraction of the sour gas species leaving stage i .) Furthermore, ϕ_i does not vary appreciably with composition and is assumed to be constant. Since P is essentially constant throughout the column, \hat{f}_i^V is a positive linear function of y_i only.
2. \hat{f}_i^L increases with x_i^L at constant temperature.
3. \hat{f}_i^L increases with T_i^L at constant x_i^L . If x_i^L increases while T_i^L decreases, you cannot determine *a priori* whether \hat{f}_i^L increases.
4. Desorption cannot occur when $\hat{f}_i^V > \hat{f}_i^L$, i.e., there is not enough bulk mass transfer to force sour gas to move from lower fugacity in liquid to higher fugacity in vapor.
5. Gas is water-saturated entering contactor such that there is not significant water transfer between phases. (High acid gas concentrations might violate this assumption.)
6. The liquid in the column reaches a maximum temperature that is higher than either the liquid or vapor feeds. $T_{max}^L > T_{feed}^L$ and $T_{max}^L > T_{feed}^V$.
7. Heat generation occurs only in the liquid phase and is transferred to vapor, so $T_{max}^V < T_{max}^L$.
8. There is no heat loss or gain from fluids to surroundings.
9. The only feeds to the column are the feed gas at the bottom and the lean amine at the top.
10. There is negligible heat generation or consumption when absorption/desorption of sour gas does not occur.

With those assumptions, several questions follow, which we will explore below. These questions are,

1. Can desorption occur on the tray with the maximum liquid temperature?
2. Can desorption occur on trays below the maximum liquid temperature tray?
3. Can desorption occur on trays above the maximum liquid temperature tray?

Each of these questions are answered below.

1. Can desorption occur on the tray with the maximum temperature?

If the maximum liquid temperature occurs on stage j (*subscript j will be used throughout to reference maximum liquid temperature tray*),

$$T_{max}^L = T_j^L \tag{1}$$

If $T_j^L > T_{j-1}^L$ then either $T_{j+1}^V > T_j^L$ or absorption occurs on tray j . According to assumption (7),

$$T_{j+1}^V < T_j^L \tag{2}$$

so absorption occurs on tray j . If the acid gas is absorbing, it cannot be desorbing, so the answer is 'no'.

2. Can desorption occur below the tray with the maximum liquid temperature?

This question is broken into two parts, 2.A and 2.B.

2.A. Can desorption occur on the tray directly below the maximum temperature?

Since we showed above that absorption occurs on maximum liquid temperature tray j ,

$$\hat{f}_j^V \geq \hat{f}_j^L \quad (3)$$

$$\hat{f}_{j+1}^V > \hat{f}_j^V \quad (4)$$

If no absorption occurs on tray $j+1$, then

$$x_{j+1}^L \leq x_j^L \quad (5)$$

Equations (1) and (5) with assumptions (2) and (3) combine to make

$$\hat{f}_{j+1}^L < \hat{f}_j^L \quad (6)$$

Equations (3), (4) and (6) yield

$$\hat{f}_{j+1}^V > \hat{f}_{j+1}^L \quad (7)$$

Indicating a tendency for absorption, which precludes desorption on the tray directly below T_{max}^L .

2.B. Can desorption occur on trays further down the column?

If no desorption occurs on $j+1$, then

$$\hat{f}_{j+1}^V \geq \hat{f}_{j+1}^L \quad (8)$$

$$\hat{f}_{j+2}^V \geq \hat{f}_{j+1}^V \quad (9)$$

Equations (8) and (9) yield

$$\hat{f}_{j+2}^V \geq \hat{f}_{j+1}^L \quad (10)$$

Let us assume desorption occurs on $j+2$, then heat is absorbed on $j+2$ and

$$\hat{f}_{j+3}^V < \hat{f}_{j+2}^V \quad (11)$$

$$\hat{f}_{j+2}^V \leq \hat{f}_{j+2}^L \quad (12)$$

$$x_{j+1}^L > x_{j+2}^L \quad (13)$$

Equations (10) and (12) yield

$$\hat{f}_{j+1}^L \leq \hat{f}_{j+2}^L \quad (14)$$

The only way for (13) and (14) to be true is if

$$T_{j+2}^L > T_{j+1}^L \quad (15)$$

Since we assumed heat consumption on $j+2$ with desorption, the following are required for (15)

$$T_{j+3}^V > T_{j+2}^V \geq T_{j+2}^L \quad (16)$$

(15) and (16) combine to give

$$T_{j+2}^V > T_{j+1}^L \quad (17)$$

If $T_{j+2}^V > T_{j+1}^L$, and there is zero or positive heat generation on $j+1$ due to absorption, then

$T_{j+1}^L > T_j^L = T_{max}^L$. This is because the liquid flowing to $j+1$ from j is heated by both the vapor from below and heat generation on $j+1$. This violates our stipulation that the maximum temperature occurs on j , which means that our assumption that desorption occurs on $j+2$ is not valid. This analysis can be extended down the column to show that the answer to Question 2 is, “No, desorption cannot occur below the stage with the maximum liquid temperature.”

This now leaves question 3.

3. Can desorption occur on trays above the maximum temperature tray?

This question is broken into two parts, 3.A and 3.B.

3.A. Can desorption occur on the tray directly above the maximum temperature tray?

Absorption on j and $T_j^L = T_{max}^L$ yield

$$T_{j-1}^L < T_j^L \quad (18)$$

$$x_{j-1}^L < x_j^L \quad (19)$$

$$\hat{f}_j^V \geq \hat{f}_j^L \quad (20)$$

(18) and (19) combine to give

$$\hat{f}_{j-1}^L < \hat{f}_j^L \quad (21)$$

(20) and (21) combine to give

$$\hat{f}_{j-1}^L < \hat{f}_j^V \quad (22)$$

Since \hat{f}_j^V is independent of temperature, (22) implies absorption occurs on $j-1$.

3.B Can desorption occur further up the column?

Once again, we will assume desorption and then search for an inconsistency caused by that assumption. Let i represent a tray at least 2 trays higher than the maximum liquid temperature tray. For desorption to occur on stage i with no desorption on $i+1$,

$$\hat{f}_{i+1}^V < \hat{f}_i^L \quad (23)$$

Since no desorption occurs on $i+1$

$$x_i^L \leq x_{i+1}^L \quad (24)$$

$$\hat{f}_{i+1}^V \geq \hat{f}_{i+1}^L \quad (25)$$

(23) and (25) combine to give

$$\hat{f}_i^L > \hat{f}_{i+1}^L \quad (26)$$

For both (24) and (26) to be true,

$$T_i^L > T_{i+1}^L \quad (27)$$

For (27) to happen concurrently with desorption, the liquid leaving i must be cooled on $i+1$. Since no cooling due to desorption occurs on $i+1$, then cooling must come from the rising vapor,

$$T_{i+2}^V < T_{i+1}^L \quad (28)$$

If (28) then

$$T_{i+1}^V \leq T_{i+1}^L \quad (29)$$

Given (27) and (29)

$$T_i^L > T_{i+1}^V \quad (30)$$

combined with energy absorbed by desorption on i yields

$$T_{i-1}^L > T_i^L \geq T_i^V \quad (31)$$

Desorption on i also implies

$$\hat{f}_i^V \leq \hat{f}_i^L \quad (32)$$

$$x_{i-1}^L > x_i^L \quad (33)$$

(31) and (33) give

$$\hat{f}_{i-1}^L > \hat{f}_i^L \quad (34)$$

(32) and (34) give

$$\hat{f}_{i-1}^L > \hat{f}_i^V \quad (35)$$

which means desorption occurs on $i-1$. Energy absorption from desorption on $i-1$ and (31) yield,

$$T_{i-2}^L > T_{i-1}^L \geq T_{i-1}^V \quad (36)$$

Similar logic can be used to show that if desorption occurs on i or any tray higher, it must occur on every tray above that tray with the liquid temperature increasing ascending the column.

The analysis above is qualitative; it has investigated only how vapor and liquid fugacities change from tray to tray with a limit that $\hat{f}_j^V = \hat{f}_j^L$ if equilibrium is reached on a tray. The analysis has shown what is impossible. To summarize:

If the maximum liquid temperature is higher than either the vapor or liquid feed temperature then,

1. Desorption cannot occur on the tray with the maximum liquid temperature
2. Desorption cannot occur on any tray below the tray with the maximum liquid temperature
3. Desorption cannot occur on the tray directly above the tray with the maximum liquid temperature
4. Desorption cannot occur above the tray with the maximum liquid temperature unless,
 - a. on the highest tray where absorption occurs the vapor rising into the tray is colder than the liquid on the tray (Equation 28)
 - b. desorption occurs on every tray above where desorption begins
 - c. starting with the tray where desorption begins, the liquid temperature increases on every tray moving up the column (Equation 36)
 - d. the amine feed temperature is the highest liquid temperature above the tray with the maximum liquid temperature

The conditions required for desorption in Condition 4 seem unlikely to occur in practice. Figure 1 gives a hypothetical temperature profile required for a column that satisfies the conditions 4a, 4c and 4d. T_{max}^L occurs on Stage 11. Absorption occurs on stages 10 – 20. There is an appreciable increase between T_{10}^L and T_{11}^L , indicating that appreciably more absorption occurs on Stage 11 versus Stage 10. This seems unlikely given that a higher temperature on stage 11 would act to increase the liquid fugacity and reduce the driving force for absorption. This study could not reproduce this profile via simulation, and it is postulated that desorption cannot occur under steady state conditions with a single sour gas species unless at least one of the original assumptions is violated. Assumptions 8 and 9 constrain the preceding analysis to a “standard” configuration with amine fed solely onto the top tray and feed gas fed solely below the bottom tray.

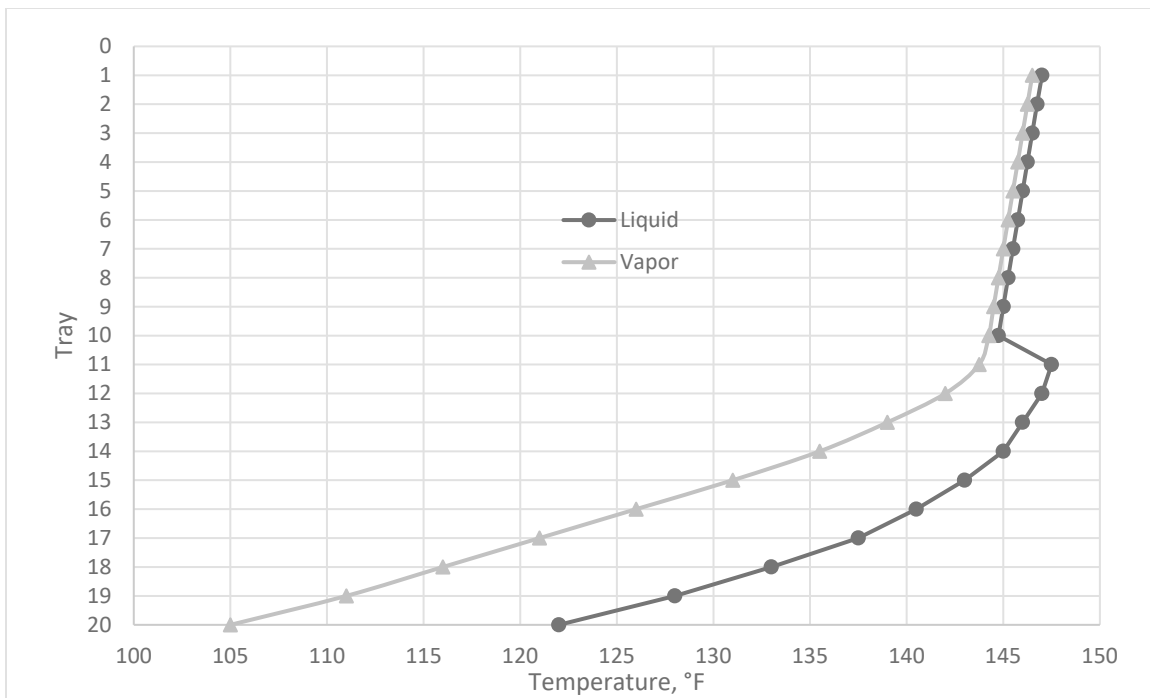


Figure 1 – Example of required temperature profile for desorption above bulge

However, desorption can occur with a single sour gas species with a non-standard configuration, such as shown in Figure 2. In Figure 2, a partially-regenerated stream leaves the middle of the regenerator, passes through a pump and cooler and feeds a middle tray of the absorber. Figure 3 shows the tray-by-tray CO₂ mass transfer rates for conditions that cause desorption in the contactor. Positive values for the mass transfer indicate absorption and negative values indicate desorption. In order to achieve desorption in the contactor, out of the ordinary conditions for the split feed case were required; the semi-lean feed to the contactor entered the contactor at 200 °F, and a low reboiler duty allowed CO₂ to accumulate in the regenerator, such that the semi-lean amine had only slightly lower CO₂ loading than the rich amine entering the regenerator. Even in this somewhat contrived case with higher than typical semi-lean temperature and CO₂ loading, the mass transfer rate dropped only slightly below zero. Under more typical conditions of a semi-lean feed configuration where the semi-lean amine feed is cooled below the liquid temperature on the semi-lean feed tray, desorption did not occur. Similar results can be seen with feed streams containing H₂S as the only acid gas.

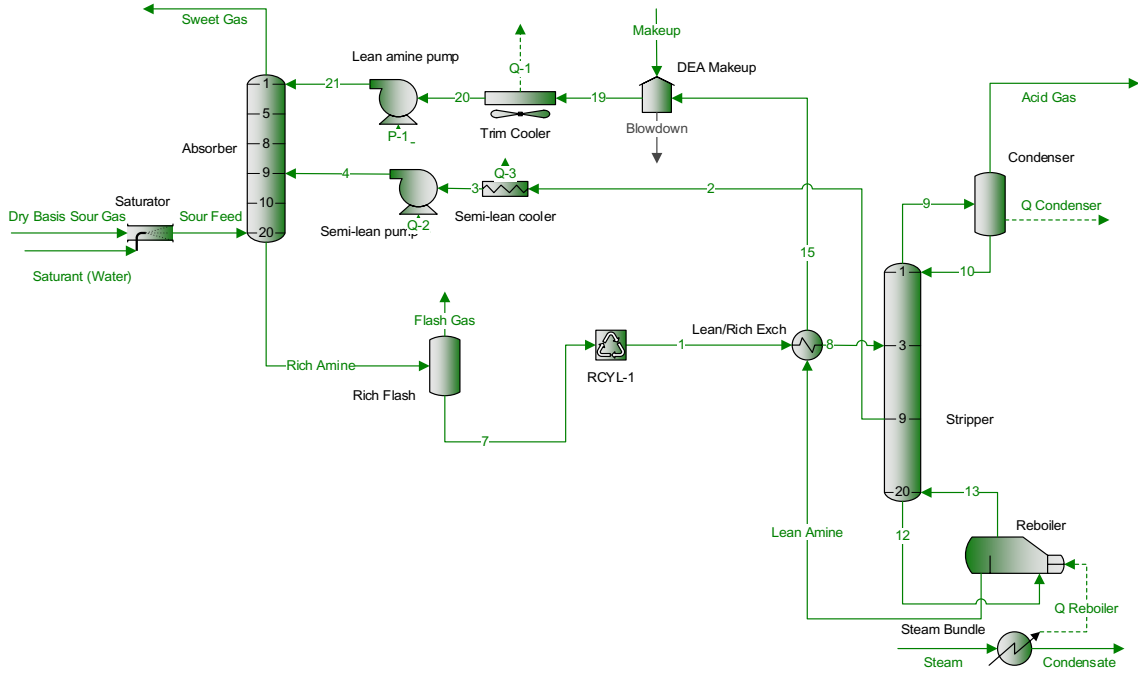


Figure 2 – Amine treating with semi-lean feed

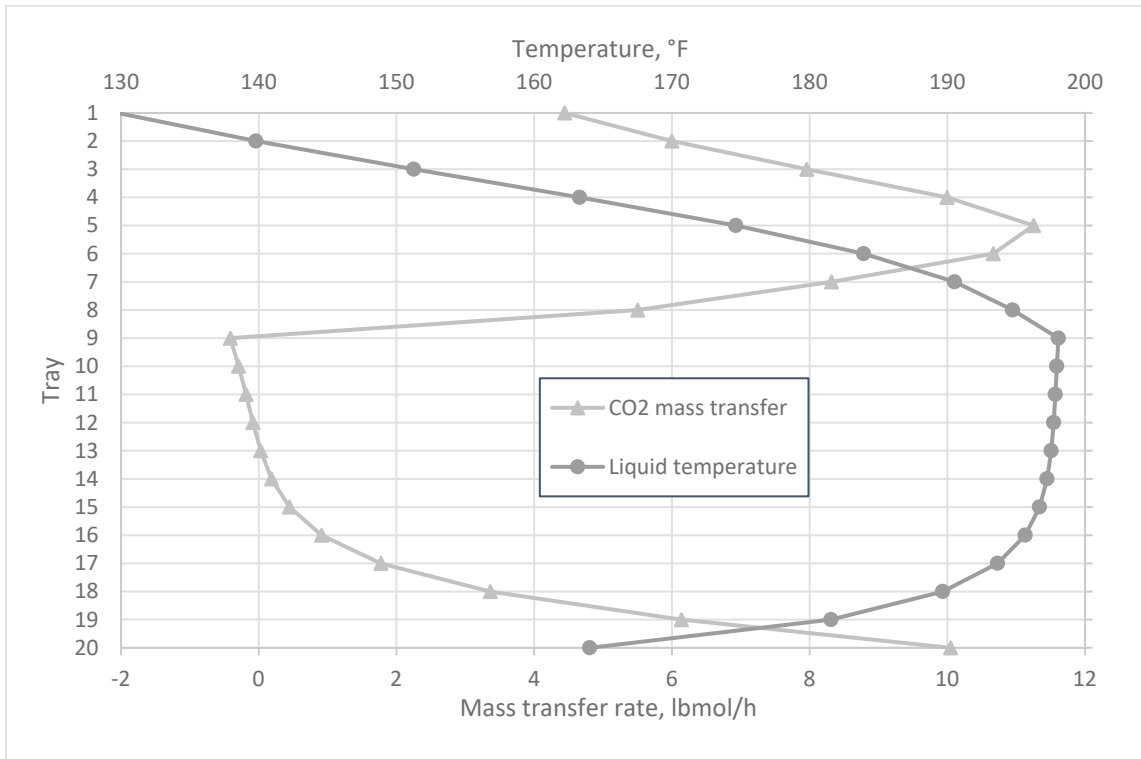


Figure 3 – CO₂ mass transfer rate and liquid temperature in absorber with semi-lean feed on Tray 9 (negative values indicate desorption)

Desorption with multiple acid gas species

The preceding analysis relied on the fact that \hat{f}_i^L is a function of only T_i^L and x_i^L . When another sour gas species is present, \hat{f}_i^L also depends on the mole fraction of the other sour gas. For instance, when the feed gas contains both H₂S and CO₂, absorption of CO₂ can increase \hat{f}_{i,H_2S}^L by increasing the liquid temperature, lowering the liquid pH, converting unprotonated amine to its protonated form, and/or consuming amine to form carbamate. Similarly, although the effect may be small, CO₂ absorption can increase \hat{f}_{i,H_2S}^V by increasing y_{i,H_2S}^V .

An analysis similar to the one above for a single sour gas species that considers two sour-gas species would be tedious. Instead, simulation can provide examples that show desorption can occur under certain conditions. Figure 4 gives an example that shows desorption of H₂S caused by CO₂ absorption in the middle of the column. In this simulation, the contactor uses a 33% DEA lean amine solution to treat a 200-psig feed gas containing 9.3% CO₂ and 0.85% H₂S. The sweet gas contaminant concentrations are reduced down to 1.96% CO₂ and 22 ppm H₂S. The column has a single lean amine feed at the top and a single sour feed at the bottom. The gas and lean amine feed temperatures are 90 and 110 °F, respectively.

The final rich loading of 0.55 mol acid gas/mol amine is high, but not extreme. However, with the relatively low feed pressure, the sour gas fugacities in the feed are relatively low. This drives the rich approaches for H₂S and CO₂ to 94% and 88%, respectively. The high rich approaches indicate that the exiting rich amine is close to equilibrium with the feed gas such that little driving force for absorption exists at the bottom of the column. The high rich-end approach appears to be a requirement for desorption in the column. If a large driving force exists at the bottom of the column, it is unlikely that this driving force will become negative further up the column.

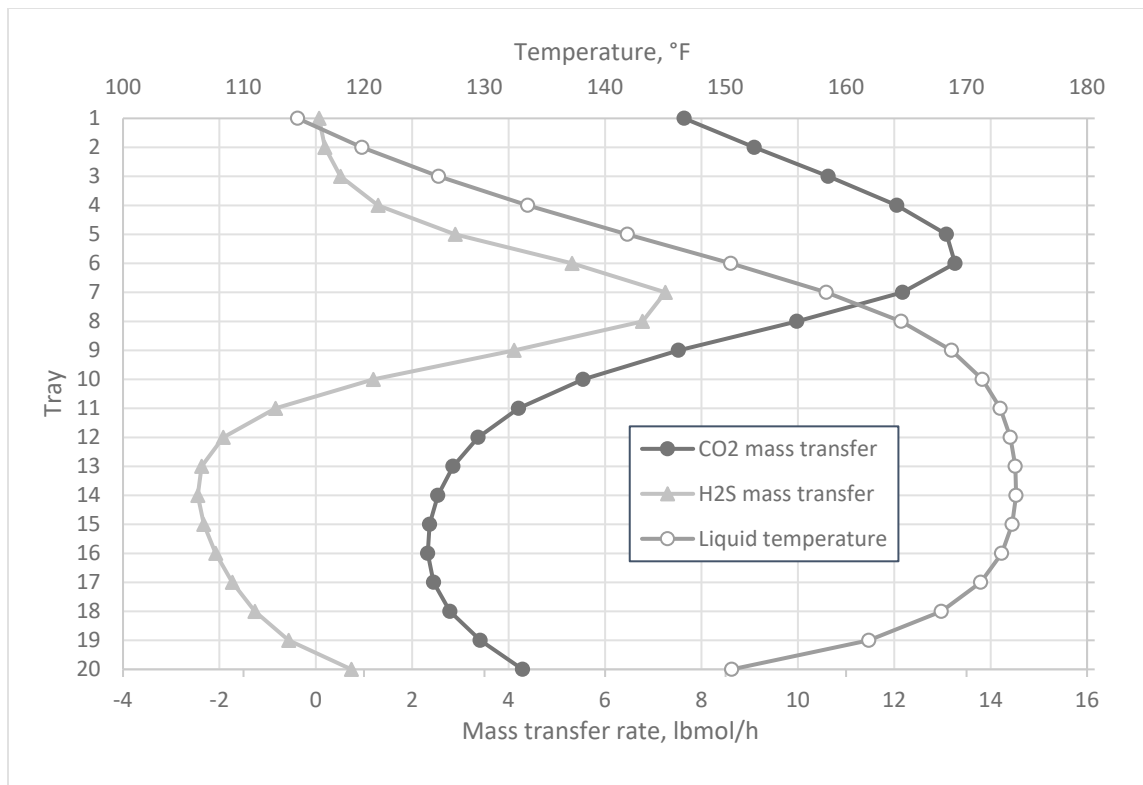


Figure 4 – H₂S and CO₂ mass transfer rates showing desorption of H₂S, along with liquid temperature. 2-in weir height

In this example, desorption of H₂S occurs concurrently with absorption of CO₂ on trays 11-19. The question arises then, is desorption caused by elevated temperature due to heat of absorption or does CO₂ force the H₂S out by some other mechanism such as lowering the solution pH.

H₂S, a weak acid, exists in amine solutions primarily in two forms, H₂S and HS⁻, governed by the equilibrium reaction,



The ratio of the forms is governed by the equation

$$K_{a,H_2S} = \frac{[HS^-][H^+]}{[H_2S]} \quad (38)$$

where the dissociation constant, K_{a,H_2S} , is a weak function of temperature and often assumed to be constant. While amines do provide buffering capacity, addition of another acid such as CO₂ to the solution will increase $[H^+]$. The $[H^+]$ increase may be small since the amine will absorb H^+ , but with a finite amount of amine, some of the additional H^+ will not be absorbed and will cause HS^- to shift to H_2S in accordance with (38). When $[H_2S]$ increases, \hat{f}_{i,H_2S}^L increases, which lowers the driving force for H₂S absorption. With enough of an increase H₂S could desorb from the liquid into the vapor.

A similar expression for K_{a,CO_2} for the CO₂ equilibrium reaction is



$$K_{a,H_2S} = \frac{[HCO_3^-][H^+]}{[CO_{2,dissolved}]} \quad (40)$$

where $[CO_{2,dissolved}]$ represents the sum of CO_2 and H_2CO_3 in solution.

The question then is what drives the absorption of CO_2 while H_2S desorbs in Trays 11-19 of Figure 4. Since K_{a,CO_2} is larger than K_{a,H_2S} , CO_2 is a stronger acid with a higher propensity to dissociate in an amine solution. This could result in the absorption of CO_2 yielding HCO_3^- and H^+ according to (39), followed by the H^+ combining with HS^- to form H_2S .

Furthermore, CO_2 can also react with DEA to form a carbamate, in accordance with



Given the dissociation constant for DEA is

$$K_{a,(C_2H_5O)_2NH} = \frac{[(C_2H_5O)_2NH][H^+]}{[(C_2H_5O)_2NHH^+]} \quad (42)$$

consumption of DEA via (41) will increase $[H^+]$.

When trying to determine whether temperature or displacement by CO_2 drives the desorption of H_2S seen in Figure 4, consideration of the kinetics of CO_2 and H_2S absorption prove useful. Reaction (37) is very quick. The corresponding reaction (39) for CO_2 occurs in two steps, with the rate-limiting step much slower than (37).

Reaction of CO_2 via the carbamate route in (41) is quicker than (39), but still considerably slower than (37). H_2S rapidly approaches equilibrium in the column, while the kinetically-limited CO_2 requires higher rates of mass transfer to approach equilibrium. Therefore, CO_2 absorption is more sensitive to mass transfer area than H_2S absorption. If the mechanism for H_2S desorption is displacement by CO_2 one would expect that allowing a greater area for CO_2 to absorb and react would increase H_2S desorption.

One can achieve greater mass transfer rates by increasing the weir height on a tray, which increases the contact area in the column. Figure 5 compares the absorption rate for weir heights of 2 in and 6 in. Figure 6 shows the liquid temperature profile in the column for the same conditions. The treated gas CO_2 and H_2S levels are 1.71% and 350 ppm, respectively, which represents a significant increase from the 22 ppm H_2S at the lower weir height. Figure 5 shows that the maximum desorption rate of H_2S increases as the contact area increases. While the shape of the temperature profiles in Figure 6 do not exactly match, the overall maximum temperatures agree to within 1 °F between the two cases. This indicates that something other than, or in addition to, temperature causes the desorption.

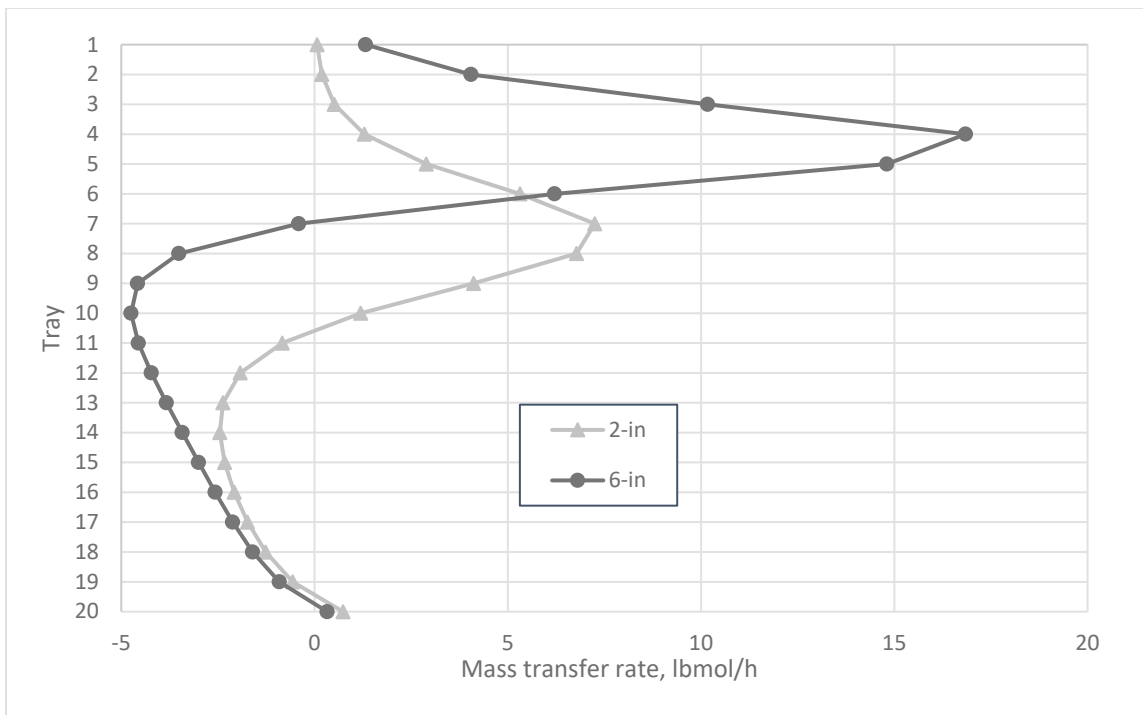


Figure 5. – H₂S mass transfer rates for 2 and 6-in weir heights

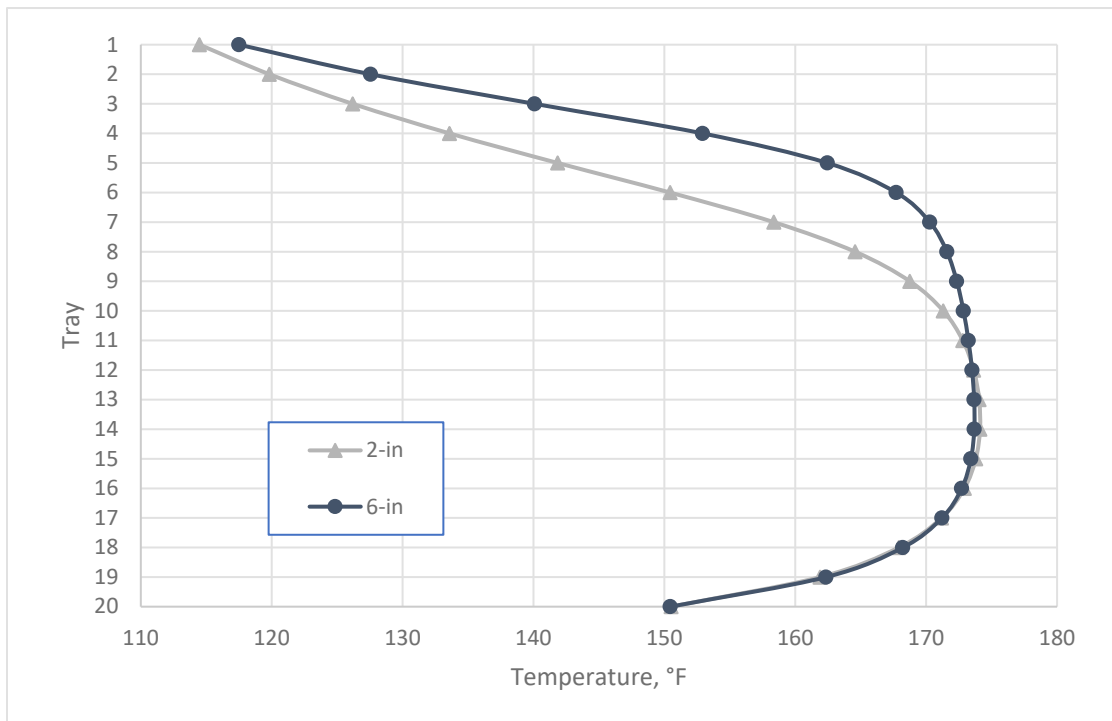


Figure 6 – Liquid temperature profile for 2-in and 6-in weir heights

To further test whether temperature drives H₂S desorption, Figure 7 shows the effect of lowering the gas and liquid feed temperatures 30 °F below the conditions in Figure 4. In this case the

maximum liquid temperature decreased approximately 25 °F to 148 °F. However, desorption still appears near the bottom of the column. In this case though, the desorption occurs low enough in the column that the remaining trays above the desorption reduce the product gas H₂S concentration to 1.2 ppm, well below typical H₂S product specification, while the CO₂ increased slightly from the base case to 2.08%. The desorption here at temperatures well below the previous cases seems to indicate that displacement by CO₂ drives H₂S desorption. One could say that the heat released by CO₂ absorption drives the H₂S release. The minimum to maximum liquid temperature change is approximately 60 °F in both cases. In reality, the pH decrease and temperature increase caused by CO₂ absorption can work together to drive desorption.

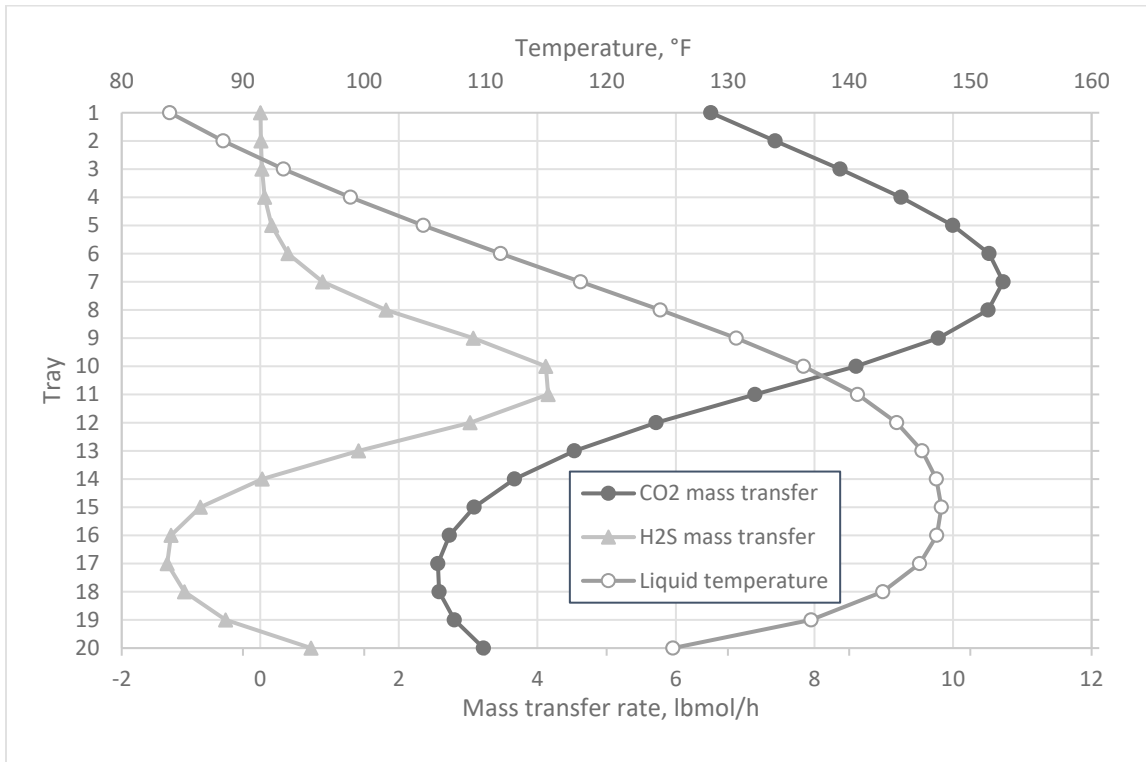


Figure 7 – H₂S and CO₂ mass transfer rates showing desorption of H₂S at reduced feed temperatures

Before concluding that equilibrium-driven absorption of CO₂ forces the H₂S out of solution, one should consider that the feed gas in this example contains 10X more CO₂ than H₂S. It could be that the quicker-reacting H₂S builds up in the descending amine at the top of the column, only to be forced out of solution once amine absorbs the much larger bulk of CO₂ further down the column. To test this idea, Figure 8 shows the results of a simulation where the H₂S and CO₂ fractions in the feed are switched from Figure 4. In this case, we once again see H₂S desorption in the column.

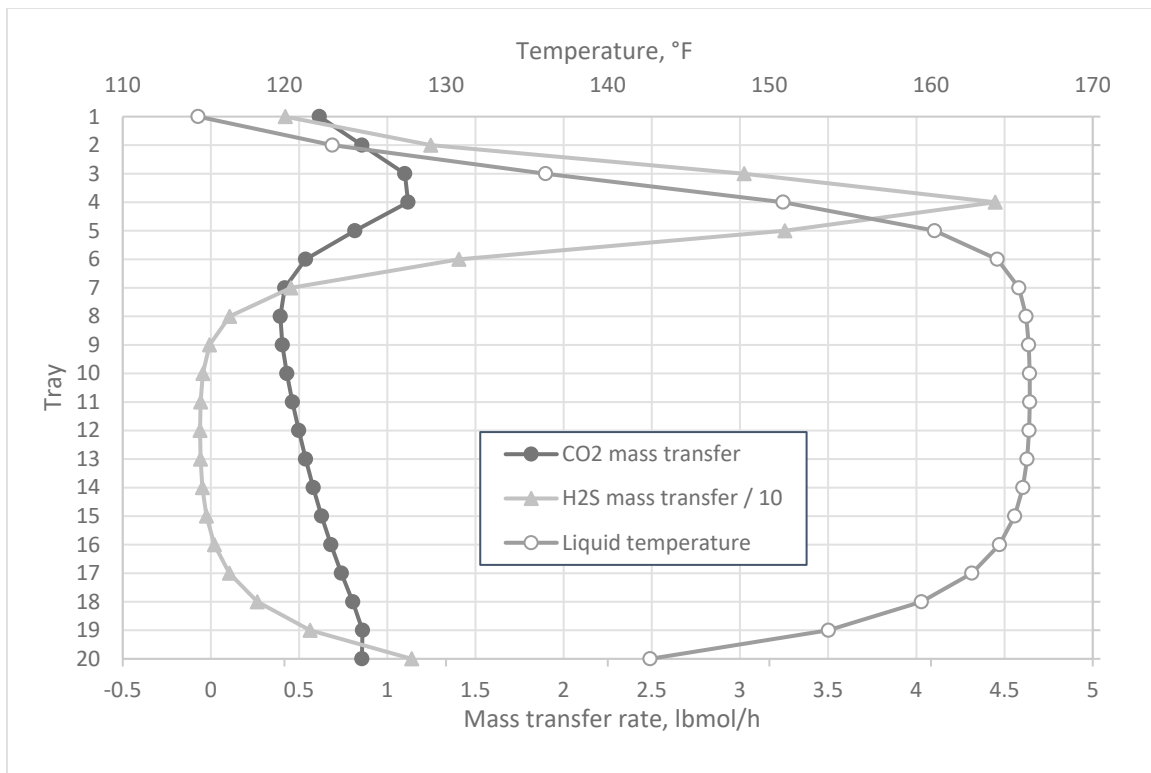


Figure 8. H₂S and CO₂ mass transfer rates showing desorption of H₂S with high H₂S/low CO₂ feed

CO₂ desorption

This study also searched for conditions where H₂S absorption occurred simultaneous with CO₂ desorption. The only conditions found where CO₂ desorption occurred had total acid gas rich loadings above 0.8 mol acid gas/mol amine. There was significant sour gas in the product gas, and it is felt that the conditions required for CO₂ desorption were unrealistic. Even under these conditions, the CO₂ desorption rate was much smaller than the H₂S desorption rates discussed above.

COS desorption

Similar to CO₂ + DEA carbamate formation in Equation (41), COS can react with DEA to form a thiocarbamate according to the following mechanism,



Reaction of CO₂ with DEA will increase $[(C_2H_5O)_2NHH^+]$ and lower $[(C_2H_5O)_2NH]$, which could lead to the reverse reaction in (43). Further, as with CO₂ and H₂S, higher temperatures caused by absorption of other species can drive COS desorption by increasing the vapor pressure of COS.

Figure 9 shows COS desorption in a contactor with a feed gas containing 10.5% CO₂, 0.4% H₂S, 0.1% COS at 800 psig. (Please note the COS desorption simulations used 800 psig as compared to 200 psig in the H₂S simulations. COS desorption did not occur at 200 psig.) The gas and lean amine feed temperatures are 90 and 110 °F, respectively. The 33% DEA feed rate is adjusted to give a rich total acid gas loading of 0.60 mol acid gas/mol amine. The sweet gas contaminant concentrations were lowered down to 0.22% CO₂, 1.2 ppm H₂S, and 0.048% COS. The desorption occurs at the bottom of the tower, and the rate is relatively small. Unlike the comparison in

Figures 4 and 7, where lower feed temperatures did not prevent H₂S desorption, lowering the feed temperatures by 30 °F did prevent COS desorption as shown in Figure 10. The amine circulation rate was maintained the same between Figure 9 and Figure 10.

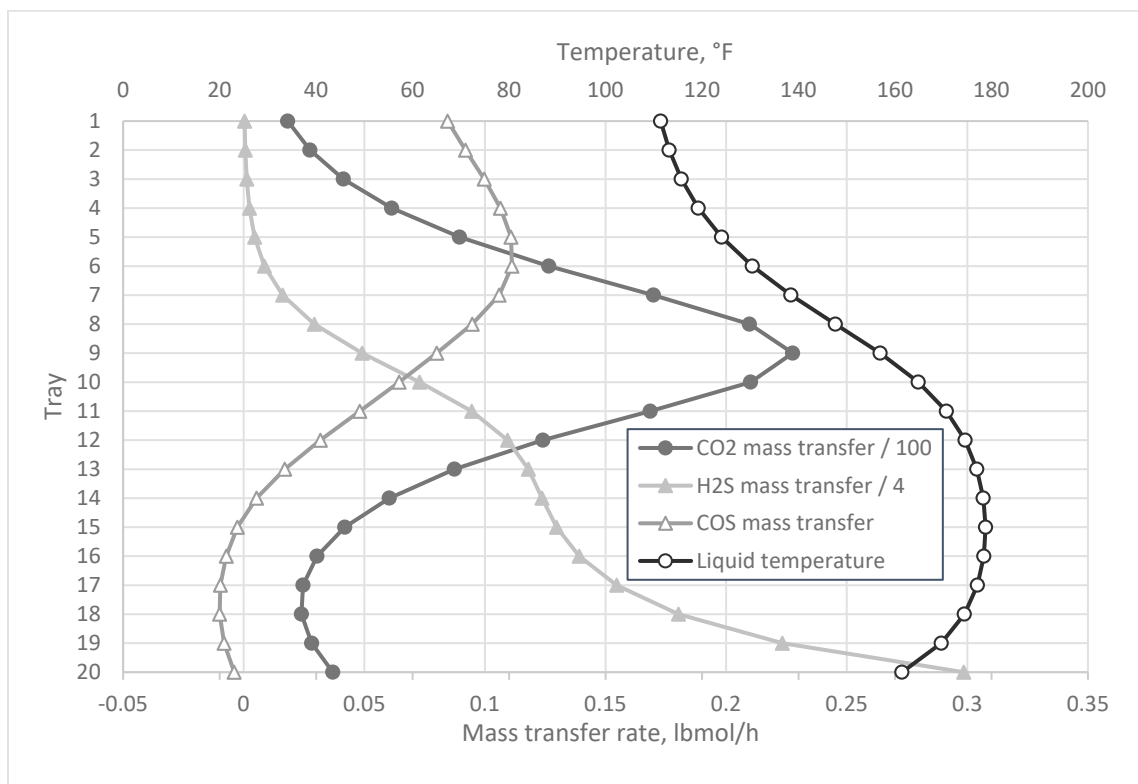


Figure 9. CO₂, H₂S, and COS mass transfer rates showing desorption of COS with high CO₂ feed

Raising the circulation rate slightly, to bring the rich amine loading down to 0.55 mol acid gas/mol amine, while maintaining the feed temperatures the same as Figure 9, also prevented desorption as shown in Figure 11. Please note that the temperature bulge has a 12 °F lower maximum with the lower loading compared to Figure 9.

Figures 9, 10, and 11 do not make it clear whether temperature or CO₂ displacement of COS causes desorption. Vaidya and Kenig¹ state that the COS to thiocarbamate conversion rate (Equation 43) is two orders of magnitude slower than CO₂ to carbamate (Equation 41), which would allow (C₂H₅O)₂NHH⁺ to accumulate and drive desorption according to (43). Higher loading in Figure 9 versus 11 would lead to higher [(C₂H₅O)₂NHH⁺], providing a larger driving force for desorption. However, the lower liquid temperatures in both Figure 10 and 11 reduce the COS liquid fugacity, which lowers the desorption driving force. As with H₂S desorption, concurrent CO₂ absorption has multiple synergistic effects on COS absorption. All of which can work together to drive desorption.

Desorption with MDEA

This study could not determine conditions that lead to desorption from MDEA solutions, except at extremely high loadings. It is likely that the slow absorption rate for CO₂ with MDEA prevents CO₂ from absorbing to an extent that would drive H₂S desorption.

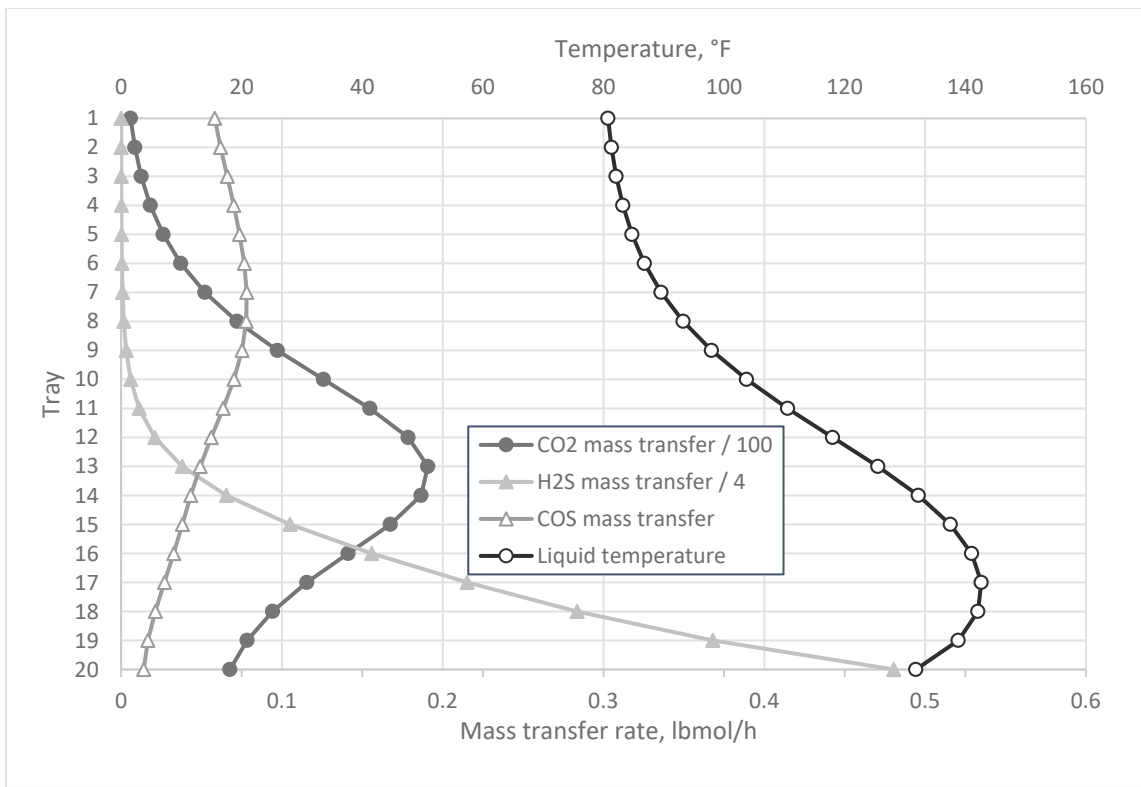


Figure 10. CO₂, H₂S, and COS mass transfer rates showing lack of COS desorption with lower feed temperature

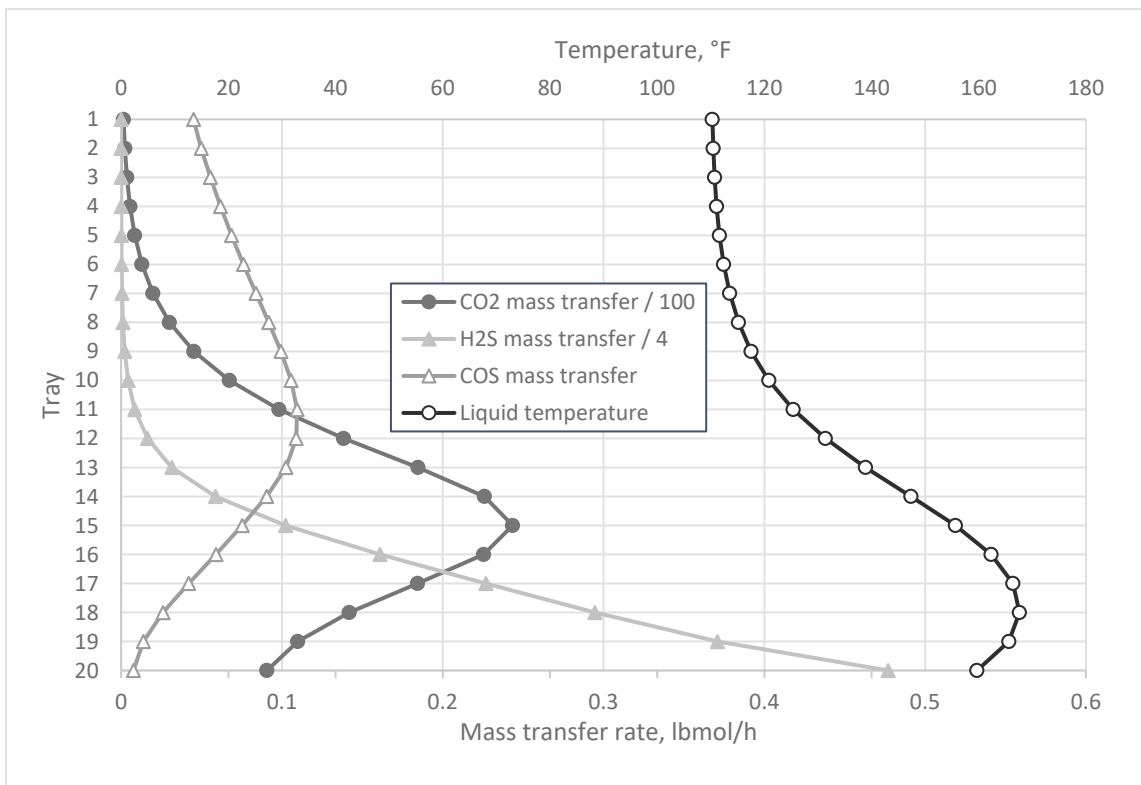


Figure 11. CO₂, H₂S, and COS mass transfer rates showing lack of COS desorption with lower rich loading

Conclusion

Desorption of a sour gas inside an amine contactor shrinks the effective height for absorption of that sour gas. In some cases, this can cause an out of specification product gas. In almost all cases desorption should be avoided as it increases the likelihood that process upsets will cause unmet specifications. The analysis presented here showed that desorption cannot occur under “standard” conditions if only one sour gas species is present in the feed gas. With only a single sour species, desorption could result from “non-standard” conditions, such as insufficiently regenerated or excessively hot semi-lean feed, or extremely high feed gas or lean amine temperatures.

Things change when the feed contains multiple sour species. With a single sour species, the only parameters that affect the driving force for absorption are the liquid temperature and mole fractions of the sour species in the liquid and vapor. With additional sour species, the liquid mole fraction of one sour species affects the absorption driving force for the other species. Specifically, an increase in the CO₂ mole fraction can reverse the driving force for H₂S such that H₂S desorbs from the amine solution. Both the heat produced and the shifting of the chemical equilibrium by CO₂ absorption affect H₂S behavior in the column. A good analogy for this behavior might be the tortoise (CO₂) and the hare (H₂S). Given enough time, CO₂ being the stronger acid can displace H₂S if the gas and amine solution are allowed to come to equilibrium. The contact times offered by a typical contactor do not allow CO₂ to reach equilibrium.

Likewise, CO₂ can also drive desorption of COS. Both H₂S and COS desorption seem to require rich amine loadings that lead to rich approaches well over 80%.

References

1. Vaidya, P. D., & Kenig, E. Y. (2009). Kinetics of carbonyl sulfide reaction with alkanolamines: A review. *Chemical Engineering Journal*, 207-211.