SENSITIVITY OF TREATING PLANT PERFORMANCE TO TOWER INTERNALS

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ABSTRACT

One of the most neglected areas in gas treating is the proper and methodical selection and comparison of tower internals. The performance of tower internals has been studied predominantly from the perspective of hydraulic capacity. Apart from efficiency measurements, notably by FRI and almost exclusively in the context of distillation, mass transfer performance has tended to receive relatively scant attention.

In distillation, the dominant resistance to mass transfer is almost always in the gas or vapor phase. In gas absorption, on the other hand, the absorption rate of *some* components under many different conditions of operation lies in the liquid phase. Furthermore, in acid gas removal, chemical reactions play a decisive role in determining the separation. These factors alone set absorption apart from distillation. A further dissimilarity is the usually large density difference

between phases in gas absorption; in high pressure distillation the vapor phase density can be a large fraction of the liquid density but in acid gas removal they are widely separated. Hydrocarbons also have different wetting characteristics on metal packing surfaces than aqueous systems. All too often, distillation data do not apply well to absorption.

When parts-per-million specifications on product gases must be met, 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. A large part of the reason was inattention to proper distributor design. Another was the persistent and still unresolved difficulty in translating ideal stages into actual packed bed depths and the selection of a particular commercial packing, especially in amine treating. One of the benefits of true mass transfer rate-based simulation (using ProTreat® for example) is that the concepts of ideal stages and HETPs are completely avoided, so translation from ideality to reality is a non-issue—no translation is needed. The separation is simulated directly and a large area of uncertainty is removed.

In this contribution, a case study is used to reveal the important tower-internals parameters that determine an absorber's mass transfer performance. This is done in the context of an LNG project, and it compares a structured packing with three random packings of different size and type, and with two types of commercially proven valve trays.

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Back to Basics: Understanding the Difference

Tower internals have natural hydraulic capacity limits imposed by such parameters as jet flooding, downcomer backup, downcomer choke flood, packed column flooding, etc. These parameters can generally be assessed through software freely available from each internals provider or through the Design and Rating Program available from Fractionation Research Inc. (FRI). All the various tools use different input and output data panels, reflecting the various differences that pertain to the different types of internals that can be selected for a particular service.

All tower internals have geometries that manipulate the hydraulics to promote contact across the interface between counter-currently flowing vapor and liquid phases. However, from the point of view of mass transfer, the rates of transfer of components between phases depend on the interfacial area available, the turbulence level created in each phase, the physical and transport properties of the phases, their flow rates, and concentration differences between phases and across phase boundaries. Of great underlying importance is the fact that mechanically different internals create different levels of turbulence for every type and size of internal even at the same flow rates using the same fluids. Turbulence levels determine phase mass transfer coefficients just like turbulence levels on the shell and tube sides of heat exchangers determine heat transfer coefficients. Equally important is the interfacial area between the gas and liquid through which the transferring species must be transported to effect a separation. An example is the behavior of packing.

As the solvent flow rate to a packed bed is increased, the wetted interfacial area rises as well, and the mass transfer rate therefore increases with solvent flow. This is shown by the right side of Figure 1 where increasing liquid rates cover more and more of the packing surface with a film of liquid. Treating improves until the tower operation runs up against another limitation, in this case a lean end pinch in which the treated gas reaches an equilibrium state with the lean solvent. Beyond that point, more solvent does not result in any change in absorber mass transfer performance. On the left side of this plot, yet another limitation is encountered, namely a solvent capacity limit where insufficient solvent is being provided to remove enough CO_2, and

breakthrough occurs. However, in the central region of the plot, mass transfer area is responding to changes in solvent flow. With trays, gas-liquid interfacial area for mass transfer is only a relatively weak function of liquid rate but a strong function of gas rate.

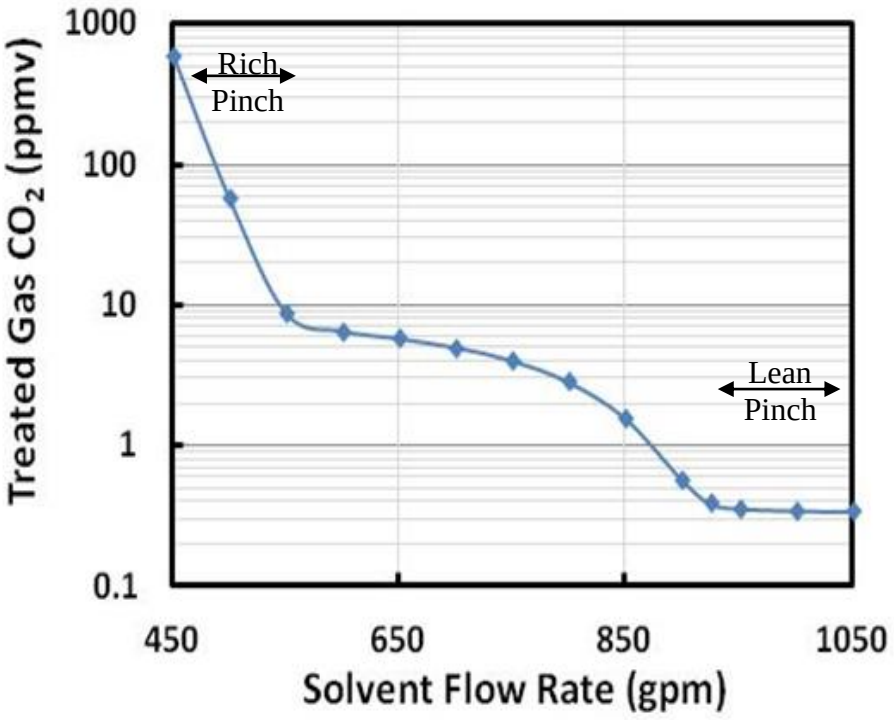

Figure 1 Performance of a Packed LNG Absorber with 0.015 Lean Loading

The case study below involves the processing of 745 kNm^3/h of a gas containing 11.5% of CO_2 supplied to an onshore contactor at a pressure of 55 barg and a temperature of 31°C. The treated gas is to meet a specification of < 50 ppmv carbon dioxide. INEOS* GAS/SPEC* CS-1120® solvent at 50% weight concentration is used in this assessment. Tower diameter was *calculated* to achieve 60% of flood[a] or, in the case of Sulzer internals, 60% of maximum operating capacity. Simulations of the full treating plant were run using the ProTreat® mass transfer rate based gas treating simulator (Version 5.5.2). In each case, the absorber contains enough internals (packing depth or tray count for the calculated tower area to operate at 60% flood) to achieve ≈50 ppmv CO_2 in the treated gas when using the same lean solvent CO_2 loading (0.0175 moles of CO_2 per mole of total amine in the CS-1120 solvent) throughout. The simulator uses established generalized correlations for mass transfer coefficients and specific effective interfacial area for structured and random packings and for trays. It has been thoroughly benchmarked against a wide range of commercial data on amine contactors using both random and structured packings as well as trays. Table 1 shows the key results of this simulation study.

[a] Raschig uses the approach to flood concept while Sulzer uses capacity such that 100% capacity is defined to occur when the pressure drop reaches 12 mbar/m of packing depth. Thus, 100% capacity is achieved before flooding occurs.

Table 1 Treating with Various Packings and Trays

		Packings				Trays	
		Structured	Random			Valve	
		2nd Gen	4th Gen	3rd Gen	3rd Gen	2nd Gen	3rd Gen
		Mellapak 2X	Raschig SR #2	Nutter #2	Nutter #2.5	Sulzer BDH	Sulzer UFM
Dry Specific Area or Weir Height	m^2/m^3 (1/m)	210.0	97.6	96.0	85.0	0.076	0.076
Lean Amine Circulation	m^3/h	1,817	1,817	1,817	1,817	1,817	1,817
	gpm	8,000	8,000	8,000	8,000	8,000	8,000
% Flood or % Capacity (max value)*		60	60	60	60	60	60
Hydraulic Load	$m^3/m^2{\cdot}h$	88.60	80.02	69.13	74.97	81.31	97.26
	gpm/ft^2	36.24	32.73	28.28	30.67	33.26	39.78
Treated Gas – CO_2 Content	ppmv	50.97	49.65	50.85	50.3	34.1	38.6
Rich Amine Loading	mol CO_2 / mol Amine	0.475	0.475	0.475	0.475	0.475	0.475
WET Interfacial Area - Total	m^2	20013	17103	17955	17964	26192	24347
- Relative to M2.X		1.000	0.855	0.897	0.898	1.309	1.217
Interfacial Area Based on Tower Volume	m^2/m^3 (1/m)	232	144	138	126	85	97
Wetted/Dry Area Ratio		1.10	1.48	1.44	1.48	n.a.	n.a.

* All columns sized for 60% flood

The treated gas CO_2 content is sensitive to the packing depth (and the number of trays) as well as to the total wetted (interfacial) area. In addition, wetted area is not the only factor determining performance. Gas and liquid side mass transfer resistance (as expressed by mass transfer coefficients play an important role, too. At given flow rates, turbulence levels within the gas and liquid phases depend on packing type, shape, and size. The comparisons shown in Table 1 are indicative; they are *not* general recommendations, *except for this specific example*. In the case of trays, 17 actual trays could not quite reach 50 ppmv CO_2, while 18 trays exceeded requirements.

A rich loading of 0.475 is just beyond the practical limit (0.40 – 0.45) for carbon steel metallurgy. However, in LNG processes, stainless cladding of certain parts of absorbers and regenerators, and some stainless piping are fairly commonly used to prevent corrosion. Interestingly, because mass transfer rate-based simulation provides a detailed map of acid-gas loading and temperature profiles in these towers, it provides a wonderful basis on which to decide where to clad and where to build in mild steel. ProTreat's corrosion coupon capability provides additional guidance.

Table 1 shows that despite the much higher *specific* dry area of Mellapak™ 2X packing, all the packings require roughly similar total wetted area to achieve the 50 ppmv CO_2 treating specification. Total wetted or interfacial area is a key parameter. Nevertheless, the random packings all require slightly less total wetted interfacial packing area than Mellapak 2X, with Raschig SuperRings® No. 2 being closest in performance to the structured packing. The reason for the slight difference between the two types of packing is subtle differences in gas- and liquid-side mass transfer coefficients because both gas and liquid resistances are important. Liquid-side coefficients are more important for CO_2 *below* the temperature bulge because increasing acid-gas loading slows the CO_2 reaction rate. But gas-side coefficients are more important *above* the temperature bulge because the very fast kinetics of the lightly-loaded GAS/SPEC CS-1120 solvent moves the dominant mass transfer resistance from the liquid into the gas phase. Another

cause of differences is temperature dependent transport properties that change in response to differing temperature profiles. There are many influences.

As shown in Table 1, Mellapak 2X structured packing has more than twice the dry specific area of any of the random packings, and 60 – 80 % the tower volume specific interfacial area. Therefore, given that all the internals need roughly equal effective interfacial area, one might expect that, in this case, a comparatively smaller total packing volume is needed for Mellapak 2X than for random packings. But this is not the case because the gas and liquid side mass transfer coefficients are different, as already described. In this case study it results in a relevant smaller packing height for Mellapak 2X compared to random packings. One might well ask the question then, if the physical volume of structured packing in an absorber can be made lower than random packings, why isn't structured packing used all the time? The answer doesn't necessarily lie in mass transfer performance, but rather in the hydraulics. Amine treating can be a fairly dirty service (especially in refineries) so structured packing, particularly *small crimp* packing, should be used only if the operator is prepared to give undivided attention to ensuring excellent amine hygiene. One might recommend, for example, full-flow particulates filtration accompanied by charcoal filtration to remove surface active agents. The deposition of solids and gelatinous substances means early death to structured packing, and the smaller the crimp, the earlier its demise. Figure 2 shows photographs provided by the INEOS* GAS/SPEC* Analytical Laboratory of amine samples from an operating facility. The left-most bottle in both photos is virgin solvent. The photographs illustrate the solids that can be present in plant solutions.

Figure 2 Amine Samples immediately after sampling (L) and after settling (R)

Some practitioners claim that structured packing is also more prone to initiate and stabilize foam—perhaps this occurs at the joints between packing layers where the gas flow undergoes sudden changes in direction. These joints are also where flooding often initiates, perhaps lending some credence to the foaming conjecture. It also possibly caused by localized disturbances to L/G ratios within the beds. Furthermore, structured packing may be somewhat more sensitive to liquid maldistribution.

Figure 3 shows a foaming test of an operating plant solution done by the INEOS* GAS/SPEC* Laboratory, indicating a probable foaming issue in this plant and, if not addressed, with a potential operating issue with the contactor internals. In any event, one may want to be a little cautious in giving an across-the-board structured packing recommendation without due regard

for the particulars of the application. Close attention must be paid to installation of distributors, redistributors, wall wipers and other design/installation issues recommended by structured packing suppliers. All in all, it is necessary to have excellent design and installation of liquid and vapor distributors to get the best performance from structured and random packings alike.

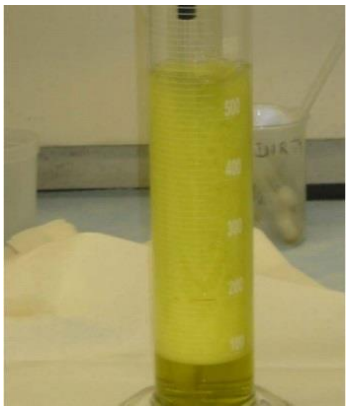

Figure 3 Foam Test of an Amine Sample following Settling

The type of analysis described here can be carried out for any type and brand of packing contained within the ProTreat database and, of course, findings will differ from packing to packing. It should be noted that the analysis for a contactor located offshore or on a floating facility will include additional considerations and, therefore, any conclusion from this work cannot directly be extrapolated to those situations. For example, on FPSOs rocking motion is a serious consideration—even here there are applications where structured packings are preferred but there are also applications where random packings are favored. Flow maldistribution must be considered in all such applications. However, even greater attention to distributor design is mandatory. ProTreat can be used to assist in troubleshooting performance issues in contactors. INEOS uses an Infrared Camera to check for distributor and maldistribution issues in a packed column after plant simulation indicates poorer than expected treating performance. Figure 4 shows a thermal image (cf. Shiveler et al.[1] and Teletzke and Bickham[2]) of the liquid flow and reaction (indicated by the red region) occurring down the wall of the contactor but with very little reaction in the packed bed. This may indicate poor distribution of amine solution or, in a large diameter column, poor gas distribution.

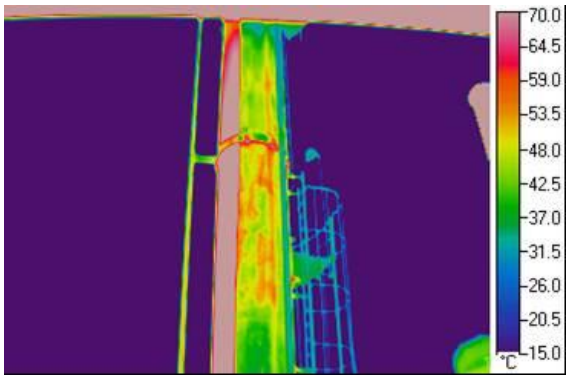

Figure 4 Infrared Camera Photo of Distribution Issue in Packed Tower

One purpose of this paper is to point out the controlling, underlying physical and chemical factors to which one should pay attention in selecting packed tower internals from a mass transfer performance standpoint. In that spirit, no endorsement of one packing over another is being made. There are always multiple choices, and indeed, multiple good choices. The best choice will depend on many factors, and in some cases mass transfer performance may not even be decisive. In any event, there is no one better qualified to help in making the best packing decisions than a reputable internals supplier.

Turning now to trays, Table 1 also shows a comparison with Sulzer BDH™ and Sulzer UFM™ (Umbrella Floating Mini-Valve) trays where it is immediately apparent that the total interfacial areas needed to achieve a specified separation using trays are similar to, albeit somewhat higher than, those needed for packings. Note also that the interfacial area in the table is based on the total tower volume occupied by the internals, not the volume over the active deck or tray bubbling area (i.e., exclusive of downcomers). In the case of the UFM tray deck, the total interfacial area is closest to structured packing and is sufficient to treat satisfactorily under these conditions because the gas-side mass transfer coefficient in the froth over the trays is a little higher than for random packing. If 17 trays had produced 50 ppmv CO2 rather than 56 ppmv actually predicted, the area would be even closer

Considering that the cost of a high-pressure vessel has more to do with its diameter than its length, it is worth noting that these trayed towers have diameters that are about 4% larger than the structured packing option for the classical BDH rectangular floating valve and about 5% smaller than the structured packing option for the newest UFM valve. This is not a large difference. For this specific application, the use of a four-pass layout is demanded by the high liquid weir loads (heavily loaded downcomers); however, this could limit turndown when compared with the packing options because multi-pass trays tend to have more limited turndown capabilities. The turndown limitation on packed beds, on the other hand, generally depends more on the distributors than on the packings themselves.

The Need and Importance of Performing a Sensitivity Analysis

A normal expectation in the course of operating an amine plant is that small changes in operating conditions should result in correspondingly small changes in plant performance. However this expectation is not always well founded[3]. Therefore, an analysis of performance sensitivity to all significant operating parameters should always be performed. This is especially recommended for amine systems operating near what is known as temperature pinch conditions where there can be stable operating regions demarcated by steep boundaries outside of which performance degrades precipitously. In other words, operationally *mass transfer* instabilities manifest as a sudden loss in treating, but there may be little or no accompanying effect on hydraulic performance *per se*. This can make treating failures hard to diagnose without a true mass transfer rate based simulation tool.

The main cause of sudden loss in treating is attempting to operate with the bare minimum of solvent absorptive capacity, whether from too low a solvent flow rate, too high a gas flow rate, too high temperature, or too high a CO_2 (or H_2S) concentration in the raw gas. With fast reactions, the reaction zone in the absorber extends only over a narrow band around the maximum temperature found in the absorber (bulge) — temperature is a strong indicator of the rate of reaction. Under the specified operation condition, this reaction zone with its telltale bulge

lies close to the bottom of the absorber. These profiles are illustrated in Figure 5. But, unlike slowly reacting solvents like DEA where the reaction zone is broader and more gradually emerges from the top of the tower (with decreasing solvent flow rate for example), a narrow reaction zone gives no warning of its impending emergence until it suddenly pops out the top of the column, and treating goes wildly off specification. Temperature bulges are also where towers find themselves hydraulically limited.

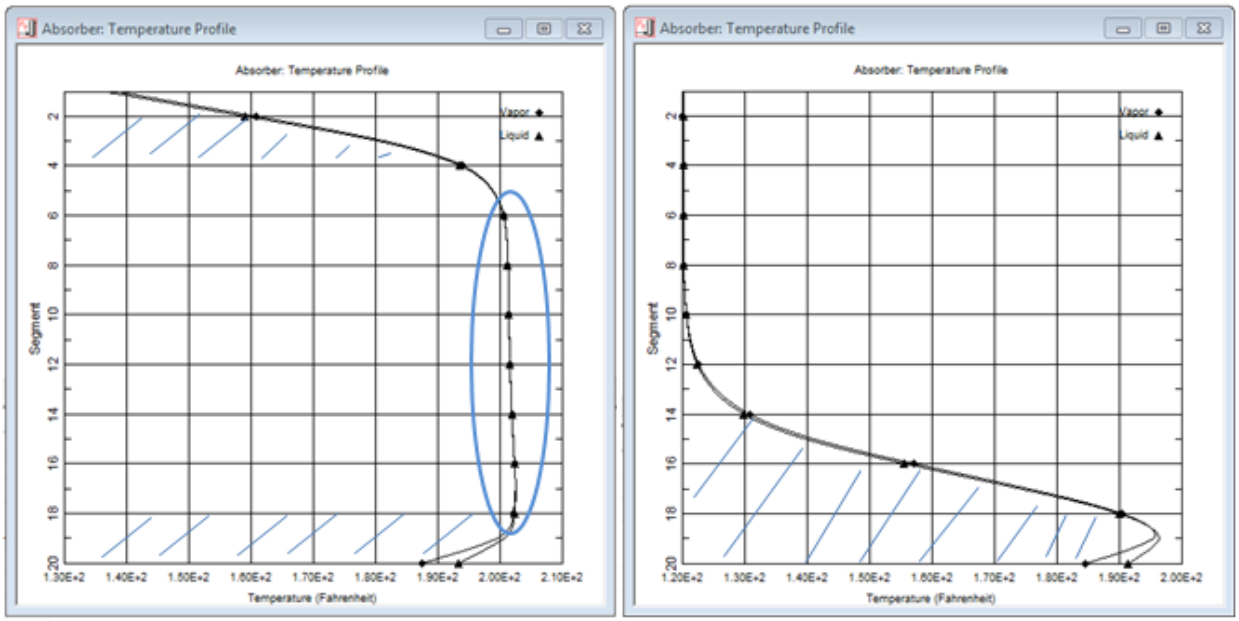

Figure 5 Bulge Pinched and Lean-end Pinched Temperature Profiles

For example, a treated gas exhibiting a carbon dioxide concentration of 10 ppmv can suddenly go to thousands of ppmv with less than a 0.1% drop in solvent flow rate[4]. Figure 6 shows an example of such a stability boundary. The design point in this particular plant is 115°F but a temperature excursion of 10°F will throw treating into a tailspin. Note too, that although 50 ppmv CO_2 may be the targeted treating level, stable operation at that value is impossible, at least by manipulating lean amine temperature. In fact, in this case treating degenerates very rapidly from 10's of ppmv CO_2 to several 1000's of ppmv with a temperature increase less of than 1°F.

The detailed reasons for the shape of the curve in Figure 6 (and Figure 1) are complex and are discussed in full detail by Hatcher et al.[5] However, the primary cause for the sudden loss in treating performance is CO_2 breakthrough when the lean amine temperature no longer permits all the CO_2 to be absorbed because of higher carbon dioxide backpressures over the solvent.

A profile of carbon dioxide concentration in the gas versus distance down the bed would show that under normal operating conditions (120°F, for example) the part of the absorber just above the bulge and very close to the bottom is where most of the absorption takes place. The upper part of the bed (more than half) merely polishes the gas. As the lean solvent is made hotter, the temperature bulge creeps up the column, the polishing section gets shorter, and eventually, with a seemingly very small further increase in lean temperature, the polishing section disappears entirely, and a high CO_2 concentration pops out the top of the bed. This event tends to occur rather unannounced because polishing deals with only a few parts per million so the treated gas

doesn't seem unusually hot, and the tower continues to meet specifications right up to the point where treating fails. When it does occur though, as Figure 6 shows, a 0.1°F temperature rise is all it takes for treating to go from a few 10s of ppmv to a 1,000 ppmv or more, completely unsatisfactory for gas liquefaction.

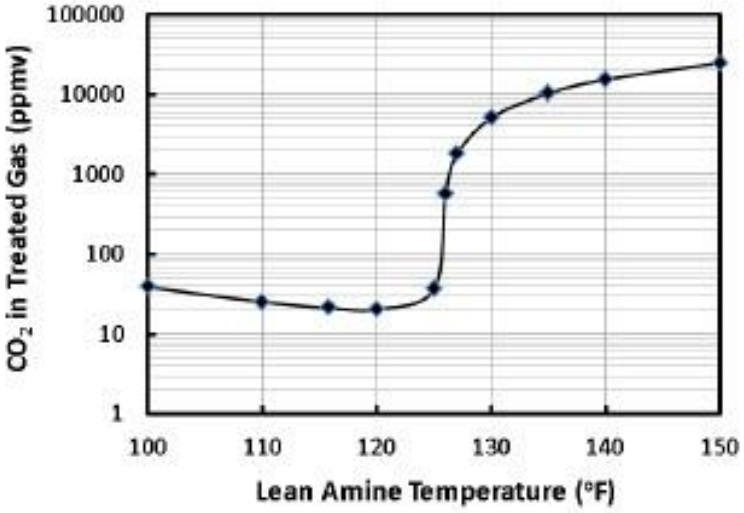

Figure 6 Extreme Sensitivity to Lean Amine Temperature in an LNG Absorber

The ability of the solvent to continue "loading up" with carbon dioxide is compromised as it becomes too hot, and what cannot be absorbed must leave in the exiting gas. Figure 7 shows contactor scans performed by INEO for operating plants with a bulge pinch (left) versus a bulge illustrating the more normal (and desired) reaction zone right in bottom of contactor (right). This

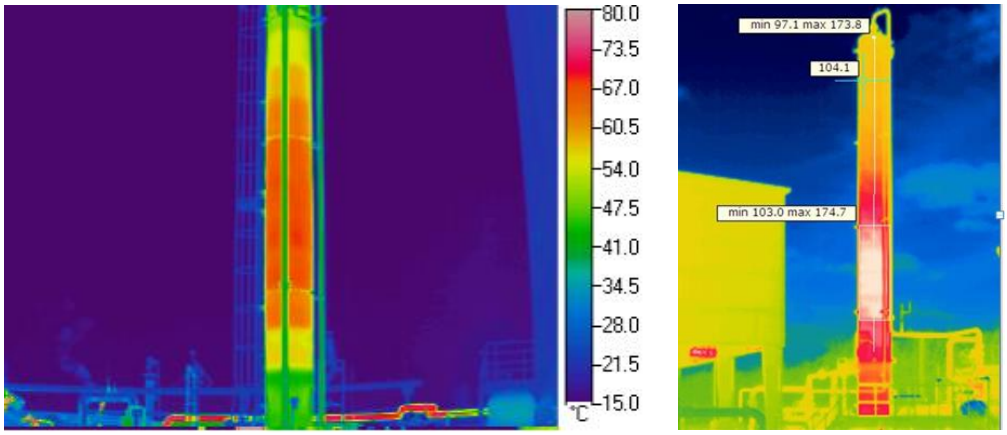

Figure 7 Infrared Scan of Contactor Profiles with Broad Bulge and Bottom Bulge

is but one example pointing to the need for a careful sensitivity study in a new gas treating unit, especially one in which a fast reacting solvent is used for deep CO_2 removal as is often the case in LNG facilities.

Finally it should be noted that operation of the contactor with a hot narrow temperature profile may result in equipment corrosion due to breakout or flashing of CO_2. Figure 8 shows photos illustrating corrosion on contactor walls and in downcomer areas caused by CO_2 flashing and vapor release in the tray liquid holdup area. In the hottest region of a contactor, turbulence can cause sudden short-term pressure fluctuations sufficient for CO_2 to flash from solution.

Figure 8 Contactor Wall Corrosion in Tray Liquid holdup Area and Downcomer

Concluding Remarks

There are many aspects that need to be considered when selecting between different internals for the same service, but ultimately the selection can be made *logically* only by using mass transfer rate-based simulation. Unlike ideal stages with their efficiencies, the mass transfer rate approach recognizes that the tower is not a black box—it really does contain internals—and it accounts for the differences between various internals and the characteristics of each. Put another way, if your simulation tool doesn't understand how one packing behaves differently from another as a mass transfer device, how can it possibly be of any assistance in making the best internals selection? Without a mass transfer rate-based simulator, the comparative table presented here could not have been generated. Indeed, quite likely a non-optimal internals selection would have been made, with higher performance uncertainty leading to a design with considerably more safety margin than necessary.

Naturally, mass transfer performance is not the only factor that must be considered. The final selection decision must also include the total installed cost, the operating costs, the availability of an experienced crew to install the selected internal at the specific location where the plant will be located, the type of service (dirty or clean), and the final technical-economical terms and conditions offered by every individual equipment supplier. Also, it must be noted that what was the best choice for one application will not necessarily be the best choice for another, regardless of apparent similarities.

The expedient selection of a particular column internal that fits within a current tower and extends hydraulic capacity *is often the very decision that defeats the mass transfer treating performance*. Rate-based simulation strongly emphasizes the fact that absorbers are truly mass

transfer equipment—not ideal stages of hydraulic contact—and simulating this way provides important support for engineers working on industrial tower designs and retrofits where mistakes may be very costly and career limiting..

This paper has focused on the sensitivity analysis that should be performed for the design of a new amine absorber, in this case in an LNG plant. However, a sensitivity analysis should also be included in retrofitting projects: no amine treating plant should ever consider a revamp from trays to packing before fully understanding the limitations already existing in the plant, including current design, and operating problems. For example, being unaware of the effect of heat stable salts (HSSs) on performance can lead to a failed revamp because the HSSs may be the real cause of poor performance, not the tower internals. In another example, when moving from trays to packing to relieve a limitation caused by vapor hydraulics, care must be taken to selecting the correct packing type to compensate for the tower volume now occupied by support and liquid distributor internals, without serious loss of mass transfer performance. This can be a terrible oversight when moving from trays to packing. Making the wrong choice of packing type, packing size, packing support, and liquid and vapor distribution for the conversion can result in unacceptable loss of mass transfer performance. Thus, the retrofit of any absorber or other mass transfer equipment must also focus on the mass transfer aspect, too, even though originally the exercise may have been driven to relieve hydraulic capacity limitations or to increase throughput.

Finally, by working in a collaborative environment, sharing common proven simulation tools, and the non-substitutable value of the experience of "lessons learned" by equipment and solvent providers, internals suppliers, process engineers, process designers, troubleshooters and plant operators, an initial design can be made much more solid with correspondingly less uncertainty, less design margin, better plant operability, and lower risk of unpleasant surprises.

There are many different opinions for designing tower internals for use in amine services, and because absorption and distillation are partly art and partly science, differing opinions can be shown to be equally serviceable in the field. Trays and packings can both be utilized for amine services. *The key to a successful design or retrofit is the proper selection and application of the tower internals to the specific project conditions been assessed*. The selection must be done from both a hydraulic *and a mass transfer* viewpoint. This paper is offered as an illustration of key facts that need to be considered in order to develop practical, serviceable column designs for amine treating systems, where operating experience must be included as one of the key factors. Consequently, this paper is a compilation of the views of authors from a variety of backgrounds, and it does not necessarily represent completely the views of the five corporations which the authors represent professionally.

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