

The CONTACTOR™

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Ammonia Distribution in Sour Water Strippers

A few months ago we put the finishing touches on a mass transfer rate model for sour water strippers (SWSs), and we've been using it to learn some things about these columns that you might find interesting, maybe even surprising. One of them is exactly where ammonia concentrations can be highest in a SWS and under what conditions. An important factor in ammonia traffic is the presence and role of hydrogen sulfide.

Sour water chemistry is important to a real understanding of stripping column performance. In sour water, unless precipitates are formed, all species except the dissolved gases themselves (H_2S and NH_3) exist as ions, not chemical compounds. Cations and anions may *appear* to form compounds; however, their association with each other is just a notational convenience and an artifact of the need to maintain electrical neutrality at every point in the solution. Thus, when hydrogen sulfide and ammonia dissolve into water, they exist mostly as bisulfide anions and ammonium cations. A higher NH_3 concentration can dissolve more H_2S , and vice versa. Thus, it is possible for sour water to contain higher concentrations of both ions than would be possible if these species were present alone.

Stripping can be driven by a reboiler (which adds no water to the system), by using live steam, or by using a stripping gas. If a stripping gas is used, the stripped ammonia and hydrogen sulfide will be diluted by a rather large volume of stripping gas, and sending this directly to a Claus plant would put quite a bit of additional load (volumetric gas flow) on the Claus unit. Furthermore, if the stripping gas were a hydrocarbon, the sulfur plant would be seriously affected. A better place for the stripped gas and its non-condensable carrier is an amine unit where the hydrocarbon can be separated out and the components of interest then sent in a more concentrated form to the sulfur plant. It may well make more sense to use methane as a

stripping gas in such cases, rather than nitrogen, which ultimately ends up with the treated gas and degrades its heating value.

In this case study, we will use live steam as seen in Figure 1 to strip the water shown in Table 1

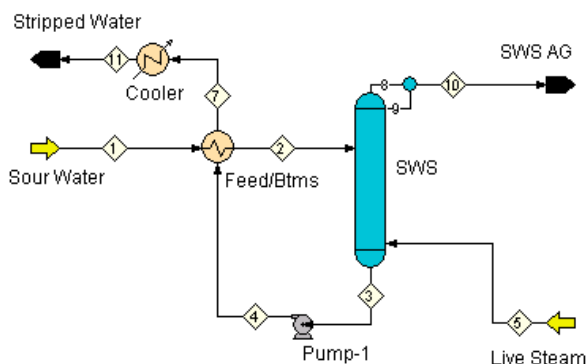


Figure 1 Stripping Sour Water with Live Steam

Table 1 Sour Water Feed Conditions

Total flow (lb/h)	150,000
Temperature (°F)	135
Pressure (psia)	70
Water (mol%)	96.4
Hydrogen sulfide (mol%)	1.5
Carbon dioxide (mol%)	0.1
Ammonia (mol%)	2.0

and the effect of heat stable salts (HSSs) will be deferred to a later issue of The Contactor. We will look at four boilup rates corresponding to 0.78, 0.92, 1.2 and 1.48 lb 50 psig steam per US gallon of water. To be specific, stripping will be done in a 40-tray column with a 20°F temperature approach in the preheater. The top 5 trays are wash trays and feed is to tray 6 (from the top). Figure 2 shows how vapor-phase ammonia concentration varies throughout the stripper. The parameter in this plot is the pounds of 50 psig steam per gallon of sour water treated. The solid lines represent the vapor

phase ammonia concentrations that would be in equilibrium with the water on each *real* tray.

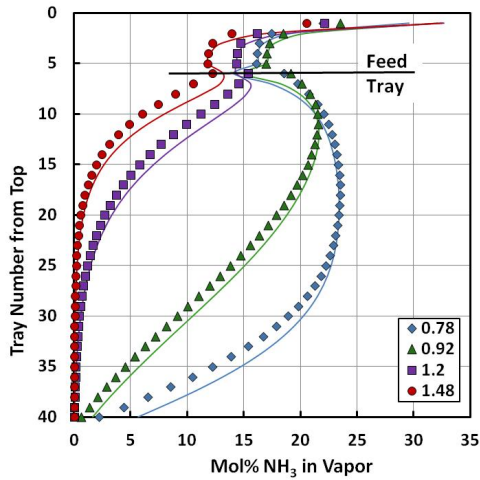


Figure 2 Stripper Ammonia Profiles

Perhaps the most striking feature is the huge ammonia bulge occupying most of the tower at the lowest steam rate. The bulge is driven back into the upper half of the tower at a steam rate of 0.92 lb/gallon and even at a steam rate of 1.2 lb/gallon there is still a small bulge immediately below the feed tray. Perhaps this is the underlying reason it is generally recommended to use a steam rate of at least 1 lb/gallon. But even if a substantially higher steam rate is used, there can still be several effectively idle stripper trays.

The curves of actual composition above each tray cross the equilibrium lines at low steam rates because NH₃ stripping takes place below the maximum and absorption above. This can lead to some seemingly strange behavior of Murphree[†] vapor efficiencies. For example, because mass transfer reverses direction at the maximum point of the bulge, the Murphree efficiency of a single tray can become negative or greatly exceed 100% there. However, away from the maximum, Murphree vapor efficiency is fairly constant at about 40%. It should be understood though, that this does *not* mean 2½ real trays are in any way equivalent to a theoretical or ideal stage. ProTreat[®] deals in real trays—not idealizations.

[†] For trays numbered top-down, the Murphree vapor efficiency of tray *n* is $(y_n - y_{n+1}) / (y_n^* - y_{n+1}) \times 100$. This is a measure of how closely individual *real* trays approach equilibrium. It is not to be confused with an overall tray efficiency to which it is completely unrelated. Indeed, the overall tray efficiency uses antiquated ideal stages and ideal stage calculations (and assumes phase equilibrium between exiting phases). It completely obscures what is being shown in Figure 4. Using overall tray efficiency loses the details by its crude averaging.

Figure 3 shows that H₂S does not exhibit a bulge under any value of steam rate. H₂S falls off quickly right below the feed tray, then falls slowly.

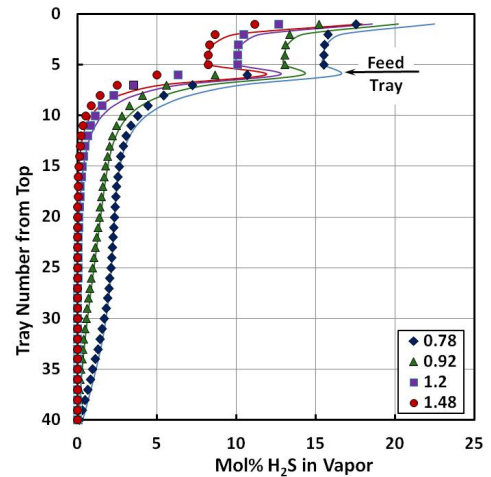


Figure 3 Stripper H₂S Profiles

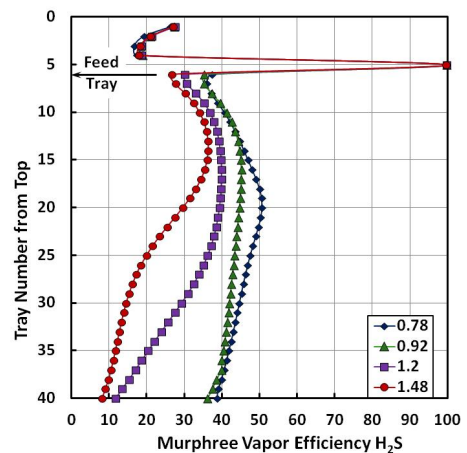


Figure 4 Stripper H₂S Murphree Efficiencies

At low steam rates each tray runs at about 45% of equilibrium vapor, but as steam rate is raised, vapor efficiency falls off to quite small values. Are these overall efficiencies? No, but they are certainly indicative of the wide tray-to-tray variation that the ideal stage with overall efficiency model completely masks. Only a ***genuine, heat and mass transfer*** rate model is capable of these details, which in themselves are indicative of ProTreat[®] model strengths and ensuing reliability.

To learn more about this and other aspects of gas treating, plan to attend one of our free seminars. Visit www.ogtrt.com/seminars for details on free seminars in Houston, Denver, or Calgary.

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