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HYDROCARBON PROCESSING®

HP | Bonus Report

LNG

Liquefied natural gas is a growing source of clean energy for gas-consuming countries and a rising source of revenue for gas-producing nations. A number of new liquefaction and regasification plants have been proposed in North America, Africa, Asia-Pacific and Eastern Europe. Engineering, design and operation of these new terminals—as well as the LNG carriers that transport and deliver the product—can be improved with intelligent simulation and data collection and monitoring, as discussed in this month's bonus report.

Photo: First LNG bunkering at the Port of Civitavecchia in Rome, Italy, May 2014. Image courtesy of LNGEurope.

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How sensitive is your treating plant to operating conditions?

A normal expectation in the course of operating an amine treating plant for acid gas removal is that small changes in operating conditions will result in correspondingly small responses in plant performance. However, such expectations are not always well founded. To establish credibility for the process simulator used in the design of a new LNG plant, the mass-transfer rate-based simulation results for the new plant are compared with performance data from an operating LNG plant.

Attention is then turned to another LNG project under study. The project's carbon dioxide (CO_2) removal system consists of a single, large absorber serviced by two regenerators in parallel. Initially, the plant would process about 1,400 million cubic feet per day (MMcfd) of gas containing approximately 16 mol% of CO_2. A sensitivity analysis shows potential susceptibility of treating performance to departures of certain operating conditions from design values, and it also provides reasons for this sensitivity.

UNDERSTANDING OPERATING PARAMETERS

There are a number of well-accepted limits on plant operating parameter values. Examples are corrosion considerations for carbon steel, which usually limit rich amine CO_2 loadings to below 0.4 mole–0.45 mole of CO_2 per mole of amine. They also place upper limits on maximum line velocities to prevent the direct scouring of surfaces and the removal of protective films, particularly sulfide layers. Tower internals have natural restrictions on gas and liquid rates. Beyond these natural hydraulic capacity limits, jet flooding or downcomer backup and choke flooding of trays, or packed column flooding, may occur. Finally, solvent capacity is itself limited by temperature and acid gas partial pressures.

Corrosion, temperature and acid gas pressures all limit rich solvent loadings, while limitations on tower hydraulic capacity restrict throughput. These limits restrict performance, but they do not cause oversensitivity of performance to small changes in the values of operating parameters. However, the "small-response-to-a-small-stimulus" paradigm has changed with the introduction of fast-reacting solvents, such as piperazine for CO_2 removal.

The development and availability of highly precise simulation tools[a] have encouraged the design and construction of new plants with lower design margins. They have also allowed engineers to assess precisely the effect of operating parameters on performance, and have revealed the existence of operating cliffs or points of instability on the performance map.[1]

Although, regions of increased sensitivity have been predicted even for moderately fast-reacting CO_2-monoethanolamine (MEA) absorbers, when piperazine is used to promote methyl diethanolamine (MDEA), reaction rates grow very large, and unexpectedly high sensitivity of performance has been observed. None of this has been predictable using the more traditional ideal-stage simulation tools, even when they have been modified with efficiencies, and even with attempts to force reaction kinetic rates into what is fundamentally an equilibrium model. What enables the mass-transfer-rate model to reveal what has, for so long, been hidden behind the façade of the ideal stage?

A real absorber contains a certain number of actual trays, or a depth of a specific packing, intended to promote contact across the interface between the countercurrently flowing vapor and liquid phases. These flows are always turbulent, to some extent, and the turbulence level depends on the tray or packing design, along with the fluid properties and flows. Turbulence affects the absorption rate because it affects the mass-transfer coefficients that prevail within the phases.

In parallel with heat-transfer coefficients in various types and designs of heat exchangers, mass-transfer coefficients for a wide range of tray and packing types have been correlated with design details, flow parameters and properties. In other words, the mass-transfer characteristics of tower internals are well understood and well established; there is no guesswork.

Absorption rates also depend on concentration *differences* between the phases. The separation achieved by a real tray, or a certain depth of real packing, depends directly on the absorption rate of the component as dictated by the mass-transfer characteristics of the internals. The model is completely integrated with the operation of real-world equipment.

The ideal stage concept, on the other hand, replaces every important detail with the single, simplifying assumption of equilibrium between the phases. However, since there are no composition differences between phases, there is no reason for a separation. Regardless of applying efficiencies or finessing the equilibrium assumption in any other manner, the ideal stage assumption eliminates reality from the calculations, leaving a model that is not only unable to perceive critically important process details, but that is also vulnerable to gross errors.

LNG CASE STUDIES

This article consists of two case studies. The first study asks the question: How closely can the performance of an LNG plant's CO_2 removal section be predicted? This question is answered by comparing performance predictions with operating data from one of the trains of an operating LNG plant. The second case study is an analysis of certain aspects of the CO_2 removal section of a project under consideration. The case study examines the sensitivity of the design to operating and design parameters.

Operating LNG plant. A proprietary CO_2 removal solvent is used at 41 wt% to treat inlet gas with the composition shown in **TABLE 1.** The process flow diagram (**FIG. 1**) is fairly conventional, although a substantial portion of the reflux water from the stripper overhead condenser is mixed with process makeup water and returned to the top of the absorber, rather than to the regenerator. At the top of the absorber, four wash trays recover the solvent from the treated gas. There are 20 contacting trays in the absorption section.

The rich amine is flashed to remove hydrocarbons, cross-exchanged with hot lean amine, and then sent to the regenerator. The regenerator contains three wash (reflux) trays and 17 stripping trays. The reboiler is energized using hot oil.

Absorber temperature and composition profiles indicate that the tower is mass-transfer-rate controlled. Simulated temperatures of various streams throughout the plant are compared with measured data in **TABLE 2.** Stream numbers refer to those shown in **FIG. 1.**

With the exception of the treated gas temperature, the simulated temperatures of all other streams match plant measurements to better than 1°C; the treated gas differs by only 1.4°C. The measured lean amine loading was 0.04 mole of CO_2 per mole of solvent, while the simulated value was 0.036 mole of CO_2 per mole of solvent. These values are almost identical, validating the accuracy of the regenerator simulation. The treated gas CO_2 level was measured at 25 ppmv, whereas the simulated value was in the range of 50 ppmv–60 ppmv—a greater discrepancy than expected. However, the feed gas is known to contain heavy ends, and plant personnel have identified the presence of foaming in the column. Even a small amount of foaming in an otherwise normally operating system can increase the vapor-liquid area for mass transfer by 10%–20%, which is sufficient to give a predicted value of 25 ppmv–30 ppmv of CO_2.

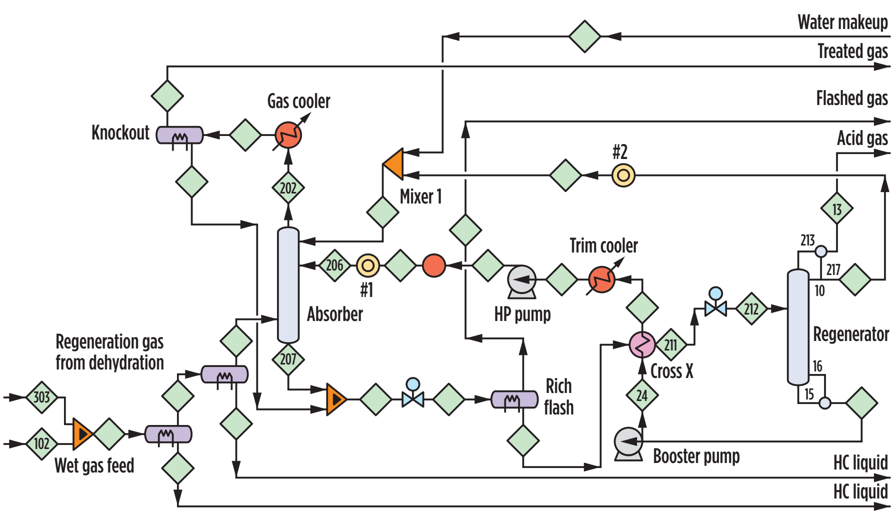

FIG. 1. Simplified process flow diagram for the operating LNG train CO_2 absorption system.

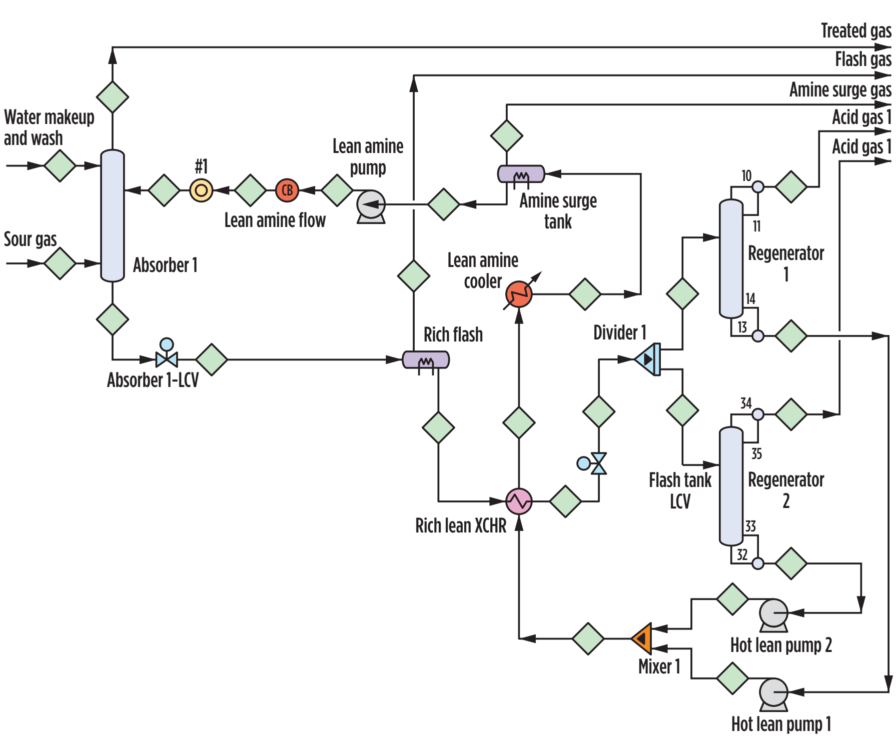

FIG. 2. Simplified process flow diagram for the study case LNG CO_2 absorption system.

LNG project under consideration. This LNG plant is in the initial study phase; therefore, the name of the project and the location are not disclosed. However, the study phase is the time to perform sensitivity studies and determine if it will be operable over the entire range of expected conditions, and whether there are regions in which the operation of the plant might be overly sensitive to one design parameter or another. If such regions exist, then the plant must operate well away from them, or it must have contingencies in place to ensure that stable operation can be maintained.

Regarding CO_2, the raw gas to the plant under consideration is at the opposite end of the spectrum from the operating LNG facility. The CO_2 concentration in the raw gas is about 45 times higher—nominally 16 mol%. **FIG. 2** shows a simplified process flow diagram. A simplified gas analysis with relevant conditions is provided in **TABLE 2.**

The absorber has been designed with three one-pass valve trays to act as wash trays and assist in the recovery of any vaporized, proprietary solvent from the treated gas. This short wash section swages into the absorption section. The main part of the absorber was designed with two identical 4-m-deep beds of proprietary packing rings selected for their excellent hydraulic capacity and high specific surface area.[2, 3] In tests, this packing combines high throughput with excellent mass-transfer performance. At 60% of flood, the absorber diameter is just over 25 feet—which, at 48 barg, is a substantial tower shell by any measure (**TABLE 3**).

The regenerator's design stripping ratio (i.e., the ratio of water vapor to acid gas in the overhead vapor line going to the condenser) is 0.8. The solvent circulation rate is set to achieve a rich amine loading of 0.5 mole of CO_2 per mole of amine in the solvent. At the design lean amine temperature of 115.8°F into the absorber, the treated gas was simulated to contain 21.4 ppmv of CO_2.

A study of how overall performance might respond to variations in design and operating parameters revealed that, for the most part, the CO_2 content of the treated gas was remarkably insensitive, so the design appeared quite solid from an operational standpoint. However, performance was found to be sensitive to both the depth of packing in the absorber and the temperature of the lean amine.

Sensitivity to packing depth. The simulated effect of packed-bed depth on treating performance, as measured by the CO_2 level in the treated gas, is shown in **FIG. 3.** (Note the logarithmic scale on the concentration axis.) The design point of an 8-m-deep packed bed appears to give a comfortable margin away from the 50-ppmv specified limit for CO_2. However, reducing the bed depth to 7 m—a difference of only 1 m—would result in the treated gas exceeding this specification by at least 20 ppmv, vs. meeting it with a comfortable 30-ppmv margin.

Using appropriate packing depth is critical to achieving a tight design. To ensure a truly safe design, it is necessary to have confidence in the reliability of the simulator. If a simulator that uses ideal stages in any form is used, it must rely on the engineer having high confidence in the accuracy of the height-equivalent-to-a-theoretical-plate (HETP) value used to translate calculations into reality. Unfortunately, information on HETP values in amine systems is often unreliable. The proprietary simulator, however, has a real mass and heat transfer rate basis, which allows it to avoid the efficiency and HETP issues. Also, it has been finely tuned to a large amount of commercial plant performance data; the mass-transfer rate model deals directly with the real internals in the tower, not with theoretical HETP values.

Absorber performance is exponentially sensitive to packed depth, mainly because the absorber's performance in this case is controlled—not by the solvent lean loading, but by the mass transfer itself. The easiest way to understand this concept is to look at the CO_2 concentration profile across the absorber, shown in **FIG. 4.** Detail is made visible by using the logarithm of CO_2 concentration, and the temperature profile is shown for reference.

The size of the temperature bulge is substantial in the bulge region of the absorber. The CO_2 concentration is only slowly changing in that region because of the high CO_2 backpressure at such temperatures. In other words, the solvent is nearing saturation toward the bottom of the absorber. However, throughout the rest of the absorber, the CO_2 concentration continues to fall exponentially. Since the CO_2 concentration in equilibrium with the lean amine is less than 1 ppmv, the absorber is far from being lean-end pinched.

Sensitivity to operating parameters. If the absorber goes off-specification, then remedial action must be taken. There are several candidates for control variables, the most obvious being lean solvent flowrate, lean solvent temperature and regenerator reboiler duty. This absorber is sized for only 60% of flood. Therefore, the solvent rate might be an excellent control variable, at least in terms of tower hydraulic capacity (provided, of course, that pumps have been adequately sized). The regenerator reboiler duty, on the other hand, is a poor control variable because treating is almost independent of lean loading (provided that the loading is low enough). **FIG. 5** shows the lack of sensitivity of treating to lean loading. As shown in **FIG. 5,** a 15% increase in reboiler duty results in a lean loading reduction of only 0.0036 loading units, and this decreases CO_2 in the treated gas by only 2 ppmv.

Lean solvent temperature is the final control variable considered. As the lean

TABLE 1. Composition of gas to the CO_2 absorber

CO_2, mol%	0.34
Methane, mol%	95.5
C_2^+, mol%	4.1
N_2, mol%	< 0.1

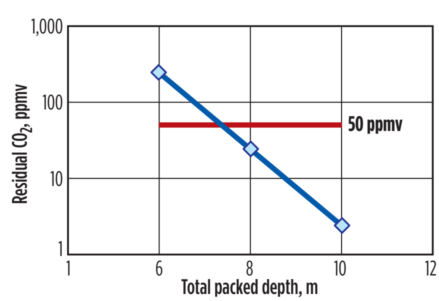

FIG. 3. Sensitivity of CO_2 in treated gas to packing depth in the absorber.

TABLE 2. Actual vs. simulated stream temperatures

Stream	Actual temperature, °C	Temperature achieved with simulation tool, °C	Temperature difference, °C
103	24.8	24.9	−0.1
202	32.5	33.9	−1.4
206	43.3	44.1	−0.8
207	24.2	23.5	0.7
211	81.8	81.8	0
213	111.9	111.8	0.2
13	44.5	44.5	0
24	126.7	126	0.7

amine temperature is increased, temperature increases are expected throughout the column. A higher-temperature solvent is unable to hold as much acid gas as a low-temperature one. In other words, its net loading decreases. This is shown in **FIG. 6**, where the CO_2 level in the treated gas and the net solvent loading are shown side by side. The rich loading remains constant between 100°F (37.8°C) and 126°F (52.2°C), regardless of the lean amine temperature, because the solvent under those conditions is able to absorb virtually all of the CO_2 (at approximately the 20-ppmv level).

However, at or above 126°F, the rich loading begins to drop with increasing temperature. The ability of the solvent to "load up" with CO_2 is compromised as it becomes too hot, and what cannot be absorbed must leave in the exiting gas. Therefore, the treated gas quality suffers very severely. The effect is noticeable when dealing with ppm specifications on the treated gas.

It is instructive to examine how treated gas composition changes with lean amine temperature in more detail. **FIG. 7** is a replot of **FIG. 6** (b) on a logarithmic basis to magnify the region where treating fails. The design temperature is 115°F (46°C). As long as the temperature is kept within 10°F (5°C) of the design point, the plant appears to maintain operational stability.

However, as **FIG. 7** shows, once the lean amine temperature approaches 125°F (52°C), the absorber will become extremely unstable; indeed, it will become inoperable. As already discussed, the reason for this significant sensitivity to an operating condition has to do with the capacity of the solvent; part of the CO_2 in the raw gas cannot be absorbed because of a solvent capacity limit, and so it passes directly into the treated gas. Continuing to decrease the solvent temperature, however, does not result in continued improvement to treated gas quality. Eventually, it starts to have a deleterious effect because the solvent viscosity is increasing with decreasing temperature, and because mass-transfer resistance to CO_2 absorption starts to increase.

The remaining question is: What is the best way to control the absorber when the lean amine is too warm? The answer is fairly simple: Manipulate the variable that will increase the solvent capacity—i.e., its net loading capacity. The choices are limited. Increasing solvent strength is always an option, but it is not a control strategy. Producing a leaner solvent would be completely ineffective because lean loading has only a tiny effect on net loading capacity; in no sense is it controlling. Solvent flowrate, on the other hand, directly affects the solvent's capacity for CO_2; this is the only short-term control variable that would be significantly effective.

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TAKEAWAY

The operating LNG facility is amenable to simulation using the proprietary simulator's mass-transfer-rate-based approach. The simulation is predictive, requiring no estimates or other input beyond what is available from data sheets, piping and instrumentation diagrams, and vendor draw-

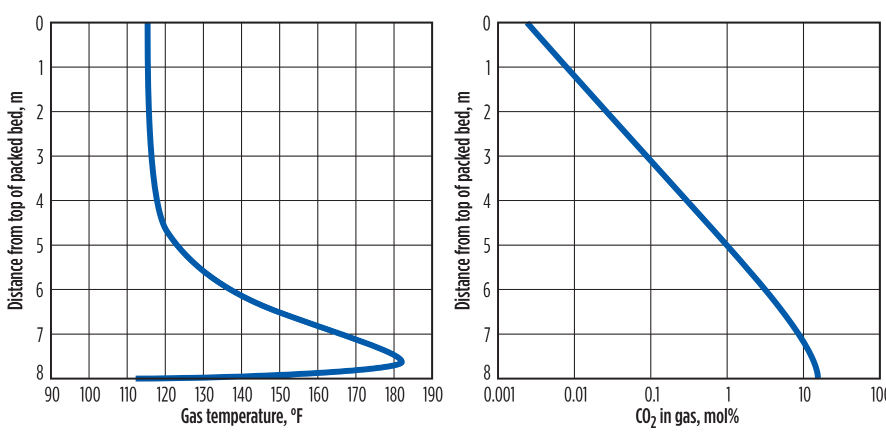

FIG. 4. Temperature and CO_2 concentration profiles in gas across the absorber.

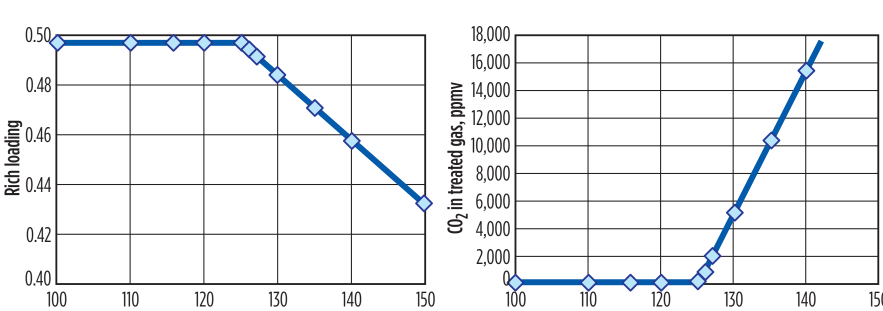

FIG. 6. Net solvent loading and response of CO_2 in treated gas to lean amine temperature.

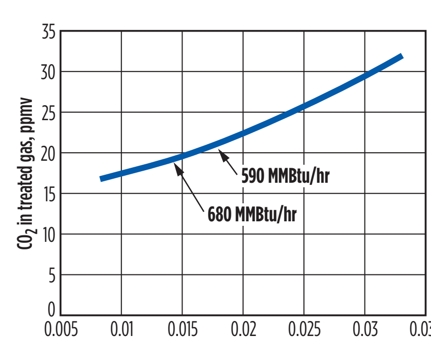

FIG. 5. Insensitivity of CO_2 in treated gas to lean amine loading, and reboiler heat duty.

ings and specifications for tower internals. All temperatures were matched to thermocouple calibrations, and the simulated lean solvent loading was almost identical to measured data. The discrepancy in treated gas CO_2 content was easily explained and accounted for by a small amount of foaming suspected to be occurring in the absorber.

The study case LNG unit appears to be quite sound and operationally stable, but with the proviso that the lean amine temperature must be kept between 100°F and 125°F. If there is any doubt that this can be

TABLE 3. Study case: LNG feed gas to the CO_2 absorber

Temperature, °C	46
Pressure, barg	54
Volume flow, kNm³/h	1,400
Composition, dry basis	
CO_2, mol%	16
Methane, mol%	79
Ethane, mol%	3
Propane, mol%	1
Butane, mol%	0.5
Other, mol%	0.5

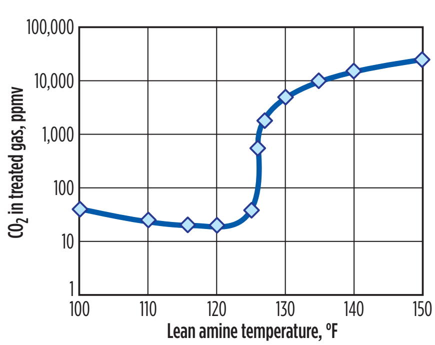

FIG. 7. Extreme sensitivity of CO_2 in treated gas to lean amine temperature.

achieved throughout the year (which may not always be the case in the region for this study case), then adequate provision must be made for operating at increased solvent flow. This contingency requires adequate design margin in pumps, in heat transfer equipment, and in the regenerator itself.

The first case study validated the ability of a proprietary mass-transfer-rate-based simulator to accurately predict amine unit performance in LNG production. In the second case study, the same model was used to explore the operability of a plant still on the drawing board. Without using such a simulator, it would not be obvious that a critical operating temperature exists that cannot be exceeded, and near which the absorber operation will become unstable. There may well be a lower operating temperature, although whether the unit will become unstable at that temperature is a moot point. HP

NOTE

[a] The proprietary software used in this study is Optimized Gas Treating's ProTreat simulator.

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