

Predicting and Mitigating Corrosion in Amine Units

*Laurance Reid Gas Conditioning Conference
February 22-25, 2015 – Norman, Oklahoma, USA*

Clayton E. Jones
Optimized Gas Treating, Inc.
112 Cimarron Pk. Loop, Suite A
Buda, TX 78610
+1 512 312 9424
Clay.Jones@ProTreat.com

Nathan A. Hatcher
Optimized Gas Treating, Inc.
112 Cimarron Pk. Loop, Suite A
Buda, TX 78610
+1 512 312 9424
Nate.Hatcher@ProTreat.com

G. Simon A. Weiland
Optimized Gas Treating, Inc.
112 Cimarron Pk. Loop, Suite A
Buda, TX 78610
+1 512 312 9424
Simon.Weiland@ProTreat.com

Ralph H. Weiland
Optimized Gas Treating, Inc.
Route 5, Box 940
Coalgate, OK 74538
+1 580 428 3535
Ralph.Weiland@ProTreat.com

ABSTRACT

Corrosion is a ubiquitous problem in gas treating in the petroleum and natural gas industries, in syngas plants, in processing unconventional gases such as shale gas and coal seam gas, and in numerous other treating applications. The primary impurities removed in the treating process are the acid gases carbon dioxide and hydrogen sulfide. The corrosion of equipment and piping is an inevitable consequence of removing these very gases with amines, and of handling sour water. There are other corrosive impurities that either enter in small amounts with the gas, such as HCN and oxygen, or that are produced in the amine system itself, mostly heat stable salts (HSSs) derived from HCN. Corrosion rates are affected by the nature of the corrosive agent, temperature, fluid velocity, the presence of solids, and the metallurgy involved¹. To prevent equipment failures, mitigate risk, and select optimal materials, it would be decidedly beneficial to be able to predict corrosion rates pertinent to the particular processing conditions. This contribution describes the underpinnings of a chemistry-based predictive corrosion model built on both public and much proprietary corrosion rate data. The model includes dependence on ionic solution composition (speciation), fluid velocity, temperature, HSSs, and metallurgy. The model is then used in a refinery case study involving treating with DEA.

Predicting and Mitigating Corrosion in Amine Units

Clayton E. Jones, Optimized Gas Treating, Inc., Buda, TX 78610, USA
Nathan A. Hatcher, Optimized Gas Treating, Inc., Buda, TX 78610, USA
G. Simon A. Weiland, Optimized Gas Treating, Inc., Buda, TX 78610, USA
Ralph H. Weiland, Optimized Gas Treating, Inc., Coalgate, OK 74538, USA

Corrosion in Alkaline Systems

The focus of this paper is corrosion in alkaline aqueous solutions typical of operating amine units. Although the concepts presented here apply equally to pH-neutral and acidic systems, these systems are not addressed because the amount of corrosion data available for modeling is not sufficiently extensive. The corrosive action of H₂S is inherently different from CO₂ in that H₂S can and does form a relatively robust protective iron sulfide layer on the metal surface. On the other hand, iron carbonate forms a more fragile layer, so it offers much less protection. There are several tenets embedded in the model:

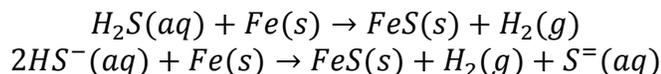
- The corrosive agents are acids
- In and of itself, the amine (or ammonia) is *not* corrosive
- The iron sulfide film can protect against further corrosion
- Iron carbonate also offers protection but to a lesser extent
- High fluid velocities physically increase corrosion rates
- Higher temperature increases corrosion rates
- Heat stable salts chemically exacerbate corrosion

Although technically incorrect, the industry continues to bandy about such terms as “amine corrosion” and “alkaline stress corrosion cracking” to describe corrosion that, at the root level of chemistry, is really caused by dissolved acid gases in various forms. For example, nearly 60 years ago, Polderman² reported that 20 wt% MEA without acid gas was actually *less* corrosive to steel than pure water. As far as the corrosive agents themselves are concerned, the important parameter is the *chemical activity of the dissolved acid gas species responsible for corrosion*. The activity (vs. concentration) changes with the amine type, amine concentration, acid gas loadings, the concentrations and identities of HSSs, and temperature — this may make some amine systems *appear* to be more susceptible to corrosion than others; however, the essential point is that it is the *activity of the corrosive species* that is of direct importance. The type of amine affects corrosion rates only insofar as it affects the speciation of the solution.

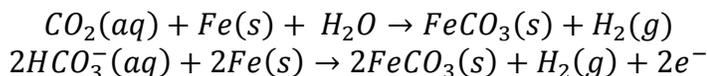
The chemical species of interest are: bisulfide ion (HS⁻), free physically dissolved H₂S, bicarbonate ion (HCO₃⁻), and free physically dissolved CO₂, all of which are oxidizing agents. These species are called protonic acids because they can either give up or produce a hydrogen ion³. Sulfide (S⁻) and carbonate (CO₃⁻) ions are also present; however, they themselves are final reaction products and are unable to provide the hydrogen ion necessary for the oxidation of iron. Molecular hydrogen sulfide and carbon dioxide react with iron only in the presence of water.

The final distribution of molecular and ionic species is found by solving the equations of chemical reaction equilibria, atom balances, and a charge balance. The resulting set of species concentrations is termed the solution's *speciation*. Speciation affects corrosion rates but corrosion products do not substantially alter the speciation.

In their simplest stoichiometric forms, the basic corrosion reactions of dissolved H₂S species with iron are:



For CO₂, the relevant reactions are



The oxidation reaction with hydrogen sulfide is faster than the reaction with bisulfide; however, the alkalinity of the amine (and ammonia) solutions means that the dissolved H₂S is predominantly in the bisulfide form, with very little remaining as free molecular hydrogen sulfide. This is also true of dissolved carbon dioxide. The concentrations of free H₂S and CO₂ are pH dependent and pH is a function of amine strength, total dissolved acid gas, temperature, and to a lesser extent HSS concentrations. However, heat stable salt species and their concentrations do affect the speciation of the solution, especially in lean solvents.

As discussed by Cummings et al.¹, the sequence of physicochemical steps in the process of oxidizing iron consists of transporting the acid from the bulk solution to the metal surface, adsorption of the acid onto the surface, reaction with iron, and transport of reaction products back into the bulk solution. The steps are analogous to what occurs in heterogeneous catalysis. The reaction of H₂S with the iron component of various iron-based metallurgies forms solid iron sulfide and hydrogen gas, and as the reaction proceeds, the surface of the iron is changed to a mosaic of iron and sulfide ions. The surface expands by addition of sulfide, and the liberation of hydrogen gas encourages the expansion. The surface layer of iron sulfide is somewhat porous, and it adheres to the surface of the free metal. The iron-H₂S and iron-HS⁻ reactions form reaction products in completely different phases, as does the reaction with carbon dioxide and bicarbonate. Thus, because the gaseous and solid reaction products are continually removed from the reacting solution, there is a *strong driving force for continued corrosion*.

What limits the corrosion reactions is primarily the amount of bare, unreacted iron that the passivating film leaves available at the metal surface. A secondary factor is the concentration of dissolved reactant gas, i.e., H₂S and CO₂, which are other parameters. Thus, H₂S, HS⁻, CO₂, and HCO₃⁻ all react with unprotected iron. To control the concentration of dissolved gas in its various forms, most practitioners adopt a rich-amine acid-gas-loading, upper limit of 0.4 to 0.5 moles of total acid gas per mole of molecular amine.

Passivation

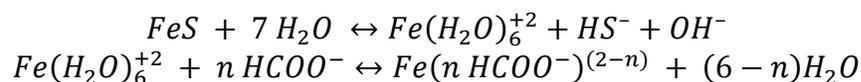
In a new amine unit or in one that has just been chemically cleaned, during the period immediately following introduction of H₂S, for example, the corrosion rate climbs exponentially.

Measurements by Cummings et al.⁵ using a bench-scale mini-amine-plant with relatively lean amine indicated in some cases over 600 mpy (milli-inches per year) corrosion rate during this phase of operation. Eventually the corrosion rate levels off, and then begins to fall as the iron sulfide layer increasingly occludes contact between H₂S and iron (passivation). Finally, once the iron sulfide layer is established, the system settles down to a nominal (and hopefully low) residual corrosion rate. The time over which the final steady corrosion rate is attained is hours to days. During these three phases of passivation, the solution first becomes nearly black, then it changes to dark green, lighter green, and finally slightly amber. Chelated iron (which results from complexation) scatters light and otherwise turns contaminant-free solution amber. The coloration depends on the size of the iron sulfide particles. The results of experiments on the time dependence of iron passivation by carbon dioxide are sparse. In one test, CO₂ was added to an amine solution that already contained H₂S. The corrosion rate quickly rose and in less than one hour a new higher, but steady, corrosion rate was established. The precipitated layer of iron carbonate is very fragile, so it is unlikely to adhere as tenaciously to metal surfaces as iron sulfide does. Different abilities of different corrosion products to adhere to metal surfaces are contained within the framework of the present model.

Managing corrosion begins primarily with protecting the iron sulfide film on the metal. Fluid velocity creates sheer stress on solid surfaces—the sheer stress can be enough to rip off the imperfect iron sulfide layer, and even a low velocity seems very likely to dislodge and remove iron carbonate deposits. When this happens, fresh iron is exposed and corrosion increases in the region of high shear stress. Subsequently, the dislodged iron sulfide and iron carbonate particles can act as scouring agents and increase corrosion in other areas of the plant via the erosion-corrosion mechanism. For these reasons, the gas treating industry has generally adopted velocity limits for carbon steel piping in amine service. In carbon steel, lean amine velocities are typically held below 7–10 ft/s while, in rich amine, velocities are commonly held below 3–5 ft/s.

Even a little flashing can enormously increase the velocity in a pipe, and the high velocity flow from the passing biphase scours the iron sulfide layer. This results in drastically shorter piping life. Although less accurately than for single-phase flows, it is still possible to provide reasonable estimates of corrosion rates in two-phase flows, and this is done in the present model.

Heat stable salts (HSSs) affect corrosion rates by altering solution speciation, and they chemically exacerbate corrosion by chelating iron and destroying the passivating iron sulfide layer. The reactions pertaining to chelation are:



There are very few quantitative measurements of the effect of HSSs on corrosion and, although data on certain HSSs are included in the current model, more data over a broader range of species and conditions would be very welcome.

Mechanistic Model

Our mechanistic, chemistry-based corrosion rate model was developed on the basis of the sequence of steps outlined earlier, namely, diffusion of the corrosive agent to the surface,

adsorption, reaction with iron, and either deposition or diffusion of products away from the surface. The parameters in the model have been regressed to a large number of corrosion measurements taken in well-controlled bench-scale flow apparatuses and static equipment using both H₂S- and CO₂-rich amine systems and sour water. Although the details of the quantitative model are proprietary to Optimized Gas Treating, Inc., the corrosion rate in the model is a function of the important factors:

$$\text{Corrosion Rate} = f(a_{H_2S}, a_{HS^-}, a_{CO_2}, a_{HCO_3^-}, v, T)$$

Here: a_{H_2S} = activity of dissolved H₂S
 a_{HS^-} = activity of bisulfide ion in solution
 a_{CO_2} = activity of dissolved CO₂
 $a_{HCO_3^-}$ = activity of bicarbonate ion in solution
T = absolute temperature
v = superficial fluid velocity in pipe

The model contains many of the elements that comprise the tenets discussed earlier and in a form consistent with the various steps occurring during corrosion, including adsorption isotherms and the temperature dependence of both reaction kinetics and turbulent diffusion. The model even includes corrosion by water alone as a limiting case in highly dilute solutions. The rate limiting step may be reaction kinetics or may be diffusional, depending on whether corrosion is in the initial phase or in the later pseudo-steady stage. Data for corrosion by H₂S were correlated over the range of conditions shown in Table 1. Although scatter in the data is sizeable if for no other reason that repeatable corrosion rate measurements are hard to make, especially by weighing coupons, Figure 1 shows that this model fits the measurements over nearly four orders of magnitude. Data for corrosion by CO₂ were taken in both a static cell and in a flow apparatus. Parameter ranges and a parity plot comparing the model with measurements of corrosion in CO₂-only amine systems are shown in Table 2 and Figure 2, respectively.

Table 1 Parameter Ranges for Corrosion by H₂S

Parameter	Range
Lab velocity	0 – 25 m/s
Temperature	55 – 120°C
Sour water	1– 30 wt% NH ₄ HS equivalent 3.5 –10 bar H ₂ S partial pressure
Amines	MEA (18–30 mass %) DEA (30 mass %) MDEA (45 mass %) 0.1 – 0.8 mole H ₂ S/mole amine
Calculated pH	6.0 – 9.5

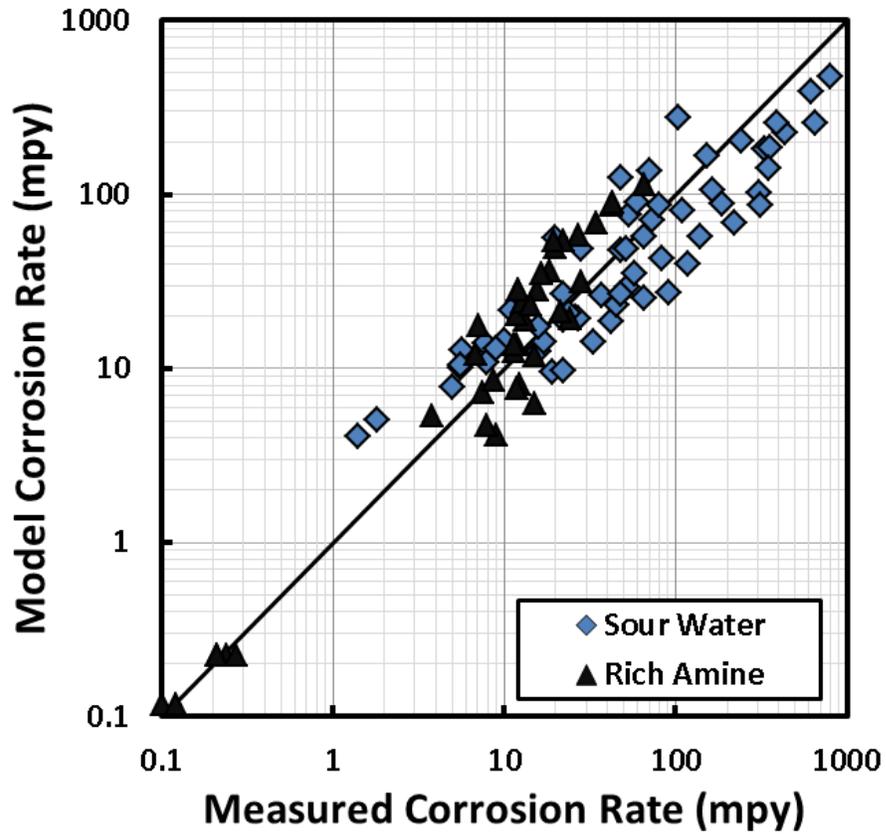


Figure 1 Parity Plot of Measured vs. Model Corrosion Rates in Rich Amine and Sour Water Systems Containing H₂S Only

Table 2 Parameter Ranges for Corrosion by CO₂

Parameter	Range
Lab Velocity	0 – 13.5 m/s
Temperature	20 – 160°C
Amines	MEA (6–30 mass %) DEA (10–40 mass %) AMP (9–35 mass %) MDEA (35 mass %) 0.0 – 0.4 mole CO ₂ /mole amine
Calculated pH	8.4 – 11.36

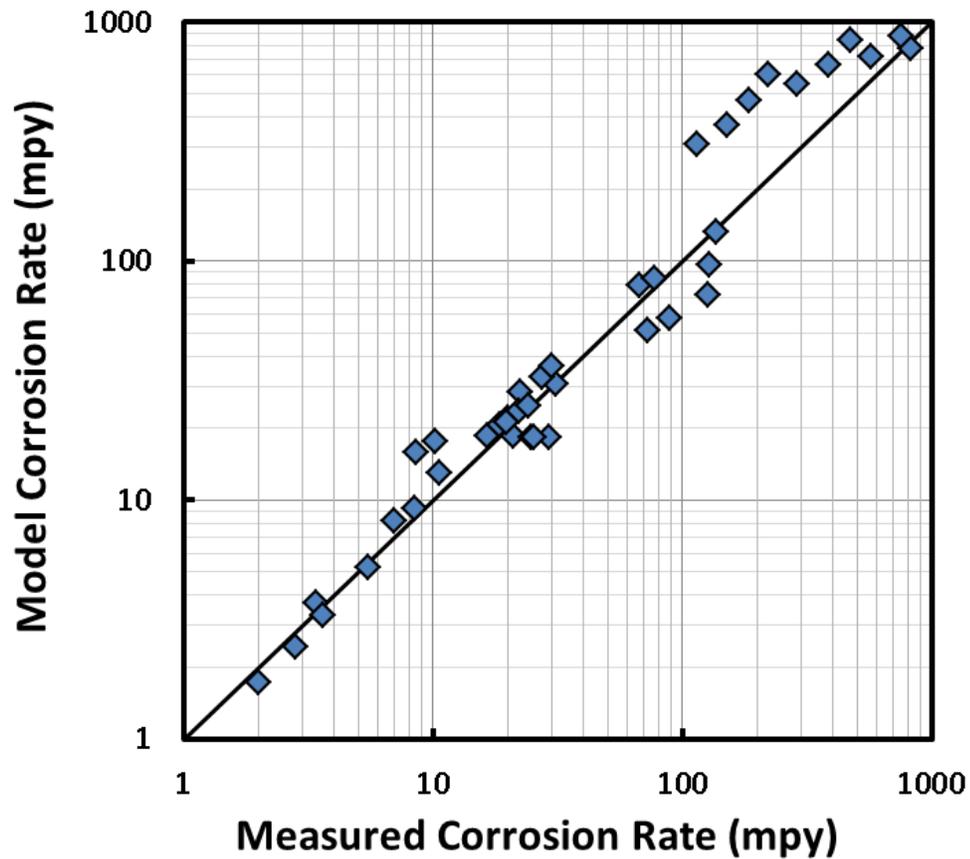


Figure 2 Parity Plot of Measured vs. Model Corrosion Rates in Amines Containing CO₂ Only

In developing the model, Arrhenius kinetics was found to fit the data well. This is in agreement with the proposal of Cummings et al.³ These authors assumed a power law kinetics expression in terms of molecular and ionic concentrations. We have found similar power law kinetics, but we have also found that the data are better represented using component activities rather than concentrations. Note that *neither amine strength, nor amine type, nor ammonia content appears explicitly in the corrosion model*, although these parameters have significant effects on the concentrations (more correctly, the activities) of the corrosive ionic species in solution because speciation depends on the amine type, its concentration, and the temperature. Corrosion of several metallurgies is included in the model's capabilities, as is the corrosion of both straight pipe and common pipe fittings. The model's mechanistic underpinning lends it not only to confident interpolation, but also to reliable extrapolation to conditions well outside measured ranges.

Recognizing the difficulty in making accurate, highly reproducible corrosion rate measurements (even in well-controlled laboratory and pilot-scale apparatuses, as witnessed by the scatter in the data), and given the fundamental basis for the current model, it appears that for amines the predicted corrosion rates are within a factor of two or so of reality. The same can be said for corrosion caused by carbon dioxide. What is unknown is the extent, if any, to which these two acid gases interact with each other in determining corrosion rates.

Case Study

The case study is of a small fuel gas treater in a refinery. Current operations treat 6 MMscfd of 5% hydrogen sulfide in predominantly methane. In preparation for a 50% increase in H₂S to 7.5 mol%, and to provide a baseline, current corrosion rates in various parts of the unit are first assessed using the corrosion rate model. Then the corrosion effects of various ways of handling the higher H₂S concentration are to be assessed. Thus, the study is in two main parts:

1. A corrosion walk-through of current operations
2. Assessing the effect on corrosion of increasing the H₂S in sour feed gas by 50% and handling the increased acid gas load by:
 - (a) Doing nothing beyond operating as normal
 - (b) Increasing DEA strength from 30% to 40%
 - (c) Increasing DEA circulation rate and reboiler duty by 50%

Figure 3 provides a flow diagram of the unit together with some of the relevant process conditions. Numbered corrosion coupons are indicated by the letters CC inside the hexagon shaped symbols. The corrosion coupons are described by the kinds of data shown in Figure 4. Either flow velocity in the pipe is calculated from a specified diameter, or the pipe is sized for a given maximum line velocity.

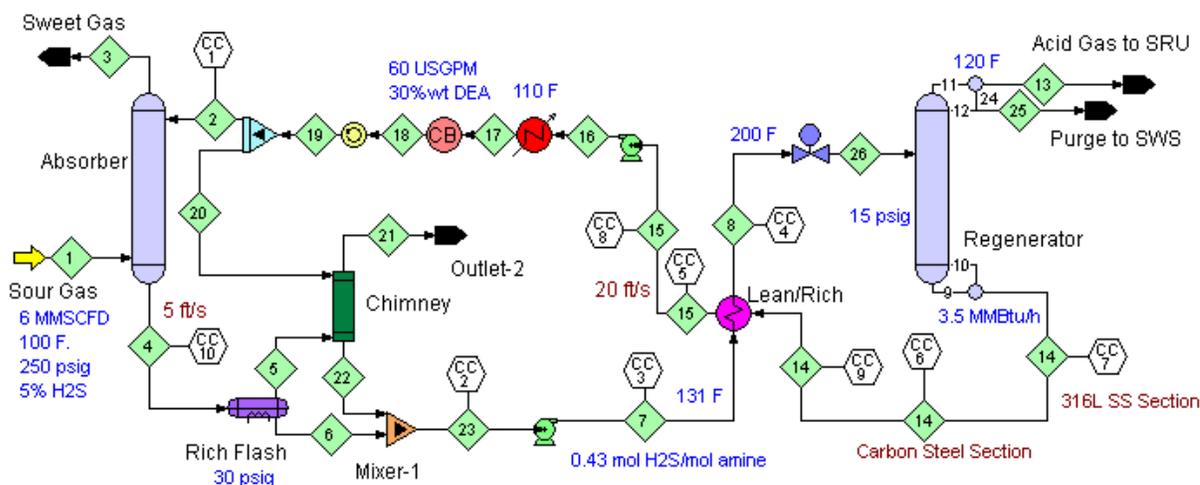


Figure 3 Process Flow Diagram of Fuel Gas Treating System

Besides carbon steel, other materials include 304L 316L stainless and the alloys 2205, 2507, 825 and C-276. Corrosion rate predictions are made at stream conditions of temperature, flow velocity and the detailed solution speciation. Note that in this particular treating unit, the original mild steel piping in the hot lean amine line going to the cross-exchanger was replaced with 316L as a result of earlier corrosion.

(a) Data Describing Corrosion Coupon

Corrosion Coupon-1 [Corrosion Coupon Block]

Stream Monitored	2	
Piping ID *	1.775	in
Velocity	7	ft/s
Roughness Factor	0.0018	in
Material	Carbon Steel	
Corrosion Allowance	0.125	in
Flow Regime	Liquid only	
Reynolds Number	56554.291	

Piping Configuration	Corrosion Rate (mpy)	Service Life (Years)
Straight Pipe	3.579	34.929
3-D Bend	3.742	33.404
90° Elbow	4.179	29.910
Weld protrusion	4.451	28.083

(b) Corrosion Rate Predictions

Figure 4 Typical Corrosion Coupon Data and Results

DEA Unit Walk-through

Table 3 shows corrosion rate predictions for straight pipe for each of the 10 corrosion coupons inserted into various sections of pipe in several locations in the plant as shown in Figure 3.

Table 3 Corrosion Rate Predictions at Current Operating Conditions

Coupon	Stream	Description	Diam (in)	Velocity (ft/s)	Temp. (°F)	H ₂ S Loading	Material	Cor. Rate (mpy)
1	2	Cold Lean	1.775	7	110	0.020	CS	3.6
2	23	Warm Rich	2.87	3	131	0.390	CS	3.7
3	7	Warm Rich	2.22	5	131	0.390	CS	4.1
4	8	Hot Rich	2.26	5	200	0.390	CS	9.1
5	15	Cooled Lean	1.125	20	188	0.020	304L	1.3
6	14	Hot Lean	1.93	7	254	0.020	CS	14.6
7	14	Hot Lean	0.93	30	254	0.020	316L	1.6
8	15	Cooled Lean	1.59	10	188	0.020	CS	9.2
9	14	Hot Lean	2.95	3	254	0.020	CS	11.8
10	4	Warm Rich	2.11	5	131	0.432	CS	4.2

The highest corrosion rate of 14.6 mpy is in a carbon steel section of pipe in the hot lean amine coming from the regenerator's reboiler. With a 0.125 inch corrosion allowance, the life of a straight section of this pipe is only 8.5 years and a weld protrusion corrodes away in only seven years. This is why part of the line was replaced with 316L metallurgy several years earlier. Even at a flow velocity of only 3 ft/s the hot lean remains very corrosive and, with a corrosion allowance of 0.125 inches, it provides a carbon-steel pipe life of only 10 years. The carbon-steel hot, rich-amine line has an expected life of 13 years, and it bears watching when the hydrogen sulfide in the sour gas increases. This base case is now used to assess the effect on corrosion of processing a more sour gas.

When the hydrogen sulfide in the sour gas is increased from 5 to 7.5 mol%, the treated gas remains at 14 ppmv H₂S because the absorber is unquestionably lean-end pinched. However, the rich amine loading increases from 0.432 mol/mol to 0.638 mol/mol. Table 4 shows predicted corrosion rates under these new circumstances in when the rich loading is allowed to increase, but no steps are taken to keep it below the recommended 0.4–0.45 mol/mol range.

Table 4 Corrosion Rate Predictions at 7.5 mol% H₂S, Other Conditions Unchanged

Coupon	Stream	Description	Diam (in)	Velocity (ft/s)	Temp. (°F)	H ₂ S Loading	Material	Cor. Rate (mpy)
1	2	Cold Lean	1.775	7	110.0	0.020	CS	3.6
2	23	Warm Rich	2.87	3	140.5	0.576	CS	4.8
3	7	Warm Rich	2.22	5	140.7	0.576	CS	5.3
4	8	Hot Rich	2.26	5	200	0.576	CS	14.0
5	15	Cooled Lean	1.125	20	200	0.020	304L	1.4
6	14	Hot Lean	1.93	7	254	0.020	CS	14.6
7	14	Hot Lean	0.93	30	254	0.020	316L	1.6
8	15	Cooled Lean	1.59	10	200	0.020	CS	10.2
9	14	Hot Lean	2.95	3	254	0.020	CS	11.8
10	4	Warm Rich	2.11	5	144	0.638	CS	6.3

The corrosion rate in the hot rich line increases from 9.1 to 14.0 mpy; the service life drops from 13 years to just 9 years for straight pipe, and to less than 8 years for 90° elbows. Very little else has changed significantly. Other than changing the hot rich line to stainless (which would also allow higher line velocity, i.e., smaller pipe size), it appears safe just to accept the increased H₂S in the sour gas and continue processing as usual (although increased attention from the inspection department would certainly be prudent). Next, it may be instructive to look at the two options of increased amine strength and increased circulation rate within the existing piping.

Table 5 shows predicted corrosion rates at 50% higher solvent flow and 50% higher reboiler duty to strip it. The corrosion rate in the carbon steel hot lean line has increased now from 14.6 to 16.4 mpy but because of the lower rich loading, the hot rich line has now a slightly lower corrosion rate.

Table 5 Corrosion Rate Predictions at 7.5 mol% H₂S, 50% Higher Circulation Rate

Coupon	Stream	Description	Diam (in)	Velocity (ft/s)	Temp. (°F)	H ₂ S Loading	Material	Cor. Rate (mpy)
1	2	Cold Lean	1.775	10.5	110	0.021	CS	4.0
2	23	Warm Rich	2.87	4.5	130	0.391	CS	4.0
3	7	Warm Rich	2.22	7.5	130	0.391	CS	4.6
4	8	Hot Rich	2.26	7.5	200	0.391	CS	10.2
5	15	Cooled Lean	1.125	30	189	0.021	304L	1.4
6	14	Hot Lean	1.93	10.5	254	0.021	CS	16.4
7	14	Hot Lean	0.93	45	254	0.021	316L	1.5
8	15	Cooled Lean	1.59	15	189	0.021	CS	10.7
9	14	Hot Lean	2.95	4.5	254	0.021	CS	12.9
10	4	Warm Rich	2.11	7.5	132	0.432	CS	4.3

Table 6 shows the effect of increasing the DEA strength from 30 to 40 wt% while maintaining the original solvent flow rate and reboiler duty. In this case, increasing the DEA strength and keeping the flow velocities fixed appears to offer less risk for increased corrosion. The reduced loading of the hot rich solvent from 0.58 to 0.43 also reduces the corrosion rate. In other words, flow velocity is a more important parameter than amine strength, all other things being roughly equal. However, it matters to keep rich loading below 0.5 mol H₂S/mol amine.

Table 6 Corrosion Rate Predictions at 7.5 mol% H₂S, 40 wt% DEA

Coupon	Stream	Description	Diam (in)	Velocity (ft/s)	Temp. (°F)	H ₂ S Loading	Material	Cor. Rate (mpy)
1	2	Cold Lean	1.775	7	110	0.022	CS	3.8
2	23	Warm Rich	2.87	3	142	0.434	CS	4.5
3	7	Warm Rich	2.22	5	134	0.434	CS	5.1
4	8	Hot Rich	2.26	5	200	0.434	CS	10.4
5	15	Cooled Lean	1.125	20	204	0.022	304L	1.5
6	14	Hot Lean	1.93	7	256	0.022	CS	14.9
7	14	Hot Lean	0.93	30	256	0.022	316L	1.6
8	15	Cooled Lean	1.59	10	204	0.022	CS	10.7
9	14	Hot Lean	2.95	3	256	0.022	CS	12.0
10	4	Warm Rich	2.11	5	146	0.480	CS	5.6

Regenerator Corrosion: Downcomers

Visual observation of flows in large-scale air-water hydraulic simulators indicates that vapor-liquid mixtures pass over outlet weirs and flow into downcomers at 0.5 to 2 ft/s velocity. This velocity range is probably also typical of liquid velocities against and across tower walls. The next question to be considered, at least in a semi-quantitative way, is how corrosion rates of downcomer aprons and tower walls vary with location in a regenerator. By drawing off a small portion of the liquid flow from a tray and passing it through a pipe that is sized for, say, 0.5 ft/s, we can assess expected corrosion rates as a function of distance down the column. For the cases corresponding to the data in Tables 3 and 4, Figure 5 show how corrosion rate is predicted to change from tray to tray in the regenerator at the two sour gas concentrations of 5% and 7.5% hydrogen sulfide. Figure 6a shows that predicted corrosion rate almost exactly tracks with tray temperature — lower temperature corresponds to lower corrosion rate.

Figures 6b and 7 confirm that the corrosion rate is independent of H₂S level in the raw gas, and is dependent solely on temperature. The H₂S loading of the solvent is changing from 0.42 down to 0.03 across the regenerator trays (Figure 6b) so one might expect to see *higher* corrosion rates on the *upper* few stripping trays; however, the opposite is predicted to occur. As shown by Figure 6, H₂S loadings are all below 0.45, and any effect of solvent loading is far outweighed by solvent temperature. It can be concluded that provided the H₂S loading of the solvent is kept below a reasonable upper limit, corrosion rates of mild steel tower shells and downcomer aprons are primarily a function of temperature.

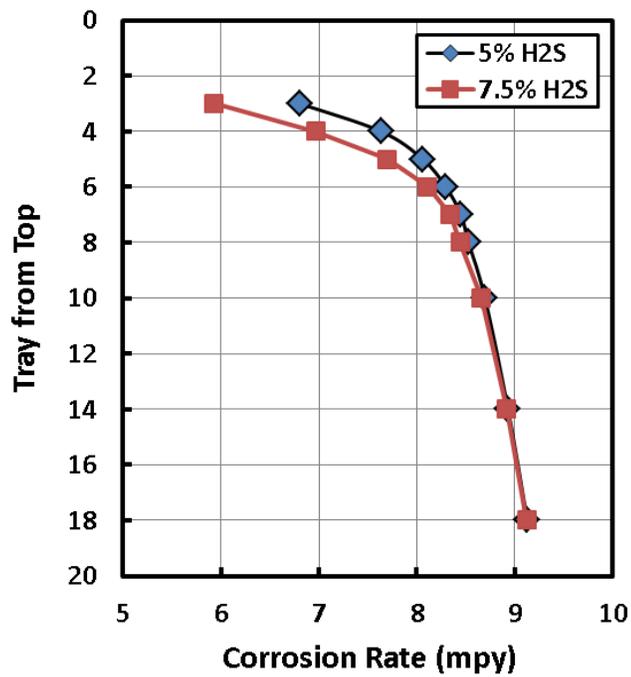
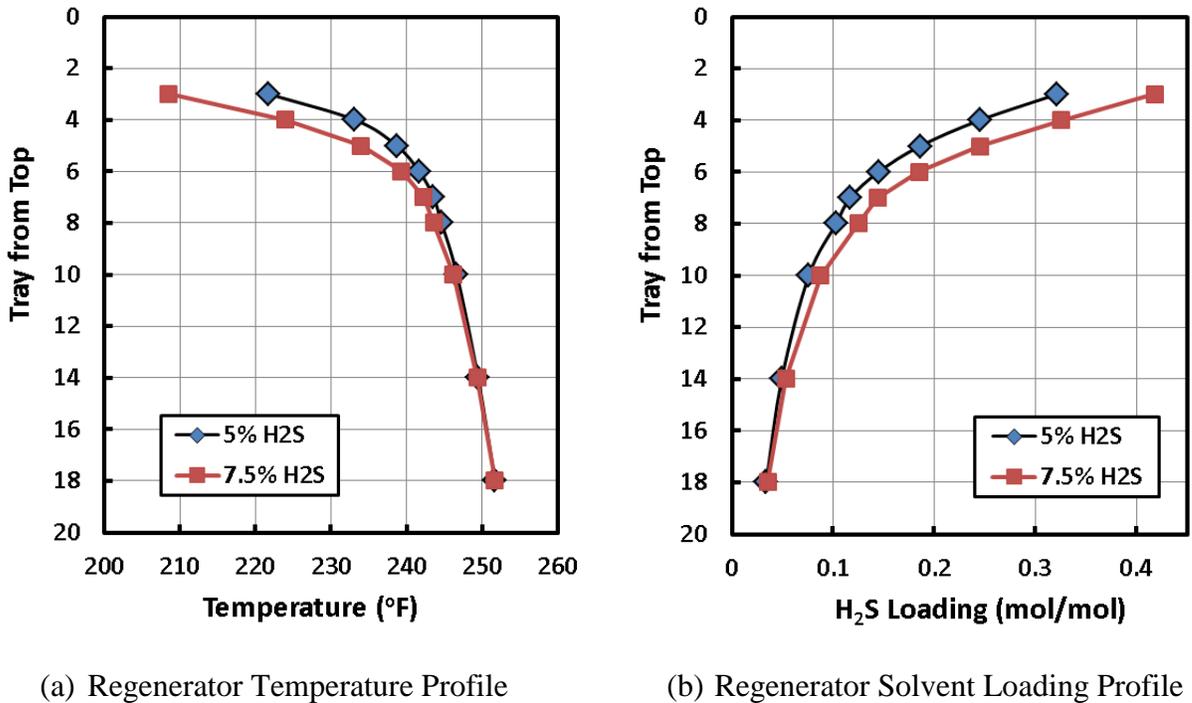


Figure 5 The Predicted Corrosion Rate Varies from Tray to Tray



(a) Regenerator Temperature Profile

(b) Regenerator Solvent Loading Profile

Figure 6 How Conditions Vary Across the Stripper

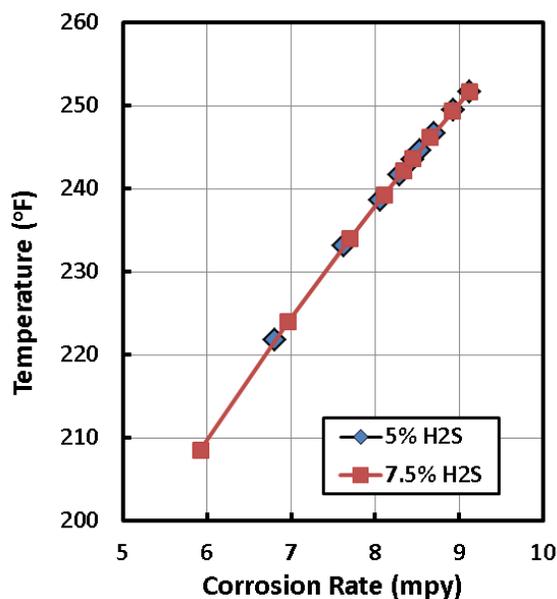


Figure 7 Corrosion Rate Dependence on Tray Temperature at Two Different Hydrogen Sulfide Loading Profiles

In Conclusion

The corrosion rate model is based on fundamental parameters: wall shear stress, activities of corrosive components (speciation), metallurgy, HSSs, and temperature. The model is correlated to mostly proprietary data, but with some public corrosion rate data as well, collected for carbon steel and alloys in several amines and sour waters containing hydrogen sulfide and carbon dioxide. In line with measurements, the model predicts the highest corrosion rates are associated with the hottest solvents, almost without regard to acid gas loadings as long as loading is below about 0.5. This is evidenced by Corrosion Coupons 2, 3 and 10 in Tables 3–6. Lean solvent following the cross exchanger is cooler and shows reduced corrosion. The fully-cooled solvent shows relatively much slower corrosion. The temperature dependence corresponds to an approximately Arrhenius relationship with temperature

Corrosion rates in the regenerator closely follow the regenerator temperature profile. It appears that provided the H₂S loading is kept below a reasonable limit, corrosion rate is independent of the variation in solvent loading on various trays across the height of the column.

References

1. Cummings, A. L., Waite, S. W., Nelsen, D. K., *Corrosion and Corrosion Enhancers in Amine Systems*, paper presented at Brimstone Sulfur Conference, Banff, Alberta, April, 2005
2. Polderman, L.D., Dillon, C.P., Steel, A.B., *Oil & Gas J.*, May 16, 1955, p. 180.
3. Cummings, A.L., Veatch, F.C., Keller, A.E., *Corrosion and Corrosion Control Methods in Amine Systems Containing H₂S*, NACE Corrosion 1997 Conference, Paper No, 341.