

Choosing Tower Internals for LNG, Shale Gas, and Tail Gas Treating¹

Ralph H. Weiland
Optimized Gas Treating, Inc.
P.O. Box 125
Clarita, OK 74535

Nathan A. Hatcher
Optimized Gas Treating, Inc.
Technical Development Center
142 Cimarron Park Loop, Suite C
Buda, TX 78610

Jenny Seagraves
INEOS Oxide
A Division of INEOS Americas LLC
Plaquemine, LA 70765

SUMMARY

It is well known that for best performance, tail gas treating requires the utmost in selectivity. Shale gas often contains a small amount of H₂S in a much larger concentration of CO₂. Sometimes this too demands highly selective treating to meet transmission pipeline specifications. This paper shows that the inherent selectivity of trays in such applications can be greatly improved by simple changes in tray design parameters.

Treating gas in an LNG plant to reduce CO₂ to a level sufficient for liquefaction of the gas (typically 50 ppmv) is a completely non-selective process and one that often uses an absorber containing either random or structured packing. This paper compares structured packings of various sizes with trays, and shows the profound effect of packing type and size on absorber performance. There is an optimal packing size for each application. Specifying too large a packing will result in failure to meet the treating goal. Choosing a packing size that is unnecessarily small is not only costly because of higher packing cost; it invites plugging and reduces tower capacity, too. This paper provides a quantitative way to determine just what the right size is.

INTRODUCTION

One of the most neglected areas in gas treating is the proper selection of tower internals in both absorbers and regenerators. The literature is replete with data on vapor-liquid equilibrium in a host of aqueous amine systems containing both single amines and mixtures. Considerable data also exist on a variety of physical, thermal, and transport properties such as solution density, viscosity, heats of reaction, heat capacity, and diffusion coefficients. The kinetics of the reaction between dissolved CO₂ and various amines has been measured and reported in terms of Arrhenius parameters numerous times for all the commonly-used amines. There is no question that such parameters are important in interpreting laboratory measurements of absorption rates and all of them have been used in that setting. Far fewer of these parameters have been applied in the design of commercial equipment using traditional methods such as equilibrium stage calculations, for the simple reason that ideal stages know nothing about what, if anything at all, is in the column. However, acid gas absorption and solvent regeneration are carried out commercially in columns *with real internals*. Apart from hydraulic considerations, the mass transfer performance of tower internals unfortunately has tended to be almost completely ignored. This is despite the fact that the mass transfer characteristics of the internals plays a central role, and at least as important a role in setting tower performance as do phase equilibrium and reaction heats. The translation from numbers of ideal stages to actual trays counts and to the required

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depths of structured or random packing has traditionally been done solely on the basis of experience. This approach seems to work fairly well in light hydrocarbon separations, for example, where tray efficiencies are fairly constant and well known and where vendor data exist on HETPs and HTUs for various packings for the systems of interest. Amine treating, however, is undoubtedly one of the processes least amenable to extrapolation into areas where experience is lacking. When parts-per-million specifications must be met on product gases, selecting the wrong packing size or bed depth can result in a failed design.

For many years, packing has had a bad reputation in absorption and distillation at high pressure. Part of the reason is inattention to proper distributor design. Another is the persistent and still unresolved difficulty in translating ideal stages to actual packed bed depths and the selection of a particular commercial packing. However, packing is now being used increasingly in gas processing for several reasons. For a given size, packed columns tend to permit higher throughput than trays. And in offshore operations such as FPSO and FLNG, columns using structured packing are much less susceptible to the effects of rocking motion caused by wave action. It is particularly critical to select the *right* packing type and size in FLNG because of severe weight and footprint restrictions, so the need for reliability and accuracy is even higher.

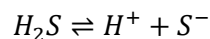
It is no longer necessary to engage in any form of guesswork when it comes to designing towers containing one or more of a large range of packing types and sizes. The key is to simulate the column as the mass transfer device it really is. This entails a mass transfer rate approach. Although rate calculations require knowledge of the mass transfer characteristics of the internals (mass transfer coefficients for both phases, and the interfacial area) this kind of information is available in the literature as well as within the ProTreat® amine treating simulator. As we shall see, ProTreat simulation makes packing performance just as easy to *predict* as trays. The emphasis here is on *prediction* because absolutely no parameters need to be guesstimated by the engineer to come up with the right answer. The simulation is out-of-the-box reliable and accurate.

In what follows, we briefly lay out the important factors that decide selectivity, then summarise previously reported findings on the effect of certain tray design parameters on selectivity. Attention is then turned to the performance of tower packings, first by comparing simulation with performance measured in an LNG plant, and then by studying the effect of packing size for a particular structured packing used in the same service.

FACTORS THAT AFFECT SELECTIVITY

The solubility of H₂S in a given amine is not enormously different from the solubility of CO₂ in the same amine. Physical solubility is not what makes one amine selective and another not. There are two major factors that control selectivity. One is *how fast CO₂ reacts* with the amine, and indeed, whether it reacts at all or not. The other is the *mass transfer characteristics of the contacting device*.

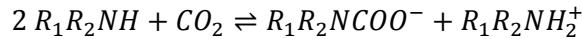
As soon as it absorbs into the solvent, H₂S dissociates instantaneously into hydrogen and bisulphide ions:



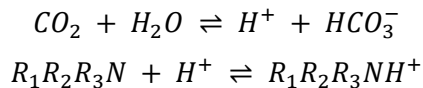
The reaction is so fast it throws the mass transfer resistance completely into the gas phase—the liquid offers no resistance to mass transfer. What is meant by the resistance being only in the gas phase is that the mass transfer resistance is caused by diffusional limitations with the mass transfer (absorption) rate determined by the gas-phase mass transfer coefficient. Both molecular and turbulent (eddy)

diffusion contribute to the value of the gas-phase coefficient, although in commercial equipment the gas flow is usually quite turbulent, so the turbulence level sets the value of the mass transfer coefficient.

Primary and secondary amines are carbamate formers, with dissolved CO₂ reacting directly with the amino group:



The reaction lowers the CO₂ concentration in the bulk liquid and it steepens the CO₂ concentration gradient by providing an additional means for CO₂ to disappear besides by diffusion. The faster is the reaction, the lower is the liquid phase resistance to CO₂ absorption. On the other hand, tertiary amines such as N-methyldiethanolamine (MDEA) are capable only of accepting a proton such as is liberated by H₂S dissociation and by CO₂ hydrolysis in water:



Mopping up the hydrogen ion with MDEA shifts the hydrolysis reaction to the right, i.e., it increases the solvent's capacity for CO₂; however, the hydrolysis reaction itself is slow and does essentially nothing to steepen the CO₂ concentration gradient and speed up the absorption process². The resistance to CO₂ absorption remains completely in the liquid phase and is just as strong as it would be without MDEA. All the MDEA does is to increase greatly the absorption capacity of the solvent.

Selectivity is determined by the reaction rate of CO₂ with the amine (or any other reactive component) and, therefore, it can be said that reaction kinetics is a controlling factor in establishing the inherent selectivity of the solvent. However, the selectivity of the process is also controlled by the mass transfer characteristics of the column internals, i.e., by the gas- and liquid-phase mass transfer coefficients and the active (wetted) interfacial area. These coefficients vary with molecular diffusion and such properties as phase viscosity, but the main influencing parameter is the intensity of turbulence in the phases near the gas-liquid interface. Thus, selectivity can be adjusted by selecting the right tower internals and operating the equipment under conditions most conducive to the treating goals in mind. Selectivity can be increased by keeping the liquid flowing smoothly and quietly, and forcing the gas into as turbulent a state as possible. This suggests structured packing as the most selective, random packing next, and trays as least selective because the liquid on a conventionally operated tray is so agitated and turbulent from being churned by the injected gas. However, as shown by Weiland and Hatcher (2012), there are tray modifications possible that will result in a very quiescent liquid flow.

DESIGNING TRAYS FOR IMPROVED SELECTIVITY

Many gas treating applications use high L/G ratios because there is a large amount of acid gas to be removed from the raw gas and the solvent has limited capacity. Such a limit is often imposed by the need to keep acid gas loadings³ low enough to avoid excessive corrosion. When very high liquid rates are used, downcomer hydraulic restrictions may dictate the use of multi-pass trays to avoid downcomer choke flooding conditions. However, high liquid loads⁴ are not the only reason to select a multi-pass tray. As shown by Weiland and Hatcher (2012), when liquid flows across a tray in the form of a spray,

² The real CO₂ reaction is with the OH⁻ ion formed by dissociation of water; however the OH⁻ concentration is small and this makes for a slow reaction.

³ Loading is defined to be moles of acid gas (individual components or total) per mole of total amine. Sometimes the basis is per mole of amino group.

⁴ The liquid load on a tray refers to the volumetric liquid flow rate per length of overflow weir (USgpm/ft or m³/m·h).

the droplets have a size of about 1 mm diameter and the liquid within each drop is hardly flowing (internally) at all. Dissolving CO₂ reaches the droplet interior mainly through the very slow process of molecular diffusion. On the other hand, gas flow around the drops becomes very turbulent reaching Reynolds numbers on the order of many hundreds (the transition to turbulence occurs at a Reynolds number of unity based on drop diameter). This is precisely what is needed to improve selectivity—much lower liquid phase coefficients and larger gas phase coefficients so that the CO₂ absorption rate will be hindered and H₂S absorption rates increased.

Whether a tray operates in the spray or froth regime is very closely connected to the weir liquid load. The transition from froth (liquid continuous, gas dispersed) to spray (liquid dispersed, gas continuous) is gradual and is generally held to begin at a weir load of around 45 m³/m·h (60 USgpm/ft). As the weir load reaches values as low as 5 or 10 m³/m·h selectivity can increase dramatically. It should be noted, however, that great improvements in selectivity are possible only for solvents that inherently have very low reactivity towards CO₂ because when they are reactive, the preponderance of the CO₂ disappears in the liquid via reaction, not diffusion, so lowering liquid turbulence has a much weaker effect on CO₂ absorption rate.

Legend has it that one should try to avoid the spray regime at all costs because in the spray regime entrainment is higher so tray hydraulic capacity will greatly suffer. This is quite erroneous⁵. In fact, the gas velocity required to entrain 1 mm droplets greatly exceeds the gas velocity at which massive liquid entrainment occurs from trays operating with froths. The reason for this ill-founded wisdom is probably (a) the errors made by tray designers in failing to provide positive vapour seals at the bottom of downcomers and (b) trying to operate trays with dynamically-sealed downcomers far below the liquid loads for which they were designed. In both cases, the downcomer bottoms will become uncovered and vapour will bypass up the downcomers and entrain the liquid they contain. In fact, to achieve remarkable improvements in selectivity, one should operate with low weir liquid loads when possible, even if that means using a multi-pass tray when hydraulically a one-pass tray is sufficient. Forcing operations into the spray regime can result in selectivity at least as good as that attained with structured packing.

PREDICTING PERFORMANCE OF RANDOM AND STRUCTURED PACKING

There are several reasons why packing may be preferable to trays:

- Pressure drop is usually lower
- Tower capacity is often higher
- Foaming is usually not as big a problem
- In applications subject to rocking motion such as FPSO and FLNG, structured packing offers better resistance to upsets caused by periodic tower tilt.
- Selectivity tends to be higher

Increased gas handling capacity can be a very significant factor in high pressure towers in situations where excessive weight and footprint have severe cost penalties. This is particularly the case in FPSO and FLNG where, indeed, tightening a design can yield large cost advantages. Reducing column diameter is certainly beneficial, but so is using only the packing depth actually required. Tower diameter is determined by hydraulics, and hydraulic performance is well documented and well understood. The

⁵ Of course, if the weir liquid load is made low enough the quantity of liquid will be so small much of it can be fragmented and entrained to the tray(s) above and nothing flows through downcomers. But this is unlikely to happen at weir loads above a few m³/m·h

same cannot be said for packed bed depth—this is determined by mass transfer, and the mass transfer performance of structured packings with chemical reactions is still a rather specialized area.

Packed columns seem always to have presented a challenge to designers, perhaps because there are so many varieties, types, and sizes of packing and perhaps even more because the experience base is so small, especially in gas treating with amines. However, today there is no reason why packed columns cannot be designed with just as much certainty and confidence as trays. The processes taking place in absorption and regeneration towers are mass transfer processes, and as long as one has access to the basic mass transfer characteristics as embodied in mass transfer coefficient correlations for the particular internals of interest, packed columns are no harder to specify and design than their trayed counterparts. The fundamental correlations contained within the ProTreat® simulator's information base have been developed from literature, vendor and research data and have been shown repeatedly to allow very accurate and reliable *predictions* of column performance without recourse to guesstimating artificial parameters such as tray efficiencies or fictitious residence times in theoretical stages. Before embarking on a case study to show the effect of the size of structured packing corresponding to a particular vendor's portfolio, it may be useful to demonstrate ProTreat's ability to predict performance in a commercial case.

Comparison with Commercial Data

The test case involves the revamp of the amine section of an LNG plant using GAS/SPEC 2020 solvent to treat a feed gas with 2.24% CO₂ at about 4 MPa. The unit was built originally with trayed columns but was unable to achieve more than 70% of the nameplate capacity, purportedly because of foaming. The process flowsheet was completely conventional with the usual absorber-regenerator combination connected to each other through a flash tank, cross exchanger, trim cooler and pumps. Analysis of the treated gas (27.7°C) showed 21.2 ppmv CO₂. Using all the known plant conditions plus the tray details, ProTreat predicted 18.2 ppmv at 27.7°C. This was what one might call an "out-of-the-box" prediction in which absolutely no parameters were guessed or estimated. The simulation data consisted of tray and column vendor drawings along with process flow, temperature and pressure measurements taken directly from the DCS. This was certainly a very encouraging result.

The cause of foaming (if indeed there really was foaming) was never determined. However, even at only 70% of the design capacity, the regenerator was running very close to the recommended 85% hydraulic flood limit so there was a question as to whether the tower was being prevented from operating at full capacity because it was perhaps undersized. Regardless of the real cause of the capacity bottleneck, the decision was taken to replace both towers with somewhat larger diameter versions, and to replace the trays with structured packing. The packing was of local manufacture, and from photographs its appearance was similar to several well-known commercial brands. The crimp size was calculated from the photographs to be 25 mm. Full-capacity plant performance data were taken in mid-2012, again using GAS/SPEC 2020 solvent. The treated gas was reported to be 7 ppmv CO₂. Out-of-the-box ProTreat® simulation indicated the unit should have been producing 1 ppmv CO₂ with the absorber and regenerator running at 23% and 36% of flood, respectively. This was more discrepancy than we would have expected based on past experience, so we set out to assess the possible cause(s). The CO₂ absorber was running in the completely mass transfer rate controlled regime—it was neither lean-end nor rich-end pinched⁶. When mass transfer rate controls absorption, and all the process parameters such as flow rates, compositions and temperatures have been verified, the reason for discrepancies must be sought in mass transfer. All that could be determined for this particular version of packing was an estimated crimp size (25 mm) and that the packing sheets were perforated and

⁶ Mass transfer rate control is indicated here by a CO₂ concentration profile that is linear on a semi-logarithmic plot

embossed, a treatment that promotes liquid spreading. We found that the reported treated gas CO₂ content, could be perfectly matched (along with all other measured data) by reducing the wetted area by about 25%. It is known (Lewis et al., 2006) that the scatter in wetted area measurements even when made in the same equipment is ± 12 to 15%. The other factor, and one that is impossible to quantify, is the efficacy of the distributor used in the column, and the frequency of liquid redistribution. Less-than-perfect liquid distribution will never generate higher interfacial area or cause better performance. In light of these factors and the inherent uncertainty in process plant measurements, a 25% reduction in area is well within the scope of reasonable (and in any event still yields very satisfactory performance from a gas quality point of view). The simulation results are pure predictions, and are certainly close enough to actual measured performance to give us considerable confidence in ProTreat's ability to predict the effect of packing type and size, for example, on CO₂ removal using reactive amines.

Effect of Packing Size on CO₂ Removal in an LNG Facility

The effect of packing size on treating for CO₂ removal can be seen by considering a range of structured packings within a particular brand series, in this case FLEXIPAC 1X, 1.4X, 2X, 3X and 4X. These packings have increasing crimp size and decreasing dry specific surface area. For the same gas and liquid flow rates, they also have somewhat different liquid-film mass transfer coefficients and, therefore, the same chemical reactions affect mass transfer rates to different extents.

Table 1 describes the raw gas common to all the cases in the study. The solvent was 50 wt% GAS/SPEC-1160 flowing at 204 m³/h and 50°C. The regenerator contained 13.7 m of type FLEXIPAC 3Y structured packing in all cases. The absorber contained 15.25 m of various crimp sizes of FLEXIPAC structured packing as described above. In all cases, both columns were sized for 70% flood regardless of the packing. The regenerator was 1875 mm diameter while the absorber diameter ranged from 2825 mm diameter with the finest packing (1X) to 1885 mm diameter with the coarsest (4X). Reboiler duty was held constant at 13.5 MW and the molar stripping ratio was typically 1.12 at the regenerator overhead.

Table 1 Raw Gas Used in Case Study

Conditions			
Temperature (°C)	37.2		
Pressure (bara)	64.4		
Flow (Nm ³ /d)	6,700,000		

Composition (mol%)			
H ₂ S	0.0001	iC4	0.020
CO ₂	2.000	nC5	0.006
C1	94.331	iC5	0.010
C2	1.900	nC6	0.030
C3	0.170	CH ₃ SH	0.0007
nC4	0.030	N ₂	1.500

Table 2 is a synopsis of the simulated treating performance with 15.25 m of each size FLEXIPAC structured packing. The lowest packing size designation has the largest specific (dry) area and also treats the gas to the lowest CO₂ level. Note that although there is an inverse relationship between area and treated gas CO₂ content (just as one should expect), it is anything but inversely linear—treated gas

quality is a *very* strong function of dry area, i.e., packing size. As ever coarser packing is used the temperature bulge moves closer to the top of the column. It also grows in size until at very large packing sizes it starts to fall again. It is also interesting to note that the ratio of wetted area to dry area of the packing grows with packing size, and can exceed unity by a considerable fraction. The same liquid flow has much larger area to spread across for small packings, thus leaving more of the packing in the dry state. But for large packings, the available area for spreading is more restricted. The wetted area can exceed the dry area because all liquid flow is not restricted just to the mechanical surface of the packing when the film becomes thick. Then the film becomes quite disrupted, large waves form and a certain amount of sparging of the gas through the liquid occurs.

In many ways, tray performance is close to the performance of the smallest packing, but this is definitely *not* because of higher surface area for mass transfer. Indeed, in this case, the equivalent surface area based on the volume between any two trays is around 130 m^{-1} whereas the column-average wetted area for 1X packing is twice as high at about 270 m^{-1} . Instead, the intense agitation on the trays gives a mass transfer coefficient almost four times larger than for the 1X packing (This also turns out to be true for all the packing sizes in the FLEXIPAC family).

Table 2 Simulated Treating Performance, Bulge Temperature, and Bulge Position

FLEXIPAK®	Dry Area	Treated Gas CO ₂	Bulge Temperature	Bulge Position from Top	Wetted/Dry Area Ratio
Packing Size	(m ² /m ³)	(ppmv)	(°C)	(m or Tray No.)	(unitless)
1X	440	0.45	88.9	13.4	0.614
1.4X	340	0.46	96.9	13.1	0.741
2X	220	1.79	114.8	9.0	0.973
3X	110	108	110.3	7.0	1.332
4X	55	516	105.4	7.0	1.691
25 Trays		0.93	88.9	Tray 20	

There are several significant learnings to be taken from this. First, structured (and random) packings are no more challenging to a true mass transfer rate based simulation than trays. Secondly, one cannot simply scale up the effect of packing size in any simple or logical arithmetic way. Packing performance is a complex function of packing size, hydraulics, and chemical reaction kinetics. Additionally, the effective wetted area active on a structured packing is not limited to the packing's dry area, so scale-up based on dry area could be wrong by several fold. In short, scale-up based on random and structured packing size is simply not feasible; for example, one cannot predict the performance of 2X packing from knowing how 1X performs. The only way forward is with a mass transfer rate-based simulation capability such as that offered by ProTreat®, soundly based on principles of mass transfer.

The variability in performance that accompanies various packings is described in greater detail in the plots of Figures 1 and 2. In Figure 1, the gas-phase CO₂ profiles for 1X and 1.4X packing are nearly coincident. Thus, it can be seen that for such small-crimp packing the final treating level is determined by the partial pressure of CO₂ in equilibrium with the entering lean amine, not by mass transfer rates. In other words, the absorber with very fine packing is completely lean end pinched. The treated gas CO₂ content is almost the same with 25 trays on 600 mm tray spacing as with 15 m of fine packing, but the absorber does not exhibit quite the same degree of lean end pinching as the packing (although it is still pinched because the top 6 trays reduce the CO₂ in the gas by less than 1 ppmv). However, with 2X and larger packing treating ceases to be lean end pinched.

As Figure 2 shows, coarser packing results in a larger temperature bulge which moves up towards the center of the column. The largest packing simply does not have sufficient area to allow rapid enough CO₂ absorption for satisfactory treating to be achieved. The temperature bulge is much larger though, because a lot of CO₂ is being absorbed locally into a relatively small volume of liquid. Insufficient physical supporting structure (dry packing surface area) for the liquid results in lower holdup volumes, therefore temperatures are higher, backpressure of CO₂ goes up and absorption rates suffer. Actually, the absorber with 2X packing is bulge pinched. This can be seen from Figure 3 where the actual CO₂ concentration in the gas at various positions almost coincides with the equilibrium values along the lower half of the absorber. There is almost no driving force for absorption and past the bottom metre or so of packing, the gas has to wait until it is nearly half way up the column before significant absorption rates can resume. It is the intensified temperature bulge that is responsible for the pinch.

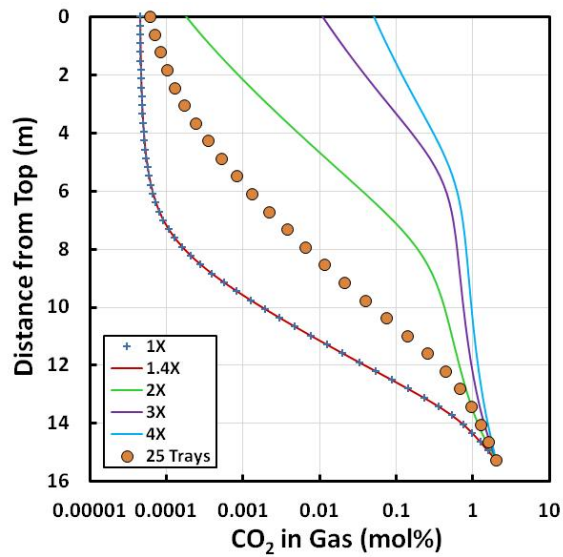


Figure 1 Absorber CO₂ Profiles in the Gas for Various Packing Sizes and for Trays

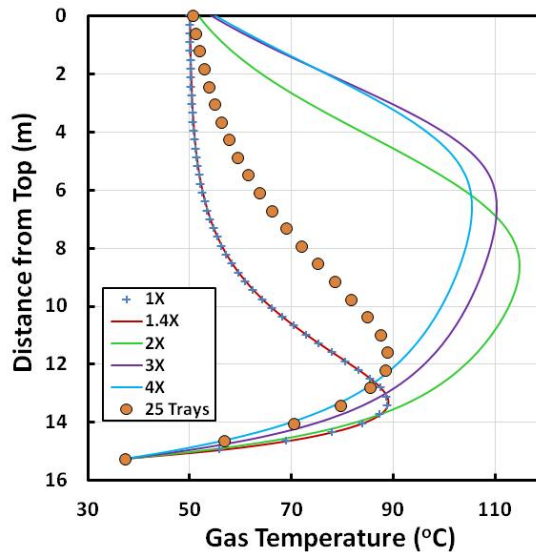


Figure 2 Absorber Temperature Profiles for Various Packing Sizes and for Trays

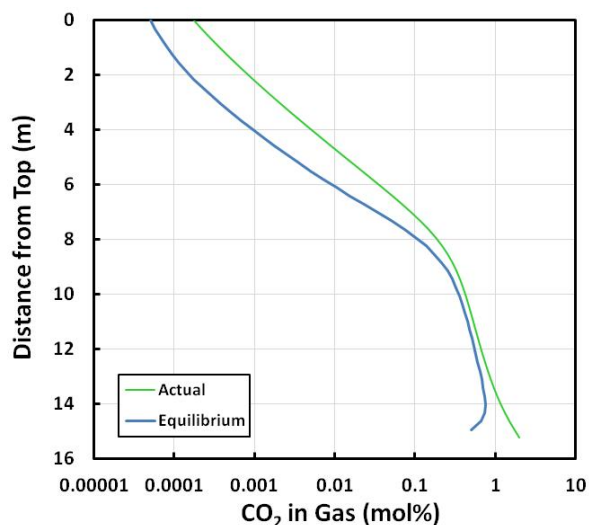


Figure 3 Actual and Equilibrium CO₂ Concentrations in the Gas at Various Positions in the Absorber. Packing is FLEXIPAC® 2X.

ProTreat's mass transfer rate model was able to match this field data closely without fitting any adjustable, artificial parameters such as tray efficiencies or hypothetical residence times on theoretical stages. This is a good indication that the simulator's predictive power can be relied upon to give the right answers with confidence for other applications. As long as accurate parameters pertinent to the particular packing of interest are in the simulator's database, that packing's performance can be reliably predicted and just the right packed depth can be determined. For the ProTreat® mass transfer rate-based simulator, structured packing performance can be predicted just as easily and reliably as for trays.

USING STRUCTURED PACKING IN GAS TREATING

In the last few years, engineers have shown increasing interest in using structured packing in amine-based gas treating applications. Consequently, structured packing is being applied more and more widely. Nevertheless, the idea is still new enough that questions are often asked as to whether structured packing should or should not be considered in a given application.

One of the most obvious application areas is where the columns are subject to periodic tilting motion as on floating structures such as FPSO and FLNG platforms. Structured packing resists liquid maldistribution brought about by rocking motion better than random packing. Trays have very poor resistance to the sloshing and seiching induced by lateral back and forth motion. In offshore applications structured packing should always be considered. Especially in the context of periodic tilting motion, however, care must be taken to use the right kind of liquid distributor. The distributor should have no free liquid surfaces and should be high pressure drop type, not a gravity flow device such as a trough distributor. This also mandates that solid amine hygiene be practiced (i.e., filtration and corrosion management).

There are very few reasons to exclude structured packing from consideration. One such reason is fouling. If the system is a fouling one, the deposits that will inevitably occur on the surfaces of the packing will be almost impossible to remove and operations may soon become plagued by plugging problems. On the other hand, unless structured packing is used too close to the flood point where liquid

holdup becomes high, it is naturally resistant to the generation and maintenance of foam. It should be recognised, however, that if the system is a *bad* foamer, structured packing may not be the answer—rather, the root cause of the foaming should be determined and alleviated. Even a good design is no match for poor amine hygiene practices.

In a revamp for higher capacity, the naturally higher vapor handling ability of structured packing may recommend it as a way to achieve higher capacity in the same shell. Except in tail gas treating and acid gas enrichment, pressure drop is not usually an issue. However, if it is, structured packing can almost always be made to work at lower pressure drop. To that end, the largest possible crimp consistent with being able to achieve the target separation within the height of the existing tower shell should be used. Again, finding out what that crimp size is can be facilitated greatly by using mass transfer rate based simulation.

SUMMARY

Packed absorption columns can operate in various modes of pinching or be mass transfer rate controlled throughout. There is just no way to tell beforehand which mode will prevail, and with what packing type and size. This makes it almost impossible to develop a design that is as truly optimized as it should be in FPSO and FLNG applications using any approach other than one with a mass transfer *rate* basis. Currently, only the ProTreat® simulator has proven this capability, as witnessed through repeated validation with real plant data.

A real mass transfer rate model is constructed from components that are soundly based in fundamental sciences and engineering, and not on approximations made to avoid what used to be impossibly complex and arduous hand computations. The enormous power of even desktop and laptop computers has turned computer time into a non-issue. Rigor can be achieved for the meager cost of computing times measured in tens of seconds. The computer models that result can be accurately described as virtual plants in which it is quite easy to investigate quite involved what-if scenarios.

Because of their wide range of sizes and the somewhat laterally compartmentalized flows, structured packings offer a great deal of flexibility in gas treating which allows them to be used where trays are very difficult, if not impossible, to apply. Possibly two of the greatest barriers to using structured packing in gas treating have been (1) the very small experience base and (2) the resulting difficulty in translating the results of more-conventional, approximate calculations into real internals. Today, the experience base is sufficient to have allowed ProTreat® to be benchmarked in a wide range of operations using structured packing, and computing power is such that approximate methods are no longer needed—all calculations can be done quickly, reliably, and in the fullest possible detail on a laptop computer. Even the effect of the flow regime on trays can be assessed with certainty.

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