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Optimising the FCC regenerator for reduced emissions

Increasing scrutiny of FCC stack emissions calls for special attention to be paid to the regenerator's combustion zone

RAY FLETCHER and MARTIN EVANS
Intercat

Continuous attention is focused upon the optimisation of the FCC conversion section for maximum profitability. However, few process engineers have placed much attention on the regenerator unless the operation is close to emissions constraints or out of compliance. The regenerator combustion zone operates with a highly complex and sometimes competing set of reactions. Distribution of spent catalyst and combustion air within the combustion zone and between cyclones adds an additional level of complexity to the FCC regenerator. The prudent FCC engineer will also ensure that the regenerator has been optimised in order that the FCC unit may operate with as few constraints as possible.

Many techniques have been developed over the years that enable today's process engineers to troubleshoot and optimise the FCC regenerator combustion zone and related hardware. The following topics will be addressed within this article:

- Regenerator bed level
- Catalyst attrition
- Cyclone integrity
- Analysis and control of afterburning
- Control of SO_x emissions
- Analysis and control of NO_x emissions.

Bed level

Bubbling bed regenerators are designed with a minimum transport disengaging height (TDH) to ensure the least possible catalyst entrainment in the primary cyclone inlets. Violating minimum TDH

constraints results in a large increase in catalyst carryover into the cyclones, leading to overloading and higher losses. It is crucial that the process engineer ensures that the bed level taps are reading accurately whenever catalyst losses exceed baseline levels. When in doubt, these level indication taps should be blown down to ensure accurate measurement. It is recommended that these level taps be continuously purged via a critical-flow restriction orifice, with a fail-safe backup having an exit velocity of 0.9 m/s (3 ft/s).

Some refiners intentionally violate the minimum TDH constraint in order to process additional charge. These refiners will, of course, observe regenerator losses that are greater than optimal. However, there are techniques to minimise loss rates, which include setting a maximum regenerator superficial velocity, and, for those units equipped with CO boilers, employing partial-burn operations.

Catalyst reformulations to a lower apparent bed density (ABD) system cause an increase in regenerator bed level. Most new units are equipped with at least one set of fully submerged taps to enable on-line measurement of bed density. These taps measure the bed density in real-time, which is then cascaded to the level indicator, enabling the most accurate measurement of bed levels. The process engineer monitoring an older unit without such taps needs to estimate the new bed density during change-out and have the bed level calculation updated to ensure the bed level is not actually higher than believed while

transitioning to a lower density catalyst.

An additional catalyst level not normally considered by those monitoring and troubleshooting the FCC regenerator is the level within the cyclone diplegs.¹ Most FCC units operate with negative pressure cyclones, in which the pressure in the cyclone is less than the pressure in the regenerator vessel. The mechanism that enables catalyst in a cyclone operating at a lower pressure to discharge into the higher pressure of the regenerator dense bed is the dipleg catalyst level. The height in the dipleg is a function of both the regenerator bed level plus the differential pressure between the cyclone and the regenerator. The dipleg level increases until these two pressures have equilibrated:

$$h_{dl} = \frac{dP_{cy} + \rho_{bed} \times h_{bed}}{\rho_{dl}}$$

Where:

h_{dl} = Catalyst height in cyclone dipleg

dP_{cy} = Cyclone pressure differential pressure

ρ_{bed} = Regenerator bed density

h_{bed} = Length of dipleg submerged in the regenerator dense bed

ρ_{dl} = Cyclone dipleg density

It is recommended that the engineer calculate the dipleg bed levels and plot these levels against catalyst losses. There is typically an inflection point at which catalyst losses increase rapidly above a certain dipleg level. The cyclone dipleg levels may be dropped by reducing the charge rate, increasing the regenerator pressure or

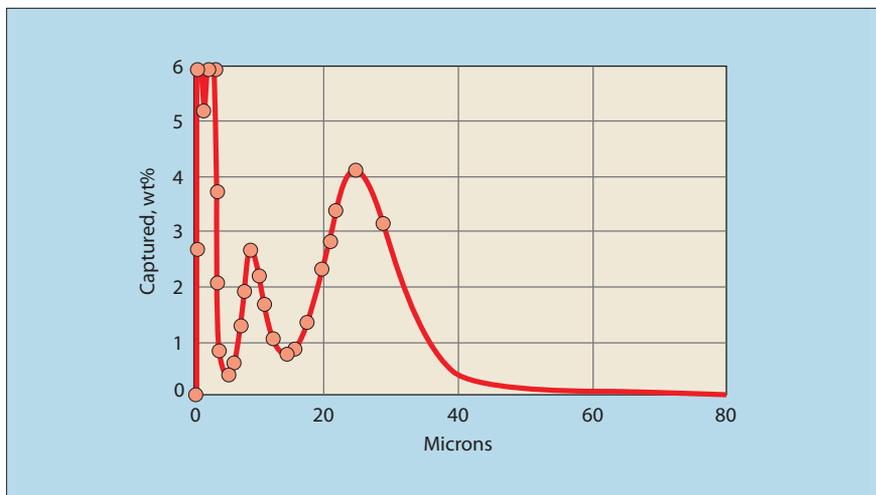


Figure 1 Normal fines distribution

reducing the combustion air rate. Dropping dipleg levels below this inflection point reduces catalyst losses if the critical dipleg level has been violated.

Catalyst attrition

One of the causes of attrition is through catalyst particles being struck by high-velocity air or steam jets within the regenerator, resulting in particle-to-particle and particle-to-wall collisions. The result of these collisions is fracturing of the catalyst particles into ever smaller particles.

A second mechanism that produces attrition fines is grinding. This is the result of particle-to-particle abrasion, which results in the breaking off of small surface nodules. Most units have both mechanisms present, with the relative ratio of the two mechanisms changing from unit to unit.

Critical jet velocities for regenerator air grid and steam nozzles are a function of both equipment design and catalyst attrition resistance. One FCC designer cites the maximum regenerator nozzle velocities for minimising attrition at approximately 30 m/s (100 ft/s). It is recommended that the process engineer consult with their unit designer for recommended maximum velocity constraints.

The best method to swiftly identify the presence of attrition sources in the regenerator is to perform a particle size distribution analysis on the third-stage separator underflow (fourth-stage catch) or on the fines collected within the first-stage bin of an electrostatic precipitator. Care must be taken to ensure that only first-stage fines are analysed. The first stage is most representative of fines being emitted from the FCC regenerator. Fines collected from

the second-stage or higher bins will be, almost entirely, very low particle size distribution (PSD) fines and lead to an improper analysis of the magnitude of attrition.

A cross plot of the weight per cent captured against particle size should be plotted. A normally operating unit produces a typical Gaussian distribution centred in the 20–30 μ range, with two additional small peaks at less than 20 μ (see Figure 1). Figure 2 represents a unit experiencing attrition. A unit experiencing attrition presents an extremely large peak in the 0–5 μ range. Furthermore, the normal peak is shifted to the left in the direction of smaller particle sizes and becomes smaller in magnitude. A comparison of the area under the attrition peak in relation to the total area under the curve provides a reliable indication of the magnitude of the attrition source.

Most units present a minimal content of small particles resulting from attrition on a continuous basis. It is strongly recommended that a baseline be established under normal operations. This serves as the reference point for comparison of abnormal to standard operations when troubleshooting catalyst attrition.

Attrition of equilibrium and fresh catalyst appears identical in the fines distribution chart described above. A comparison of the surface area and contaminant metals content of the fines with the equilibrium catalyst enables the process engineer to distinguish between soft catalyst and an attrition source. If the attrition product originates primarily from fresh catalyst, the surface area of the fines is higher than the equilibrium catalyst. Furthermore, the contaminant metals (Ni, V, and so on) levels of the fines are lower than the equilibrium catalyst. Again, regular monitoring of these values while the unit is operating normally provides the baseline for comparison during periods of high losses.

Catalyst attrition is most often observed at the start of run and near the end of a long operating cycle. Start-of-run attrition sources are most often the result of missing

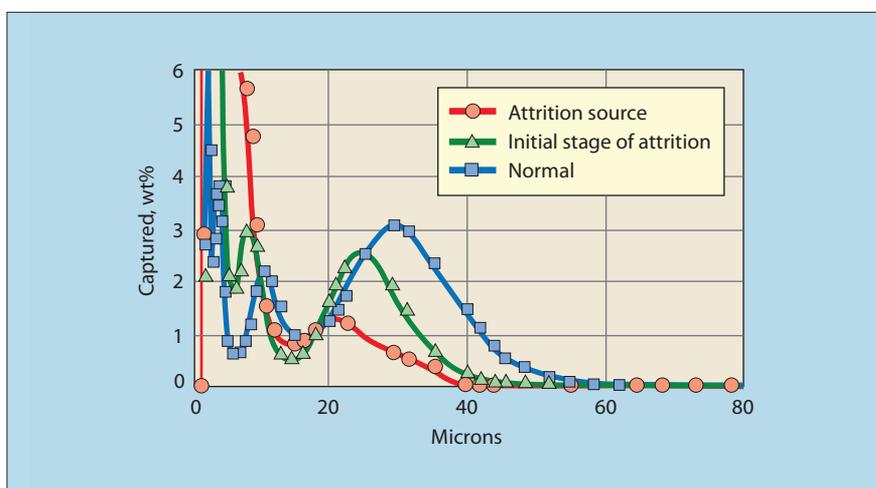
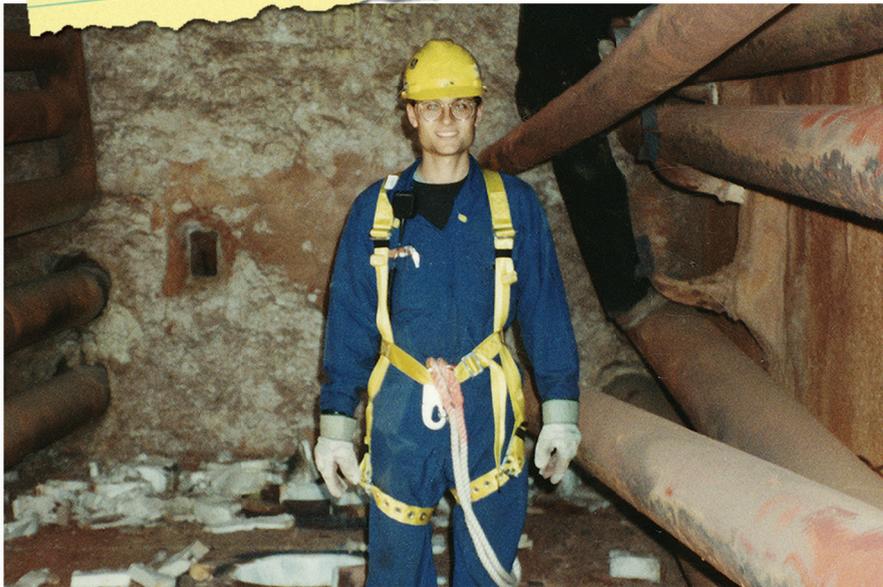


Figure 2 Attrition source



Process Notes



Designing Deepcut Vacuum Units That Really Work

Every barrel of vacuum gas oil (VGO) you can save from being reduced to coke in the delayed coker unit is a barrel more that can go to the FCCU. That's a good reason to raise HVGO cut-point. But how to do it? Some people think the job can be done just by running computer models in the engineering office, relying on vendors and their data sheets

for a clue to true equipment performance. Experience, however, shows it just ain't so. For either a grass roots project or a revamp, designing a deepcut vacuum unit to handle heavy sour crudes that are becoming more common today demands a lot more.

It takes personal experience in doing tube-by-tube design of the fired heaters and knowing where and how to inject coil steam, managing the hydraulics of the heater including critical two-phase flow in the tubes and transfer line, how to deal with corrosion and fouling in heat exchangers, balance the intricate details of tower internals and

vacuum ejectors- and above all, mate these interdependent components into a fully integrated system that reliably and consistently maximizes VGO yields low in microcarbon and vanadium.

Difficult? Not a job, certainly, for anyone who only sits in the office doing nothing but running simulations. This isn't to say that models are not important. They are. But to produce anything of real value these process and equipment models need to have been tested and checked against actual measured pressures, temperatures, flows and stream compositions. And the person who runs the simulations must always keep in mind the dictum of one of the grand old men of refining: "Fluids obey the laws of physics and not the whims of the process designer." In these times when every barrel of crude needs to be converted to highest value products, one must look to the expertise not just of the engineers who run the models, but of those who also wear Nomex® and get dirty measuring real unit performance.



To learn more about designing and operating deepcut vacuum units, ask for Technical Papers 185, 199, and 205.



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or improperly sized restriction orifices. These orifices are generally found on purge streams associated with torch oil nozzles and instrument taps, for instance. Having an up-to-date listing of all orifices together with their proper sizes and locations speeds up the troubleshooting process. Orifice sizes are typically stamped on the orifice handle, enabling swift verification of proper sizing. All orifices should be removed and inspected for wear during each turnaround.

End-of-run attrition sources are frequently attributable to damaged air nozzles on the air distribution grids. Great care must be exercised during periods of operation at reduced charge rates. A reduced charge rate results in lower combus-

possessing multiple bends coupled with the presence of high velocities (>100 ft/s) frequently undergoes load line attrition. Multiple designs exist for load line configuration and include long-radius sweep Ls, blind Ts and 90° elbows. The load line design should contain as few bends and pipe swedges (a transition from a larger to smaller diameter in the catalyst transfer line) as possible.

Cyclone integrity

Cyclones are typically designed for a lifetime of 15 years or more. Monitoring cyclone integrity is well within the capability of the process engineer. Monitoring the condition of the regenerator cyclones requires the engineer to regularly acquire samples of the fines captured in the

approximately 25 μ , with a range of 20–30 μ . A highly efficient set of cyclones present the primary peak at about 20 μ , while a set of cyclones having a poor efficiency present the primary peak at approximately 30 μ . Those units that are damaged, significantly exceeding design conditions, or are near the end of their normal lifespan, present a primary peak greater than 30 μ . Additionally, it is typical to observe a small attrition peak at between 0 μ and 5 μ . A third peak is also observed at approximately 10–15 μ . These two peaks are likely to be the result of attrition and particle collisions within the secondary cyclones.

A unit experiencing attrition presents a large peak at approximately 0–5 μ , together with a shifting of the primary peak to the left in the direction of smaller particle sizes coupled with a reduction in peak magnitude. These shifts are demonstrated in a unit presenting the primary peak at about 28 μ under normal operations (see Figure 2). With the initiation of attrition, this primary peak was shifted to the left, eventually equilibrating at 20 μ . Additionally, the magnitude of the attrition peak gradually increased until it overwhelmed even the primary peak.

A hole in one of the primary cyclones is indicated by an abnormal peak shift to the right of the normal peak. This abnormal peak is typically positioned at 60 μ or greater (see Figure 3). The two smaller peaks to the left of 20 μ continue to be present.

A hole in one of the secondary cyclones is also indicated by an abnormal peak shifted to the right of the normal peak. However, this abnormal peak typically presents itself at 45–50 μ . Figure 4 represents a unit that developed a hole in one of its secondary cyclones. In this case, the intermediate curve demonstrates the expected bimodal distribution. As the damage to the cyclone progressed, the second abnormal peak eventually overwhelmed all other peaks and was centred at 45 μ . In these cases, the two peaks at less than 20 μ disappear.

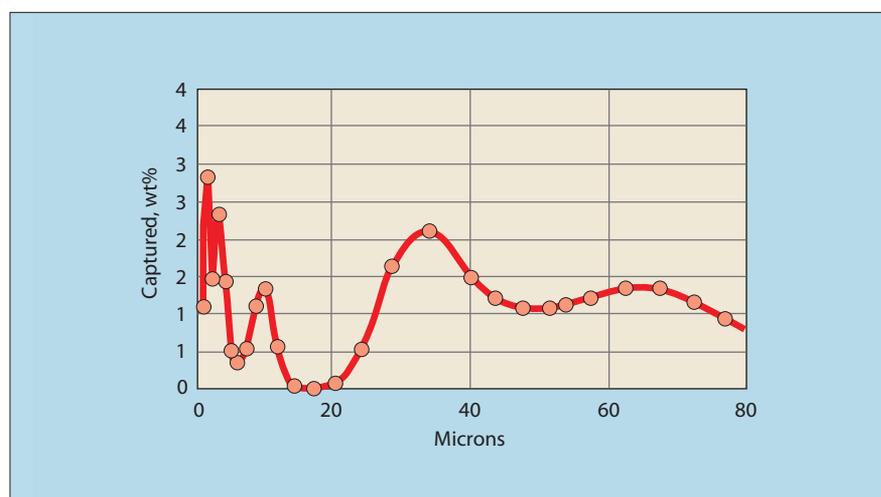


Figure 3 Primary cyclone hole

tion air requirements, leading to a reduced air grid back pressure. With insufficient airflow, catalyst begins to back flow into and circulate through the air nozzles at the extreme ends of the air grid arms. This flow rapidly leads to holes being formed within the air grid since the interior surfaces are not protected with abrasion lining. Once the unit is returned to normal rates, this hole produces a very high-velocity jet, leading to immediate attrition. It is recommended that the unit's pressure be reduced during periods of turndown to ensure normal airflow through the entire length of the air grid arms.

Additionally, the process engineer should not ignore the fresh catalyst transfer line. A configuration

first stage of the electrostatic precipitator or the third-stage separator fines underflow.² A detailed particle size distribution analysis is required. The weight per cent captured is then plotted against the particle size distribution.

The use of particle size distribution charts for analysis and troubleshooting of both regenerator and reactor cyclones is most effective for troubleshooting if a baseline has been established during normal operating conditions. This baseline establishes normal operations whenever cyclone performance is being investigated.

A normal fines distribution is presented in Figure 1.³ A normally operating set of cyclones produce a primary peak typically centred at

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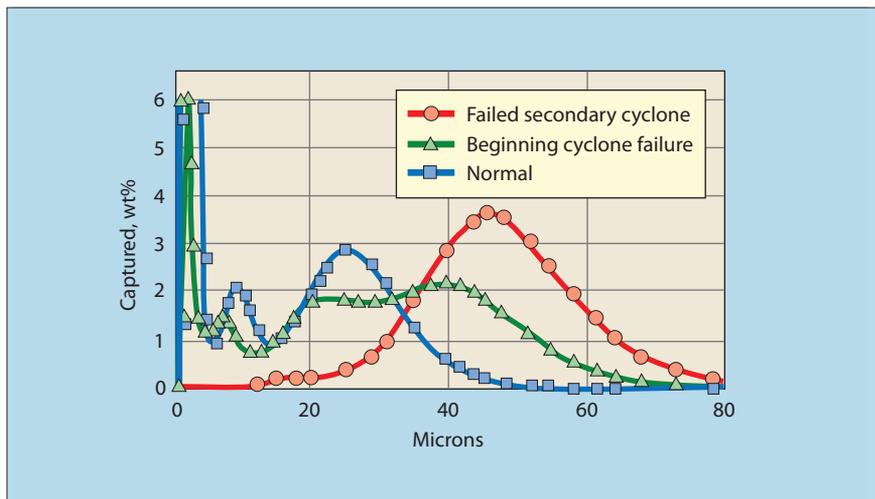


Figure 4 Secondary cyclone hole

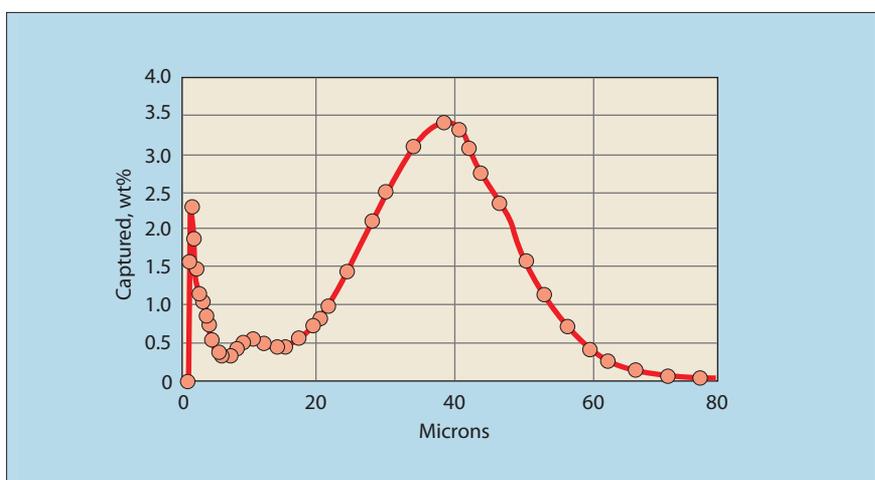


Figure 5 Flooded secondary cyclone

Analyser measurements to detect regenerator oxygen deficiency					
Position 1	0.1	16.6	>10 000	482	17
Position 2	1.4	15.6	4276	127	51
Position 3	5.1	12.6	23	4	86

Table 1

Cyclone flooding (see Figure 5) presents a particle size distribution curve very similar to a hole in a secondary cyclone. The chief distinguishing characteristic between these two curves is that the two peaks at less than 20 μ remain present.

It is strongly recommended that the fines distribution curves for the regenerator cyclones be monitored consistently over the life of the cyclones. This forms a baseline for comparison with abnormal operations if losses are observed. Monthly sampling is adequate for this

step once a baseline has been established.

Afterburning

In FCC units, afterburn is defined as any temperature increase in the flue gas following the dense bed. This can take place in the dilute phase, the cyclones or the flue gas line. Afterburn can limit feed rate or reduce feedstock flexibility. Severe afterburning can damage regenerator internals or flue gas system components, leading to premature shutdown of the FCC unit and costly repairs.⁴

Afterburning is the result of the combustion of CO to CO₂ in the regenerator cyclones or flue gas system. This is caused by incomplete combustion of carbon in the dense bed.

Two types of afterburn are commonly seen commercially. The first of these, thermodynamically limited afterburn, normally responds well to the addition of CO promoter. The second type of afterburn is caused by poor catalyst and/or air distribution and typically responds less well to the use of CO promoter.

These types of afterburn are usually fairly easily distinguished. The first type occurs where there is insufficient residence time in the regenerator dense bed for the CO to CO₂ reaction to be completed. This can occur in a unit that has high superficial velocities and low bed levels, and is frequently seen in units that have low dense bed temperatures. With this type of afterburning, the high temperatures are normally fairly well distributed around the regenerator. In this case, all that is required to eliminate the problem is to increase the rate of reaction of CO to CO₂. This can be done by increasing the dense bed temperature (the rate of reaction doubles for every 10°C increase), or by the addition of CO combustion promoter. Historically, most CO promoters have been platinum based, although many refiners are now using non-platinum promoters such as COP-NP to eliminate the increase in NO_x emissions seen when platinum is used.

Regenerators that are experiencing afterburning caused by poor distribution frequently exhibit a region of the dense bed that is carbon rich and in partial combustion, while another section of the dense bed is carbon poor and in full combustion, with excess oxygen exiting the dense bed. The carbon-rich zone delivers CO to the cyclones, while the carbon-poor zone delivers excess oxygen to the cyclones. A portable flue gas analyser may be used to take oxygen concentration measurements around the circumference of the regenerator to identify regions

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where oxygen deficiencies exist. An example of this technique is provided in Table 1.

Regenerator combustion gas analysis

Combustion commences immediately when the gases from these two zones begin to mix in the cyclones or plenum. At this point in the regenerator, there is little catalyst to act as a heat trap to absorb the heat of combustion. This likely leads to very high temperatures in the regenerator overhead, which may potentially become a limiting factor in the operation of the unit. This is often the case when the flue gas temperatures begin to approach metallurgical constraints.

Afterburning that is localised in nature may be intermittent or continuous. Additionally, the localised afterburning may shift to other regions within the regenerator. The elevation of afterburning is a function of regenerator temperature. If temperatures are low, afterburning usually appears in the flue gas line. As temperatures increase, afterburning is shifted in the direction of the dilute phase.

The root cause of afterburning in many cases is directly related to decisions made during the design of the FCC unit itself. Vertical and curved spent catalyst lift lines tend to deliver high-density spent catalyst into the centre of the regenerator dense bed. If the bed is shallow, there may be insufficient time for the carbon to be completely combusted to CO_2 , leading to CO breakthrough. The regions of the regenerator surrounding this dense zone are likely to be in full burn with excess oxygen. This excess oxygen results in afterburning immediately upon mixing with the CO-rich flue gas from the other parts of the bed. Side entry regenerators, either tangential or radial, tend to result in dense carbon-rich zones near the point of entry, together with oxygen-rich zones elsewhere in the regenerator, resulting in afterburning.

Empirical evidence has shown that the most effective means of spent catalyst distribution with respect to minimising afterburning

is with spent catalyst distributors. These distributors take advantage of the higher density of the spent catalyst. The spent catalyst is often distributed across the cross-sectional area of the regenerator via troughs. There are many other potential solutions to afterburning, which include splash plates for side entry regenerators. To be effective, the splash plate must be located above the regenerator bed; splash plates submerged in the bed provide very little catalyst distribution.

Localised afterburn appearing in a normally well-behaved unit is most likely due to air distributor damage. Damaged air distributors generally show a reduction in pressure drop. This change in pressure drop may sometimes be detected by a valve position change in the combustion air control valve. Damage affecting the flow of spent catalyst into the regenerator may be observed in the spent slide valve position. An increase in catalyst losses or a shift in the particle size distribution of the equilibrium catalyst together with the appearance of afterburn indicates internal damage.

The most common response to afterburning is to use a CO promoter. There are two types of promoters in use today: platinum and non-platinum based. Non-platinum-based promoters are used when NO_x emissions need to be controlled.

CO promoters are extremely effective at reducing and, in many cases, eliminating thermodynamically induced afterburn. CO promoters are usually added to achieve an equivalent platinum concentration on total catalyst additions of about 1 ppm. Most units do not show any additional reduction in afterburning above this level, although NO_x emissions continue to increase with added platinum. CO promoters usually rapidly eliminate afterburn in cases where the afterburn is generalised across the cross-section of the regenerator. CO promoters will be somewhat less effective in those cases in which the afterburn is localised or caused by poor distribution. In these cases, the promoter reduces afterburn until all

oxygen is consumed in the carbon-rich zone. After this point, additional platinum is less effective. Careful monitoring of the localised hotspot versus CO promoter additions is strongly advised. CO promoter additions after the hotspot fail to show temperature reduction can sometimes result in the dilute-phase temperatures in the remainder of the regenerator falling below the average dense bed temperature.

In units with poor distribution, some refiners have seen a benefit from using a lower platinum content CO promoter — that is, adding the same amount of platinum distributed over a larger number of particles. This effect is, however, not seen in all units.

Many refiners utilise antimony as a method to reduce the impact of nickel on hydrogen and delta coke. Antimony is a poison to platinum. Injection of antimony while utilising CO promoters often results in a 50% or larger reduction in CO promoter effectiveness.

SO_x emissions control

Approximately 10% of the sulphur present in the FCC charge will be emitted as SO_x (SO_2 and SO_3) in the flue gas stack. The typical range is 5–35% for most operating units. SO_x -reducing additives were developed by ARCO in the 1970s and were magnesium aluminate spinel-based technology that underwent further improvements.⁵ These additives have remained basically unchanged since the early 1990s and are still in use today. A further development in SO_x reduction additives was introduced by Akzo Nobel in the 1980s and was based upon hydrotalcite technology. In 1997, Intercat developed and patented a self-supporting hydrotalcite that overcame many previous technology limitations. The resulting technology is a hydrothermally stable, attrition-resistant material with superior performance compared to spinel-based additives.

Feed quality is the most significant factor affecting SO_x emissions from an FCC unit (see Figure 6). The sulphur content and the particular sulphur species present in the

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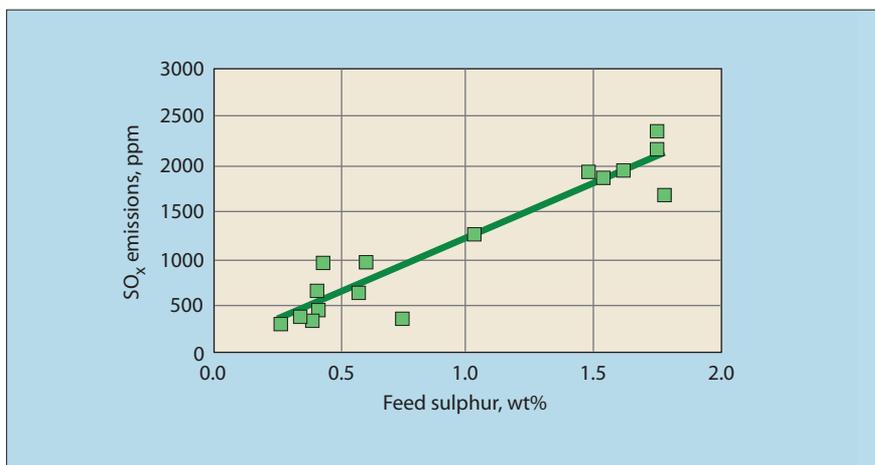


Figure 6 Effect of feed sulphur

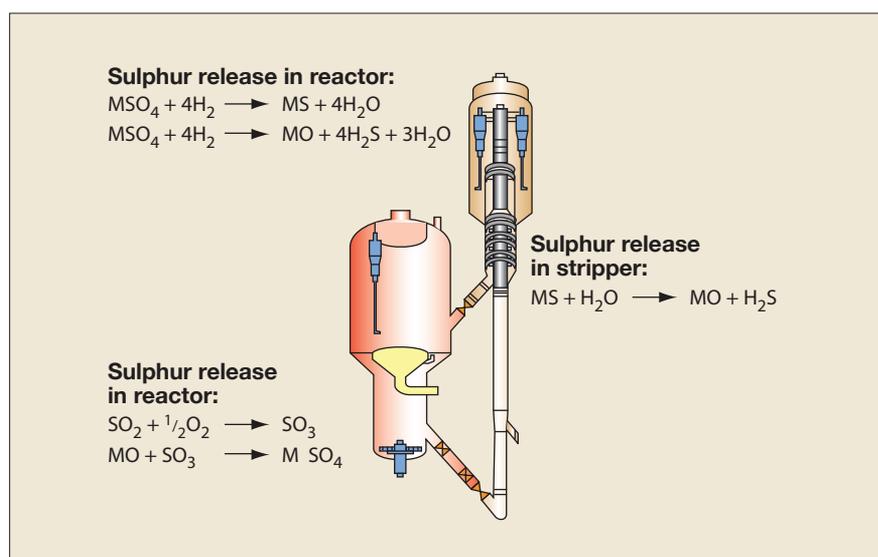


Figure 7 SO_x additive reactions

feed strongly determine the extent of potential SO_x emissions.⁶ Careful monitoring of feedstock quality will alert the process engineer to potential increases in SO_x emissions prior to occurrence.

The content of excess oxygen in the flue gas can strongly influence additive efficiency.⁷ Oxygen is required to drive the SO₂-to-SO₃

reaction that must take place in the regenerator before the SO_x additive can pick up the SO₃ and transport it to the reactor (see Figure 7). Higher concentrations of SO₃ will be produced in the presence of excess oxygen, so SO_x reduction additives tend to be more effective in full combustion regenerators. However, increasing excess oxygen levels in

the flue gas above certain concentrations will have little or no impact on SO_x additive efficiency. For many units, this level is reached at approximately 2%. In some units, increasing excess oxygen may be more effective in reducing SO_x emissions than adding more SO_x additive. See Table 2 for examples of SO_x reduction in full combustion units.

Lower regenerator temperatures tend to favour SO₃ formation, while good air distribution and mixing in full-burn regenerators enhance SO₃ pick-up. Large regenerator inventories will generally reduce the efficiency of an additive.

Higher levels of SO_x emissions are observed in units with inefficient strippers. Poor stripper efficiency increases the sulphur loading to the regenerator, since the heavy oil carryover contains high sulphur concentrations.

Increasing the catalyst circulation rate increases the availability of fresh metal oxides for SO₃ pick-up, thereby reducing SO_x emissions. Careful monitoring of the catalyst addition rate is important while adding SO_x-reducing additives.

The presence of CO promoter can help catalyse the oxidation of SO₂ to SO₃ and, in situations in which the rate of oxidation is the limiting step, can enhance the SO_x removal process.

The FCC catalyst itself may also have a very minor influence on the reduction of SO_x emissions. The active alumina in some FCC catalysts plays a limited role as a pick-up agent for SO₃ (similar to magnesium oxide). However, the fresh catalyst lacks the oxidants that enhance the effectiveness of SO_x

SO_x additive performance in full-burn units

Additive	SG	SSG	SG	SSG	SSG	SSG	SSG	SSG	SSG	SSG
Feed rate, TPD	7566	7559	2308	2319	15 194	4503	2404	2210	2120	5114
Sulphur, wt%	0.42	0.40	0.80	0.80	1.08	0.60	0.50	0.75	0.80	0.72
Slurry sulphur, wt%	1.41	1.34	1.90	2.00	1.99	1.45	1.10	2.10	1.89	1.60
Flue gas O ₂	3.46	3.41	1.80	1.90	1.38	1.24	2.80	2.10	2.60	2.50
Additive addition, kg/day	63	63	151	92	276	98	188	170	147	125
SO ₂ uncontrolled	282	277	1138	1189	689	1665	775	1255	1061	744
SO ₂ controlled			476	498	241	324	122	137	238	288
SO ₂ reduction, %			58	58	65	81	84	89	78	59
Additive efficiency, kg SO ₂ /kg additive	16	29	22	36	33	75	41	29	28	35

Table 2

reduction additives. Furthermore, the aluminas in FCC catalysts are unable to be regenerated in the reducing atmosphere of the reactor vessel as with SO_x additives.

One interesting example of the observation cited above is that the FCC catalyst itself has at times contributed more SO_x reduction than expected in residue operations, with high catalyst turnover rates coupled with high vanadium levels. The high catalyst turnover rate supplies fresh alumina for SO_x absorption, while the high vanadium contaminant levels enable some minor release of H₂S in the reactor vessel. This has been observed in several units when switching from low to high alumina-bearing formulations, or vice versa.

The limiting reactant in full-burn operations is the concentration of magnesium oxide. The higher concentration of magnesium oxide in hydrotalcite-based additives has resulted in more applications globally than spinel-based technologies. The limiting reactant in partial-burn operations is oxygen. In partial burn, the ratio for the combustion of sulphur to SO₂ versus COS is critical. An extremely well-mixed bed with countercurrent flow can lead to nearly all the sulphur being combusted to COS. It is very difficult for SO_x additives to further oxidise COS to SO₂ and SO₃. It is therefore advised that a portable gas analyser be utilised to measure SO₂ concentration in the flue gas line upstream of the CO boiler prior to utilising SO_x additives in partial-burn operations.

Oxygen availability is usually the factor limiting SO_x additive efficiency in partial-burn regenerators. Special oxidation packages have been developed and included in certain SO_x technologies to enable efficient SO₃ capture. Therefore, technologies for absorption of SO₂ have been developed and are being utilised successfully in these more challenging operations. Examples of such technology are provided in Tables 3 and 4.

In most partial-burn regenerators, there is a practical limitation to the

Example of LoSO _x -PB commercial data			
	Super SOXGETTER	Super NoSO _x	Lo-SO _x PB
Dates of trial	12 Oct-2 Nov 06	1 Jan-5 Feb 07	28 Feb-5 Mar 07
Fresh feed rate, TPD	3108	3328	2971
Fresh feed sulphur, wt%	0.53	0.55	0.43
Slurry sulphur, wt%	1.99	2.08	1.84
SO ₂ uncontrolled, kg/hr	79.4	87.5	67.1
SO ₂ controlled, kg/hr	56.7	54.9	27.2
Reduction, %	28.7	37.4	59.5
Additive efficiency steady state (PUF), kg/kg	Base	1.2 x base	2.2 x base

Table 3

amount of SO_x that can be removed. SO_x additives remain effective until all of the available SO₂ has been absorbed. Little or no additional SO_x reduction is achieved when adding beyond this level. Monitoring of the SO₂ concentration with a portable gas analyser in the flue gas line during the early stages of injection may be helpful to ensure that excess SO_x additive is not injected. It is also worth noting the importance of exercising care during base loading operations in partial-burn units. Since the oxidation components contained in SO_x additives can also catalyse the CO-to-CO₂ reaction, these must be carefully controlled to avoid disturbing combustion patterns and the CO-to-CO₂ ratio in the regenerator. Intercat has developed a set of base-loading criteria for partial-burn operations.

NO_x emissions control

During regeneration, some of the nitrogen present in the coke on the spent catalyst is converted to NO and is observed as NO_x emissions from the FCC unit.⁸ However, under normal regenerator operations, most of the nitrogen evolved from coke is converted to N₂. Only about 10% of the nitrogen in coke is

actually emitted as NO_x. The regenerator contains a variety of reductants, including CO and unburned coke on catalyst that can react with NO_x to reduce it to N₂. CO is present in the regenerator in relatively high concentrations, especially in unpromoted or partial-burn operations.

Although NO_x can react with both coke and CO, the reaction with CO to form N₂ and CO₂ is probably the most important, since it occurs as a homogeneous gas-phase reaction. This may explain the observed increase in NO_x emissions with the addition of a CO combustion promoter (less CO in the regenerator), and the extremely low NO_x emissions that are seen in the CO-rich flue gas from partial-burn regenerators.

The level of excess O₂ measured in the FCC flue gas is probably the most important operating parameter affecting the level of NO_x emissions from a particular FCC unit. In full-combustion units, a strong dependence has been observed in practically all trials between NO_x emissions and excess O₂ in the flue gas. Figure 8 shows a typical NO_x-to-O₂ relationship. Poor distribution of O₂ in the regenerator bed can result in large variations of

SO _x additive performance in partial-burn units			
Additive usage (kg/day)	Super SOXGETTER	Super NoSO _x	Lo-SO _x PB
Charge rate, BPD	22147	23714	21174
Feed sulphur, wt%	0.53	0.55	0.43
Slurry sulphur, wt%	1.99	2.08	1.84
SO ₂ uncontrolled, ppm	175	193	148
SO ₂ controlled, ppm	125	121	60
Reduction, %	29	37.4	59.5
Additive efficiency (PUF)	Base	1.2 x base	2.2 x base

Table 4

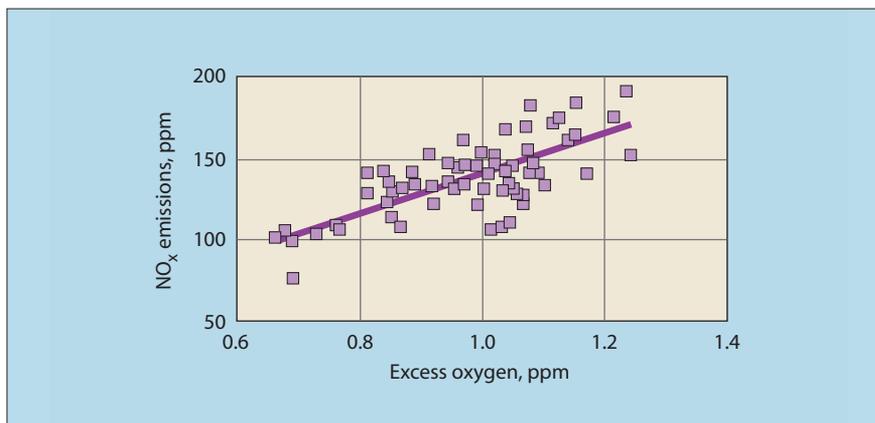


Figure 8 Effect of excess oxygen on NO_x emissions

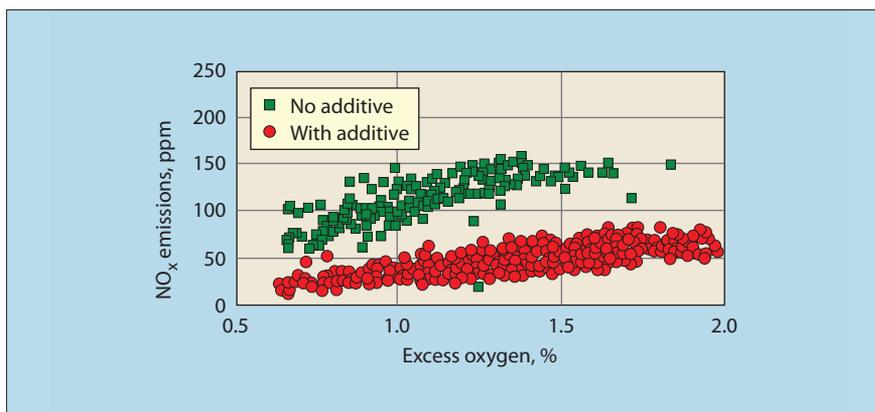


Figure 9 NO_x additive performance

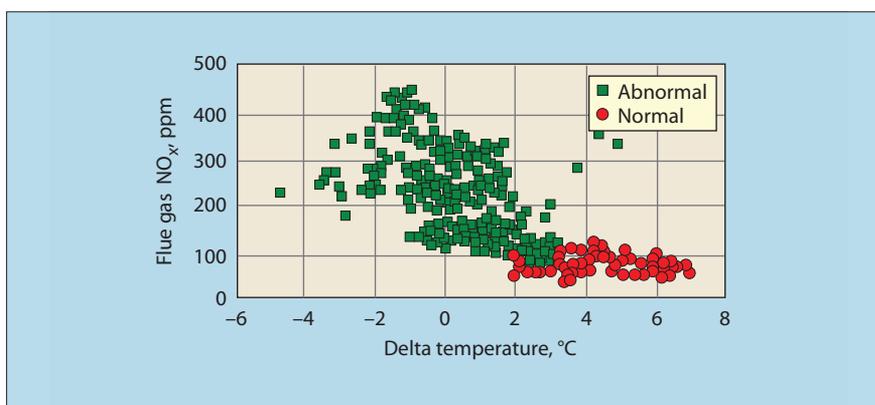


Figure 10 Cyclone delta temperatures 1 to 2

NO_x within various regions of the regenerator. Competition between other gaseous species can also affect NO_x levels. This implies that manipulation of the NO_x-to-O₂ relationship is crucial for reducing NO_x from the FCC unit.

NO_x emissions from the FCC unit are highly dependent upon regenerator design. Controlling the air-catalyst mixing is the most important design parameter for minimising the formation of NO_x. Combustor-style regenerators with superior air-catalyst mixing have

been shown to control afterburn and emit substantially less NO_x than typical bubbling bed regenerators.⁹ The air-catalyst mixing in bubbling bed regenerator designs may also be improved. Counter-current combustion may be initiated via the uniform distribution of spent catalyst across the top of the regenerator bed in conjunction with a symmetrical air grid for uniform radial airflow up through the bed. This design with the spent catalyst moving downward against the upward-flowing combustion air

and gases has resulted in lower levels of NO_x emissions in some regenerators.¹⁰

NO_x reduction additives are effective for limiting NO_x formation in full-burn regenerators (see Figure 9). However, most of today's NO_x additives remain ineffective in partial burn. The objective of additives is to provide the refiner with a simple, low-cost alternative to the capital-intensive hardware options. The use of additives can be implemented quickly so as to effect immediate reductions in NO_x emissions. In addition, when added to the FCC unit separate from the catalyst, the additive concentration in the unit inventory can be adjusted to determine the optimum additive level and achieve the desired reductions in NO_x emissions.

Unlike SO_x additives, there is always a maximum effective concentration for NO_x additives. The NO_x reduction effect will increase with additive concentration until this maximum effective concentration is reached. Increasing the NO_x additive concentration above this level will give no additional benefits and can actually lead to increased NO_x emissions at very high concentrations.

NO_x emissions are strongly affected by maldistribution of air and/or catalyst within the regenerator bed. A sudden increase in NO_x emissions in a regenerator that has been operating stably is generally an indication of maldistribution within the regenerator. A simple technique enabling the process engineer to accurately determine where the maldistribution exists, and possibly its source, can be achieved via analysis of radial temperature profiles in the regenerator. It is recommended that the delta temperature between cyclone inlet and outlet temperatures be plotted versus NO_x emissions for each cyclone set.

Plotting the delta temperatures between cyclone set outlets against NO_x emissions is recommended. For example, a regenerator containing three sets of cyclones should be analysed for delta temperatures between sets 1 and 2, sets 2 and 3, and sets 1 and 3. Analysis of these

plots for step changes will provide an indication when the maldistribution began. Time plots of typical regenerator variables such as combustion air rate, coke loading, pressure, space velocity and unit charge rate will usually provide an indication of what has changed in the unit.

As an example, one unit observed a large, unexplained increase in NO_x emissions. Comparison of the delta temperature plots described above enabled the process engineer to identify the presence of maldistribution within the regenerator. The delta temperature patterns observed between cyclone sets 1 and 2 and sets 1 and 3 were similar, whereas the delta temperature patterns between cyclone sets 2 and 3 were markedly different. Additionally, multivariable linear regression of standard operating parameters together with delta temperatures between cyclone sets identified set 2 as being suspect. The process engineer was then able to begin troubleshooting the variables present within the FCC regenerator to identify what change initiated the maldistribution. Figures 10 to 12 provide an example of these cyclone delta temperature plots.

Finally, one negative aspect of the use of platinum-based CO promoters for the control of afterburn is their effectiveness at also increasing NO_x emissions. NO_x emissions will increase as the platinum content in the circulating inventory increases. Technology has been developed and commercialised that catalyses the reaction of carbon directly to carbon dioxide while substantially reducing the concentration of NO_x in the flue gas. Figure 13 demonstrates the value of state-of-the-art, low- NO_x CO promoter technologies.

References

- 1 Tenney E D, FCC cyclone troubleshooting, *Catalagram*, European Edition 1/93.
- 2 Fletcher R P, Stepwise method determines source of FCC catalyst losses, *Oil & Gas Journal*, 28 Aug 1995.
- 3 Oberlin J, personal conversations plus source of charts.
- 4 Wilson J W, FCC regenerator afterburn causes and cures, 2003 NPRA Annual Meeting, AM-03-44.

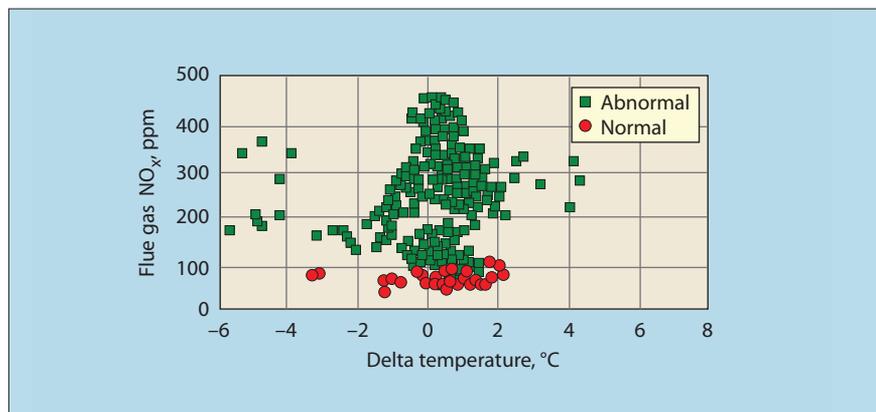


Figure 11 Cyclone delta temperatures 2 to 3

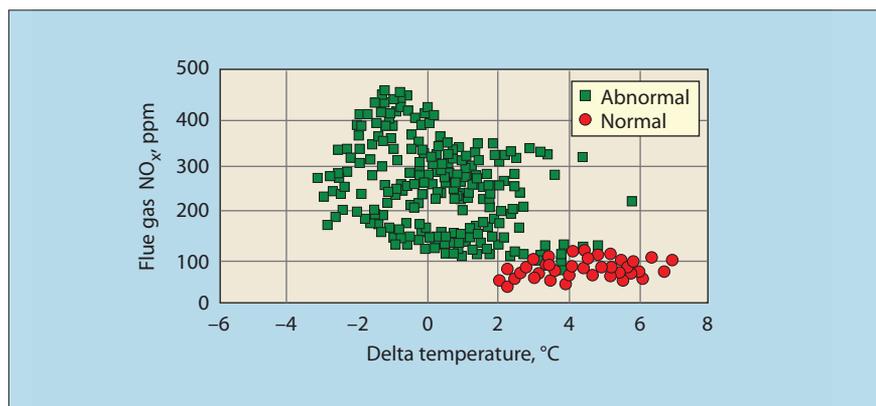


Figure 12 Cyclone delta temperatures 1 to 3

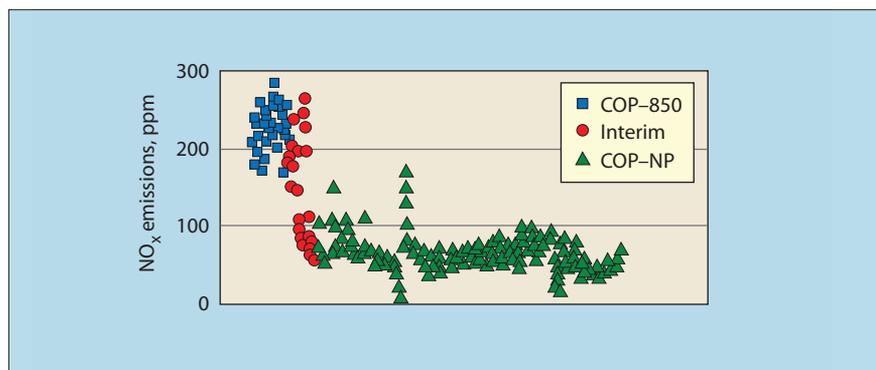


Figure 13 Non-platinum CO promoter performance

- 5 Powell J, et al, Advanced flue gas desulphurization technology, NPRA Annual Meeting, San Antonio, TX, Mar 1988.
- 6 Huling G P, et al, Feed sulphur distribution in FCC product, *Oil & Gas Journal*, 19 May 1975.
- 7 Vierheilg A, et al, The role of additives in reducing FCC emissions to meet legislation, 2003 NPRA Annual Meeting, AM-03-97.
- 8 Fletcher R P, et al, An alternative to FCC flue gas scrubbers, NPRA Annual Meeting, AM-09-38.
- 9 Rosser F S, et al, Integrated view to understanding the FCC NO_x puzzle, AIChE Annual Meeting, 2004.
- 10 Miller R B, et al, Solutions for reducing NO_x and particulate emissions from FCC regenerators, NPRA Annual Meeting, San Antonio, TX, Mar 2004.

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Debottlenecking hydrogen plant production capacity

Retrofitting a steam methane reformer with oxygen enhanced reforming can substantially boost hydrogen production capacity at low capital investment

GREGORY PANUCCIO, TROY RAYBOLD, JAMES MEAGHER, RAY DRNEVICH and VENKATARAYALOO JANARTHANAN
Praxair

Stricter environmental emissions regulations and increased processing of heavier and more sour crude oil fractions are causing hydrogen demand to grow across the refining industry. Costs associated with new hydrogen plant construction have been escalating, so refiners are increasingly looking for low capital cost retrofit solutions that can expand hydrogen production capacity at their existing facilities. This article discusses Praxair's oxygen enhanced reforming (OER) technology, which can expand hydrogen plant production capacity by enriching the steam methane reformer's (SMR) combustion air with oxygen. An OER retrofit can be installed with low capital investment, quick turnaround and minimal plant downtime.

Options for increasing hydrogen supply capacity

A refinery manager has many options to consider when operations are limited by hydrogen supply from an on-site SMR. For example, hydrogen could be purchased from a liquid hydrogen supplier or the hydrogen supply capacity could be increased by constructing an additional on-site SMR or by replacing the existing SMR with newer and larger equipment. These solutions can be costly, so it is often most economical to debottleneck the existing hydrogen plant. Approaches typically considered for debottlenecking include increasing the firing rate of the reformer, replacing existing tubes with ones with a larger diameter or improved metallurgy, adding a low temperature shift (LTS) reactor

Comparison of common SMR retrofitting methods with Praxair's OER retrofit ^{1,2}			
Method	H ₂ rate	Cap cost	Comments
Modify WGS reactor	+3–5%	Medium	Single HTS reactor changed to two-stage HTS, HTS+LTS, or MTS design
Upgrade reformer	+5–15%	Med-high	New catalyst; replace tubes with better metallurgy; modify pigtail/tunnel design; upgrade controls
Install pre-reformer	+8–10%	Medium	Requires significant changes to reformer convective section; considerable drop in steam export
Install OER	+10–15%	Low	Simple installation, equipment and controls; non-invasive
Install post-reformer	+20–30%	High	Large footprint; capital intensive

Table 1

downstream of the existing high temperature shift (HTS) unit, and adding a pre-reformer or a post-reformer.^{1,2} Characteristics of these retrofit solutions are summarised in Table 1.

The first concept usually evaluated is to increase the firing rate of the primary reformer by burning more fuel and increasing the heat available in the radiant section of the reformer furnace. As a result, more heat is transferred to the reforming reaction within the tubes and additional feed gas can be processed. A new induced draft fan may have to be purchased to accommodate the additional flow of flue gases, and a significant increase in flue gas stack temperature can result unless the convective section heat recovery system is modified. Furthermore, the resulting increase in reformer tube wall temperature will reduce the life of the tubes and significantly add to plant maintenance costs. Increased maintenance costs can be defrayed by upgrading the reformer with new tubes made of better metallurgy that is tolerant

to higher temperatures. Taking these steps can increase hydrogen production capacity by 5–15%, but will require significant capital investment and plant downtime.

The hydrogen production rate from an SMR can be increased by 3–5% by modifying the water-gas shift (WGS) reactor system in order to convert more of the residual carbon monoxide in the syngas into hydrogen and thereby improve hydrogen output without increasing the flow of process gas or flue gas, or raising the reformer firing rate. A two-stage HTS configuration can be installed with interstage cooling; the existing HTS reactor can be replaced with a medium temperature shift (MTS) reactor; or a LTS reactor can be installed immediately downstream of the HTS reactor. LTS units can be difficult to operate and additional make-up fuel is required to replace the lost PSA tailgas. Furthermore, modifying the WGS design will require a moderate capital investment and significant downtime for installation.

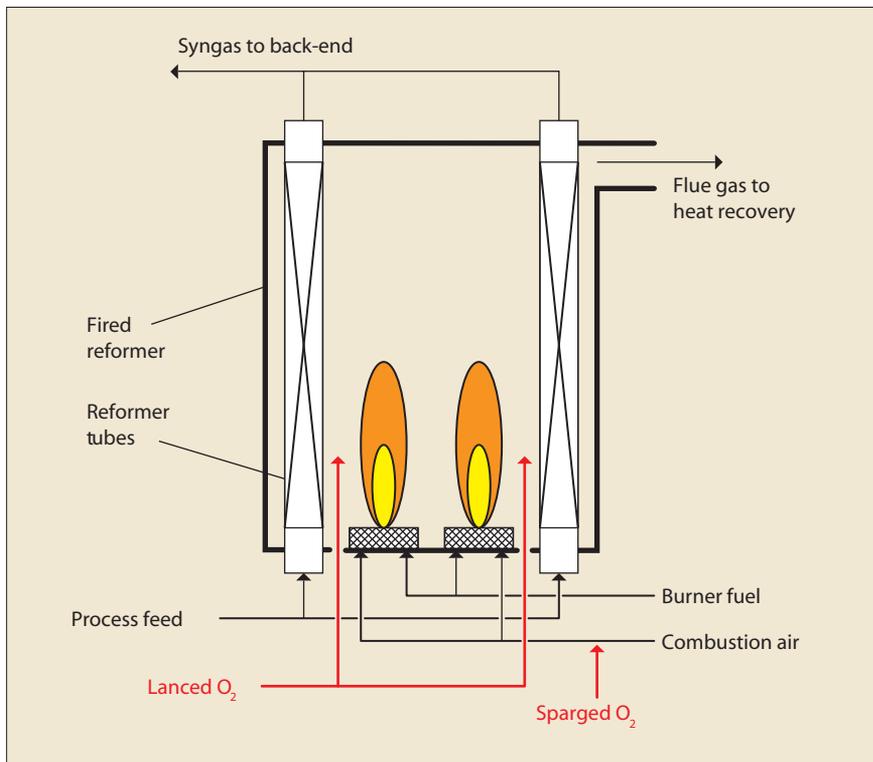


Figure 1 Simplified schematic of an SMR furnace radiant section with an OER retrofit. Combustion air is enriched by either premixing oxygen with air in the burner air feed duct (sparger) or directly injecting oxygen into the burner flame (lance). This illustration is for a bottom-fired upflow cylindrical reformer, but OER can be used in any reformer furnace design (eg, terrace wall, side-fired, top-fired)

Retrofitting an existing SMR with a pre-reformer can boost production capacity by 8–10%. The pre-reformer reactor utilises heat from the hot flue gas that was previously used for steam production to instead reform a portion of the hydrocarbon and steam feedstock prior to introducing it into the reformer tubes. A pre-reformer retrofit will require a moderate capital investment and significant downtime, as both a new catalytic reactor and a flue gas convective section reheat coil must be installed. Pre-reforming will also result in lower production rates of valuable steam by-product.

A post-reformer retrofit (for instance, a secondary autothermal reformer or product gas heated reformer) can increase hydrogen production by as much as 30% if the SMR is not bottlenecked by the downstream syngas processing equipment. With a post-reformer, a significant portion of the feed is being reformed outside of the primary reformer, so hydrogen production is increased without increasing the firing rate in the

reformer. However, installation of the retrofit will require significant investment of capital, plant downtime and plot space.

The characteristics of an OER retrofit are also shown in Table 1. With OER, oxygen is used to enrich combustion air in the reformer furnace, which can improve hydrogen production capacity by 10–15% without increasing maximum reformer tube wall temperatures, modifying the induced draft (ID) fan or convective section heat exchanger design, or appreciably affecting by-product steam production rates. Further-more, an OER retrofit can be installed with little plant downtime and capital investment.

Oxygen enhanced reforming

A simplified schematic of the radiant section of an SMR furnace that includes OER is shown in Figure 1. The combustion air is enriched with oxygen by one of two means: either oxygen is premixed with the air via sparger in the air feed ductwork or the oxygen is injected directly into the burner flame via a lance. For

either delivery method, OER increases oxygen concentration and decreases the concentration of inert nitrogen in the combustion air. Since the oxygen concentration is higher than in normal air, the furnace firing rate can be increased without raising the volumetric flow rate of flue gas that is processed in the convective section of the reformer. And because the flue gas flow rate is unchanged, no modifications to the flue gas heat exchange equipment or the ID fan are required to maintain the reformer's performance. A secondary effect of increasing the oxygen concentration in the combustion air is that the additional heat that is generated is released over the same distance or even over a shorter distance than with air. This results in an increase in the heat flux into the tubes near the inlet of the reformer where the tubes are coldest. Process gas flow to the reformer tubes is increased to utilise the additional heat that becomes available and to maintain the maximum tube wall temperature at the desired value. Typical reformer tube heat flux and maximum tube wall temperature profiles with and without OER are shown in Figure 2.

The extent to which oxygen enrichment of combustion air can increase hydrogen production capacity in an SMR depends on many factors, including whether other bottlenecks exist in the equipment upstream or downstream of the reformer furnace. Increasing the oxygen concentration in the combustion air to between 22% and 23% can typically result in a 10–15% increase in hydrogen production capacity. Field testing and simulation results show that between 14 and 18 tonnes of oxygen are typically required to produce 1 million standard cubic feet of incremental hydrogen from a bottlenecked SMR. One of the other advantages of OER technology is the operational flexibility inherent to the retrofit. If incremental hydrogen production is not required for a period of time, the flow of oxygen can be turned off and the SMR will operate as it did prior to the retrofit.

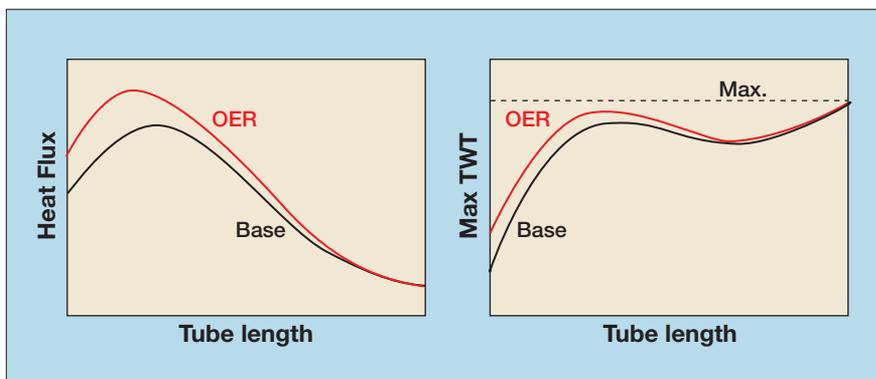


Figure 2 Approximate profiles of radiant zone reformer tube heat flux and maximum tube wall temperature (Max TWT) with and without OER. With OER, hydrogen production can be increased without increasing Max TWT and decreasing tube lifetimes

Oxygen delivery: lance vs sparger

Oxygen can be injected into the combustion air by either premixing it with the air via a sparger in the air feed ductwork, or by directly injecting oxygen into the combustion zone within the furnace via a lance. There are advantages and disadvantages to both methods.

A sparger is simpler to install than a lance. A sparger requires a single penetration into the air ductwork, whereas lancing requires multiple penetrations into the SMR furnace near the burners. It is more difficult to install all the piping required to feed oxygen to multiple lances than the single pipe required to feed oxygen to a sparger. Furthermore, the lance nozzle's orientation may have to be adjusted once it is installed, to optimise operating conditions.

One disadvantage of the sparger is that NO_x production within the furnace will increase. The adiabatic flame temperature within the burner rises as the concentration of oxygen in the air increases, which in turn rises the production of thermal NO_x . The extent to which NO_x emissions increase will depend on the design of the burners and the recirculation patterns within the furnace. Field testing has confirmed that lances can be installed so that NO_x emissions (on a lb/mmBtu fired basis) do not increase as oxygen is added to the SMR furnace. A lance installation can be NO_x emission neutral because the lances can be designed to stage the oxygen consumption within the combustion zone and increase recirculation of cooler flue gases within the furnace.

Incompatibility issues arising from air ductwork materials may limit the air enrichment level to 23.5% oxygen for the sparger case. According to Compressed Gas Association (CGA) standards, air that contains more than 23.5% oxygen is subject to different safety standards compared with normal air. For lances, the enrichment is not limited to the equivalent of 23.5% oxygen in air because oxygen is added to the furnace separately from the air supply. Therefore, there are no compatibility issues with air duct materials because the air duct is only exposed to ordinary air. Hydrogen production capacity can usually be increased to desired levels without raising the concentration of oxygen in air beyond 23.5%, so either a sparger or lances would be suitable.

The oxygen injection method is custom designed for each retrofit. Although a sparger is much easier to install than lances, NO_x emissions or oxygen enrichment limits may dictate that lances are the best option.

OER implementation

OER is a low-cost solution because, unlike most other reformer retrofitting solutions, there is no new expensive equipment to install (such as catalytic reactors) and no significant modifications to the existing reformer furnace are required as part of the installation. OER retrofit equipment is simple, safe and reliable. New equipment to be installed includes an oxygen supply system, an oxygen flow

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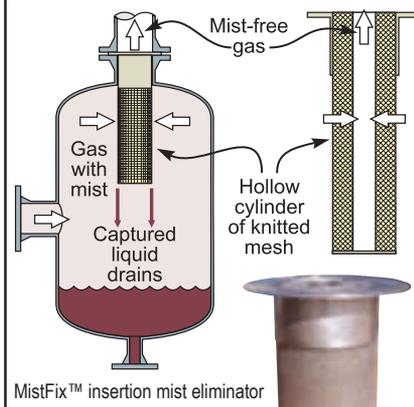
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Summary of OER field testing results

Operating condition	Base	OER	OER
O ₂ injection device	None	Lances	Sparger
H ₂ production rate	100%	111%	110%
O ₂ in air ¹	20.8%	21.9%	22.1%
O ₂ /H ₂ ² , tonne/mmscf	N/A	20	18
Natural gas feed + fuel	100%	109%	108%
ΔMaximum TWT ³ , °F	Base	-1	+2
ΔReformer outlet temp, °F	Base	+1	+4
PSA feed CH ₄	4.9%	5.6%	5.4%
NO _x emissions ² , lb/mmbtu	100%	88%	120%

¹ Equivalent O₂ enrichment level for OER cases with lances
² Corrected for equivalent flue gas flow rate and flue gas O₂ concentration ³ Calculated (not measured)

Table 2

control skid and an oxygen injection device (a sparger or lances). The oxygen flow control skid is sized for the required oxygen flow rate and may be integrated with an existing control system. The flow skid can be installed in almost any convenient location, it has a small footprint and can be built for compliance with most electrical classifications. The nature of the oxygen supply system will depend on the size of the application and location of the refinery. If the refinery does not currently have oxygen available in bulk on-site, a liquid storage tank or an oxygen generation plant may be installed. Depending on the refinery location, it may be possible to draw product from a nearby oxygen pipeline network. The supply system and flow control skid can typically be installed without shutting down hydrogen plant operations. Installation of the sparger or lances could take up to one day of SMR outage.

Safety and environmental reviews of the proposed design are conducted before construction begins. Standards and procedures are applied to ensure that the OER retrofit is installed and operated safely. New interlocks and alarms are programmed into the control system to monitor critical operating parameters. The flow of oxygen to the furnace will automatically be shut down if design limits are exceeded.

An OER retrofit with a liquid oxygen supply system can be installed and commissioned within six to 12 months from contract

completion. Detailed SMR design and operating data are required to engineer the retrofit, since each OER installation is custom configured for the SMR in which it will be used. To expedite the project schedule, an assessment of retrofit performance can be performed prior to contracting and the results can be reviewed along with the first draft of the contract.

OER demonstration

OER technology was field tested at one of Praxair's cylindrical, up-fired, up-flow SMRs that utilises a PSA for hydrogen purification. Both premixed (sparger) and lanced oxygen delivery methods were tested. Most of the OER equipment and controls were installed over a period of several weeks. Only a single day's outage was required to make the final oxygen connections: the sparger into the air duct and the lances into the furnace floor.

The results of testing are summarised in Table 2. The results for each operating condition are from experiments representative of those operating conditions. Each of the values reported is given from a heat and mass balance process simulation of the SMR that was reconciled to the experimental plant data. Some of the values that are reported in Table 2 (for instance, NO_x emissions in lb/mmbtu fired) are not directly measured, but are determined from a combination of direct measurements and calculations made in the plant process simulation. When practical, observed and calculated values were renormalised to equivalent operating

conditions within the corresponding process model for the executed experiment, as it was much easier to hold key operating parameters constant in a process simulation than it was during actual SMR operation. Note that these results may not reflect achievable outcomes for other OER installations. Actual OER operating conditions will vary, dependent on the SMR design and the way in which the SMR is operated, but the results shown here should be fairly representative.

Field testing showed that both the lance and sparger cases were able to increase SMR production capacity by approximately 10% by enriching the combustion air to an equivalent 22% oxygen. Enrichment oxygen was consumed at a rate of 18–20 tonnes per mmscf of incremental hydrogen produced; it was estimated prior to testing that it would take 19.4 tonnes of oxygen per mmscf hydrogen. The SMR hydrogen output increased by 10–11% in the OER tests, but the natural gas (NG) consumption rate only increased by 8–9%. The efficiency of using OER to produce incremental hydrogen is higher than the baseline hydrogen production efficiency. The calculated maximum tube wall temperature and observed reformer outlet syngas temperature changed only slightly from test to test. However, reformer methane slip increased because the residence time of the process gas within the reformer tubes decreased and the pressure in the reformer tubes rose as the hydrogen production rate increased.

At equivalent excess airflow, NO_x emissions actually decreased slightly when the lances were used to inject oxygen directly into the burner flame. As expected, NO_x emissions increased for the sparger experiment, where oxygen was premixed with the combustion air upstream of the burner inlet. Stable furnace operation was observed during the OER testing as well as during the transitions into and out of the OER operating condition. Transitions into and out of OER were accomplished within 20 to 60 minutes.

Cost of incremental hydrogen produced

Incremental hydrogen from an OER retrofit can be produced on a cost-competitive basis with the baseline hydrogen. For example, assume that economic parameters (cost of NG, power, make-up water and credit for steam) are such that the total variable cost of producing the baseline quantity of hydrogen from the plant above is \$2.25/mscf hydrogen (thousand standard cubic feet). Based on current market prices for liquid oxygen, the cost of producing incremental hydrogen from the OER retrofit during the field demonstration would be approximately \$2.90/mscf hydrogen. Furthermore, previous experience has shown that the total installed cost of an OER retrofit for the SMR operator can range from \$500 000 to \$1.5 million, depending on the size and design of the retrofit. The low capital investment along with the reasonable cost of incremental hydrogen produced leads to short payback times on the retrofit investment.

Conclusions

Demand for hydrogen within the refining industry will continue to grow as emissions limits for transportation fuels become more stringent and the nature of the crude oil supplies becomes more severe. The economics associated with building new on-site supply or replacing old hydrogen production equipment dictate that refiners focus on debottlenecking existing SMRs. Simple, reliable equipment, low capital investment, incremental hydrogen production capacity and competitive incremental hydrogen production costs make OER technology a good choice for retrofitting an existing bottlenecked SMR.

References

- 1 Drnevich R F, Fenner G W, Kobayashi H, Bool L E, 2006, Production Enhancement for a Reactor, US Patent 6 981 994.
- 2 Ratan S, Vales C F, 2002, Improve your hydrogen potential, *Hydrocarbon Processing*, 81, 3, 57–64.

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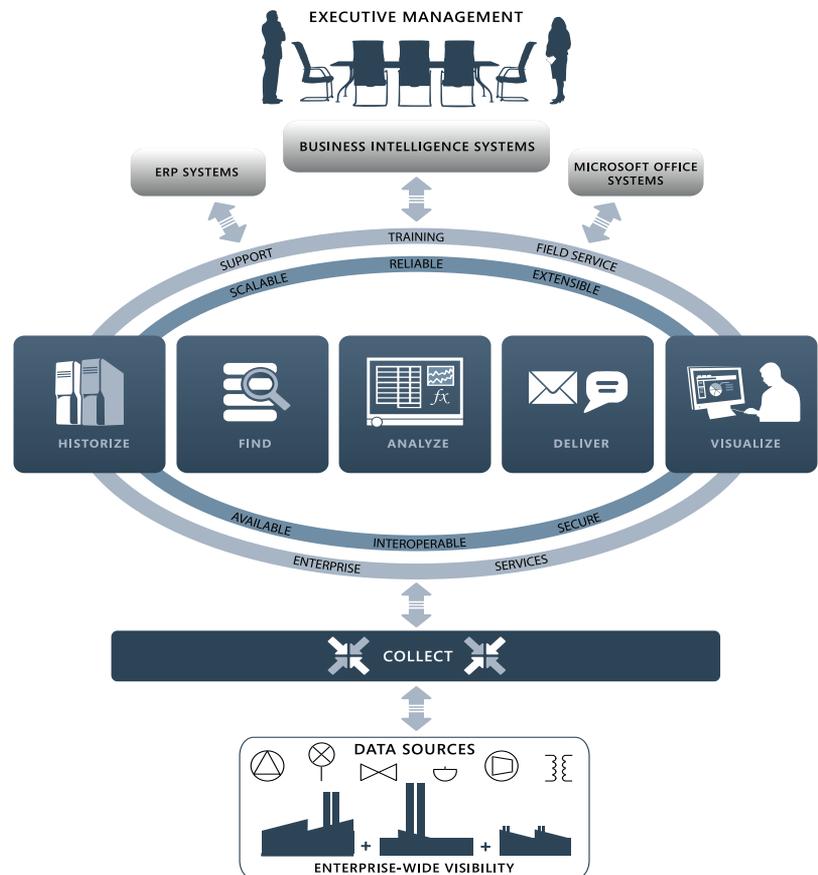
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Debottlenecking a refinery fuel gas absorber

A refinery fuel gas absorber was revamped to increase capacity while maintaining H₂S in the product well below specification

DARIUS REMESAT *Koch-Glitsch Canada*
MICHAEL BESHARA *Irving Oil Refining*

The purpose of a fuel gas absorber is to selectively remove components, primarily H₂S and to a lesser extent CO₂, using a solvent (amines) that absorbs these specific components. The product fuel gas can then be burned with reduced environmental impact.

One of the fuel gas absorbers at the Irving Oil Refinery in Saint John, New Brunswick, Canada, had a maximum sustainable rate of approximately 980 mscfh. Increasing the gas flow rate beyond this point had resulted in increased column differential pressure (an indicator of the onset of column flooding) and amine carryover (increasing operating cost and operational challenges). This column was limiting the ability to increase overall plant capacity, since Irving Oil Refining has strict operating requirements for environmental stewardship.

Design objectives and path

Irving Oil Refining wanted to process as much material through the column while maintaining product quality (H₂S in fuel gas not to exceed 50 ppm[v]) with minimum modifications to the plant during a planned shutdown in the autumn of 2009. It considered multiple options to debottleneck the column and settled on studying changes to column internals for increased throughput while maintaining or improving product quality. The newly designed high-performance trays would need to address the following criteria:

- The new design will take into consideration the foaming tendency of amine

- The expected rich amine loading shall not exceed API guidelines for carbon steel in specific amine service at the anticipated temperatures.

Based on past successes with high-capacity trays at the site and from other references,^{1,2,3} Irving Oil Refining commissioned Koch-Glitsch to:

- Model the operation of the fuel gas absorber (C14001) and validate current operation versus design, based on a comprehensive unit test

Numerous programs are available to assist in representing a column that uses amines to remove H₂S and CO₂ from fuel gas streams

run conducted in January 2009 and on existing internal drawings

- Recommend and model internal changes to increase column capacity while retaining 60% turndown capability (maximum throughput with given constraints is desired)

- Limit the extent of modifications to reusing the existing tray ring supports, including downcomers. Tray number and spacing to be retained, with 25 trays in total at 2ft spacing

- Retain current absorbent (amine at 25–30 wt%) and limit the flow and temperature that can be provided with existing equipment,

such as recirculation pumps and exchangers

- Revamp work to fit within the set turnaround schedule.

Methods and tools

The first and most important step in any revamp study is to generate an accurate characterisation of the process.³ The test run performed in January 2009 gathered data using calibrated instrumentation, creating a closed mass and energy balance. The next step is to take the data from the test run and to create a representative model of the plant that can be used to predict the future performance with the new tower internals.

Choice of modelling program

Numerous programs are available to assist in representing a column that uses amines to remove H₂S and CO₂ from fuel gas streams. From the authors' experience, rate-based models provide the best overall representation for new columns in this service, especially for packed columns. As an example, the rigorous, mass transfer rate approach used for all column calculations eliminates the need for empirical adjustments to simulate new applications correctly.

However, for column revamps, especially with trays, the use of an equilibrium-based model that has the necessary, proven adjustable parameters from operating experience is a suitable alternative to rate-based models, provided the necessary specific equipment characteristics of the high-capacity tray can be appropriately represented in the simulation model.

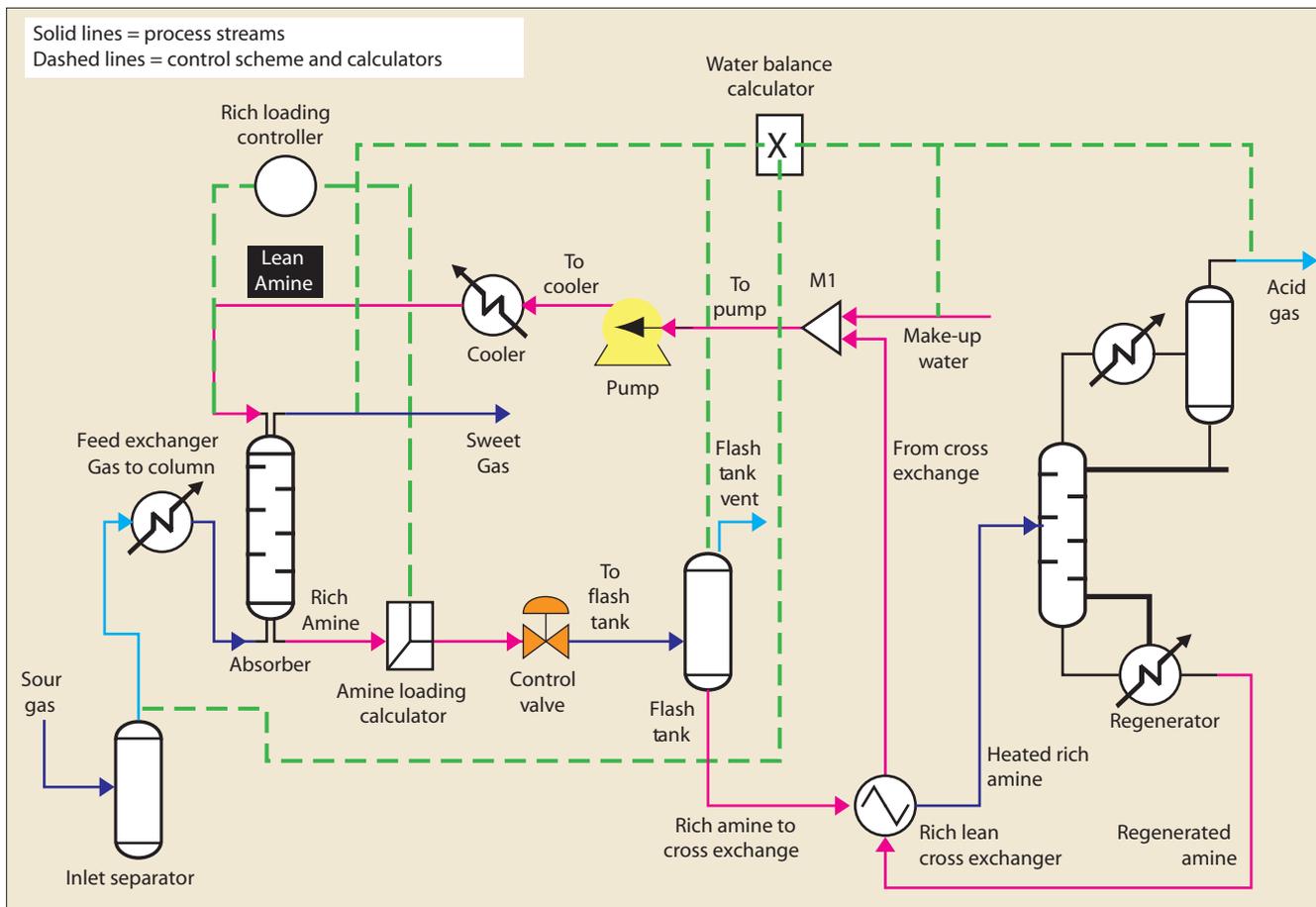


Figure 1 Primary simulation topology used

The simulation user needs to be sensitive to the fact that even the most sophisticated equilibrium-stage model uses only two of five elements employed in the rate-based model; namely, mass and energy balances around an entire ideal stage, plus thermodynamic-phase equilibrium. Programs that include reaction kinetics by empirical modelling via an adjustable parameter (H_2S and CO_2 tray efficiencies and/or liquid residence times) that forces the simulation to reproduce a conventionally operated column's treated gas composition can only be effective if comprehensive operating

experience has been gained and validated. In addition, the equilibrium-based program should have a reliable feature to include tray efficiencies to convert ideal stages into actual trays so that the tray characteristics can be represented pre- and post-revamp. VMGSim⁴ uses specific mass transfer multipliers that can be tuned to match tray and component efficiencies in the model. As a result, VMGSim has been used successfully to model existing plants and to accurately predict tray revamps in this service.

Of note, the solvents used in

amine absorbers are rarely pure solutions of water and amine. Contaminants entering with the feed gas or makeup water can change the chemistry of the solvent significantly. This can both worsen and, in some cases, enhance the absorption efficiency. To improve the accuracy of the simulation, the impact of heat-stable salts and other contaminants on the performance of the amine should be factored into the evaluation.

Process evaluation

A simulation using VMGSim (equilibrium-based model) with an appropriate amine thermodynamic package (validated with both Protreat and Ratefrac rate-based models) was developed based on plant data provided from January 2009. The fuel gas absorber was running at ~921 mscfh charge to the unit. Simulation cases were run at:

- 921 mscfh to match plant data
- 980 mscfh demonstrated sustainable limit of absorber column performance
- 1175 mscfh based on expected

VMGSim results — increased flow through plant

Case (constant lean amine, %)	H ₂ S, ppm	H ₂ S/amine, mol/mol	Acid gas/amine, mol/mol	Lean amine, BPD
920 mscfh - plant data	5	0.49	0.573	10 500
980 mscfh - 5 ppm H ₂ S	4.88	0.49	0.569	11 200
980 mscfh - max H ₂ S in amine	22	0.527	0.605	10 500
1175 mscfh - 5 ppm H ₂ S	5.27	0.49	0.562	13 600
1175 mscfh - max H ₂ S in amine	17.7	0.523	0.6	12 700

Table 1

acid gas loading limit of 0.6 (moles acid gas/moles amine).

Acid gas is primarily H₂S and CO₂. The hydraulic limit of a proposed tray change would be 1234 mscfh, which is 5% above the expected revamp design value of 1175 kscfh. The primary objective of the simulation work was to determine what maximum flow the absorber could handle within the existing 5ft (1.5m) shell diameter and the supporting equipment (coolers and pumps) while still meeting desired product specifications. The regenerator was included in the evaluation and simulation to provide a closer representation of the plant (see Figure 1) and to better extrapolate the performance of the unit at higher rates.

In addition to tray modifications, process modifications can be considered (see Figure 2) to increase further the capacity of the fuel gas absorber. An approximately 4% decrease in amine flows for the same outlet H₂S ppmv value can be realised by increasing the lean amine concentration from 23.5–25.5%.

Simulation results (possible increased charge rates)

Taking the base representative simulation (VMGSim) for the 921 mscfh plant data of January 2009, which was within 5% of the plant data, and adhering to the design criteria, the following cases were reviewed:

- Maintaining sweet gas H₂S at approximately 5 ppm(v)
- Minimising lean amine rate to a maximum of 0.6 (mol acid gas/mol amine) acid gas loading.

Feed rates of 980 and 1175 mscfh were used in the evaluation.

The reduced amine circulation was reviewed to determine how much more capacity the tower had by offloading liquid to allow more vapour while still meeting minimum product specifications. The cases used the identical thermodynamics and tuning developed to match the plant data.

The 980 mscfh simulation was developed to determine a baseline for the limit of the trays, because operational feedback indicated that amine carryover began to occur at

Tray hydraulics evaluation for 920 and 980 mscfh				
Trays	Top	Top	Btm	Btm
Description	920 mscfh	980 mscfh	920 mscfh	980 mscfh
System factor	0.83	0.83	0.83	0.83
Hydraulic data				
Jet flood, %	85	93	89	100
Downcomer flood, %	41	44	41	44
Downcomer backup, in liq	9.2	10.0	9.6	10.7
Total tray, ΔP * in liq	5.9	6.7	6.4	7.5
Total tray, ΔP * mm Hg	11.3	12.3	12.1	13.7

Table 2

this feed rate. The 1175 mscfh case simulation was developed to reflect the expected maximum feed rate that the absorber could handle hydraulically (after a revamp to

of 0.6 for the existing case and the expected revamp case.

Evaluation of existing internals

Using the simulation output results for the 920 and 980 mscfh cases, the existing tray internals were evaluated. A system factor (foaming) of 0.83 was used in calculating the tray performance. From past experience, a typical system factor of 0.73–0.85 for heavy foaming systems such as amine absorbers is applied. Using 0.83 for the study was well within what is expected for this service and provided a representative match to the plant performance. Table 2 shows the tray evaluation results for the 920 and 980 mscfh cases. It appears that the primary limit on the trays was the active area, with a jet flood of 100% for the 980 mscfh case. Such a high jet flood value matches with the observation of amine carryover due to high froth heights on the trays. From the plant data at 980 kscfh and the hydraulic evaluation, trays may not still be at incipient point of flood, yet operation and evaluation indicates that entrainment may be the primary issue and thus this phenomenon needs to be

The impact of heat-stable salts and other contaminants on the performance of the amine should be factored into the evaluation

higher capacity internals). For the 1175 mscfh case, the total acid gas load (mol acid gas/mol amine) of 0.6 was the limiting process variable when trying to keep amine flow to a minimum. Table 1 shows the simulation output results for the four cases evaluated. The sweet gas H₂S composition is below the specification of 50 ppm(v) for all cases. As noted previously, the limiting parameter was the acid gas loading

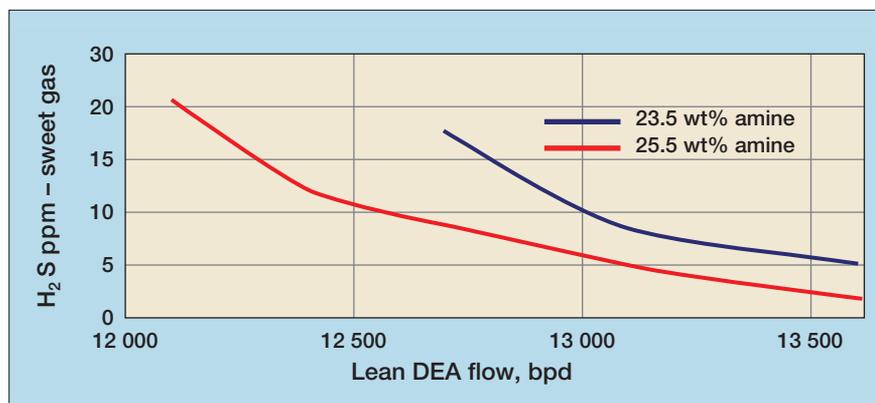


Figure 2 Amine concentration vs product H₂S purity @ 1175 mscfh plant feed rates

Debottleneck case (1175 and 1234 mscfh) high-performance tray hydraulic rating						
Trays	Top	Top	Mid	Mid	Btm	Btm
Description	1175 mscfh	1175+ 5%	1175 mscfh	1175+ 5%	1175 mscfh	1175+ 5%
System factor	0.83	0.83	0.83	0.75	0.83	0.83
Hydraulic data						
Jet flood, %	75	79	75	78	81	85
Downcomer flood, %	79	83	76	80	81	85
Downcomer backup, in liq	6.7	6.9	6.6	6.8	6.9	7.2
Total tray, ΔP^* in liq	3.5	3.6	3.4	3.6	3.6	3.8
Total tray, ΔP^* mm Hg	6.6	6.8	6.5	6.7	6.7	7.0

Table 3

considered during the revamp design.

Revamp considerations

By using a high-performance tray device, the increase in capacity over the 980 mscfh current maximum was determined to be 1175 mscfh (an increase of 20% over current maximum sustainable rates). The revamp product type chosen maintained the overhead H₂S to below 50 ppm (v) and the acid gas loading on the amine below an acceptable maximum (0.6 mol acid gas/1.0 mol rich amine).

Using the information from Tables 1 (to set the column process performance) and 2 (to set the internals hydraulics performance), a debottleneck evaluation resulted in the recommendation that the column could favourably (that is, maintain H₂S on specification) support a process gas feed rate of up to 1175 mscfh.

Superfrac trays with valve push

features and enhanced contact (for example, the Minivalve movable valve - MV-1) and enhanced downcomers were considered for this revamp. These trays, illustrated in

Superfrac trays reduce the dead space on trays to increase capacity yet still maintain the necessary contact time between the amine and gas

Figure 3, reduce the dead space on trays to increase capacity yet still maintain the necessary contact time (with bubble promoters and other features) between the amine and gas to ensure optimum absorption occurs.

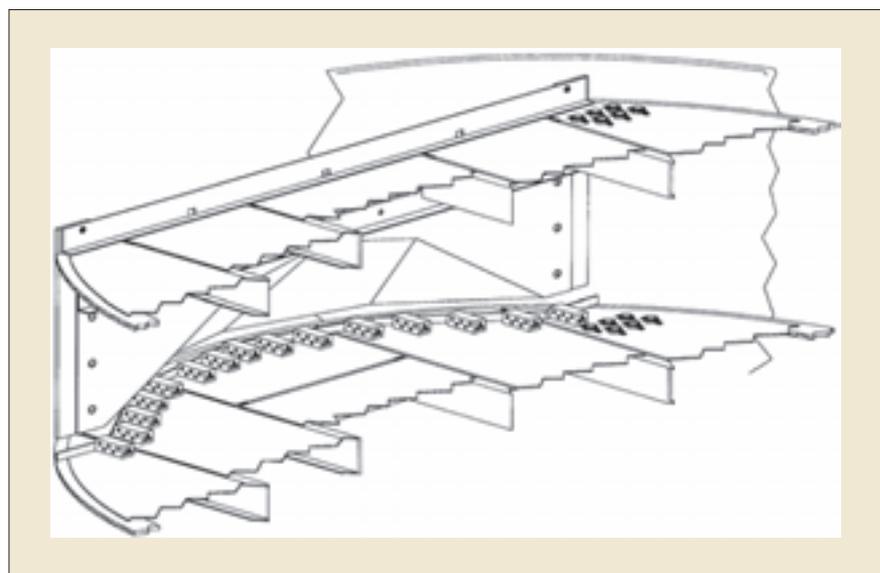


Figure 3 Sketch of Superfrac tray setup

Tray geometry		
	Existing	Proposed
Net top DC area, ft ²	3.5	2.7
Active area, ft ²	12.7	15.8
Valve type	Sieve	MV-1

Table 4

Tracerco scan results ⁵ summary		
Tray # (1-bottom)	Tray clear liquid height, in	Froth height, in
11-25	2-3	8-9
1-10	2	7-9

Table 5

Table 3 shows the results from the tower internals hydraulic analysis using high-performance trays. Using 85% for both downcomer and jet flood limits in a revamp, the hydraulic limit is at 5% above the 1175 mscfh revamp design feed rate, or 1234 mscfh. Factoring in inaccuracies in field measurements, fidelity of the simulation and hydraulic calculations, the tower limit should be expected to easily handle 1175 mscfh of feed and meet sweet gas product specification.

To meet the increased flows, the downcomer size and shape would need to be optimised, along with changes to the active area, with the use of Minivalve high-performance valves included in the active area changes. Table 4 shows a generic comparison of the characteristics of the existing and proposed tray designs.

The tray evaluation was based on reusing all existing tower attachments, with no welding required on the vessel shell or tower attachments to meet the shutdown schedule. Since the existing trays are one-pass, cross-flow trays, any modification to increase capacity preferably should be based on cross-flow trays so that the existing tower attachments can be reused. The foaming factor for the revamped tray evaluation will remain at 0.83 due to uncertainty in the level of foaming in the future. If the same conditions persist after the revamp, the Superfrac trays using movable valves could be evaluated



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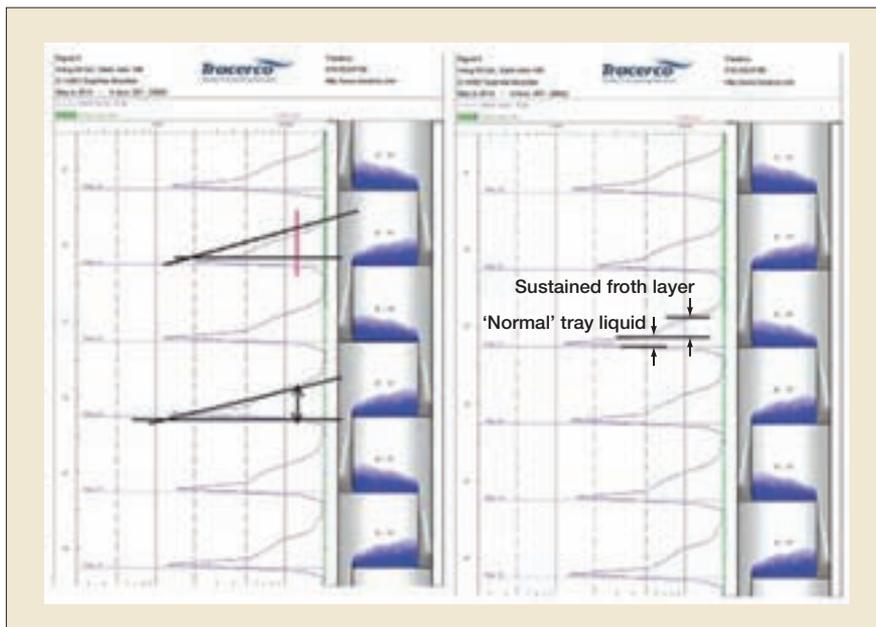


Figure 4 Excerpt of Tracerco's gamma scan of the tower

with a foaming factor of 0.9 (in other words, a further increase of capacity of around 8%, or up to 1460 mscfh feed rate). The features of the MV-1 valve's active area are configured in such a manner as to reduce the promotion of foam in this service. Since a definitive analysis of foaming tendencies was not performed, the foaming benefit was not included in the revamp's design expectations.

Post-revamp test results

A test run and tower scan were performed in May 2010, to validate the performance of the column post-revamp. Since startup post-revamp, the column feed had reached up to 1160 mscfh at 20 wppm H₂S with no operational

issues. Using a material balance with sulphur, the column flow rate for the test run was calculated/confirmed to be in the range 1150–1170 mscfh, with no indication of amine carryover and a sweet gas H₂S concentration of 18–20 wpmw. The VMGSim simulation was updated and corroborated the post-revamp plant data. The tower scan indicated that, at these rates, there was still ample room on the trays to handle flows of up to 1340 mscfh. Table 5 shows the calculated activity on the trays in the form of clear liquid height on the tray and froth height. Adding these two values together gives the total height occupied by the liquid (clear and aerated) on the tray deck, which is in the 9–12in (22.5–30cm) range

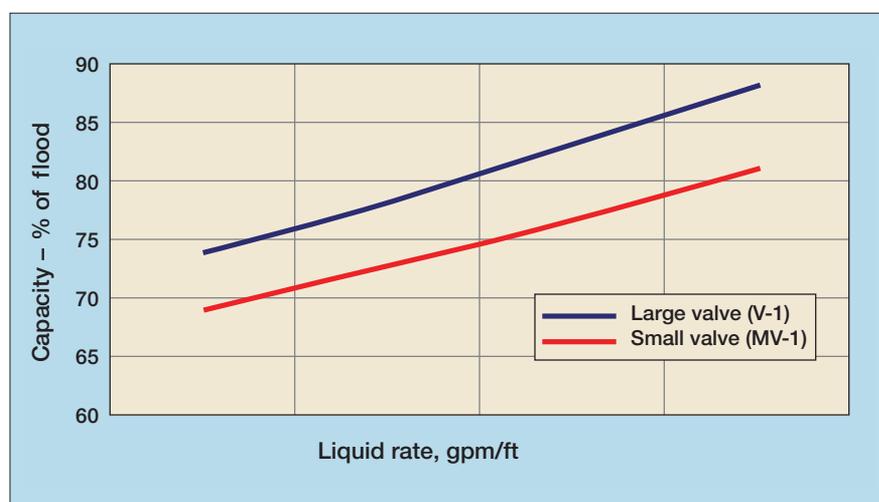


Figure 5 Capacity impact from different valve types

Source: KG-Tower rating software

across the tower. Using the same process data from the test run in the simulation to generate the internal loads, the KG-Design hydraulic rating program from Koch-Glitsch provided 70% jet flood results. Both values, the froth height and the jet flood, would tend to indicate that the trays still have room to process more material.

Figure 4 shows an excerpt from the gamma scan of the top section of trays, performed during the test run, capturing the level of activity on the high-performance trays and providing an indication of how much room is left hydraulically on the tray. Tracerco's mid-peak calculation is shown on the left, and the tray and froth height calculation is shown on the right of Figure 4. These tools help to convey how the high-performance tray functions at such high rates. Even factoring in the potential high foam generation, there appears to be approximately 45–50% disengaging space still available for further processing of gas above the test run rates.

The plant test run values, post-revamp simulation, tower scan and hydraulics evaluation appear to be in line with each other, giving similar results. Using the scan and plant data, and calculating the trays at 80% jet flood, the flow to the column can safely be 15% more than the test run, which is approximately 1340 mscfh.

The expected design flow rate post-revamp was set to 1234 mscfh at 20 wppm H₂S (85% jet and downcomer flood), with reasonable expectations of reaching up to 1351 mscfh at the amine carryover point. If foaming/froth is still proportional as the rates increase further, based on the test run evaluation, an upper rate of 1460 mscfh through the tower is possible.

Performance likely better than expected

A plausible reason for why the performance of the revamp trays is currently better than expected with the test run rates was that the full benefit of the Minivalve valve to mitigate foaming (over large sieve and large valve trays) in the column was not factored into the revamp

design. As was noted previously, a foaming factor of 0.83 was used for the conventional trays because it resulted in a good match of the plant data and pre-revamp gamma scan results. For the revamp study, the foaming factor was kept at 0.83, not accounting for the benefit of the Minivalve, which can reduce tendency to foam.

Considering the tower scan results post-revamp, a foaming factor of 0.9 could be used for the Superfrac trays with MV-1 movable valves. With the tower exhibiting an improvement over design expectations, the difference could be attributed to the reduced foam generated by the valve type or simply that the added capacity of the valve's active arrangement on the tray provides even more capacity in this service than is normally anticipated. For a set open area, when the hole/valve size decreases, the capacity of the tray increases (see Figure 5). The increase in capacity comes from a reduction in froth height. With reduced froth height, there is more disengaging room to deal with foam, and thus more capacity. This phenomenon, arising from the different valve size, helps to deal with and/or address foaming issues in the column and thus further increase capacity in the column.

Conclusions

The revamped fuel gas absorber has met and exceeded the design objectives to enable Irving Oil Refining to increase overall refinery performance while maintaining strict environmental objectives. The absorber has been able to operate consistently above pre-revamp rates, and with expected post-revamp rates at the same product quality levels as before the revamp. Collaboration between the operating company and the tower internal company enabled a low-cost and effective tower revamp.⁶

MINIVALVE, SUPERFRAC and KG-TOWER are marks of Koch-Glitsch LP.

References

1 Nieuwoudt I, Penciak J, Best of both, *Hydrocarbon Engineering*, July 2007, 85–89.

2 Nieuwoudt I, et al, Revamp & retune, *Hydrocarbon Engineering*, Jul 2009, 14, 7, 56–60.

3 Remesat D, Improving performance through low-cost modification of tower internals, *PTQ*, Q3 2010, 37–42.

4 VMGSim website, www.virtualmaterials.com

5 Mak R, Tracerco Internal report for Irving Oil Refining GP, Tru-Scan of the Sulphide Absorber, May 2010.

6 Remesat D, Inside-out design approach, *Hydrocarbon Processing*, August 2006.

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A step-by-step approach to managing emissions

A comprehensive review of combustion systems and controls underpins a strategy for emissions reduction

BARNEY RACINE and BRENDAN SHEEHAN *Honeywell Process Solutions*
WILLIAM DE LOS SANTOS *Callidus Technologies* JEFFREY RAFTER *Honeywell ECC Maxon*

Nitrogen oxides, or NO_x, result when the combustion of a mixture of air and fuel in a combustion engine or device produces temperatures high enough to drive endothermic reactions between atmospheric nitrogen and oxygen in the flame. In areas with large concentrations of industry or heavy traffic, the amount of NO_x in the atmosphere can reach significant levels and negatively affect the environment and human health.

In light of these environmental and health hazards, the hydrocarbon processing and petrochemical industries face ever-increasing environmental regulations. Over the past two decades, a variety of measures aimed at mitigating the negative impact of emissions such as NO_x have also emerged. These include a number of Federal NO_x emission reduction programmes and the Clean Air Act Amendments of 1990, which aim to address these health concerns and ultimately reduce common air pollutants, including overall NO_x.

To remain successful, companies must simultaneously meet the rising demands of these emissions regulations while keeping operating costs at a minimum. Striking this balance can be challenging. Various emissions requirements have made the need for timely and accurate emission data collection and reporting more important than ever. And companies must address ageing and inefficient technologies, which can become a drain on resources.

Still, it is possible for companies to address the balancing act of cutting emissions — specifically NO_x — across the board while

managing the cost of compliance. Choosing a comprehensive, single-source solution for reducing NO_x emissions can help lower the total cost of ownership and achieve emission reduction goals. The ideal solution combines new and innovative process technologies that optimise the combustion process — from start to finish — while providing online control and monitoring in real time.

When applied to a process heater

To remain successful, companies must meet the rising demands of emissions regulations while keeping operating costs at a minimum

during the production of ammonia, for example, these types of solutions can effectively reduce NO_x emissions while maintaining an optimal level of system performance. To successfully implement a NO_x emissions-limiting solution, you must examine all points in the ammonia production process, including:

- Fuel composition and control of the gas train
- Burner design and management
- Flue gas recirculation
- Air and temperature control.

Beyond these points in the production process, a comprehensive emissions-reduction strategy

can include post-combustion solutions, such as selective catalytic reduction (SCR), stack emissions monitoring, and regulatory compliance and reporting.

Fuel gas control

The first step in reducing any process heater's NO_x emissions begins by focusing on the combustion of fuel. As hydrocarbon fuels combust, the high-temperature reaction forms pollutants when the combustion conditions achieve peak flame temperatures or when fuel and air mixtures stray from optimal stoichiometry in any flame zone of a specific burner design. Although the flames are contained inside the heater, controlling the formation of pollutants, such as NO_x, carbon monoxide (CO) and sulphur oxide (SO_x), actually begins by looking outside the heater at the fuel gas control skid and air control devices.

Most installations consolidate fuel gas instruments and controls to a single location in a fuel gas skid or fuel train. The heater and skid can be fed with a variety of fuels, ranging from commercial-quality natural gas to refinery gas, to process off-gases. The mix of constituents in the fuel gas can dramatically affect heater emissions. Heavier hydrocarbons tend to be more difficult to mix with air and burn more slowly, and this decreased rate of reaction can increase the impact of a NO_x formation pathway called prompt NO_x. The presence of hydrogen in fuel gases can counter this by accelerating combustion rates and, with the proper burner design and air-to-fuel ratio control, can mitigate reactions that increase NO_x.

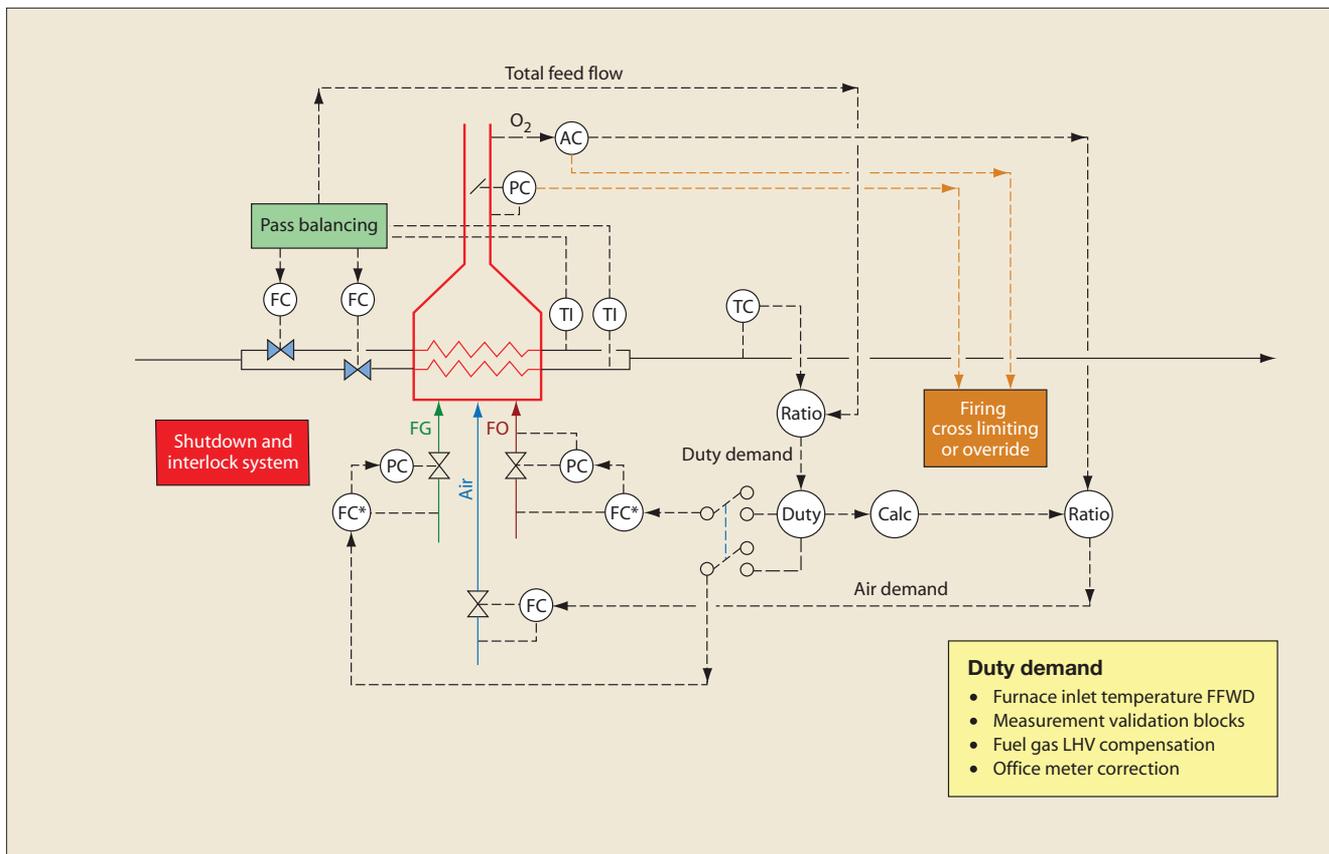


Figure 1 Fired heater advanced regulatory controls

To reduce emissions from hydrogen or heavier hydrocarbons, you must also understand how to detect the range of fuel composition changes. Active adjustment of fuel and air ratios can combat the formation of emissions by reducing variations in flame temperature and combustion conditions. A variety of methods can help measure fuel composition or quality, including complex gas calorimeters for widely varying fuels and tools to measure simple mixture density for slight changes in hydrocarbon content. Overall, diligent design and control system compensation can enable active air-to-fuel ratio adjustments to minimise peak flame temperatures and provide the correct feed to burners for pollutant control.

Once measured, changes in fuel gas can then lead to adjustments in air-to-fuel ratio with high-resolution control actuators in the fuel skid. Air registers located on the burner “windbox” assembly can enable combustion air control, while dampers in the stack or air piping can control available heater draught. By finely tuning air-to-fuel ratios, burners can operate at opti-

mal conditions and reduce peak flame temperatures and burning speeds.

The ideal actuators combine high-resolution control architecture capable of 0.1-degree positioning accuracy with intelligent positioning feedback. This enables the integration of high-performance valves with advanced control algorithms. You can then apply actuators to fuel control valves, pressure-reducing valves, stack dampers, air control valves and process control valves. Today, many available valves operate on a 24 VDC power supply and a digital communication network. This architecture provides the advantage of distributed intelligence for increased reliability and improved safety through error reporting.

In addition, high-accuracy fuel flow meters applied to fuel skids can provide real-time adjustment capabilities to air-to-fuel ratios or fuel blending control. Such meters offer precision flow reading and are corrosion resistant, making them ideal for most hazardous environments where NO_x emissions frequently originate.

Multiple air-to-fuel ratio control

Many industrial process heaters may use multiple fuels to provide heat to the furnace. Refineries, for example, often use a combination of fuel oil and fuel gas, and the pulp and paper industry often taps a combination of gas, coal or biomass such as wood chips to provide thermal energy.

In cases that involve multiple fuels, the control strategy may employ an overall duty control that takes contributions of duty from each fuel source. Ideally, each source has its own air flow so you can determine and provide individual air-to-fuel ratios. This is not always possible in practice, however. As a result, it becomes necessary to estimate the overall air-to-fuel ratio and the overall excess air required.

Typically, furnaces will operate with one fuel as a base load and another type of fuel that you can adjust to achieve the desired outlet temperature. In process heaters with both oil and gas fuel flows, the oil flow often serves as a base load but, as it is likely to be a heavy oil residue from the bottom of one of

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the cracking units, it can be very viscous and difficult to control.

Pressure control maintains the fuel gas flow into the process heater. As with the single-fuel conditions, measuring the lower heating value (LHV) or calorific value of each fuel is beneficial, as you can apply these values as a disturbance or feed-forward effect in the control strategy.

Figure 1 demonstrates this type of process heater control strategy from a functional perspective, which can help set the air-to-fuel ratio and control the process heater outlet temperature. In addition to implementing this in traditional regulatory control strategies, you can also implement this strategy using a multivariable controller.

In this specific strategy, the temperature of the process fluid leaving the heater is normally either the master controller or the primary controlled variable. This controlled variable cascades onto the duty controller, which in turn sets the required fuel flow of each of the fuels fired in the furnace.

Also in this control strategy are feed-forward or disturbance variables that include the LHV of each fuel and the pre-furnace inlet temperature of the process fluid, as well as the impact of changing the process feed flow rate into the furnace. In Figure 1, the desired air-to-fuel ratio sets the air flow into the furnace heater, while adjustments to the stack damper position maintain the draught pressure. In a natural draught furnace, directly adjusting the stack damper position helps control excess air.

Other control strategies commonly seen on process heaters include cross-limiting controls, which ensure that as the process outlet temperature calls for more duty, the air will be increased ahead of the fuel, thus making sure sufficient air is always present. Conversely, when the process calls for less duty, the dynamic response ensures, cutting the fuel flow before reducing the airflow.

Heater burners

After establishing optimal control and metering of fuel, the next step in reducing NO_x emissions involves

examining a process heater's burner design and integration with heater parameters. Every process heater design provides different conditions that can affect the operation of burners and variables, such as burner spacing, distance between burners and process tubes, radiant heat flux, downstream flow field development, and even the location of stacks and sight windows can all impact the formation of pollutants. Taken as a whole, successful emissions control requires designing burners for each specific application and modelling them into the process conditions for optimal emissions performance.

An ideal burner design emits the lowest level of NO_x emissions over a wide range of operating condi-

An ideal burner design enables lean pre-mixing and deep fuel staging, which can fit with both retrofits and new heaters and reformers

tions by featuring a smaller visible flame profile and burner size when compared to conventional raw gas burners. In addition, an ideal burner design also enables lean pre-mixing and deep fuel staging, which can fit with both retrofits and new heaters and reformers.

At work, the ideal lean pre-mix burner design achieves lower emissions through several steps. First, these types of burners maintain a larger portion of the available momentum of combustion reactants along the burner's centre line, which allows for better resistance to furnace currents. A steep velocity gradient along the burner axis also provides more rapid mixing for the remainder of the fuel and air. Increased fuel along the central axis of the burner removes duty from the secondary nozzle-mix fuel stage, which is influenced more easily by furnace currents.

An ideal lean pre-mix burner design also provides increased application flexibility to allow for high hydrogen content in the fuel gas, wide turndown ranges and stable flame shapes. When successfully implemented, such burners provide increased heat release without producing increased flame length, non-compliant NO_x levels or excessive draught requirements common with other burners.

Finally, in addition to adequate burner design, furnace modelling services and advanced computational fluid dynamic tools can help ensure successful integration with all furnace parameters. Such services and tools enable the modelling of burner installations in specific conditions so that burners can provide optimal emissions and heat transfer performance. Modelling combustion inside the furnace also prevents excessive heat flux to process tubes and reduces the likelihood of any unexpected furnace currents that cause flame drift.

Burner management systems

The next step in reducing a process heater's emissions is to examine the operation of a burner management system (BMS), which is responsible for the safe startup, operation and shutdown of a boiler or fired heater. The proper operation of a BMS is crucial to the safety of any industrial boiler. Such systems monitor and control igniters and main burners, among other tasks, and can apply flame scanners to detect and discriminate between the igniter and main flames. A BMS also employs safety shut-off valves and pressure, temperature, flow and valve position limit switches.

A BMS is required to meet the applicable codes for large furnaces and boilers (such as NFPA 8502 in the US). Historically, a BMS has been a standalone relay or PLC system integrated into a plant automation strategy using a communications gateway. Many control platform suppliers do not offer certified BMS solutions, while most BMS suppliers do not provide systems that exhibit the capabilities of a modern control system.

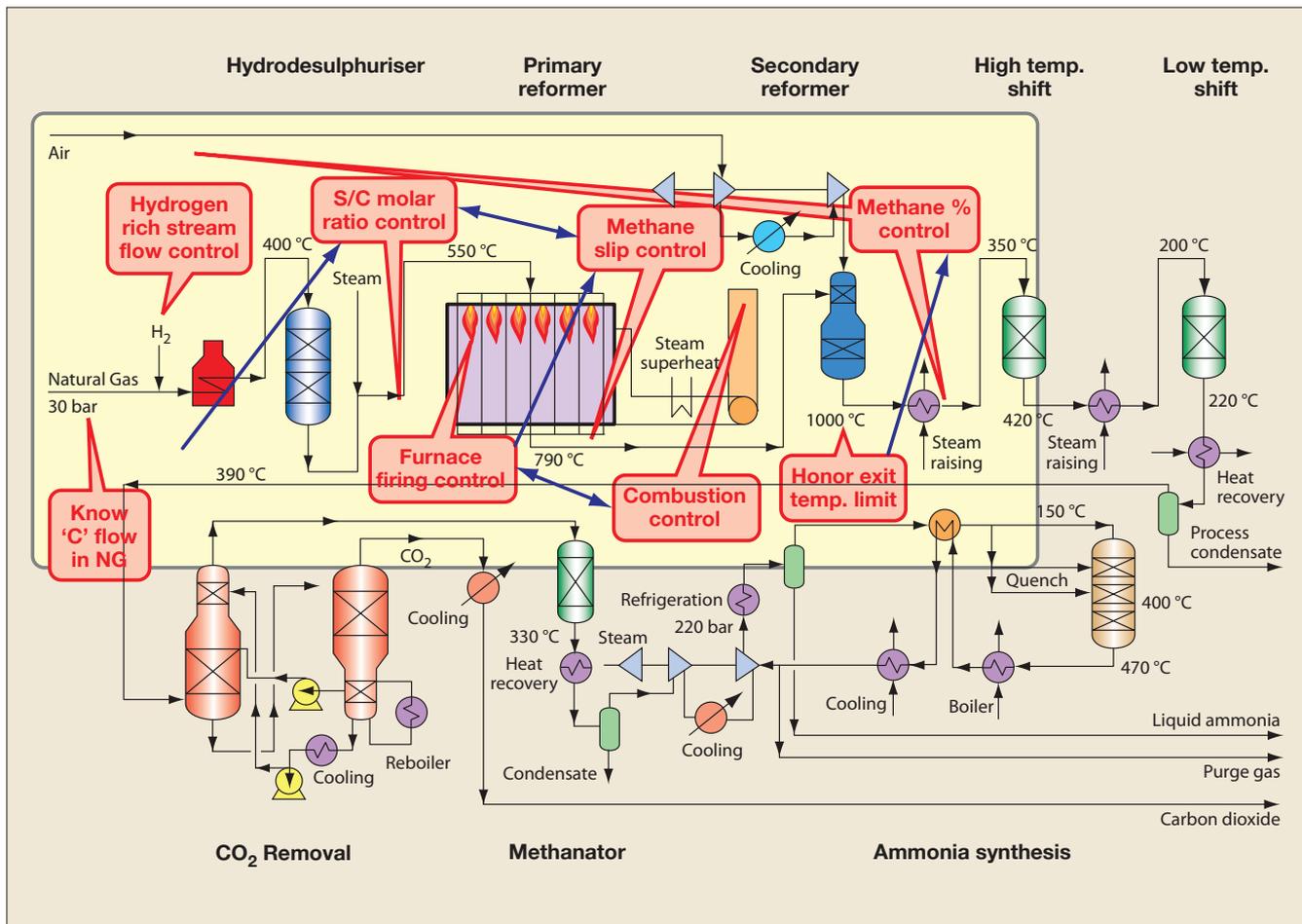


Figure 2 Example of a typical controller for an ammonia primary reformer

Engineering a gateway to support automation and a human-machine interface (HMI) with acceptable response times can also be challenging. However, solutions have recently been developed that provide tight integration of a BMS into a modern control platform. This can provide many functions, including: peer-to-peer communication between the BMS and the control system; use of a common fault-tolerant Ethernet; integration of alarms and events; and use of a common HMI. This type of solution addresses the challenge of providing improved integration while maintaining a dedicated separate safety management system and enables improved control.

This approach results in:

- **Improved profitability** Increased up-time due to the improved availability of the burner management system; elimination of unreliable relay-based systems that tend to fail, causing spurious trips of the process; and reduced maintenance costs inherent in self-checking and

diagnostic capabilities readily available to control systems, information systems and plant personnel

- **Improved safety** Risk reduction for people, plant and environment; and personnel safety improved by not having to access the old relay boxes, which may be located in hazardous and high-temperature areas of the boiler

- **Regulatory compliance** Meeting required regulatory codes associated with the safe operation of large process heaters and boilers.

Process heater outlet temperature control in a steam methane reformer

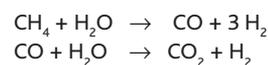
Improving a boiler or process heater's overall efficiency of service is an indirect yet effective way to reduce NO_x emissions. Improving process efficiency results in less fuel consumption per tonne of process feed, which means more product can be made in accordance with the same NO_x emissions limits that may exist at the plant.

For example, with a steam methane reformer, the first step of the

steam reforming process takes place in the primary reformer, where the hydrocarbon and steam mixture that is preheated to 505–520°C is passed downwards through vertical tubes containing a catalyst (typically nickel). The primary reformer is a fired heater, where the sensible heat and the heat of reaction are transferred by radiation from a number of wall burners to the catalyst tubes.

To ensure complete combustion of the fuel gas used in the burners, the burners operate with 10% excess air, which corresponds to 1.8% oxygen in the flue gas. The hydrocarbon in the gas, which leaves the primary reformer, then converts to methane. The exit temperature of the primary reformer is about 800°C, which is also the inlet temperature to the second step of reforming.

The following primary reactions occur in this process:



To achieve the right level of conversion of feed gas to CO₂ and H₂ (syngas), it is necessary to tightly control the ratio of natural gas and steam added to the reformer. Usually producers aim to control steam to carbon content in the natural gas feed. You must not operate the reformer with a low steam-to-carbon ratio, as this could thermodynamically lead to carbon formation, especially inside catalyst particles. It is also important to avoid poisoning the catalyst with sulphur compounds, as this can also promote carbon formation. At the same time, a steam-to-carbon ratio that is too high affects the activity of the catalyst and also means wasted energy and fuel.

To tackle these challenges, the following control strategies can be implemented with a multivariable predictive controller application:

- Minimise steam consumption by optimising the steam-to-carbon ratio in the primary reformer in tandem with the overall carbon number predicted from gas composition or molecular weight or specific gravity or manually entered lab analysis
- Reformer exit temperature control along with furnace firing and combustion control in tandem with the off-gas flow from the purge gas recovery unit
- Minimise CH₄ slip from secondary reformer subject to maximum exit temperature constraint. This ensures that the primary reformer is running at a reasonable efficiency. Methane slip is the amount of unreacted methane that passes through the reformers and typically occurs at a level below 1%.

Although controllers do not directly maintain NO_x emissions, they can be included as a constraint to prevent violating overall NO_x emissions restrictions. This may limit the amount of feed processed.

Overfiring air for staged combustion

In some large industrial boilers or waste gas incinerators, NO_x emissions can become a limiting factor to operation. In these instances, you can lower the peak combustion temperature by performing a staged combustion. The peak temperature

of combustion significantly affects the amount of NO_x a boiler or heater produces. Higher temperatures lead to more oxidised nitrogen.

Consider the following staged scenario: primary air — typically 70–90% of the total air required — is fed in a sub-stoichiometric ratio with the fuel, which results in a lower peak temperature. As a result, the combustion products contain more nitrogen than NO_x. This process will also result in incomplete combustion and high levels of CO. The second-stage area, just above the primary combustion zone, completes conversion of the CO in the flue through the addition of more secondary air into the boiler by overfiring the air ports.

This converts all of the CO to CO₂, and lowering the overall peak temperature results in the production of lower amounts of NO_x for the equivalent amount of energy supplied or steam generated in the boiler. The relatively low temperature in the secondary stage helps prevent the formation of NO_x, and the location of the secondary air ports, along with the mixing of overfire air, are all critical to maintaining efficient combustion.

This modified process allows for more degrees of freedom within the overall control strategy because it adds two areas of air to the process. Control of this process is enhanced when measurements of flue gas conditions exist in more than one area.

The control strategy is well suited to multivariable controllers, but it should also include CO levels as a constraint. Poor operation of staged combustion will result in significant levels of CO in the flue gas. Thus, the CO constraint must be handled as a non-linear input, as it will move very quickly to high levels if the air-to-fuel ratio becomes sub-stoichiometric, or if channelling in the heater occurs. This can result from poorly adjusted burners, which can lead to sub-stoichiometric conditions in a boiler, even if the stack analyser indicates an overall excess of air.

Flue gas recirculation

Flue gas recirculation (FGR) involves recycling and redirecting 15–30% of the flue gas to the burners, which dilutes combustion gases. Doing so also reduces the peak flame temperature.

FGR is not universally applicable. Since FGR limits thermal NO_x formation but has little effect on fuel NO, it is more effective in natural gas-fired heaters than in oil-fired heaters. Only mechanical draught heaters with burners that can accommodate increased gas flows are amenable to this technique. However, conversion of natural draught heaters to mechanical draught operations as part of a FGR retrofit is possible. Required FGR retrofit components include ductwork, recirculation fans and controls to vary damper settings on variable-load heaters. The difficulty of retrofit in crowded plants may be greater.

For FGR to be successful, process needs must be compatible with the lower flame temperatures it generates. Achievable emissions reductions are a function of the amount of flue gas recirculated, and they are thus limited by efficiency losses and flame instability at higher recirculation rates. Limited performance data and experience of industrial boilers suggest that reductions of 50–60% may be expected with natural gas-fired heaters, and somewhat less with oil-fired heaters.

Selective catalytic reduction

For very large energy-consuming processes, the total tonnage of NO_x emissions may require a more aggressive treatment than process, burner and control optimisation can offer. In these circumstances, post-treatment devices such as selective catalytic reduction (SCR) units can provide immediate solutions. In a SCR unit, pollutants undergo further reaction in the presence of a catalyst to greatly reduce the concentration of pollutants in the stack gases. Each unit is custom-designed for handling specific stack inputs, including pollutants, concentrations and flow rates for optimal conversion to harmless stack gases.

SCR systems have mostly been limited to large industrial installations. Over time, however, the technology and cost of SCR systems have evolved, making the device applicable to a wider range of processes and devices. As emissions regulations become more aggressive, a SCR system provides a positive assurance of compliance.

SCR systems are applicable to heaters that have both a flue gas temperature appropriate for the catalytic reduction reaction, as well as enough space for a catalyst bed to provide sufficient residence time for the reaction to occur. Specifically, several different available catalyst formulations make the temperature window fairly wide, ranging from approximately 250°C to more than 550°C. The installation of a SCR system on natural draught heaters requires conversion to mechanical draught in order to overcome the pressure drop that occurs across the catalyst. Finally, sufficient space must be available for ammonia storage.

The control system modifies the amount of ammonia that is added to the SCR system and used to convert NO_x to N_2 — all while maintaining NO_x emissions levels below their limits. The flow of flue gas acts as a disturbance variable within the control application.

Excess air control

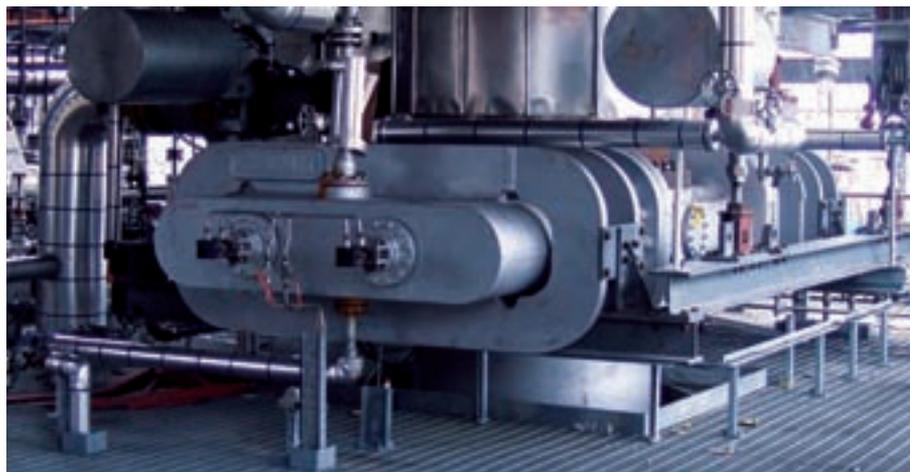
Maintaining low excess air minimises the air level above what is needed for complete combustion and lowers operating conditions. As a result, it also limits thermal and fuel NO_x formation and is a critical component in reducing overall emissions from process heaters.

Today, many process heater operators already minimise excess air levels in order to increase heater efficiency and decrease fuel requirements. You can reduce excess air levels on all process heaters, but this approach is most effective on mechanical draught heaters. Achieving better control of air flow — coupled with the higher pressure drop across the burners caused by the higher air flow — results in improved air-to-fuel mixing and

Over time, the technology and cost of SCR systems have evolved, making the device applicable to a wider range of processes and devices

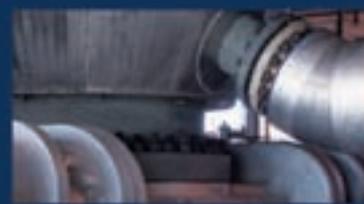
allows for greater reductions in excess oxygen concentrations before they reach levels that negatively affect flame stability.

Lowering excess air levels normally requires minimal capital investment, although retrofit controls may be necessary for some older heaters. Achieving emissions reductions using low excess air depends on the initial excess air level, the fuel used and other heater-specific factors, with a



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probable reduction range of 5–20%.

Also, very low excess air levels can lead to flame instability, as well as the formation of soot and increased emissions. A reducing atmosphere in the heater may also result in corrosion.

In addition to excess air control, a typical control strategy may include minimising pressure in the firebox to prevent over-pressuring. The same strategy may also monitor CO levels to ensure there are no pockets of incomplete combustion control in the heater.

A typical solution involves the use of a multivariable controller that aims to control 2–3% of O₂ in the flue gas by primarily manipulating the air damper in a natural draught furnace. In a forced draught furnace, the controller can be more precise by directly controlling the air-to-fuel ratio, which in turn adjusts the air flow to the heater.

Operating constraints are included in the multivariable controller and may act either to prevent further movement in an undesirable direction, or to directly adjust the operation to alleviate the constraint.

Examples of common operating constraints found in the multivariable control strategy include stack temperature, which impacts heater efficiency, as well as CO levels, draught pressure and any valve or positioner limits.

Monitoring

For an ammonia plant, emissions monitoring is typically required on steam methane reformer units. Natural gas and steam react in the steam reformer to produce hydrogen and CO₂. The reformer also introduces air to supply nitrogen. The process train then converts CO to CO₂ and removes the CO₂ while converting any remaining CO and CO₂ to methane. This yields a gas stream that is three parts hydrogen and one part nitrogen. The reformer then synthesises the process gas under high pressure to form ammonia.

The remaining process emissions from the reformer typically move to a stack, where the outflow is continuously monitored by gas

analysers for NO_x and CO, by-products of the process. Oxygen is also typically monitored to adjust emissions concentrations to nominal conditions. In cases where emissions are combined from multiple emission sources, stack flow monitors can help measure overall stack volume. To satisfy other criteria air pollutant requirements, periodic stack test measurements are often required for particulate matter, including SO₂, NO_x, CO and/or visible emissions.

A monitoring system may include some or all of the following components:

- Sample probe
- Heated sample line (to prevent condensation)
- Sample transport lines

Proper quality assurance is critical to accurately calculating emission rates, which are typically capped at hourly concentration levels or mass emissions limits

- Calibration assembly (including calibration gas bottles)
- Moisture removal system
- Particulate filter
- Sample pump
- Sample flow rate control
- Gas analysers.

The probe and sample line are typically heated, and the sample transport lines are made of non-reactive tubing such as Teflon, stainless steel or glass to transport the sample from the moisture removal system to the sample pump, sample flow rate control and electrochemical cells. The moisture removal system is typically a chilled condenser or similar device that removes the moisture while maintaining minimal contact between the condensate and the gas sample.

The gas analysers can be assembled in a rack and installed

within a cabinet, or they can be enclosed in an air-conditioned weatherproof enclosure known as a shelter, depending on available space and environmental concerns. While calibration of the gas analysers can be initiated manually by operators or technicians, most calibration control is done through an automated, optimised sequence that is initiated through either a distributed control system or programmable controller designed to minimise the downtime associated with the calibration checks.

These calibration checks, which are essentially real-time data and quality assurance checks, are typically analysed and collected by a dedicated data collection system that is designed to analyse and store emissions data and related quality assurance data. Proper quality assurance is critical to accurately calculating emission rates, which are typically capped at hourly concentration levels or mass emissions limits.

For most situations, implementing an ammonia plant emissions monitoring system that includes built-in calibration control can minimise costs by eliminating the need for a separate programmable controller. This is ideal for situations where the shelter or cabinet is isolated from the primary control network.

Compliance reporting

In the US, requirements for emissions data collection, record keeping and reporting for ammonia plants are typically detailed in Title V operating permits issued by individual states. These permits specify the monitoring frequency requirements around quality assurance and testing, as well as applicable state and federal regulations.

An integrated data collection and reporting system can help meet such requirements. The actual configuration of the data collection system depends on the permit specifics, but there are some common themes that exist for specific emission units. For a steam methane reformer unit, if a stack monitor is not installed on the stack to measure the overall volume of

gas out of the stack, the individual fuels consumed in the reforming process (or other combustion processes) must be monitored.

All fuels that contribute to stack emissions must also be monitored, including combustion fuels if combustion sources such as auxiliary boilers or heaters are routed to the stack. These fuel measurements are used to derive the overall stack flow rate, which is in turn used to calculate mass emissions of NO_x and CO in pounds per hour.

In many cases, the fuel measurements, if properly corrected for temperature and pressure, can also be used for calculation of the greenhouse gas emissions required under 40 CFR Part 98 based on emission factors, heating values or fuel sampling. If the combustion sources, such as auxiliary boilers, are large enough, stack flow monitors along with CO_2 and moisture analysers may be required on the stack for proper reporting of greenhouse gas emissions.

Combining combustion sources with reformer emissions can also require the NO_x and CO monitors provided to be capable of multiple ranges, depending on which units are operating. Range filters and appropriate scaling must be applied at the initial point of data collection, and daily calibration data from each range must be captured as well, in aggregate 15-minute or hourly segment averages. These averages are then converted to mass emission values using the stack flow calculation or measurement. Emission limits can also be placed on annual or rolling 365-day averages, which are put in place to effectively limit overall fuel consumption or unit operating times.

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Revamping centrifugal compressors at an ethylene plant

Technical improvements and increased throughput justify the revamp of a Russian ethylene plant's compressors

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Many applications of centrifugal compressors may involve process conditions that are quite different from those for which the machine was originally designed. In some cases, these off-design conditions can be highly inefficient and increase overall energy consumption. In many instances, however, new internals — specifically designed for the revised operating conditions — can be installed in an existing compressor casing, thus achieving a significant improvement in the overall performance of the compressor. Most of the time, the reason for revamping is dictated by the need to increase plant production. This article describes the benefits of revamping existing equipment at the largest ethylene plant in Russia, illustrated by a recent project.

In 1995, Thomassen came into contact with the largest ethylene plant in the former Soviet Union. The ethylene plant is part of a conglomerate of petrochemical plants that also produce butadiene, ethylene oxide, butyl rubber, polystyrene and linear alpha olefins. In fact, the ethylene plant is a supplier of raw material to many of the other plants within the conglomerate.

The ethylene plant was put into operation in 1976. The plant underwent its first uprating in 1992 in a project under the control of contractor Toyo Engineering. At that time, the compressors were revamped by

Ebara and new Mitsubishi Heavy Industries (MHI) steam turbines were installed.

Based on site inspections and discussions with maintenance and operational personnel, it became clear that the compressors experienced mechanical and operational problems. In 2000, Thomassen took up the opportunity to carry out modernisation of the existing compressors to try to solve these problems. This modernisation process has involved several years' work and is, in fact, a continuing project.

From 2001, technical and economical feasibility studies investigating various possibilities for capacity expansion were initiated by the ethylene plant's management and ABB Lummus Global in Germany (see Table 1). These feasibility studies (termed "basic engineering studies") were based on scenarios involving both the installation of new equipment as well as the revamping of equipment.

In December 2002, Thomassen was invited to quote for the revamp of the charge gas, propylene refrigerant and ethylene refrigerant steam turbines and compressors at the ethylene plant.

Ethylene plant production capacity increase	
Plant production capacity, MTA	Increase
Original plant output (1976) 350 000	} +29%
1st uprate plant output (1992) 450 000	
Uprated plant output (2007) 600 000	} +33%

Table 1

Modernisation

This ethylene plant is operated according to established practice in Russia. Every year, there is a complete plant shutdown in line with current state legislation, while maintenance methods tend to be rather conservative.

The modernisations are basically aimed at solving problems that date back to the previous uprate in 1992. In other words, they are troubleshooting activities implemented via root cause analyses. The modernisations are improvements that contribute greatly to increasing the reliability of the equipment and to establishing its more efficient operation. Examples of such improvements include:

- Greatly reducing seal oil leakage by replacing carbon seals with new mechanical oil seals (see Figure 1)
- Replacing conventional buffer gas labyrinths by installing positive seal barriers to prevent oil migration into the process and to eliminate pollution of the mechanical oil seal by process gas (see Figure 2)
- Supplying a new buffer gas system including stainless steel piping, valves, duplex filter sets, and so on
- Replacing outdated radial bearings with keys using customised spherical bearings (five pads - load between pads) in order to obtain stiffness and damping values for optimum rotor stability
- Installation of abradable inter-stage seals to an improved design (impeller eye and shaft seals) to reduce secondary leakage flows
- Raising customer awareness by emphasising the importance of correct internal alignment of end



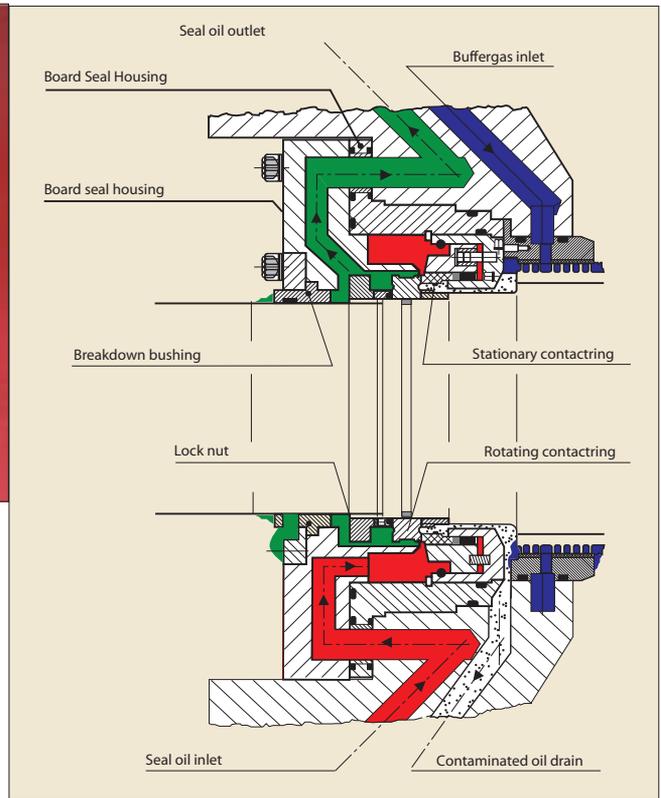
Figure 1 New design of mechanical oil seal

walls to casings, bearing retainers to end walls and diaphragms to bearing retainers. Thrust bearing load, inter-stage seal losses, balance piston leakage and diaphragm split line leakages are all greatly influenced by the quality of internal alignment

- Improving the existing combined oil system for control-seal-lubrication oil supply to compressors and steam turbine by replacing outdated critical components in order to obtain a better-responding and therefore more reliable oil system
- Generally, supplying spare parts such as dry gas seals, compressor rotors, steam turbine parts and oil pump parts in order to enable control of overall quality and design integrity, and thus guaranteeing maximum reliability and availability.

Immediately after the first uprate in 1992, the compressors experienced high rates of seal oil leakage. As a consequence, the compressors had to be stopped three to five times a year, which meant significant loss of production.

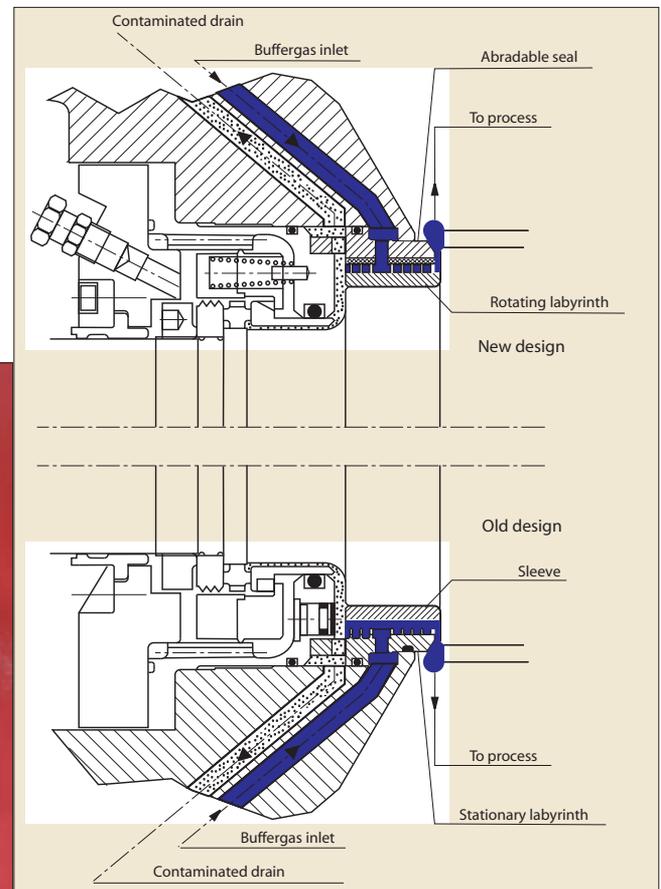
The situation needed urgent attention. However, once the new mechanical oil seals were installed, the excessive oil leakage disappeared. Note that only a minor rotor modification was required for the



compressors to enable the installation of the oil seals. The compressors operated in uninterrupted fashion between the yearly overhauls, meaning continuation of production and savings in investment and maintenance costs.



Figure 2 Positive seal barrier



Revamp project

The results of the basic engineering studies of ABB Lummus Global were presented to the ethylene plant's management in 2002. The next step — a detailed study — was never started. The customer decided to perform the complete ethylene plant revamp through a process of step-by-step debottlenecking using the results of the basic engineering study.

Thomassen was invited to quote for the revamp project involving the capacity increase for charge gas, propylene refrigerant and ethylene refrigerant compressor trains. Thomassen worked with MHI for all steam turbine-related issues. The quotation phase lasted from 2003 until 2004.

The capacity expansion project involved revamping of the equipment, and it also included arranging for a "permit to operate" issued by Rostechnadzor (Russian Federal Mining and Industrial Inspectorate). Revamping of the existing equipment was beneficial because the casings, process piping and foundations were all retained. In addition, the revised operating conditions could be met within the casing constraints in combination with applying state-of-the-art technology for reliable and efficient operation.

The intended capacity of the ethylene plant after revamp was 600 000 t/y, a figure that is based on the comparative assessment of several options within the technical and economical feasibility studies.

The following main components were replaced in order to meet the revised operating conditions:

Charge gas train E-GB-201

- Steam turbine rotor (ten stages)
- Steam turbine stator parts (ten stages)
- Flexible element couplings between steam turbine and compressors.

The steam turbine (an extracting/condensing type) was revamped in response to the increase in shaft power. In addition, during the quotation phase, it became apparent that the customer wanted to limit the condensing steam flow in order to reuse the vacuum surface



Figure 3 Steam turbine rotor GT-201, 10 stages

condenser and condensate pump system.

The existing internals of all three charge gas compressor bodies (five sections) could be reused in combination with the revised process conditions (see Table 2). Due to an increase in suction pressure of section one, the volume flow and head requirement for each compressor section could be handled by a slight speed increase only (± 100 rpm). The combined seal, lubrication and control oil system did not require modification. The revamped

equipment was put into operation in September 2006 (see Figure 3).

Propylene refrigerant train E-GB-501

- Footprint steam turbine (12 stages)
- Compressor rotor (four sections, five stages)
- Compressor stator parts
- Flexible element coupling between steam turbine and compressor.

In this case, the shaft power increase was too high (+38%) for revamping the existing steam

Comparison of existing main operating conditions vs revamp

Unit	Charge gas E-GB-201		
	Existing	Revamp	Deviation, %
Mass flow, kg/hr	222 537	267 929	+20
Suction pressure, bara	1.45	1.66	+14
Discharge pressure, bara	37.26	38.19	+2.5
Volume flow, m ³ /hr	151 450	153 751	+1.5
Shaft power, kW	27 775	31 506	+13
Unit	Propylene refrigerant E-GB-501		
	Existing	Revamp	Deviation, %
Mass flow, kg/hr	308 839	389 057	+26
Suction pressure, bara	1.24	1.24	+0.0
Discharge pressure, bara	16.22	16.22	+0.0
Volume flow, m ³ /hr	109 750	137 955	+26
Shaft power, kW	19 655	27 125	+38
Unit	Ethylene refrigerant E-GB-601		
	Existing	Revamp	Deviation, %
Mass flow, kg/hr	26 675	33 953	+27
Suction pressure, bara	1.02	1.02	+0.0
Discharge pressure, bara	27.45	27.45	+0.0
Volume flow, m ³ /hr	12 805	16 268	+27
Shaft power, kW	5020	6350	+26

Table 2



Figure 4 Rotor for propylene refrigerant compressor GB-501, five stages

turbine (see Table 2). Therefore, the next best solution was mutually agreed, which was replacement by a footprint steam turbine (extracting/condensing type). All piping connections and casing supports have the same spatial positions, while the vacuum surface condenser and condensate pump system can be reused.

The compressor casing is able to handle the capacity increase by means of newly designed internals. The combined seal, lubrication and control oil system does not require

modification. The revamped equipment was put into operation in September 2007 (see Figure 4).

Ethylene refrigerant train E-GB-601

- Steam turbine rotor (three stages)
- Steam turbine stator parts (three stages)
- Compressor rotor (three sections, eight stages)
- Compressor stator parts
- Flexible element coupling between steam turbine and compressor.

The steam turbine (a condensing

type) was revamped in accordance with the increase in shaft power. In this instance, the vacuum surface condenser and condensate pump system could be reused.

The compressor casing is able to handle the capacity increase by means of the newly designed internals. The combined seal, lubrication and control oil system does not require modification. The revamped equipment was put into operation in September 2008 (see Figure 5).

In general, the change-out of existing hardware was carried out during a routine overhaul in each respective year. In many cases, these compressors can be accommodated with newly designed rotors and stationary diaphragms within the restrictions of the existing casing, either originally manufactured by Thomassen or by any other compressor vendor. This makes a revamp very competitive when compared to the investment required for new compressors, including civil work and the cost of auxiliary equipment.

Conclusion

This revamp project proved to be beneficial from both a technical and an economic perspective. The technical improvements, in combination with the increased throughput of the plant, justified the decision to revamp the existing equipment. The old compressor parts and steam turbines could be replaced by new internals and steam turbines during a routine maintenance period.

The route to an extensive revamp began, in this case, by solving both the minor and major mechanical and operational problems that had affected the customer's rotating equipment for many years.



Figure 5 New internals for steam turbine GT-601, three stages

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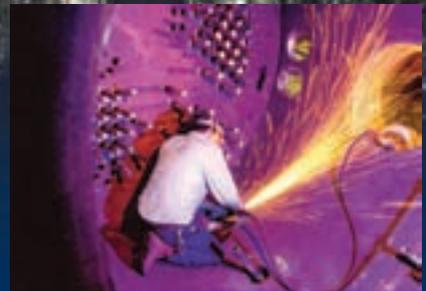
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Additional hydrogen production by heat exchange steam reforming

Applying the heat exchanger principle in hydrogen manufacture can significantly reduce the consumption of hydrocarbon feedstock

JACK HESELER CARSTENSEN
Haldor Topsøe

Many refiners are in need of additional hydrogen in order to process more feed or lower-quality crude. Over the past 20 years, Haldor Topsøe has developed a number of steam reforming technologies that can be implemented for additional new hydrogen plant capacity and also as an add-on to an existing hydrogen plant to provide extra hydrogen capacity.

These technologies are based on heat exchange steam reforming and are characterised by efficient heat transfer, resulting in feed and fuel savings of up to 20% compared to a traditional box-type hydrogen plant.

Heat exchange steam reforming in hydrogen production is not a new development. Back in the 1980s, Haldor Topsøe developed the first small-scale steam reformers based on the heat exchange principle, encouraged by increasing energy prices. The capacity of the first hydrogen units ranged from 100 000 scfd to 1 million scfd and were typically used for fuel cell applications. As the technology matured, it became possible to increase the capacity and, in 1997, the first industrial-scale (5 million scfd) hydrogen units were successfully put into operation. The development of the heat exchange reformer technology has continued, and Topsøe has licensed hydrogen plants, including a heat exchange reformer, with capacities of up to 185 million scfd. At the same time, a number of different variants of heat exchange reforming technologies have been developed, enabling the construction of new, tailor-made



Figure 1 HTCR reformer layout

hydrogen plants or the revamp of existing units to create maximum value. Industrial feedback has confirmed that the use of heat exchange reforming can save up to 20% on feed and fuel consumption (and corresponding savings in CO₂ emissions) compared to conventional steam reforming.

Heat exchange reforming: principles

Being an endothermic reaction, the steam reforming of hydrocarbons requires a significant heat input to obtain the desired conversion to hydrogen. In a conventional steam reformer, heat transfer takes place by radiation, which leads to a limited thermal efficiency, as evidenced by a high flue gas temperature — typically more than 1800°F (980°C). The thermal efficiency of a conventional steam

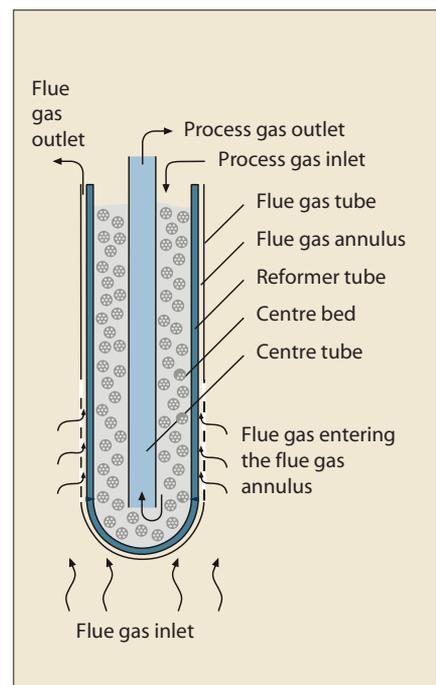


Figure 2 Principle of HTCR tube

Consumption figures for a HTCR and a conventional 27 million scfd hydrogen plant

27 million scfd hydrogen plant	HTCR process	Conventional SMR process
Feed, BTU/scf H ₂	354	352
Fuel, BTU/scf H ₂	11	53
Feed + fuel, BTU/scf H ₂	365	405
Steam export, BTU/scf H ₂	0	60
Cost of feed and fuel, \$ million/year	18.5	20.5

Table 1

Consumption and TIC figures for a 185 million scfd hydrogen plant

185 million scfd plant	State-of-the-art SMR process + HTER	State-of-the-art SMR process
Relative total installed cost (TIC)	100	100
Feed + fuel, BTU/scf H ₂	372	390
Steam export, BTU/scf H ₂	43	71
Feed + fuel minus steam, BTU/scf H ₂	329	319
Relative power consumption	86	100
Cost of feed and fuel, \$ million/year	129.9	136.4

Table 2

reformer is around 50%, and the surplus heat is used for steam production. Many refiners have little or no use for the steam export generated in a hydrogen plant, which is therefore considered of low value.

Heat exchange reformers are very compact and have a high thermal efficiency. The majority of the heat transfer takes place by convection with hot flue gas or hot process gas, whereby the thermal efficiency can be increased by as much as 60–70% compared to the radiant solution. You could say that, in a heat exchange reformer, the waste heat energy is used for producing extra hydrogen instead of surplus steam.

Heat exchange reforming is well suited to both standalone units and as a revamp option for increasing the capacity of existing plants.

Heat exchange reforming technologies

HTCR

The Haldor Topsøe Convection Reformer (HTCR) is a heat exchange reformer in which the process gas is heated mainly by hot flue gas. The HTCR is very compact and suited to new hydrogen units and to add-on revamps for increasing the capacity of existing plants.

The reformer is shown in Figure 1 and the principle is shown in

Figure 2. The reformer consists of a vertical refractory-lined vessel containing the tube bundle with bayonet tubes. The heat from the flue gas is transferred to the process gas inside the bayonet reformer tubes, resulting in low feedstock consumption and zero steam export. In an HTCR reformer, the

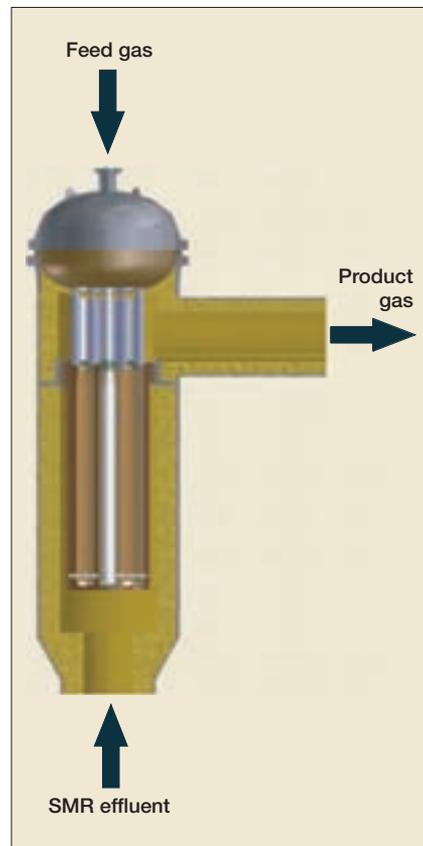


Figure 3 HTER principle

heat input is provided by only one burner, which ensures a very easy operation and a fast load response. The easy operation implies that only a minimum of operator attendance is required, and there are examples of unattended operation of HTCR plants.

The unit is to a high degree skid-mounted and the reformer is shop-lined, minimising erection time and cost on site. Industrial experience with HTCR includes more than 30 plants and design capacity of up to 27 million scfd.

Case study 1

Grassroots 27 million scfd HTCR hydrogen plant

In connection with an extensive revamp of an existing refinery in the Russian Federation, an analysis of the future hydrogen balance showed that an additional 160 million scfd of hydrogen would be required. The majority of the hydrogen was needed for a new hydrocracking unit, whereas the remaining part would be needed for hydrotreating purposes. The new hydrotreating unit was scheduled to come on stream one year before the hydrocracking unit, and the refinery decided to build two separate grassroots plants to cover the hydrogen requirement: a 27 million scfd unit and a 130 million scfd unit. HTCR technology was chosen for the smaller unit due to its low feed and fuel consumption and fast implementation time, and requirements to export steam would be covered by the larger hydrogen unit.

Table 1 shows the consumption figures for a 27 million scfd HTCR hydrogen plant compared to a conventional SMR process typically used for this capacity range.

The example clearly illustrates the advantage of the HTCR process in the case of low or no value of steam export. Based on a feed and fuel cost of \$5.36/million BTU,¹ the annual savings for a HTCR plant amount to \$2 million compared to a conventional steam reformer process. Furthermore, the 11% lower consumption of feed and fuel results in a correspondingly lower emission of CO₂.

HTER

In the Haldor Topsøe Exchange Reformer (HTER), the reaction heat is provided by hot process gas. The HTER can be used in grassroots hydrogen plants in combination with a radiant wall steam reformer and also as an add-on unit for additional capacity in an existing plant.

The HTER utilises a bayonet tube or a two-bed system (see Figure 3), allowing for optimal utilisation of the heat transfer areas. The hot reformer effluent is used as the heating medium and the high pressure results in a very compact design.

The inclusion of an HTER unit reduces fuel consumption and steam production, and can increase capacity by up to 30%. The HTER is well suited to both capacity revamps in existing plants as well as to new units, where factors such as low feed and fuel consumption and compactness are important.

Case study 2

3 x 185 million scfd grassroots hydrogen plants incorporating HTER

In connection with a large refinery project in the Middle East, a refiner was in need of an additional 555 million scfd of hydrogen, which was to be covered by three grassroots plants. The three plants should be designed to operate on a range of different feedstocks, depending on cost and availability in the refinery.

Based on analysis of investment cost and consumption figures, the refiner chose a solution with three new grassroots plants of 185 million scfd, each plant a state-of-the-art SMR process incorporating an HTER. Table 2 compares the key figures for the chosen HTER solution compared with an alternative without an HTER.

For the same investment, the solution based on an HTER results in a 5% saving in feed and fuel, which, based on a price of \$5.36/million BTU,¹ corresponds to annual savings of \$6.5 million, correspondingly lower CO₂ emissions and a lower power consumption.

TBR

The Topsøe Bayonet Reformer

Consumption and TIC figures for a 43 million scfd hydrogen plant based on TBR			
43 million scfd plant	State-of-the-art SMR process	TBR with steam export	TBR no steam export
Relative TIC	100	95	95
Feed + fuel, BTU/scf H ₂	390	361	350
Steam export, BTU/scf H ₂	71	30	0
Feed + fuel minus steam, BTU/scf H ₂	319	331	350
Relative power consumption	100	90	90

Table 3

(TBR) is the newest addition to the Topsøe family of heat exchange reforming technologies, combining the principle of convection heat transfer known from HTCR and radiant heat transfer known from the radiant wall steam reformer.

The inclusion of an HTER unit reduces fuel consumption and steam production, and can increase capacity by up to 30%

The TBR consists of a series of double tubes in a single row in a radiant furnace box, similar to conventional steam reforming technology.

Feed gas flows downward

through the catalyst bed located in the annulus between the two tubes (see Figure 4). At the bottom of the catalyst bed, the gas turns and continues upward through the inner, empty bayonet tube. The bayonet tube arrangement is the same as that used in HTCR technology. The gas exits at the top of the bayonet tubes and is collected in a common header.

Improved heat utilisation, in combination with a high average heat flux in the TBR tubes, significantly reduces the size and capital cost of the hydrogen plant and provides hydrogen production with low hydrocarbon consumption and little or no steam export.

Table 3 illustrates the main features of a 43 million scfd TBR hydrogen plant compared to a state-of-the-art SMR process. For the TBR process, we have considered two cases: with and without steam export.

The table illustrates that, in the case of little or no requirement for steam export, the TBR technology

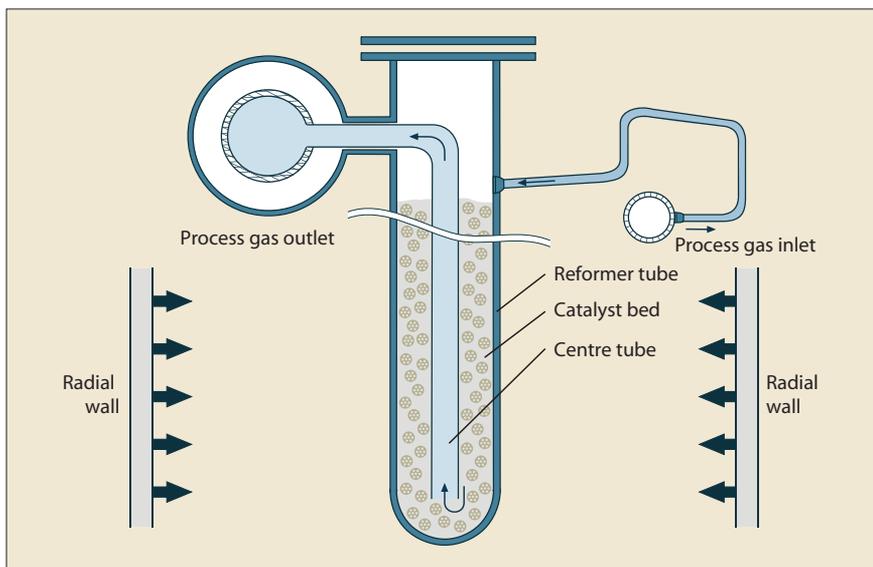


Figure 4 Simplified sketch of the TBR principle

offers a solution for the refinery with a 5% lower investment cost and, at the same time, up to 10% lower consumption of feed and fuel compared to a state-of-the-art SMR process.

Revamping of an existing hydrogen unit

Revamping an existing hydrogen unit with an HTER or an HTER is an economically attractive solution to increase capacity and reduce specific feed consumption.

When revamping with an add-on heat exchange reformer, the capacity can typically be increased by up to 30% with a much lower investment compared to a new plant. At the same time, the compactness of a heat exchange reformer ensures that the additional required plot area is kept to a minimum.

In order to determine the feasibility of a revamp, Topsøe performs a revamp feasibility study to provide the client with the optimal solution. The study takes into account the status of the existing unit, the various requirements of the client, as well as prevailing conditions such as availability and price of feedstock, plot plan and available downtime for revamp.

Revamping with an HTER

As a revamp option, the HTER is a

flexible solution, whereby the HTER is operated in parallel with the existing reformer (see Figure 5). This layout results in a revamp package consisting of a few pieces of equipment; namely, an HTER with a waste heat section, a process boiler and a combustion air blower that allows for independent operation of the two reformer sections.

The on-site implementation of an HTER revamp requires minimum downtime because basically only

HTER results in a hydrogen plant with lower specific feed and fuel consumption and lower steam export

three tie-ins need to be performed during a shutdown. The remaining construction can be performed during operation of the existing unit. An HTER revamp can typically increase production capacity by up to 30%. Following a revamp, the specific feed consumption decreases as a result of the higher thermal efficiency of the HTER.

Revamping with an HTER

Compared to the HTER revamp, an HTER revamp is completely integrated into the existing unit (see Figure 6), resulting in an economically attractive solution requiring less plot area. The benefits of an HTER revamp are best utilised when the process gas exit temperature from the existing steam reformer is approximately 1630°F (890°C) or higher. For an HTER revamp, a longer downtime must be foreseen, as the HTER is connected to the existing reformer.

An HTER revamp can typically increase production capacity by up to 30%. Integration of the highly efficient HTER results in a hydrogen plant with lower specific feed and fuel consumption and lower steam export, as illustrated in the example below.

Case study 3

Increasing capacity 25% by adding an HTER

A refinery in Asia operates a 65 million scfd, an 80 million scfd and a 110 million scfd hydrogen plant, all designed by Topsøe. In 2009, the client wanted to increase hydrogen consumption by 27 million scfd. It was important that the project could be implemented during a normal four-week turnaround and

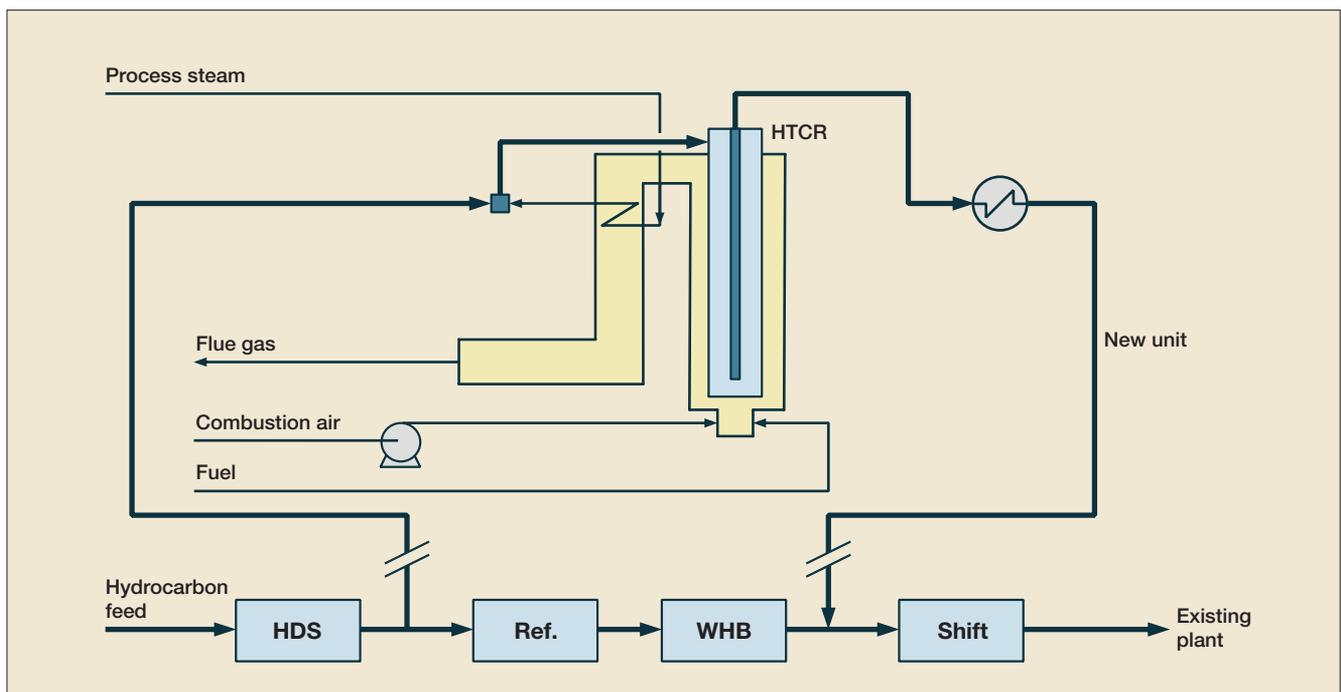


Figure 5 Schematic layout of an HTER revamp solution

that the additional plot area was minimised. The following revamp alternatives were considered:

- Adding an HTER reformer to each of the smaller (and older) units
- Adding an HTER reformer to the 110 million scfd plant
- A new 27 million scfd grassroots plant.

A feasibility study concluded that adding an HTER reformer could yield the 27 million scfd capacity increase at the minimum investment. The feasibility study also identified the required modifications/replacement of existing equipment necessary to achieve the desired capacity increase. Total installed cost (TIC) for the complete revamp including a new HTER and the required modifications/replacement of existing equipment amounted to approximately 60% of the TIC for a new separate hydrogen plant with the same capacity. The results in terms of energy consumption are shown in Table 4.

The results show that the capacity could be increased by 25%, with a 10% lower specific consumption of feed and fuel.

Conclusion

Heat exchange reformer solutions developed and industrially

Consumption figures before and after an HTER revamp of a 110 million scfd hydrogen plant		
110 million scfd plant revamp	Before revamp	After revamp
Capacity, million scfd	110	137 (+25%)
Feed and fuel consumption, BTU/scf H ₂	425	384 (-10%)
Steam export, BTU/scf H ₂	96	52 (-46%)
Net energy consumption, BTU/scf H ₂	329	332 (+1%)

Table 4

demonstrated by Topsøe are solutions for adding additional hydrogen capacity, either in the form of standalone plant or as a revamp of an existing hydrogen unit. By utilising the heat exchanger principle in hydrogen manufacture, it is possible to significantly reduce the consumption of hydrocarbon feedstock compared to traditional radiation-based technology and reduce CO₂ emissions accordingly. Furthermore, heat exchange reforming is compact, characterised by easy operation and low investment cost. With the range of technologies presented here, it is possible to provide tailor-made compact and energy-efficient solutions for additional hydrogen production in refineries.

References

- 1 Natural Gas Futures NYMEX price (Henry Hub) average for Dec 2009.

2 Dybkjaer I, Winter Madsen S, Compact hydrogen plants, *Hydrocarbon Engineering*, Nov 2004.

3 Gyde Thomsen S, Han P A, Lock S, Werner E, The first industrial experience with the Haldor Topsøe exchange reformer, *AICHE*, 2006.

4 Hedegaard Andersen K, Hydrogen agenda, *Hydrocarbon Engineering*, Nov 2006.

5 Winter Madsen S, Olsson H, Steam reforming solutions, *Hydrocarbon Engineering*, Jul 2007.

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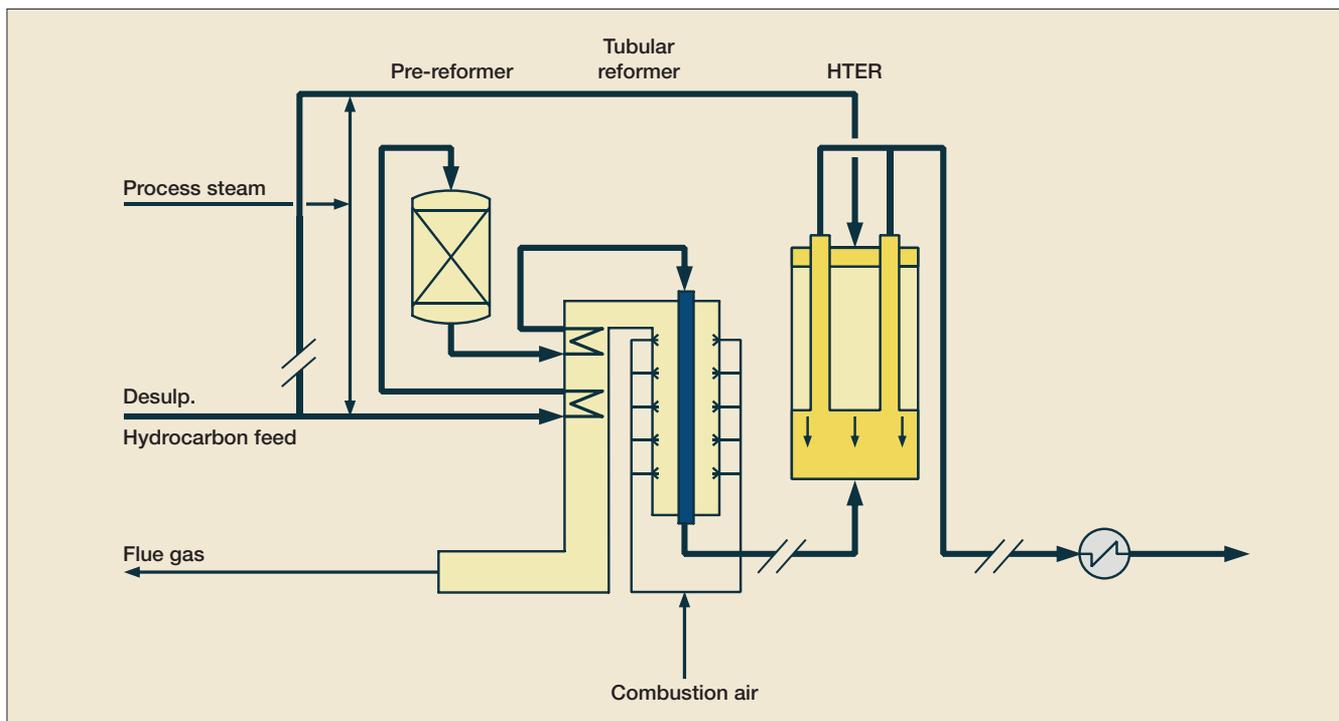


Figure 6 Schematic layout of an HTER revamp solution

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Amine contactor revamp

With the old trays in the scrap yard and no working drawings available, an amine still revamp called for engineering detective work

VILAS LONAKADI
Amistco Separation Products Inc

The amine still column at an energy plant was designed to treat 300 million scfd of sour gas containing H₂S and CO₂ with alkanolamine. The sweet gas after treatment is routed to an energy company. The unit, on the Gulf coast of the US, was shut down in September 2009 for regular maintenance. On opening the amine still, engineers found that the trays were all damaged, corroded and plugged with iron sulphides and scale.

Almost all of the trays had fallen to the bottom of the still. Due to a very tight schedule, the turnaround contractor removed all of the damaged trays and threw them away in a scrap yard. The operator then realised that it no longer had drawings of the existing trays and there was not enough time to carry out a complete process and equipment design for replacement trays. Hence, the company decided to replace the trays with new versions to the same design. However, no process data were available and there were no existing tray drawings, factors that together posed great challenges in duplicating the design and fabricating replacements. Amistco was called in to design and supply new trays.

Conventionally, the design of distillation column internals involves two main steps: process design and optimisation, and equipment design. The process design, after a number of iterations and optimisations, specifies the internal vapour and liquid flow rates and properties across the number of theoretical stages required to achieve separation efficiency, while the equipment design and

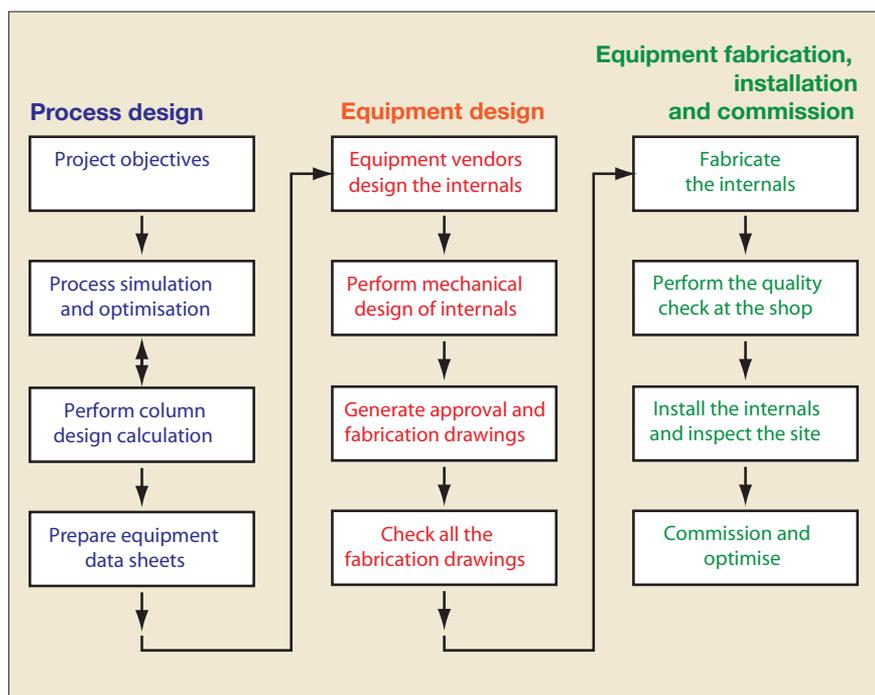


Figure 1 Conventional steps in the design of distillation column internals

fabrication step uses this data to design the actual hardware (see Figure 1). Ideally, the equipment's

Conventionally, the design of distillation column internals involves two main steps: process design and optimisation, and equipment design

design and fabrication evolves from an earlier set of design, operations and drawings.

The process design and optimisation step mainly determines how close the tray geometry is to reaching maximum capacity and efficiency. However, in this case, due to the non-availability of process data, this first step was bypassed. Subsequent detailed discussions with the operational and process staff of the gas plant concluded that the column and existing trays were working to their satisfaction in terms of capacity and efficiency, and there were no plans to either increase or decrease the unit's operations in the near future. Hence, the decision was taken to duplicate the existing geometry in new trays.

Since drawings for the existing trays were not available, an inspection crew consisting of engineers,

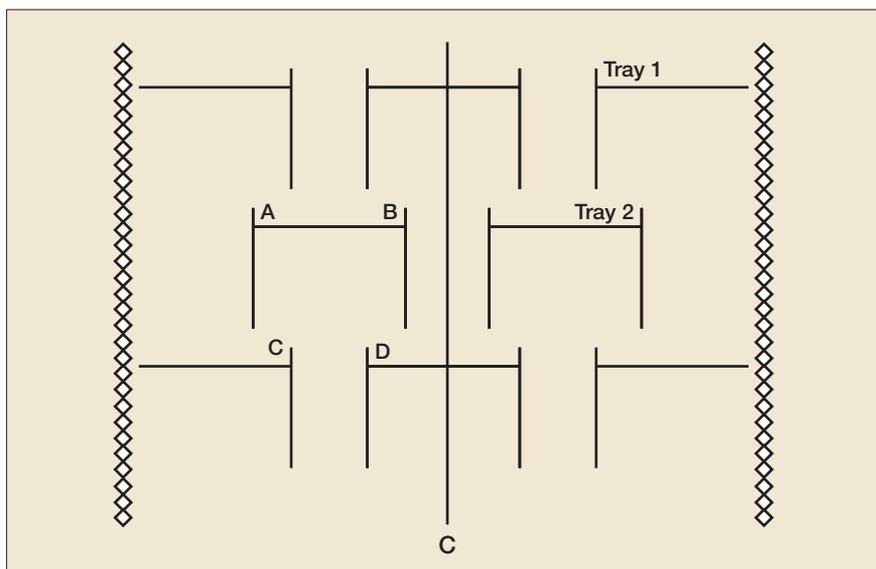


Figure 2 Schematic of a four-pass tray arrangement showing panel designation

designers and drafters was dispatched to the plant to inspect the tower. The tower diameter was measured at several locations and confirmed to be 156in (400cm). Although, at some locations, the vessel's internal diameter was not uniform, the inspection crew recognised that the "out of roundness" was within the tolerance level of 1% of vessel internal diameter, in accordance with ASME UG80.

An inspection of the weld-ins inside the tower revealed that the tower consisted of 20 four-pass trays with side and centre downcomer trays at the top tray, ending with off-centre downcomer trays at the bottom. Although some corrosion was observed at the support rings and downcomer bolting bars, with some cleaning required, they were deemed acceptable for the next run. An ultrasonic inspection of these weld-ins revealed that the support rings were 0.5in thick and

2.5in wide. Based on a turnaround schedule of five years, the support rings' thickness was deemed adequate to withstand the maximum allowable working pressure.

Given that this tower has four-pass trays, balancing the flow paths became critical, to avoid poor distribution of vapour and liquid

Based on the chord lengths of various support rings, width of inlet panel and the minor beam dimensions, it was concluded that the side downcomer widths were

11.75in (30cm), the centre downcomer widths were 12in and the off-centre downcomer widths were also 12in. At the same time, and based on the recovery of some material from the scrap yard, some of the downcomer panels were reassembled, to the point where it was possible to verify these dimensions. It proved interesting to note that the judgment made through chord lengths matched closely the widths of the downcomer panels. Measuring between the support rings provided an idea of tray spacing. An inspection of the tray panel indicated that the trays had round floating valves with 0.4375in (1.1cm) lift and they were 1056 in number.

Following discussions with the operator to confirm that the existing trays did not have any operational problems, the tray layout was decided upon, considering that the existing trays were operating in froth regime and there was no excessive entrainment, excessive downcomer backup or excessive downcomer choke, with reasonable pressure drop. Although some corrosion and fouling was observed on the trays, it was felt that their condition was acceptable following a five-year run length.

The column was required to operate in different cycles with varying gas rates. As a result, the percentage turndown would be most important. Hence, the decision was taken to continue using floating valves.

Given that this tower has four-pass trays, balancing the flow path became critical, to avoid poor distribution of vapour and liquid, which would reduce the efficiency and/or the capacity of the trays. It is important to ensure that the vapour and liquid contact each other uniformly across each panel and to make sure that the vapour-to-liquid ratio is as close to unity on each of these panels. All four-pass trays have two different sets of configurations. One set consists of two side downcomers and a centre downcomer, while the other set consists of two off-centre downcomers, and these alternate in a given column. As a result, the trays will have four active panels, with

Tray design specifications for an amine still revamp

Tray geometry	Measurement
Tower diameter	156in
# of passes	4
# of valves	1056
Type of valves	Round floating valve
Valve lift	0.4395in
Side downcomer width	11.75in
Centre downcomer width	12in
Off-centre downcomer width	12in
Distance of off centre downcomer from tower wall	45.5in
Metallurgy	SS 304L, 14 Ga./12Ga.

Table 1

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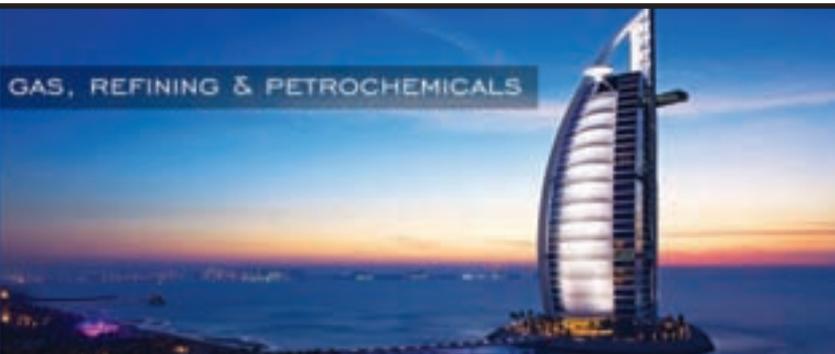
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panels A and B designated for the side and centre downcomer trays, and panels C and D designated for the off-centre downcomer trays (see Figure 2).

Within the column, the vapour and liquid streams are split, but recombined on each path. Hence, if the split is not uniform across each of these paths, the tray will flood prematurely or it will lose its efficiency. The liquid flows on the tray deck and downcomer are controlled by modifying the downcomer clearances and/or outlet weirs, while the vapour flows are balanced using the vapour tunnels or by providing the same bubbling area. Four-pass trays are balanced either by the equal bubbling area method or by providing equal flow path length.

In this case, the active bubbling area concept was used and the outlet weir lengths and weir heights for panels A and B were kept the same, with the same number of valves on each panel. As these downcomer panels were available, a positive material identification (PMI) revealed that they were made

of SS 304L, and ultrasonic measurements indicated that the downcomer trusses were 7 gauge thick and the tray panels were 14 gauge thick.

Thus, even in the absence of process data and drawings of the trays, a systematic evaluation of all the

Four-pass trays are balanced either by the equal bubbling area method or by providing equal flow path length

recuperated scrap, inspection of the tower and experienced engineering judgment helped to finalise the tray geometry for this four-pass mass transfer tray for an amine still (see Table 1).

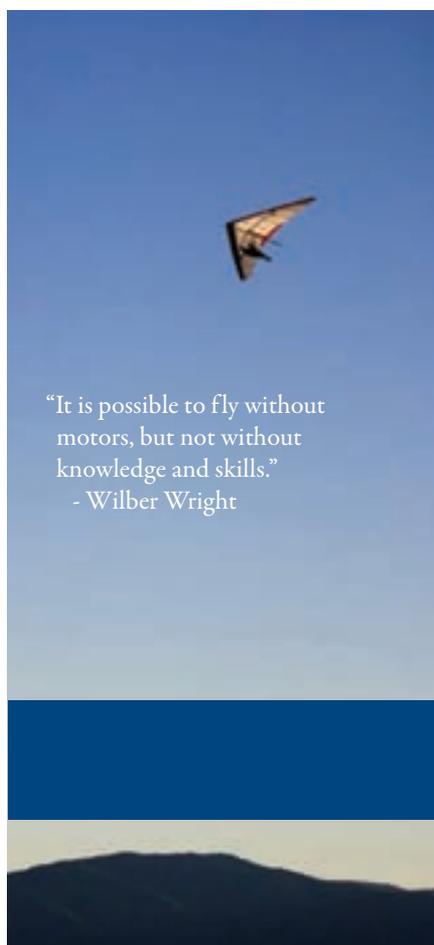
Based on the calculated geometry, mechanical designs such as stress calculations and deflections were

calculated and found to be 90% allowable at design conditions and 0.173in respectively. The trays were then fabricated and delivered to the customer a week later. The amine column has been commissioned and is operating satisfactorily.

Conclusion

In the absence of process data and existing tray drawings, an innovative and systematic evaluation of all the damaged internals, inspection of the tower, and application of experienced engineering judgment and teamwork resulted in the effective design and fabrication of replacement fractionating trays for an amine still.

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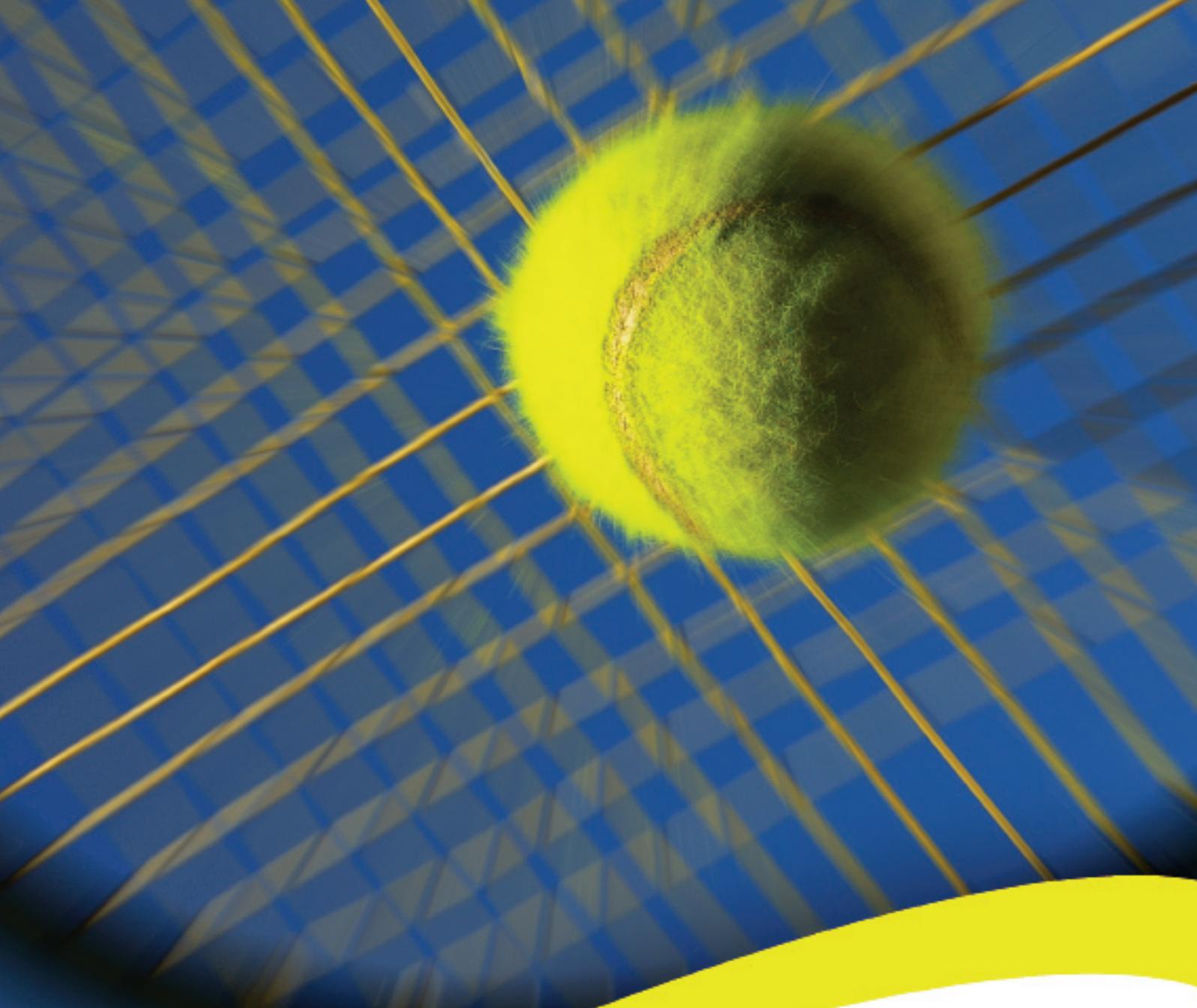
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