



## REMOVING THE GUESSWORK

**S**tructured packing has been successfully leveraged in many applications to drive higher unit capacity, reduce pressure drop, or handle low liquid loads. These factors all come into play with the glycol dehydration process. However, similar to amine units and sulfur plants, glycol dehydration units have been historically viewed as 'utilities'. In the past, overdesign of columns using 6 – 8 bubble cap trays was the mainstay. However, with revamps becoming increasingly more prevalent and increasing offshore applications, the penalties for over-design are amplified. In particular, the rocking motion in FPSO applications demands a better option than conventional trays.

Until very recently, estimating the height of packing required to achieve a given gas dryness level has been done using rules of thumb for heights equivalent to a theoretical plate, without much certainty as to the impact of packing type, brand and size. These methods work fine in the world of overdesign, but new applications require more precise

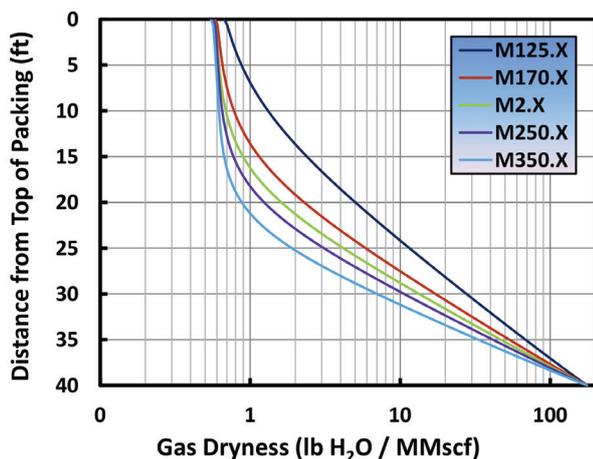
**Ralph H. Weiland and Nathan A. Hatcher, Optimized Gas Treating, Inc., USA, explain how precision mass transfer rate-based simulation removes reliance upon 'rules of thumb' in glycol dehydration.**

engineering. Mass transfer rate-based modelling uses proven engineering science to provide the precision needed to reliably and consistently meet engineering and plant objectives without recourse to the height equivalent to a theoretical plate (HETP) or height of transfer unit (HTU) estimates and other rules of thumb.

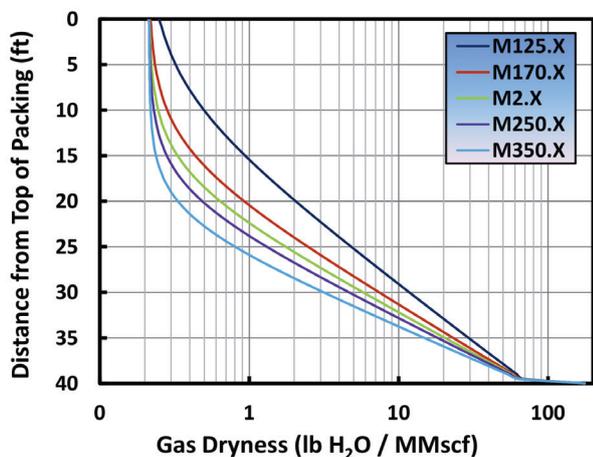
There are numerous commercial software packages capable, at some level, of simulating glycol dehydration along with a host of other

separation processes including conventional, azeotropic, extractive, and three-phase distillation, gas absorption and liquid extraction. Simulation of any process for separating chemical components between immiscible phases rests on two central foundations:

- ▶ Knowing the distribution of components between the phases at equilibrium.
- ▶ Modelling how the separation takes place in the piece of equipment itself.



**Figure 1.** Water removal depends on packing size and packed bed depth. Dehydration of water saturated methane at 500 psig.



**Figure 2.** Dehydration of water saturated methane at 1500 psig.

Phase equilibrium is a question of equilibrium thermodynamics and models for the individual phases sufficient for adequately representing solution and phase non-idealities. Models for the phases can range from ideal gas and ideal solution approximations to equations of state and activity coefficient models complete with interactions between different molecules. Phase equilibrium is in some ways least taxing; modelling the separation process itself is the real challenge.

The actual separation brought about by the phase contacting action of the equipment can be modelled in one of two different ways:

- ▶ As a set of countercurrent equilibrium stages that, conceptually at least, can be stepped off on an equilibrium diagram.
- ▶ Using a more scientific approach grounded in the concept of the mass transfer rate process as opposed to an equilibrium stage process.

Phase equilibrium in glycols has been addressed in some detail elsewhere,<sup>1</sup> particularly in the context of dehydrating streams with very high acid-gas content. The

model presented there was quite general and accounted for all the vapour and liquid phase interactions and nonidealities and it compared favourably with data from the Engineering Data Book, GPSA (2004).<sup>2</sup>

The authors note in passing that glycol dehydration is not usually considered part of an LNG train liquefaction process; rather, it is used to treat gas following acid gas removal. This permits the gas to be moved by pipeline in the relatively dry state so that condensation and subsequent corrosion do not occur. However, if extra water can be removed using glycol, less water will have to be removed in the liquefaction train.

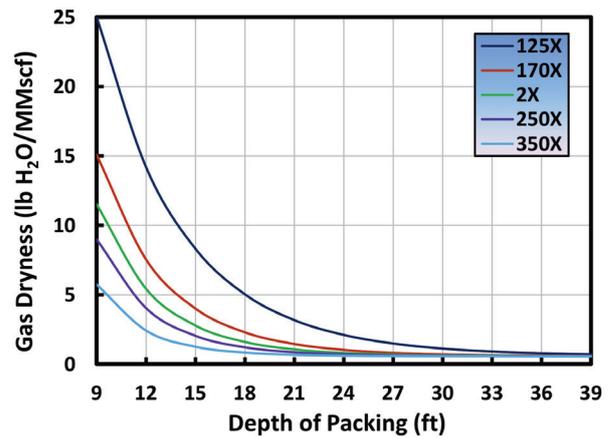
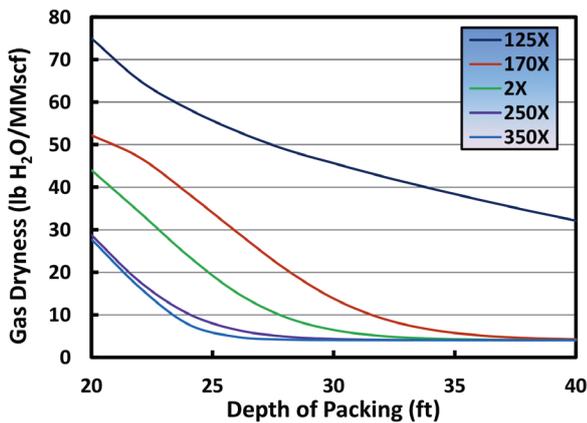
## Mass transfer rate model

Mass transfer rate modelling is a relatively recent (mid-1980s) departure from the classic ideal- or equilibrium-stage model of McCabe-Thiele vintage. Over the past 30 years, this type of model has gained increasing acceptance as 'the right way to do things', especially amongst the distillation fraternity. It takes advantage of modern high-speed computing power to calculate directly the transfer rates of components from one phase to another. Its closest analog is heat transfer, which uses film coefficients on the two sides of the rigid heat transfer surface (tubes or plates). There are extensive correlations of film coefficients for heat transfer that enable heat exchangers to be designed, not on the basis of hypothetical, 'thermal-equilibrium stages', but by actually calculating heat transfer rates at points along the length of the exchanger. The mass transfer analog is a very close one, more complex because many components (not just heat) cross phase boundaries that are themselves flexible and moving, not rigid. Modern digital computing makes the computations a non-issue. And today there is a good database of mass transfer coefficient information for various types of tower internals as functions of process flows and phase properties – certainly sufficient to permit accurate prediction of mass transfer rates and, therefore, an accurate prediction of the separation that a given tower will produce under actual process conditions. For a more detailed discussion of the mass transfer rate model, please refer to Weiland et al.<sup>3</sup>

## Using structured packing

In the last few years, engineers have shown increasing interest in using structured packing in various gas treating applications. Consequently, structured packing is being applied 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 seicheing 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



**Figure 3.** Depth of mellapak structured packing required to achieve various dryness levels. a) Dehydration at 15 psig (left). b) Dehydration at 500 psig (right).

as a trough distributor. This also mandates that solid unit hygiene be practiced (i.e., filtration and corrosion management).

There are few reasons to exclude structured packing from consideration. One reason, however, 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.

In a revamp for higher capacity, the naturally higher vapour handling ability of structured packing may be recommended as a way to achieve higher capacity in the same shell. Except in sulfur plant 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. 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.
- ▶ Low water-content glycols are extremely viscous, which recommends against using trays.

The advantage of increased gas handling capacity can be a 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 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 same cannot be said for packed bed depth – this is determined by mass transfer, and the mass transfer performance of structured packings (especially with chemical reactions) is still viewed by some as a rather specialised area, if not an art.

Packed columns always seem to have presented a challenge to designers, perhaps because there are so many varieties, types, and sizes of packing, and because the experience base is so small, especially in gas treating applications such as amine treating and glycol dehydration. Traditionally, these units have been viewed as utilities, and in years past, building in over-design was not critical in grassroots situations. Offshore units and revamps are another matter. Here the penalty for over-design can be much more costly or just simply not practical.

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 accurate and reliable predictions of column performance without recourse to estimating artificial parameters, such as tray efficiencies or meaningless residence times of theoretical stages.

## Process simulation

Attention is restricted to the use of structured packing in glycol dehydration. Until now only an equilibrium stage

model has been available for calculations involving the performance of structured packing. However, the HETP of the particular packing must surely be related at least to packing size and type. For structured packing, size can be expressed in terms of specific surface area and crimp size, characteristics that are geometrically related. Under otherwise identical process conditions, one should expect that use of large crimp packing will require a deeper bed to give the same performance as a relatively short bed of small crimp packing, if for no other reason than because the surface area (for mass transfer) of the small crimp material is so much larger.

The rest of this article uses a case study to elucidate how packing size and bed depth affect the performance of a particular structured packing in a glycol dehydration application. Packings from the Sulzer Mellapak X-series were selected without prejudice for this study. Simulations were run for treating gases at low pressure (15 psig) and higher pressures (500 and 1500 psig) saturated with water at 120 °F. The absorber was set up to dehydrate 49 000 lbmol/hr of wet sweet methane (trace CO<sub>2</sub> and H<sub>2</sub>S) using slightly more than 3 US gallons of 99.95 wt% TEG per pound of water removed. This TEG dryness of 99.95 wt% represents the approximate upper end of practical glycol purity in an application using stripping gas for solvent regeneration. Simulation results are discussed from two perspectives:

- How gas-phase profiles of water content vary with position across the height of packing.
- The actual depth bed of various packings needed to achieve a given gas dryness.

Figure 1 shows the water content of the intermediate-pressure gas (500 psig) at various positions along the height of a column containing 40 ft of packing. The glycol flow was 12 US gal./min and the tower diameter for 50% flood ranged from 2.2 to 2.8 ft depending on the packing size. It is evident that after traversing the bottom 20 ft of M350.X packing, the gas is nearly as dry as 99.95 wt% TEG at 100 °F can get it, certainly below 1 lb/million ft<sup>3</sup>. With M125.X packing on the other hand, it takes 35 ft of packing to achieve the same dryness and water is still being removed even after the gas has passed through the full 40 ft bed. Regardless of what level of dryness is to be achieved, twice the height of M125.X packing is required compared with M350.X packing for the same water removal. (In this example, the water content of the inlet gas was 176 lb/million ft<sup>3</sup> in all cases.) So the required bed height is very much a function of the packing size. However, it is not in simple linear proportionality to the specific surface area of the packing. It is not that rules of thumb cannot be made to work; rather, it is that the right

rules of thumb, at least for packing, depend on too many parameters (not just packing type and size, but also on the gas and solvent fluxes through the column, and fluid properties) and this makes them rough at best and unreliable at worst. Until now the answer to this dilemma has been to over-build the columns; however, in a competitive environment, being able to avoid over-design gives the knowledgeable contractor and the astute internals vendor a commercial advantage. Mass and heat

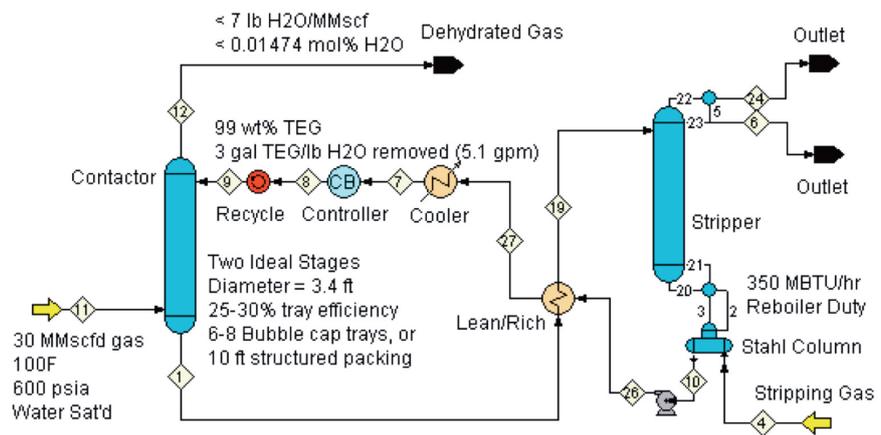


Figure 4. GPSA Data Book Example 20-11.

transfer rate-based simulation is the precision tool that allows this to be done.

Figure 2 shows similar profiles of the gas water content for dehydration of water-saturated gas at 1500 psig. The required glycol flow is lower in this case (5 US gal./min vs. 12 US gal./min at 500 psig) because there is less water to remove from such a high pressure gas – the saturated water contents are 176 and 75 lb/million ft<sup>3</sup> at 500 and 1500 psig, respectively. At 1500 psig only 14 ft of M350.X packing are needed to achieve 1 lb of H<sub>2</sub>O/million ft<sup>3</sup> of gas; whereas, 19 ft are needed for the same gas at 500 psig. However, once again nearly twice the bed depth of M125.X packing is required to achieve the same result.

Figure 3b (right) shows results corresponding to the same conditions as for Figure 4, while Figure 3a (left) corresponds to dehydration of the same gas at only 15 psig. This form of presentation is more directly related to the design question: ‘how much packing of what type is needed to achieve a stated dryness?’

At 15 psig, the specific area of Mellapak M125.X is so low that no reasonable amount of this packing will come even close to reaching anything like the 4 lb/million ft<sup>3</sup> dryness possible with 99.95 wt% TEG. On the other hand, M2.X needs about a 35 ft bed while M350.X needs only 25 ft. Choosing the wrong packing size can result in a requirement for significantly more packing, or maybe even setting up an impossible design if the packing is just too coarse. All of this begs the question of even estimating the required bed depth at all from rules of thumb applied to ideal stages. For example, it should be

quite apparent that using a rule of thumb for the HETP developed for dehydration at low pressure will give completely the wrong answer at high pressure. A real mass transfer rate-based simulation provides an accurate bed depth with great ease, something that is quite impossible using ideal stages and rules of thumb to estimate HETP values.

## Conclusion

Reliance upon rules of thumb in glycol dehydration is no longer a necessary handicap with the availability of precision mass transfer rate-based simulation, such as that available in the ProTreat® simulator. As this article has suggested, size does matter in the context of structured packing. Quoting or recommending a single HETP or HTU in these cases would have been nonsense. The right value depends on the particular packing brand, size, as well as on the operating conditions and the gas dryness sought. When dealing with rules of thumb, the rules may simply not apply, or they may be misapplied because their basis is not well enough understood. A well-known example from the tray literature on dehydration uses a blanket tray efficiency of 25% (suggested in the GPSA Engineering Data Book, 2004), which is close to the truth only some of the time. But in deep water removal, 25% is optimistic and unless one adds several additional 'safety' trays, failure will threaten. The situation with random or structured packings is much worse. For example, the same GPSA Engineering Data Book suggests an HETP value without

even referring to the packing type, let alone its size. Rules of thumb once had their place when the best one could do was an equilibrium stage calculation, and reliance had to be placed on experience as expressed (and mis-expressed) in such rules of thumb.

With a true rate-based heat and mass transfer approach to simulation, one never has to worry about guessing tray efficiencies, phony residence times, HETPs, HTUs, and other approximations, estimates, and arm waving. The mass transfer rate-based approach to modelling does not use rules of thumb – it uses science and good sound engineering to predict performance. As with all well-based mass transfer rate calculations, these results were obtained without any correction factors whatsoever. They are true out-of-the-box predictions in every sense of the word and do not leave the process engineer wondering whether their design is even adequate, let alone optimal. **LNG**

## References

1. Carroll, J.J., Hatcher, N.A., and Weiland, R.H., 'Glycol Dehydration of High-Acid Gas Streams', *PTQ Gas*, 43, 2011.
2. *Engineering Data Book*, Gas Processors Suppliers Association, 12<sup>th</sup> Edition, Vol. II, 20, 2004.
3. Weiland, R.H., Sivasubramanian, M.S., and Dingman, J.C., 'Effective Amine Technology: Controlling Selectivity, Increasing Slip, and Reducing Sulfur', paper presented at the Laurence Reid Gas Conditioning Conference, Norman, Oklahoma, USA, February 2003.