

# HYDROCARBON ENGINEERING

## SIMULATE GAS TREATING HOLISTICALLY

EXCLUSIVE  
3-PART SERIES

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OPTIMIZED GAS TREATING,  
INC., USA**, PROVIDE A HOLISTIC  
PERSPECTIVE ON GAS TREATING  
SIMULATION OVER A THREE PART  
SERIES OF ARTICLES.

PART ONE FOCUSES ON AMMONIA  
SYNGAS PURIFICATION.

# PART 1

## SIMULATE GAS TREATING HOLISTICALLY

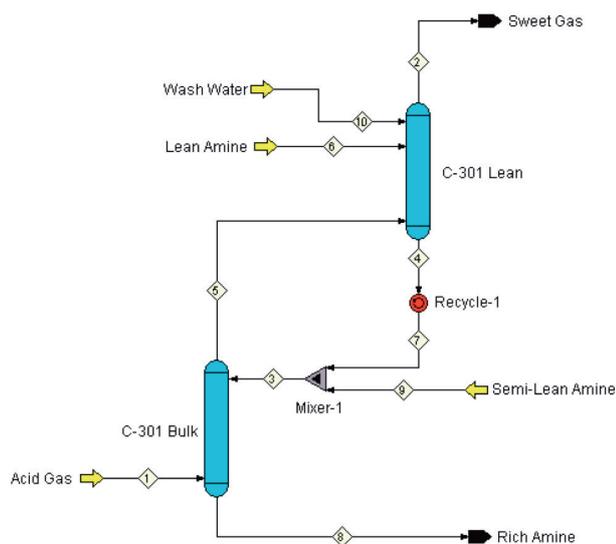
**D**eveloping a good understanding of the behaviour and performance of any gas treating plant depends on being able to look at the plant in its entirety rather than focusing on some specific detail or set of details too soon. Although especially true in troubleshooting, this applies equally well to grass roots plant design. This article is the first in a three part series analysing real world cases where the behaviour of the gas treating plant is somewhat counterintuitive.

The main tool used in the analysis is mass transfer rate simulation to create a virtual plant on a computer. This virtual plant is a digital image of the real plant on a scale and to a level of detail that permits performance and

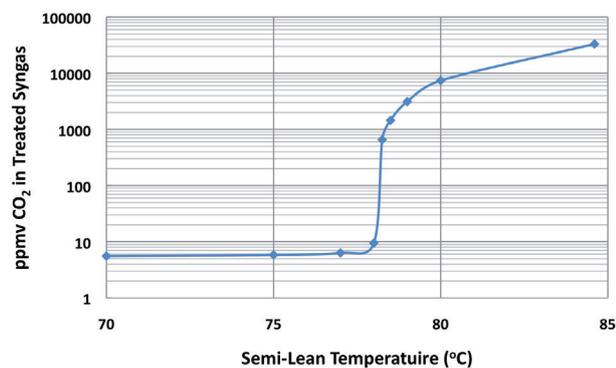
behaviour to be examined as under a microscope. What such a model is capable of revealing can be quite intricate, detailed and satisfying from an engineering science perspective. Perhaps of more pragmatic importance, it can lead to the solution of difficult troubleshooting exercises on the one hand, or to the selection of correct operating conditions for a new plant on the other. Each case centres on CO<sub>2</sub> removal: part one looks at a specific situation of piperazine promoted MDEA in an ammonia plant; part two considers a performance map for a CO<sub>2</sub> capture plant using MEA; and in part three, a sensitivity analysis of the simulated performance of a proposed LNG plant using piperazine activate MDEA is assessed. These cases are all connected by the commonality of fairly reactive solvents and, in every case, unusual behaviour is predicted and explained, with paramount emphasis on the importance of parametric studies in amine unit design so that difficult operating conditions are avoided in the as built plant.

## Role of piperazine

Functionally, MDEA acts as a sink for the hydrogen ions produced by CO<sub>2</sub> hydrolysis so it has an important effect on the vapour/liquid equilibrium of CO<sub>2</sub> in MDEA solutions. However, MDEA itself does not react with CO<sub>2</sub> despite it being alkaline, which therefore increases the reactive hydroxyl ion



**Figure 1.** Two stage absorber for CO<sub>2</sub> removal in an ammonia syngas plant.



**Figure 2.** Effect of semi lean temperature on performance of the two stage absorber shown in Figure 1.

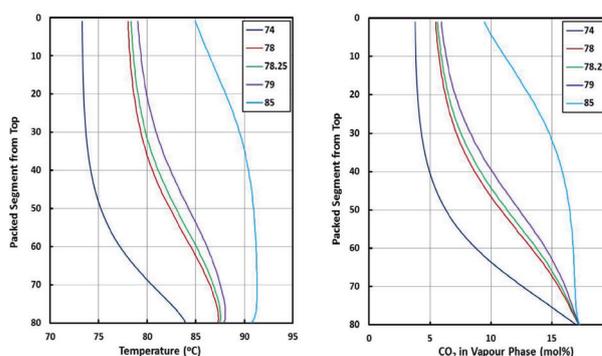
concentration in aqueous solution. Consequently, MDEA has very limited ability to influence the CO<sub>2</sub> absorption rate beyond what it would be into just pure water. At best, MDEA might be said to catalyse CO<sub>2</sub> hydrolysis by providing a more alkaline environment than water, but its inherent lack of reactivity makes it incapable of removing CO<sub>2</sub> quickly. For this reason it simply cannot reduce CO<sub>2</sub> to low concentrations in reasonable packed column heights or numbers of trays. This makes it an excellent choice for slipping CO<sub>2</sub> but a bad choice even for moderately deep CO<sub>2</sub> removal.

Piperazine, on the other hand, reacts extremely rapidly with CO<sub>2</sub> (some 10 times faster than MEA), which makes it an excellent promoter when used even in relatively small concentrations with MDEA. It increases solvent reactivity enormously. Thus, piperazine promoted MDEA is a very commonly used solvent in LNG and syngas (hydrogen and ammonia plant) applications. It is offered by major solvent suppliers under a variety of trade names.

## Piperazine promoted MDEA

It has already been established that there is a cliff like boundary limiting a treating plant's stable operating region when using blends of piperazine and MDEA. The instability manifests as a sudden loss in treating ability. Its main cause is the fact that the reaction zone in the absorber extends over only a few trays, and it is in this very zone that the majority of the CO<sub>2</sub> is absorbed. Under normal conditions, well away from the unstable region, this reaction zone exists over the bottommost trays, with the upper trays serving to polish the gas and remove the last vestiges of CO<sub>2</sub>. However, if the solvent rate is reduced to a low enough value, the point is reached where the reaction zone starts to creep up the column. The column still treats quite satisfactorily, so there is no warning or indication at all of what is about to happen with a further reduction in solvent flow. The zone suddenly pops out the top of the absorber and, as it is quite narrow (because of high solvent reactivity), breakthrough results in sudden and unexpected failure to come even close to meeting treating specifications. For example, a 10 ppmv CO<sub>2</sub> gas near the critical area can suddenly go to 1000s of ppm with less than a 0.1% drop in solvent flow rate.

Figure 1 shows a simplified process flow diagram of the ammonia absorber system operating in a Middle Eastern ammonia plant. This is a two stage (split flow) absorber using semi lean and fully lean piperazine promoted MDEA streams to treat syngas to less than 500 ppm CO<sub>2</sub>. (Each column contained 11 m of 50-5 Hiflow Rings.) The fully lean amine flow rate is fairly



**Figure 3.** Bulk absorber temperature and CO<sub>2</sub> profiles with semi lean temperature as parameter.

low, being just enough to polish the gas leaving the bulk column where the most of the CO<sub>2</sub> is removed. The semi lean flow is approximately six times the fully lean value.

The process setup consists of two separate columns in series: it could have been simulated more simply (avoiding an external recycle stream) as a single, albeit quite tall, column, but it was actually built as two columns. The plant suffers from inadequate cooling of the very high flow rate, semi lean stream because of an undersized trim cooler. The problem is worst in summer when cooling air temperatures are highest. Plant operations were unable to get the semi lean below approximately 84 °C, even with the lowest of ambient air temperatures. The plant was failing to meet the 500 ppm treated syngas specification by an extremely wide margin (i.e. the treated gas was many 1000s of ppm CO<sub>2</sub>).

The process licensor recommended a maximum semi lean stream temperature of 74 °C, so the operators were quite surprised that only a 10 °C deviation resulted in such a huge loss in treating; they felt something had been seriously compromised in the plant. The usual culprits such as packing disintegration, foaming and system contamination were suspected, but no plausible reason for the failure could be found. The packing supplier felt the semi lean temperature was too high, but was at a loss to explain the magnitude of its effect on treating.

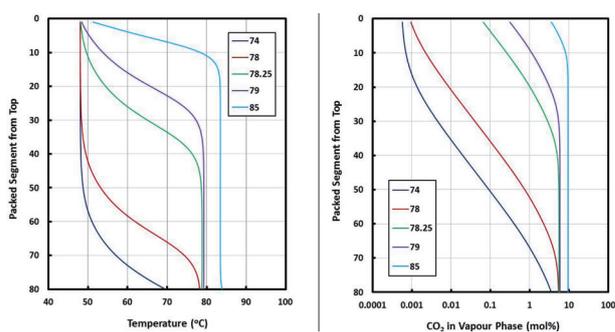
Figure 2 shows the simulated performance of this unit as a function of semi lean solvent temperature. There is an operational cliff at approximately 78 °C where performance changes by two orders of magnitude with a mere 0.25 °C temperature change. As long as the temperature of the high flow rate semi lean is kept below approximately 78 °C, the treated gas will more than meet a 500 ppmv CO<sub>2</sub> treating specification by quite a wide margin. Thus, the recommended maximum temperature of 74 °C would be far enough away from the cliff that stable operation could certainly be maintained. However, it is not possible to come even close to meeting the treated gas specification with the semi lean feed temperature in the 80 °C range. To understand the reason for this remarkable and seemingly counterintuitive behaviour, it is important to examine temperature profiles in the bulk and lean absorbers at semi lean temperatures in the vicinity of the cliff.

Figure 3 shows temperature and CO<sub>2</sub> profiles in the bulk absorber with semi lean solvent temperature as parameter. Apparently, there is plenty of solvent capacity and mass transfer surface for the bulk column to be lean end pinched and therefore to be removing as much CO<sub>2</sub> as the solvent strength, loading and temperature will allow, at least at a low semi lean temperature. When the CO<sub>2</sub> partial pressure at the top of the column in equilibrium with the entering solvent is setting the treating level, the operation is said to be lean end pinched. When the treating level is determined by solvent capacity, the operation is termed rich end pinched. In the present case, as the semi lean temperature climbs, the bulk column stops being lean end pinched and starts to become rich end pinched with the amount of CO<sub>2</sub> removed determined by solvent capacity; at high semi lean temperatures the solvent has less and less capacity for CO<sub>2</sub>, to the point where the CO<sub>2</sub> rejection rate from the column is quite high. What does this mean for the lean polishing column?

Figure 4 shows temperature and CO<sub>2</sub> profiles for the lean column as a function of semi lean solvent temperature. As the

semi lean temperature rises, the lean column is increasingly being fed with more CO<sub>2</sub> than it can absorb. At low semi lean temperatures, the gas entering the lean column is fairly low in CO<sub>2</sub> and the fully lean solvent has no trouble absorbing all of it, right to the solvent lean loading's own limit. However, as the semi lean temperature rises, the bulk column is able to absorb less, so more and more CO<sub>2</sub> is presented to the lean absorber: it becomes overwhelmed and exhibits sudden failure at semi lean temperatures only moderately higher than the process licensor's recommendation. The CO<sub>2</sub> profile breaks through with little or no warning. Thus, ultimately the lean column's performance becomes fixed strictly by solvent capacity limits, not mass transfer rates, and the treated gas shows extremely high CO<sub>2</sub> levels and extraordinary sensitivity to the semi lean temperature.

In a sense, this is all caused by trying to squeeze the last drop



**Figure 4.** Lean absorber temperature and CO<sub>2</sub> profiles with semi lean temperature as parameter.

of performance from the system, which pushes one or more operating parameters (be it temperature, flow or composition) over the edge. Part of the key to understanding what is really going on is using a simulator of such high quality that it reveals the inner workings of the real plant via digital translation into a virtual plant.

The temperature profiles in both the bulk and lean absorbers show radically different forms when the semi lean temperatures are separated by only 0.25 °C. When the semi lean temperature is slightly too high, the bulk absorber cannot adequately handle the CO<sub>2</sub> load and the excess CO<sub>2</sub> spills over into the lean absorber. Because the lean absorber is expected to perform its final polishing duty with only a small solvent flow, this small flow is immediately overwhelmed and the lean column goes into a rich end pinch condition. This sends the CO<sub>2</sub> in the final treated gas from 10 ppmv to approximately 700 ppmv. Operation with a semi lean temperature of 78 °C is completely unstable in the sense that even the slightest variation of inlet CO<sub>2</sub> content, solvent rate, gas rate or temperature in the wrong direction will cause the outlet gas to go completely off specification. This plant did not have the luxury to operate differently, because the original plant was designed much too tightly.

This type of failure can also occur in LNG applications using piperazine promoted MDEA, where a slight reduction in solvent flow too close to the stable operating limit can send the treated gas from a few ppmv CO<sub>2</sub> to well over 1000 ppmv.<sup>1,2</sup> This kind of behaviour is apparently well known to the more astute solvent vendors and process licensors, but the never ending incentive to provide competitive bids always leaves the temptation to offer

designs that may be just a little too tight. The result can be a plant inadvertently designed with built in instabilities.

In part two of this series, simulation of a pilot plant for post combustion CO<sub>2</sub> capture will be examined. An appreciation of the difference in operating conditions for carbon capture versus conventional deep CO<sub>2</sub> removal will be developed to understand how reboiler duty varies with circulation rate through the plant in order to achieve a specified percentage of CO<sub>2</sub> removal. It is important to think in a holistic manner and not simply to apply conventional wisdom with a view to understanding seemingly odd behaviours.

## Conclusion

Plant simulation and analysis are holistic when the entire treating plant is examined all at once, or at least the isolated equipment item is examined and analysed using a simulation tool that is based on considering all the factors that affect its performance, without idealisations or unwarranted approximations. Using ideal stages, or any approach that refers to ideal stages, immediately leaves the simulation analysis open to debate because guesswork is inherently involved, either in assumed or calculated efficiencies applied to ideal stages, or even sometimes by introducing a phony residence time or thermal efficiency per theoretical stage. Regardless of the exact approach, at least two assumptions have to be made when ideal stages underlie a model, and there is no way to prove the assumptions are valid unless the answer is already known. This immediately disqualifies the resulting analysis as nothing more than a gross approximation to the truth, and frequently a completely faulty one.

Using an ideal stage model with either overall or Murphree efficiencies begs the question, 'Where do the efficiencies come from?' Neither tables of efficiencies nor reliable correlations of such efficiencies exist, so they can come only from what is termed 'experience'. Efficiencies vary widely from tray to tray, from component to component and with specific operating conditions of temperature, pressure and flow rates, as well as the tower internals details (tray design, packing type and size). Experience will not necessarily come in handy to solve new problems when the answer is not already known. Therefore, simulation based on necessarily overall or component efficiencies will yield results that leave an unsettled feeling when the analysis is completed.

There is no longer any need to use such antiquated and unreliable models. Modern computing power is more than adequate to solve columns and flowsheets in only a few seconds using models that are very soundly based in engineering science. The ProTreat® simulator is a proven commercial tool that allows this to be done for gas treating in a computationally robust and reliable manner. There is no reason to continue accepting the risk of failure by using inadequate tools to develop unreliable designs when cost effective tools are readily available. 

## References

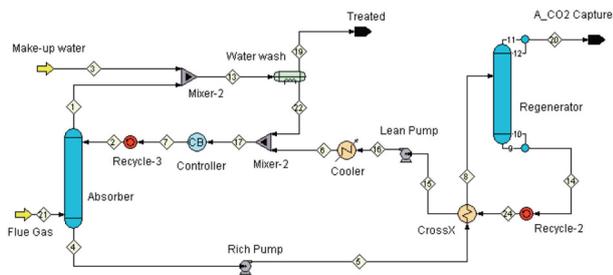
1. WEILAND, R.H. and HATCHER, N.A., 'Foundations of Failure', *Hydrocarbon Engineering*, December 2011.
2. WEILAND, R.H. and HATCHER, N.A., 'Stable Operating Limits in Amine Treating Units', *Proceedings of the Lawrence Reid Gas Conditioning Conference*, Norman, Oklahoma, February 2011.

PART TWO CONSIDERS A PERFORMANCE MAP FOR A CO<sub>2</sub> CAPTURE PLANT USING MEA.

## PART 2 SIMULATE GAS TREATING HOLISTICALLY

**T**his article is the second in a three part series examining real world cases in which the behaviour of a treating plant is somewhat counterintuitive. Part one focused on a treating problem experienced in the CO<sub>2</sub> removal section of an ammonia plant. There, a seemingly small departure from an operating condition recommended by the process technology's supplier resulted in failure to treat adequately by a very large margin. In the present article, attention is turned to the surprising relationship between reboiler duty and solvent circulation rate in a pilot plant for CO<sub>2</sub> capture operating to meet a specified fractional CO<sub>2</sub> removal. To understand the behaviour, one must appreciate the difference in operating philosophy between conventional deep CO<sub>2</sub> removal and the removal of a limited amount of CO<sub>2</sub>.

As pointed out in part one, developing a good understanding of the behaviour and performance of any gas treating plant requires examination of the plant in its entirety rather than focusing too soon on some specific detail. The main tool used in the analysis of the present case is mass transfer rate simulation, specifically the ProTreat® simulator, which creates a virtual plant on a computer. The virtual plant is a digital image based on all the real factors that affect performance. The scale is microscopic and quite comprehensive. The model reveals intricate detail and is pleasing from an engineering science perspective. More pragmatically, it can lead to the solution of difficult troubleshooting exercises on the one hand, or to the selection of correct operating conditions for a new plant on the other. After use, the engineer is left with the knowledge of what makes a given plant tick. Here in part two, a performance map for a CO<sub>2</sub> capture plant using MEA is considered.



**Figure 1.** CO<sub>2</sub> capture pilot plant.

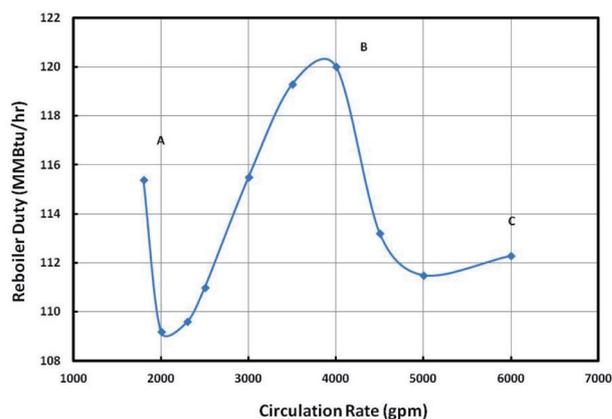
## Post combustion CO<sub>2</sub> capture

Figure 1 shows the pilot plant processing scheme. The plan was to remove 90% of the CO<sub>2</sub> from the entering flue gas; the engineers were interested in exploring the extent to which reboiler duty (regeneration energy) depended on solvent circulation rate. The absorber contained 10 m depth of structured packing with a specific surface area of approximately 250 m<sup>2</sup>/m<sup>3</sup>. The cross exchanger was to operate with a 10 °F temperature approach. When a series of simulations were run over the solvent flow range from 1800 – 6000 gal./min and the reboiler duty was adjusted to achieve 90% removal, the rather unusual looking curve shown in Figure 2 was obtained. There are two minima, a maximum and two asymptotes at the far left hand and right hand sides. The engineers conducting the study initially believed the simulator predictions were in gross error. Based upon results from an equilibrium stage, efficiency based simulator, they expected to see a single minimum in the reboiler duty followed by a steady increase in reboiler energy input. However, this odd and surprising behaviour is quite real.

The reason for this seemingly strange behaviour is that the absorber moves from a rich end pinch condition in region A at the far left of Figure 2 (the desired mode of operation for carbon capture), to a lean end pinch in region C at the far right (the desired mode of operation for deep CO<sub>2</sub> removal). The solvent flows in region C are far higher than needed for only 90% CO<sub>2</sub> removal: these flows are much more typical of a conventional MEA column used to treat to low CO<sub>2</sub> levels. To make such high flows work economically for CO<sub>2</sub> capture, the reboiler duty must be greatly reduced so that the lean solvent loading is very high. At a high solvent rate, a high lean loading is completely consistent with what is needed for the equilibrium partial pressure of CO<sub>2</sub> over the lean solvent to have a high enough value that it results in only 90% of the CO<sub>2</sub> being removed.

A significant amount of information can be gleaned from a seemingly simple temperature profile. Accurate temperature (and composition) profiles are the purview solely of a genuine mass transfer rate model, and they indicate quite accurately just how a given column is operating and where improvements, including what other treating strategies such as solvent additives, might be worthwhile.

In this case, the regions marked A, B and C in Figure 2 highlight areas where it might be worthwhile to look at temperature profiles for clues. It is no surprise that the combination of low reboiler duty and low solvent rate is an efficient way to remove 90% of the CO<sub>2</sub>. Indeed, the rich end pinch is purposely produced by low solvent flow and is exactly what is used to limit CO<sub>2</sub> absorption. However, why does a higher solvent flow eventually require more reboiler energy to be expended?



**Figure 2.** The interaction between reboiler duty and solvent flow rate to achieve 90% CO<sub>2</sub> removal in a carbon capture pilot plant.

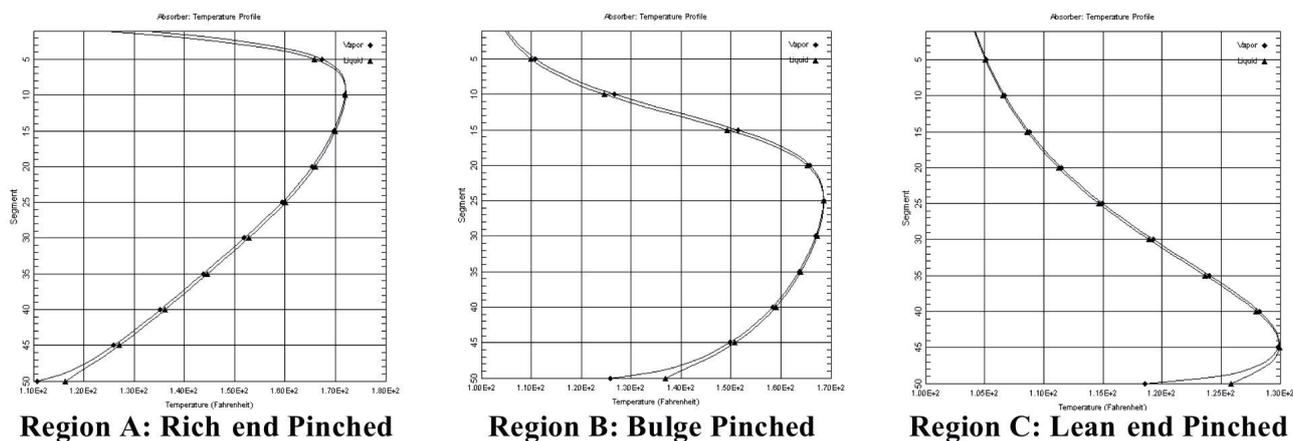
The answer is that as solvent flow increases, the temperature bulge spreads to much of the interior of the absorption column making the centre region so hot that it can do little or no absorbing. Therefore, it leaves more CO<sub>2</sub> in the gas, unless the solvent gets stripped cleaner by more reboiler steam. At point B, the bulge temperature is nearly 170 °F and only the ends of the column are effective in removing CO<sub>2</sub>. As the solvent rate goes still higher, beyond the peak at 4000 gal./min, the bulge continues to move down the column and treating becomes increasingly lean end pinched. The column has entered an operating region more suited to deep CO<sub>2</sub> removal than carbon capture, but it is forced to remove only limited CO<sub>2</sub> by being grossly under stripped.

Under lean end pinched conditions, solvent lean loading controls the outlet CO<sub>2</sub> concentration and as the solvent rate increases, higher and higher lean loadings are adequate for treating to remove only 90% of the CO<sub>2</sub>. Despite the fact that the solvent flow is higher so more reboiler energy will go into heating the rich feed to its bubble point on the stripper's feed tray, less stripping still requires less energy and the curve falls through another minimum. It should be apparent that to the left of the first minimum a further decrease in solvent rate will need to be offset with a substantial decrease in lean loading, hence higher reboiler duty. To the right of the second minimum, a higher solvent flow will require a gradually increasing reboiler duty to heat the solvent to the feed tray temperature, the solvent lean loading being fairly constant beyond 6000 gal./min.

Maxima and minima are always caused by the existence of (at least) two factors opposing each other, with one factor dominating on one side of the minimum and the other dominating on the opposite side. The factors at play here are:

- Lean end versus rich end pinch.
- Solvent net loading capacity versus the solvent flowrate.

There are essentially two pairs of factors. There is more than one minimum and the situation is obviously complex. In part, the complexity is a result of one trying to examine carbon capture (left hand side) and high purity treating (right hand side) at the same time. These two regions and goals are at opposite ends of the treating spectrum and looking at them simultaneously (i.e. holistically) leads to the suspicion that the simulator is just plain wrong. However, if these two regions are considered separately



**Figure 3.** Temperature profiles in the three regions marked in Figure 2.

from each other, the analysis becomes much simpler, and a lot of the complexity goes away.

The performance curve shown in Figure 2 is for 30% MEA, a fairly reactive solvent. There is no operational cliff and there are no instabilities, although at first glance the operating curve as mapped out by the engineers involved was a little surprising. Having two regions of lowest energy treating is simply a result of lean end versus rich end pinching in the absorber at the two extremes of operation. However, to appreciate this, it must be possible to look not just at the absorber, but holistically at the whole treating plant, including the regenerator, which plays at least as important a part in determining the operating curve as the absorber does. Thus, the regenerator in this case was crucial in allowing an understanding to be gained of these two extreme operating regions and how one transitions into the other. It also

allows engineers and researchers to see what goes on in the intermediate region where the absorber is neither rich end nor lean end pinched, but is instead bulge pinched. Coming to terms with a seemingly strangely shaped operating curve also required a close look at how temperature profiles responded to changing solvent rates and reboiler duties. Yet the real key was being able to identify and distinguish between rich end, lean end and bulge pinch conditions.

Part three of this series will attend to an apparent anomaly in performance in a plant using piperazine promoted MDEA to treat LNG. Again, the performance curve in terms of how treated gas CO<sub>2</sub> level responds to total solvent flow rate with all other conditions (including regeneration) remaining fixed shows some surprising twists. Furthermore, it will be shown that packing behaves very differently from trays in this application. [f3](#)

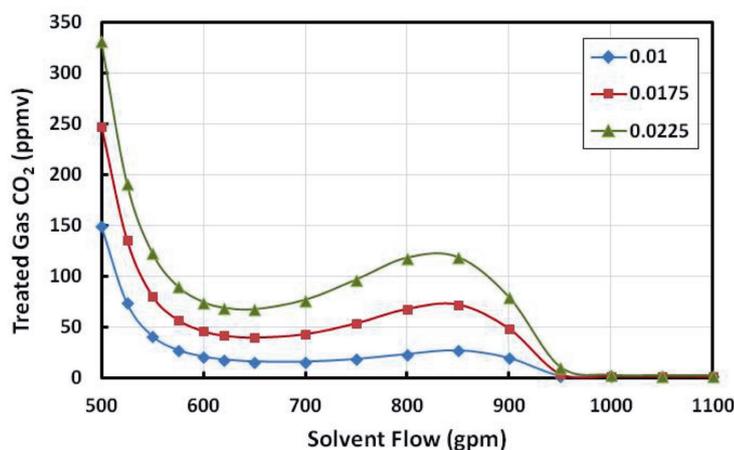
THIS THIRD AND FINAL PART ANALYSES THE REMOVAL OF CO<sub>2</sub> WITH A PIPERAZINE PROMOTED MDEA BASED SOLVENT IN AN LNG PLANT.

## PART 3 SIMULATE GAS TREATING HOLISTICALLY

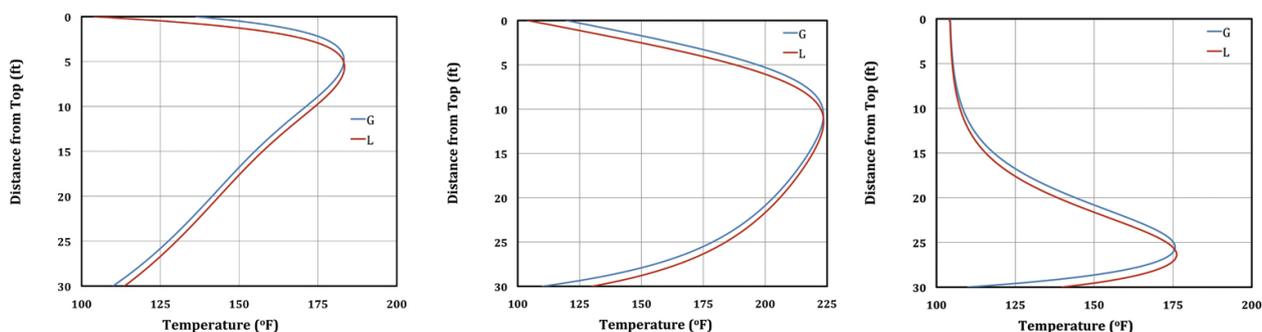
This is the final article in a three part series dealing with CO<sub>2</sub> removal in a variety of real world cases in which the behaviour of the gas treating plant is somewhat counterintuitive. Part 1 focused on a treating problem experienced in the CO<sub>2</sub> removal section of an ammonia plant. There, a seemingly small departure from an operating condition recommended by the supplier of the process technology resulted in failure to treat adequately by a huge margin. Part 2 looked at the surprising relationship between reboiler duty and solvent circulation rate in a pilot plant for CO<sub>2</sub> capture operating to meet a specified fractional removal. To understand the behaviour in this case, one had to develop

and appreciate the difference in operating philosophy between conventional deep CO<sub>2</sub> removal and the removal of a limited amount of CO<sub>2</sub>, specifically only 90%. Part 3 will analyse the removal of CO<sub>2</sub> with a piperazine promoted MDEA based solvent in an LNG plant where the target gas purity is <50 ppmv CO<sub>2</sub>.

As this article will show once again, developing a good understanding of the behaviour and performance of the treating plant depends on being able to look at either the whole plant or a specific tower and not focusing too soon on some specific detail or set of details. Although especially true in troubleshooting, this philosophy applies equally well to grass roots plant design, the focus of this article. The primary tool for the discussion in this article is mass transfer rate simulation, specifically the ProTreat® simulator, which creates a virtual plant on a computer. The virtual plant is a mirror image of the real plant on a scale and to a level of detail that permits examination of detailed performance and behaviour. What such a model is capable of revealing can be quite intricate and complete, and very satisfying from an engineering science perspective. Perhaps of more practical importance, it can lead to the solution of a difficult troubleshooting exercise, on the one hand, or to the selection of correct operating conditions for a new plant or column on the other.



**Figure 1.** LNG absorber performance as a function of solvent rate at a series of solvent lean loadings (see Legend).



**Figure 2.** Temperature profiles corresponding to solvent flows in Figure 1. (From left to right: Rich end pinched at 500 gpm, bulge pinched at 800 gpm and lean end pinched at 1000 gpm).

## Piperazine promoted MDEA

Although MDEA is alkaline and therefore increases the reactive OH<sup>-</sup> ion concentration, MDEA itself does not react with CO<sub>2</sub>, so MDEA has very limited ability to influence the CO<sub>2</sub> absorption rate. At best, MDEA might be said to catalyse CO<sub>2</sub> hydrolysis by providing a more alkaline environment than water, but its inherent lack of reactivity makes it incapable of removing CO<sub>2</sub> quickly. For this reason, MDEA simply cannot reduce CO<sub>2</sub> to low concentrations in reasonable packed column heights or numbers of trays. This makes it an excellent choice for slipping CO<sub>2</sub> but a terrible choice even for moderately deep CO<sub>2</sub> removal. Piperazine, on the other hand, reacts extremely rapidly with CO<sub>2</sub> (10 times faster even than MDEA), which makes it an excellent promoter when used in relatively small concentrations with MDEA. Thus, piperazine promoted MDEA is quite commonly used in LNG applications because it allows deep removal but has the low energy benefits of MDEA. Major solvent vendors offer it under a variety of trade names.

## LNG absorber

This case is a sensitivity analysis of the design of an absorber in an LNG plant to treat a feed gas containing mostly methane (84%), with 10% ethane, 4% propane and 2% CO<sub>2</sub> on a dry basis to a specification of <50ppmv CO<sub>2</sub>. The proposed absorber

contains 60 ft of IMTP-50 random packing and it was sized in each simulation for 80% of flood. Three sets of simulations were run at a series of solvent rates, but with each set having a constant value of CO<sub>2</sub> lean loading, as shown by the legend in Figure 1.

At each lean loading, the absorber fails to treat adequately if the solvent rate is too low. This is as one should expect, because at too low solvent flows the solvent has inadequate capacity. This makes the column rich end pinched and allows a significant amount of CO<sub>2</sub> to pass through the column without being absorbed. The temperature profile at 500 gpm is shown in Figure 2. Increased solvent flow improves treating, and if the lean loading is low enough there is adequate absorbing capacity to achieve <50 ppmv CO<sub>2</sub>. But as solvent flow is increased further, the temperature profile deforms into the shape shown in the middle plot of Figure 2 for 800 gpm. At this flow rate, the absorber

is bulge pinched, in that the temperature in the central part of the column is so high that only the ends are effective in removing CO<sub>2</sub>. The centre part of the column does nothing, and the column behaves as though it has perhaps 30 or 40 ft of packing, not the 60 ft that is really there. As the solvent rate is increased further, the temperature bulge gets pushed further down the column, and the absorber becomes lean end pinched, where treating is determined primarily by the solvent's lean loading, i.e., by vapour/liquid equilibrium conditions at the lean (top) end.

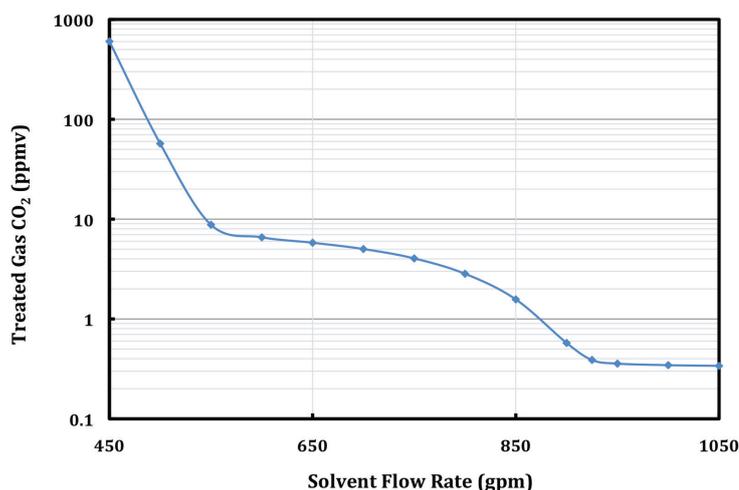
In this particular case, the treated gas was 40 – 45 ppmv CO<sub>2</sub> over the flow range from 600 – 700 gpm but the way to respond to the gas going off specification may not be to increase solvent rate, but rather to decrease it, or perhaps to increase reboiler steam or hot oil flow. Without a detailed operating diagram such as the one in Figure 1, operators could probably not do much more than guess at the correct response and hope for the best. And if the engineering contractor uses an equilibrium stage simulator of whatever ilk (ideal stages with Murphree efficiencies, or ideal stages with kinetic corrections and user estimates of ideal stage residence times and stage thermal efficiencies), none of this would be apparent at all and the design would have an unwelcome element of uncertainty and surprise.

The behaviour of trayed columns is a little different from ones containing packing. With packing, as the solvent flow is increased, the wetted, interfacial area rises as well, and the mass rate therefore increases with solvent flow for this reason. With trays, gas/liquid interfacial area for mass transfer is only a relatively weak function of liquid rate and the performance curve typically looks like Figure 3. (Note the logarithmic scale). There is no maximum because the liquid rate dependence of area is insufficient to drive higher absorption rates and torment the central, flat region, into becoming a peak. Nevertheless, there are still lean end, bulge, and rich end pinch conditions at the low, medium and high solvent rates like those shown in Figure 2. Under the conditions for the simulation results shown in Figure 3, none of this really matters. However, if the lean loading were to become too high, one might have to operate at nearly twice the flow of very lean solvent to achieve the <50 ppmv specification.

The difference between packing and trays and, indeed the effect of packing type and size can be very important in the design of an LNG facility. If the simulator is not genuinely mass transfer rate based, all of this will be missed. Trays and packing will all be treated as ideal stages; the differences between them will not be apparent (indeed, one packing will be as good as another and all packing will give the same results), the design will be subject to great uncertainty, and the plant may not work at all.

## Conclusion

Plant simulation and analysis are holistic when the entire treating plant is examined all at once, or at least the isolated equipment item is examined and analysed using a simulation

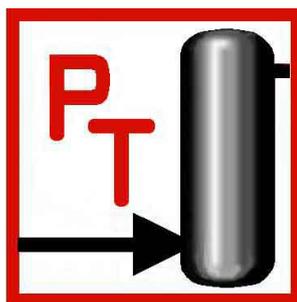


**Figure 3.** Performance of a trayed LNG absorber with 0.015 lean loading.

tool that is based on considering all the factors that affect its performance, without idealisations or unwarranted approximations. Using ideal stages or any approach that refers to ideal stages immediately leaves the simulation analysis open to debate because guesswork was inherently involved, either in assumed or calculated efficiencies or through the number of ideal stages assumed or even sometimes insofar as introducing a specious residence time per hypothetical stage. Regardless of the exact approach, when ideal stages underlie any model, at least two assumptions have to be made, and there is no way to prove an assumption was valid unless the answer is already known. This immediately disqualifies the resulting analysis as nothing more than a gross approximation to the truth.

Using an ideal stage model with either overall or Murphree efficiencies begs the question: 'Where do the efficiencies come from?'. Neither tables of efficiencies nor reliable correlations of such efficiencies exist, so they can come only from what might be termed 'experience'. Efficiencies vary widely from tray to tray, from component to component and with specific operating conditions of temperature, pressure and flow rates as well as the tower internals details (tray design, packing type and size). Experience will not necessarily come in handy to solve new problems when the answer is not already known. Therefore, simulation based on necessarily overall or component efficiencies will yield results that leave an unsettled feeling when the analysis is completed.

There is no longer any need to use such antiquated and unreliable models. Modern computing power is more than enough to solve columns and flowsheets in only a few seconds using models very soundly based in engineering science. The ProTreat<sup>®</sup> simulator is the only available, proven, commercial tool that allows this to be done for gas treating in a computationally robust and reliable manner. There is no reason to continue taking the risk for failure using inadequate tools to develop unreliable designs when the right tool is available at modest cost. 



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