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LyondellBasell's ChannelView petrochemicals complex near Houston, Texas
Photo: LyondellBasell

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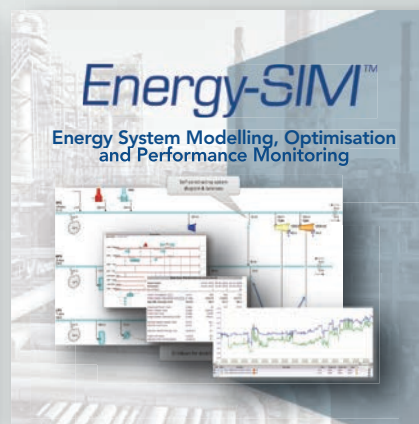
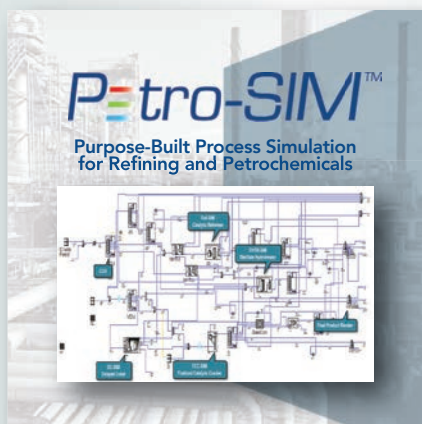


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answers@kbc.com • www.kbc.com • blog.kbc.com

Editor
Chris Cunningham
editor@petroleumtechnology.com

Production Editor
Rachel Storry
production@petroleumtechnology.com

Graphics Editor
Rob Fris
graphics@petroleumtechnology.com

Editorial
tel +44 844 5888 773
fax +44 844 5888 667

Business Development Director
Paul Mason
sales@petroleumtechnology.com

Advertising Sales Office
tel +44 844 5888 771
fax +44 844 5888 662

Publisher
Nic Allen
publisher@petroleumtechnology.com

Circulation
Jacki Watts
Louise Shaw
circulation@petroleumtechnology.com

Crambeth Allen Publishing Ltd
Hopesay, Craven Arms SY7 8HD, UK
tel +44 844 5888 776
fax +44 844 5888 667

PTQ (Petroleum Technology Quarterly) (ISSN No: 1632-363X, USPS No: 014-781) is published quarterly plus annual *Catalysis* edition by Crambeth Allen Publishing Ltd and is distributed in the US by SP/Asendia, 17B South Middlesex Avenue, Monroe NJ 08831. Periodicals postage paid at New Brunswick, NJ. Postmaster: send address changes to *PTQ (Petroleum Technology Quarterly)*, 17B South Middlesex Avenue, Monroe NJ 08831. Back numbers available from the Publisher at \$30 per copy inc postage.



Diesel in decline

The aftershock of the Volkswagen emissions scandal continues to rumble on in Europe. And while tailpipe technology is chiefly to blame in that case, the incident has done nothing for the reputation of diesel by adding to public concerns about particulate levels in the streets. It is as well to note, though, that the fashion for diesel is in decline across the globe.

In India, diesel is the bargaining chip in a game of environment vs economy. The nation's cities, and the Delhi mega-conurbation in particular, carry the unwelcome label 'among the world's most polluted'. As a result, moves are afoot to restrict sales of diesel-engined vehicles as well as larger gasoline consuming cars in major centres of population.

At the same time, greater than 7% growth in the national economy has had to be fuelled somehow. Hence India has the world's fastest growing refining industry. The vehicles targeted in the attempt to clean up metropolitan India's air account for around one-third of reported total diesel sales in the country, which amounts to a mighty problem for refiners juggling with unavoidable distillate production. Perhaps the much-vaunted introduction of cleaner, Euro VI specification diesel will help to allay the fears of state regulators and so avoid the need for refiners to wind down their throughput.

City centre pollution also plays a part in China's issue with diesel, although a general slowing down of the national economy and a shift of emphasis away from manufacturing as the mainstay of growth are more significant factors.

Production of refined products in China is beginning to outstrip domestic consumption significantly as the delayed impact of massive refinery expansion greets declining demand passing in the opposite direction. The result has been the PRC's emergence as a net exporter of diesel and the rest of the product slate, the past year alone accounting for an increase in exports of around 30%. A component of such rocketing growth is the change in government policy, previously outlined on this page, which allows smaller, independent refiners to export alongside the much larger, state-owned corporations.

As always, a surge in export-available product has to find somewhere to go. But the trend in diesel sales (for instance in India) is likely to be less than firm for at least the foreseeable future. Significant Iranian diesel production is due on stream later in the year, and in the distant markets of Europe diesel is most definitely no longer the leading motor fuel of choice.

One measure to counter a soft market for diesel is to reduce the price of refined products significantly, a policy already well practised on behalf of Chinese steel. While this might help to shift surplus diesel, it would also apply downward pressure to refining margins, to the point where refiners are cutting back their throughput and ensuring that the tank farm is not too full of raw material that may plunge in value before it ever reaches the crude unit. Demand from one of the world's largest importers of crude falls further. A significant dip in the price of crude worldwide results.

CHRIS CUNNINGHAM

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While you've been getting to know the scientists behind our advanced research and technology the past few years, we've been increasing the performance and quality of our catalysts. We've continued to surge ahead with industry leading results and the next generation of catalysts are here. Our advanced array of products protect operations and bring all the technology together into practical, customized solutions for your refinery.



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CRITERION
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Q Is there any way we can increase feed rate to our FCC unit without losing stripper efficiency?

A Yorklin Yang, Global Process Technology Advisor, BASF Refinery Catalysts, yorklin.yang@basf.com

There are several ways to maintain the stripper efficiency even at increased feed rate in an FCC unit. But these are dependent on the current stripper design and operation points.

Here are typical catalyst stripper design parameters:

- Diameter is based on catalyst flux: 3900 kg/m²-min (800 lbs/ft²-min); potential flooding at 6000 kg/m²-min
- Residence time should be at least 45 seconds, which controls length of stripper
- Cone section design should avoid long residence time – concern is catalyst fluidisation
- Typical total stripping steam is between 2-4 kg of steam/1000 kg catalyst.

An efficient catalyst stripper design is one that maximises mass transfer between the two phases (the stripping steam flowing up and the fluidised catalyst flowing down the stripper vessel). Hydrocarbons in the catalyst phase need to be replaced with steam. To enhance this mass transfer phenomenon, the stripper internals should have the following characteristics:

- Maximum surface area for mass transfer per unit volume of the stripper vessel
- Maximum cross-sectional area of the vessel available for catalyst and steam to flow through it
- Maximum active volume (no stagnation or dead zones)
- Uniform distribution of catalyst and steam to avoid channelling and by-passing
- Increased contact time and mixing between the two phases
- Excellent fluidisation quality
- Plug flow conditions
- Mechanically robust to withstand service.

We would recommend the following approaches on optimising stripping efficiency at increased feed rate.

Optimising steam rate in the catalyst stripper

The optimum steam rate to ensure proper stripping depends on the catalyst circulation rate, the stripper design, the catalyst diffusivity. Good stripping occurs when the delta coke is minimised. For a given design, the stripping steam rate is optimised by adjusting it to a value slightly above the one that minimises the regenerator temperature. This is determined by ramping steam rate upward until regenerator temperature stops dropping.

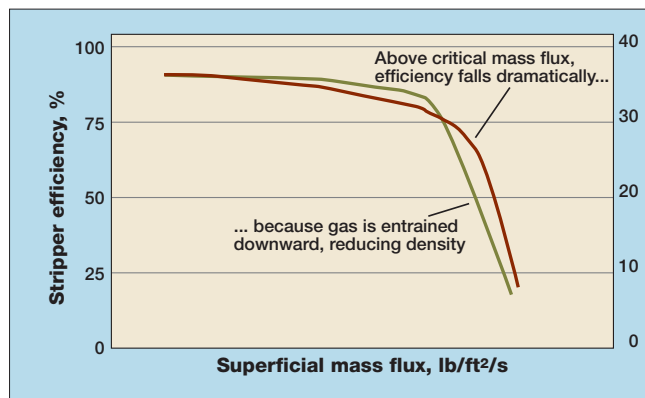


Figure 1 Stripping efficiency and density vs catalyst circulation

Optimising feed preheat temperature

Feed preheat temperature can be used to manipulate feed rate and/or conversion. Typical preheat temperature is between 350°F and 600°F (177-315°C). Each 25°F increase in feed preheat temperature could reduce catalyst circulation by about 4.0%. This will reduce catalyst flux in the stripper and move the stripper to operate at above critical flux point (see Figure 1).

Reformulate catalyst to have better coke selectivity and optimise catalyst diffusivity

While the majority of stripping is from hydrocarbon entrained or between catalyst particles, using a catalyst with good diffusivity will directly affect the observed stripper performance. Entrainment of hydrocarbons in the catalyst pores and the resulting carry-over to the regenerator will be minimised using a catalyst with high diffusion. With such a catalyst, it is easier to remove the hydrocarbons from the catalyst.

Further reading

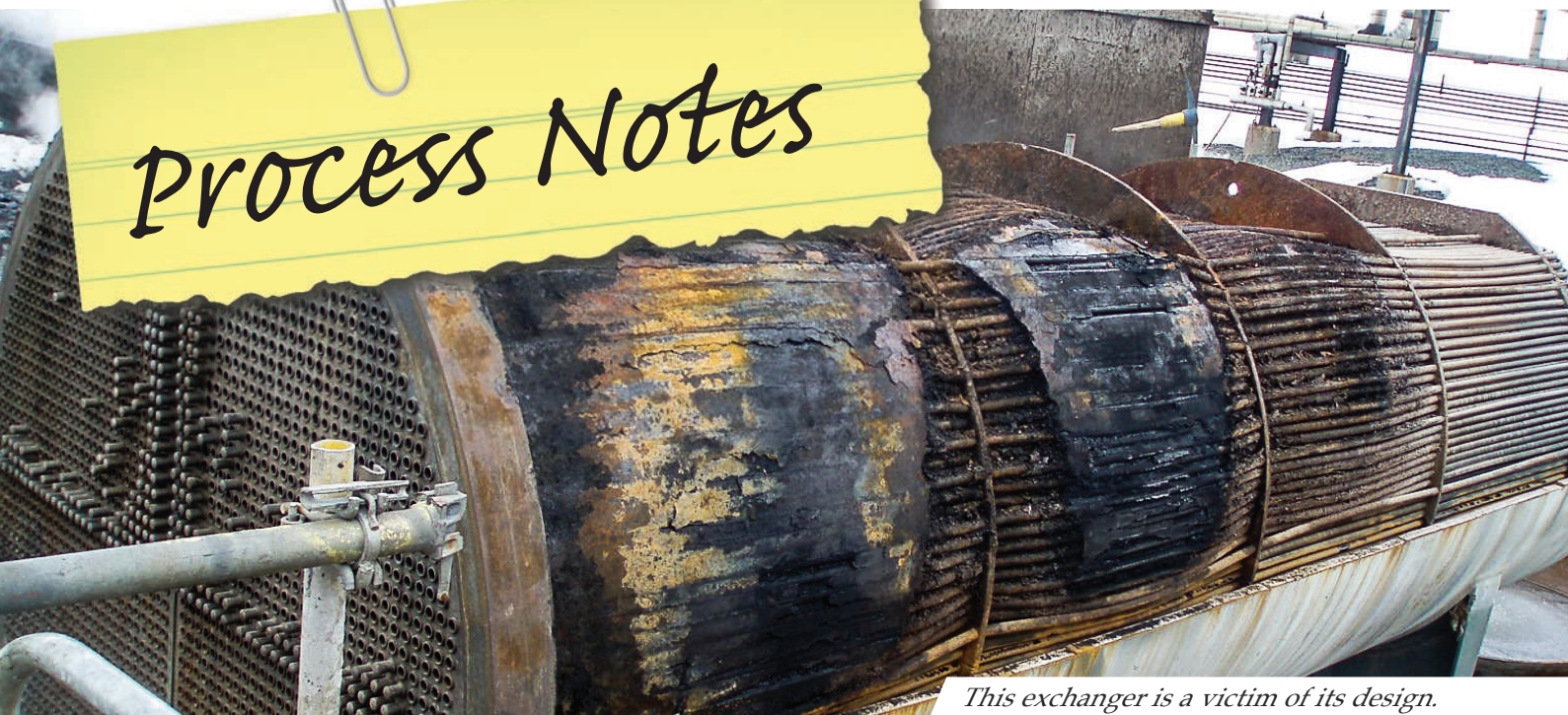
- 1 Miller R, Yang Y, Gbordzoe E, New Developments in FCC Feed Injection and Stripping Technologies, AM-00-08.
- 2 Marri R, Soni D, Catalyst stripper improves FCC unit performance, PTQ Q3 2012.

A Tom Ventham, Senior Technical Services Manager, Johnson Matthey, Tom.Ventham@matthey.com

As feed rate to the FCC unit is increased, the regenerated catalyst slide valve will open further, increasing catalyst circulation to maintain constant riser outlet temperature. The additional catalyst volume circulated is required to heat and vaporise the incremental feed as well to compensate endothermic reactions taking place in the riser. Ultimately, the required catalyst circulation rate is set by the heat balance, of which

Additional Q&A can be found at www.eptq.com/Qanda

Process Notes



This exchanger is a victim of its design.

Opportunity or Annoyance?

Price differentials between conventional and opportunity crudes compel refiners to process increasing percentages of lower valued opportunity crude. However, as many refiners have learned the hard way, opportunity crudes are tied to unique processing challenges. Furthermore, existing crude unit configurations may limit high-profit opportunity crude to a disappointingly small proportion of the total unit blend.

Processing a changing slate of opportunity crudes of varying compositions requires a CDU/VDU design that is flexible, reliable, and commercially proven. A good design must control chronic problems associated with many of these crudes including corrosion, exchanger fouling, tray plugging, vacuum heater and wash bed coking, and unreliable product quality prediction. CDU/VDU designs for challenging crudes should be based on proven best-practices rather than simply on a low-CAPEX strategy. For example, a poorly designed heat exchanger may operate well initially, but high fouling will quickly lower efficiency and eventually limit throughput. Up front investment in engineering experience and know-how pays dividends when units meet capacity and run-length targets.

Over the course of 20 years, Process Consulting Services has completed more than 130 revamp and grass-roots designs supported by over 75 detailed test runs. An extensive collection of test run equipment performance data and feed/product analyses enables confident prediction of real-world opportunity crude performance. PCS has enabled refiners worldwide to extend crude unit run lengths from months to years while improving yields and operability.



Test run feed and product samples



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FCC feed rate is an important independent input variable.

The issue some FCC operators experience when catalyst circulation increases is that residence time in the stripper section becomes insufficient to fully purge all strippable hydrocarbons from the FCC catalyst. In more extreme cases, steam flows can be 'dragged down' by the high catalyst flow rate. In both these cases the rate of coke entering the regenerator increases (the rate of coke divided by the catalyst circulation rate being defined as 'delta coke'), a fact that is made worse by the unstripped hydrocarbons being of a relatively high hydrogen content further boosting regenerator temperatures (due to the high energy of combustion of hydrogen) and leading to accelerations in the hydrothermal deactivation of the FCC main catalyst (accelerated by presence of elevated levels of hydrogen (further exacerbated if quantities of steam from the stripper are also pulled under). Therefore, the overall net effect is an increase in the temperature of the regenerator, potentially exceeding design conditions and resulting in undesired feed rate or severity reduction, and an increase in the fresh catalyst addition rate to compensate for losses in activity.

Improvements can be made in stripper efficiency either by increasing the stripping steam rate when FCC feed rate is high (stripping steam rate should be set above the point at which reductions in steam rate result in increases in regenerator temperature) or by redesigning the stripper, perhaps moving to structured packing, to increase the throughput of the stripper.

Another approach can be to employ other methods of decreasing delta coke to offset any regenerator temperature escalations caused by stripper inefficiencies at high feed rates. FCC delta coke is made up of four different components: occluded coke (strippable coke), contaminant coke (metals coke), feed coke (Concarbon) and conversion coke (catalytic coke). When stripper operation becomes de-optimised due to high catalyst circulation rates the occluded coke component increases.

Johnson Matthey has developed the Intercat FCC additive, LO-Coker, which is effective in reducing delta coke. It enables enhanced matrix pre-cracking of Concarbon type molecules and passivates the effects of metals that promote dehydrogenation reactions. This will have the effect of reducing feed coke and contaminant coke to generate an overall delta coke reduction of approximately 10% based on laboratory studies.

The purpose of LO-Coker is to lift a constrained FCC unit away from the normal limitations (be it high regenerator temperature, maximum air blower rate, maximum cat cooler duty or wet gas compressor throughput) and take advantage of this shift by making operational adjustments that are most economically attractive for the refiner (for example, increasing residue processing, increasing equilibrium catalyst activity, and so on). In this case, use of LO-Coker would allow the refiner with a potential stripper efficiency issue (manifesting as maximum regenerator temperature limitation) to operate at higher feed rates without compromising other aspects of the FCC operation.

Q Demand for our diesel has dipped. What process steps do you recommend for increasing gasoline make?

A **Andrea Battiston, Senior Technical Specialist, Hydroprocessing, Albemarle, Andrea.Battiston@Albemarle.com**
Refiners can potentially take a number of steps to increase gasoline-to-diesel production ratio. The simplest one, depending on limitations in the refinery's process design and operations, is to increase naphtha make from the atmospheric distillation tower. This can in principle be done by running lighter crude oil with higher yields of naphtha and/or by tweaking distillation cut points to gain more naphtha volume.

Another approach consists of intervening in specific refinery processes by optimising operations and catalysts to maximise products that can be blended into the refinery's gasoline pool.

A diesel hydrotreater (HT), provided pressure is sufficiently high (say, $\text{ppH}_2 > 45$ bar), can be operated in a mild hydrocracking mode to increase naphtha yield while producing ultra-low sulphur diesel. For this case, depending on the operation and on the nitrogen slip, part of the hydrotreating catalyst bed in the bottom section of the reactor can be substituted by a dedicated mild hydrocracking (MHC) catalyst. MHC NiMo catalysts supported on amorphous silica alumina (ASA) have a broader application range, as they can be applied even at a higher nitrogen slip. Alternatively, NiMo or NiW catalysts based on zeolite-containing supports can be used, which typically allow a higher naphtha yield, but should be applied with a lower nitrogen slip. Albemarle's solution for higher nitrogen slip operations is KF 1022, while our solution for lower nitrogen slip and maximum naphtha production is KC 2715. Alternatively, KC 2601, KC 2610 and KC 2710 grades can be applied also depending on the requirements for hydrocracking activity (conversion) and products' selectivity and properties.

Hydrocracking (HC) and fluid catalytic cracking (FCC) operations can be optimised as well with the objective of increasing the gasoline-to-diesel ratio in the refinery.

For HC, the pretreat section (HC-PT) can be operated at more severe conditions or with improved catalyst systems to reduce nitrogen and aromatics slip to the HC section. This allows a more efficient hydrocracking unit (HCU) operation with higher conversion and, where desired, higher selectivity toward naphtha/kero. Albemarle's HC-PT catalytic solutions in this case include the use of KF 870, a new premium NiMo supported catalyst that can be combined with Nebula for maximum performance at reactor bottom. Nebula is Albemarle's bulk catalyst developed jointly with ExxonMobil that is widely used also for medium to high and high pressure distillate hydrotreating applications for additional desulphurisation activity and/or volume swell. In the HC-PT unit, KF 870 and Nebula can also be applied to free up reactor volume that can then be filled with additional HC catalyst, and thus increase overall conversion. At the same time, the



Process Notes

Crude units can be designed for reliability.

Maximize Reliability in Grassroots Crude Units

Crude unit operators are far too familiar with a long list of crude unit reliability problems including fouling in heat exchangers and fired heaters, poor desalting, corrosion of piping and equipment, and coking in the vacuum column wash zone. Many millions of dollars have been spent fighting these problems, yet they continue to force unplanned shutdowns with depressing regularity.

Revamps must address reliability issues, but project scope is hindered by the limitations of existing equipment. Grassroots design of crude and vacuum units presents an opportunity to get everything right the first time. Here are a few tips for designing a reliable and profitable crude/vacuum unit.

HEAT EXCHANGER AND HEATER FOULING

High velocities in heat exchanger tubes produce high shear at the walls, preventing foulants from accumulating. High shell-side velocities, coupled with exchanger designs that minimize dead zones in the flow, eliminate shell-side fouling. In fired heaters, high mass fluxes maximize wall shear, shorten residence time, and lower wall film temperatures, all of which reduce coking. Furthermore, reliable heaters must have correctly sized burners with proper burner-to-burner and burner-to-tube spacing.

DESALTING

Desalter size is highly dependent on crude blend due to dramatic variation in required centerline velocity. A unit must be designed with the flexibility to carefully control desalter temperature, which can range from 110°C to 150°C, by shifting heat from upstream to downstream of the desalters. Vendors are often judged on cost alone, which results in minimum sizing for the design crudes and rates. Carefully consider whether long-term crude trends will soon render these desalters inadequate.

CORROSION

In grassroots design, be realistic about metallurgy. Because modern refineries do not run a steady diet of the same crude, consider the sulfur and TAN numbers of potential crudes outside the unit's design blend. Chronic corrosion issues, or the inability to process high-margin opportunity crudes, will quickly overshadow the initial savings from choosing too low of a metallurgy.

VACUUM COLUMN WASH ZONE COKING

Wash zones are not for fractionation, they are for de-entrainment! Pursuing fractionation efficiency by specifying a deep bed with small crimp packing is a recipe for rapid coking. The correct choice of packing combined with the right wash rate and good distribution will properly de-entrain while preventing coke formation.



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cracking catalyst in the HC section can be optimised by selecting a higher cracking catalyst with more selectivity toward naphtha/kero. Thus, there are several opportunities in the combined HC-PT and HC process to increase gasoline make by leveraging the latest catalytic technologies.

Similar considerations apply to FCC operations. The pretreat section (FCC-PT) can be run at higher severity and/or with an improved catalyst system to increase desulphurisation and hydrogenation activity to reduce nitrogen and aromatics slip to the FCC unit. Depending mainly on pressure, feed properties and unit limitations, various catalytic solutions can be applied. Albemarle's latest generation FCC-PT catalysts to maximise performance in the FCC process include KF 780 for low and medium pressure operations and KF 905N, KF 907, KF 861 and KF 870 for medium to high and high pressure operations. At the same time, a highly gasoline selective catalyst can be applied in the FCC unit to increase gasoline production. This can be achieved by high rare earth, high matrix, high accessibility technology as can be found in Albemarle's Upgrader, Amber, Opal, Coral and other FCC products. The FCC unit can also be operated at an optimised severity/conversion to increase naphtha make.

Finally, to maximise gasoline make, cut points for both the FCC and hydrocracker products can be modified to increase naphtha and decrease the mid-distillate fraction.

A Alexis Shackelford, Technical Marketing Specialist, BASF Refining Catalysts, alexis.shackelford@basf.com

The FCC unit is well designed to convert gas oils and resid to high octane gasoline. There are three main controls on a FCC unit that can increase gasoline and lighter products (lighter products used for alkylate production): catalyst activity, reactor temperature (ROT), and catalyst circulation rate. Maximum gasoline octane barrels are produced by optimising these variables until unit conversion is limited. First, catalyst activity should be targeted at a high level, but not too high to limit over-cracking. Gasoline over-cracking, caused by excessive catalyst activity, is illustrated by the laboratory data in **Figure 1**. Reactor temperature is the least effective way to increase gasoline yield, but is the most effective way to increase gasoline octane. Increasing ROT also has the benefit of increased cat to oil ratio, which will further increase conversion. Operating at the highest reactor temperature up to the wet and/or dry gas production limit will produce maximum octane.

After ROT and activity are established, increase catalyst circulation by reducing feed preheat until the FCC reaches an air blower limit or LPG constraint. Longer term options to increase gasoline include selecting the proper catalyst with low delta coke and good pore architecture such as BASF's NaphthaMax III catalyst to maximise gasoline production. Also, ensure optimum rare earth content of the catalyst. A catalyst with a higher rare earth content will give higher gasoline selectivity, but at an octane penalty. Your catalyst

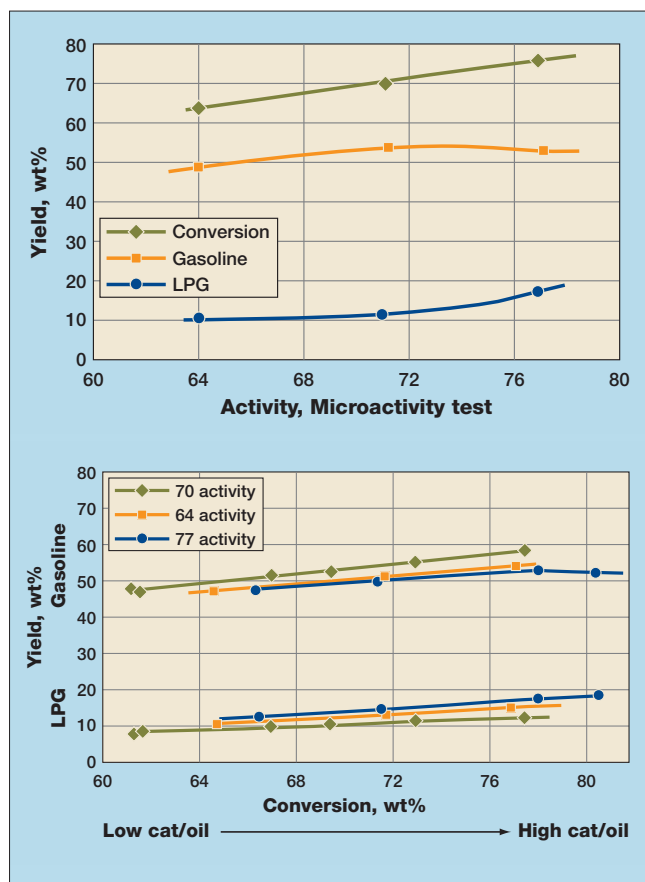


Figure 1 Gasoline over-cracking caused by excessive catalyst activity (top), and optimum activity (bottom)

supplier will be able to help determine the best fit for your unit.

Reference

Optimum FCC conditions give maximum gasoline and octane, *The Catalyst Report*, Engelhard 1994.

A Kate Hovey, Senior FCC Technical Service Engineer, Johnson Matthey, Kate.Hovey@matthey.com

There are many operating variables on the FCC that can be manipulated on a day-to-day basis, which gives this unit the ability to be much more flexible than other units of a refinery. The problem is that the FCC is normally operating at one or more constraints, which limits the level of flexibility available.

The most common initial response for maximising gasoline make is to raise the conversion by increasing the cat to oil ratio. This can quickly be achieved by increasing the riser outlet temperature; however this can be limited by dry gas make and wet gas compressor throughput. Other ways to increase conversion though increased cat to oil include reducing the feed preheat temperature or increasing the catalyst cooler duty (if possible). Lowering the feed preheat temperature can be limited by feed vaporisation, which is particularly important for units operating with atmospheric residue as part of the feedstock blend. Additionally, increasing cat to oil ratio can be limited by catalyst circulation rates and slide valve differential pressures.

Aside from the unit operation, the feedstock blend

and quality will have an impact on the FCC product slate. However, this is typically not something that can be adjusted easily so will be discounted in this discussion as a process step for increasing gasoline make.

The circulating Ecat properties and selectivity play the biggest role on the product slate and many FCC units that have been targeting a maximum distillate operation will have a fresh catalyst suited to maximum bottoms upgrading with a high matrix surface area. Reformulation of the fresh catalyst is always an option and fresh catalyst suppliers should be able to propose their most suitable catalyst for the current FCC objectives; however changing the fresh catalyst is not a fast response action to take. Once the chosen fresh catalyst formulation has been made, it first needs to be delivered to site and, depending on the size of the FCC unit inventory and typical fresh catalyst make-up rate, the changeover can take a few months, not to mention the consumption of fresh catalyst stock already held in the fresh catalyst hopper.

Delays resulting from changing the fresh catalyst formulation and full circulating inventory can mean maximum unit profitability can be missed. Johnson Matthey's Intercat FCC additive, HI-Y, can quickly change the circulating inventory activity and selectivity to push more conversion and target increased yields in the gasoline range. This means the FCC product slate can easily be tuned to track the changes in market prices and allow maximum unit margin and total profit to be achieved.

HI-Y is a highly concentrated source of rare earth stabilised Y-zeolite that has the ability to quickly increase unit conversion. The key to its performance lies in the special synthesis technique used to manufacture the additive. Through manipulation of the pore diameters of the zeolite support structure during the manufacturing process, HI-Y provides a potent alternative for converting bottoms into gasoline and lighter products without the usual increase in coke and gas make.

For the fastest response to changes in mode of operation, HI-Y can quickly change the Ecat selectivity towards maximisation of gasoline. Additionally, its physical properties, including particle size distribution, attrition resistance, and bulk density, have been optimised to be fully compatible with all commercially available FCC unit designs and catalyst technologies.

A final point for consideration is the flexibility to switch the selectivity back in the opposite direction if the market prices dictate. A highly attractive quality of using an additive such as HI-Y is that the FCC selectivity can easily be switched in either direction, without the need for a full fresh catalyst reformulation.

A **Jeffrey Zurlo, Marketing Leader – Refining, GE Water & Process Technologies, Jeffrey.Zurlo@GE.com**

As refineries have complex operations, with variation in unit processes that produce gasoline blend components, there are many answers to this question so a difficult one to answer definitively without knowing

the particulars of your plant. As such, I will provide a few items to consider for shifting refinery yields from diesel to gasoline, outside of considering capital outlay for new operating units.

- **Switch crude diet to increase gasoline range materials produced** The Economics & Planning group for your refinery/company should be able to provide guidance on crude oil purchases and crude oil blend recipes that can be used to promote gasoline production. This could come from direct increases in naphtha production at the crude distillation column and/or from increases in components that ultimately produce gasoline components deeper into the refinery process. An example of the latter would be increased gas oil feed rate or sending some vacuum residuum to the FCC unit to produce more FCC gasoline. Of course, when considering new crude oils or crude blends, the potential impact on operations and reliability should be considered. Changes in crude blend stability, contaminant loading, types of contaminants, and contaminant distribution into the refinery operating units could have negative impacts on desalter performance, overhead corrosion reliability, equipment fouling, catalyst life, and/or wastewater quality when changing crude oil blends. Another potential issue at the crude unit is shifting the heat recovery balance of the cold and hot exchanger train that could impact desalter temperature and create limits in product rundown temperatures. An 'eyes open' approach to understanding the full impact on operations should be part of the analysis when considering significant crude blend changes.

- **Alter crude distillation tower operation to maximise naphtha yield** As there is quite a bit of flexibility in the distillation ranges for straight run naphtha vs light and middle distillate streams, some of the kero/jet or diesel cut can be 'lifted' up into the naphtha cut to increase gasoline production. There are several operational handles available including increased tower temperatures in the naphtha and distillate sections of the tower, control via altering pumparound rates, reduced tower pressure, increased stripping steam to the tower to reduce partial pressures and promote additional 'lift'. As above, the impact of the operational changes should be well understood in terms of operations, reliability and operating costs.

- **Maximise conversion in the cracking units** Modifying feeds and operations to the cracking units that produce gasoline – such as the FCC unit, coker, hydrocracker, and visbreaker – is another common technique for increasing gasoline yield. Typically, this can have impacts on catalyst requirements or shorter catalyst life, shorter operating runs, and more frequent heater decoke cycles so the economics should be balanced with changes to unit performance expectations.

As a final cautionary note, when production levels are altered, one also needs to keep an eye on the gasoline blend recipes, as well as product specifications, as both can require changes to keep the blended gasoline quality on specification and meeting all the requirements for pipeline and finished fuel properties



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Q We are getting organochlorides in some product streams and have traced this back to the catalytic reformer. Causes and prevention measures please.

A Andrea Fina, Process Development & Marketing, CHIMEC, process@chimec.it

An increasing number of refineries are experiencing corrosion and fouling issues, due to the presence of severe spikes of chloride in the crude distillation unit (CDU) feed.

Chlorides are usually removed via a desalter, nevertheless there are some types of chlorides known as 'non-extractable chlorides' or 'non-desaltable chlorides'.

These 'non-extractable chlorides' can be inorganic (inorganic chloride salts encapsulated in high melting point waxes or flocculated asphaltenes, due to blend incompatibility) and/or organic (chemical cleaning solvents, halogenated biocides, filming amine hydrochlorides, amine chloride polymer coagulant, waste-chlorinated oil [PCBs from transformers] and so on).

When organic chlorides pass through the CDU heater, they partially undergo hydrolysis, thus forming acid chlorides that concentrates in the water of the main fractionator as it condenses in the overhead system.

The non-hydrolysed organic chloride portion, depending on the boiling range, usually travels up the atmospheric tower as vapour and dissolves in the naphtha as it condenses. Presence of organic chloride in the naphtha promotes fouling and corrosion issues in the downstream naphtha hydrotreating (NHT)/reforming complex.

Moreover, in a catalytic reforming unit with activated alumina based chloride traps on the export hydrogen line, there is a possibility of forming organic chloride as the alumina is approaching its end of life. It is thought that the alumina promotes formation of organic chloride from inorganic chloride on the surface. The alumina reacts with acid chloride to form aluminium chloride that promotes polymerisation of olefins (gums and green oils). It is also thought that the Lewis acid ($AlCl_3$) can promote the formation of organic chloride from the reaction between olefins and HCl.

As a result, the hydrogen, which may contain periodic high chloride content, introduces further organic chlorides in the NHT.

When organic chlorides pass through the NHT reactor, acid chloride is formed. In the reactor effluent system, due to the cooling, the acid chloride can dissolve in any condensed water droplets, thus forming an aggressive and corrosive environment, or can precipitate as ammonium chloride salt, by reacting with the ammonia gasses present in the reactor effluent stream before the water dew point.

Ammonium chloride deposit creates fouling and, being hygroscopic and acidic, under-deposit corrosion in the coldest feed/effluent heat exchangers and in the air cooler located before the high pressure separator.

A common practice to solve this issue is to inject water before these equipments to dilute the acidity and to wash out precipitated salts.

The water washing programme can be both on an intermittent or continuous base, depending on the nitrogen-to-organic chloride content.

Nevertheless this water wash, if not properly neutralised or due to insufficient quantity, often leads to high corrosion rates, caused by a very low pH at the injection point and throughout the system to the high pressure separator.

During the water wash procedure, it is very important to intensify the monitoring and controls to avoid salt reallocation in the downstream reforming unit: when chloride salts are removed from the reactor effluent system, they must be completely discharged with the tail water from the high pressure separator.

The water wash programme must be set in a proper way:

- The water injection point has to be placed before the water dew point or salt point and has to be designed to achieve proper mixing with the reactor effluent stream.
- The wash water amount should be enough to have 20-25% of liquid water at the injection point conditions (pressure, temperature, and so on)
- The neutraliser used to neutralise the acidity has to be characterised by a low salt point, the lowest solubility in naphtha (to avoid salt reallocation in the downstream reforming unit), low vapour liquid equilibrium constant (VLE), and the salt formed between the chloride and amine must have a low melting point. Chimec can boast many successful experiences in contrasting organic chloride issues in the naphtha hydrotreater and provides a complete portfolio of solutions, such as:
 - 'Top View' simulating software able to design the water wash programme (injection point, wash water amount, water dew point and NH_4Cl salt point)
 - Proper neutralising amines characterised by low salt point, negligible solubility in naphtha, low vapour liquid equilibrium constant (VLE) and low melting point.
 - Reforming catalysts are continually dosed with organic chloride compounds such as ethylene dichloride (EDC) to promote the acidic sites that enhance isomerisation reactions and improve reformate quality. These chloride compounds decompose to hydrochloric acid (HCl), which then reacts with ammonia present from the small amounts of nitrogen compounds in the reformer feed. The resulting formation of ammonium chloride (NH_4Cl) leads to fouling and corrosion once process temperatures fall to the salt point or below. Since the organic chlorides tend to 'slip' through the catalyst bed, they can create downstream problems such as poisoning of metal based catalysts and reformate specification issues. Continuous feed of the chloride compounds is necessary due to the reversible bonding of chloride on the catalyst surface, so some chloride will always be found throughout the process. Keeping this contamination to a minimum can be achieved by closely monitoring and controlling the organic chloride addition rate at stoichiometric requirements. The use of traditional fouling and corrosion

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inhibitors, such as neutralising and filming amines, is usually contra-indicated due to the nitrogen present in these products. So, to further reduce the threats of fouling and corrosion, the chloride compounds must be collected in strategically located reactor guard beds loaded with activated alumina or other chemical absorbents. This technology is well developed and available from suppliers who should be contacted prior to pursuing this solution.

A Alec Rockwell, Technical Service Engineer, Johnson Matthey, Alec.Rockwell@matthey.com

Chloride issues outside of the crude unit typically originate from the catalytic reformer unit. Catalytic reforming catalysts are dosed with chlorides to contribute to activity. These chlorides are not permanently bound, meaning that they are free to leave with the product streams at low ppm levels. Traditionally, the focus on preventing operational problems from the chloride compounds in the catalytic reformer product streams was to remove HCl. However, more recently a growing concern for many refinery operators has become the removal of organic chloride species as these compounds are less easy to detect, measure, and are less readily removed by many adsorbents. Increased organic chloride content is linked to increasing catalytic reforming severity and has been validated with field testing results.

Strong acid function, available chlorine, and olefins/aromatics are the constituents necessary to form organic chlorides. In the catalytic reformer, the metal function is provided by platinum or multi-metallic formulations, which are permanently held onto the catalyst. The acid function is provided by the chlorides that are dosed onto alumina to condition the catalyst. These chlorides are less permanently bound and are therefore available to leave in the product streams at low ppm levels. If untreated, these chlorides, containing both inorganic (HCl) and various organic chlorides, cause operationally related problems. These problems include formation and deposition of ammonium chloride, chloride related corrosion, poisoning of downstream catalysts, and product specification issues. As a result, many operators install chloride guard beds to remove the chloride.

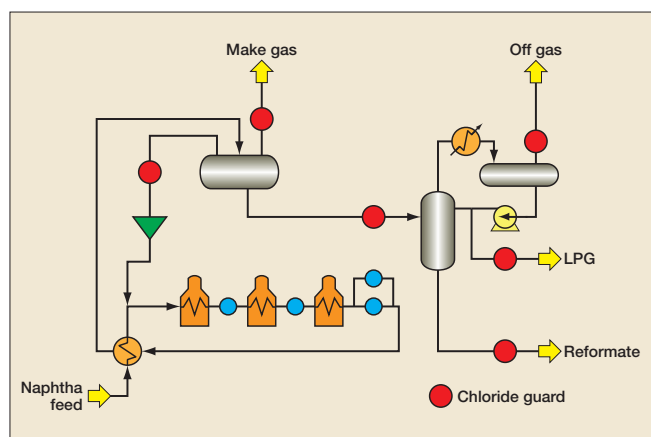


Figure 1 Possible locations of chloride guard beds in a catalytic reforming unit

Chloride guard beds are an adsorbent and/or absorbent that can be installed to remove chlorides and thus prevent operational issues. Depending on the material employed, these beds can be designed to remove both HCl and organic chlorides down to <0.1 ppm. The location(s) of the chloride guard bed(s) will depend on the operator's issues and preferences. **Figure 1** shows possible locations of chloride guards in a typical catalytic reforming unit.

In order to remove organic chlorides, chloride guard products that are capable of total (inorganic and organic) chloride removal must be selected. This can be achieved by selecting a bimetallic product formulation that is capable of total chloride removal by itself, or by customising a split bed design that is capable of removing inorganic and organic chlorides in separate layers. In this set-up, products such as zeolite or specialised alumina are utilised as a small layer to remove the organic chlorides.

It is important to monitor the inorganic and organic chloride concentrations surrounding the unit in order to confirm effective removal and plan bed changeouts. Organic chlorides are more difficult to detect using traditional testing methods, and therefore testing is not routinely conducted. For gas phase applications, Johnson Matthey offers the only methodology to detect inorganic and organic chlorides accurately to <0.1 ppmv. Online, onsite, or coordinated options are available for liquid phase duties.

A Jim Trigg, Senior Applications Engineer, GE Water & Process Technologies, Jim.Trigg@GE.com

Reforming catalysts are continually dosed with organic chloride compounds such as ethylene dichloride (EDC) to promote the acidic sites that enhance isomerisation reactions and improve reformate quality. These chloride compounds decompose to hydrochloric acid (HCl), which then reacts with ammonia present from the small amounts of nitrogen compounds in the reformer feed. The resulting formation of ammonium chloride (NH_4Cl) leads to fouling and corrosion once process temperatures fall to the salt point or below. Since the organic chlorides tend to 'slip' through the catalyst bed, they can create downstream problems such as poisoning of metal based catalysts and reformate specification issues. Continuous feed of the chloride compounds is necessary due to the reversible bonding of chloride on the catalyst surface, so some chloride will always be found throughout the process. Keeping this contamination to a minimum can be achieved by closely monitoring and controlling the organic chloride addition rate at stoichiometric requirements. The use of traditional fouling and corrosion inhibitors, such as neutralising and filming amines, is usually contra-indicated due to the nitrogen present in these products. So, to further reduce the threats of fouling and corrosion, the chloride compounds must be collected in strategically located reactor guard beds loaded with activated alumina or other chemical absorbents. This technology is well developed and available from suppliers who should be contacted prior to pursuing this solution.

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Q We are experiencing severe corrosion attack in our atmospheric tower fractionation trays. How can we deal with this?

A Collin Cross, Center of Excellence Leader – Refinery Corrosion, GE Water & Process Technologies, Collin.Cross@GE.com

Problems in the internal portions of a naphtha section are often caused by combinations of operational variables and corrosive chemical species present in the crude. When the production of heavier side-cut products is selected, problems can begin to occur at higher temperatures. Operational changes such as kero, diesel or gas oil production can push salt formation, both inside the tower and sometimes into distillate side circulations. Modern crude oils often come with a changing slate of tramp amines and hydrochloric acid liberating salts. While chemical corrosion control programmes are commonplace, utilising inhibitor chemistries such as neutralisers and filmers, frequently problems can be exacerbated by various combinations of operational, mechanical or crude related factors.

The first step towards solving such a corrosion problem is to understand the specific factors driving the corrosion. Once the factors are outlined, they can be addressed on a prioritised basis. New techniques are available that can be used to diagnose, maintain and improve visibility into integrated chemical, operational and engineering based drivers for such problems, but also allow their optimisation and engineering improvements over time to improve unit flexibility in these regards. By using combinations of chemical, monitoring and sophisticated data processing techniques, situations leading to corrosion can be computed and identified on a dynamic basis. Such visibility and response time allows root drivers to be identified, controlled and optimised in a proactive way. The benefit of such techniques is that they enable improved operational flexibility while allowing the refinery to enjoy an improved ability to make distillate products with less overall impact on overhead reliability. Until a unit is diagnosed, however, the specific combinations of drivers manifesting the corrosion are often not obvious.

While corrosion problems might be complex, or even subtle, they are always the result of specific, measurable and controllable events. The goal of a corrosion control provider is to provide chemical and service based solutions based on rapid and quantitative data analysis. This approach makes it possible to more quantitatively outline the cost/benefit relations for available options which best fit changing and individualised needs.

A Vivek Srinivasan, Sr.Engineer, Technical Services, Dorf Ketal Chemicals, viveks@dorfketal.com

Corrosion attacks are not a common phenomenon in fractionator trays, but when corrosion occurs it is likely due to reduced top temperatures, poor desalter performance, high amine concentrations and/or use of high acid crudes. The desalter is the first line of defence and

if performance is compromised, severe dew point corrosion can result from hydrolysis of salts to corrosive hydrochloric acid (HCl). In addition to dew point corrosion, ammonia/tramp amines react with HCl to form salts, causing potential under deposit corrosion. Both types of corrosion can be effectively treated in the overhead with a combination of chemical treatment and operating practices.

The situation is worse when the corrosion shifts inside the column, especially in top trays when operating conditions are pushed beyond limits to achieve specific production product mix targets.

Examples of scenarios conducive to corrosion in fractionator trays include:

A. Maximising jet fuel is a common strategy to increase refining margins. Increasing jet fuel production can require significant reduction of top temperatures, which can result in breaching the ammonia/tramp amine salting temperatures. Sources of ammonia/tramp amines are under-performing sour water stripper feed to the desalter and opportunity crudes. These basic amine species partition to the crude phase, depending on the wash water pH in the desalter, and ultimately land in the atmospheric column. The ammonia/amine and chloride content governs the salting points and hence risk of corrosion.

B. Top pumparound/naphtha pumparound designs are effective ways to reduce the overhead vapour load; however, depending on the crudes processed and operating conditions, this section and associated circuits/exchangers are prone to salt deposition and hence under-deposit corrosion.

C. Maximising aromatic feedstock: in some integrated aromatic refinery complexes, the atmospheric columns are operated at lower temperatures to reduce the naphtha FBP for aromatic feedstock, which may breach the ammonia/amine salting point, leading to corrosion.

D. Water ingress in the column: stretched operating parameters impacts oil-water separation in the overhead reflux drum, causing water/moisture carry-over in the reflux back to the column. Eventually, amine/ammonia salts are washed from the drum and accumulate in the top tray via water carry-over in reflux, causing corrosion and the potential for overhead fouling. Refiners around the world target <500 ppm moisture content in the reflux.

E. Naphthenic acid introduced by processing high acid crudes plays a major role in high temperature oil phase corrosion. This can potentially occur in high temperature/high TAN sections, including trays and associated piping at temperatures in the range 220-350°C.

Corrosion risk mitigation strategies include:

(i) Breaking the ammonia/tramp amine cycle: mitigating/reducing the ammonia/tramp amine partition in desalter helps overcome the salting and hence corrosion issues in column trays. Acidification of the desalter is a current practice. However, use of acid to control desalter pH requires capital investment, is difficult to control, and is vulnerable to upset in the desalter and at the wastewater treatment plant. There is the added risk of corrosion, salt deposition, and

operating safety problems with storage and handling of the acid. Dorf Ketal has introduced new 'reactive adjunct' chemistry added to the desalter wash water to supplement the emulsion breaker. The chemical reacts with tramp amines, reducing desalter pH and allowing the amines to be efficiently removed in the desalter as imines. The reactive chemistry also improves metals and solids removal, reduces rag layer, and improves brine quality. The cost for the reactive chemistry is offset by reducing the dosage of the emulsion breaker. The reactive adjunct chemistry increases operational flexibility for jet fuel production with reduced corrosion risk.

(ii) Choice of metallurgy: commonly used MOCs like CS, 300 Series SS, Duplex SS and Monel cladding in locations such as overhead, column top trays and dome are not effective against ammonium chloride salt corrosion (see Refining Industry Damage Mechanism, Section 5, under 5.1.1.3). Use of alloys like Inconel is more suitable for resistance to such corrosion.

(iii) Use of top pumparound corrosion inhibitors helps overcome corrosion inside the column and associated pumparound exchangers.

(iv) Use of robust monitoring tools to determine dew point and ammonia/amine salting point temperatures keep a check on the corrosion rates. Dorf Ketal's proprietary Ionic Equilibrium Model (IEM) helps estimate the dew point, amine/ammonia salting point and selection of suitable amine chemistry for overhead neutraliser.

(v) Naphthenic acid corrosion; use of high temperature corrosion inhibitors (HTCI) will help form a protective barrier on the metal surface. Conventional phosphate ester based inhibitors are highly acidic and require very high phosphorus loading, which is a source of potential phosphorus fouling. Dorf Ketal's Tancsient is a breakthrough in high temperature corrosion inhibition. It is a phosphate triester, which means it is non-acidic. It includes a custom synthesised alcohol, which has superior solubility and thermal stability. This advanced chemistry produces safer, more reliable high temperature corrosion inhibition with significantly reduced potential for phosphorus fouling.

Q What is the role of dispersants in preventing exchanger fouling?

A **Xiomara Price, Center of Excellence Leader – Fouling Control, GE Water & Process Technologies, Xiomara.Price@GE.com**

Fouling precursors that exist in fluids as particles, whether they are organic or inorganic in nature, will agglomerate. As the particle size increases, the potential for precipitating out onto the surface of the metal also increases. This deposition rate is not only impacted by particle size, but also by other variables such as fluid velocity and viscosity. Dispersants contain polar and non-polar ends. The polar end interacts with the particle and the non-polar end interacts with the fluid. Enough of the dispersant molecules have to surround

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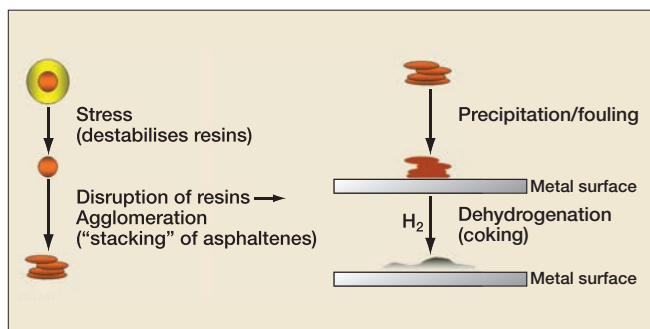


Figure 1 Destabilisation/agglomeration of asphaltene

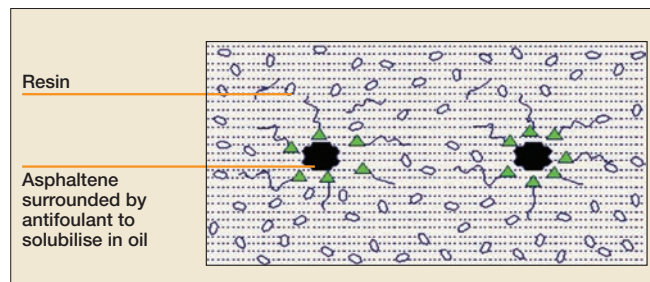


Figure 2 Organic dispersant

the particle to keep it from agglomeration with other particles. It is important to have a high enough concentration of the dispersant in the fluid because it becomes tied up by the particle with which it is interacting. Since the particle cannot grow through agglomeration, it will be easier to move it through the system by the velocity of the fluid. If sufficient chemistry is not present to interfere with the agglomeration of enough particles, the rate of deposition will not decrease and the fouling rate will not be mitigated.

It is important to point out that some dispersant molecules work best with organic materials and some with inorganic particles. Additionally, certain dispersants can interact with the metal surfaces creating a barrier and facilitating the movement of the particles through the system.

A Parag Shah, Deputy Technical Director, Technical Services, Dorf Ketal Chemicals, paragshah@dorketal.com

The main precursors for exchanger fouling can be divided into organic and inorganic. Organic precursors include asphaltene; inorganic includes corrosion products, metals and salts. At high temperatures organic precursors, especially asphaltenes, phase separate from resins, agglomerate with other species, and settle on metal surfaces to cause fouling (see **Figure 1**). Inorganic precursors form nucleating sites for additional organic deposition to occur.

Dispersant chemistries are selected as a function of the type of precursors and application. Organic dispersant chemistries are used for asphaltene related fouling. These organic dispersants keep the destabilised asphaltene dispersed to prevent agglomeration, which would result in deposition on the metal surface (see **Figure 2**).

Inorganic dispersants are used to form a steric barrier, which limits the size of the particle and the

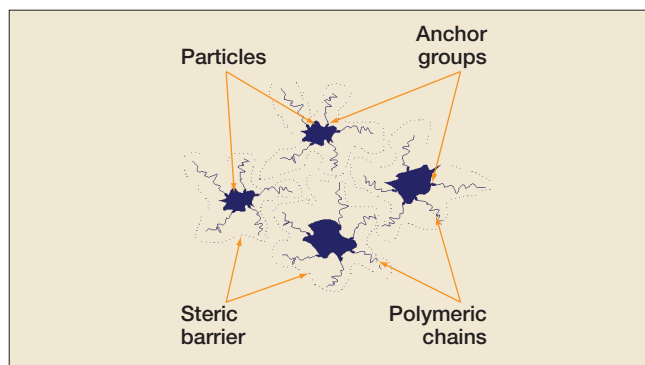


Figure 3 Inorganic dispersant

resulting deposition. This is achieved by the inorganic dispersant containing an anchoring group, which attaches itself to the inorganic particle, and a polymer chain, which acts as an oil soluble tail, repelling it from other particles/metal surface (see **Figure 3**).

A Marcello Ferrara, Chairman, ITW, mferrara@itwtechnologies.com

Normally, dispersants have the role of preventing insoluble coke and polymer-like material agglomerating into large particles, which can settle out of the process stream and adhere on the surface of process equipment. Dispersants are believed to absorb the insoluble materials in organic fluid on the surface, converting them to a colloidal suspension. They can delay, in some cases, the fouling mechanism, which leads heavier components to go out of solution, but they cannot inhibit the fouling mechanism and, most importantly, they have no effect on fouling that has already formed.

Fouling can be better addressed by dissolving the same. ITW Online Cleaning is capable of solubilising, not dispersing, the fouling that settles on equipment surfaces and of transforming the same into a reprocessible product. This is achieved by means of proprietary chemicals and process steps.

Online Cleaning can be applied proactively, whenever required, to clean heat exchangers from a fouled stated. This will allow the unit to run under cleaner conditions and avoid losses arising from fouling, both in refinery and petrochemical units.

Far more importantly, ITW Online Cleaning can be applied to clean an entire process unit in 24 hours on a feed-out/feed-in basis, without the need for deinventorying the unit. After Online Cleaning, the unit will immediately resume operation.

Q Can you advise how to minimise hydrocarbons in our tank water draw-off before it reaches the treatment plant?

A Jim Trigg, Senior Applications Engineer, GE Water & Process Technologies, Jim.Trigg@GE.com

This question could have several answers depending on which 'tank' water draw-off is creating the problem. The most common area of concern involves draining water from crude oil tanks since that draw

usually goes directly to a wastewater treatment plant, potentially creating a hydrocarbon load on the primary separation section (API separator or similar equipment). Since the incoming crude oil contains water in varying amounts, there is really no opportunity to limit the water quantity. Thus, implementing source control requires several components to achieve minimal hydrocarbon contamination. First, a distinct oil/water interface needs to be created by having sufficient residence time for the water droplets to coalesce and settle. This process can be accelerated using robust demulsifier chemicals similar to those used in crude oil desalting. Product selection, addition locations and dose rates can be quite specific to each location and crude oil quality, so engaging a reputable specialty chemical company is advisable. Second, the tank operator must determine the interface level in relation to the tank drain. That is a fairly simple operation known as 'water-cutting', by which a flexible string is coated with a water finding paste (WFP). The paste turns colour (usually to red or green) in the presence of water and permits the operator to accurately gauge the water level. Third, knowing the water height allows the operator to drain the tank reliably using a 'drain' rate, which can be estimated by line size and draw-off valve opening. Since the last operation still involves some guesswork, tank operators need to become proficient in visually assessing the cleanliness of the draw, a learned skill, which can be taught and mastered by draw sampling and testing to create a visual library of hydrocarbon content.

A **Marcello Ferrara, Chairman, ITW,**
mferrara@itwtechnologies.com

Water usage in the hydrocarbon processing industry is of primary importance due to large scale consumption and related wastewater generation. Hydrocarbon entrainment in water is a quite common problem during processing and storage of hydrocarbons.

A high content of hydrocarbons in the water delivered to the wastewater treatment plant (WWTP) will have a serious impact on the WWTP's performance since the unit is designed to handle a defined maximum amount of hydrocarbons.

Near or above such a maximum allowed target, the bacteria will be killed and the WWTP will be put out of service. The hydrocarbons need therefore to be separated from water before this reaches the WWTP.

Besides some physical separations means, the easiest way of implementing such separation is using reverse emulsion breakers. ITW has developed a specific reverse emulsion breaker to quickly separate hydrocarbons from water. This can be added at the inlet of the API separator in order to facilitate separation of oil from water. The clear water phase can be sent to the WWTP as it will not contain hydrocarbons.





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Developments in hydrotreating catalyst

How a second generation hydrotreating catalyst was developed for high pressure ultra-low sulphur diesel units and hydrocracker pretreaters

MICHAEL T SCHMIDT
Haldor Topsoe

The worldwide market demand for more active NiMo hydrotreating catalysts is extraordinary. Despite tremendous improvements in catalyst technology for the past 20-30 years, ultra-low sulphur fuel legislation and the shift towards VGO hydrocracking to maximise diesel production are more than ever forcing refiners to search for the absolute top tier NiMo catalyst for their ultra-low sulphur diesel (ULSD) or hydrocracker pretreat reactor.

The dramatic drop in natural gas prices observed particularly in the US has resulted in a low production cost for hydrogen. The low cost of hydrogen makes it very profitable to add hydrogen catalytically to middle distillate fractions, thereby increasing the liquid volume swell, resulting in higher yields of valuable diesel.

To address these requirements, refineries continuously need better, yet cost-efficient, alumina based catalysts with the highest possible activity in order to obtain the desired boost in performance and hydrogen consumption. Furthermore, alumina based hydrotreating catalysts will help minimise the operating cost when targeting volume swell in comparison with higher cost unsupported catalyst formulations.

Continuously improving alumina based catalysts

Topsoe has been at the forefront of important technological breakthroughs in the hydroprocessing industry for decades and continuously explores the possibilities

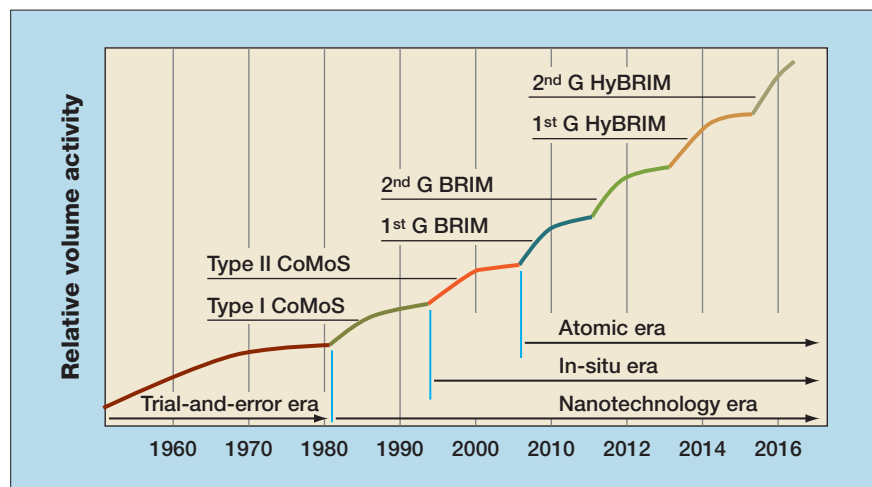


Figure 1 Topsoe's catalyst development progress – from Type I through second generation HyBRIM

within this area. In the 1970s, Topsoe discovered the active CoMoS phase in hydrotreating catalyst, which revolutionised the catalyst world by applying fundamental research in the catalyst development approach. In the 1980s, Topsoe researchers, headed by Dr Henrik Topsøe, discovered the difference between Type I and Type II hydrotreating catalysts and gave them the names that are known throughout the industry today. With this breakthrough, the hydroprocessing catalyst development entered the nanotechnology era, and the Type II hydrotreating catalysts became the industry standard for high-activity catalysts. In the early 2000s, Topsoe's research in surface science paid off again, and a new activity site was discovered: the BRIM site. With this finding, Topsoe developed BRIM technology within both the CoMo and NiMo type catalysts, which fuelled Topsoe's unparalleled growth in market share

globally due to a top tier catalyst portfolio.

From BRIM to HyBRIM

Topsoe's latest catalyst technology, HyBRIM, involves an improved production technique for both NiMo and CoNiMo hydrotreating catalysts. It combines the BRIM technology with a proprietary catalyst preparation step. The synergistic effect of merging the two technologies has enabled Topsoe to design an advanced metal slab structure that is characterised by an optimal interaction between the active metal structures and the catalyst carrier. The activity of the Type II sites is positively influenced to a high degree by this interaction between the metal slab and the carrier. HyBRIM technology exploits this interaction and substantially increases the activity of both the direct sites and the hydrogenation sites without compromising the catalyst stability.

Topsoe's NiMo catalysts that are

developed today are around three times more active than the catalysts produced in the 1990s. **Figure 1** illustrates the development of the company's many catalyst generations. Since Topsoe's scientists employed tools such as electron microscopes, in situ monitoring, and high throughput screening in their research programmes, we have made considerable progress within hydrotreating and hydrocracking catalyst development. The BRIM and HyBRIM catalyst technologies are the direct outcomes of this scientific approach.

HyBRIM technology was originally introduced with Topsoe's TK-609 HyBRIM in 2013 and has since been extended to include several different hydrotreating catalysts covering medium to high pressure refinery applications. As



Figure 2 Image of TK-611 HyBRIM

mentioned above, the need for ULSD production from low quality crudes and higher severity hydrocracking is creating a need for even better catalysts.

During the past three years, HyBRIM technology has been broadly recognised by the industry

to be at the forefront of what is possible within hydrotreating. In addition, more than 100 hydrotreating units around the globe have a HyBRIM catalyst installed right now. However, Topsoe's researchers recently discovered even more potential within the HyBRIM method – a potential utilising the active metals to an even higher extent and securing a dispersion of active sites to a level never seen before. Therefore, Topsoe is launching the second generation of HyBRIM technology – the TK-611 HyBRIM catalyst with 25% higher activity for both sulphur and nitrogen removal.

Higher activity, same stability

When applying the TK-611 HyBRIM catalyst in either ULSD or hydrocracker pretreatment service, the improved activity can be exploited in several ways. Obviously, a higher activity is often used to operate the unit as in previous cycles, only at a lower reactor temperature, which then yields longer cycle lengths. However, the new high activity can also be utilised in terms of increasing unit throughput or, in the case of a hydrocracker, to lower the nitrogen slip from the pretreat section to the cracking section, resulting in higher conversion and better yields. In addition, TK-611 HyBRIM will increase the volume swell due to better hydrogenation functionality. Refiners also benefit from purchasing more opportunity crudes or processing more LCO and they will have a stronger and robust catalyst candidate to handle these more challenging feedstocks. In any case, TK-611 HyBRIM will significantly improve the profitability of refinery assets.

Figure 3 shows ULSD pilot plant testing and compares TK-611 HyBRIM with TK-609 HyBRIM side-by-side. It is seen that if the two catalysts are operated at exactly the same conditions and at a reactor temperature giving 10 wt ppm product sulphur for TK-611 HyBRIM, then TK-609 HyBRIM will simultaneously deliver a product with 32 wt ppm sulphur. Hence the activity advantage of TK-611 HyBRIM corresponds to a delta

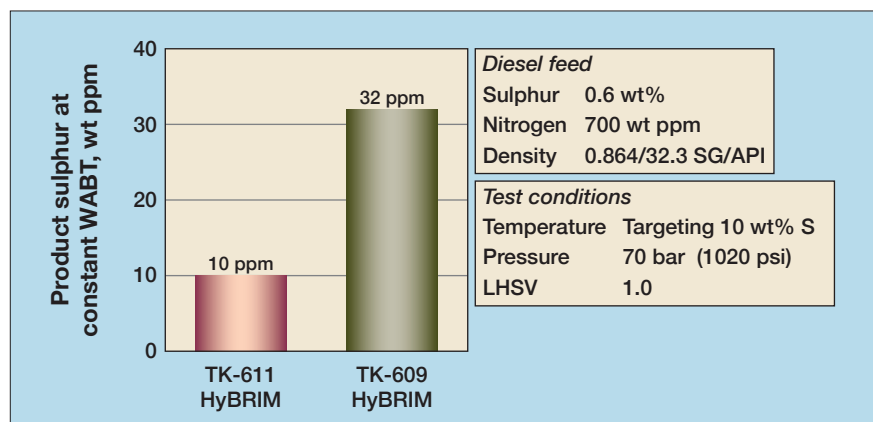


Figure 3 Comparing TK-611 HyBRIM with TK-609 HyBRIM in ultra-low sulphur diesel service

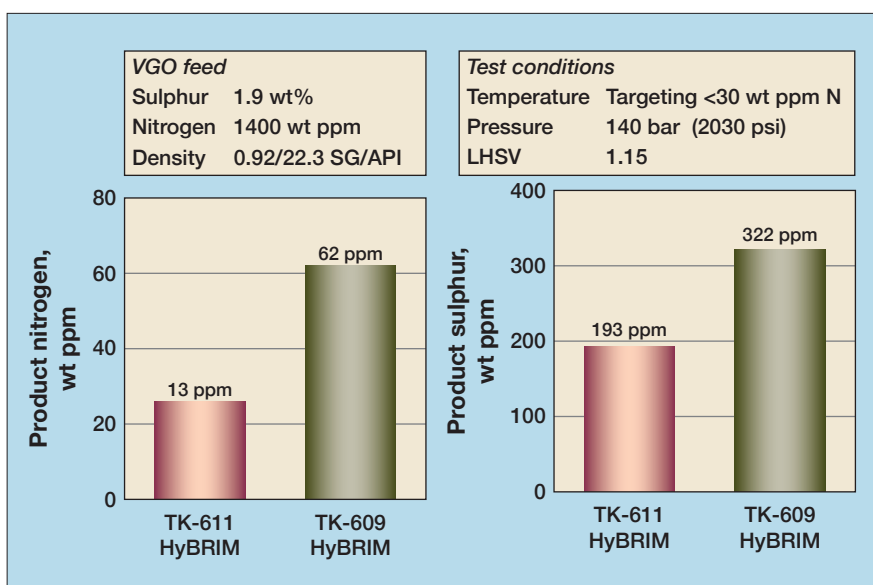


Figure 4 Data for hydrocracker pretreatment service for TK-611 HyBRIM versus TK-609 HyBRIM

product sulphur of 22 wt ppm at ULSD conditions, which is a remarkable step-change in activity.

The same type of experiment is illustrated in **Figure 4** for vacuum gas oil (VGO) at hydrocracker pretreatment conditions. While 62 wt ppm product nitrogen slip is achieved with TK-609 HyBRIM, TK-611 HyBRIM, at exactly the same conditions, is able to deliver a product nitrogen slip of only 26 wt ppm. Such a difference is a substantial improvement for a hydrocracker pretreating unit.

High start-of-run activity is obviously important; however, activity has no real meaning unless it is accompanied by high catalytic stability, ensuring that improved performance is maintained over the projected cycle. Some catalyst formulations on the market display an impressive fresh activity; however, due to the nature of these catalysts, there is an initial line-out deactivation caused by the nitrogen species present in the feed passivating the most active sites. When these types of catalysts are installed, they will typically have lost their activity benefits after four to six weeks on-stream.

During the development of BRIM and subsequent HyBRIM technologies, Topsoe has successfully been able to modify the alumina structure and catalytic surface. By employing scanning electron microscopy (SEM) and transmission electron microscopy (TEM) techniques, Topsoe's researchers have observed how the catalyst preparation steps influence the catalyst functions. This knowledge has led to an improved alumina pore structure and an optimised interaction with the alumina support, providing very active and stable CoMoS/NiMoS catalyst formulations. The data shown in **Figure 5** compares the stability of the new TK-611 HyBRIM with TK-609 HyBRIM in VGO hydrocracker pretreat service at the same operating conditions. The testing reveals that even though TK-611 HyBRIM is operated at higher sulphur and nitrogen conversion levels, due to its higher activity, the two catalysts exhibit exactly the same performance stability.

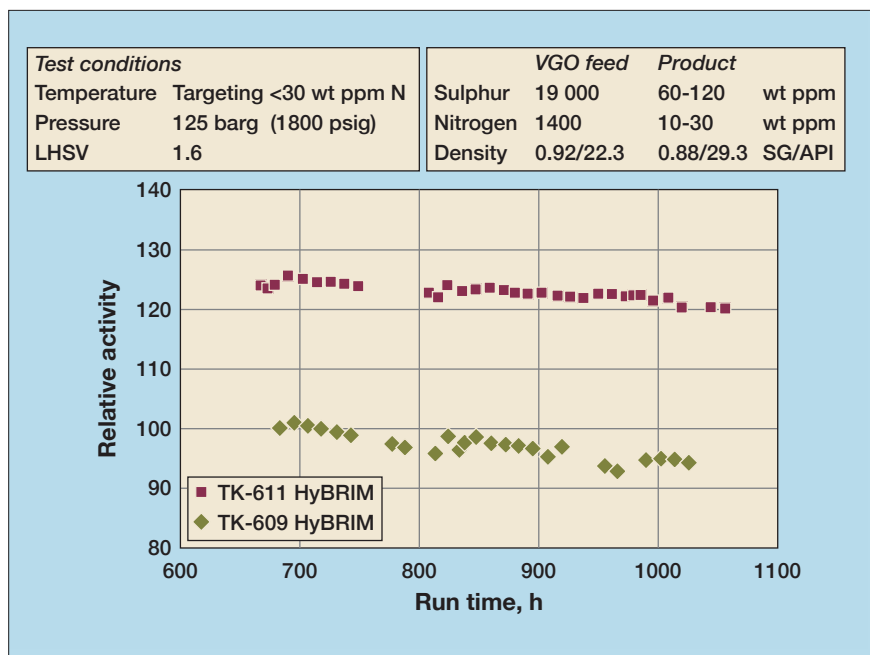


Figure 5 Catalyst stability for TK-611 HyBRIM versus TK-609 HyBRIM in hydrocracking pretreat service

More volume swell

Introducing as much hydrogen into the distillate fractions as possible can, as previously mentioned, be very profitable when excess hydrogen is available. The term 'volume swell' refers to the increase in the liquid volume when the product density and distillation are lowered by hydrotreating. This includes contributions from removing sulphur, nitrogen (only a very small contribution), and hydrogenation of olefins. However, the main contributor is without doubt the saturation of poly-, di-, and mono-aromatics.

Topsoe has previously demonstrated that there is a strong correlation between the observed density improvement and the degree of aromatic saturation (%HDA). Aromatics saturation is controlled by the catalyst's hydrogenation activity, the presence of inhibitors, such as nitrogen species, and by thermodynamic equilibrium. The overall amount of hydrogen consumed by hydrodearomatization (HDA) reactions is of course also dependent on the actual amount of aromatics present in the feed (see **Figure 6**).

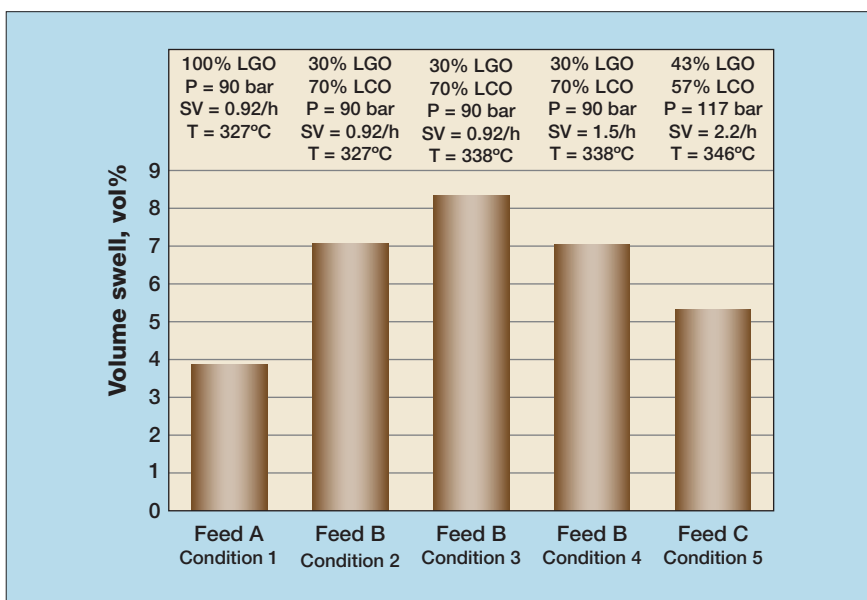


Figure 6 Obtained volume swell with HyBRIM catalyst in a LCO hydrocracker

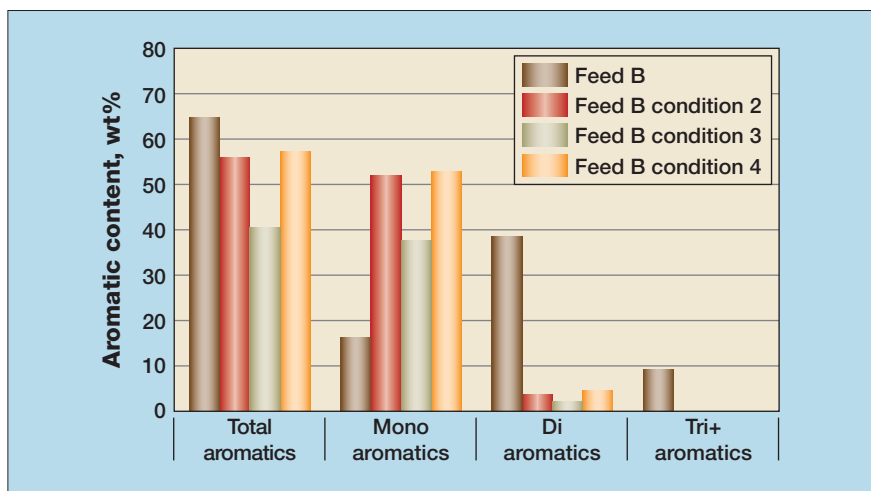


Figure 7 Aromatics content in feed and products in Feed B case

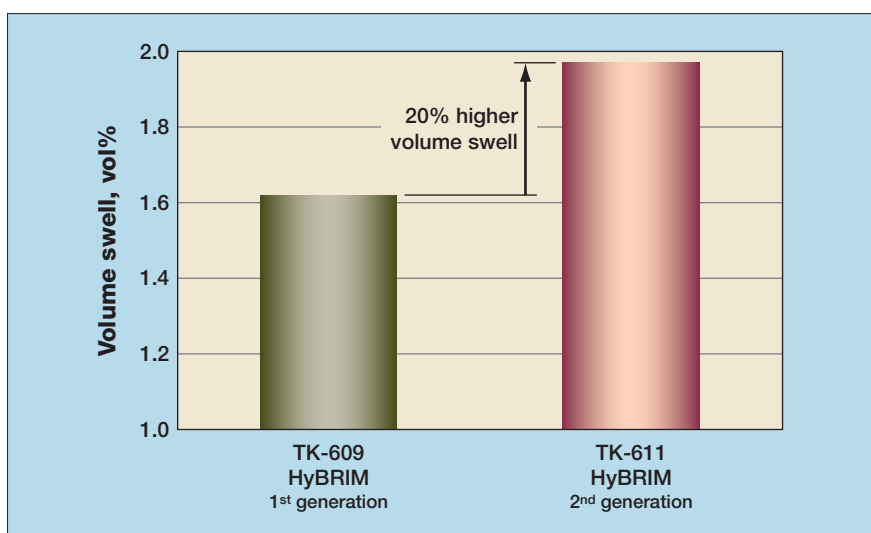


Figure 8 Volume swell comparison of TK-611 HyBRIM versus TK-609 HyBRIM at ULSD conditions

A simple way to quantify the volume swell is to compare the liquid product density with the feed density. This is an easy approach and is based on available data; however, it does not take the yield losses into account. The correct way is therefore to compare the increase in the C_5+ volume yield with the fresh feed. This method is accurate; however, these data are not always readily available from commercial units or from simple pilot tests, as it requires fractionation and the ability to close the mass balance.

Topsoe's researchers have previously published that the presence of nitrogen and, in particular, basic nitrogen compounds strongly inhibits HDA reactions. Therefore it is beneficial to remove nitrogen to very low levels, lower than 2–3 wt

ppm N, in order to increase HDA and the associated volume swell. Consequently, TK-611 HyBRIM with maximum activity for nitrogen removal (HDN) will also provide the highest volume swell.

A simple way to quantify the volume swell is to compare the liquid product density with the feed density

Figure 6 illustrates the effect on volume swell when changing the operating conditions and feed blend. The pilot plant test was conducted

simulating a LCO hydrocracker pretreat unit using a HyBRIM catalyst as the main treating catalyst. It included five different conditions, where pressure, temperature, space velocity, and LCO amount in the feed blend were varied. The results show that the HyBRIM catalyst removed the nitrogen to very low levels, below 0.2 wt ppm N, which is the detection limit. Consequently, this indicates that saturation of aromatics and, in particular, monoaromatics should take place in the last part of the reactor. In this pilot plant test, the temperature has been kept in the low range in order to avoid thermal cracking. As expected, Feed B, being very rich in aromatics due to the high LCO content, obtained the highest volume swells at test conditions 2, 3, and 4.

In Figure 7, the corresponding aromatics content from Feed B is plotted. It is seen that test condition 3, where the highest volume swell is reached, is also where the most aromatics are saturated. This can be explained by the higher reactor temperature and low space velocity. Monoaromatics especially are saturated the most at condition 3, and there is actually a correlation between the obtained volume swell and the residual content of monoaromatics. In refinery terms, the highest volume swell observed with Feed B corresponds to gaining more than 3300 b/d of additional liquid out of a 40 000 b/d unit. This is remarkably high in conventional hydrotreating without any cracking catalyst and is the result of a high degree of aromatic saturation. This volume swell represents an additional profit of about \$40 million per year even after subtracting the cost of the extra hydrogen that is consumed in the hydrotreater. The value is based on an oil price of \$45/bbl, and profitability will obviously increase with the increasing cost of crude oil.

As part of the data shown in Figure 3, we also obtained volume swell data for the two catalysts, enabling a direct comparison of the hydrogenation and hydrodearomatization difference between first and second generation HyBRIM cata-

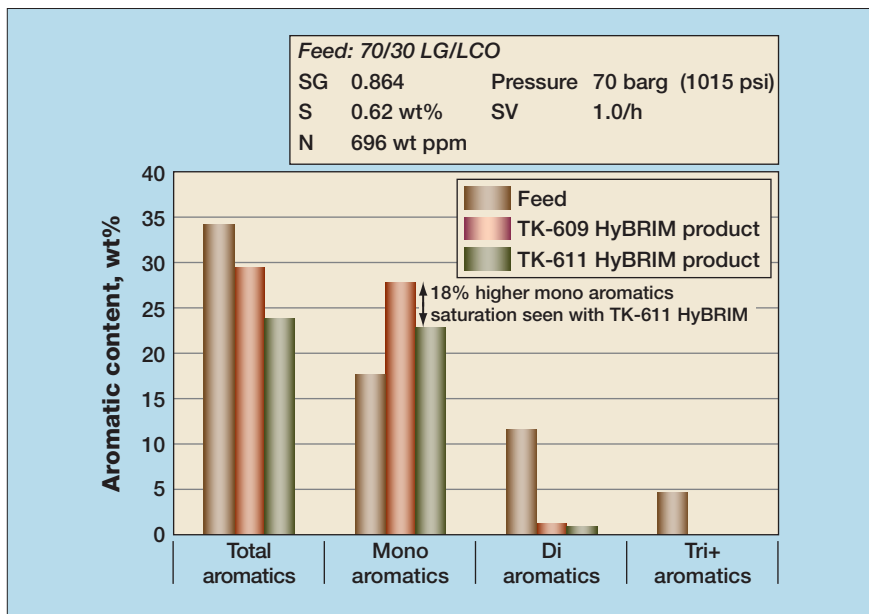


Figure 9 Aromatics content before and after ULSD hydrotreating with TK-611 HyBRIM and TK-609 HyBRIM

lysts at the same reactor conditions. In **Figure 8**, the difference in volume swell based on density reduction is depicted. Despite a quite low feed aromatic content, TK-611 HyBRIM delivers 20% higher volume swell compared to TK-609 HyBRIM at ULSD conditions. This is a significant difference, and the improvement is remarkable because TK-609 HyBRIM is already a high volume swell catalyst.

Figure 9 reveals the main reason for the 20% higher volume swell. At ULSD conditions, the product nitrogen for both catalysts is low and actually below the detection limit. This means that the 'monoaromatics saturation high-way' is fully open with the nitrogen inhibitors removed. At such conditions, the hydrogenation potential of the catalyst is providing the volume swell, and it is therefore observed that TK-611 HyBRIM is saturating 18% more monoaromatics than TK-609 HyBRIM.

Conclusion

The second generation of HyBRIM technology is now available with the introduction of TK-611 HyBRIM catalyst. The 25% higher activity of TK-611 HyBRIM compared to first generation TK-609 HyBRIM catalyst unlocks additional flexibility to obtain longer catalyst cycles, more throughput, better product qualities

or even processing of more severe feedstocks. The activity boost can be translated into lowering the reactor temperature significantly in hydrocracker pretreating and ULSD units, while simultaneously obtaining the same conversion of sulphur and nitrogen. Equally important, it is established that second generation HyBRIM technology exhibits the same high stability that refiners have come to expect from the BRIM and HyBRIM catalyst series. It has been demonstrated that TK-611 HyBRIM catalyst has significantly higher hydrogenation activity than TK-609 HyBRIM catalyst due to its higher activity for nitrogen removal, an ability that will provide higher volume swells at similar conditions, resulting in substantially increased profitability for the refinery asset. In conclusion, the advantages of the new TK-611 HyBRIM can be utilised to improve the overall profitability and economy of all hydrocracker pretreating and ULSD units in multiple ways.

HyBRIM and BRIM are trademarks of Haldor Topsoe.

Michael T Schmidt is Product Manager with Haldor Topsoe. He is responsible for the development and quality of Topsoe's FCC pretreatment and hydrocracking pretreatment catalysts and holds a degree in chemical engineering from the Technical University of Denmark (DTU).

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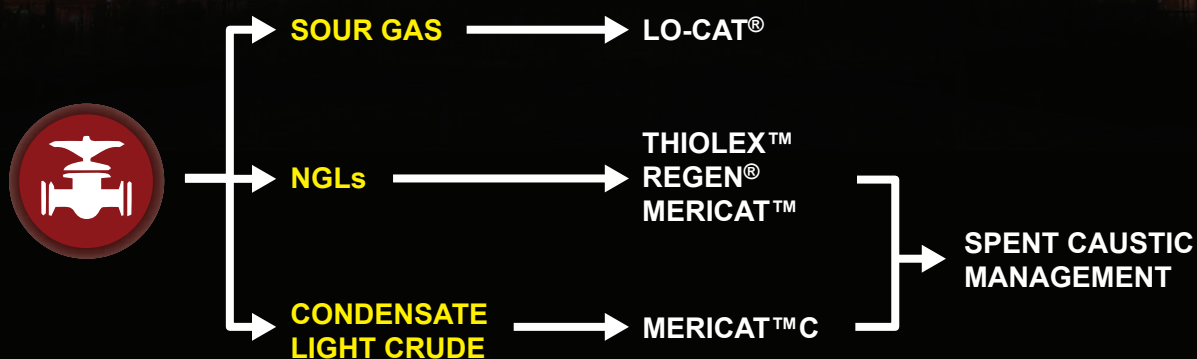
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Loss into gain in high-capacity trays

Part 2: reverse vapour cross flow channelling

Systematic troubleshooting diagnosed two complex problems that limited the capacity of an atmospheric crude distillation tower

HENRY KISTER *Fluor*

NEASAN O'SHEA (ret'd) and DAN CRONIN *Phillips 66 Whitegate refinery*

In the 1990s, Irish Refining Company operated a 52 500 b/d (7070 t/d) crude unit processing North Sea crudes (this refinery is now owned by Phillips 66). A detailed description of the crude tower and products is in Part 1 of this article (*PTQ*, Q2 2016).

Initially, the four-metre ID atmospheric crude tower contained 34 single-pass jet tab trays. At a later time, the tower was revamped by others to increase fractionation efficiency and to permit the refinery to intermittently take a kerosene side-cut while maintaining the same throughput of 52 500 b/d (7070 t/d). Trays 11-20, the trays in the heavy naphtha (HN) diesel fractionation section, were replaced by single-pass, high-capacity Nye trays. Structured packings were installed in the wash and stripping sections. A new kerosene side draw was installed at the inlet to Tray 16, approximately mid-way between the HN and diesel draws. A side stripper was added to strip the kerosene. Initially, the kerosene rundown system was incomplete so the kerosene side draw was not operated.

After the retrain, the tower was not able to achieve the 52 500 b/d (7070 t/d) it had previously achieved. At throughputs exceeding 38 000 b/d (about 5200 t/d), the tower flooded with a rise in pressure drop and the HN product going off-specification with heavy components. Initial gamma scans showed the flood initiated near Trays 14-16, building up until Tray 23. The pumparound Trays 21-23 were jet tab trays and were not changed in the revamp.

In addition, a severe instability in

the lower part of the column, which we termed an 'excursion', would usually occur at about the same time and rates as the tray flooding was initiated. This excursion was caused by the diesel draw nozzle and rundown line running into a self-venting flow limitation following a revamp modification that reduced the liquid residence time of the diesel draw pan. The diagnosis and correction of this problem are described in detail in Part 1 of this article. This Part 2 of our article

At throughputs exceeding 38 000 b/d, the tower flooded with a rise in pressure drop and the HN product going off-specification

focuses on troubleshooting and correcting the cause of the premature flood.

Channelling to explain the premature flood

Figure 1 shows the tower with the post-revamp mass balance and operating conditions at a reduced crude rate of 38 700 b/d (5200 t/d) in black print. Also shown on the diagram, in red print, are the mass balance and operating conditions following the operational debottleneck to 49 100 b/d achieved by taking a kerosene draw that will be discussed later in this article.

Trays 11-20 are high-capacity Nye trays² (see Figure 2). These trays differ from conventional trays by having a 'Nye unit' at the tray floor, right under the downcomer. The Nye unit is closed at the top, perforated at the side and underneath. Vapour entering the unit from the tray below via the bottom holes is turned horizontally towards the tray and exits via the side holes. This horizontal wind imparts a horizontal 'push', or horizontal momentum, to liquid droplets in the inter-tray space, blowing them towards the outlet downcomer and preventing their entrainment into the tray above. By reducing this entrainment, the horizontal push is the capacity enhancement mechanism on the Nye tray. The same horizontal push mechanism, produced by a variety of devices and geometries, is employed by most of today's high-capacity trays.

Figure 3 (roughly to scale) shows the tray dimensions. The active areas of Trays 11-16 contained long-legged venturi (smooth orifice) uncaged moving valves with 16.2% open area (the most restrictive open area of the fully open valve, expressed as a percent of active area), while Trays 17-19 contained small-diameter sharp-orifice moving valves with 14.6% open area. There were also small differences in downcomer and active area dimensions (see Figure 3).

Hydraulic calculations both by the vendor and the troubleshooting team (see Table 1, the column headed 'Before operational debottleneck') showed that at 38 700 b/d crude charge, the most highly

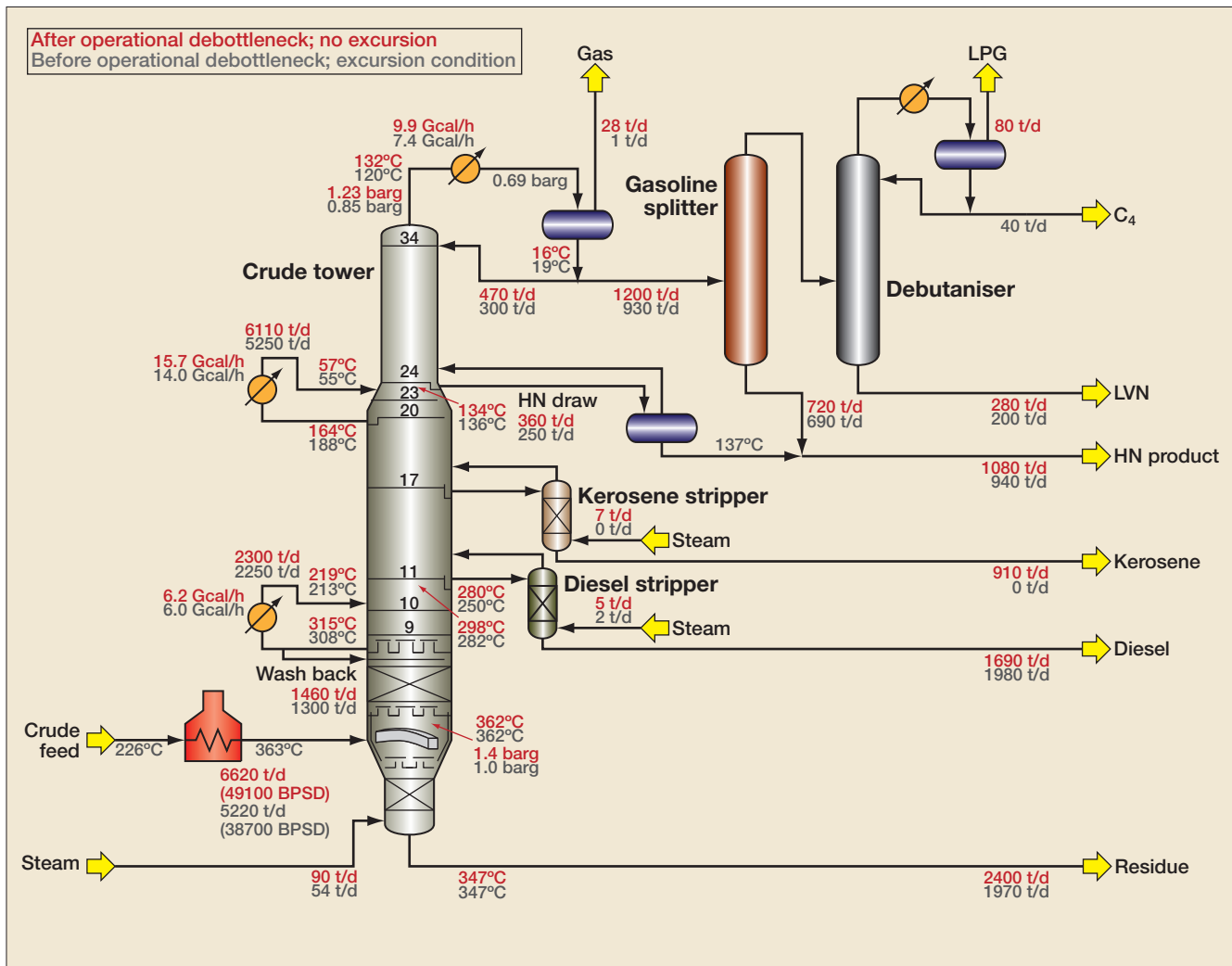


Figure 1 Crude tower mass balance

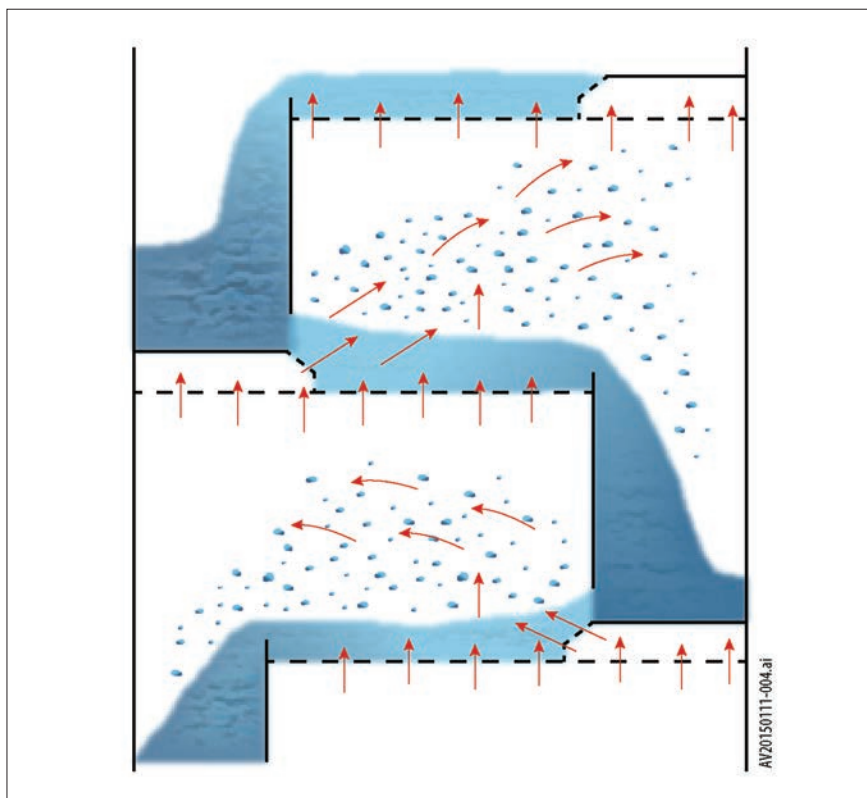


Figure 2 Schematic of a Nye tray

loaded trays above the diesel draw and below the top pumparound (Trays 11-20) should have been operating comfortably below any capacity limit. The calculations showed that at these loads even conventional trays should have operated without flood.

Eye-catchers in **Figure 3** and **Table 1** are the very long flow paths, giving a huge ratio (3.6-4) of flow path length to tray spacing, the very large weir loads (100-116 m³/h m of outlet weir), the venturi valves, and the large open areas (14.6-16.2%). Previous work showed that venturi valves, large open areas (>13-14% with sharp orifice moving valve trays), large ratios of flow path length to tray spacing (> 2:1), and high weir loads (>50-60 m³/h m of outlet weir) render trays prone to vapour cross flow channelling (VCFC).^{3,4} VCFC is encountered when, under the influence of a hydraulic gradient,

vapour preferentially channels through the tray outlet and middle, generating a high vapour velocity region with high entrainment and premature flood. At the same time, the tray inlet region remains vapour deficient, which promotes excessive weeping and poor tray efficiency.

VCFC is not the only form of channelling experienced on distillation trays. There are reports^{5,6} of other forms of channelling, such as due to vapour maldistribution or multi-pass tray maldistribution. One thing they have in common is that they have only been experienced at large tray open areas, high ratios of flow path lengths to tray spacing, and high weir loads, conditions that apply also to the trays involved in this case.

An argument against VCFC is that forward push was used as a cure to VCFC. In VCFC, the channelling is caused by an excessive hydraulic gradient.^{3,4,7} Using 'push valves', slots or valves that enter some of the vapour onto the tray with a horizontal velocity in the direction of the liquid flow, can reduce the hydraulic gradient and alleviate VCFC⁸ as has been demonstrated by one published case study⁹ and some others experienced by the authors. The Nye trays used in this tower entered about 10-15% of the hole area in a horizontal direction, providing considerable forward push. This forward push should have been sufficient to counter the hydraulic gradient.

It therefore became necessary to seek further insight into the froth pattern on Trays 11-20. Channelling is difficult, often impossible, to diagnose with conventional troubleshooting techniques such as vendor software, ΔP measurement, and conventional single-chord qualitative gamma scans. Judicious multi-chordal gamma scans with quantitative analysis is the best tool for diagnosing maldistribution on trays. The principles and practice of this technique were recently described in detail elsewhere.¹ This technique was applied here to give a concise definition of the trays' hydraulics and to diagnose the root cause of the premature flood.

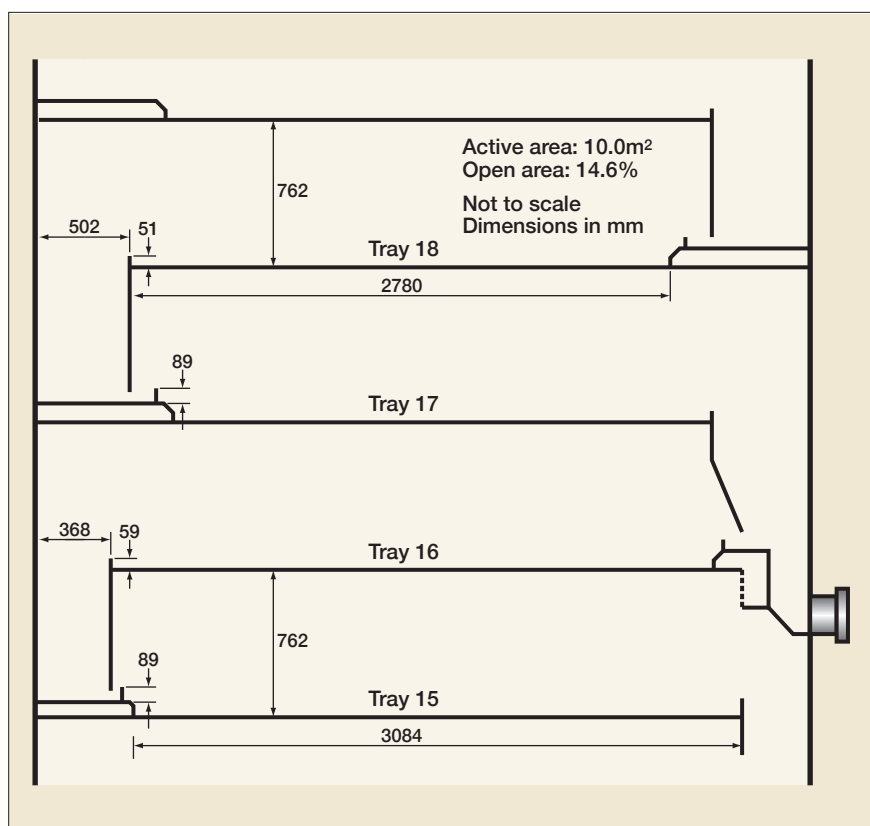


Figure 3 Dimensions of Trays 11-20

Internal flows and hydraulic evaluation, Trays 11-20				
Operation	Before operational debottleneck		After operational debottleneck	
Crude charge rate, t/d	5220 (38 700 BPSD)		6620 (49 100 BPSD)	
Kero drawn?	No		Yes	
Tower top pressure, bar g	0.85		1.23	
Tray number	16	18	16	18
Moving valve type	Venturi	Small-valve	Venturi	Small-valve
Temperature, °C	213	203	238	203
Vapour rate, t/h	232	225	245	261
Liquid rate, t/h	172	171	134	188
C-factor**, based on active area, m/s	0.091	0.097	0.092	0.107
Weir load, m³/h m outlet weir	116	100	89	110
% Jet flood, modified FRI	78	78	74	86
% Jet flood, Glitsch (eq.13)*	82	86	78	95
% DC backup	40	45	37	51
DC inlet velocity, m/s	0.128	0.082	0.098	0.088
Dry pressure drop, mm liquid	44	51	45	63

* Not accounting for capacity enhancement by Nye units

** $C = U_a \sqrt{\frac{\rho_g}{\rho_L - \rho_g}}$ where U_a is the vapour velocity based on the active area, m/s, and ρ_g and ρ_L are vapour and liquid densities in consistent units.

Table 1

Multichordal gamma scans with quantitative analysis

Four gamma scan chords were shot through the active areas of Trays 9-22 parallel to the downcomers. In addition, two gamma scan chords were shot perpendicular to the downcomers. These perpendicular chords were intended to pass

through the active areas only, and to miss the downcomers. All chords were shot when the tower was just below flood at 37 500 b/d charge rate ('incipient flood') and repeated when the tower was flooded at 44 300 b/d charge rate. During the scans, column operation was kept extremely stable and steady. This

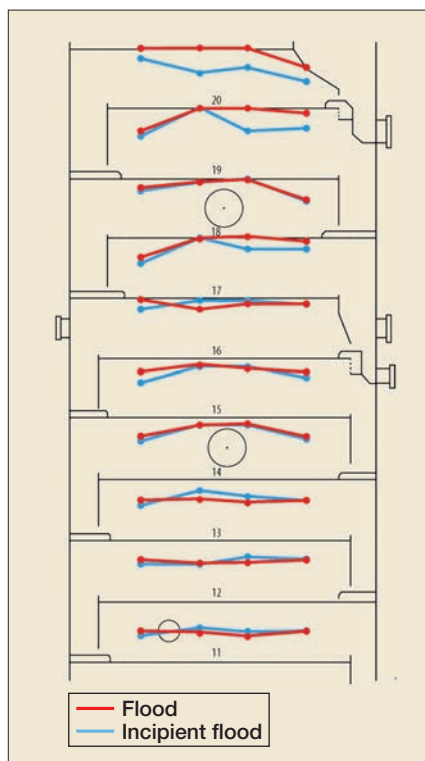


Figure 4 Froth heights, Trays 11-20. Blue, incipient flood; red, flood

was vital as for the scans to be consistent with one another no change could be allowed to the process during the shooting period. At only one time instability was experienced for a short time, and the data obtained during this instability were discarded.

Additional chords were shot through the downcomers and showed no downcomer bottlenecks. They were validated by downcomer neutron backscatter measurements. It was concluded that the problem was in the active areas, not in the downcomers.

Verification of data validity and consistency is a must for multichordal gamma scanning¹ as one set of bad data is sufficient to lead to an incorrect diagnostic. Several consistency checks were conducted. Some chords were repeated and were shown to produce near-perfect repeatability.

Over the three months preceding the tests, several scans were conducted, at both flooded and unflooded conditions, along the same or similar scan chords to those used in the tests. Those served as invaluable consistency checks. The agreement was nothing short of amazing, commending the

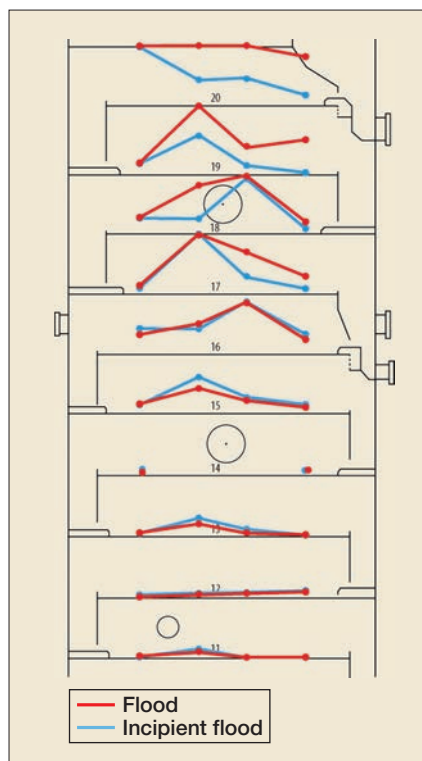


Figure 5 Entrainment index, Trays 11-20. Blue, incipient flood; red, flood

excellent work of the scanners. It also showed us that in this tower the flood state appeared to be uniform and stable.

Among the earlier scans, there were scans shot at 35 000 b/d under completely unflooded conditions, not even incipient flood. These were also analysed, and provided invaluable insight, in particular to the clear liquid heights and hydraulic gradient evaluation.

The two scans perpendicular to the downcomers and equidistant from the tower centre-line were near-identical to each other at incipient flood, demonstrating absence of channelling in this direction, and giving one more consistency check.

Gamma scan results

Scan peaks obtained in a quantitative gamma scan were analysed to provide froth heights, froth densities, clear liquid heights and an entrainment index. The procedures used were discussed in great detail elsewhere.^{1,7,10}

Multichordal quantitative analysis produces a considerable volume of data that are difficult to keep track of. These data need to be put together with the details of the tray

geometry in a way that provides a clear picture. We use the 'Kistergrams', which are elevation diagrams of the column and trays drawn to scale, with superimposed quantitative information derived from gamma scan analysis. The information plotted includes froth (or spray) heights (see **Figure 4**), entrainment (see **Figure 5**), froth densities (see **Figure 6**), and clear liquid heights (see **Figure 7**). The Kistergrams here use the following conventions:

1. All scales are linear.
2. The x-axis shows the exact position of the scan chords on the trays.
3. Froth height data are plotted at the same scale as the diagram. If the determined froth height equals or exceeds the tray spacing, it is plotted right on the tray above. Zero froth height is plotted as a point on the tray floor.
4. The froth density scale shows a froth density of 0.5 as a point on the tray above (the full tray spacing).
5. Clear liquid height is plotted so that a height of half the tray spacing is plotted as a point on the tray above. Zero height is plotted as a point on the tray floor.
6. An entrainment index of 2 (flood) is plotted as a point right at the tray above. An entrainment index of zero is plotted as a point on the tray floor.

Froth heights (see **Figure 4**)

Flood is indicated when a froth height point is plotted on the top of the tray spacing, at the floor of the tray above. **Figure 4** shows:

- Flooding took place only on Trays 16 and up. Trays 11-15 were not flooded.
- Froth heights tended to be low at tray inlets. In both the incipient flood and the flooded scans, the inlet froth heights were all below about 500-550 mm, the only exception being Tray 16 where it was about 700 mm. There was no flooding in any inlet region.
- At incipient flood, flood was approached at the middle of the trays, but not at the tray outlets. On Trays 16-19, the froth heights reached the tray spacing in the middle, but the froth heights



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remained about 100 mm below the tray spacing at the outlets. Only Tray 20 had a froth reaching the tray spacing at the outlet, but note the different shape of its inlet downcomer.

- Flood occurred in the middle of the trays.

Entrainment (see Figure 5)

The entrainment index (EI) is defined as the ratio of the gamma ray transmission count (as picked by the detector) at clear vapour, divided by the maximum transmission of gamma rays at the vapour space above the tray, less 1.0. If there is no entrainment, the gamma ray transmission at the vapour space above the tray would reach the transmission count at clear vapour, the ratio would be unity, and the entrainment index would be zero. In our experience, an entrainment index of 0.5 signifies light entrainment, 1.0 moderate entrainment, 1.5 heavy entrainment, and 2.0 or greater flood.

Flood is indicated when an entrainment index point is plotted on the top of the tray spacing, i.e., at the floor of the tray above. This point corresponds to an entrainment index of 2. **Figure 5** shows:

- Flooding took place only on Trays 16 and up. Trays 11-15 were not flooded and show low entrainment indices.
- Flooding took place only in the second quadrants of Trays 16-19, and in all but the inlet quadrant of Tray 20. In the incipient flood scans, entrainment from other quadrants was light to moderate (entrainment indices 0.1-0.8), indicating that these quadrants were not even close to flood. This provides strong evidence to support flood initiation in the second quadrant of the trays.
- Entrainment was light at tray inlets. In both the incipient flood and the flooded scans, the inlet entrainment indices were all below 0.5, the only exception being Tray 20 at the flood where it was about 1.6. There was no flooding in any inlet regions.

Froth densities (see Figure 6)

A froth density of 50% liquid is

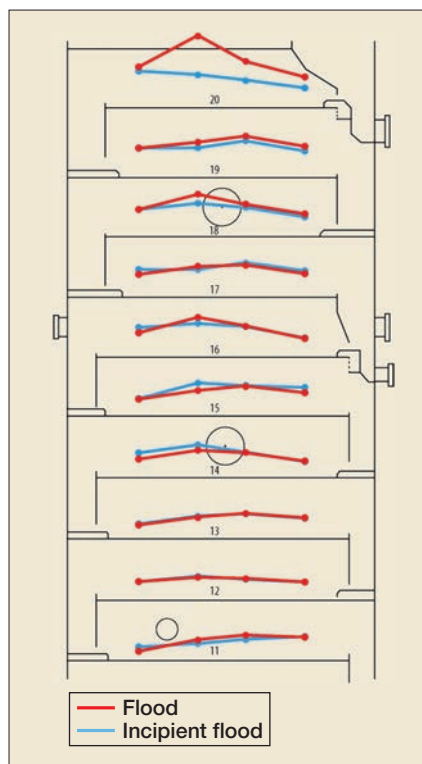


Figure 6 Froth densities.
Blue, incipient flood; red, flood

plotted on the top of the tray spacing, i.e., at the floor of the tray above. A point half way between two trays indicates a froth density of 25% liquid. **Figure 6** shows:

- Froth densities were low at the inlets and outlets, both in the flooded and the incipient flood scans, all in the range of 10-25%. The only exception is Tray 20 where the outlet froth densities reached 30% in both the flooded and incipient flood scans.
- Froth densities in the middle of the trays at incipient flood (16-19) were higher, in the range of 22-30%.
- Froth densities in the middle of the flooded Trays 16-19 rose from 22-30% in the incipient flood scans to about 26-38% in the flooded scans.

Clear liquid heights (see Figure 7)

With clear liquid heights, the unflooded scans provided invaluable insight so they were added to the diagram along with the flooded and incipient flood scans. A clear liquid height of 380 mm is plotted on the top of the tray spacing, at the floor of the tray above. A point halfway between two trays indicates a clear liquid height of 190 mm. **Figure 7** shows:

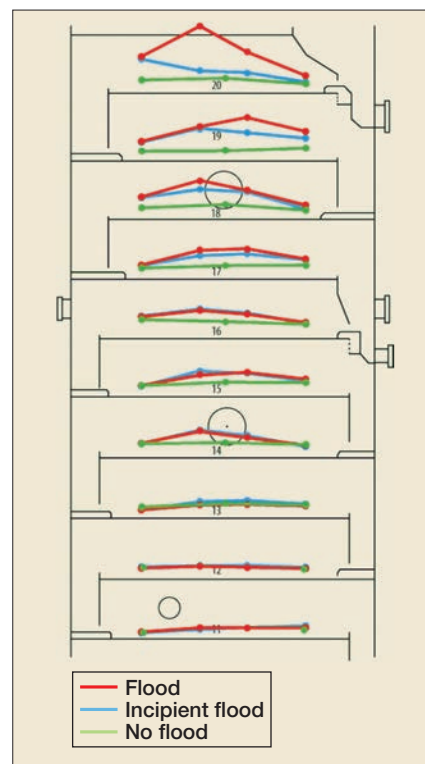


Figure 7 Clear liquid heights, Trays 11-20.
Green, no flood; blue, incipient flood; red, flood

- In the unflooded scans, the clear liquid heights on Trays 13-20 were 15-30 mm lower at tray inlet than at the tray middle and outlets. At incipient flood, the gradient upward from inlet towards the middle of the trays intensified. There was a large rise in clear liquid height in the middle of the trays, about 50-100 mm on Trays 13-19. This is an increase by a factor of 1.5-2. The rise was smaller, about 20-50 mm, at the Trays 16-19 outlets, and was zero for Trays 13-15 outlets. The inlet clear liquid height rise from non-flooded to incipient flood was small, of the order of 20-30 mm in Trays 17-19, and none on Trays 13-16.
- At incipient flood, clear liquid heights in the middle of Trays 16-19 reached 180-230 mm, and a little lower, 150-180 mm, on Trays 14-15.
- Going from incipient to full flood, there was little change in the clear liquid heights on Trays 13-16. On Trays 17-19, there were no major changes at the inlet, outlet and second quadrant, but the clear liquid heights on the third quadrant of Trays 17-19 and at the outlet of Tray 19 jumped by 80-100 mm.

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Reverse vapour cross flow channelling

Figure 8 shows how the hydraulic gradient on Trays 11-20 changes from non-flooded to flooded conditions. The hydraulic gradient plotted is the maximum clear liquid height on the tray less the minimum clear liquid height on the tray. We term this a 'reverse hydraulic gradient'. In all the scans, the minimum clear liquid height was always at the tray inlet. For the incipient flood and the flooded scans, the maximum on Trays 12-19 was always in the tray middle. For Tray 20 at incipient flood, as well as for many trays in the non-flooded scan, the maximum was at the tray outlet. So for the incipient flood and flood conditions on most trays, the reverse hydraulic gradient is the difference between the clear liquid heights at the middle and inlet.

Figure 8 shows that for non-flooded conditions, the hydraulic gradients were of the order of 20-30 mm. This is well in line with values reported from measurements on conventional trays⁷, except that on normally operating conventional trays the gradient is from inlet to outlet: clear liquid is highest at the inlet and lowest at the outlet. Here the gradient is reversed: the clear liquid is highest in the middle of the tray, and lowest at the inlet, so the hydraulic gradient is from the middle to the inlet. At incipient flooding, there was a sudden buildup of this reverse hydraulic gradient on Trays 14-20. This means a lot of liquid channelled towards the middle of the tray, raising the reverse hydraulic gradients to 80-120 mm. This is a huge hydraulic gradient. Going from incipient flood to full flood, there was no further change on Tray 16 and below. However, as liquid accumulated on Trays 17-20 above, the reverse hydraulic gradient increased by another 50-100 mm, indicating intensification of the channelled pattern. The resulting channelling pattern had vapour breaking through the inlet regions, with liquid building up in the middle of the trays. We term this

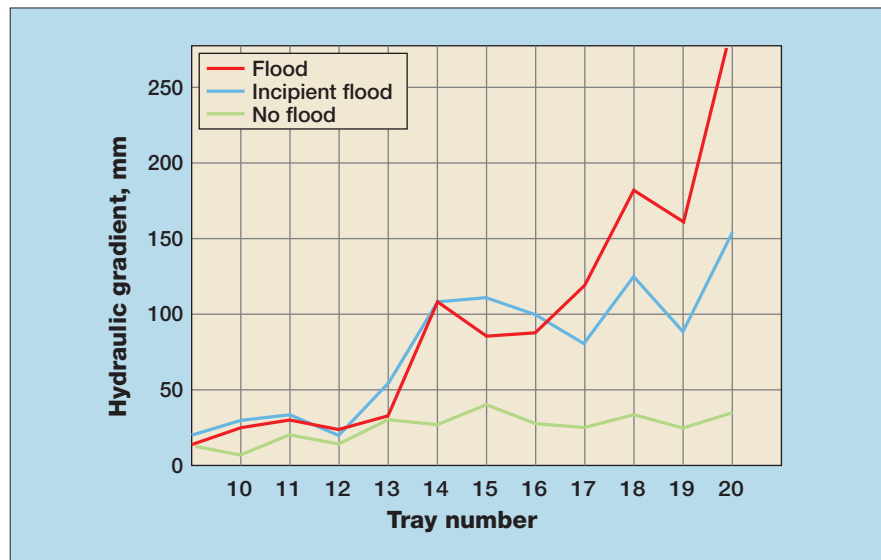


Figure 8 Reverse hydraulic gradients, Trays 11-20

type of channelling reverse vapour cross flow channelling (RVCFC).

Figure 9 assembles this information to construct the likely hydraulic pattern on the trays. The Nye unit at the inlet aerates the incoming liquid and shoots a lot of it forward, generating a low liquid head region near the tray inlet. The liquid from this region is blown to the middle of the tray, generating a high-liquid head region there. At low rates, this is likely to lead to excessive weep from the middle of the tray, which

may negatively impact tray efficiency and turndown. The efficiency loss with RVCFC is smaller than in vapour cross flow channelling (VCFC), because in VCFC the weep occurs from the tray inlet, which is much more detrimental to tray efficiency than the weep from the middle of the tray that occurs with RVCFC.

At higher rates, more vapour channels through the tray inlet, generating a high-velocity region at the tray inlet, with liquid blown towards the tray above and the

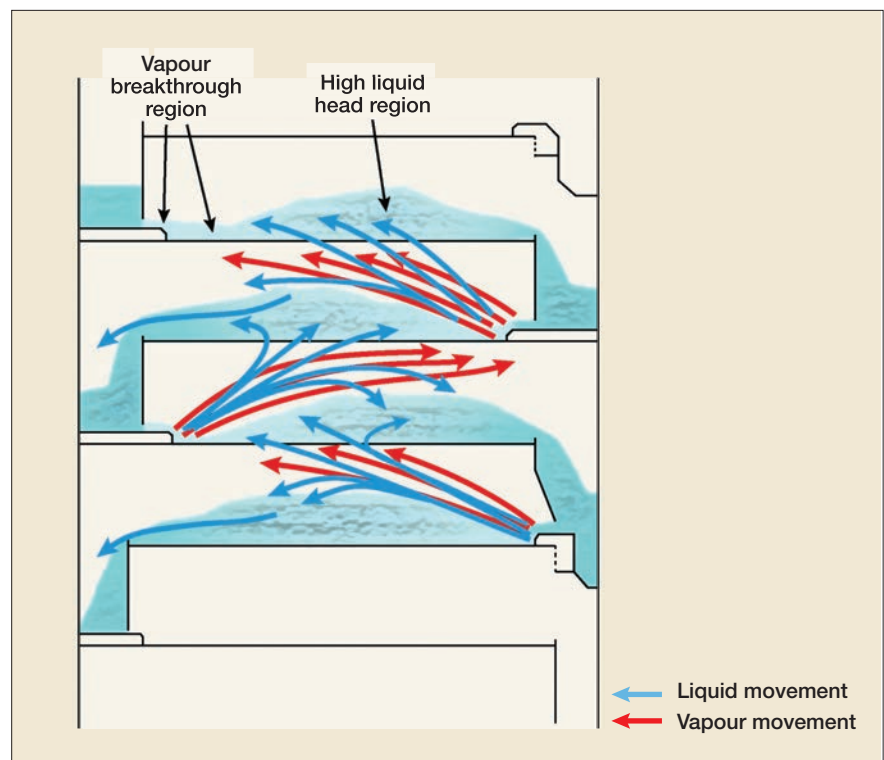


Figure 9 Reverse vapour cross flow channelling

middle of the tray. This shows up as increases in entrainment and froth height in the middle of the tray. Some of the blown liquid falls to the tray floor at the middle of the tray, generating higher liquid heads there. This channels more vapour to the tray inlet. This starts a self-accelerating mechanism which ends up with low froth heights, low liquid heads and low clear liquid heights at the tray inlet, and high froth densities, froth heights, entrainment and clear liquid heights at the tray middle. The result is premature flooding, as experienced in this crude tower.

Similar to VCFC, per the mechanism proposed by Davies¹¹ and later discussed for sieve and valve trays,^{3,4} the channelling tendency is countered by the dry pressure drop of the tray. When the dry tray pressure drop is high relative to the hydraulic gradient, this pressure drop will distribute the vapour evenly across the tray and prevent channelling from setting in. In the current case, Trays 11-16 had venturi valves, which have rounded orifices that halve the valve pressure drop compared to those with sharp orifices. Also, these valves had long legs, leading to excessive open slot areas, which also lowered dry pressure drop.

The channelling criterion proposed by Davies¹³ is that the hydraulic gradient needs to be kept below 40% of the dry pressure drop to avoid channelling. **Table 1** shows that Tray 16 has 44 mm of liquid dry pressure drop, so as long as the hydraulic gradient is less than about 20 mm, there will not be significant channelling, as seen in the unflooded scan. However, should the forward push action generate a hydraulic gradient exceeding 40 mm, it is likely that the dry pressure drop will not be large enough to counter the channelling, which will then initiate the self-accelerating mechanism. The best prevention is to boost the dry pressure drop.

The small valves on Trays 17-20 had sharp orifices and lower open areas, so they had higher dry pressure drop (**Table 1**) and were more resistant to channelling. However,

these are also lower pressure-drop valves and are also prone to channelling, albeit to a lesser extent.

Like in VCFC, long flow path to tray spacing ratios generate the geometry shown in **Figure 9** (these are drawn to scale), and promote RVCFC. Short flow path compared to tray spacing will not permit development of the RVCFC pattern. Finally, like in VCFC, high weir loads are likely to promote this type of channelling.

The RVCFC requires a source of forward push to generate this pattern. So while push valves have been effective in countering VCFC, with some geometries the forward push can generate its own channelling pattern. We have seen other cases of RVCFC. In one case,

The key to preventing RVCFC from forming is following the good design practices that prevent VCFC

RVCFC in a refinery fractionator pumparound caused excessive weeping, generating cavitation at the pump. In a recent case, RVCFC appeared to have combined with multi-pass trays maldistribution to give RVCFC in the middle panels of four-pass trays.⁶

The key, therefore, to preventing RVCFC from forming is following the good design practices that prevent VCFC^{3,4}: avoiding excessive open areas, venturi valves, and large ratios of flow path length to tray spacing with tray geometries that may be prone to channelling.

Operational debottleneck

The solution to the excursion problem in Part 1 also provided an operational debottleneck for the current flood problem. That solution was operating in the kerosene mode. Following the tests, the kerosene mode became the normal mode of operation. There was little economic penalty to drawing a relatively small amount of kerosene product stream, as it could have

been blended with the diesel when not desired as a product.

Operating at the kerosene mode circumvented not only the diesel draw bottleneck, which caused the excursion problem, but also the flooding bottleneck. With both bottlenecks overcome, the tower charge rates could be raised from 38 700 b/d (5200 t/d) to 49 100 b/d (6600 t/d), much closer to the design 52 500 b/d (7070 t/d). At 6600 t/d, the flood limit re-occurred but the excursion did not. At 6600 t/d charge rate, the capacity loss was tolerable and an imminent shutdown was avoided.

The red print in **Figure 1** shows the tower material balance following the operational debottleneck. The diagram permits comparison to the material balance before the debottleneck.

One puzzle is how drawing kerosene debottlenecked the flood in the tower. **Table 1** shows that drawing kerosene greatly reduced the liquid rate and weir load on Tray 16, where the capacity bottleneck was. However, channelling phenomena such as VCFC and RVCFC produce vapour floods, which are not greatly affected by liquid loads.

Table 1 shows that even though the tower charge rate rose by more than 25% in the operational debottleneck, the internal vapour rate on Tray 16 only rose by 6%. At the same time, the tower pressure for the operational debottleneck was higher, reducing internal vapour velocities. **Table 1** shows that the C-factor for Tray 16 at the operational debottleneck was only 1% higher than before. Considering the lower liquid load, the trays percent jet flood diminished from 78% of flood to 74% despite the rise in tower charge rate.

The next question is why the internal vapour flow rate only increased by 6% when the tower charge rate increased by over 25% with the same crude. Our simulation (see **Table 1**) shows that before the kerosene draw was operated, Tray 16 operated at 213°C. Drawing the kerosene made Trays 11-16 much hotter, with Tray 16 operating at 238°C. This 25°C



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temperature increase carries a lot of heat up the tower. With the Tray 18 temperature remaining unchanged, this means a much smaller but hotter internal vapour rate was used to maintain the heat balance.

While this helped debottleneck Tray 16, Trays 17-19 did not enjoy these large temperature increases, and the internal vapour rate through them rose. On Tray 18, the internal vapour flow rate rose by 16% following the operational debottleneck, and the C-factor increased by 10% (Table 1). With this tray being above the draw, it also saw a 10% increase in weir load.

Following the operational debottleneck, the tower was again gamma scanned using the same technique of multichordal gamma scans with quantitative analysis.

Figure 10 (equivalent to **Figure 4**) shows the froth heights for the operational debottleneck conditions at incipient flood (47 000 BPSD) and at flood (51 000 BPSD). **Figure 10** shows no flooding below Tray 17. The flood initiated on Tray 18.

The flood experienced was still premature, and occurred at 86% of the normal jet flood. This suggests that the RVCFC still occurred, this time in the small valve section. It occurred at higher hydraulic loads than it did prior to the operational debottleneck, and at higher C-factors than previously. The reason is that the small valves with

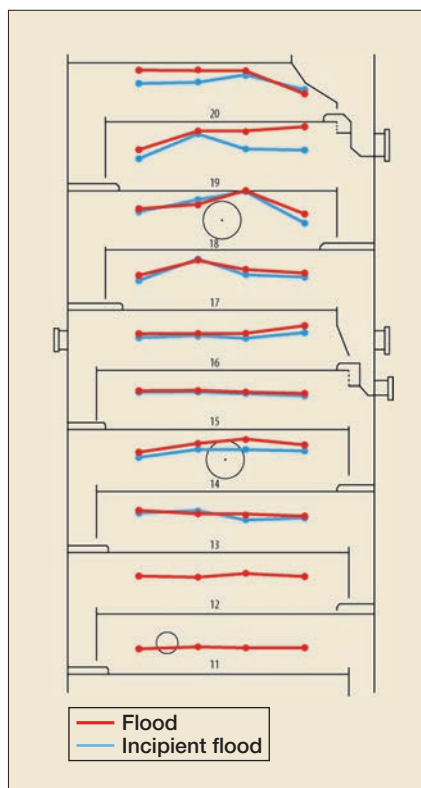


Figure 10 Froth heights, following operational debottleneck. Blue, incipient flood; red, flood

the sharp orifices and lower open area were significantly less prone to channelling than the venturi valves with the larger open area. This is supported by the higher dry pressure drop shown in Table 1 for these valve trays. The hydraulic gradients shown on **Figure 11** for the operational debottleneck scans confirm that RVCFC was still in action, this time on Tray 18.

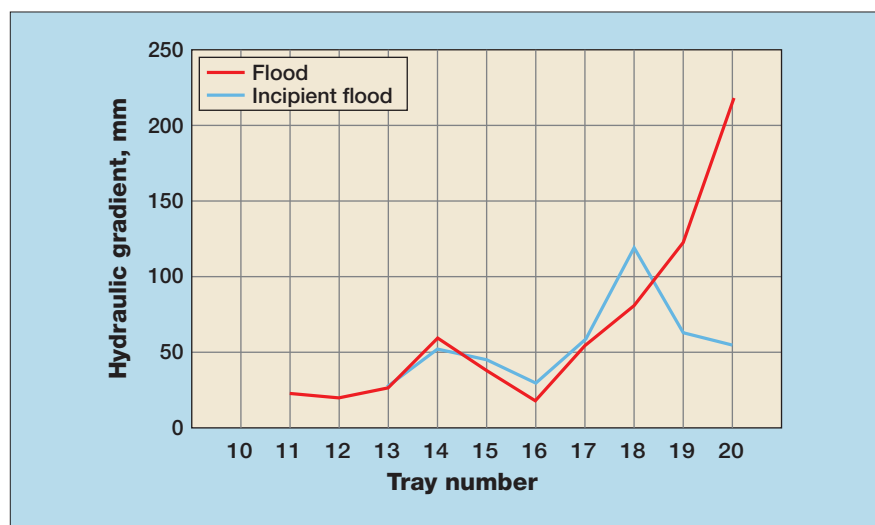


Figure 11 Reverse gradients, following operational debottleneck. Blue, incipient flood; red, flood

Permanent fix and takeaway

Based on this diagnosis, for a permanent fix, the RVCFC needed to be eliminated. The tower was retrayed with the same type of tray, Nye trays, but with a design that gave significantly larger friction pressure drop. The new trays also had two passes, which halved the ratio of flowpath length to tray spacing and lowered the weir loads. The modifications completely eliminated the bottleneck and the RVCFC, permitting higher than design capacities to be achieved.

Modifications were also made to fully eliminate the excursion, with and without the kerosene side draw. The diesel draw from Tray 11 was replaced by a seal-welded chimney tray. The diesel draw and diesel rundown line were increased to 16 inches diameter.

Following these modifications, the tower continuously operated well at a feed capacity of 52 500 b/d (7070 t/d), with or without the kerosene side draw. The tower was tested and shown to operate well at a feed capacity as high as 57 500 b/d (7750 t/d). The high-capacity Nye trays (which were the same type, but different design, compared to the originals) well exceeded their original design capacity. Excursions and flooding no longer occurred, soon to become a faded memory.

The investigation unveiled a previously unknown phenomenon, which we termed reverse vapour cross flow channelling (RVCFC). This phenomenon is closely related to vapour cross flow channelling (VCFC), but is induced by the forward push on high-capacity trays rather than by the normal hydraulic gradient on the trays. This forward push generates reverse hydraulic gradients, with the clear liquid heights highest at the tray centre and lowest at the tray inlets. Like VCFC, this phenomenon sets in when there is a significant channelling initiator, while the friction pressure drop of the trays is too low to counter the channelling, and when the ratio of flowpath length to tray spacing and the weir loads are high.

Acknowledgement

As stated, the quantitative gamma scanning requires high-quality gamma scan data. With this in mind, the authors wish to specially acknowledge a gamma scanning maestro, Dave Fradgley from Tracerco, and his Tracerco team, whose excellent work provided us with outstanding-quality data that led us to a correct diagnosis and solution. Our multitude of validation checks demonstrated that these were among the very best quality scans we have ever seen. Many problems along the way were solved thanks to their ideas and expertise. Without their expertise, we would have been unable to unveil RVCFC.

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Henry Z Kister is a Fluor Corp. Senior Fellow and Director of Fractionation Technology. He is the author of three books, the distillation

equipment chapter in *Perry's Handbook*, and over 100 articles, and has taught the IChemE-sponsored Practical Distillation Technology course over 450 times in 25 countries. A recipient of several awards, he obtained his BE and ME degrees from the University of NSW in Australia, and is a Fellow of IChemE and AIChE, and a member of the NAE.

Email: henry.kister@fluor.com

Neasan O'Shea retired from P66 having served nearly 40 years with Irish Refining Company as Technical Manager and latterly Commercial Manager prior to Phillips 66's acquisition of the company. He holds a BE in chemical engineering from University College Dublin and a BComm from University College Cork, is a Chartered Engineer and a Member of IChemE and Member of the Energy Institute.

Dan Cronin has over 30 years of experience in the oil industry, in a variety of technical and operational roles based at the Whitegate Oil Refinery. He served as Process Design Lead for several projects and more recently as Advanced Process Control and Optimization Lead. His areas of expertise include fractionation, process modelling, APC and dynamic real-time optimisation. He is a Chartered member of the IChemE.

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The future of automation

The Industrial Internet of Things promises to be the biggest influence on plant automation systems since the advent of distributed control systems

ROHAN MCADAM and PAUL MCLAUGHLIN
Honeywell Process Solutions

The hype around the Internet of Things (IoT) is even getting to technology companies. One Samsung vice president recently suggested the industry stop using the phrase. “[F]or consumers it’s just the Internet. I think we need to get over ourselves,” he said.¹

Whatever the language however, its continued growth looks assured. The number of connected devices is forecast to reach 34 billion by 2020, up from 10 billion in 2015, according to one recent forecast. Nearly \$6 trillion will be spent on IoT solutions over the next five years, researchers predict.²

Moreover, businesses rather than consumers are leading adoption. This year will be an important one for uptake, according to technology consultants Gartner. Its research finds that the number of organisations adopting the IoT will increase 50% in 2016. And, while only 29% of organisations overall are using the IoT, according to its survey, the proportion in heavy industries such as oil and gas is much higher. By the end of 2016, 56% of such businesses are expected to have implemented the technology.³

Such forecasts are not fanciful. The Industrial Internet of Things (IIoT) has promise to be the biggest influence on automation systems since the advent of microprocessor based distributed control systems (DCSs). Combining ubiquitous sensors with Cloud-borne data analytics and storage systems, it offers the opportunity to bring rapid improvements in safety, reliability, efficiency, productivity and, ultimately, profitability. Intelligent application of digital technology

can meet fundamental automation needs for more informed, quicker and better decisions. Indeed, the IIoT could equally stand for the intelligent Internet of Things.

To secure the benefits, however, it must be intelligently applied, and roll-out of the IIoT could yet stall.

For the IoT, simply connecting vast numbers of objects from daily life to the internet will not, in itself, create interesting and useful new ways of doing business; likewise the IIoT will only deliver if there are platforms, tools, algorithms and applications to analyse, distribute, and act on the huge amounts of data that result.

To secure these benefits will require an intelligent application of the technology, which in turn requires an intelligent approach to IIoT architecture. This must cost-effectively secure the benefits of IoT technologies while also meeting the distinct requirements of automation systems in terms of performance, availability, reliability, safety and security.

We are, in many ways, just at the start of this process.

A new technological ecosystem

The primary ingredients of the IoT are fairly obvious: the Internet and ‘things’ – in the context of the process industries encompassing the digital automation system.

Intelligent devices and machine-to-machine interfaces are not new concepts for these industries. The parallels between the IIoT and the microprocessor based DCS (conceived by Honeywell in 1969) are obvious. The latter’s network of sensors, actuators, controllers, and

computational capabilities is an essential precursor to the IIoT. However, internet connectivity significantly expands both the possibilities for connectivity and the computational and storage capacity available. In short, it allows more sensors to be connected, more measurements to be recorded and stored, and more computational power to analyse and act on the data.

The IoT and IIoT therefore encompass a range of elements, including ubiquitous device connectivity, internet storage for the very large amounts of data produced and statistical and machine learning algorithms to analyse and act on that data. The IoT has both been enabled by and precipitated the rapid emergence of a series of technologies that are now integral to the new types of systems typically categorised under its banner:

- **Virtualisation** technology allows software workloads, whether entire operating systems or individual applications and tools, to be decoupled from the hardware on which they run. Using this technology to deploy solutions can significantly reduce cost, simplify management, improve availability and avoid problems associated with churn in the underlying hardware platforms.
- **Cloud computing** provides virtualised platforms with elastic computational and storage capabilities. Usually run as a service based offering supported by large data centres, the Cloud eliminates the requirement for firms to manage their computing infrastructure, and significantly reduces the costs

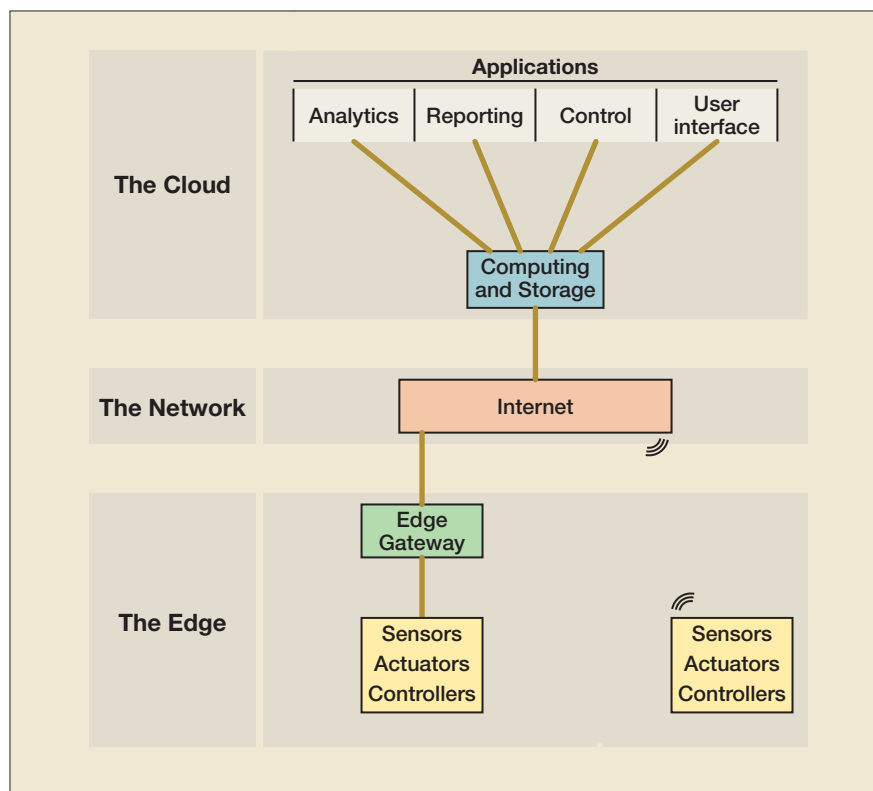


Figure 1 The Internet of Things

associated with adding CPUs and disk space.

- **Pervasive networking** is an essential part of the Internet of Things, enabling devices to participate in Cloud based services. This can be through direct connectivity or through gateway devices such as smartphones connected to a Cloud based service. It can therefore take the form of a direct 3G and 4G cellular network internet connection or an indirect connection through Wi-Fi, Bluetooth or the Near Field Communication protocol.

- **Big data analytics and machine learning** are essential to extract value from the increased computational and storage capacity provided by the Cloud and the large amounts of data collection made possible. Big data is characterised by large volumes of data (in the order of terabytes and petabytes) requiring new techniques for storage and analysis, such as massive parallel data stores and statistical techniques for identifying correlations and patterns. The availability of data for analysis has also spurred development of machine learning techniques, including artificial intelligence algorithms that are applied to big data stores.

- **Smart devices** have proliferated thanks to the availability of small, low-power processors now embedded in many devices, allowing them to act as more than mere sensors or actuators. This local computational resource enables devices to become more interactive and autonomous.

- **Mobile computing** has been facilitated by smart phones and tablets to enable a wide range of highly interactive context-sensitive (location, time, task, and so on) applications. More recently, smart phone capabilities are being distributed across a series of smart, wearable devices; the current crop of smart watches and head-up displays are a good example of this. Smart wearables are able to collect more data from a wider variety of sensors (cameras, voice, gesture, biophysical) and render information across a wider range of modalities (vision, speech, haptics). This dissemination of task-specific technology enables richer applications that are usable in a wider variety of circumstances.

- **Cyber security** has become a pervasive concern in all computing systems, but perhaps particularly systems that leverage the technology

inflections outlined above. The increasing pervasiveness of networking and connectivity creates more complex webs of communication that need to be secured to support the operation of systems and protect intellectual property (IP).

Architectures: addressing misconceptions

The architecture of the Internet of Things (see Figure 1) shows how these various technologies and trends interact. It also illustrates significant differences from traditional systems, including the DCS (whatever debts the IIoT owes to that development).

The consumer IoT architecture consists of three main domains: the Cloud, the network and the Edge:

- **The Cloud** hosts the computational and storage capabilities as well as applications for analysis reporting, control, and the user interface (although the user interface may sometimes run locally on devices such as smartphones at the Edge).
- **The network** provides the connectivity for the components of the architecture through IP based protocols
- **The Edge** is where the things in the IoT are connected – the sensors, actuators, and controllers – either directly or through an Edge gateway (as described above).

Two mistakes can be made when considering the IIoT. The first, as stated, is to underplay the differences between IIoT and traditional automation solutions.

The second is to underestimate the distinctions that must be made between the IoT familiar from consumer applications and the IIoT, which must meet the particular requirements of process control and safety applications. To put it another way, it is a mistake to think that the IoT and IIoT are the same and that the principles that inform the traditional DCS architecture can be ignored.

The primary consideration in industrial settings is safety and security. These are obviously factors in any IoT architecture, but when dealing with process control,

volatile substances, exothermic reactions and potentially dangerous operations, they are critical.

In practice, this means not everything in the IIoT can be pushed up into the Cloud for reasons of cyber security but, more significantly, because of the response times and robustness required. To put it plainly, most control elements need to be close to the device they control to ensure fast signal processing and resilient connections for process safety. The emergency shutdown system cannot rely on a potentially unreliable internet connection.

This is not a new observation; it is the philosophy that informed the development of the DCS. It is a large part of why the DCS is distributed. It is also interesting to note the recognition of similar principles in the development of Fog computing⁴ and Mist computing⁵, both of which push storage and computing capabilities down from the Cloud towards the Edge for a more robust solution.

Another important consideration

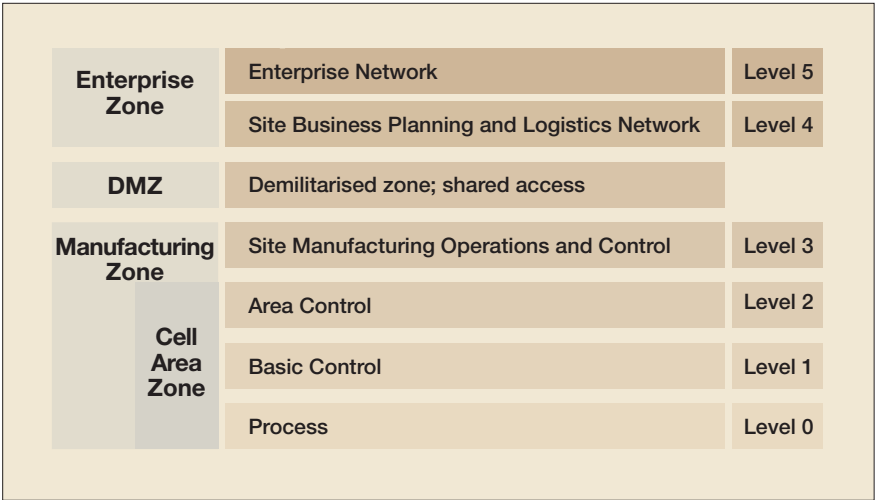


Figure 2 Purdue Enterprise Reference Architecture model

is the installed base. Again, connectivity and integration with existing devices is undoubtedly an issue in many IoT applications. In the journey towards the IIoT, however, the existing, huge installed base of automation systems and its legacy software and hardware is going to play a much greater role. An industrial plant is a long-lived, capital-intensive asset requiring long term support in the face

of rapid technological advances.

In practice, this means that a lot of the Edge environment – the sensors, field devices and human machine interfaces (HMI) – already exists and will continue in place. The big question is how you build on that installed legacy to create an IIoT system.

The IIoT architecture

In broad terms, the IIoT is simply

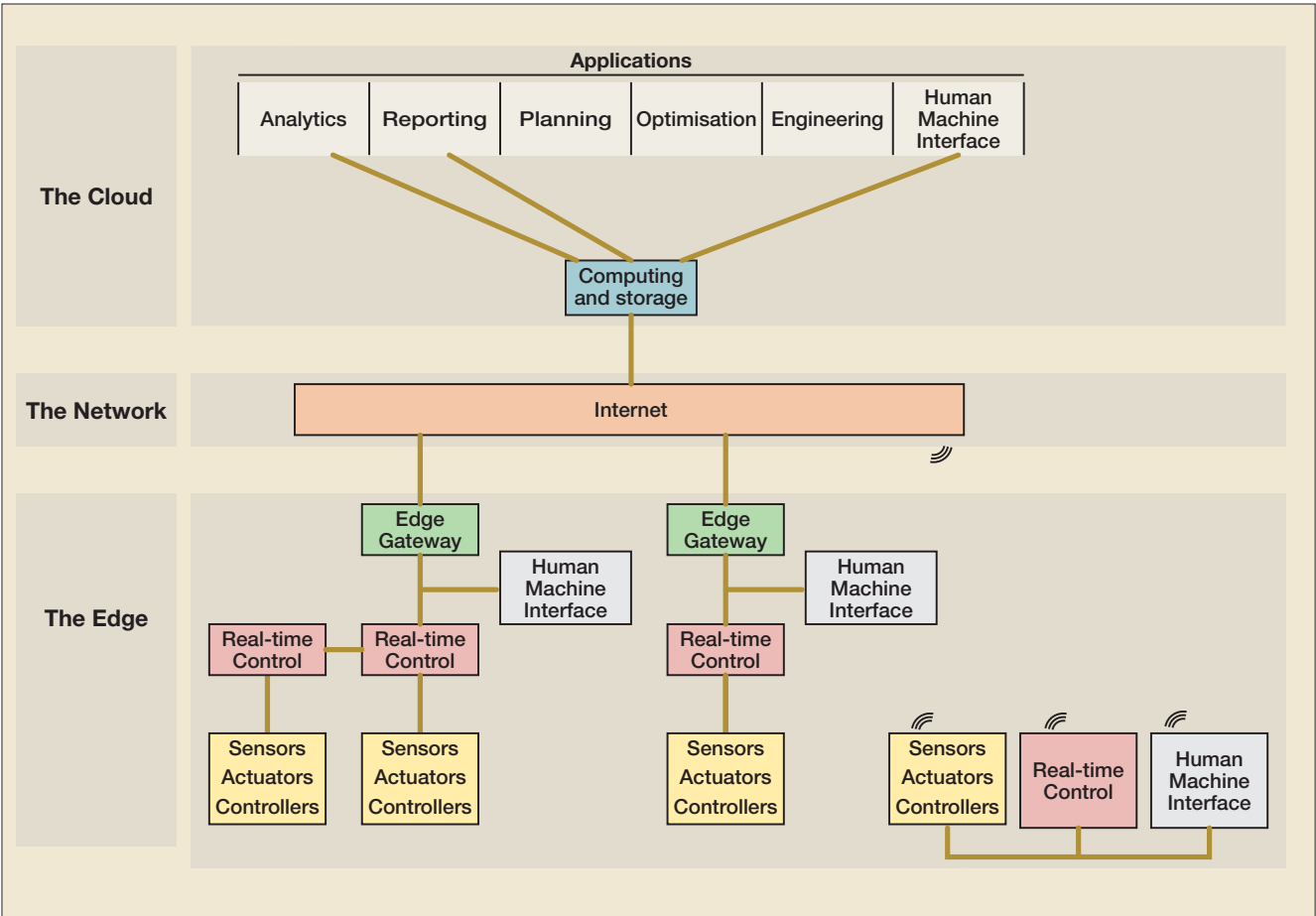


Figure 3 The Industrial Internet of Things

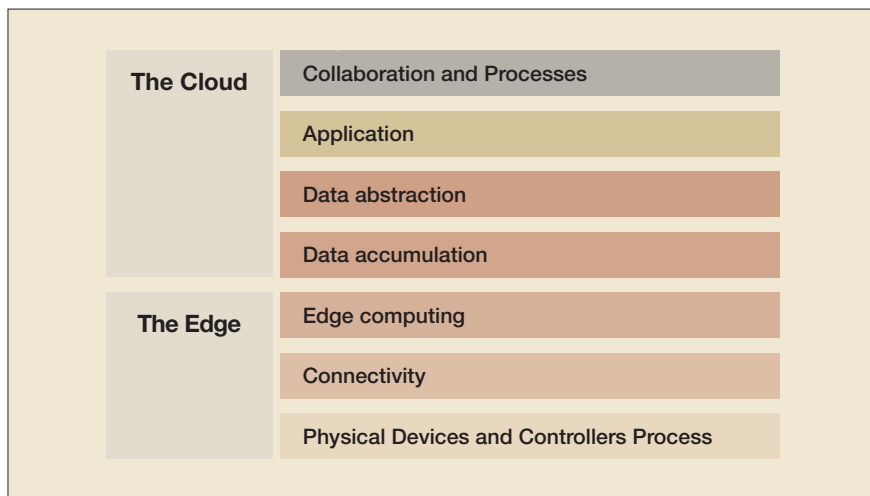


Figure 4 IIoT reference model

the application of IoT ideas to the planning, running, analysis and optimisation of industrial enterprises. The IIoT connects the world of industrial ‘things’ – the sensors, actuators, controllers, robots – to computational capabilities residing in internet based storage and analytics. In doing so, it significantly enhances the capabilities and scope of current automation systems.

In line with the discussion above, however, the architecture must take into account the key principles and existing equipment traditional systems are built on. Any discussion of the IIoT architecture should therefore properly begin with reference to the traditional Purdue Enterprise Reference Architecture developed in the 1990s for the DCS and associated applications. This is shown in **Figure 2**, with the physical process at Level 0; basic control (Level 1); area control (Level 2); site manufacturing operations and control (Level 3); and business planning and logistics (Level 4). Enterprise wide business systems such as ERP systems are often considered as Level 5 of the Purdue model.

Now consider the architecture of the IIoT (see **Figure 3**). As can be seen, this is different from the IoT architecture.

Compared with the IoT architecture, a key difference is the nature of the Edge computing environment. In the IIoT, the Edge computing environment provides the opportunity to address key

performance and robustness requirements for industrial process control. This accounts for the large number of real-time controllers at the Edge.

The Edge in the IIoT also hosts a much wider range of devices and associated communication protocols. The devices here will include sensors, actuators, controllers, and HMIs. They are located in close proximity to the production process and may communicate directly with Cloud based services or via an Edge gateway that acts as a data concentrator and/or filter and protocol converter. Edge devices may also act collectively in a federation of devices to provide an autonomous coordinated set of capabilities at the Edge. For example, a federation of sensors, actuators, controllers and HMIs could provide real-time control and management for a process unit or area. Such a federation would utilise peer-to-peer communication amongst devices using a variety of protocols.

While there is a trend toward open IP based protocols in the IIoT, such as OPC UA, there will continue to be a role for existing protocols such as HART, FieldBus, Modbus and so on, particularly for existing installed devices.

The IIoT and Purdue

The differences between the IIoT architecture and the traditional DCS are more starkly illustrated if you look at the Purdue model, however.

At the highest level, the IIoT can be separated into two major subdivisions – the Edge and the Cloud. This may be subdivided (see **Figure 4**), but all the control system and automation elements must be placed in one of these two basic levels.

Given the safety and security considerations discussed, much of the computational work must be placed at the Edge, close to the process control elements, rather than pushing it up into the Cloud. This is not just a necessity, however; it is an opportunity.

There is now, due to smart devices and increasingly efficient and powerful processors, much more computational power in the Edge environment. That means that more data processing can occur there. As well as basic control, Level 3 capabilities such as advanced process control, optimisation and even analytics in the form of asset management analytics can also be pushed to the Edge, being built into the controllers. The capabilities will still operate in essentially the same way, but they can be located at a more appropriate level, closer to the assets and equipment they relate to.

This is really just an extension of the logic informing the DCS, and these developments may have come anyway. The review of architecture prompted by the IIoT, however, accelerates the process.

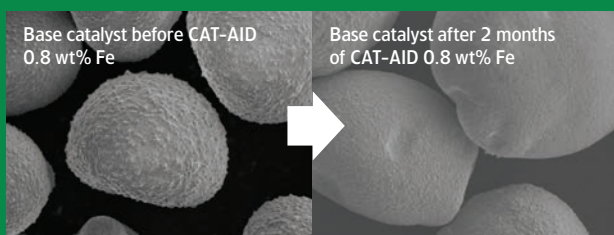
The placement of different capabilities will vary considerably even within the oil and gas sector. In upstream applications with geographically distributed systems, for example, it is much more difficult to locate some of that work at the Edge, and functions must be more centralised, pushing them to the Cloud. In refineries, much more local capability is possible at the Edge. Applications that are not safety critical, however, can usually be put into the Cloud. In many cases, there is a choice to be made.

Figure 5 represents one potential architecture that reconciles the traditional DCS structure with the IIoT while maintaining the key qualities provided by the Purdue

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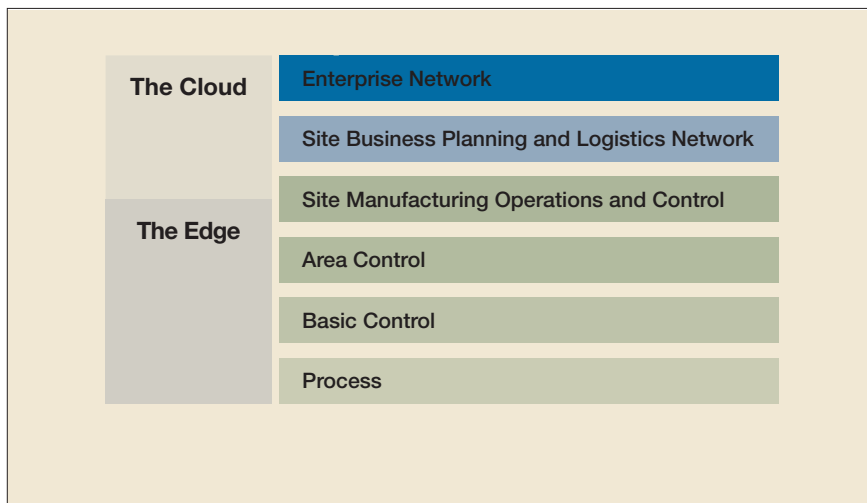


Figure 5 Basic structure of the IIoT in the Purdue model

model (safety, security, reliability, efficiency):

- Level 1 of the Purdue model, basic control, moves to the Edge in the IIoT model
- There is also a strong argument for moving much of Level 2, area control, to the Edge to keep it close to the process being controlled for performance, security, and reliability.
- Functionality represented by Level 3 – site manufacturing operations – will be pulled up into the Cloud and pushed down to the Edge depending on the balance of key system quality attributes.
- Level 4, business planning and logistics, moves to the Cloud.
- History, APC, S88 Batch, and alarm management can all be deployed either in the Cloud, on premises in embedded devices, or between both.

Architectural benefits

As stated, plants moving to an IIoT structure are forced to make decisions to relocate elements of the control system, either pushing capabilities to the Edge or the Cloud. This is necessary to realise the benefits to the production process and operation in terms of efficiency, reliability and safety. These have been covered extensively in other papers, but briefly result largely from increased intelligence and automation possibilities due to big data and analysis; and from the increased opportunities for collaboration, situational awareness and mobility provided by

internet connectivity, the Cloud and mobile devices.

There are, however, a number of benefits directly linked to the revisions in the system architecture. In terms of security, for instance, many existing DCS components have no inherent security built in. They may lack explicit access control mechanism and transmit data on the network in plain text. In an IIoT based system, these legacy components are confined to the Edge, where Edge gateway devices provide access control and secure communications.

Likewise, the IIoT architecture addresses the vulnerability in current automation systems stemming from the use of open systems platforms, particularly in Levels 2 and higher in the Purdue model. Patching these platforms in traditional systems to address vulnerabilities is a major maintenance cost. Even then, ‘zero day’ attacks for which patches are not available, and attack vectors including standards based interfaces such as USB ports mean vulnerabilities remain.

The IIoT, however, pushes automation system functionality either down into the hardened Edge computing environment or up into the Cloud, which, again, has rich access control and communications security mechanisms built in. The centralised nature of the Cloud infrastructure also makes it much easier to maintain in order to address vulnerabilities that are discovered.

Finally, this delivers benefits in terms of reliability, maintenance and safety. Moving functions into the Cloud allows them to be more easily managed, maintained and upgraded, while the decoupling of Edge and Cloud-based functions allows them to be managed much more independently. As a result, the system can remain operational through a wider range of life-cycle events. Pushing control functions to the Edge, meanwhile, allows those functions to act more autonomously. They therefore have fewer dependencies on other components, reducing the potential causes of failure.

First things first

The journey towards the IIoT will not be completed overnight, and existing investments must be accommodated and integrated.

Development of standards to allow connection of data from a wide range devices and systems is a key part of the development of the IIoT. OPC Unified Architecture (OPC UA), for example, which extends the widely used OPC communication protocol to allow products to easily interconnect and share data effectively, is being embedded in an increasing number of devices. This can be extended to existing devices using OPC UA proxies or UA wrappers that allow traditional OPC servers to communicate with a new UA client, or an HMI without an OPC UA to interface with UA devices. With this, existing equipment and devices need to be securely integrated into the IIoT architecture.

Development of the IIoT will need to accommodate legacy systems in a number of other respects, too:

- It must take account of existing customer IP investments, preserving control strategies, supervisory applications, and HMI graphics as far as possible as the automation system evolves.
- It will have to ensure that equipment and systems’ Safety Integrity Levels are maintained; and the impact of additional demands for data from IIoT applications on existing automation system components will have to be carefully managed.

After all, there is little benefit in new applications if they compromise the core capabilities of the automation system. These challenges are not insurmountable, however, and IIoT technologies will also help smooth the transition and implementation of new capabilities by removing burdens from maintenance and engineering teams.

An appliance based approach to new Edge capabilities, for instance, will significantly cut down set-up and maintenance work: plug-and-play boxes can connect to the Cloud to automatically configure and update, relieving teams of the responsibility to set up a PC, and install, configure and maintain software. Virtual engineering platforms, meanwhile, enable entire automation systems to be engineered in the Cloud. Equipment such as computers and networking hardware only need to be procured at the end of the process, dramatically streamlining the automation system engineering process.

These efficiencies and cost savings will enable plants to focus on the areas where greater thought is required: whether that is accommodating and integrating legacy equipment and IP, or considering carefully where in the architecture to place capabilities.

As said, in many ways we are still at the start of this journey. The IIoT remains an undiscovered country, waiting to be explored. However, the lessons from decades of development of automation systems and DCSs can serve us well. If we draw on them wisely, we can be confident of mapping out a route to bring us safely to the new world.

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Rohan McAdam is Chief Architect with Honeywell Process Solutions and has worked primarily on the development of human machine interfaces for HPS industrial process control products. He now leads the development of IIoT architecture in HPS. He holds a mathematics degree and a PhD in computer science from Charles Sturt University and a master's degree in cognitive science from the University of New South Wales.

Paul McLaughlin is Chief Development Engineer with Honeywell Process Solutions and leads the HPS Chief Engineering Council and the Global Architecture Team, working on system design, product line architectures, formal roadmaps, investment prioritisation, strategic planning, intellectual property management, and cross-business major initiatives. He holds a master's degree in engineering from the University of Pennsylvania, and a bachelor of science degree in mathematics and computer science from the University of Delaware.

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Connected data improves refinery operations

Connecting data with a common knowledge management platform improves visibility, response, and analysis to benefit refinery operations

JEFFREY ZURLO
GE Water & Process Technologies

The transition from board-mounted, pneumatic refinery controls to electronic distributed control system (DCS)-centred control and data historian systems that started to become popular in the 1970s has helped refinery operations improve production, reliability, and product yield optimisation. However, still today much of the data originating from vendor service companies remains locked away in proprietary systems that are shared typically only as periodic snapshots of performance – collectively referred to as ‘dark data’. Dark data in a refinery often includes information about the performance of specialty chemical programs, equipment vibration monitoring, corrosion measurements, inspection readings, and many other ‘non-core’ areas of refinery operations. In the last several years, new sensors, controllers, and data handling infrastructures have been under development to connect this once dark data with the traditional refinery operating data as part of the movement by dozens of companies to create an ‘Industrial Internet of Things’. Connecting these data sources allows more rapid, more transparent, and remote evaluation of system performance, automated application of proprietary analytic algorithms, and other system analyses that were impossible to do previously. The increase in analysis and performance measurements provides significant opportunities to better optimise reliability and profitability for these once isolated systems.

There are several factors that are

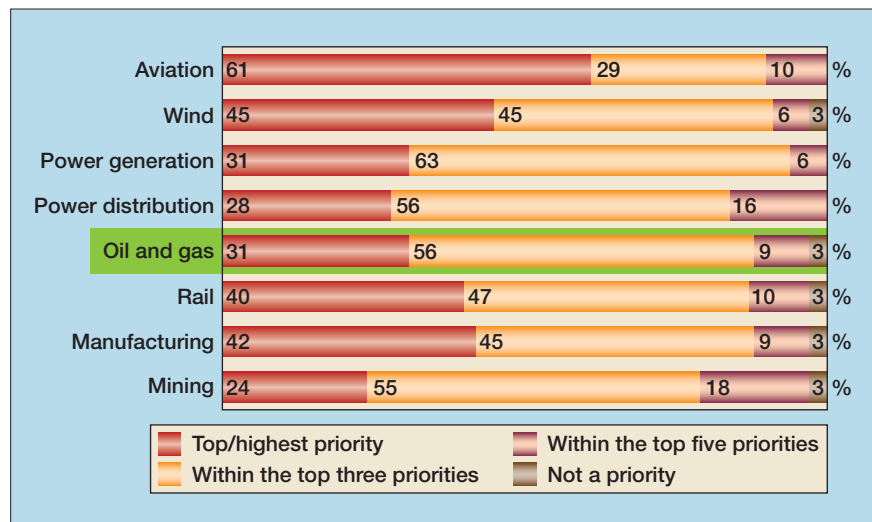


Figure 1 Industry survey on Big Data Analytics

leading to this leap in technological capability. First, the cost to produce sensors continues to drop and the variety of operational and physical parameters that can be measured with sensors continues to increase. This allows more sensors to be deployed in more places in a refinery, making visible what was previously unmeasurable. In addition to sensor technology, wireless transmission has become cheaper and more reliable, allowing reduced cost of installing sensors and maintaining connection without the need for running miles of wires and cables. The increase in sensors is already happening today – the average refinery has seen a five-fold increase in the number of sensors deployed over just the last five years.¹ The trend to increase measurements via the use of sensors and other on-line equipment is expected to continue for some time as the digitisation of industrial plants provides another

leap in productivity and efficiency.

Robust data systems and large data storage capacity are needed to handle the additional data streams generated as sensors and measurements increase. Luckily, the cost of data storage has also dropped considerably to economically accommodate the increased data load. Additionally, cloud based storage solutions are becoming more popular daily, it seems, opening up even greater storage and processing capacity. While cyber security is still a concern for refiners and industry,² as security measures are improved there is a trend in the industry to allow data to be transmitted ‘beyond the fence’ and stored remotely, thus making data available to people and systems not physically located at the refinery. This movement contributes to the tight integration of the physical and digital worlds, which many are calling the Industrial Internet of Things, and

places us on the verge of a new productivity revolution to transform industry on the scale of the industrial and digital revolutions of the past.³ According to a 2014 study by Accenture and GE, 'Big Data Analytics' is within the top three corporate priorities for 87% of oil and gas companies surveyed and 'not a priority' for only 3% (see **Figure 1⁴**).

The growing abundance of data provides great opportunities, but also additional challenges, in efficiently sifting through large amounts of data and extracting valuable information to improve refinery performance. The emergence of more advanced analytics is a growing trend in the industrial world, including refineries, and may indicate a shift in hiring practices. Forward-looking companies have already started investing in data scientists, programmers, and data analysts to provide the value that big data analysis can offer. More sophisticated analysis due to better data visibility and advances in analytics helps to more easily anticipate impacts of operational or feedstock changes on system performance and reliability. This field is growing rapidly in many areas of industry, so much so that 'data scientist' was listed as the number one job in America for 2016 by Glassdoor.⁵ It is quite possible that in the near future data scientists will join the regular ranks of refinery professionals like engineers, instrument and controls teams, and reliability staff. In a recent interview with SAP, Peter Reynolds from the ARC Advisory Group argues that adopting the technologies provided by the Industrial Internet of Things could be a key to survival for many oil and gas companies by taking advantage of optimising infrastructure to lower operating costs. "Why monitor 50 pumps when you can monitor 50 000 pumps [from a centralised location]."⁶

The bottom line: connecting dark data enables greater insight to the system conditions, performance, and pitfalls to better turn data to information and information to action.

Common connected data types

- **Real-time (streaming) data connection:** making a constant, always-on, connection between the local sensors/controllers and a cloud based control system. This would provide the most dynamic data stream for analysis and action and theoretically enable remote closed-loop control of a system outside the refinery fence. However, this method is the most difficult to keep secure from a cyber security perspective and requires an absolutely stable signal connection to avoid dropped signals compromising system control. Additionally, utilising some of the other methods described below minimises the need to have an 'always on' data connection for the scope discussed in this article.

Emergence of more advanced analytics is a growing trend in the industrial world, and may indicate a shift in hiring practices

- **Near-real-time data communication** blends local closed-loop control and monitoring with cloud based analytics and performance monitoring. The approach uses local, smart devices to record sensor data and perform closed-loop control on systems. The system can periodically establish a connection to the cloud based platform and bilaterally transmit data between the local device and the platform. This increases data security since the information is only transmitted in 'bursts', and maintains closed-loop control locally, which is more stable. This method still enables the platform to monitor system performance rapidly and is usually sufficient to provide feedback to out-of-control conditions to enable timely corrective actions.

- **Edge computing:** a variant on the above near-real-time communication and a growing trend, it increases the intelligence of the local smart device

to enable more complex analytic execution locally and autonomously. This approach takes advantage of the greater processing power and lower power requirements of today's computer processors to unload communications and server bandwidth while enabling controls beyond traditional PID-type control schemes. Data transparency and system performance are preserved by communicating critical data and analytic outputs to the platform, while pushing more 'data crunching' down to the local level.

- **Manual, or on-demand, data upload:** intended mostly for local data obtained on a spot basis, such as laboratory results for field readings not captured by on-line sensors or instruments. Data of this type is updated either on some set frequency, or as it is generated by a person. There are several mechanisms to incorporate this data, such as manual template upload, live forms, or file transfer/email. Historically, this has been some of the most difficult data to incorporate into the data stream for refinery operation and is therefore a huge, mostly untapped resource of information.

Advantages of connected data

In a 2012 AFPM Technology Forum presentation, Sam Lordo outlined the need to define key performance indicators (KPI), key control parameters (KCP) and key stress indicators (KSI) as essential elements of a well-defined performance management system.⁷ For a system to be valuable to refinery operations, a performance management and communication system needs to be able to not only identify when a system is out of compliance but also provide a simple mechanism to ensure action is taken to bring a system back into control. A common platform utilising connected data can provide a single data view that is transparent, can automate specialised analytic analysis to allow more advanced performance assessment, and can allow for defined control limits for KPIs, KCPs, and KSIs. This type of comprehensive system provides a clear mechanism to turn data to

information, and information to corrective actions.

Utilising a system that allows all the pertinent data to be collected on one platform and available to a wider group of support personnel has several advantages for the refinery. Pairing connected data with the increased scope of on-line sensors and increasing capabilities of local smart devices benefits refinery operations and reliability at a number of different levels. The following sections will describe the main types of advantages seen so far in refinery operations and include examples that highlight how these improvements were achieved and the value provided to the refiner. Please note that this is a rapid-growth area, so it is expected that the scope of what can be achieved using connected data will grow rapidly over the next 5-10 years. We are only very close to the starting line today.

Improved oversight and communication - automating performance assessment and dashboards

With a connected data system everyone looks at the same data pool. Plant personnel, vendor field teams, and corporate subject matter expert staff in both the refinery and vendor organisations have visibility to system status and data at the same time, from the same system. This breaks the historic linear communications chain from the past that slows communication and often results in some parties using out-of-date information (see **Figure 2**). Also, dashboard ‘health checks’ are requested more frequently by plant and corporate technical personnel to provide visibility to current system performance and anticipate any potential future operating issues, often needing daily updates. A connected data system can accommodate creation and maintenance of dashboard systems for any level of organisation. In addition, analytic outputs and data feeds from vendor measurement and control systems can be directed back into the plant’s DCS or data historian system. This enables expansion of the data typically captured in the plant’s historian to more detailed vendor

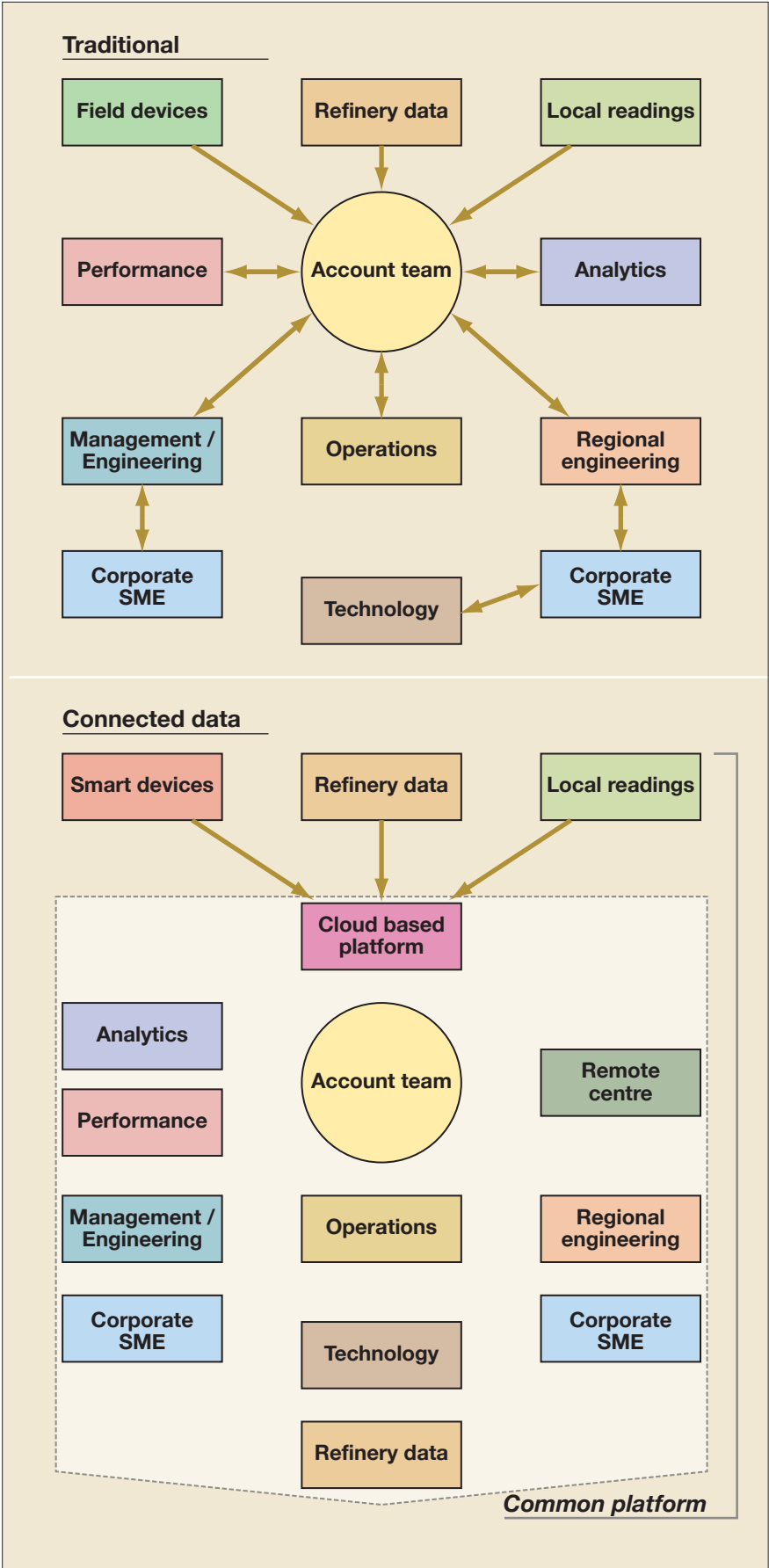


Figure 2 Changing communication avenues

information, preserves the ability for plant personnel to be able to see all their operating data in one applica-

tion, and still take advantage of the improved communications that connected data systems provide.

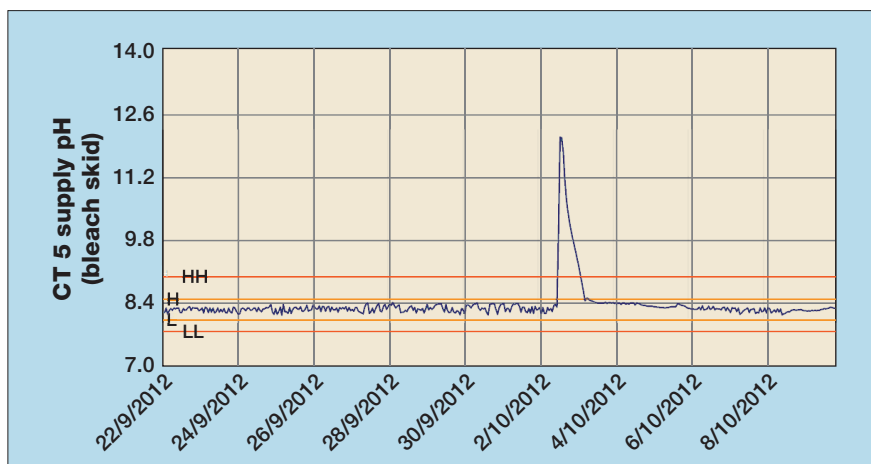


Figure 3 Cooling tower pH measurement

Tighter system control through faster recognition and response to out-of-control conditions

Triggering alarms from the automated data gathering and performance assessment of KCPs and KPIs against established operating limits means that loss of control is recognised as it happens, not at the next operator or vendor visit to the system. While this is nothing new for a refinery's core operating and control functions that feed through the DCS and data historian systems, vendor systems often use proprietary, disconnected systems and spot data joined with plant operational data to assess system performance. Historically this has been done via pooling of this data locally using spreadsheets and local analytic packages, which is not very efficient and sometimes introduces significant lag time between data generation, issue recognition, and corrective actions.

Case 1: reducing upset impact through rapid response protocols

This example highlights the advantage of connected data in greatly improving response to an event-driven, upset condition. A refinery cooling tower utilises local automated closed-loop control of the cooling water chemistry and on-stream monitoring of several operating parameters such as pH, chlorides, conductivity, and chemical residual testing. Due to a valve misalignment outside the battery limits of this operating unit, concentrated caustic was fed to the make-up line of this cooling tower

instead of the normal treated surface water. Replacing water in the make-up line with caustic overwhelmed the control system, resulting in an almost immediate increase in pH of the tower above 12 pH (see **Figure 3**). Sensing the system was out of control, the cloud based platform issued an alarm to the local account team via email, text, and direct system alert.

The account team was on-site and immediately contacted the unit operators, who were unaware of the issue, as a field inspection round had been completed just before the incident occurred. The local vendor team and the operators responded to the condition by verifying the condition was real via hand measurement at the cooling tower basin, making aggressive system changes to reduce the pH of the cooling water back to the normal control range to prevent scaling, and communicating to other plant personnel to identify and correct the root cause of the problem. Since cooling water pH was identified as a critical alarm for this system, the vendor's remote central monitoring centre was also directly alerted by the system to ensure that the local personnel and customer were aware and corrective actions were being taken.

As a result of the systems in place, the upset condition was identified in minutes instead of hours; it would have been several hours until the operator was scheduled to physically monitor the cooling tower again. Rapid response brought the system back

under control in a shorter time than would have otherwise been possible, and good plant communication enabled quick identification and correction of the root cause to remove the caustic from the make-up water line. A high pH event of this magnitude in a cooling system would typically result in widespread scaling of heat exchange equipment and reduced unit run length. This would cost a refinery significant lost opportunity dollars from energy efficiency loss or unplanned outages, but was avoided by quick event recognition and rapid response to get the system back under control.

Increase the layers of protection through additional monitoring and analysis by off-site personnel

Connecting data on a cloud based platform allows the creation of a remote monitoring centre or centres, staffed by vendor personnel to provide an additional layer of oversight on system performance and act as a command centre during system upsets to help ensure local personnel are aware of the problem. The function of a remote monitoring centre is similar to the operations central control room staff at a refinery, but focused on the systems the vendor helps to service and generally do not have direct control on the process manipulated variables, just visibility to the system conditions. Also, support staff can more easily and more routinely access the information for a given refinery, typically through a secure password-protected web portal. This allows additional oversight by people with the right skill set to supplement the expertise provided by the local service and support team.

Case 2: identifying and correcting a chronic operating issue via pattern recognition

Shortly after commissioning a continuous, on-line chloride analyser in a crude unit atmospheric tower overhead system at a North American refinery, one of the vendor's off-site regional engineers noticed a repeated pattern of high

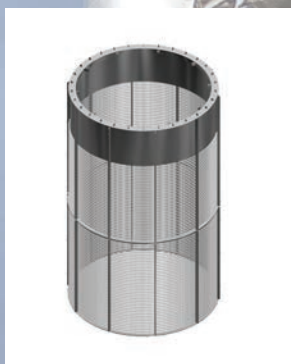


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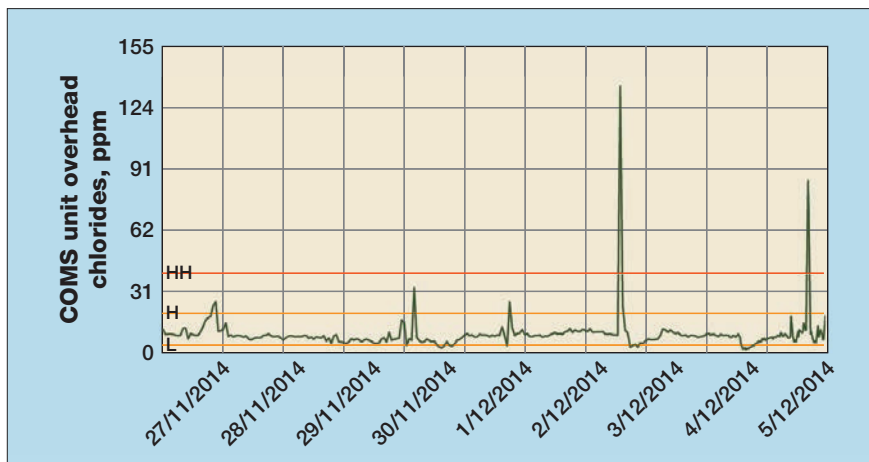


Figure 4 Overhead chloride excursions

chloride excursions on an otherwise fairly well controlled system. As can be seen in **Figure 4**, these excursions were occurring once or twice a week, have very sharp peaks and a quick recovery. This system had a history of salt-induced corrosion and fouling, resulting in extensive equipment damage and short equipment life, unplanned shutdowns, and high maintenance costs, so the short-term spikes were very concerning and a root cause investigation was launched.

The root cause was identified: the operating procedure used for many years to blend a new batch of caustic injected to control overhead chloride concentration. In brief, when the tank inventory was low, a set amount of make-down water was charged to the caustic batch tank, followed by a concentrated caustic feed to achieve targeted dilute caustic strength. Adding the water first caused injection of essentially pure water into the

crude preheat and resulted in a loss of chloride control in the overhead until the batch blending was completed. Although this practice had been in place for many years, it was not identified as an issue until continuous chloride measurement and visibility to the data stream was achieved. As a result of the investigation, the plant now purchases pre-diluted caustic to avoid periodic spikes (see **Figure 5**). As a side note, the refinery has further increased control of caustic injection by implementing closed-loop control of the injection rate using the overhead chloride measurement and to reduce sodium load in the reduced crude and vacuum residuum from the crude unit fed to the FCC unit.

Expanded analytics improve insight to system performance

Proprietary analytics previously required someone to collate and organise data from a variety of

sources, input data to run the analytic, and then analyse the output. A common data platform can automate this process whenever applicable new data is received into the connected platform, allowing vendor and plant personnel to focus on interpreting the outputs and engineer corrective action plans instead of spending time in data handling activities. Additionally, a common data handling platform that uses appropriate timestamp protocol eases the ability to export synchronised data when the rhythm of data acquisition is not the same between data parameters, for instance pairing continuous operating flows with periodic laboratory testing. This improves usability of the data for analysis and review.

Case 3: improving waste treatment plant performance

Refinery wastewater treatment units often are the areas with the darkest data. Many refineries that have moved to DCS systems decades ago still use manual log sheet entries to capture wastewater treatment data. This limits the extent of data analysis and can impede troubleshooting efforts during root cause and performance optimisation activities.

An exception is a refinery on the US Gulf Coast that had transitioned to using its data historian system to capture and store the operational data at its waste treatment unit. Thus, when the refinery started to share the data with the treatment vendor's cloud based knowledge management system, it was easy to include the data captured at the waste treatment plant as well.

This proved to be very useful when the waste treatment area started experiencing significant but intermittent operational issues in the activated sludge clarifier area. The biological system was generating an excessive amount of bulking bacteria, swelling clarifier beds and preventing effective sludge settling. This caused difficulty meeting waste treatment final effluent water quality, which required emergency measures to be taken to avoid environmental penalties.

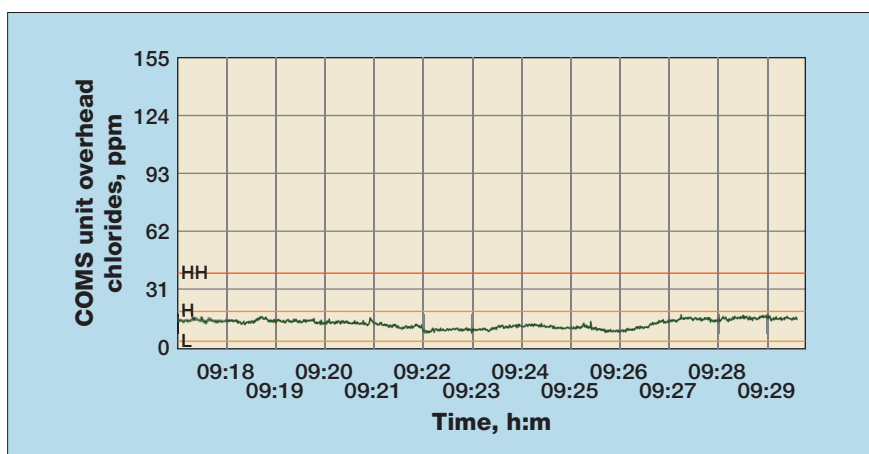


Figure 5 Overhead chloride in control

The vendor's corporate refinery wastewater SME, in addition to the local vendor personnel and refinery staff, investigated the issue to determine root cause and corrective actions. Using the connected data system allowed rapid application of multivariate regression analytics for data associated with refinery operations and the waste treatment performance. Among all the operating data considered, a strong correlation was identified between injection of organic acids at the crude unit desalters for pH and contaminant control with the increase in clarifier bulking (see **Figure 6**). Due to the large lag time between the injection at the desalter and the effect at the clarifiers, it was difficult to detect the correlation using just historical trend analysis. Organic acid programmes are used at the desalter to improve desalter performance in response to a need for controlling desalter pH due to increased desalter pH and contaminant levels at the crude units as a result of expanding the crude diet to process tight oils and other crude oils not historically processed at this refinery. Due to the strong economic incentive to process these opportunity crudes, a solution that would preserve crude oil flexibility and improve waste treatment operations was desired.

As a result of the analysis, tighter control of acid feed was initiated at the crude units to limit total stress on the waste treatment plant and communications were improved by adding total organic acid feed rates to the daily dashboard reporting system. When high injection rates at the crude units are necessary based on crude unit conditions, an operational response plan was developed at the waste treatment plant to better handle high organic acid feed rates that involved more aggressive operating changes and use of the acid injection dashboard data as an 'early warning' system. As a result of improved operational control and increased communication across operating areas of the refinery, waste treatment clarifier operation has improved significantly and is no longer considered

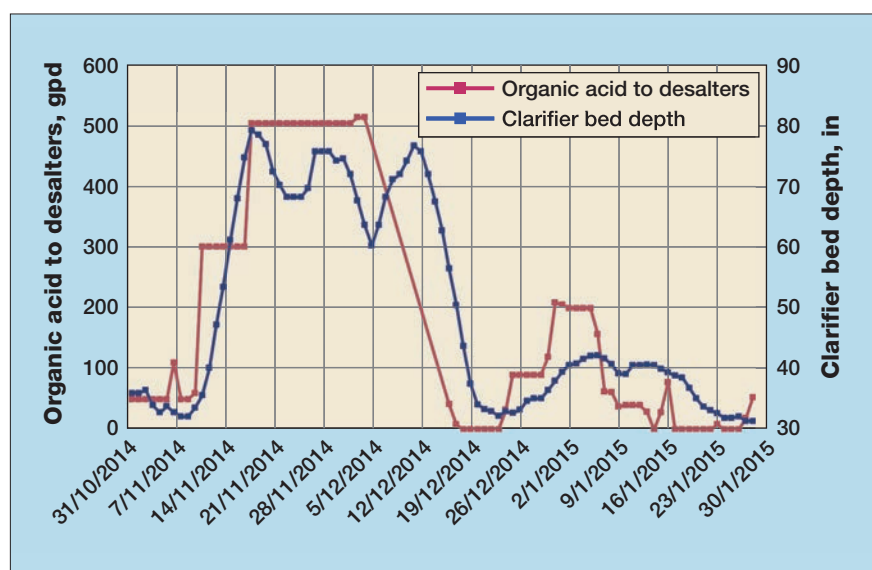


Figure 6 Desalter to waste treatment correlation

an operational bottleneck, preserving flexibility in operation and crude oil purchases and maintaining good quality brine from the desalters.

Improve integrity and economic operating windows

Crude tower operations are faced with several constraints, including operating windows, economic demands, and several factors that can affect reliability and efficiency. Using connected data and taking advantage of Big Data analysis techniques, an analyst can now develop more dynamic analytic analyses to simultaneously expand operational flexibility whilst maintaining mechanical integrity within the bounds of these constraints. The analyst should be able to more easily identify the operating points to maximise production goals, such as producing more diesel fuel, while avoiding conditions that impact reliability, such as crude tower overhead corrosion and salting. Further system improvements can also be made by identifying correlations and root causes, and opens up the ability to perform predictive analytics on refinery operations. Simply put, utilising connected data and advanced analytics can help to better explain the state of a system, determine when action is needed to improve the system, and define the factors that are the main drivers to affect positive change.

Case 4: improving reliability by improving integrity operating window definition

A North American refinery was experiencing reliability issues in a crude unit overhead system. Fouling and corrosion impacted unit reliability, causing short equipment life and increased unplanned maintenance costs. A solution was desired that identified the root causes and helped to determine effective corrective actions.

Traditional techniques deployed identified salt-induced fouling, under deposit corrosion, and high overhead velocities as root causes. Amine speciation testing revealed over a dozen types of amines present in the overhead system originating from multiple sources contributing to salting. Sources of the problem amines included those injected into the refinery for neutralisation, introduced as contaminants in the crude oil stream, and recycled and degraded amines from various refinery circuits.

A number of actions were implemented to reduce or eliminate the problematic amines and chlorides contributing to salting to reduce salting from a chemistry perspective: the desalter operation was modified to improve crude oil contaminant removal, and the refinery steam amine programme was changed to reduce impact on overhead salt formation.

In parallel, an analytic routine was implemented to better quantify the

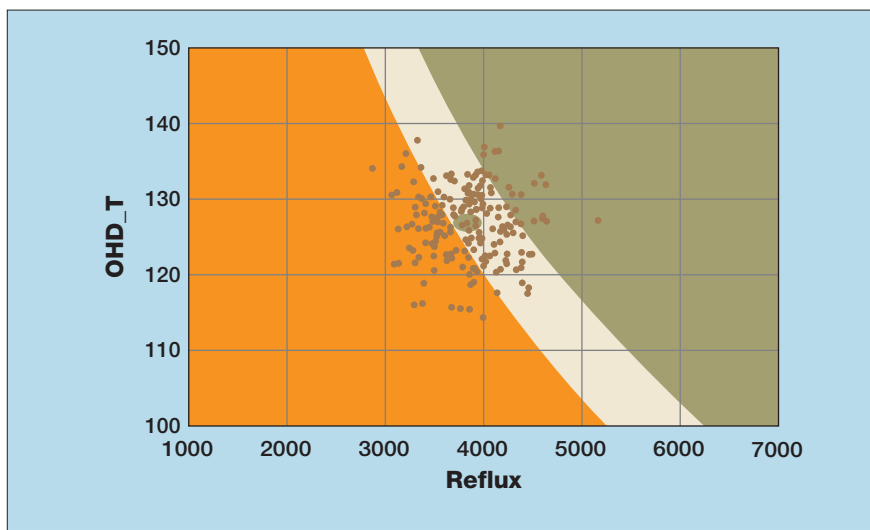


Figure 7 Base case: salting condition, 40% fractional salting

time salt formation is occurring in the system, resulting in an output variable coined as 'Fractional Salting'. Whereas traditional techniques only allowed salt point computation at specific points in time, the new analytic toolset enables visualisation of the entire 'salt potential' for the system, its sensitivities to all input variables, and their changes over time. Using the fractional salting approach, one is able to dynamically calculate the fraction of time a system is in a salting condition based on both a given set of operating conditions and the variability of each operating parameter. This provides a method to rapidly assess the current state of the overhead system in terms of salting, quantitatively determine if further improvements are needed, and numerically show the level of

improvement that can be achieved by changes made on the system to reach specific goals within the constraints of the system. Further, as the operating objectives and crude diet change over time, it can now rapidly be determined when the system leaves safe operating conditions and moves into a higher salting risk profile. When this occurs, the relative ranking of all the drivers responsible for the excursion can be easily and clearly identified, providing a path to most effectively mitigate the risk and re-establish safe operating conditions.

Fractional salting requires several operating parameter inputs in addition to what is normally considered for traditional salt point analysis, such as additional crude column conditions, product fraction flow rates, pumparound rates and physi-

cal conditions like temperatures and pressures. Since this refinery had already been providing data to the vendor's cloud based system, expanding the data sent to run the analytic was fairly straightforward. This provides several potential benefits to the refiner, including being able to:

1. Improve visibility to the risk of salting corrosion and fouling by moving from spot analysis to describing the boundaries of the salt formation given the variability in operating parameters, not just current data or averages. **Figure 7** shows the calculated salt points for a particular operating condition at this refinery. The highlighted point is the operating point at the time of measurement and the 'cloud' of data points around it indicates the amount of variability seen during an operating mode. In this case, the fractional salting was calculated at just over 40%, with most of the operating points being either in the salting zone (orange area) or below the margin of safety for salting (beige band). This is an area of concern for salting and represents a high potential for negative impact on reliability. In fact, the plant confirmed that overhead fouling and corrosion did take place during this period of operation. Note that the chart compares overhead temperature to reflux rate, but the analytic calculates salt formation potential scalar fields around the variation of all considered operating parameters.

2. Calculate the sensitivity that changing any specific operational handles, both average value and variation, makes on fractional salting. This can be used to advise on changes to improve reliability, improve yield profile, or both. **Figure 8** shows the salting potential after changing the operating point and improving variability of several operating parameters in the system. As can be seen graphically, the potential for salt precipitation has been significantly reduced with the operating point moving into the safe zone (green area) and the spread of data points shrinking. Specifically, the concentration of diglycolamine (DGA) in the overhead was reduced

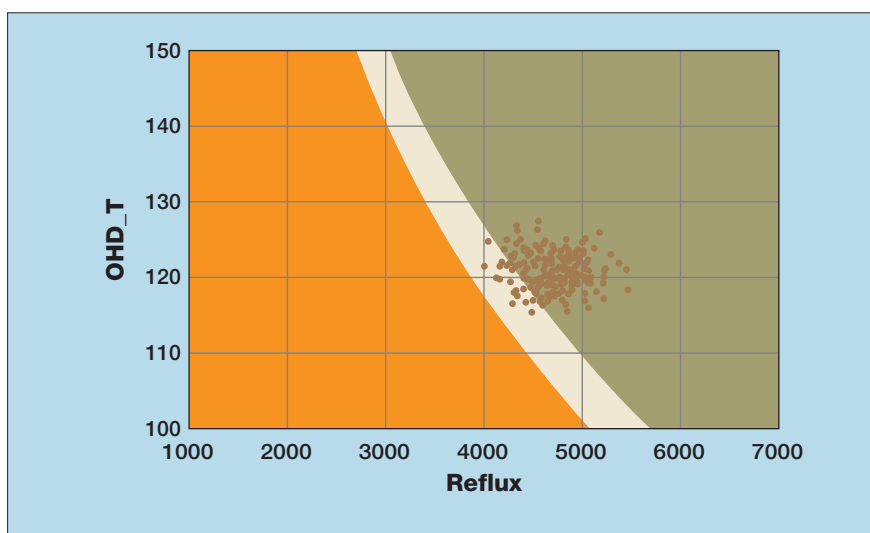


Figure 8 Improved case: fractional salting reduced to 6.5%

and tighter control of several operating parameters to reduce variability improved fractional salting from over 40% to 6.5%.

3. Given a set of defined constraints around any or several operating parameters, calculate an optimised case that can achieve multiple goals, such as reducing fractional salting below 5% while maximising the amount of distillate produced. This allows advisory control for the crude unit overhead, to enable meeting operational goals while avoiding negative impact to reliability due to salt formation.

Conclusion

Connecting data with a common knowledge management platform allows improved visibility, accelerates response, and enables deeper analysis to improve refinery operations, reliability, and profit potential. Bringing dark data to light reveals trends that would otherwise go unnoticed, drives visibility to the current state of system performance, and clearly identifies efficient paths toward optimised systems to

improve yield profiles and preserve system integrity. As this field grows in industry, and specifically in refineries, new opportunities for data scientists and analytics are emerging, with the potential to change the face of refinery operations and usher in a new era of achieving operational calm in the face of increasing variability and tighter regulations.

Acknowledgements

The author would like to recognise the significant contributions of Dr Collin Cross, David Mason, Henry Grichuk, and Steve Bakas for their help in providing data and feedback for this work.

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Jeffrey A Zurlo is Senior Strategic Project Manager with GE Water & Process Technologies. With over 25 years' experience in process engineering, refinery treatment sales and services, hydrocarbon product applications and technical support, he provides marketing and technical support to the refining and gas processing industries. He holds a BE in chemical engineering from Stevens Institute of Technology, New Jersey.



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Handling delayed coker disturbances with APC

An advanced process control system provides automatic detection of disturbances during drum switch in a delayed coker to raise performance levels

DINESH JAGUSTE
Yokogawa India

A delayed coker unit (DCU) is an established candidate for the application of advanced process control (APC) with an attractive return on investment. The challenge in DCU-APC lies in handling the big disturbances that occur during drum switch-over and vapour heating events when the vapour enthalpy feed to the main fractionator is suddenly reduced. Timely actions need to be taken on several control loops to minimise the effects of sudden cooling of the column, which leads to large variations in coker gas oil quality as well as flow rates. The exact occurrences of these events are unmeasured and a conventional control system is not adequate to handle such big disturbances. The resulting economic losses due to quality give-away and off-spec product generation are substantial, not only in the DCU but also in downstream units where the disturbances are propagated. Hence continuous operator attention is required for managing these events.

Effective disturbance handling with APC requires DCS logic for unambiguous detection of the various events that lead to major disturbances in the downstream fractionation sections. Detected discrete events are then used for generating continuous disturbance functions, which, in turn, are utilised for multi-variable modelling as well as for predictive feed forward control, honouring multiple constraints. Since manual actions in the field are also involved in these events, the extent of disturbances varies in each coking cycle. This article also describes how intermediate variable and state

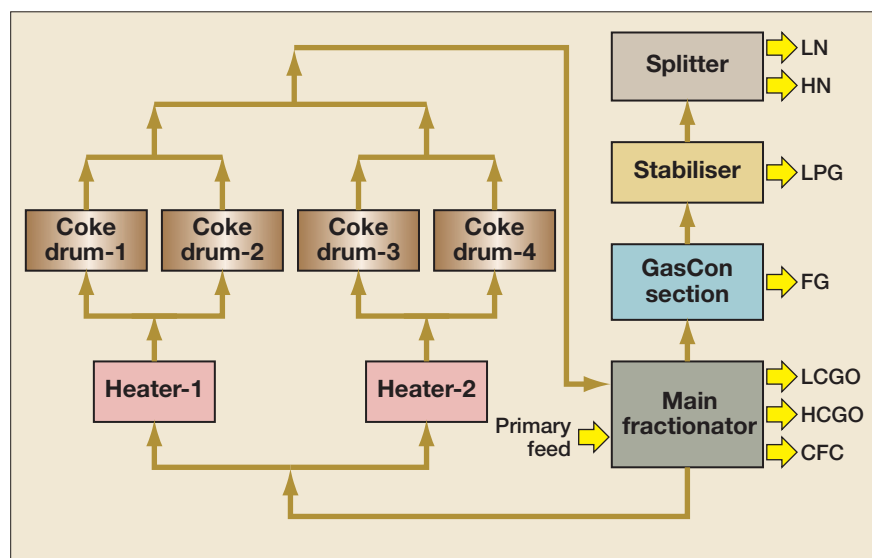


Figure 1 Block diagram of a typical delayed coker unit with two pairs of coke drums

observer (Kalman filter) concepts are utilised for robust control.

A DCU is one of the most profitable refinery units. The process involves thermal cracking for upgrading (converting) asphalt-like residue, typically from the vacuum distillation unit, into lighter distillates, coker gas oils and solid coke, which are further processed into marketable fuel products such as LPG, gasoline, diesel, fuel oil and petroleum coke.

Delayed coking is a semi-batch process where one or more pairs of coke drums are used for the thermal cracking and coking process. Simultaneously in each pair of coke drums, one drum is online for the coking process while the other drum is offline undergoing decoking. Figure 1 is a simplified block diagram of a typical delayed coker with two pairs of coke drums. Vacuum residue (fresh feed) after preheating (by exchanging heat

with run-down streams), is injected into the main fractionator bottom. The fractionator bottom is heated again in the two coker furnaces to a high cracking temperature (about 500°C), and hot, partially cracked feed flows from the coker furnace into the coke drums, where cracking continues. Cracked distillate vapour ascends in the coke drum and flows into the fractionator where it is separated into wet gas, unstabilised naphtha, light coker gasoil (LCGO), heavy coker gasoil (HCGO), and recycle oil. Coke is deposited in the drum. Two drums (one from each pair) are online at one time, accumulating coke until almost full. About every 24 hours, the filled coke drum is switched off for coke removal and the empty drum is connected. The drum that was just filled goes through a cycle of steaming out, cooling, opening, coke removal, closing, steaming, pressure testing, heating, and

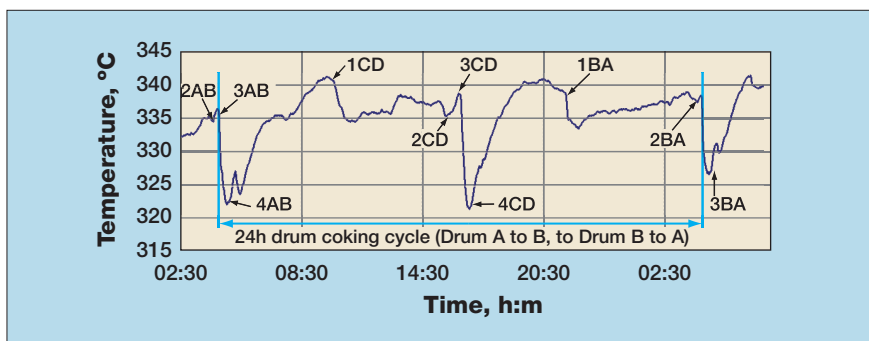


Figure 2 Variation in HCGO draw temperature during drum coking cycles of two drum pairs

finally reconnecting to the furnace and fractionator.

The semi-batch process poses unique challenges for APC in DCUs, and the challenges are discussed in this article.

Disturbances due to drum switch events

The operations of two pairs of coke drums are staggered, and the coking-decoking cycles in each pair are scheduled in such a manner that drum switch-over happens every 12 hours (twice in every day). Vapour from the furnace at about 500°C is passed through a specially designed four-way valve to the bottom of the drum which is in operation. Vapour leaving the top (after cracking and coke deposition) is quenched on temperature control to 425°C, and fed to the main fractionator bottom for separation of the product streams. About six hours before the drum switch-over, a portion of (quenched) hot vapour from the operating (on-line) drum is diverted from flowing to the main fractionator towards the offline empty drum for gradual 'vapour heating'. The hot vapour flows from the top vapour line of the empty drum and the condensate is removed from the bottom. The empty drum is gradually heated from about 150°C to about 300°C for six hours before eventual complete switch-over after every 24 hours. During drum switch-over, the flow of hot vapour to the top of the empty drum is stopped and is instead diverted to the bottom of the empty drum through a four-way valve gradually in three steps. In the first step, one-third of the vapour is diverted to the empty drum and two-thirds to the filled drum. In the second

step, two-thirds of the vapour is sent to the empty drum and one-third to the filled drum. Eventually, the empty drum is taken on-line by diverting the entire (100%) vapour flow towards the empty drum and isolating the filled drum, and taking it off-line.

The load on the main fractionator thus varies widely due to drum switch events and feedstock quality fluctuations. The effects of vapour heating and drum switch events on the main fractionator temperature profile are shown in **Figure 2**, where the disturbances in the trend of HCGO draw temperature are shown for one complete coke drum coking cycle (24 hours). Two drum switches between the AB pair, from drums A to B and from B to A, occur at 24-hour intervals. Since the drum switches between two pairs of drums, AB and CD, are staggered (see **Figure 3**), two drum switches occur in a 24-hour period at 12-hour intervals. Each drum switch involves four major disturbance events (see **Figure 2**):

Event 1

The hot vapour is diverted for the purpose of vapour heating (of the empty offline drum) about six hours before the switch-over between the drums of a pair (AB or CD). The sudden and substantial reduction of vapour flow from one of the coke drum pairs disturbs the main fractionator.

Event 2

Once the drum is warmed up with vapour heating for about six hours, the vapour flow to the empty drum is stopped by closing the condensate drain (at bottom). Then, about one-third of the total hot vapour

from the on-line drum is diverted to the bottom of the empty drum, by partly opening the four-way valve (at the bottom). As the direction of flow is changed, the lighter, uncondensed vapour hold-up inside the empty drum is flushed out, thereby suddenly increasing the vapour load on the main fractionator and increasing the temperature profile.

Event 3

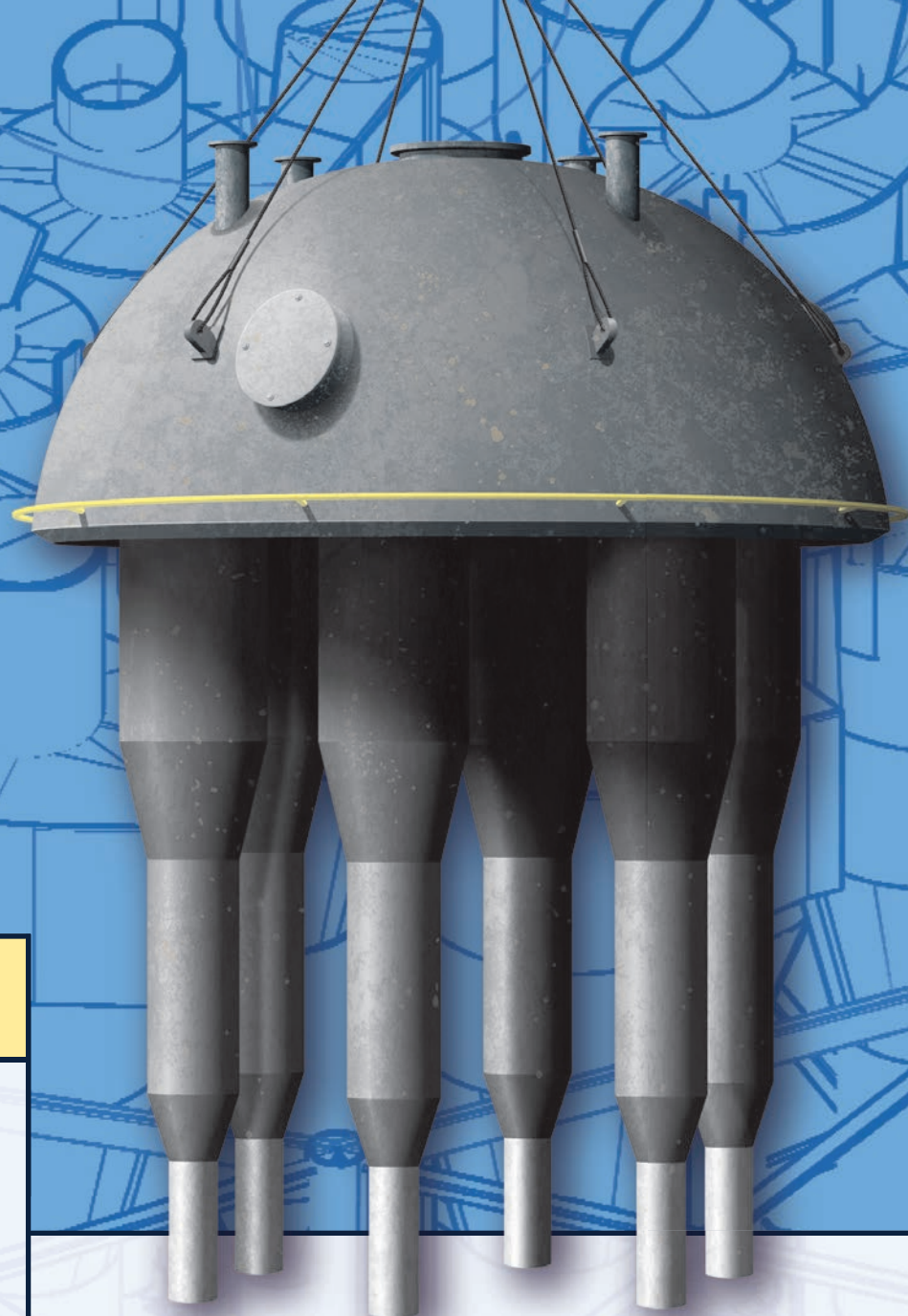
The drums are completely switched. Vapour from the furnace is completely diverted towards the empty drum and the filled drum is isolated, thereby taking the empty drum on-line. The temperature in the drum is around 400°C, which is much less than that required for cracking. The cracking reaction is abruptly quenched, causing a major disturbance to the main fractionator as both the heat and vapour mass flow are suddenly reduced.

Event 4

As the empty drum heats up, the cracking reaction gradually resumes. The temperature and vapour flow to the main fractionator increase gradually over the next 3-4 hours.

APC design challenges

The main fractionator is the key unit operation in the DCU and determines product yields and quality. The major disturbance to the main fractionator occurs during switch-over (see Event 3 and **Figure 2**) where the vapour flow from one of the drum pairs is almost cut off for a few minutes as the thermal cracking reaction is quenched due to lower temperatures in the empty drum. The heat as well as mass flow to the main fractionator bottom are reduced abruptly, resulting in sudden cooling of the column. The effect of this disturbance can be minimised by pre-emptive and quick manipulation of the circulating refluxes, product flows and other heat duties in order to restore the mass and heat balances. If timely actions are not taken, the column tends to cool down, resulting in quality give-away and off-spec generation due to condensation of the valuable



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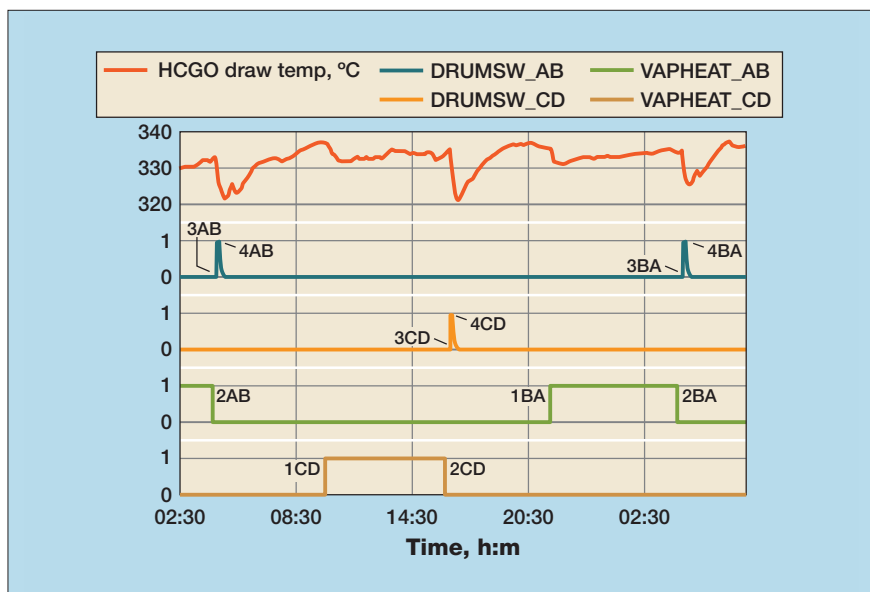


Figure 3 HCGO draw temperature vs (inferred) drum switch disturbances

lighter components into heavier product streams. The sudden changes in product flows can also lead to upsets in tray and side stripper levels. The large variations in product qualities and flows result in economic losses not only in the DCU but also in downstream units where the disturbances propagate.

The conventional (feedback) control system is not adequate for handling these sudden and big disturbances. Hence continuous operator attention is essential. APC comprising multivariable predictive control (MVPC) and inferential quality estimators has long been established as a promising solution for stabilisation of the DCU. However, design and implementation of APC for the DCU involves unique challenges discussed in the following sections.

Automatic detection of drum switch events

The automation of drum switch disturbance rejection is the most important part of DCU APC. However, there are no specific measurements for directly identifying the various drum switch events. These events must be logically inferred from various switch positions and measurements in each pair of coke drums. Furthermore, events detection must be unambiguous as false detection would lead to unnecessary control actions, which may cause upsets during stable operation. The various events described above and shown on the HCGO draw temperature trend in Figure 2 are detected (using logical conditions) as pulses, and continuous disturbance functions are generated to represent disturbance

dynamics. The HCGO draw temperature trend shown in Figure 2 is reproduced along with drum switch disturbance trends in Figure 3.

Step test and model identification

Dynamic models are the backbone of MVPCs. Hence, step test in stable plant conditions is crucial for APC design and performance. However, coker units are rarely at a steady state. Drum switches occur every 12 hours and each switch is associated with two major disturbances: vapour heating and drum switch. As seen in Figure 2, the temperature profile in the main fractionator is rarely steady throughout the day due to major mass and enthalpy disturbances. Furthermore, the vapour composition keeps changing with states of coker operations and also due to feedstock changes.

Consequently, the plant is available in a relatively stable condition for just 3-4 hours in a 24-hour period, requiring excessively high step test time. Also, as the operating conditions keep changing during different modes of coker operation, the dynamic models are expected to be different. Hence for robust design, it is essential to account for disturbances as inputs for modelling and obtain the 'average models' that would be valid for all stages of the coking cycle.

Since disturbance events are measured, it is possible to include these events as inputs for multiple input single output (MISO) model identification. Figure 4 shows a step test on HCGO circulating reflux (HCGO-CR) with the response of HCGO draw temperature. It can be seen that the variation in HCGO draw temperature, which is completely dominated by disturbance variables, could, in fact, be satisfactorily represented by a MISO model.

Conflicting constraints and change in control priorities during drum switch

In the main fractionator, the HCGO components are partly condensed with a circulating reflux stream, wherein part of the vapour

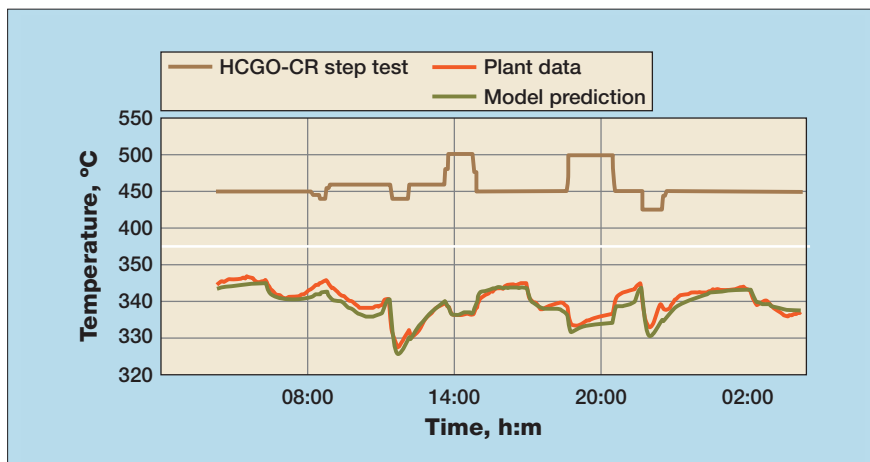


Figure 4 MISO model identification with drum switch events as measured disturbances

enthalpy is recovered at a higher temperature, mainly by preheating the feed stream and by providing reboiler heat duty for the debutaniser column.

During drum switch-over (see Event 3 and **Figure 2**), the vapour flow and enthalpy of the feed vapour are substantially reduced. In order to maintain heat balance, heat removal from the HCGO circulating reflux needs to be reduced proportionately. However, reduction in HCGO-CR is mostly constrained by the feed preheat requirement. Though the heat input to the column is reduced during the drum switch, the fresh feed to the DCU remains unchanged, which results in cooling of the feed stream. The reduction in feed preheat due to the reduction in HCGO-CR further cools the feed stream, resulting in a higher furnace heat load, which in many cases is limited by skin temperature constraint. The limitation on the reduction in HCGO-CR results in excess heat removal from the

HCGO-CR circuit, which in turn leads to cooling of the lower section of the column. The valuable LCGO components are condensed with HCGO, resulting in quality giveaway and the loss of internal reflux from the LCGO draw tray.

During stable operation, when the draw temperature is reduced, the product flow is increased to restore the temperature and maintain the quality of the product. The steady state material and heat balances are ensured with inventory and temperature (or quality) control strategies. However, as previously explained, as the column tends to cool down during the drum switch, the control priorities must be changed and, despite the reduction in temperature, the LCGO product flow must be reduced to ensure minimum internal reflux flow from the LCGO product draw tray. The priorities for temperature control should be restored once the vapour flow increases sufficiently after the drum switch-over. These issues must be

addressed satisfactorily in APC design to ensure robust automation during drum switch.

Unmeasured disturbances

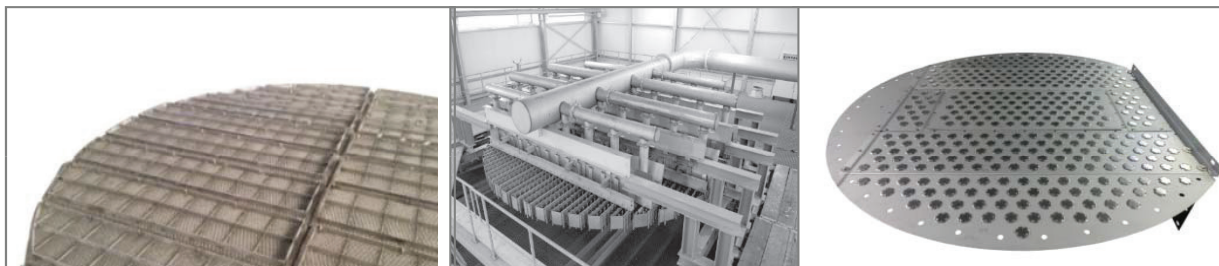
Process dynamics in the DCU are dominated by unmeasured disturbances, mainly due to drum switch events and feedstock changes. The occurrence of particular types of discrete events can be inferred from various operating conditions. However, as some of the drum switch operations are carried out manually in the field, the extent of disturbance for each event differs in each coking cycle. Hence the models obtained for these disturbances are approximated. A robust control strategy must deal effectively with these modelling errors, process or measurement noise, and unmeasured disturbances. The control algorithm for handling unmeasured disturbances forms one of the important differentiating features of MVPC technologies.

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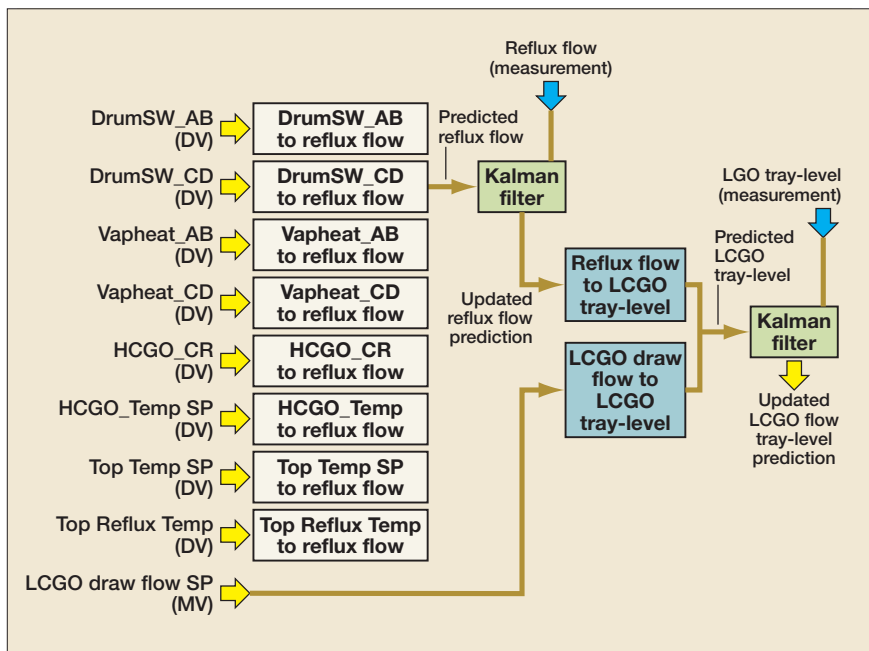


Figure 5 Grey box representation of a LCGO tray level prediction model with reflux flow as an intermediate variable

unique way of handling unmeasured disturbances with grey box model representation, involving feedback from measurements of intermediate process outputs (called ‘intermediate variables’), and also for its use of the Kalman filter (state observer) as an output feedback mechanism. There follows an example of an intermediate variable for robust control of the LCGO tray level in the presence of unmeasured disturbances in the main fractionator.

The LCGO tray tends to lose level rapidly during the drum switch event as LCGO components condense in the HCGO section due to over-cooling caused by excess heat removal in the HCGO circuit-

ing reflux circuit. The level is controlled by aggressive manipulation of the LCGO draw flow. The level loss implies a loss of internal reflux from the LCGO tray, which affects fractionation between LCGO and HCGO. However, for aggressive manipulation of the LCGO draw flow, it is essential to predict accurately the future behaviour of the LCGO tray level, which is only possible if all the variables affecting the LCGO tray level are measured and their effect on the level is accurately modelled.

The LCGO tray level is affected by several factors. While the most important variables like feedstock

composition are unmeasured, the following variables are measured:

1. Drum switch for pair AB
2. Drum switch for pair CD
3. Vapour heating for Pair AB
4. Vapour heating for Pair CD
5. LCGO draw flow
6. HCGO-CR flow
7. HCGO-CR return temperature
8. Total secondary feed flow to furnaces
9. Top temperature set point
10. Top reflux temperature

Among these variables, the most important models involving drum switch related disturbances are only approximated; hence the level control is not expected to be robust. However, with a better process understanding, it is known that all the inputs listed, except for LCGO draw flow, do not directly affect the tray level. These inputs affect the vapour flow to the top section, which in turn, depending on the top temperature set point, determine the reflux flow from the top section of the column. Thus, only reflux flow and LCGO draw flow directly affect the LCGO tray level. This process knowledge can be utilised for grey box representation with measured reflux flow as an intermediate variable, relating several inputs to the level output (see **Figure 5**).

The feedback from reflux flow (intermediate variable) is used for correcting the model prediction using a Kalman filter, which mini-

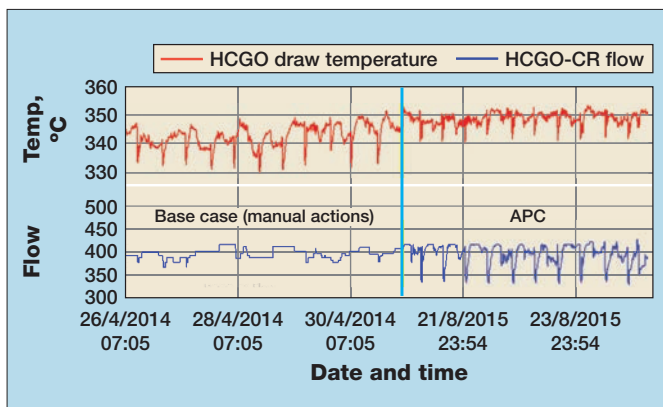


Figure 6 Improved control and maximisation of HCGO draw temperature with aggressive manipulation of HCGO-CR flow with APC

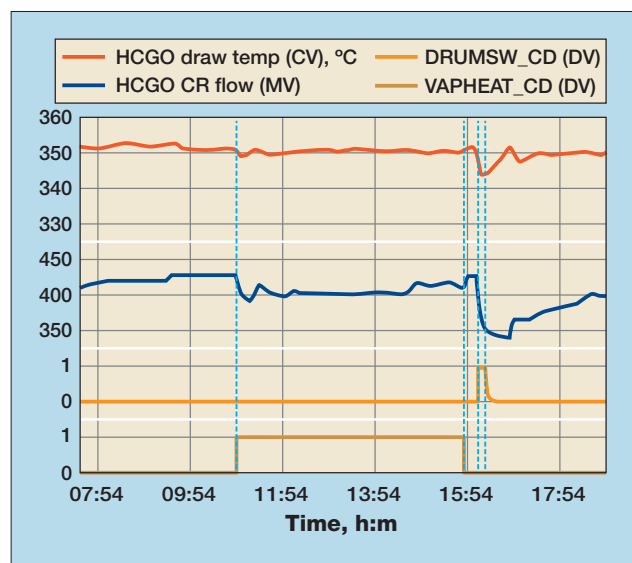


Figure 7 Control of HCGO draw temperature by feed-forward actions on HCGO CR flow for each drum switch related event

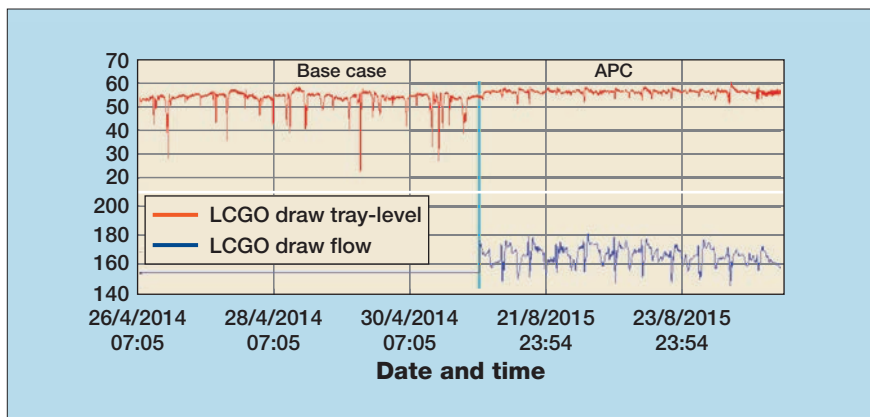


Figure 8 Better control of LCGO tray level with feed-forward actions on HCGO-CR and LCGO draw flows

mises the effect of unmeasured disturbances. Kalman filtering, also known as linear quadratic estimation (LQE), is an algorithm that uses a series of measurements observed over time, containing statistical noise and other inaccuracies, and produces estimates of unknown variables that tend to be more precise than those based on a single measurement alone. Thus the concept of intermediate variable and Kalman filter, along with process knowledge, is used for robust control in the presence of unmeasured disturbances.

Disturbance rejection with APC

The inferred drum switch events are used as disturbance variables for predictive feed-forward control. The onset of these disturbances predicts the wide variation in column temperature profile and product qualities, thereby providing feed-forward compensation by making aggressive moves on circulating refluxes and product draw flows. As **Figure 6** shows, the HCGO draw temperature control significantly improved due to timely and aggressive actions with APC (standard deviation reduced by 45%). The draw temperature is maximised by a reduction in heat removal from the HCGO-CR circuit, thereby increasing LCGO yield (by about 2%).

Figure 7 shows how timely feed-forward actions are taken with APC on HCGO-CR flow for each drum switch related event for effective disturbance rejection.

Figure 8 shows that, before APC,

due to excess cooling of the column, LCGO was condensing on the HCGO section, resulting in a loss of LCGO tray level and internal reflux. With APC, pre-emptive actions are taken on HCGO-CR and LCGO draw flows, which ensures that the level is stable with sufficient internal reflux. This ensures better fractionation between LCGO and HCGO products, with a maximum yield of valuable LCGO product.

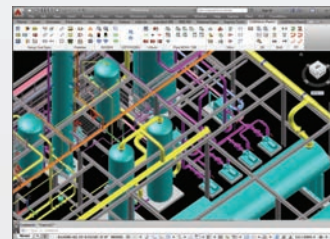
Conclusion

An APC system has been successfully implemented in a DCU with very high uptime (exceeding 95%). Automatic and unambiguous detection of various types of events involved in the drum switch, and the reliable handling of these disturbances, are the main challenges in APC design and implementation. Process stabilisation with effective disturbance rejection, while honouring conflicting constraints, results in substantial tangible benefits from product upgrading and throughput maximisation.

Dinesh Jaguste is Principal Consultant in APC with Yokogawa India Limited in Bangalore, playing a key role in the Yokogawa-Shell Global Solutions alliance for APC applications. He has a background in mathematical modelling and the design and implementation of APC systems, and is considered an expert in regulatory control troubleshooting and PID loops. He holds PhD and master of technology degrees in chemical engineering from the Indian Institute of Technology (IIT).

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Automation of tank farm systems

Automation helps to streamline operations as well as supply and demand planning, sales orders and inventory in tank management

JIJNASA PANIGRAHI and ARUN KUMAR S
Helium Consulting

Refinery tank farms are the largest units in terms of numbers of plant items in any refinery. Refineries handle a huge stock of crude oil and product inventories, which invariably involves high capital investment. Accurate inventory is critical to terminal operations and to reconciling hydrocarbons within the plant. However, the complex and dynamic nature of the operation can make it difficult to pinpoint current tank levels or predict a refinery's future inventory. This can result in increased operational expense to fulfill existing commitments and an inability to execute future orders. This article focuses on new horizons in the field of automation in tank farm management systems and also highlights the challenges arising.

Challenges in refineries

Refineries continuously strive to better their standards and uphold effective best practices in the areas of operation, accounting, loss control and overall hydrocarbon management to deliver significant cost savings and reduce their environmental impact. To improve their hydrocarbon management, refineries should continuously evaluate their perception of the principles and effects of the measurement of oil movement and storage to understand the errors involved. This should help them towards better management of hydrocarbon inventory.

Refiners are under pressure to increase refining margin by reducing inventory, optimising crude mix and reducing product quality

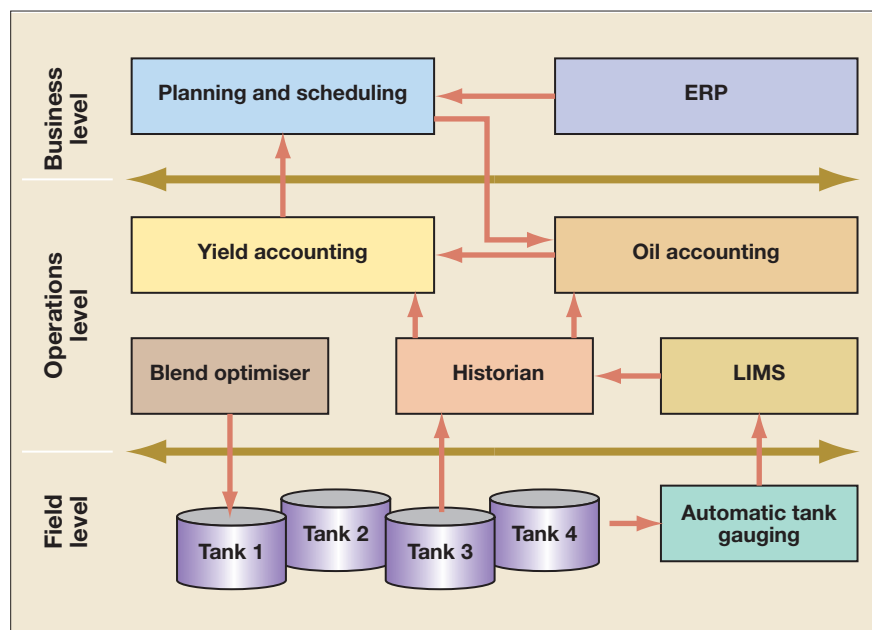


Figure 1 Typical map of integrated automation systems in the refinery

giveaway at the same time as producing more grades to tighter product and environmental specifications. Economic benefits result from improved planning and scheduling, coupled with tighter control and increased operational assurance and flexibility.

Refinery tank operations involve a list of tasks and oil movements like crude receipt and storage, process unit feeds and crude blending, process unit rundowns, intermediate product storage, component storage, blended product storage and dispatch through ships, pipeline, road trucks and rail wagons. Significant improvement in the operation and management of the off-sites can be achieved through refinery automation.

The main challenges faced are:

- Executing an effective daily reconciliation of oil movements

- Handling huge data collection and manual entry
- Quantification of individual unit-wise loss recorded
- Unaccountable loss due to errors
- Eliminating property giveaway in product blending
- Reconciliation of estimated production data with actual data from the refinery
- Optimum production planning objectives.

Automation is the solution

Automation of oil movement and storage and control system helps to:

- Reduce operating cost
- Optimise blending
- Provide better planning by opening up opportunities for optimisation in feedstock selection
- Reduce inventory
- Reduce routing error

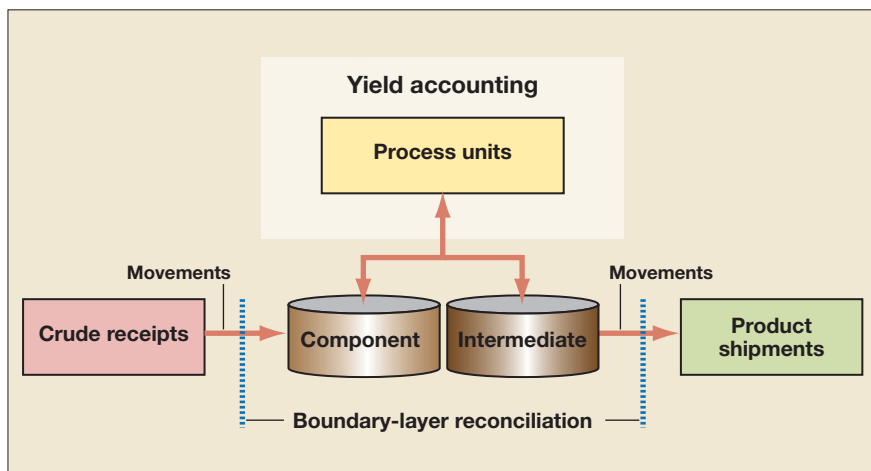


Figure 2 High level view of the components of oil and product movement in the refinery

- Provide incident and slop analysis.

Oil movement and storage automation enables all tank farm and loading bay processes to be managed and automated with maximum consistency and according to standards. This can be done on three levels:

- Field: automation at the field device level
- Operations: automation at the management execution system (MES) level
- Business: closing the loop by integrating oil accounting into the enterprise systems (ERP).

This would allow the refinery to monitor all its automated processes and ensure rapid identification of problems as well as savings in both time and money. Many companies have already implemented field automation and enterprise ERP systems that are capable of handling the first and third levels. In this article, we are discussing the second level, the operations level (see **Figure 1**).

Data gathering, reconciliation, yield accounting or oil accounting, and refinery information reporting systems are essential in automating refinery oil movement and storage systems. Automation in oil movement and storage calls for a detailed definition of complete system integration for effective functioning of the automated system. Automated oil movement and storage might include a complex business procedure involving comprehensive daily analysis of all flows, oil movement data, lab

data and inventory data. Normally, these different data types are from different sources, and integration of all these data into a single system as an input for an oil movement and storage business process is an absolute necessity.

MES helps in tracking and optimising operations with rapid, accurate and transparent movement of data in real time. It helps in better understanding how tank operations are being performed in real time.

MES helps in tracking and optimising operations with rapid, accurate and transparent movement of data in real time

This would have a positive impact on operations and business, as it would help in timely and informed decision making about inventory and production performance.

Data historian

Every refinery generates significant volumes of data but encounters difficulty in leveraging information effectively without a good MES layer. A data historian makes it easy for the refinery to handle large amounts of data generated by the plant and other areas of operation. It ensures that historical data is

readily available to users for analysis. In this case, it ensures that data is readily available for other systems and applications for oil movement and storage. For example, for movement and yield calculation, flows of streams and their respective temperatures and pressures are normally obtained from a refinery historian, and inventory data is obtained from the tank gauging system. MES is the foundation for integrating and connecting plant control systems and the shop floor with business systems.

Tank operation

The key to refinery profitability is being able to plan, schedule and optimise operations based on real-time information with feedback on what is currently happening and what has actually been done, in order to regulate refinery operations:

- Starting, stopping and tracking of all movements
- Improved automation sequences, thereby increasing operator efficiency and reducing contamination of products
- Scheduling automation of oil movements
- Reporting and monitoring of actual quantities
- Improved oil loss accounting, using better quality real-time data for material balancing
- Automated workflows to support production processes
- Automated data collection from tank and related auxiliary equipment and sensors
- Management and tracking of material within work centres and across work centres
- Generation of tank area documentation and management reporting on fuel and losses.

Automation of tank operation involves oil movement planning, oil movement order management, oil movement line-up and oil movement order monitoring and reporting. This allows the planner/scheduler to import, create, modify and monitor oil movement orders (see **Figure 2**).

Following approval, the operator executes an oil movement order.

This monitors the progress of the oil movement, terminates the movement when it is complete, then starts the next order. A major advantage is that the destination properties are calculated based on the tank properties and the source properties, and are updated by laboratory results from the LIMS system interface. The operators also have to dispatch the product, and these oil movements are dependent on the blend order being completed and the product quality being certified.

Blending optimisation

The final component in the tank tracking and optimisation sequence involves managing the blending process of various components both on the crude side as well as on the product side. A blend optimisation tool manages both real-time blending and oil movements, and is used by schedulers, technical services and by operators/supervisors in the control room. It is designed to optimise the blending operation based on component cost and component specification, reduce inventory requirements, and maximise valuable product production (see **Figure 3**).

Blend optimiser applications have always been attractive investments because of their benefits; though not always straightforward, these benefits are almost always quantifiable. The key areas where tangible benefits are realised include:

- Optimised blend recipes and reduced product quality giveaway
- Reduced margin between sales and target product specifications.

Hence total blend cost is reduced.

Oil accounting or yield accounting

Comprehensive oil accounting should feature a standard business process for data gathering and validation, as well as producing an accurate material balance for the refinery units and for publishing the required data. Four successive steps involved in oil accounting are: data gathering; identification of data discrepancy or gross error; reconciliation; and reporting.

Daily reconciled unit production will help the refinery in reaching

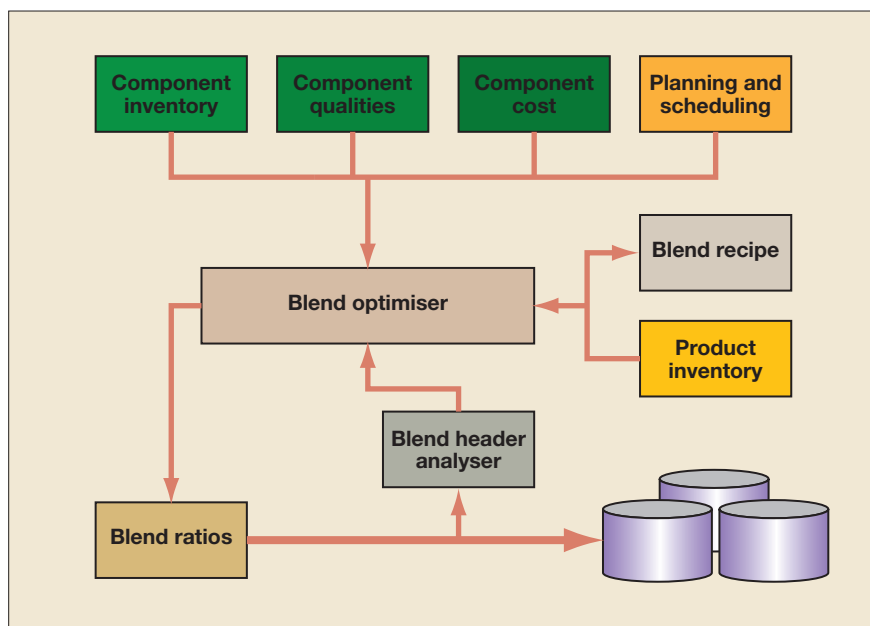


Figure 3 Typical blend optimisation flow in the refinery

commercial decisions, in price risk management, and in fuel and loss monitoring and control with more accuracy and assurance.

Reporting

An integrated MES for tank farms and terminals provides extensive monitoring and control at the enterprise level for optimal response in marketing and manufacturing, enabling optimum data transparency and accessibility at the execution level. It explicitly represents tank terminal operations as a graphical workflow and effectively synchronises, coordinates, analyses and optimises the entire tank terminal, thus enabling various operations to compile and process large amounts of data into meaningful reports.

Conclusion

Automation offers a comprehensive range of functionalities for the typical needs of tank management, to help streamline operations and for other supply and demand planning as well as sales orders and inventory management.

Robust and comprehensive oil accounting is beneficial to refineries as it can improve refinery economics in terms of optimised planning and scheduling. The most important benefit that oil accounting systems provide is the direct impact of reduced losses.

From the refinery management's point of view, direct benefits that can be achieved at the operational level include: higher productivity through better asset utilisation, shorter waiting times, and fewer errors caused by manual operation; and better control through improved distributed visibility of centralised information on all terminal operations (both real-time and historical).

Automation also offers indirect benefits, not necessarily commercial. It provides reconciled and accurate production figures that can be used by different business processes within a refinery for improved operational management and decision making.

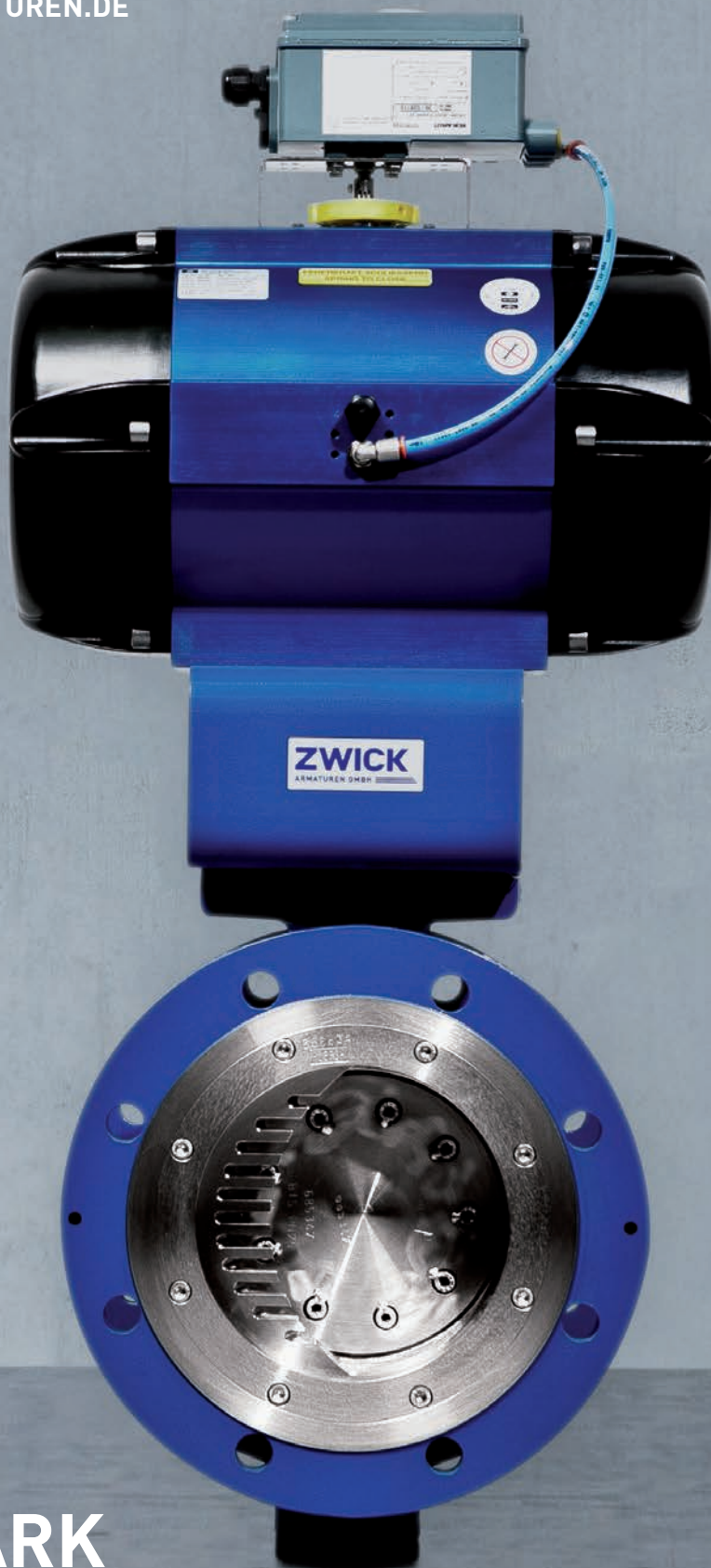
Jijnasa Panigrahi heads the Technical Services & Consulting group with Helium Consulting, Pune, India, and specialises in process modelling, production and technical services. She holds a bachelor's degree in chemical engineering from Indira Gandhi Institute of Technology (IGIT) Orissa.

Arun Kumar S is Technical Lead with Helium Consulting Private Limited, Pune, India. He has eight years of experience in refinery operation and MES. He has vast experience in the area of refinery operation, commissioning, and process software implementation and support. He holds a bachelor degree in chemical engineering from St. Josephs College of Engineering, Tamilnadu.

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Corrosion resistance in tropical environments

Tests at NASA demonstrate how super-duplex tubes provide corrosion resistance at lower cost in a challenging environment

BARINDER GHAI
Sandvik Materials Technology

Corrosion problems are creating serious issues for refineries and petrochemical suppliers that are up against fierce competition in growing economies like Asia. Because they are among the most prized industrial assets in the world, refineries and petrochemical plants must be equipped to meet changing industry demands. Suppliers are turning to innovation and collaboration to address the most pressing issues, including the latest economic developments and requirements for greater cost efficiency.

New technologies are at the forefront of this, particularly in China and India as refining and petrochemicals applications move increasingly from West to East in what some are calling the 'Asian Petrochemical Boom'. Major consumption centres are shifting to Asia mainly because of growing demand for compressed natural gas (CNG), market pressures and fiercer competition. These factors are driving companies into deeper wells with higher pressure environments.

The challenges make it ever more necessary for operators to combat problems of corrosion on the process side. Corrosion phenomena have caused numerous failures to occur in conventional steels, including in hydraulic and instrumentation tubing and particularly chemical injection lines, which are today used by most of the key oil and gas producers.

Such problems typically arise due to the presence of hydrogen sulphides (H₂S) and chlorides in process fluids, issues that are

Pitting resistance equivalent (PRE) numbers of different stainless steel grades				
Alloy	% Cr	% Mo	% N	PRE
Sandvik SAF 2507	25	4		42.5
ASTM 904L	20	4.5		34
AISI 304L	18.4			18
AISI 316L	17	2.2		24

Table 1

becoming more pronounced as refineries must process sourer and more corrosive 'poor quality' crude oils.

To address these issues, it is necessary that refinery materials evolve to yield superior performances, improved corrosion prevention and better overall production economy. The answer therefore lies in upgrading the metallurgy of key components.

The challenges make it ever more necessary for operators to combat problems of corrosion on the process side

Metallurgy and anti-corrosion

Indian refineries have traditionally opted for conventional materials such as carbon steels and 304L and 316L grade stainless steels. This is despite the fact that carbon steels are prone to general corrosion while AISI 304 and AISI 316 can succumb to stress corrosion cracking (SCC) in chloride bearing

environments. This is more so the case with the lower carbon equivalents, 304L and 316L, when they are used at high temperatures.

The normal life cycle of AISI 316L steels can exceed 100 years in non-corrosive environments. Yet the same grade has exhibited a service life of less than five years in corrosive chloride-containing environments, or less than one year in some cases as confirmed both in extensive tests and operational experience. These kinds of corrosion failures cannot be afforded by refineries and petrochemical suppliers as they face fiercer competition and increasing performance demands.

Higher content of alloying elements in general, or the introduction of nickel or ferrite into the metallic structure, can greatly improve the resistance of a steel to SCC. Steels with an austenitic-ferritic metallurgy are found to be superior to conventional austenitic grades like 316 because they have higher mechanical strength and greater stability.

By examining the Pitting Resistance Equivalents (PRE) values of stainless steels, it is possible to gain a better understanding of how resistance to corrosion phenomena, like SCC or pitting, is effected by



Figure 1 AISI 316L stainless steel and Sandvik SAF 2507 super-duplex tubing installed side by side in the Gulf of Mexico. Whereas the AISI 316 tubing showing extensive corrosion, the super-duplex tubing (grey strips in the image) showed none due to its superior austenitic-ferritic metallurgy (Schiroky, Dam, Okeremi and Speed, 2009)

the chemical composition. The PRE number is widely accepted as the best method to rank the pitting resistance of stainless steels (see **Table 1**).

PRE values are measured in accordance with exact testing procedures specified in the ASTM G48 standard. Generally, the higher

the PRE number, the more corrosion resistant the steel. The formula used to calculate PRE takes into account the content of chromium (Cr), molybdenum (Mo) and nitrogen (N) within the steel's metallurgy defined in weight as: $\%PRE = \%Cr + 3.3\%Mo + 16\%N$.



Figure 2 These images show the levels of corrosion present on the 316L tubing sample after just 18 months' exposure

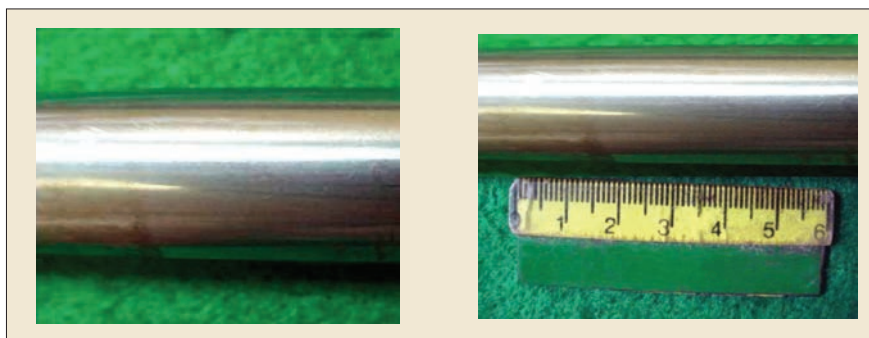


Figure 3 By comparison, the Sandvik SAF 2507 tubing sample showed no traces of additional sulphur or chlorine and no visible signs of corrosion after 18 months' exposure

The figures in **Table 1** show that AISI 304L and AISI 316L have lower resistance to pitting. Austenitic ASTM 904L (UNS S08904) has a comparably higher PRE number because it contains more molybdenum and nickel. However, the limited market availability of these elements makes the grade more expensive; this is the case with 6% molybdenum or '6Mo' steels, which are frequently chosen for H₂S processes by operators in the Middle East and Asia.

Tropical marine tests at NASA

Hydraulic and instrumentation tubes used to process oil or gas either in offshore or shoreline installations are typically subjected to high levels of salt, humidity, ultraviolet (UV) light and temperature.

This is particularly the case in tropical marine installations, where seawater evaporates, leaving a higher concentration of chlorides on the tubing. Hydraulic and instrumentation tubing made from the super-duplex material Sandvik SAF 2507 was exposed to such conditions in a test conducted in the demanding tropical marine atmosphere at the NASA Research Center located in Florida, USA.

The study saw stainless steel seamless tube samples in 316L, 904L and 2507 positioned 150ft (45m) from the mean high water mark on the east coast of Florida, directly facing the Atlantic Ocean, for a period of one year. Each tube sample was bent at 90 degrees and the ends were terminated with fittings and mounted at 30 degrees on a test rig facing the ocean.

Proximity to the ocean was typical of a shoreline petrochemical installation and close enough to ensure that the test samples would be affected by salt spray and mist throughout the 12 months. The conditions on the Florida coastline with its prevailing tropical climate provided an ideally aggressive and corrosive environment for the test samples.

After conclusion of one year's exposure, the test samples were returned to the laboratory for detailed inspection and analysis.

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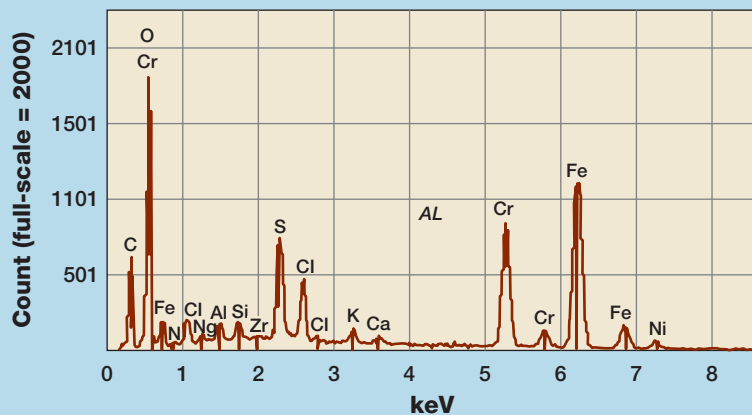
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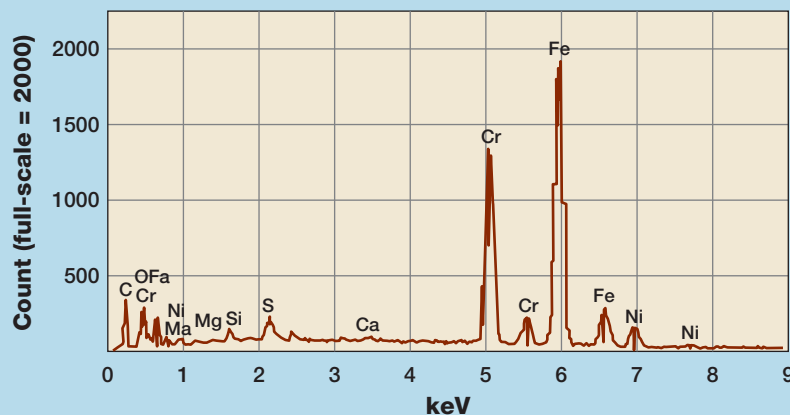
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Element	Weight concentration %
O	22.19
Na	3.01
Mg	0.64
Al	1.06
Si	0.86
S	5.96
Cl	3.82
K	1.09
Ca	0.60
Cr	18.72
Fe	39.54
Ni	2.19

Figure 4 Analysis carried out with SEM/EDS showed the presence of sulphur and chlorine in the corrosion areas of 316L tubing



Element	Weight concentration %	Element	Weight concentration %
O	1.85	O	2.93
Al	0.28	Na	1.15
Si	0.62	Mg	0.35
Al	25.42	Si	0.63
Cr	62.08	S*	1.50
Fe	6.29	Ca	0.30
Ni	3.46	Cr	25.55
Mo	1.09	Fe	60.46
		Ni	7.13

Figure 5 SEM/EDS scan readings and tables taken from the Sandvik SAF 2507 tube sample after 18 months' exposure, confirming there was no reading of chlorine or additional sulphur

Visual inspection revealed that the most corroded samples were the 316L tubes, more so on the section that had faced seawards. Corrosion damage, discoloration and pitting were found.

Some discoloration and pitting were also observed on the 904L samples, although to a lesser extent. The 2507 super-duplex material, however, exhibited no pitting whatsoever. Some slight discoloration was observed, but this easily washed off after cleaning in water. No signs of SCC were present on any of the tube samples.

When the fittings and ferrules were dismantled, corrosion products were found on the treading of the fitting and the tube showed red corrosion products where they had been in contact with the treading. Once the fittings had been removed, close examination of the tubes with an optical microscope revealed crevice corrosion on the 316L and 904L test samples, but not on the super-duplex 2507 tubes.

Gulf of Mexico tests

A more in-depth understanding of the material properties exhibited at NASA can be gained from the results of other tests performed in the Gulf of Mexico. Sandvik SAF 2507 was installed alongside AISI 316L and the corrosion behaviour of the grades was then assessed after 18 months (see **Figure 1**).

Subsequent extensive laboratory tests found that AISI 316L experienced heavy corrosion, while no signs of corrosion were detected in the super-duplex tubing (see **Figures 2 and 3**).

Super-duplex SAF 2507 and AISI 316L grades were subjected to detailed analysis using energy dispersive X-ray spectroscopy (EDS) and also scanning electron microscopy (SEM). This revealed the presence of sulphur and chlorine in the corrosion area on the 316L tube sample, measured in the graph and table in **Figure 4**.

The SEM/EDS scans revealed no visible signs of corrosion in the Sandvik SAF 2507 tube sample (see **Figure 5**). The scans confirmed that no harmful corrosion agents were present after 18 months' exposure

in the demanding Gulf of Mexico environment. There was no chlorine present and only the original sulphur content of the material.

The proven corrosion resistance of Sandvik SAF 2507 makes the grade increasingly justifiable as a solution for economic considerations, such as operational reliability and long service life. The steel's Mo content is 4%, which gives it performance levels comparable to more costly 6Mo austenitic stainless steels.

Conclusions

The NASA Research Center test demonstrated clearly that super-duplex 2507 delivers a similar performance in seawater to more expensive, highly alloyed austenitic stainless steels, despite the lower overall content of costly constituents such as nickel and molybdenum.

The coastline of Florida – with its tropical marine conditions of high air temperature, high average

relative humidity and high precipitation volumes – provided an ideal test environment. In these conditions, super-duplex SAF 2507 was proven to be an outstanding and cost-effective material for hydraulic

The proven corrosion resistance of SAF 2507 makes the grade increasingly justifiable as a solution for operational reliability and long service life

and instrumentation tubing when the marine environment proves too severe for 316L and 904L materials.

The findings are especially relevant as companies seek to upgrade

their materials to more advanced metallurgies that are able to better withstand general corrosion, SSC and pitting, while also achieving enhanced performance and cost advantages.

While these requirements are addressed by 6Mo grades to some extent, the materials are too expensive to be truly cost-effective. Instead, the alternative super-duplex SAF 2507 is better equipped to help operators improve their production economy against increasing global competition.

Sandvik SAF 2507 is a mark of Sandvik Group.

Barinder J S Ghai is the Regional Technical Marketing Manager, EMEA with Sandvik Materials Technology based in Birmingham, UK. He has over 20 years' experience in industry including project execution, maintenance, and customer service and sales. He holds a degree in production engineering and post-graduation qualifications in sales and marketing.



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Troubleshooting standpipe aeration

Standpipe aeration plays a critical role in smooth catalyst circulation in the riser and reactor of a FCC unit. A refiner details guidelines for effective aeration

RAHUL PATIL, AJAY GUPTA and ASIT DAS
Reliance Industries Limited

Fluidisation of catalyst is most important in catalytic cracking units since the rate at which catalyst is circulated in the riser reactor configuration decides the overall performance of the unit. Difficulty in catalyst circulation can lead to a lower than desired gas to oil ratio, resulting in less conversion. Besides lower conversion, improper aeration can lead to vibrations in the riser-reactor structure, which is cause for concern from the point of view of safety and reliability of operation. Standpipes play a major role in establishing the smooth circulation of catalyst from the FCC riser-reactors to the combustor-regenerators and vice versa.

A standpipe is essentially a pipe connecting the reactor to the combustor and the regenerator to the riser through which solids flow. The solids in standpipe flow downward due to gravity. The riser-reactor and combustor-regenerator system needs to be in perfect pressure balance to get smooth circulation of catalyst in the system. The solids in the standpipe must therefore be well fluidised to ensure a high pressure build-up between the two ends of the standpipe.

The fluidised FCC catalyst emulsion descends from the top of the standpipe and undergoes a form of compression due to the increasing pressure head seen at any given depth in the standpipe. This increasing pressure compresses interstitial gas as well as gas entrapped in the pores of the catalyst. The net result is a reduction in the volume of fluidising gas

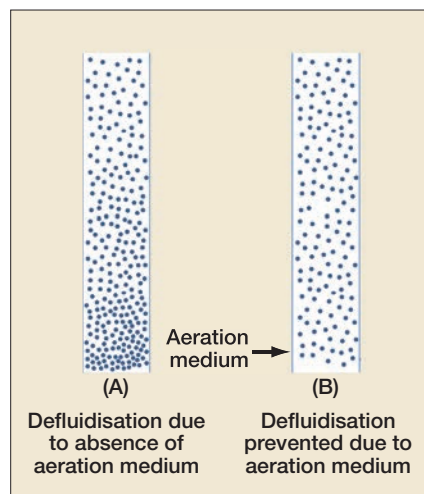


Figure 1 Effect of the aeration medium on fluidisation in a section of standpipe

surrounding catalyst particles, resulting in an increase in emulsion density. Aeration along the length of the standpipe is thus essential to keep the solids in a minimum fluidisation state. Failing to maintain the proper aeration rate to the standpipe may lead to irregular flow and poor pressure build-up. Inadequate or excessive aeration are both undesirable for fluidisation in the standpipe. Inadequate aeration may lead to a packed bed regime of catalyst due to loss of fluidising gas below the incipient fluidisation regime, and can cause non-uniform catalyst flow through the standpipe. Excessive aeration may lead to large bubbles inside the standpipe, which can cause obstruction to catalyst flow, resulting in non-uniform catalyst flow. Optimum aeration and aeration locations are therefore critical for operation of the FCC unit. The aeration rate should be just enough to restore

the catalyst emulsion to its desired volume at each location in the standpipe (see **Figure 1**).

There are methods available in the literature to calculate the required aeration rate. These require separate calculation for each small section of the standpipe to obtain the total aeration rate by adding the results of each section. We could obtain an accurate, unified, integrated expression for the whole length of the standpipe to calculate the aeration rate in one go.

Method from literature

The simple method given in the literature can calculate the aeration rate in the standpipe.¹ The following steps are involved:

Calculation of the volume of catalyst emulsion travelling down the standpipe per minute:

$$V_{emulsion} = \frac{M_c}{\rho_{mf}} \quad (1)$$

Calculation of the total volume of interstitial and inter particle gas that is circulated with the catalyst:

$$V_{gas} = V_{emulsion} \left(1 - \frac{\rho_{mf}}{\rho_{sk}} \right) \quad (2)$$

Calculation of the absolute pressure at the outlet aeration tap (see **Figure 2**):

$$P_{outlet} = P_T + \rho_{mf} g H \quad (3)$$

Calculation of gas volume change due to pressure increase across the aeration tap:

$$\Delta V_{gas} = V_{gas} - \frac{V_{gas} P_T}{P_{outlet}} = M_c \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left(1 - \frac{P_T}{P_T + \rho_{mf} g H} \right) \quad (4)$$

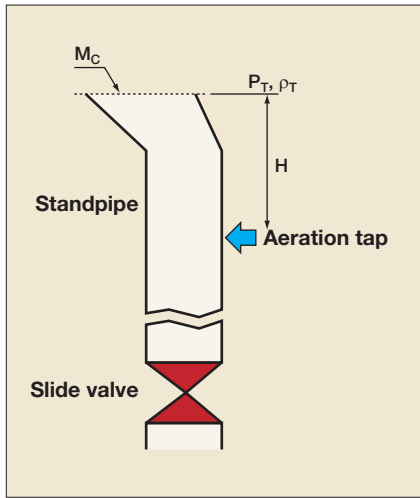


Figure 2 FCC standpipe with an aeration tap at distance H from the top

If the distance between each aeration tap, the catalyst skeleton and the minimum fluidisation density are known, one can easily calculate the aeration rate at each tap.

Calculation of aeration rate

Consider a small element of height dZ at height Z from the inlet of the standpipe, as shown in Figure 3.

The total pressure at the inlet is P_T (kg/cm^2) and H (m) is the total elevation of the standpipe up to the slide valve. Catalyst is entering the standpipe at a circulation rate of M_C (kg/h) with a flowing density of ρ_T (kg/m^3). The stripper is operated such that the catalyst enters the standpipe at minimum fluidisation condition i.e. with minimum fluidisation density ρ_{mf} (kg/m^3). So, $\rho_T = \rho_{mf}$ (normally in the range of 550-600 kg/m^3). P_i is the pressure at height Z and P_{i+1} is at height $Z+dZ$. It is required to calculate the compression of the gas and thus the additional aeration rate needed to maintain constant density of bed throughout the standpipe. dQ_a is the aeration rate required at height $Z+dZ$ to maintain element of height dZ at a desired density of ρ_{mf} . Once the requirement of the aeration rate for a small element is determined, integration along the height of the standpipe will give total aeration rate Q_a for smooth circulation of catalyst from the standpipe.

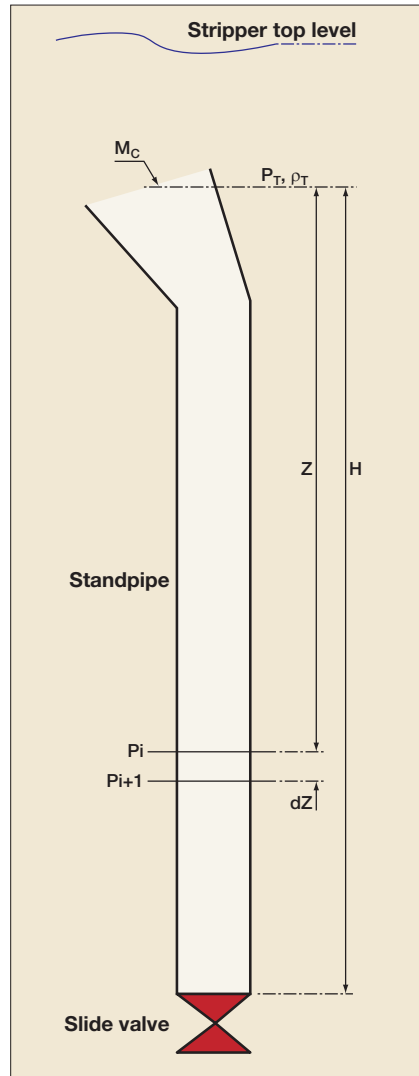


Figure 3 Vertical standpipe from stripper to regenerator in a FCC

Aeration for a small element of standpipe

The volume of catalyst emulsion at height Z in the standpipe can be calculated easily by the following relationship:

$$V_{Emulsion,Z} = \frac{M_C}{\rho_{mf}} \quad (5)$$

The volume of gas at height Z in the standpipe can then be calculated by the following relationship:

$$V_{gas,Z} = \frac{M_C}{\rho_{mf}} - \frac{M_C}{\rho_{sk}} = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \quad (6)$$

where, ρ_{sk} is the skeletal density of the catalyst.

The volume of gas at height $Z+dZ$ in the standpipe can be calculated by:

$$V_{gas,Z+dZ} = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left(\frac{P_i}{P_{i+1}} \right) \quad (7)$$

Thus, the change in the volume

of gas (dQ_a) across height dZ of the standpipe can be calculated by subtraction of $V_{gas,Z+dZ}$ from $V_{gas,Z}$:

$$dQ_a = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left(1 - \frac{P_i}{P_{i+1}} \right) \quad (8)$$

The pressure at height Z and $Z+dZ$ can be calculated by:

$$P_i = P_T + (\rho_{mf} g Z) \quad (9)$$

$$P_{i+1} = P_T + [\rho_{mf} g (Z + dZ)] \quad (10)$$

where g is gravitational acceleration (9.81 m/s^2).

The correlation for change in the volume of gas (dQ_a) across height dZ of the standpipe can be modified as:

$$dQ_a = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left[\frac{\rho_{mf} g dZ}{P_T + \rho_{mf} g Z + \rho_{mf} g dZ} \right] \quad (11)$$

It is apparent that the term in denominator ($\rho_{mf} g dZ$) is very small as compared to ($P_T + \rho_{mf} g Z$) and can thus be neglected:

$$\therefore dQ_a = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left[\frac{\rho_{mf} g dZ}{P_T + \rho_{mf} g Z} \right] \quad (12)$$

Aeration rate for the entire standpipe

dQ_a is the reduction in the volume of the gas in a small element of height dZ and is equal to the quantity of aeration required for element dZ . To determine the total aeration rate (Q_a) required for maintaining the desired minimum fluidisation condition in the standpipe, Equation 12 needs to be integrated along the height of the standpipe.

$$\therefore Q_a = \int_0^H dQ_a = \int_0^H M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \left[\frac{\rho_{mf} g dZ}{P_T + \rho_{mf} g Z} \right] \quad (13)$$

$$\therefore Q_a = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \ln \left[\frac{P_T + (\rho_{mf} g H)}{P_T} \right] = M_C \left(\frac{1}{\rho_{mf}} - \frac{1}{\rho_{sk}} \right) \ln \left[\frac{P_B}{P_T} \right] \quad (14)$$

The aeration rate required for the entire standpipe can be calculated by Equation 14. It can be seen that the aeration rate is proportional to the logarithmic ratio of pressure at the bottom and top of the standpipe.

Comparison with the literature method

A simple example is illustrated here, which uses the above relation in Equation 14 to calculate the total aeration rate for the standpipe in a FCC unit. If, in a standpipe, density of 575 kg/m^3 (ρ_{mf}) is to be maintained for smooth catalyst circulation, with catalyst skeletal density of 2500 kg/m^3 (ρ_{sk}), circulation rate of 100 T/min (M_c) and top pressure of 3.8 kg/cm^2 (P_T) for a standpipe with height of 16 m (H), then from Equation 14:

$$Q_a = 1789 \text{ m}^3/\text{hr} \text{ (29.81 m}^3/\text{min)}$$

Table 1

If one calculates the gas requirement from Equation 4 directly considering inlet pressure and outlet pressure in the standpipe, one would get a total aeration rate of $1604 \text{ m}^3/\text{h}$ or $26.73 \text{ m}^3/\text{min}$. But the sum of aeration rate (total aeration rate) calculated from Equation 4 at different aeration points along the height of the standpipe (see Table 1) comes to a value of $29.34 \text{ m}^3/\text{min}$ ($1760 \text{ m}^3/\text{h}$). This would be still clearer from a calculation for determining aeration rate at different aeration taps by both correlations.

Location of standpipe aeration

Once the desired amount of aeration medium flow is established, it is necessary to add it at appropriate locations so as to get uniform density across the length of the standpipe. For FCC particles, it is necessary to add aeration medium uniformly along the standpipe to get uniform void fraction in the entire standpipe. If aeration is added at only one location, a large bubble may form at the aeration point. The large bubble may obstruct the path of solid catalyst particles. Therefore the aeration medium is added at several locations to avoid formation of a larger bubble and provide sufficient area for the aeration medium to dissipate through the very fine particles at a rate almost the same as the rate of addition at that time.

In a commercial FCC unit, the amount of aeration is divided at several parts via aeration taps located approximately two to three metres apart. If the distance between several aeration taps is known, one can easily determine the aeration flow rate at every aeration tap, an example of which is shown in Table 1.

It is clear from Table 1 that the total aeration rates calculated from Equations 14 and 4 determined from the aeration rate for individual aeration taps are close enough compared with the total aeration rate calculated directly considering inlet and outlet pressure for the standpipe. Integration over the length of the standpipe enables one to calculate the total aeration rate required for the standpipe directly from Equation 14 as against Equation 4.

Conclusion

From this discussion, it is apparent that the smooth operation of a standpipe requires appropriate aeration to maintain constant void fraction throughout the length of the standpipe. A simple expression has been derived for determining the aeration rate at different

Aeration rate at aeration taps

Tap number	Tap location from Inlet	ΔH , metres	Pressure, kg/cm^2	Aeration required, m^3/min of aeration medium	
				Eq. 14	Eq. 4
1	1.5	1.5	3.89	3.00	2.97
2	4.0	2.5	4.03	4.86	4.77
3	6.5	2.5	4.17	4.69	4.61
4	9.0	2.5	4.32	4.53	4.45
5	11.5	2.5	4.46	4.38	4.31
6	14.0	2.5	4.60	4.24	4.18
7	16.5	2.5	4.75	4.11	4.05
Total	16.5			29.81	29.34

aeration taps. It can be used to evaluate the theoretical aeration rate at each aeration tap in the standpipe.

Reference

1 Mott R W, Troubleshooting FCC standpipe flow problems, *Catalagram* 106, 11-20, 2009.

Rahul C Patil is Senior Manager of the Refining R&D group with Reliance Industries Ltd in Navi Mumbai, India. He holds a masters in chemical engineering from IIT Kanpur and a degree in chemical engineering from ICT Mumbai.

Ajay Gupta is Assistant Vice President, Refining R&D at Reliance Industries Ltd, Jamnagar, Gujarat, India. He has bachelor's, master's and doctoral degrees in chemical engineering from IIT, Delhi, India.

Asit K Das is Vice President and heads the Refining R&D at Reliance Industries Ltd, Jamnagar, Gujarat, India. He holds a degree in chemical engineering from Jadavpur University, West Bengal, a master's in chemical engineering from IIT, Kanpur, and a doctoral degree in chemical engineering from Ghent University, Belgium.



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A refrigeration compressor requires the same high-performance and fast-acting surge control technology as other gas compression circuits

In a 'normal' gas compression circuit, the fluid that is compressed in the compressor, flowing in the immediate upstream and downstream process lines and vessels, is all in a gaseous phase without any change of state from liquid to vapour or from vapour to liquid. In this 'normal' gas compression circuit, the compressor could experience surge-inducing events due to changes in the process conditions downstream of the compressor (for instance, sudden increase in process resistance), due to changes in the suction of the compressor (such as changes to gas composition or inlet pressure and temperature), or sudden changes in the driver performance (sudden drop in shaft power or speed, for instance).

In the refrigeration circuit however, there is a change of state (from liquid to vapour) that takes place immediately upstream of the compressor in the LP process chiller and another change of state (from vapour to liquid) almost immediately downstream of the compressor in the condenser. In a refrigeration circuit, the process is 'closed loop' and the gas composition does not usually change. Also, the change of state in the condenser means that the discharge resistance changes very slowly (in fact seasonally) with the changes in ambient temperature around the condenser. A possible cause of compression surge can be due to a drop in vapour available for compression, which occurs when the process load drops, thus less vapour in the process chiller is produced.

The diagram illustrates a two-stage refrigeration cycle for ethylene liquefaction. The cycle consists of the following components and flow paths:

- LP stage (Low Pressure stage):** Compressor at the top left, drawing from the LP suction drum.
- HP stage (High Pressure stage):** Compressor at the top right, drawing from the HP suction drum.
- Condenser:** A green rectangular unit receiving high-pressure gas from the HP stage at 1500 kPa. It cools the gas to 45.0°C.
- Accumulator Liquid propane:** A purple cylindrical tank that collects liquid propane from the condenser.
- Flash economiser:** A purple vertical cylindrical unit that separates liquid from gas. It receives liquid from the accumulator and the process stream.
- Process stream:** A yellow arrow indicates the flow of the process stream into the flash economiser.
- Pre-cooler:** A blue horizontal cylindrical unit that pre-cools the process stream before it enters the flash economiser. It is controlled by a Low Control (LC) loop.
- Chiller/Evaporator:** A blue horizontal cylindrical unit that provides the final cooling for the process stream. It is controlled by a Low Control (LC) loop.
- Suction Drums:** Two orange cylindrical drums. The LP suction drum draws from the chiller/evaporator (150 kPa, -32.0°C). The HP suction drum draws from the flash economiser.
- Pressure and Temperature Points:**
 - Between LP and HP stages: 480 kPa
 - Between HP stage and condenser: 1500 kPa
 - Between condenser and accumulator: 45.0°C
 - Between accumulator and flash economiser: 150 kPa, -32.0°C
 - Between flash economiser and pre-cooler: 480 kPa, +4.0°C
- Control and Safety:** The system includes several control loops (UIC LP, UIC HP, LC) and safety valves (red symbols) to manage the cycle's operation and prevent overpressure.

process designers to consider that any surge-inducing events on a refrigeration compressor are limited and slow occurring and require only simple and 'low performing' surge control technology. This is not a correct perception as it ignores the surge-inducing events that could occur because of driver problems, which tend to be fast and sudden.

Thus a refrigeration compressor requires the same high-performance and fast-acting surge control technology as 'normal' gas compression circuits.

In general, each compressor stage that supplied refrigerant to a process chiller requires a dedicated anti-surge control loop (controller plus valve).

In the example given in **Figure 1**, where the LP process chiller provides refrigerant vapour to the LP stage of the compressor and the pre-cooler provides refrigerant vapour to the HP stage of the compressor, two dedicated anti-surge control loops would be required.

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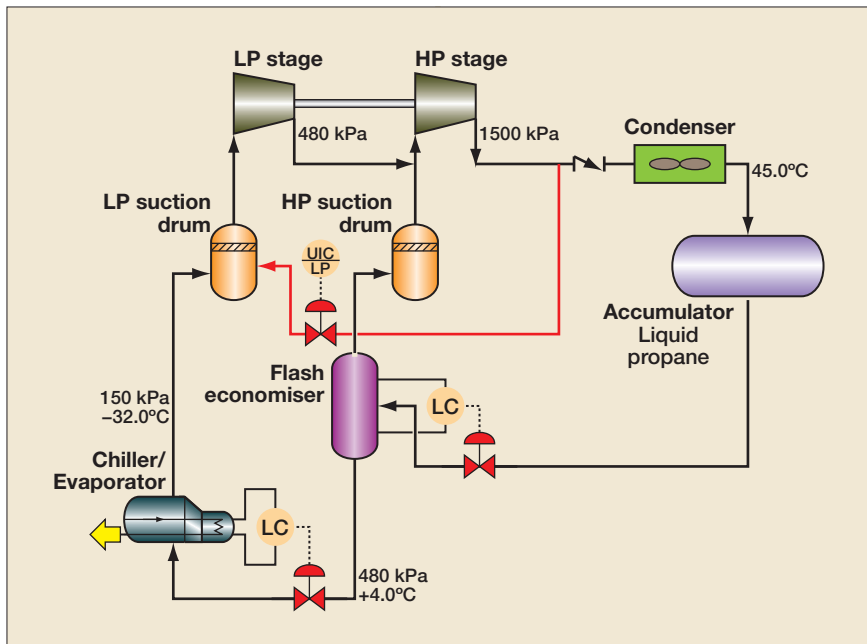


Figure 2 If a compressor stage is linked to a flash economiser only, an anti-surge loop is not required

operating conditions of the compressor when it is running. This is illustrated in **Figure 2**.

The need to provide recycle gas cooling

Method 1 – Quenching

In refrigeration compressors, it is advisable to design the anti-surge

(recycle) piping so that it commences upstream of the discharge condenser. The reason is straightforward. If recycle gas was taken from downstream of the condenser, it would be, by definition, mixed phase with mostly a liquid content and so would be unsuitable for use as recycle gas.

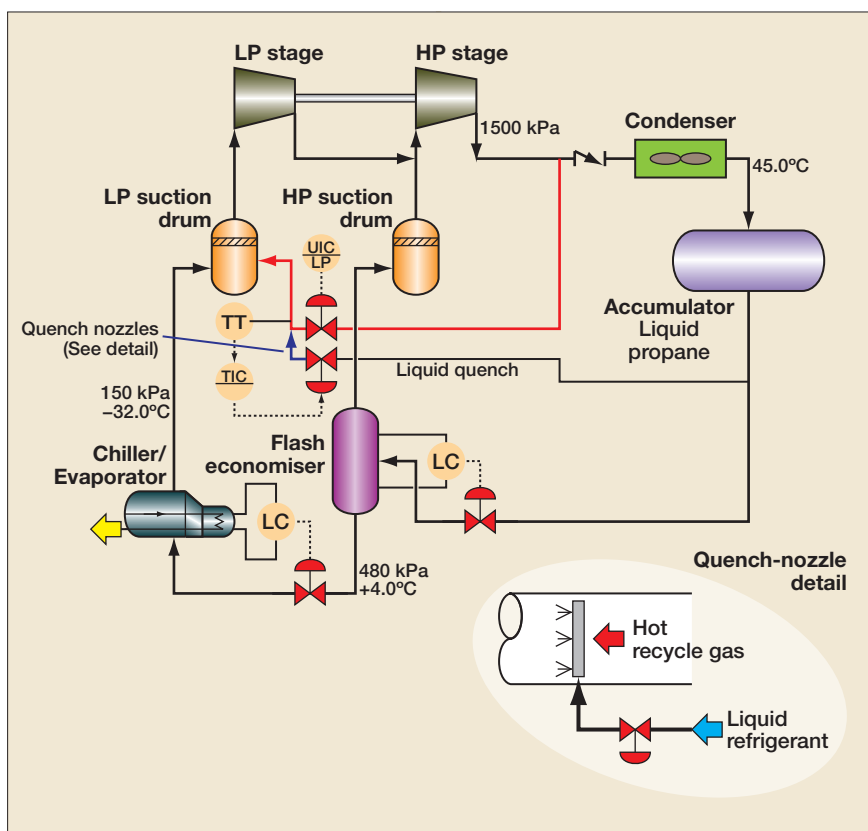


Figure 3 Evaporative cooling for recycle gas

In a propane refrigeration application, the temperature of the propane gas at the discharge of the compressor may easily reach 130-140°C, and if recycled back to the LP stage suction drum, would be problematic on two levels:

- First, the replacement of cold and denser vapour originating in the process chiller with hot recycle gas would initially drive the operating point of the LP stage further into surge
- Secondly, it would be just a matter of time before the machine tripped on high LP stage inlet temperature.

Therefore, good design practice dictates that the recycle gas be cooled to a temperature approximating the vapour coming from the process chiller. A relatively straightforward method of doing this is to apply an evaporative cooling concept in the sense of releasing an atomised fine spray of liquid refrigerant into the hot recycle gas stream downstream of the anti-surge valve. Since this almost instantaneously evaporates, it sufficiently cools the hot recycle gas stream. This is typically achieved by installing a specially designed atomising nozzle system into the recycle piping, downstream of the anti-surge valve. This is illustrated in **Figure 3**.

When quenching is used as the method to cool the hot recycle gas, it is not good design practice to simply modulate the quench valve with a slow-acting temperature controller. For example, if the surge control system step-opens the anti-surge valve, providing a sudden increase of hot refrigerant recycle gas, the much slower quench controller would allow the recycle gas to be delivered too hot to the LP suction drum for an unacceptable amount of time.

Dynamic decoupling between the anti-surge controller and the quench controller is required so that when the anti-surge valve opens by a significant amount, the quench controller would also immediately open the quench valve by a suitable and proportionate amount and then allow the slower temperature control response to

determine the required final opening of the quench valve based on the set-point.

The quench controller's control action needs to be further coordinated with the anti-surge control system so that when the anti-surge valve is closed (or open up to a configurable minimum opening) the liquid quench valve is kept closed regardless of its recycle gas temperature measurement. Also, when the compressor is stopped, the liquid quench valve needs to be forced closed.

Traditional quench controllers operated with a fixed downstream recycle gas temperature set-point. However, the actual saturation temperature of the vapour in the LP suction drum mentioned above is a function of its operating pressure. Ideally, the temperature set-point of the quench controller must be adjusted as the operating pressure in the LP suction drum changes. Not only that, it should typically be calculated slightly higher (but close to) the current

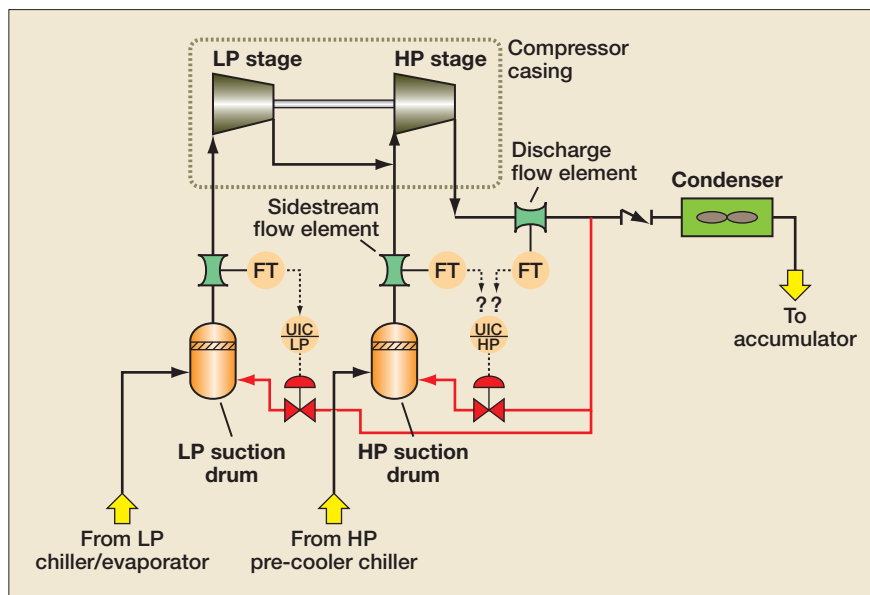


Figure 4 In many cases, it is necessary to provide dedicated anti-surge controllers for each stage of the refrigeration compressor

saturation temperature in the LP suction drum.

The latest state-of-the-art quench controllers actually calculate the temperature set-point as being offset (or biased) slightly from the saturation temperature as expressed

by the pressure-enthalpy saturation curve characteristics for that particular refrigerant.

There are, however, considerable limitations on the use of quenching that must be considered.

If the compressor is operated



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'blocked-in', that is isolated from the process chillers for any length of time, the addition of liquid quench into the necessary recycle gas will simply cause the total refrigerant inventory to accumulate and build up in what is in practice a closed circuit. Hence the shaft load of the compressor will rise and possibly exceed the capacity of the driver, causing a compressor trip.

Method 2 - Sparging

Recently, process designers started to utilise the fact that there is a continuous supply of liquid refrigerant in the LP (and even the MP) suction drums. It is possible to introduce the hot recycle gas into the lower liquid containing portion of the suction drum and design a sparging system that allows the hot recycle gas to bubble through the liquid inventory. This causes increased conversion of the liquid inventory into refrigerant vapour and, since the operating pressure of the drum does not change appreciably, the produced vapour temperature sent to the compressor inlet remains unchanged.

The sparger needs to be designed to efficiently allow the dispersion of the maximum amount of hot recycle gas as small bubbles into the liquid inventory without causing the back pressure onto the anti-surge valve to build up. In terms of sizing the anti-surge valve and the design differential pressure across it, one must consider the operating pressure of the LP suction drum plus the static pressure introduced by the liquid level that must be maintained as liquid inventory in the drum. This method eliminates the need for quenching and the associated control complexity.

Calculating suction flow rate in a multi-stage refrigeration compressor

In many cases it is necessary to provide dedicated anti-surge controllers for each stage of the refrigeration compressor. This means that the suction flowrate for each stage must be measured directly or inferred accurately.

The problem arises with the traditional design of these multi-stage machines where a single

casing envelopes the multiple stages with incoming sidestream lines (or outgoing extraction lines as the process design dictates). This is illustrated in **Figure 4**.

As may be seen in **Figure 4**, measuring the flow rate into the LP stage of the compressor is straightforward. However, for the HP stage, it is a different story. The flow element located in the side-stream line does not measure the total HP inlet flow rate and the differential pressure generated across that flow element cannot be simply added to the differential pressure generated across the LP incoming line flow element because of the considerably different pressures and temperatures experienced where these two flow elements are located.

By the same reasoning, the differential pressure generated across the discharge flow element, while directly representing the discharge flow rate of the HP stage, is also operating at completely different pressure and temperature conditions compared to the inlet of the HP stage. Therefore, a properly designed surge controller for the HP stage must include dedicated HP stage inlet flow rate inferential algorithms that produce adequately accurate results for the whole range of compressor operation.

Medhat Zaghloul is Regional Technology Manager for the Europe, Middle East and Africa Regions with Compressor Controls Corporation (CCC), based in the Abu Dhabi Office where he is responsible for the development of technical solutions and control applications and providing technical guidance and support. He has over 38 years of experience in the oil and gas industry with 15 years specific to instrumentation and controls for the petrochemical industry. He holds a BSc in electrical engineering from the Cairo Institute of Technology in Egypt (now Helwan University).

Wayne Jacobson is Global Technology Manager for the Technology Group in the Des Moines, IA, headquarters of CCC where his responsibilities include providing technical guidance and support throughout the organisation, development and improvement of control applications, maintaining and improving standard engineering design and practices, and overseeing the overall technical development of engineering staff. He holds a BS in mechanical engineering from the University of Wisconsin, Platteville.



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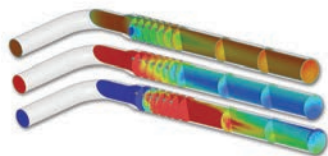
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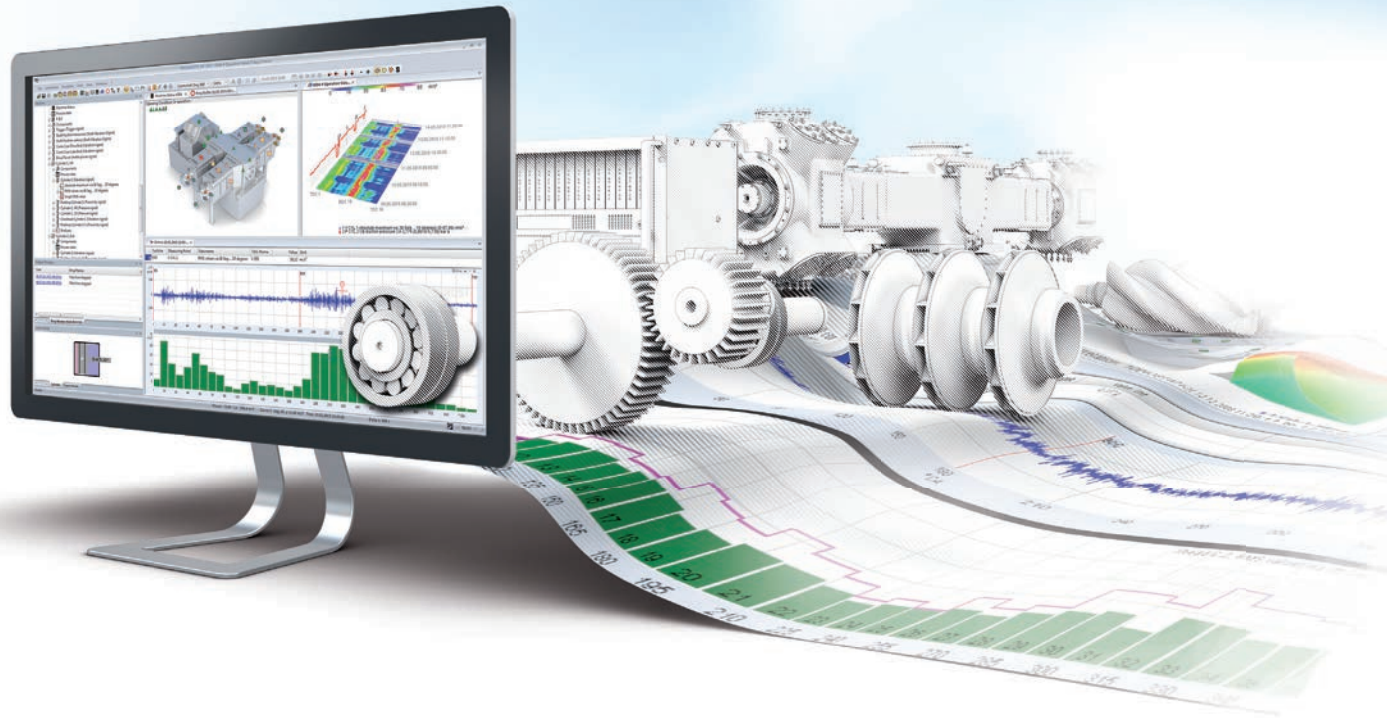
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Reciprocating compressors in a hydrogen plant

A hydrogen producer chose moderate speed reciprocating compressors for its natural gas feed and hydrogen product services

BENJAMIN WILLIAMS *Ariel Corporation*

LAURENT RICHAUME *Air Liquide Engineering & Construction*

Since the early 1990s, emissions regulations throughout the world have led to increased demand for hydrogen. This is because hydrogen is required for many desulphurisation processes. Industrial gas companies have constructed a number of hydrogen plants throughout the world to meet the demands of the refining and petrochemical industries.

The majority of reciprocating compressors used in these hydrogen plants have been lubricated, long stroke, slow speed types. This is due to historical preference and the perception that these are the most reliable compressors available. It should be noted that other types of reciprocating compressors have also been used successfully in these facilities. These include vertical compressors and horizontal, short stroke, moderate speed types.

Although lubricated compressors are the most common type used in these facilities, there are times when the design of the hydrogen plant or its components cannot tolerate oil carry-over in the gas stream. In those cases, the decision must be made whether to use a lubricated compressor with a downstream coalescing (oil removal) system or a non-lubricated (dry piston) compressor.

When determining lubricated vs non-lubricated compressors, among the items that need to be considered are the impact of oil downstream of the compressor, initial and life cycle cost, and frequency of required maintenance. Typically non-lubricated compressors will have a higher initial capital cost due to the higher cost

Feed gas compression requirements	
P suction, BarA	11
P discharge, BarA	41
Gas specific gravity	0.5727
Capacity required, Nm ³ /hr	6800

Table 1

of wear part materials. Non-lubricated compressors typically require more frequent maintenance than a lubricated compressor.

Air Liquide made the decision to purchase packaged, non-lubricated, horizontal, balanced opposed, short stroke, moderate speed reciprocating compressors for the natural gas feed and hydrogen product services at a new hydrogen plant it was constructing in Campana, Argentina.

The hydrogen plant went on-line in November 2009. This article describes the compressor selections, provides a brief overview of their design characteristics and summarises their operational history.

Hydrogen production unit

Air Liquide built a new hydrogen unit (steam methane reformer technology) in Campana, Buenos Aires, Argentina. The unit supplies hydrogen and steam to the Esso Petrolera Argentina refinery for use in the production of gasoline.

Prior to this project, Air Liquide would typically purchase vertical, slow speed compressors for this type of facility. The decision to purchase moderate speed, short stroke compressors for the Campana project was based on delivery, compact packaging and

Hydrogen compression requirements	
P suction, BarA	28
P discharge, BarA	54
Gas specific gravity	0.0741
Capacity required, Nm ³ /hr	15400

Table 2

capital cost. Spare parts availability and references for similar applications also contributed to the decision to purchase moderate speed reciprocating compressors.

The decision to use non-lubricated compressors for both the natural gas feed and hydrogen product services was based on differing requirements. Primarily, the natural gas feed service required non-lubricated compressors to prevent oil contamination of the Air Liquide process and process equipment. The hydrogen compressors needed to be non-lubricated due to customer requirements.

Compressor requirements

Operating details of compression requirements for the natural gas feed and hydrogen product are shown in **Tables 1** and **2**.

Compressor selections

Air Liquide purchased two 100% capacity, non-lubricated, non-cooled, short stroke, moderate speed, electric motor driven reciprocating compressors for each service. (Typically, Air Liquide will purchase either two 100% or three 50% units for hydrogen plants.)

The following is a brief description of the compressor selections and details of the calculated performance of each (see **Tables 3** and **4**).

Natural gas feed

Non-lubricated, short stroke, moderate speed, two stage, two throw compressor driven by a 500 kW, 750 rpm squirrel cage induction motor.

Cylinder data:	Stage 1	Stage 2
Cylinder bore dia., mm	358.78	244.48
Cylinder MAWP, BarG	56.2	87.6

Compressor performance calculations

Stage	1	2
Calculated flow, Nm ³ /hr	6876	6860
Gas specific gravity, S.G.	0.5727	0.5726
Ratio of specific heat "N"	1.2864	1.2938
Suction pressure, BarA	11	21.72
Suction temperature, C	37.1	40
Discharge pressure, BarA	22.15	41
Discharge temperature, C	96	96

Table 3

Non-lubricated, short stroke, moderate speed, two stage, two throw compressor driven by a 500 kW, 750 RPM squirrel cage induction motor.

Cylinder data:	Throw 1	Throw 2
Cylinder bore dia., mm	244.48	244.48
Cylinder MAWP, BarG	87.6	87.6

Hydrogen product specification

Stage	1
Calculated flow, Nm ³ /hr	15503
Gas specific gravity, S.G.	0.0741
Ratio of specific heat "N"	1.4047
Suction pressure, BarA	28
Suction temperature, C	45
Discharge pressure, BarA	54
Discharge temperature, C	119

Table 4

As noted previously, the compressor frame models are the same for both services. The following overview applies to both the natural gas feed and hydrogen product compressors unless otherwise noted.

Compressor frames

Model: JGT/2

Compressor stroke, mm: 114.3

Piston rod diameter, mm: 50.8

Rated speed, rpm: 1500

Operating speed, rpm: 744

Average piston speed, m/s: 2.8

Rated tension rod load, kN: 165

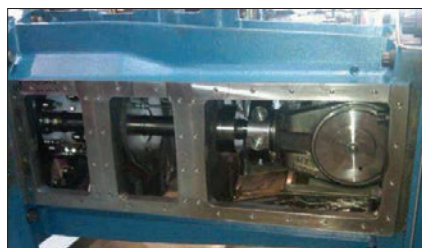


Figure 1 API-618 type "C" distance piece

(Average calculated tension rod loads were 119 kN and 103 kN, respectively)

API-618 Type "C" distance pieces (see Figure 1)

Cylinder bodies

Material: ASTM A395 ductile iron

Non-cooled

Unlined: ion-nitride hardened to approximately 57 Rc.

Surface finish: 0.15 μ m

Piston and rod assemblies (see Figure 2)

Piston material: ASTM A48 Class 30 grey iron (two piece)

Piston rod material: Low alloy carbon steel with tungsten carbide coating

Surface finish: 0.2 μ m

Piston wear bands

1 piece, angle cut

Pressure balanced



Figure 2 NL piston assembly

0.035 N/mm² wear band loading
Special polymer alloy for oil-free service

Piston rings

1 piece angle cut

Special polymer alloy for oil-free service

Packing (see Figure 3)

Water cooled packing case: carbon steel

Packing purged with nitrogen

Combination of cut and uncut packing ring sets

Special polymer alloy for oil-free service

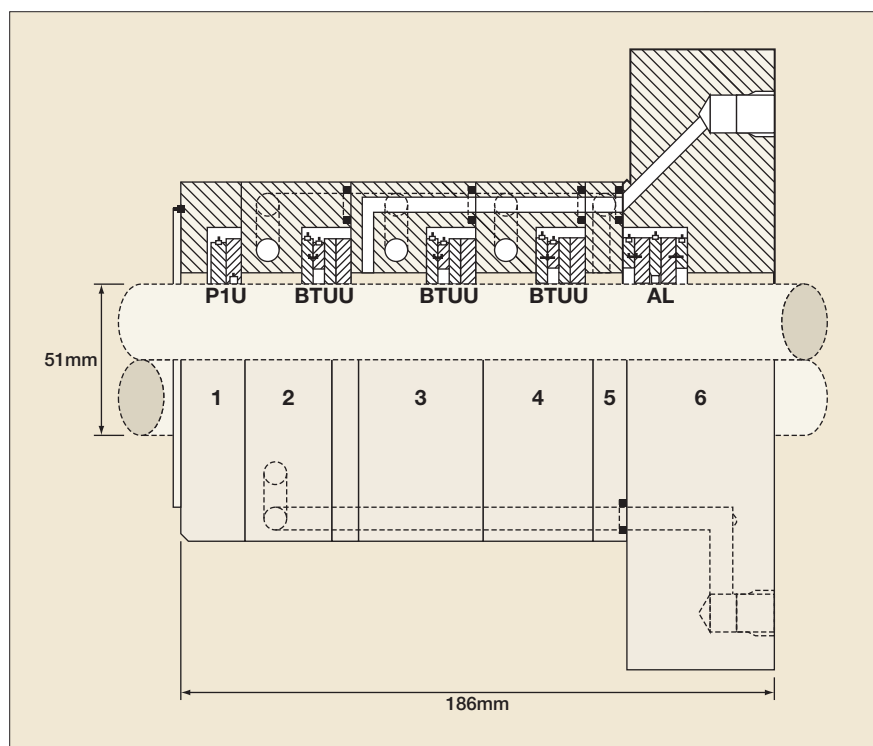


Figure 3 Pressure packing case

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Cylinder bore diameters recorded during inspections				
Dimension, mm	NG feed stage 1	NG feed stage 2	Hydrogen cylinder 1	Hydrogen cylinder
A	358.80	244.52	244.05	244.05
1	358.78	244.50	244.00	244.02
B	358.90	244.89	244.09	244.10
2	358.84	244.97	244.05	244.11
C	358.78	244.51	244.11	244.11
3	358.78	244.52	244.05	244.11

Table 6

Intermediate and wiper packing sets include buffered packing rings.

Valves

Natural gas feed compressor

Non-metallic plate valve

2.59 mm lift

Hydrogen compressor

Non-metallic concentric ring valve

0.99 mm lift

Designed for non-lubricated service

Head end valve depressor type suction valve unloaders

Thermocouples for monitoring discharge valve temperatures.

Operating history

The compressors are operating on a rotational basis; generally the units are switched every six months. The natural gas compressors are run at 100% capacity; the hydrogen compressors operate at 50% load.

As of March 2012, the plant had been in operation for approximately 18 000 hours total. The individual compressor operating hours were as follows:

Natural gas feed

Unit 1: 12 086 hours

Unit 2: 6726 hours

Hydrogen

Unit 1: 9202 hours

Unit 2: 9334 hours

Inspection results

Inspections have been performed on two of the compressors, one hydrogen and one natural gas feed. Each had approximately 9300 hours run-time at the time of the inspections.

Running gear

The running gear for both compressors was inspected and all clearances were within specified tolerances.

Cylinder inspection results

Visual and dimensional inspections were performed on each cylinder. A brief summary of these inspections follows (see **Figures 4-8**):

Piston and piston rod assemblies



Figure 4 Campana hydrogen compressor piston after 9300 hours of operation

Piston rods (nominal dimension: 50.8 mm)

Piston rods were inspected for damage and wear. Rod diameters (as measured in **Figure 5**) all were acceptable at 50.76-50.78 mm. All piston rods were cleaned and reinstalled, with no repairs needed.

Pistons

Piston ring grooves and wear band grooves were inspected. All groove

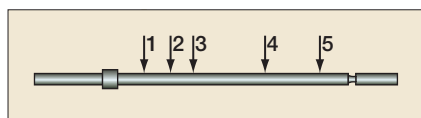


Figure 5 Piston rod inspection points

Piston ring and wear band measurements			
Dimension, mm	Cylinder 1	Cylinder 2	New
WB radial thickness	8.8-9.2	8.8-9.2	9.6
WB width	40.8	40.8	40.8
WB end gap	10.5	10.5	7.0
PR radial thickness	9.3	9.3	9.3
PR width	11.5-11.9	11.5-11.9	12.9
PR end gap	8.0	8.0	8.0

Table 5

widths and depths were within new tolerance.

Piston rings and wear bands

Piston ring and wear band radial thickness, side clearance and end gaps were measured. All showed normal wear. The hydrogen compressor piston rings (PR) and wear band (WB) measurements are shown in **Table 5**, along with the same measurements for the new replacement rings and wear bands.

It is a common perception that cooled cylinders are needed for critical service machines, particularly non-lubricated types. As noted previously, the cylinders on these moderate speed compressors are non-cooled. The worn piston ring and wear band dimensions shown in **Table 6** (see also **Figure 6**) are within the manufacturer's acceptable range. These rings and wear bands could have been reinstalled. However, as spare piston rings and wear bands were available, the decision was made to replace them.

Cylinder bore dimensions

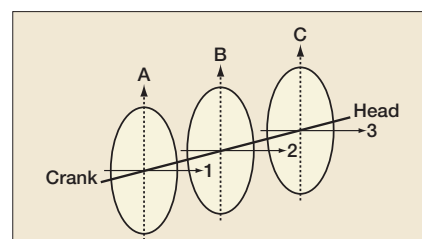


Figure 6 Cylinder bore measuring points

Packing rings

The packing rings showed normal wear. However, there were some broken springs and packing rings in one of the hydrogen compressor packing cases (see **Figure 7**). All packing cases were cleaned up and packing rings were replaced with rings of the same material.

Valves

Due to the rotating speeds of moderate speed reciprocating compressors, a common industry concern is the number of times the compressor valves open and close, as compared to those of a longer stroke, slow speed compressor. In this case, the compressor valves are cycling at 744 times per minute. The valves for both units were inspected and all valves looked good. The valves were re-installed in both compressors. Air Liquide plans to replace the valves at the 16 000 hour inspection.

Conclusion

Moderate speed compressors have performed well in hydrogen and natural gas feed services at an Air Liquide hydrogen plant in Argentina. After 9300 hours of operation, two of the four compressors were inspected and the cylinder wear parts (piston trim, packing and non-metallic valve components) showed little or no wear. This can be attributed to a number of things, including clean, dry gas at the compressor suction, wear part material selection, conservative design and application, as well as reduced piston speeds.

In simple terms, piston speed affects the 'rubbing parts'. Although the rotating speed is higher than normal for these types of applica-



Figure 7 Packing ring set – shows broken spring and packing rings

tions, the short stroke, moderate speed compressors installed at Campana have very conservative piston speeds. The low piston speeds have a very positive effect on the life of the wear parts.

Although the valves in these cylinders cycle more often than what would be considered 'typical' for the industry, these valves were found to be in very good condition and were reinstalled for continued use.

It is important to note that the moderate speed compressors detailed herein use non-cooled, unlined cylinders. The commonly held industry viewpoint is that cylinder cooling and liners are necessary for successful operation, especially in those applications requiring non-lubricated compressors. However, as the inspection results described in this article show there is an alternative design suitable for these applications.

With a conservative design and

application along with proper material selection and package design, moderate speed reciprocating compressors are very well suited for use in critical services, such as hydrogen plants.

Dimensional data included herein was taken from Air Liquide inspection/maintenance reports.

This article is based on a paper given at the 8th Conference of the EFRC, Düsseldorf, Germany.

Acknowledgements

The author would like to acknowledge and thank the following people for their invaluable contribution to this paper:

Mr Thierry Ott, E&C Project Director, Air Liquide Engineering and Construction

Mr Guillermo Tear, Chief Maintenance Engineer, Air Liquide Argentina SA

Mr Marcelo Ledesma, Technical Service Engineer, Hoerbiger Argentina, SA

Mr Greg Phillippi, Director Process Marketing & Sales, Ariel Corporation

Ben Williams is the Process Application and Account Manager with Ariel Corporation in Mount Vernon, Ohio. His responsibilities include the sizing and selection of Ariel's API618 Moderate Speed Process compressors. He has over 30 years of experience with reciprocating compressors. He is a veteran of the U.S. Navy Submarine Force and a graduate of the US Naval Nuclear Power School.

Laurent Richaume is currently Rotating Equipment Engineer with TOTAL and previously Rotating Equipment Manager with Air Liquide Engineering. He holds a degree in mechanical engineering from Institut National des Sciences Appliquées de Toulouse, France.



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SRU simulation: getting the properties right

Reliable, predictive simulation software for SRUs requires recognition of many of the unique properties and behaviours of sulphur

ANAND GOVINDARAJAN, NATHAN A HATCHER, CLAYTON E JONES and G SIMON WEILAND
Optimized Gas Treating, Inc.

There are several commercially available tools for simulating the performance of sulphur recovery units (SRUs). The general basis for these simulators is often regression to a collection of plant performance data. However, the results are only as good as the data on which they are based.

Collecting accurate SRU performance data is fraught with difficulty, much of it associated with the collection and analysis of samples. For example, gas samples from the reaction furnace must be very rapidly quenched, otherwise hydrogen keeps reacting. Samples containing water must have the water quickly removed to prevent further reaction. Careful attention must be paid to the metallurgy of sample containers because they can have catalytic activity — early catalysts were iron based, and even stainless steel is reactive towards sulphur dioxide. Material balances around SRUs require special techniques to work around analytical data limitations. Even when the data are accurate, simulators that depend on this kind of regression have only limited reliability when extrapolated to conditions outside the range of the basic data.

We have recently completed development of a simulation module for SRUs that has a more fundamentals oriented basis, including reaction kinetics of COS and CS₂ formation and ammonia, BTEX, and hydrocarbon destruction in the furnace. The furnace model includes non-equilibrium conversion of hydrocarbons to COS, CO, and other species as well as burner mixing characteristics, which affect

ammonia destruction. The Sulphur Converter model predicts profiles of dew points and conversion through converter beds as well as COS and CS₂ destruction profiles. Sulphur condensers have rigorous sizing and rating integrated into their simulation. The solubility of H₂S, H₂S_x and SO₂ in all liquid sulphur streams is calculated, with particular application to sulphur pits where air or another carrier can be used to sparge and sweep these volatile species from liquid sulphur. The SRU model can be fully integrated with upstream acid gas removal and enrichment processes, and with downstream TGTU quench and tail gas amine treating. Through the use of detailed, highly non-ideal aqueous ionic chemistry, it includes the effects of SO₂, heat stable salts, and monomethyl-monoethanolamine (MMEA, a degradation product of MDEA) on the performance of these units.

Developing reliable, predictive, simulation software for SRUs necessarily requires recognition of many of the unique properties and behaviours of sulphur. To a large extent, the uniqueness of these properties stems from the fact that sulphur occurs predominantly in three forms: S₂, S₆, and S₈. In liquid form, as temperature increases, the S₈ allotrope will polymerise to higher molecular weight forms. Its distribution amongst these allotropes is temperature dependent, and transition between forms occurs spontaneously, accompanied by substantial enthalpies of reaction. This article addresses several properties of sulphur.

Molecular formula and molecular weight of sulphur vapour

The most fundamental property of any compound is its molecular weight. Sulphur exists in several forms under conditions prevalent in SRUs. In the temperature range of interest in SRUs, sulphur vapour exists predominantly in the allotropic forms S₂, S₆, and S₈ with temperature dependent distribution amongst these forms. Thus, its molecular weight is temperature dependent and can be expressed in terms of the average number of sulphur atoms per molecule of sulphur. The enthalpy changes associated with the conversion of one species form into another makes the Claus reaction endothermic in the reaction furnace where the S₂ form dominates, but exothermic in converter beds where the S₆ and S₈ forms dominate. Conversion between forms also makes the molar flow rate of sulphur bearing streams variable because sulphur is not simply 'elemental sulphur' — it is actually a mixture of the different forms with variable molecular weight. This effect, however, is normally a minor one because sulphur concentrations in gas streams usually are not high, with vapour composition dominated by water and nitrogen from the air used in the combustion of H₂S to SO₂ in the furnace.

Figure 1 shows a comparison between the temperature dependence of the average number of atoms of elemental sulphur per mole of molecular sulphur as reported in the *GPSA Data Book*,¹ the experimental data^{2,3} and the simulator model. The model is

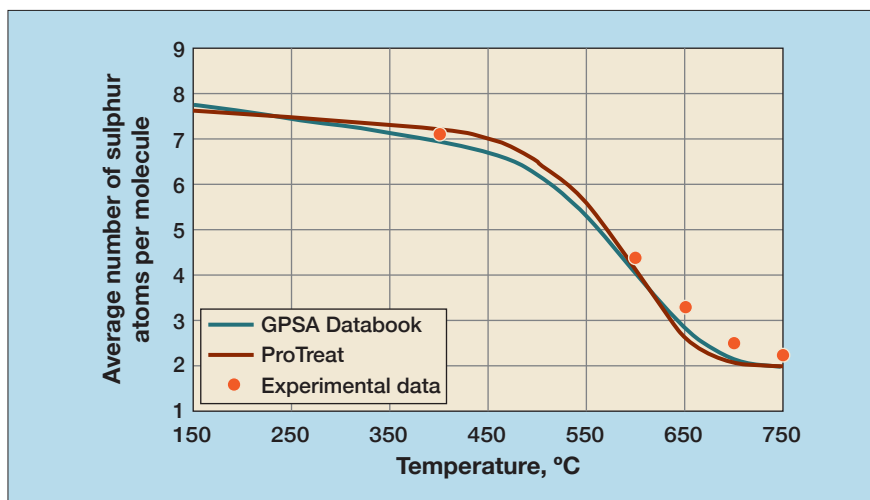


Figure 1 Temperature dependence of number of sulphur atoms in average molecule

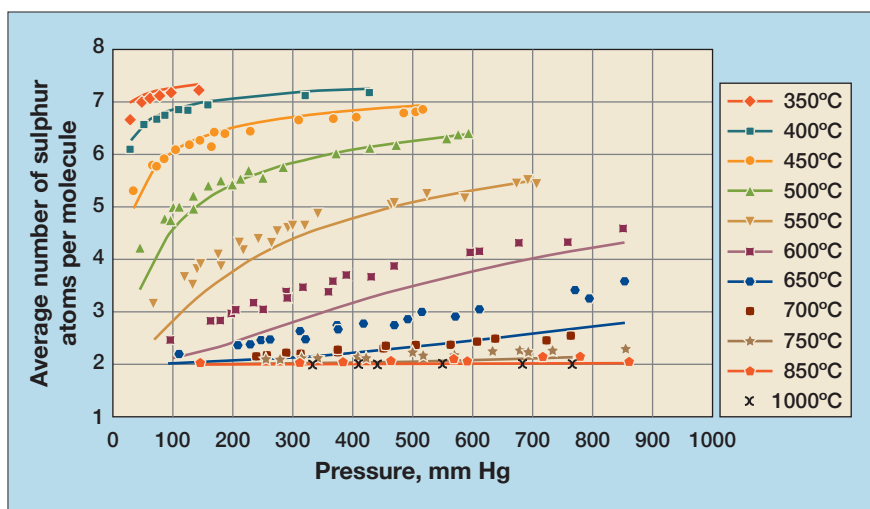


Figure 2 Experimentally measured sulphur molecular composition. ProTreat simulation predictions are solid lines

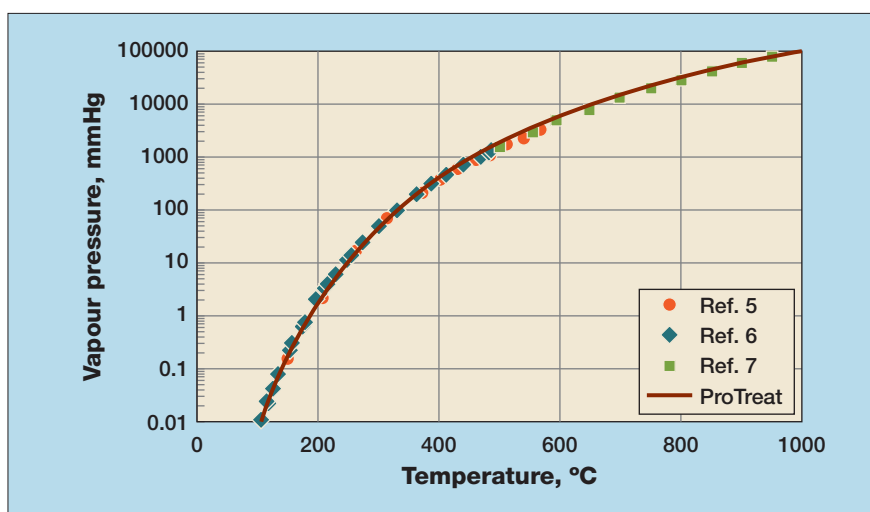


Figure 3 Experimentally measured vapour pressure of molten sulphur and the simulation model calculations

based on the reaction equilibria:



The reaction equilibrium constants

have the following form where, because pressures are low, vapour phase mole fractions rather than fugacities can be used with equal validity:

$$K_{P,1} = \frac{[y_{S6}]^{1/3}}{[y_{S2}]} P^{-2/3} \approx P^{-2/3}$$

$$K_{P,2} = \frac{[y_{S8}]^{1/4}}{[y_{S2}]} P^{-3/4} \approx P^{-3/4}$$

Here P is the total pressure and the y_i are mole fractions. The temperature dependence of the equilibrium constants obeys a van't Hoff type of equation with temperature dependent Gibbs free energies of formation regressed from tabular data⁴.

Without looking at the original references, it is not commonly known that **Figure 1** is at a pressure equal to the vapour pressure of sulphur at temperatures below its normal boiling point, and equal to one atmosphere at higher temperatures. The points in **Figure 1** have been interpolated from the experimental data of **Figure 2**. The data in **Figure 2** are from two sources^{2,3} and are the result of P-V-T measurements of a known mass of sulphur to determine the average molecular weight of the sample of gaseous sulphur. These measurements were not included in the original tabular data⁴ that form the basis for the ProTreat model. Thus, the lines in **Figure 2** are model predictions and have been made independently from the data plotted in the figure. The simulation model is in reasonably close agreement with the measured data.

Liquid sulphur

Liquid sulphur is normally dominated by S_6 and S_8 rings, but at higher temperatures (above $\sim 160^\circ\text{C}$, 320°F) the rings open and the short linear chains begin to polymerise. As will become apparent, the opening of these rings and the subsequent polymerisation of liquid sulphur has a significant effect on H_2S and H_2S_x solubility, viscosity, and heat capacity. Dissolved H_2S and H_2S_x also affect viscosity.

Vapour pressure of liquid sulphur

An important property, and one from which the latent heat of vaporisation can be derived, is vapour pressure. **Figure 3** shows vapour pressure data from several sources, together with calculations from the simulator.

Latent heat of vaporisation

Heat of vaporisation is an important property in sulphur condenser calculations. Data have been taken⁵ and regressed for use in the simulation model. The data (Figure 4) form an unusual curve, first decreasing and then increasing with temperature. This is the result of the changing distribution of the S_2 , S_6 , and S_8 allotropes of sulphur with temperature. In reality, what is measured and presented as latent heat of vaporisation also includes the heats of reaction associated with the chemical reactions that occur as the sulphur composition changes between its allotropic forms.

Viscosity of pure liquid sulphur

Pure liquid sulphur exists mostly in the forms of S_6 and S_8 rings, but around 160°C the ring structure opens and the sulphur polymerises. This is reflected graphically in Figure 5 where viscosity data⁵ are compared with simulation results. In the vicinity of the transition temperature, the viscosity undergoes a three to four orders of magnitude change, going from being relatively easy to pump to almost non-pumpable. The simulator accurately represents this behaviour.

The basis of the solubility model is polymer chemistry coupled with vapour-liquid equilibrium. The model accounts for the anomalous behaviour of H_2S in sulphur, which shows H_2S becoming more soluble in sulphur as the temperature increases. In the region around 160°C, the normal six- and eight-membered ring structures of sulphur open, becoming linear segments. These polymerise, with the polymer chains being terminated by H_2S molecules. The polymeric sulphur chains are apparently of maximum length at about 187°C (368°F) and they start to shorten with further rises in temperature. As they shorten, the viscosity falls and the chains become shorter but more numerous and, because they all use H_2S for termination, the apparent solubility of H_2S rises as temperature increases. The reason for increasing solubility is chemical bonding of

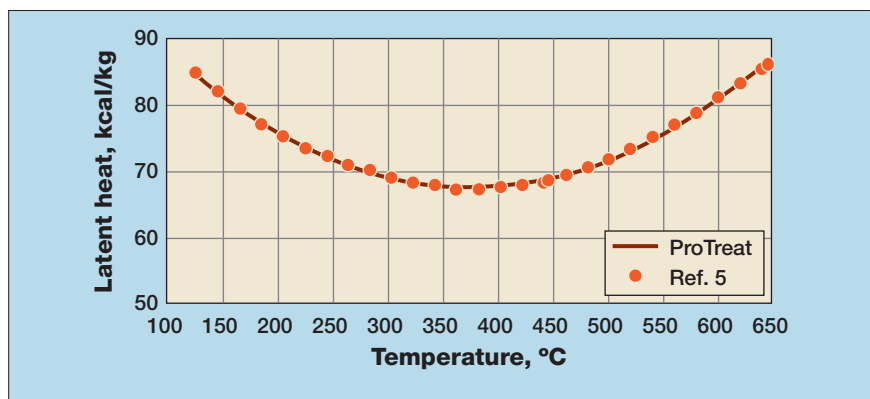


Figure 4 Latent heat of vaporisation of sulphur

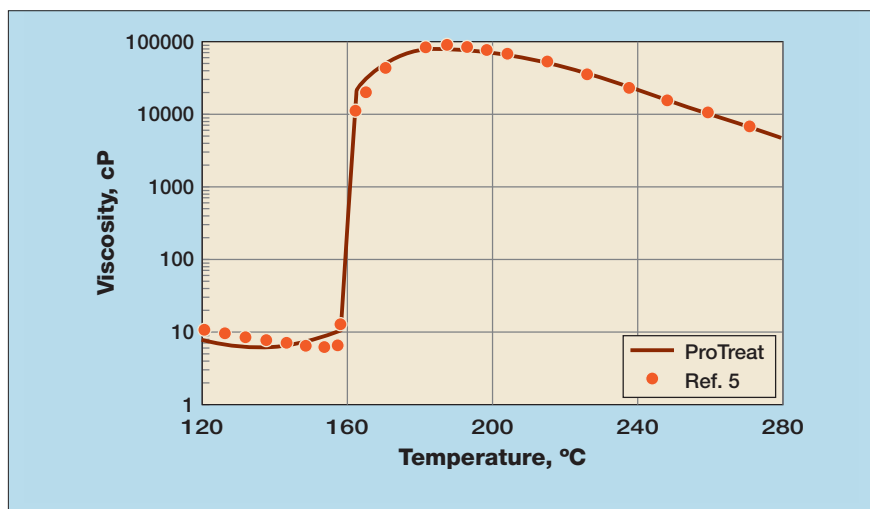


Figure 5 Viscosity of pure liquid sulphur

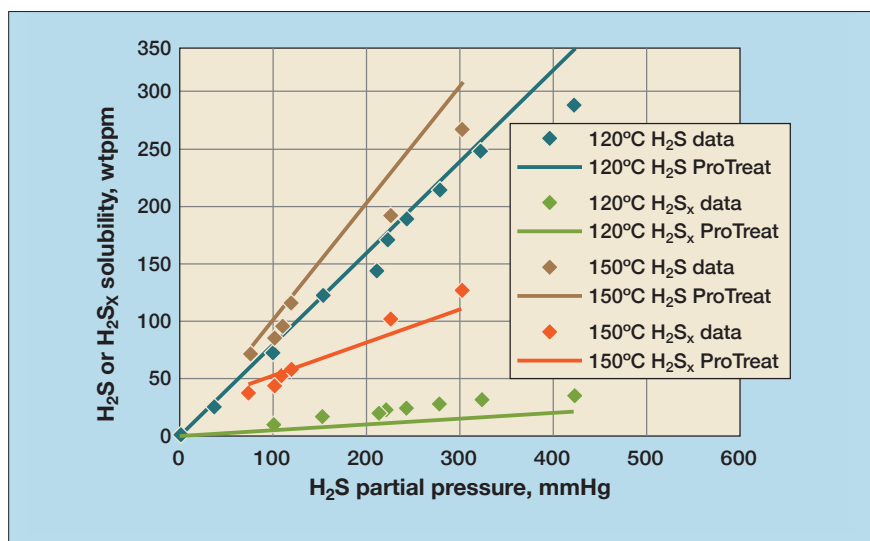


Figure 6 Solubility of H_2S and H_2S_x liquid sulphur.¹¹ Lines are model predictions

H_2S to sulphur. If solubility were purely physical, it would tend to fall with increasing temperature.

The model uses data from several sources.⁸⁻¹² A sample of the solubility data¹¹ is shown in Figure 6, while Figure 7 compares model predictions with viscosity data.^{3,5} In both plots, the lines have been calculated from

the same model. As expected, the higher the H_2S partial pressure, the higher the solubility. The linearity of Figure 6 might be interpreted as adherence to Henry's Law; however, Figure 6 says otherwise. The solubility is not that simple. Increased H_2S partial pressure results in reduced viscosity. One might argue that

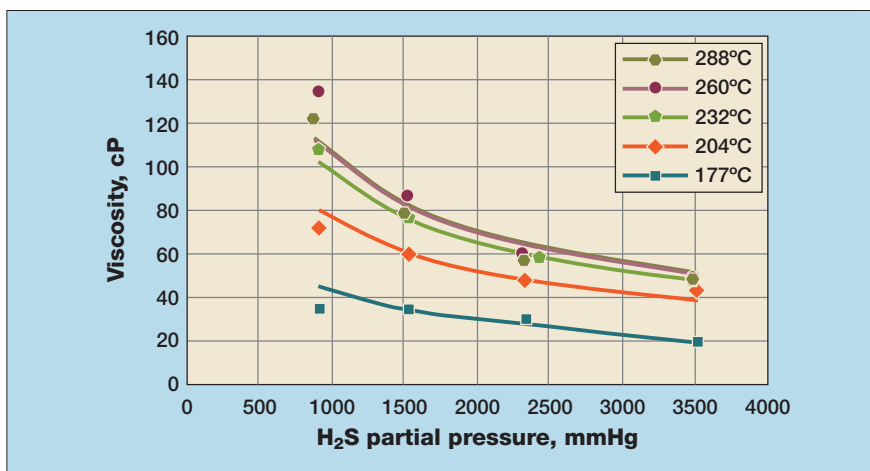


Figure 7 Effect of dissolved H_2S on the viscosity of liquid sulphur. Data^{3,5} vs model heat capacity

higher H_2S concentrations reduce the viscosity by diluting the polymer; however, a more likely explanation is shorter polymer chains that use more H_2S for chain termination. So the solubility is partially explained through a Henry's Law type of relationship, and partially through chemical reaction of H_2S as a sulphur chain terminator. Agreement between model and data is quite satisfactory.

Above 232°C (450°F) liquid sulphur is a sticky gum that cannot be processed, and below 120–121°C (248–250°F) it is a solid. Thus, it is only inside this range that the liquid heat capacity has any practical significance. **Figure 8** shows heat capacity data¹, an artist-drawn line from the *GPSA Data Book*,¹ and the current simulation model.

The reason for the peak at 160°C is the same as for the sudden jump in viscosity at the same temperature, namely the opening of S_8 and

S_8 rings and polymerisation, with subsequent chain breakup and the attachment of H_2S at both ends as chain terminators. These are all reactions that act as additional sinks or sources for any heat added to the system, rather than having the added heat generate sensible heating alone.

Summary

Sulphur is one of nature's most unusual elements. It exhibits what at first glance seems like bizarre behaviour with sudden changes in viscosity and heat capacity, and counter-intuitive solubility of H_2S , all as functions of temperature and H_2S partial pressure. When understood in terms of transitions between multiple forms and sulphur's ability to polymerise, the behaviour of its properties can be understood.

It is important that any simulation tool at least reproduces the

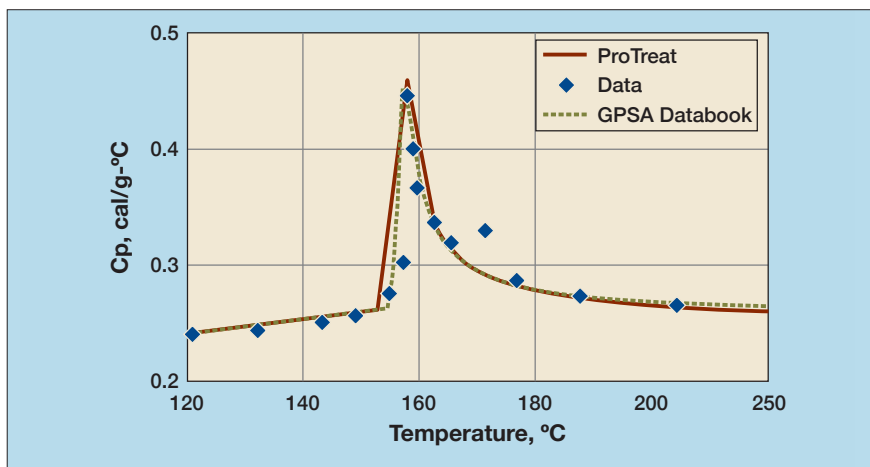


Figure 8 Heat capacity model (—) and data¹ for liquid sulphur

known properties of the components contained in the system. The ProTreat simulator calculates properties that correspond to measured data and uses chemistry in model development.

ProTreat is a mark of Optimized Gas Treating, Inc.

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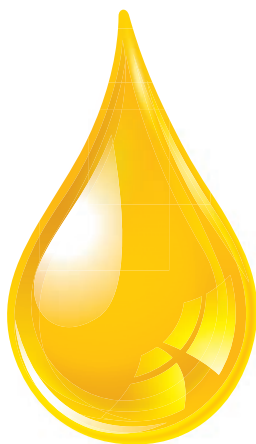
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Anand Govindarajan joined Optimized Gas Treating, Inc as Development Engineer in 2014. He holds a B.Tech degree from Anna University (India), MS and PhD from Oklahoma State University, all in chemical engineering.

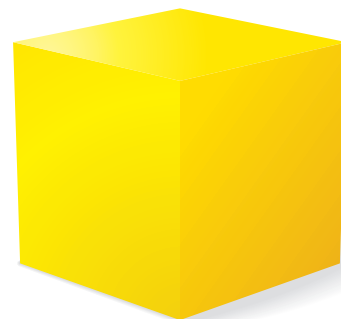
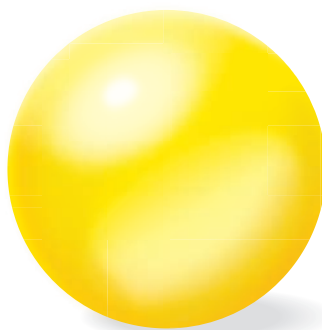
Nathan (Nate) A Hatcher is Vice President, technology development for OGT. He holds a BS in chemical engineering from the University of Kansas. Email: Nathan.Hatcher@ogtr.com

Clayton E Jones joined Optimized Gas Treating, Inc as a Software Development Engineer in 2012. He holds a BS in chemical engineering from McNeese State University and a MS in chemical engineering from the University of New Mexico.

G Simon Weiland recently joined Optimized Gas Treating as a Software Development Engineer. He holds a BS in chemical engineering from the University of Oklahoma.



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Development of a new hydrogen production technology based on steam methane reforming

Liquid operates a large fleet of hydrogen production plants worldwide providing H_2 , CO and syngas to users in the refining and chemical industry. The Global Engineering and Construction offers proprietary hydrogen production technologies to the market. Steam methane reforming is the technology most commonly applied to produce hydrogen from natural gas and light hydrocarbons. The technology typically produces high pressure steam as a by-product. This article describes the challenges for the development of a hydrogen production technology based on SMR. SMR-X technology is dedicated to locations where steam is of low economic value. The technology as well as the development effort are described, and the results are shared, and the technology is explained.

Steam reforming is extensively applied in industry to convert natural gas and hydrocarbon streams into pure hydrogen. **Figure 1** shows a typical configuration for a hydrogen plant. The feedstock is converted catalytically in the presence of steam in tubular reactors at high temperature. The energy for the endothermic reforming reactions is provided by heat transfer from the firebox in which fuel is burned. The hot reformed gas and the flue gas released from the firebox are used for pre-heating various streams such as the feed, the combustion air and the fuel, and also to produce high pressure steam. This steam is utilised for the reforming process itself while the surplus is exported to other, nearby users. The amount of export steam can be adapted by process optimisation to the user's needs over a wide range.

The SMR process can be considered as a heat exchanger network. The design of the heat exchanger network for optimised energy recovery of the hot streams is pivotal for an efficient plant layout. The energetic efficiency of the overall SMR process can be defined by the following equation:

$$\eta = (n_{H_2-out} \times h_{H_2} + n_{Steam} \times \Delta h_{Steam}^{LV}) / n_{NG} \times h_{NG} \quad [1]$$

where the nominator represents the energy flows of the product hydrogen and the export steam respectively and the denominator is the sum of the energy of natural gas feed and fuel streams. It can be shown that the overall theoretical efficiency of the SMR process increases with increasing steam export.¹

A pinch study enables analysis of

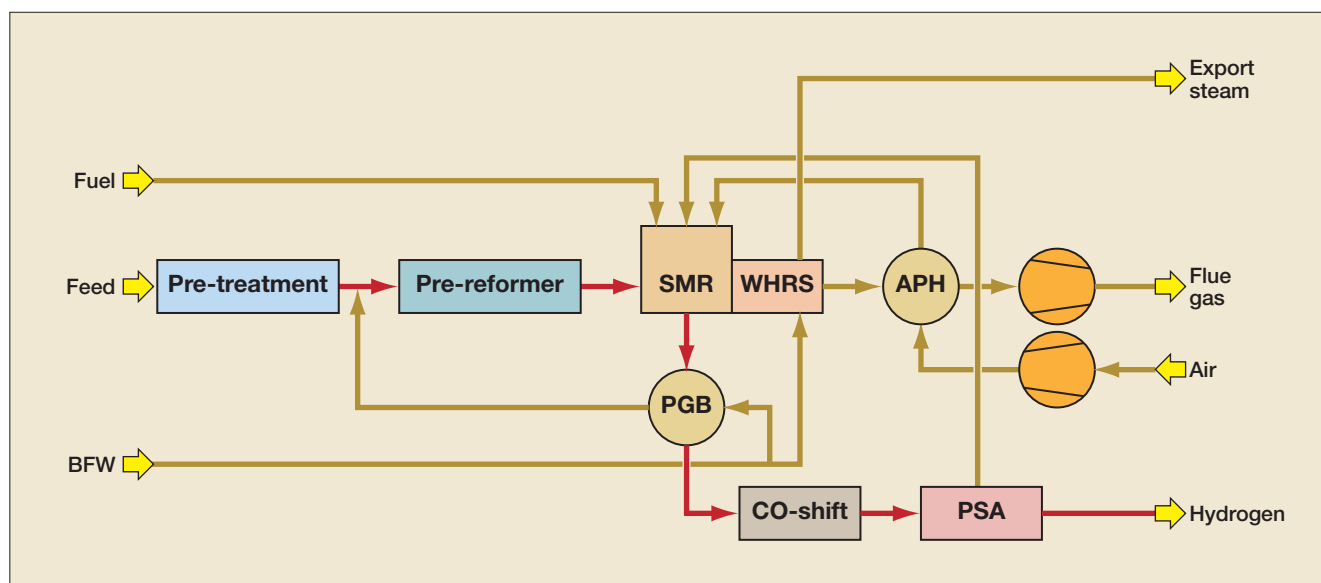


Figure 1 Typical H₂ SMR configuration: WHRS – waste heat recovery section, APH – air preheat, PGB – process gas boiler, PSA – pressure swing adsorption, BFW – boiler feed water

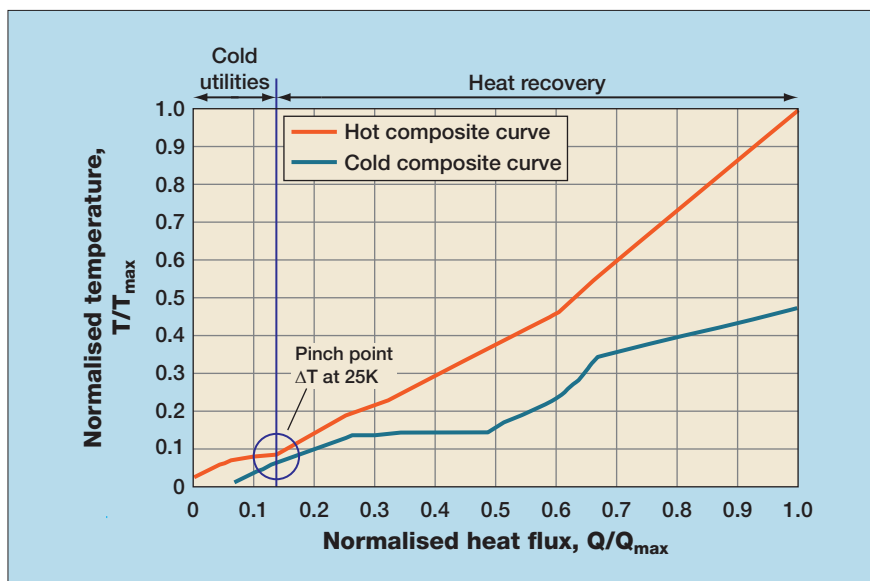


Figure 2 Typical composite curve of SMR heat exchanger network

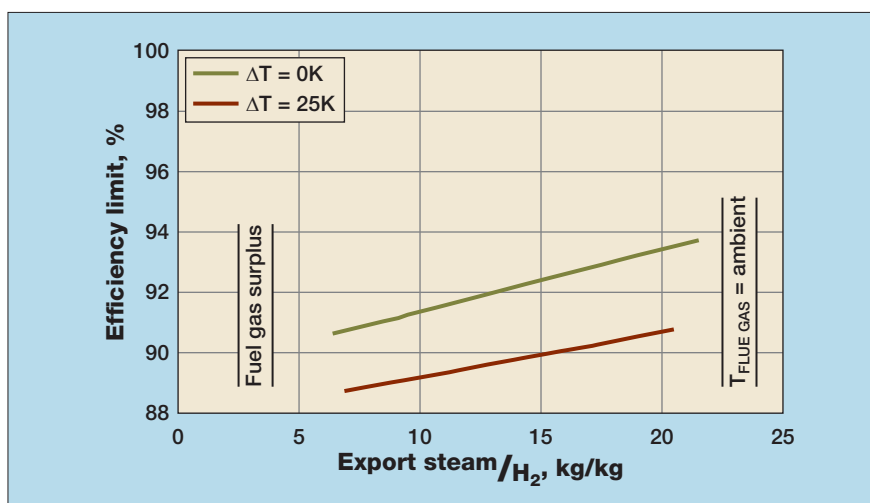


Figure 3 Theoretical efficiency limit of the steam reforming process

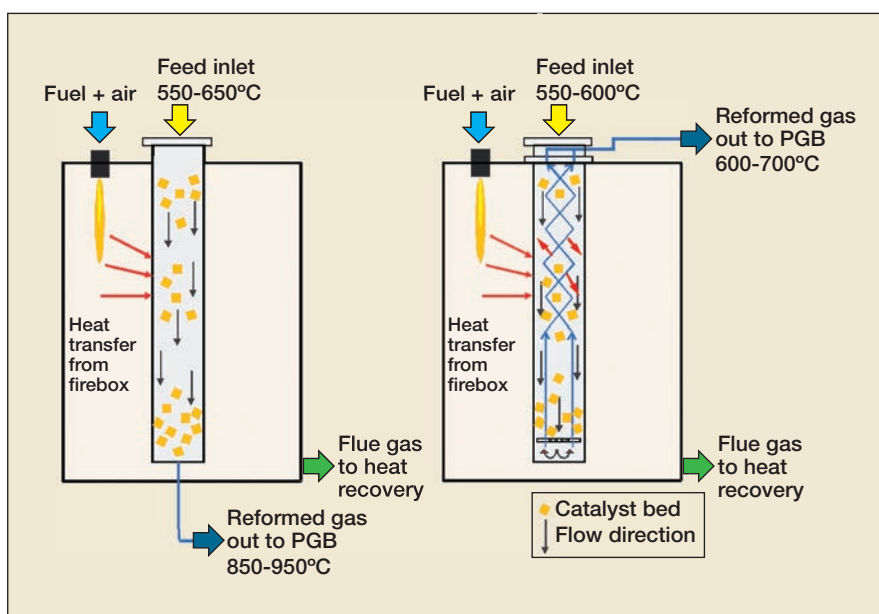


Figure 4 Classical SMR reformer tube arrangement (left) compared to SMR-X technology based on internal heat exchanger reformer tube arrangement (right)

the energy efficiency of a heat exchanger network. The composite curve displays the amount of heat that is transferred at each temperature from the hot streams to the cold streams. **Figure 2** shows a typical composite curve for the SMR process, which exhibits a pinch point typically below 200°C.

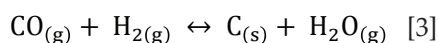
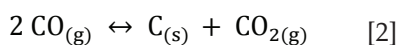
Figure 3 depicts the theoretical efficiency of a standard SMR configuration, assuming an ideal heat exchange with 0K pinch temperature and a commercial case with 25K pinch temperature, respectively. The amount of export steam per volume of hydrogen product has two practical limits. Reducing steam export below 5 kg/kgH₂ results in a surplus of tail gas from the hydrogen purification unit, which is used as fuel gas in the firebox. Export of excess fuel gas is normally neither an efficient nor an economic option. Maximum steam export is reached when the flue gas is released with ambient temperature. Higher steam export can only be reached by additional firing, for instance by duct firing.

Improvement of physical efficiency often has no economic advantage. Energy integration and optimisation of heat exchanger networks reduce the steam demand on process plants, often leading to low economic value of export steam coming from the hydrogen plant. Therefore plant layouts become more attractive when they provide highest efficiency for hydrogen production while minimising export steam. This can even mean reducing export steam to zero or making the hydrogen plant a net importer of steam.

Various options exist to reduce the export steam of a standard SMR plant. For instance, a pre-reformer with reheat can be introduced, operating conditions can be fine tuned, or the high temperature reformed gas can be used to preheat the reformer feed and to produce steam. It is more efficient to utilise the high temperature heat of the reformed gas for endothermic natural gas reforming reactions. This leads to the concept of a heat exchange reformer, which is the basis of SMR-X technology.

In this technology, the hot reformed gas flows in an inner tube arrangement counter-current to the feed flow through the catalyst bed, thereby providing a portion of reaction heat (see **Figure 4**). Approximately 20% of the energy required for the SMR reactions can be provided by this internal heat exchange. The lower temperature of the reformed gas leaving the reactor leads to significantly lower steam production in the process gas boiler. In addition, less energy has to be transferred from the firebox to the reformer tubes, resulting in significantly lower flue gas flow and consequently lower steam production in the flue gas boiler. Zero export steam SMR plants can be easily designed with reduced steam production in both boilers.

The challenge of such technology is to specify the material grade of the inner heat exchanger tubes since they are operating in process conditions prone to metal dusting (MD) corrosion risk. MD is a complex and catastrophic corrosion phenomenon affecting alloys exposed to reducing and highly carburising gases in the temperature range 400-800°C. This results in alloy disintegration to a fine dust of metal and carbon particles.² MD corrosion is thermodynamically possible below the carbon formation temperature; this can be expressed by the chemical equilibrium temperature of the Boudouard reaction (see Equation 2) and the carbon monoxide reduction reaction (see Equation 3), respectively:



Since both of these reactions are exothermic, MD corrosion potential increases during gas cooling while material degradation occurs in a narrow temperature range with a maximum rate between 600°C and 700°C, depending on gas composition, process parameters and the alloy (see **Figure 5**). The lower temperature limit is kinetically controlled mainly by the activation energies of the reactions. The higher temperature limit is controlled either

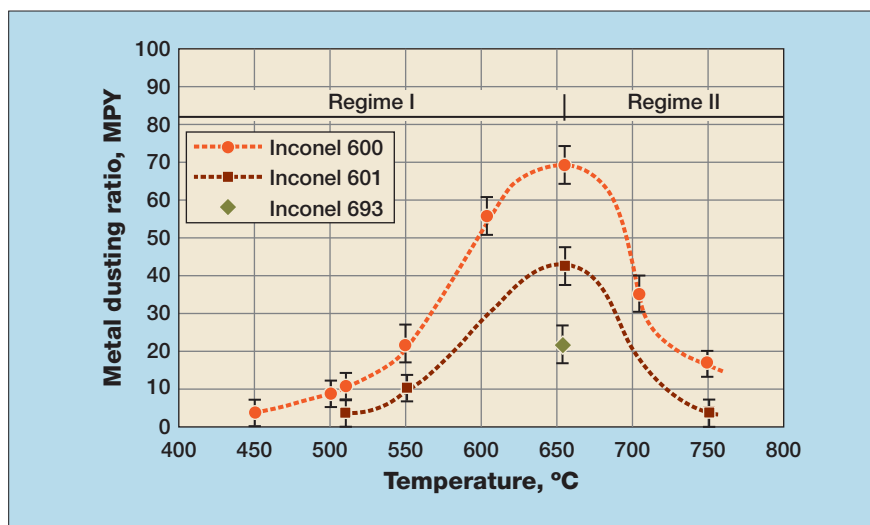


Figure 5 Corrosion rate vs temperature for three Ni based alloys⁴

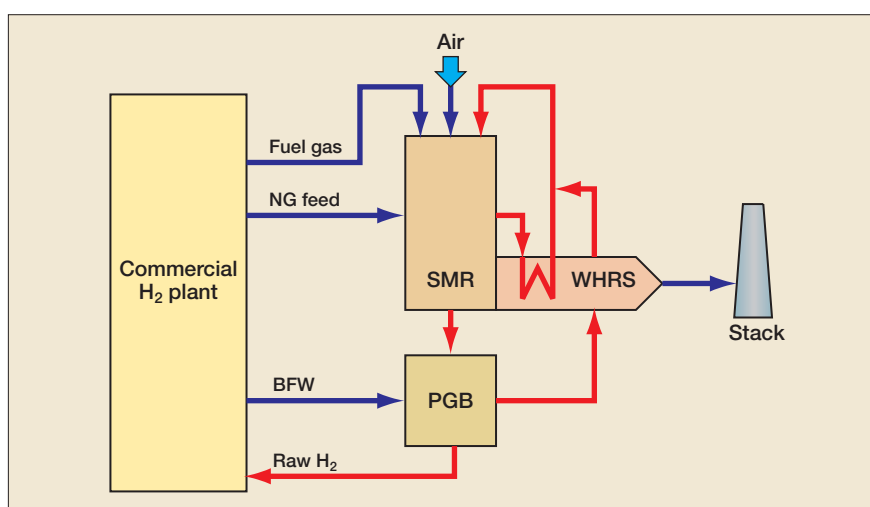


Figure 6 Block flow diagram of the multipurpose SMR demonstration plant; PGB – process gas boiler, WHRS – waste heat recovery section

by the thermodynamic potential or, as proposed in [3] and [4], by the formation of a protective chromium oxide layer, which is favoured at high temperature.

Therefore the design of a reliable heat exchange reformer requires the selection of an appropriate material for the inner tubes. In addition, an accurate model is required to predict the gas and metal temperatures in order to assess MD corrosion potential for the whole range of the plant's operating conditions. A typical engineering approach of adding design margins bears the risk that the equipment has to be operated in the range of the highest level of MD corrosion attack, leading to potentially short equipment lifetime and premature failures.

To address the technical chal-

lenges involved, Air Liquide decided to execute long term demonstration tests in its multipurpose, commercial scale SMR pilot plant. The results validate SMR-X technology's performance in the long term and provide reliable data sets used for model validation.

Demonstration plant

The multipurpose SMR demonstration plant comprises the main process steps of a standard SMR and is connected to a commercial plant (see **Figure 6**). The pilot plant consists of a radiant firebox featured with a variable number of commercial size reformer tubes (see **Figure 7**). The reformed gas is cooled in a process gas boiler before it is routed to the commercial plant. The flue gas released from the firebox is cooled in a

waste heat recovery section and pre-heats the desulphurised feed gas. The pilot plant is connected to the steam system and the utility network of the commercial plant and is operated by the commercial plant's operators. Numerous sensors and sampling points enable calculation of the heat and mass balances and determination of heat transfer for all sections of the plant.

Test programme

The reformer tubes are equipped with helical heat exchanger coils made of different nickel base alloys. One of the base materials was selected from the group of lower MD corrosion resistant alloys, while the second belongs to the group of superior corrosion resistant alloys. In addition, some of the inner tube arrangements were equipped with a protective diffusion coating.

The test programme comprised 8000 hours of operation at commercial reforming conditions. At regular time intervals, the inner tubes were inspected in order to identify the various MD corrosion steps, from incubation to generalised corrosion. After the 8000-hour operation test, the inner tubes were removed from the firebox and fully characterised by metallographic techniques. The status of the various materials and protective coatings were determined along the entire tube length.

A test matrix was executed covering a wide range of operating conditions (see **Table 1**).

The instrumentation of the plant enables recording of the temperature profiles of the reformer tubes. In addition to the data from the installed instrumentation, pyrometer as well as IR camera measurements were performed for each operating point. The results were treated with in-house correction tools to convert



Figure 7 Air Liquide's commercial scale demonstration plant

the raw temperature measurements into accurate tube wall temperature profiles, considering reflections and emissivity of various materials in the firebox. The plant data were processed with statistical data reconciliation, taking into account the uncertainty of sensors and analysis. Very accurate data sets were generated for each operating point. They served as the basis for improvement and validation of the process and equipment design tools as well as sophisticated CFD simulation.

Results

Pilot plant operation was stable for the SMR-X configuration and similar to the operation of the normal reformer tube arrangement. The process parameter test matrix was easily executed over the described broad range of process parameters, demonstrating the flexibility of SMR-X technology and the demonstration plant. Full data sets were collected at stable operation together with the analysis of the relevant streams. The pilot plant's instrumentation delivers redundant information, which was used to feed

the data reconciliation model. This model considers the uncertainty of the measured data and analysis, including all redundant data, to estimate the most probable combination of true plant data. Data reconciliation is a powerful tool to identify sensors with false readings and drift, as well as deterioration of the plant's performance.

The reformer tube simulation comprises models for external heat transfer from the firebox via the reformer tube to the catalyst bed and the internal heat exchanger tubes, respectively, combined with detailed reaction kinetics and flow calculation. Detailed CFD simulation was applied to represent the reactive flow in the packed catalyst bed as well as the firebox including the burners. The parameters of the heat exchanger models were calibrated to the measured packed bed heat transfer prevailing in the reformer tube with the data from the pilot plant.

The simulation model represents the plant data with high accuracy. **Figure 8** shows the calculated temperature profile in the reformer tube for a typical case. The model was used to predict the potential for MD corrosion along the inner tube arrangement. Carbon activity could be calculated with the known gas composition and the temperature profile and is also shown in **Figure 8**. Only those tube sections exposed to carbon activity greater than one are thermodynamically prone to MD corrosion. Comparison of the predicted region potentially affected by MD corrosion with the findings of the heat exchanger material characterisation after 8000 hours of operation confirms the thermodynamic calculations.

The heat exchanger tubes were optically inspected by boroscopy on a regular basis. In addition, a sample of the internal tubes was taken each time and characterised by optical microscopy. The most information about the process of corrosion can be identified from the material with the known lowest resistance to MD corrosion attack. After 1500 hours of operation, the MD corrosion incubation period had already passed and the first

Test matrix with range of operating conditions

Process parameter	Normal	Range
Feed temperature	T-Feed, design	+/- 30 K
Reformed gas temperature	T-Reformed, design	+/- 50 K
Load	100%	50 – 110%
Steam/carbon	S/C – design	+/- 1

Table 1

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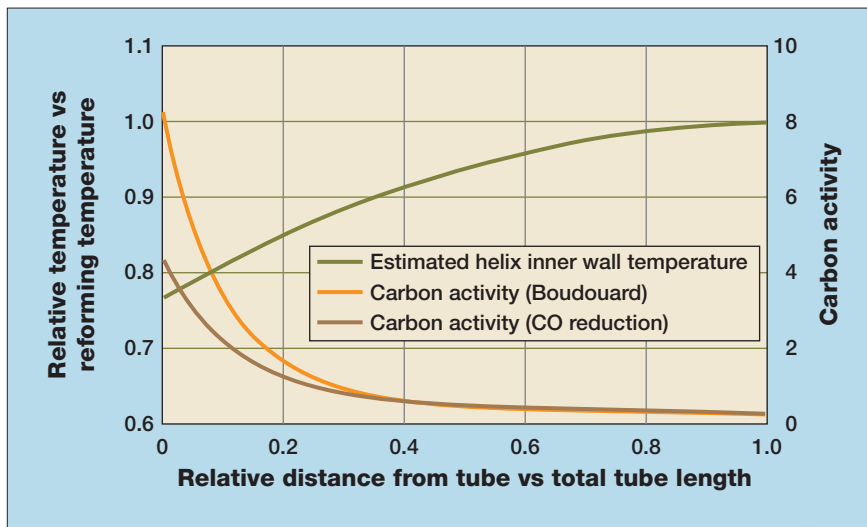


Figure 8 Estimated inner wall temperature and carbon activity as a function of elevation

corrosion pits were identified. Tube sample characterisation revealed carbide formation and confirmed MD attack on this material. Subsequent inspections after 3000 hours and 5000 hours confirmed ongoing corrosion attack, resulting in fast germination and growth of corrosion pits, which led to uniform corrosion and loss of wall thickness.

Full physico-chemical analysis of the helical tubes after 8000 hours of operation, in combination with intermediate analyses of sampled sections, led to some conclusions on the integrity of both the base material and the diffusion coating under conditions promoting metal dusting.

For the less resistant alloy, the number of pits and their depth become smaller with increasing temperature, in line with the theoretical prediction shown in Figure 8. No indication of corrosion such as carbide formation or pits could be detected close to and above the elevation where the model predicted a carbon activity equal to 1 and below. Figure 9 shows surface morphology and cross-sections after etching of the low resistant alloy from various heights of the tube length affected by MD corrosion.

The alloy more resistant to MD corrosion showed some MD attack – carbide formation and some small

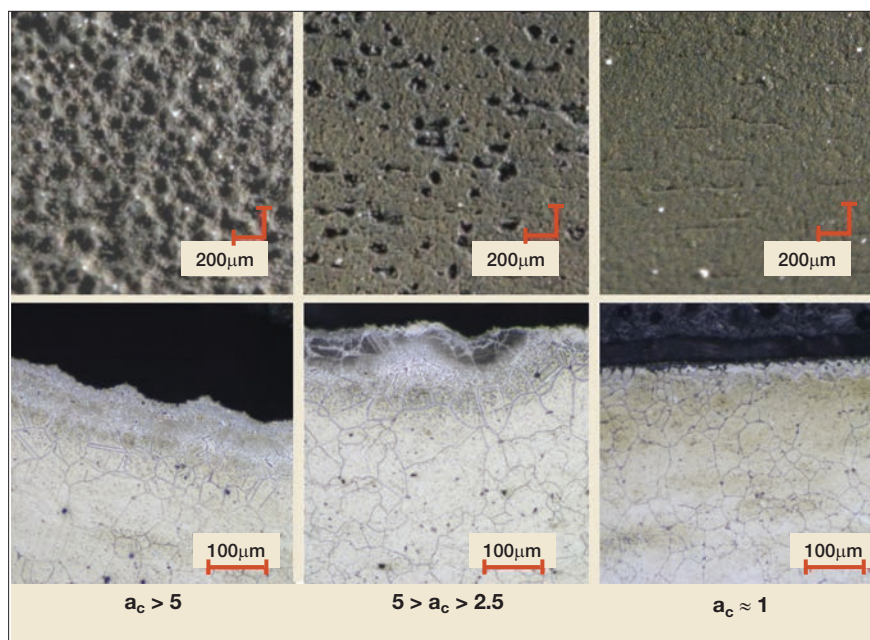


Figure 9 Overview of internal surface morphology (top pictures) and cross-sections after etching (bottom pictures) of the helical tube made in less resisting alloy after 8000 hours of operation as a function of carbon activity

corrosion pits – but to a lesser extent compared to the lower grade alloy. The much higher resistance of this material to MD corrosion attack was proven; however, the material would also suffer from MD corrosion over a longer exposure period.

Two sets of heat exchanger tubes with different base materials were coated with a protective diffusion bond layer. The coating procedure resulted in a homogeneous, defect free layer with a constant thickness. This protective layer was stable over 8000 hours' operation and no carbon diffusion was found in the base material. The protective coating provides stable protection against MD corrosion attack and is the best choice for long life of the internal heat exchanger tubes operating in conditions promoting MD corrosion.

The catalyst particles are too big to fill the gap between the helical heat exchanger coils and the reformer tube completely. This void on the inner side of the reformer tube wall reduces heat transfer locally since no endothermic reactions absorb the energy provided from the firebox at these points. The resulting temperature variance of the reformer tube skin was simulated with an established reformer tube model integrated into the existing full scale 3D SMR CFD simulation. The simulation predicted a maximum tube skin temperature difference of 10K at high heat flux. This small temperature fluctuation has no impact on the tube's life and cannot be seen with the naked eye. Tube skin temperatures were measured with an IR camera (see Figure 10). The measured temperature fluctuation matched very well the predicted fluctuation, which is another indication of the accuracy of the established simulation model.

Process advantages of SMR-X technology

The internal heat exchange reformer tubes are beneficial for a zero export steam plant layout. The high temperature heat from the reformed gas is utilised directly for endothermic reactions. Consequently, less heat is available for steam production. In addition, a lower amount of

fuel gas is required, resulting in lower flue gas flow and smaller convection section equipment such as heat exchangers and fans. These savings over-compensate the slightly more complex design of the reformer tubes.

The investment costs are compared for various zero export steam concepts in **Figure 11** for large H_2 plants. A steam turbine driven electrical generator can be added to a standard SMR to consume the surplus steam and convert it to power. The steam turbine generator producing power adds 21% to the investment of the SMR unit and its economics depend strongly on the spread between power and natural gas prices. The SMR process parameters, reformed gas temperature, and steam to carbon ratio can be tuned to consume all of the steam produced in the reformed gas and flue gas boilers. This results in very similar investment costs but suffers from low efficiency reflected in higher hydrogen production costs. The SMR-X concept has a clear advantage for a zero export steam plant in terms of investment and operating costs, resulting in lower H_2 production costs.

Conclusion

A new SMR technology is presented, which applies reformer tubes with internal heat exchange. Internal heat exchange contributes up to 20% to the energy required for the endothermic steam methane reforming reactions. This advantage is used to propose a SMR technology with zero export steam. Long term demonstration tests at full commercial scale demonstrate the lifetime of the material concept for the internal heat exchanger tubes, which are operated in a MD corrosion prone atmosphere. The plant was operated over a wide range of process conditions, showing the flexibility of the technology. Accurate data sets were established, providing the basis for improvement and validation of the simulation model.

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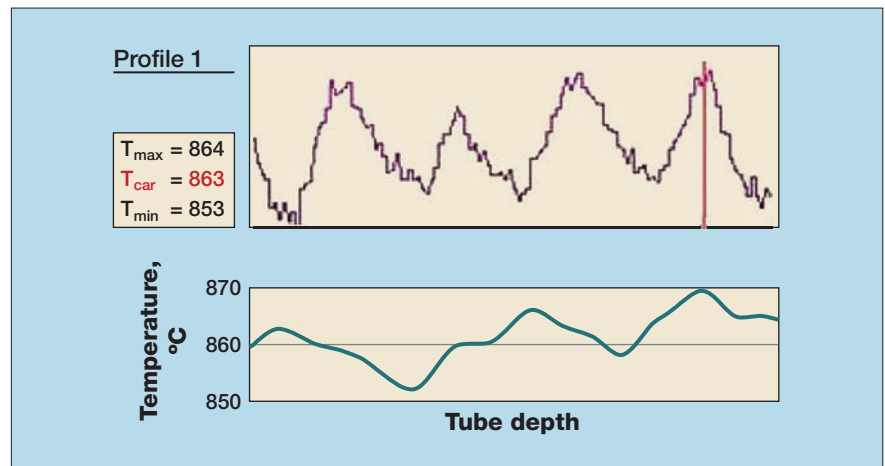


Figure 10 Comparison of predicted and measured variation in tube skin temperature

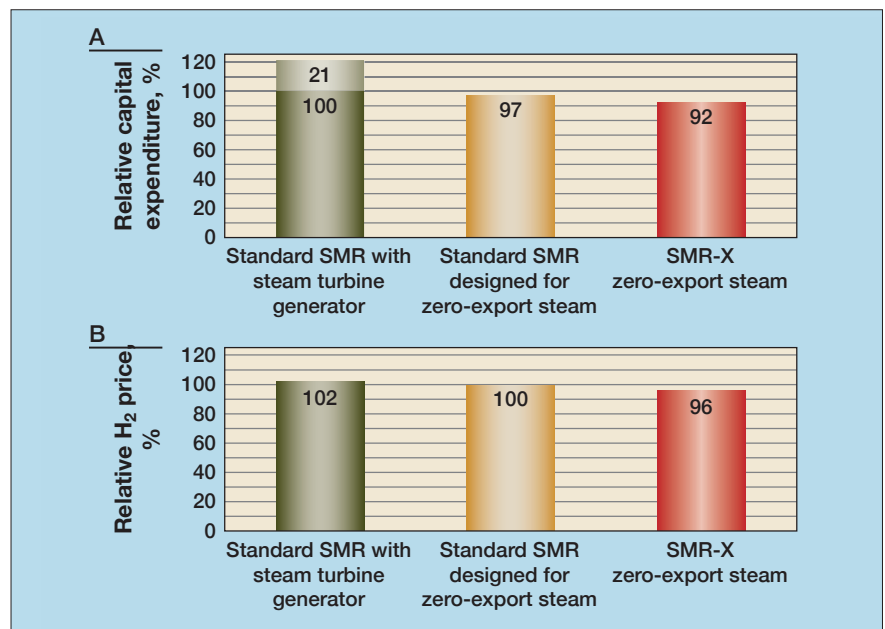


Figure 11 Comparison of investment and H_2 production costs of various SMR concepts for zero export steam (top) and related hydrogen production costs (bottom)

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Holger Schlichting is Program Director Syngas Production with Air Liquide Research & Development. He has more than 20 years' experience in process engineering and holds a PhD degree in chemical engineering from Technical University Berlin, Germany.

Dieter Ulber is Director Product Development in the HyCO Product Line with Air Liquide Global E&C Solutions. He has more than 15 years' experience in syngas production and holds a PhD in mechanical engineering from the University of Technology RWTH Aachen,

Germany and graduated from RWTH Aachen and Thayer School at Dartmouth College, USA.

Sébastien Cadalen is a Research Engineer with Air Liquide Research & Development. He has over 10 years' experience in fluid mechanics modelling and simulation, and holds a PhD in hydrology and hydrochemistry from Toulouse National Polytechnic Institute, France.

Sébastien Doublet is a Research Engineer in Materials Selection with Air Liquide Research & Development. He has more than 10 years' experience in materials engineering and corrosion management, and holds a PhD in materials science from the University of Technology RWTH Aachen, Germany.

Laurent Prost is Senior Project Manager and Manager of the Metallic Materials Pole at Air Liquide Research & Development. He has 20 years' experience in engineering and research, and holds a Master's degree in applied mathematics and a Master of Science in computer engineering from the National Polytechnic Institute of Grenoble, France.



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Dealing with desalter cleaning challenges

How to overcome lengthy and potentially dangerous challenges when cleaning a crude unit desalter

JUSTIN WEATHERFORD and MARCELLO FERRARA
ITW

Let us play a hypothetical yet very real scenario. Like any well operated refinery, you have made many strides in the previous decade to lower operating costs. You are very aware that 85-90% of a refinery's costs to do business is in the purchase of crude oil – feed to the refinery from which all your valuable products are derived.

The vast majority of your efforts to reduce expenses were realised by a capital project eight years ago that allowed your refinery to begin processing more 'advantaged' crudes. Advantaged crudes represent a lower cost for your refinery, which is worth millions of dollars in additional profits and value to your shareholders each year, but compared to your 'normal' crude feed slate they are typically lower in API gravity. They also tend to contain high levels of solids, unstable asphaltenes and difficult to remove chloride salts.

Just four years ago, during the previous unit maintenance turnaround, your refinery performed mandatory inspection of the desalter vessels. In the pre-turnaround planning phase, a great deal of time and energy was spent developing the plan for how to safely clean and reach safe conditions for man entry of the vessels.

As part of your due diligence, you had met with several chemical cleaning companies that promised they could help you quickly reach safe-entry conditions of the equipment throughout the unit. Time was the main driver for you. Of course you would be willing to pay a little extra for that one supplier

that could deliver the unit 'clean' and ready to enter most quickly.

One supplier clearly stood above the rest as they were very professional in their initial presentation and they provided a fairly substantial finely bound binder with example after example of jobs they had performed with great results and happy customers.

Going into the turnaround, you had five days planned for degassing/decontamination of the equipment alone and they promised this could be achieved in only 18 hours with their chemistry. At about \$2 million per day of lost profit opportunity when the unit is out of production, you were happy to pay some more for this company over the low cost suppliers because you had to go with someone who you had confidence in (and the extra represented only a fraction of the true cost of doing the job).



Figure 1 A typical stable emulsion created by using emulsifiers

The plan called for a chemical injection to be performed during the steam-out of the equipment. The chemical would be injected in several locations throughout the unit and would 'guarantee' quick removal of all the hydrocarbon and other contaminants present so that your maintenance activities and inspection could begin a full four days ahead of schedule.

In addition to its formulation for handling H_2S and a few other contaminants, the chemical essentially works by forming an emulsion with the hydrocarbon contaminants present in the unit (see **Figure 1**). As the steam condenses, the chemical and the hydrocarbons follow the path of the condensate. The first injection point for the degassing chemical was into the steam upstream of the first preheat exchanger. Because the chemical base is a heavy hydrocarbon based emulsifier, it effectively binds the condensate to the hydrocarbon in the unit and follows the path of the condensate, which is blown down from low points throughout the unit and collected for disposal.

In order to get the chemical dispersed throughout the unit, a high steam flow rate was specified by the chemical cleaning provider, which to meet that requirement was a large job in itself. Several additional steam out and drain lines were installed in the weeks leading up to the turnaround.

One of the first surprises was that the unit did not seem to heat up as fast as you had expected based on previous experiences. Of course the unit was to be hot (approximately

steam temperature) prior to initiating the chemical injection.

You were certainly aware that the heat exchangers in the preheat train would probably be more fouled than normal due to the additional advantaged crude the unit had been running as this was evident from the lower furnace inlet temperature and hydraulic limitations that operations had been managing for the six months or so prior to the turnaround.

What you could not have predicted was that, even with more steam rate specified than normal, the unit would take over 14 hours just to heat up to steam temperature. The reason for this was clearly evident as the mechanical cleaning portion of the turnaround began a few days later and the hot preheat train exchanger heads were removed. They were packed with sludge and over half of the tubes were plugged on several exchangers, especially those just upstream of the fired heater. At least this explains the long heat-up time, but unfortunately the job was already 12 hours behind schedule and you had barely begun.

Following each condensate collection spot, an additional chemical injection was utilised. Altogether over 30 chemical injection locations were scattered throughout the unit.

Collecting the condensate, chemical and contaminants required approximately 12 full size frac tanks. This material could not be sent directly to oily water sewage because the nature of the chemical and the bound hydrocarbons from the unit (tied up in the emulsion) would overwhelm the biological process at the wastewater treatment plant (WWTP).

For this reason, a separate demulsification step was required. This step included adding an additional chemical to the frac tanks, which then had to settle for several days before the first amount of liquid could begin to be dribbled off into the water treatment system. Due to the high load placed on the WWTP, and its already fragile state following other turnaround activities, the frac tanks remained on-site for the

next six months until they were finally emptied.

Other than the initial 12-hour delay, most of the unit was essentially degassed and ready for mechanical cleaning and other maintenance following 18 hours of chemical injection as promised. Most of the unit...

Due to an unknown but likely large volume of sludge in the desalter vessels, they were isolated and decontaminated independently of the remainder of the unit. Because the desalters required mandatory inspection, as previously discussed, the operators and chemical cleaning supplier made them a priority. Prior to initiating any steam-out of the unit, the vessels were floated with water to displace liquid hydrocarbon still

The minimum requirement for man entry into a vessel containing flammable vapours has been accepted as 10 ppm of lower explosive limit

present following unit shutdown.

Once the vessels were liquid full, they were drained of most of the free liquid water and steam-out was initiated. This step was started concurrently with steam-out of the rest of the unit.

Due to severe steam limitations during the turnaround (other steam producing units were already out of production), steam was in short supply; however, refinery management had already declared the crude unit to be the critical path to the refinery getting back online. The other downstream units also included in the turnaround had to wait until steam was available to begin decontamination efforts.

While it was indeed known that sludge was accumulating in the desalters more rapidly than in previous runs due to the additional advantaged crude in the crude

slate, perhaps no one could have predicted the issues that would be encountered trying to clean the desalter vessels.

After 24 hours of steam-out operations, of which 18 hours included chemical injection, there was still no sign of any progress.

In order to certify that a vessel is sufficiently degassed/decontaminated, operators take regular four-gas measurements of the steam from a vent on the top of the desalter.

The desalter vessels are confined spaces and, by definition, they have limited openings for entry and exit, unfavourable natural ventilation which could contain or produce dangerous air contaminants, and are not intended for continuous employee occupancy. The hazards associated with confined spaces can cause serious injury and death to workers. In fact, one of the leading factors in fatal injuries is failure to recognise or control the hazards associated with confined spaces.

To declare a vessel safe for man entry, the vessel must be free of the vast majority of liquid hydrocarbon and the flammable atmosphere inside the vessel must be effectively eliminated.

Three components are necessary for an atmosphere to become flammable: fuel and oxygen (in the proper proportions) and a source of ignition. Below a given proportion of fuel to oxygen, the mixture is below the lower explosive limit (LEL).

LEL defines the lowest concentration of a gas or a vapour in air capable of producing a flash of fire in the presence of an ignition source. At a concentration in air lower than the LEL, gas mixtures are 'too lean' to burn. The minimum requirement for man entry into a vessel containing flammable vapours is expected to be 10 ppm of LEL (or, stated another way, the concentration of hydrocarbon in the air within the vessel is at or below 0.1%). This was indeed the target of your degassing/decontamination effort. Of course, you would have preferred having 0% LEL (no hydrocarbon in the air within the vessel), but in your experience this

would have taken many days of steam-out, which was not affordable as additional downtime.

In addition to eliminating the flammable atmosphere, the vessel must be cleared of all toxic gases. Most notable in this case is the H_2S present in the sludge. H_2S is a highly flammable, explosive gas that can cause a wide range of health effects, typically by breathing it. Even at very low concentrations, the effects can be severe. The odour threshold for H_2S is first noticeable to some people as a 'rotten egg' type odour at less than 1 ppm, and becomes more offensive at 3-5 ppm. At this concentration level, prolonged exposure to H_2S can cause nausea, tearing of the eyes, headaches and loss of sleep.

At approximately 100 ppm H_2S , a person may experience severe coughing, eye irritation, loss of smell (which means the hazard may still exist even though you can no longer personally detect it), altered breathing and drowsiness after about 15 minutes of exposure with a gradual increase of symptoms with time. 100 ppm H_2S is generally considered and defined by the US National Institute for Occupational Safety and Health (NIOSH) to be immediately dangerous to life and health (IDLH).

At 500-700 ppm H_2S , a person will begin to stagger and ultimately collapse within about five minutes. Serious eye damage will occur in approximately 30 minutes and death will occur in 30-60 minutes. At 700-1000 ppm H_2S , a person will experience rapid unconsciousness and immediate collapse. Breathing will cease after 1-2 breaths and death will occur in minutes. Above 1000 ppm H_2S , death occurs nearly instantly.

To be sure, working around H_2S is serious business for serious, well trained, competent workers. Your refinery had a significant near miss in the past, when two pipe-fitters were exposed. Luckily they were okay but everyone knows it could have been much worse. Since then, your refinery has initiated a procedure as part of your efforts to eliminate all injuries. The procedure

is based on one that several European refiners have implemented (as part of a new regulation), which calls for a minimum seniority level for any contracting crew entering a confined space with a potential hazardous environment. In your case, 80% of the crew must have a minimum of five years of on-the-job experience. It has been very difficult to even find contractors that qualify.

Achieving safe levels of H_2S in the desalter vessel has always been a challenge in previous turnarounds, but it was indeed high on everyone's mind this particular turnaround. The degassing efforts could not stop until the goal was met.

Refineries normally require a limit of zero ppm for H_2S for work-

Refineries normally require a limit of zero ppm for H_2S for workers to enter a confined space safely

ers to enter a confined space safely, thus the degassing/decontamination target for this job targeted at least achieving H_2S levels below 10 ppm.

After 48 hours into the chemical cleaning, there was still very little sign of progress, if any; 72 hours.... 96 hours.... Nobody expected this.

Finally, after a full five days of steaming with chemical, the H_2S level was beginning to decrease and, by the middle of day six, the target of 10% LEL and 10 ppm H_2S in the steam was finally met. After another 12 hours or so for cool-down of the vessel, the manways were opened and air ventilation was set up to further assist cooling and maintain sufficient oxygen levels inside the confined space for subsequent manual sludge removal efforts.

Feeling that all the hazards of working inside the confined space had been mitigated, the trained

labour crews entered the vessels, buckets and shovels in hand, to begin the laborious task of manual sludge removal.

It was discovered that nearly 4 ft of compact sludge was present in the first-stage desalter and another 2.5 ft of sludge in the second-stage desalter vessel. Working around the clock, with three workers per shift inside each vessel, the sludge removal effort would take at least 4-5 days if everything went smoothly. Everything did not go smoothly.

The compacted sludge inside the desalter vessels was a mixture of inorganic material held together by a vast matrix of organic sludge. Indeed, the greater part of the volume of the sludge was the organic matter, some hard, coke-like material, but for the most part a fairly soft oily sludge.

Even after the extensive steaming and chemical cleaning efforts of the previous days, it quickly became evident that working inside the vessels without a much more elaborate HSE plan in place was not going to be possible. Even upon first agitation of the sludge, the personal H_2S monitors of two workers inside the first vessel alarmed. All workers were rapidly evacuated from both vessels and operations personnel discovered that even slightly disturbing the sludge in the desalter released significant quantities of trapped vapours, which could cause the atmosphere to become toxic again or even re-enter the flammable zone.

A management review of the incident the following day determined that, from this point on, all work inside the desalter vessels must be done utilising a nitrogen blanket and workers fully equipped with personal protective equipment (PPE) capable of keeping them safe in an oxygen-deficient atmosphere, meaning the already difficult task now had to be accomplished with full self-contained breathing apparatus in place.

The unit was hydraulically constrained already due to fouling in the preheat exchangers and vacuum tower packing, but margins were driving the refinery

and therefore management had decided to move the turnaround up three months, placing it directly in the heat of the summer.

Managing the safety of the workers inside the equipment with all the required PPE worn would be challenging enough, but even more so given that dehydration played an additional factor.

Desalter challenges

We feel it is important to play out a hypothetical, yet possible case study so that you can have a good understanding of the operating decisions made by many refiners and how ITW technology can be utilised to eliminate all of these challenges. ITW is aware of many cases in the industry where desalter cleaning is pushed out of the turnaround scope or avoided altogether until operations are simply no longer sustainable due to high sludge level.

Desalters are included in the crude preheat train upstream of the hot section of the preheat train and the fractionation columns.

On a very simplified level, water is mixed with the crude oil just upstream of the desalter. Certain contaminants are removed from the crude and leave with the water. The desalter performance can best be described by the degree to which the desalter removes the contaminants from the oil and the degree of oil and water separation that occurs in the desalter vessels (which is very important to the operation of the unit).

The main purpose of the desalter is to wash out the salts present in the crude in order to prevent plugging and fouling of downstream process equipment by salt deposition and to reduce corrosion caused by the formation of hydrochloric acid (HCl) from the chloride salts present in the raw crude.

In addition to removal of salts, the desalter removes other contaminants such as sand, clay, drilling mud, iron sulphide, iron oxide, and other solids from the crude. Along with a natural build-up of hydrocarbon sludge that deposits from destabilisation of asphaltenes and/or precipitation of paraffins, these

contribute to sludge in the desalter vessels. As the sludge level builds, the working volume of the desalter effectively decreases, leading to an increased quantity of oil in the brine (desalter effluent water) and an increased quantity of water and associated contaminants (mainly chloride salts) in the crude oil.

Many refiners utilise a mud wash system to assist removal of sludge and contaminants from the bottom of the desalter vessels. Mudwashing is fairly effective but has the downside of increasing the discharge of hydrocarbons in the brine and thus to the water treatment plant. Even with systems that utilise the best industry practices, sludge build-up is common and, at some point, desalter cleaning is inevitable.

From an environmental standpoint, the desalter effluent, or brine, is typically the largest load on the WWTP. Oil present in the effluent

Even with systems that utilise the best industry practices, sludge build-up is common and, at some point, desalter cleaning is inevitable

water is called oil carry-under. Changes in crude or other unit changes can result in desalter upsets that inadvertently discharge oil, emulsions and solids to the WWTP. Of course this is exacerbated by large amounts of sludge in the vessel.

Formation of a stable emulsion is a major problem that many refiners deal with on a regular basis. They normally utilise an electric grid in the desalter vessel to assist the separation of the natural oil/water emulsion that forms. As the sludge level increases, the sludge and solids present in the sludge can actually work to stabilise the emulsion, leading to poor oil/water separation. Ultimately, the sludge level can be high enough

that the water layer in the desalter can reach the electric grid, thus causing the grid to short out and lead to an extremely poor desalter performance.

On another note, poor desalter performance has the downside of increasing water carry-over – water concentration in the crude leaving the desalters for processing in downstream equipment.

Each barrel of water takes roughly 6-7 times the amount of energy to heat compared to a barrel of crude oil. Even slightly elevated levels of water carry-over will result in very poor energy efficiency of the unit – an increase in fired duty and an increase on the duty of the atmospheric overhead condensers. As mentioned, water is the solvent for the extraction of salt from the crude; therefore, carry-over of water from the desalter means an associated increase in salts in the crude. These salts carried with the water entrained in the oil dramatically increase the amount of chlorides in the downstream equipment and ultimately lead to accelerated fouling and corrosion.

Additionally, water has an expansion ratio of approximately 1200:1 when vaporised from liquid to steam. This is many times the expansion ratio of hydrocarbon.

If a severe desalter upset occurs, which can be worsened by a reduced desalter working volume, a slug of water can be sent downstream. In some cases, this can cause a pressure surge downstream of the fired heater, at the flash zone of the atmospheric column, and can bend/damage the trays or packing near the bottom of the column. In a worst case scenario, severe damage and even total destruction of the fired heater is possible, depending on the volume of the water slug.

Common industry approach to desalter cleaning

ITW is aware of many cases in the industry where desalter cleaning is pushed out of the turnaround scope or avoided altogether until operations are no longer sustainable due to sludge level. As discussed above, this approach can have serious



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In many cases, cleaning can be achieved outside of the turnaround when a unit has a two-stage desalting configuration by having only one desalter vessel in service at a time while cleaning and maintenance is performed on the other. This is done because the nature of desalter cleaning is very nasty, manpower intensive, and time consuming. Of course, this significantly reduces the overall desalting efficiency and can have severe operating and reliability consequences in itself as the work takes several days, often several weeks. Additionally, operators try to avoid by all means having cleaning personnel inside vessels while the unit is running, due to obvious safety concerns. In particular, desalter cleaning is not welcome because manual cleaning, besides being difficult to perform (for instance, the cleaning personnel cannot stand inside the vessel because of the grids), implies handling of the sludge (shovelling and packaging) while the refinery is in operation.

Eliminating desalter cleaning challenges

ITW has developed, patented and successfully implemented two technologies that solve all the problems associated with cleaning desalters and other process equipment: Online Cleaning and Improved Degassing/Decontamination

ITW Online Cleaning is a patented cleaning technology that includes the following:

- Method and process steps
- Chemicals
- Monitoring system.

This cleaning method utilises an oil based chemical (a patented asphaltene stabiliser) that will dissolve sludge and remove sludge and deposits from metal surfaces. Asphaltenes are dissolved by injecting a chemical during a closed-loop circulation and are chemically re-peptised (the process responsible for the formation of stable dispersion of colloidal particles). In other words, they are chemically transformed from a sludge to a liquid and stabilised.



Figure 2 The original sludge level was above the manway in each desalter vessel. Following ITW's 2-Step Approach, over 98% of the sludge was removed, with only 2-3 inches of sludge remaining in the bottom of the vessel, primarily composed of sand and other inorganic material left behind

ITW's approach is comprehensive and provides many HSE benefits compared to the conventional approach. The first phase is to utilise ITW Online Cleaning technology to remove the vast majority of the sludge and contaminants inside the vessels. The desalter vessels are cleaned separately from the rest of the unit and this job lasts 24 hours or less, oil to oil.

Following the application of ITW Online Cleaning, the desalter vessel(s) can be placed immediately back into service because the sludge volume will be so dramatically reduced (up to 98%). While a small amount of inorganics (sand, soil, and so on) will be left behind in the vessel(s), this quantity is



Figure 3 No emulsion formed after treatment with ITW chemicals. Clear water can be sent for processing

insignificant to the original sludge volume and there is no need for mechanical cleaning. Since the sludge is converted to a fully reusable and reprocessible product, tens of thousands of dollars can be saved by recovering the product compared with hazardous waste disposal.

In the case of a turnaround/inspection application, and in other situations where safe entry and/or hot work is required, the second phase is to utilise ITW Improved Degassing/Decontamination technology.

This process utilises a single multifunctional chemical, which contains, among others, compounds that improve hydrocarbons' affinity to water or steam.

The degassing/decontamination chemical does not create any emulsion with hydrocarbons as we do not use any emulsifier to accomplish the task. Better, the chemical is capable of transforming water into a temporary 'solvent' for hydrocarbons. The other technologies available in the market create a stable hydrocarbon/water emulsion that needs to be addressed by disposal or by having the same on hold for many months and carefully routing it to WWTP.

While quickly removing hydrocarbons from the equipment, the chemical will simultaneously:

- Eliminate H_2S and mercaptan emissions/noxious odours
- Reduce benzene inside the equipment to <1 ppm
- Eliminate pyrophoric solids issues.

As the chemical does not create any emulsions with hydrocarbons, the effluent can be routed directly into the oily sewage. No frac tanks are needed to store the emulsion and no additional demulsification step is needed. In fact, by simply settling, a spontaneous and immediate separation of hydrocarbons from water will occur. The WWTP will not be overloaded as the hydrocarbons will be easily separated in the API separator and the chemical is fully biodegradable.

That said, an additional benefit is achieved when Improved Degassing/Decontamination is

applied following Online Cleaning. The vessel is nearly bare-metal clean to begin with so there is no requirement to degas a huge volume of sludge, and there are no HSE concerns as you would have with the typical approach.

Combination of technologies

By applying ITW Improved Degassing/Decontamination after Online Cleaning, safe entry conditions will be achieved in a shorter time, and steam-out time/steam consumption will be further reduced.

In one recent job in a 100 000 b/d crude unit, the following was achieved:

- The first- and second-stage desalter vessels were cleaned simultaneously with the remainder of the unit (including the entire preheat train, atmospheric and vacuum towers) prior to a turnaround.
- ITW Online Cleaning lasted 16 hours. Some 98% of the original sludge volume was dissolved and removed from the vessels (see **Figure 2**).
- The washing fluids were sent to a crude tank and fully reprocessed (see **Figure 3**).
- The subsequent degassing/ decontamination of the desalter vessels was achieved with 0% LEL and 0 ppm H₂S within six hours of initiating steam-out operations.
- Only 2-3 inches of inorganic material remained in the vessels and no mechanical cleaning was required.

ITW helped this refining company in solving all the conventional problems associated with cleaning their desalter vessels. Because the desalters can be cleaned so quickly and efficiently utilising ITW technology, the following options are available for future consideration:

- The job can be done during the turnaround simultaneously while online cleaning the remainder of the unit.
- The job can be done during the turnaround following the cleaning of the remainder of the unit.
- The job can be pushed outside of the turnaround altogether with minimal operational consequences because the time frame is dramatically less than the conventional approach (24 hours oil-out/oil-in only).

All of the above is a powerful tool to improve current operations and to effectively overcome the current challenges of desalter cleaning.

Justin Weatherford is Sales Manager North America with ITW. With 15 years' experience in the refining/petrochemical business he previously worked for a major oil company and engineering company and holds a BS in chemical engineering from Kansas State University.

Email: jweatherford@itwtechnologies.com

Marcello Ferrara is Chairman of ITW. With 30 years' experience in the petroleum business, he holds international patents for new processes and additive compositions for environmental control and for improving petroleum/petrochemical processes, and holds a PhD in industrial chemistry.

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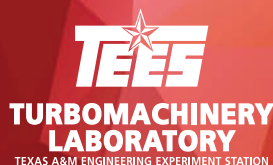
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Automated column start-ups

Automation at start-up, supported by dynamic simulation, is desirable if it can be accomplished with a modest amount of engineering effort

MARTIN SNEESBY
APESS

Most oil and gas processes run continuously for several years between shutdowns. These plants are typically optimised for 'normal' operation and fine-tuned to run up against throughput and/or efficiency limits, often utilising multi-variable controllers and sometimes even dynamic optimisers. The technology is advanced and developed on an ongoing basis. In contrast, start-up procedures can remain virtually unchanged throughout the life of the plant and are rarely considered except in the immediate period around a shutdown or turnaround. At that stage, old procedures and operating instructions that consist of a long series of manually executed steps are dusted off and given to operators who are mostly unfamiliar with the process plant in this transitory state. The prevalence of operating incidents associated with start-ups is, therefore, unsurprising. Kister lists start-up issues as the fourth most common type of column malfunction behind plugging and coking (#1), sump issues (#2) and internals damage (#3).¹ Since that report, the Texas City disaster in 2005 highlighted how easily problems can occur when a column is being operated without its usual automatic controls.

Batch and heavily sequenced processes, as found in some other process engineering sectors, are typically operated quite differently. The focus in these cases is commonly on getting the plant to a particular condition and then keeping it there. Shutdowns and start-ups are often more frequent,

and it is more likely that these process transitions will contain elements of automation and be supported by coded sequences.

There are opportunities for aspects of both styles of operation to be incorporated into continuous improvement plans within the other sector. For example, the potential benefits of applying automation (either fully or partly) to start-ups for continuously operated plants include:

1. **Safer start-up:** better definition of the start-up route and controls to keep the start-up on track; automated detection and handling of possible disturbances; support for operators working in unfamiliar territory.

Zero risk is not possible and every operating plant carries with it some risk, however well managed

2. **Faster start-up:** normal operation achieved quicker, resulting in increased production; less chance of an operating incident that might compromise the start-up.

3. **Increased flexibility:** freeing up of operating resources to concentrate on other activities during a period of intense activity and pressure; however, the potential need for some process control engineering resources to be on hand might partially offset this benefit.

The total value of these benefits will always be difficult to estimate accurately but the potential avoidance of a single incident could easily be sufficient to justify some investment, or at least an exploration of possibilities. The production benefits (for instance, an extra day of production at full rates) are likely to be enough to sweeten the deal but possibly not more. Yokogawa issued a white paper in conjunction with FRI and claimed time savings of 30% as well as improved operational safety and increased margins to safety and design limits.²

Improved safety is always a worthy goal. Engineers are often required to make a professional judgement with regard to the 'acceptable level of risk'. Of course, zero risk is not possible and every operating plant carries with it some risk, however well managed. Plant start-ups bring this equation into focus because it is undoubtedly safer not to start the plant at all, yet that philosophy is self-defeating. The real questions are:

- How do I ensure that the risks are fully understood?
- Do I need additional layers of protection against risk?
- If required, what additional layers of protection are practicably possible?
- Do the necessary process operations meet international standards of risk management after all layers of protection are considered?

Costs always need to be considered whenever benefits are being discussed. The costs of a start-up automation project depend on the scope and the execution model.

Several approaches are possible. ‘Getting a few heads together’ is always going to be a good starting point, but will it be sufficient given a probable lack of collective experience of infrequent events? And does this method provide enough detail for a control engineer to create an automated sequence? One exciting alternative is to use dynamic simulation or an existing operator training simulator (OTS) to explore and optimise the start-up procedure. The technology to support this approach exists and the right skills are available from specialist and niche suppliers to deliver such a project.

A relatively simple example is considered here: a C₄-C₅ splitter with non-equilibrium condenser and some thermal integration via preheating of the feed (see **Figure 1** and **Table 1**). However, the principles are extensible and more complex sequences can be developed from the building blocks and methodology. Indeed, a new project is probably best trialled in a test bed arrangement on a relatively small section of plant in order to gain

Debutaniser operating conditions	
Approximate equipment and operating specifications	
Cold feed rate, T/hr	4
Hot feed rate, T/hr	8
Reflux rate, T/hr	25
Overheads rate, T/hr	4
Bottoms rate, T/hr	8
Vent rate, T/hr	0.1
Condenser pressure, barg	13
Reboiler duty, MW	2.7
Total tray inventory, T	1
Sump inventory, T	4
Reflux inventory, T	8
Sump temperature, C	159
Reflux temperature, C	68
nC ₄ in bottoms, wt%	2
iC ₅ in overheads, wt%	1

Table 1

confidence and experience before rolling out across multiple units.

A detailed dynamic simulation model was created for the purposes of studying the start-up in detail. For an operating plant, it would be necessary to validate and tune the model against historical plant data to ensure that the model has sufficient detail to be a good (but not necessarily perfect) representation

of the plant and its process dynamics. After validating the model, the simulation must then be migrated to a ‘cold and empty’ condition. This is achieved by performing a process shutdown on the simulation, including venting, draining and purging, where appropriate. The migrated simulation provides a virtual test bed to develop an automated start-up sequence.

The recommended approach from here is to draft a sequence based on typical operating practices with a focus on automation and standardisation. This sequence can then be programmed into the simulation and iterated until an acceptable procedure is established within the simulation environment. The emphasis should be on robustness rather than accuracy.

Ruiz *et al* described the general principles of distillation column start-up:³

1. The discontinuous phase. This is the initial phase of the start-up, where liquid and vapour inventory is established. The trays are weeping and vapour can bypass the liquid via downcomers. Heating is applied.

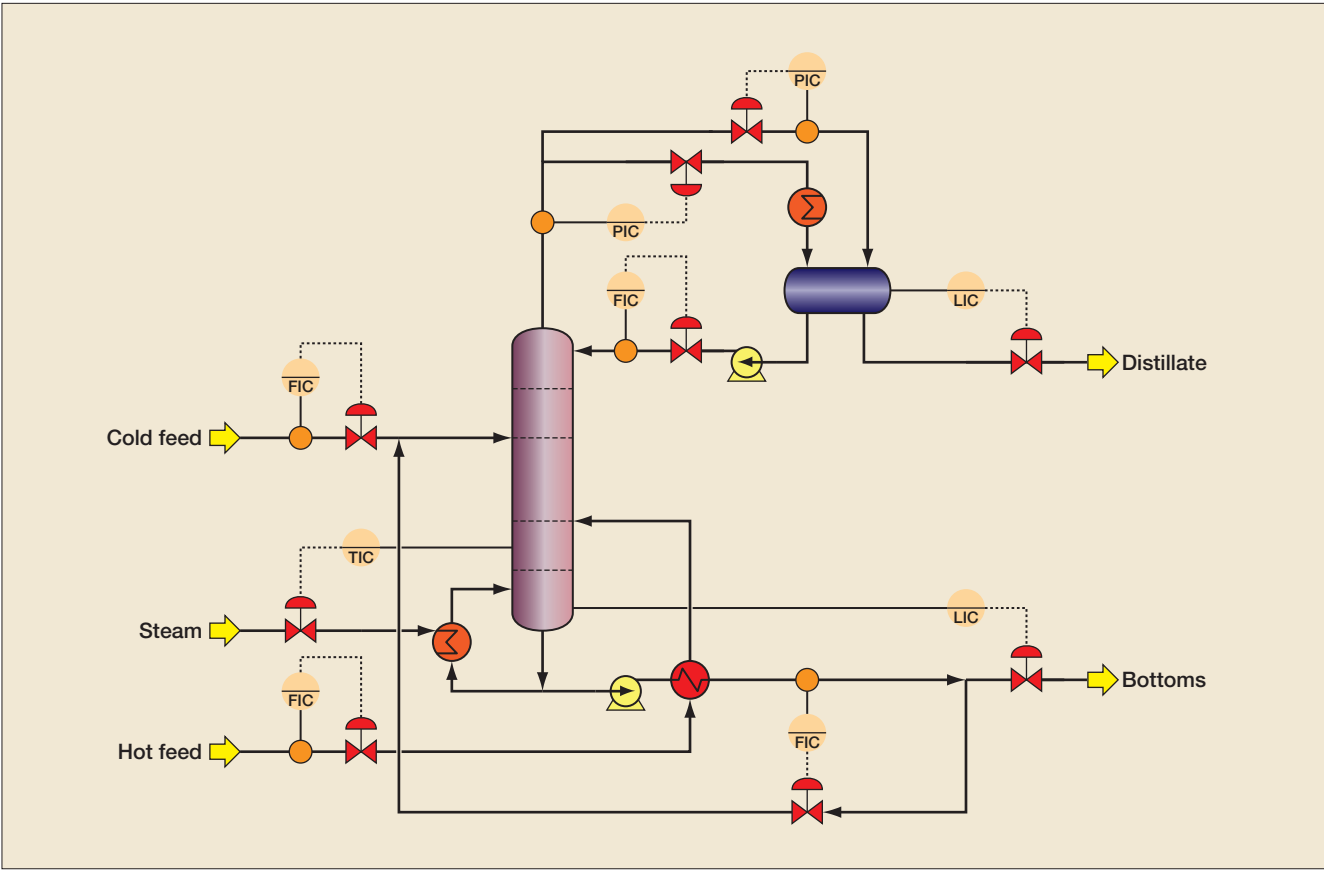


Figure 1 Debutaniser process flow diagram

2. The semi-continuous phase.

Liquid levels are established on the trays and vapour-liquid mass transfer starts. Pressures and levels come under control. Heating continues and the column approaches total reflux operation – stable operation without product flows.

3. The continuous phase. The column reaches the vicinity of the desired steady state. Temperature and composition profiles are established. Steady state controls become active to drive the column towards its expected operating condition.

A preliminary, generic start-up procedure for this distillation column can be drafted from these principles:

1. Supply liquid to the column to provide an initial inventory.
2. Apply heat to drive temperature and pressure towards normal operation.
3. Apply cooling to generate liquid inventory in the reflux drum and trays.
4. Stabilise operation with total reflux.
5. Establish a suitable composition profile.
6. Move to continuous operation with fresh feed and product draws.
7. Stabilise operation at full throughput.

It should be no surprise that the process control system design strongly influences the start-up procedure. However, just as the process design will have been developed from and for steady state operation, so, too, will the process control system. The implications of the control system design for start-up are rarely considered because the column will be expected (quite rightly) to operate at steady state much more than in its transitory states so that this mode is the appropriate focus.

A brief review of distillation control systems highlights some of the inter-relationships. A two-product distillation column will typically have five degrees of freedom (five final control elements) that can be manipulated to achieve the desired operating condition:

1. Heat input to the reboiler (V or Q_r)

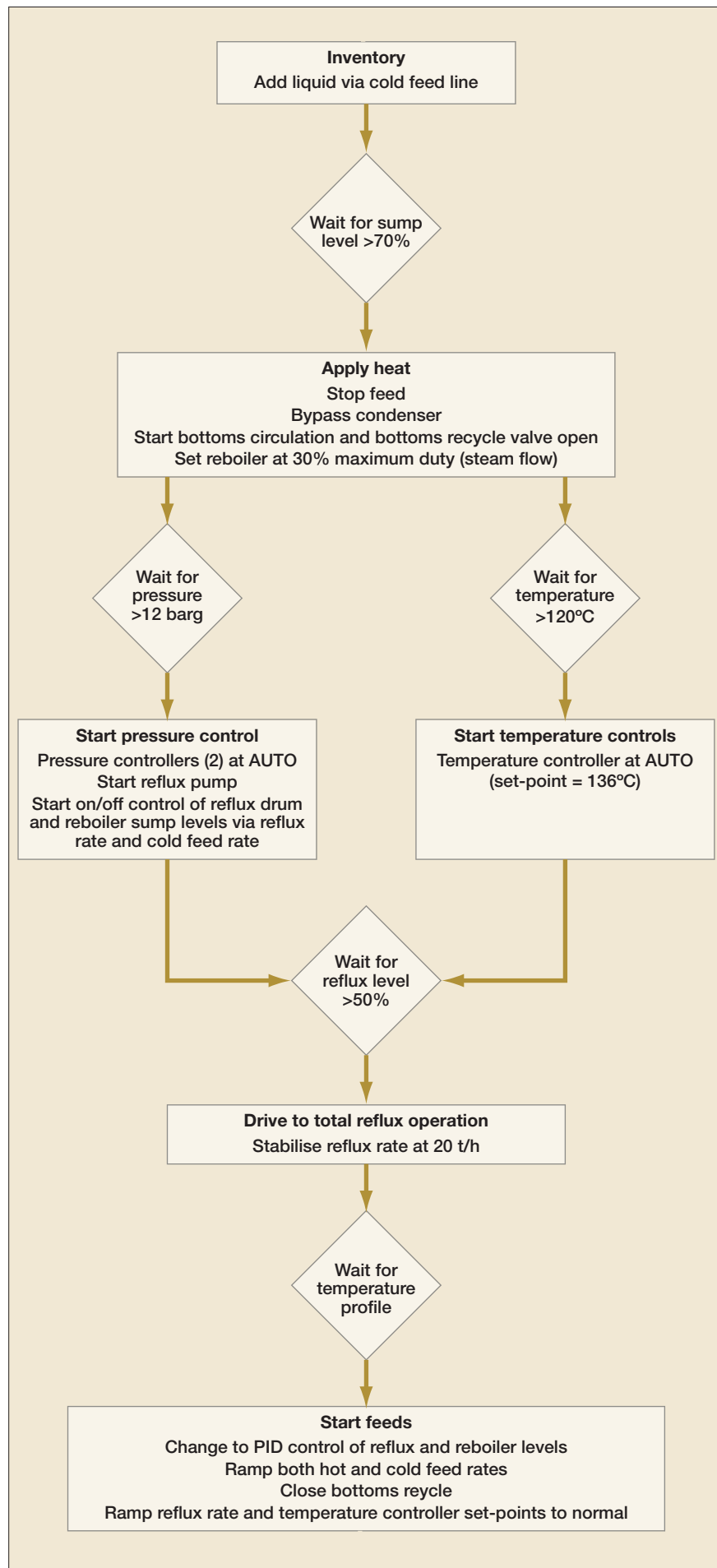


Figure 2 Basic procedure for automated start-up

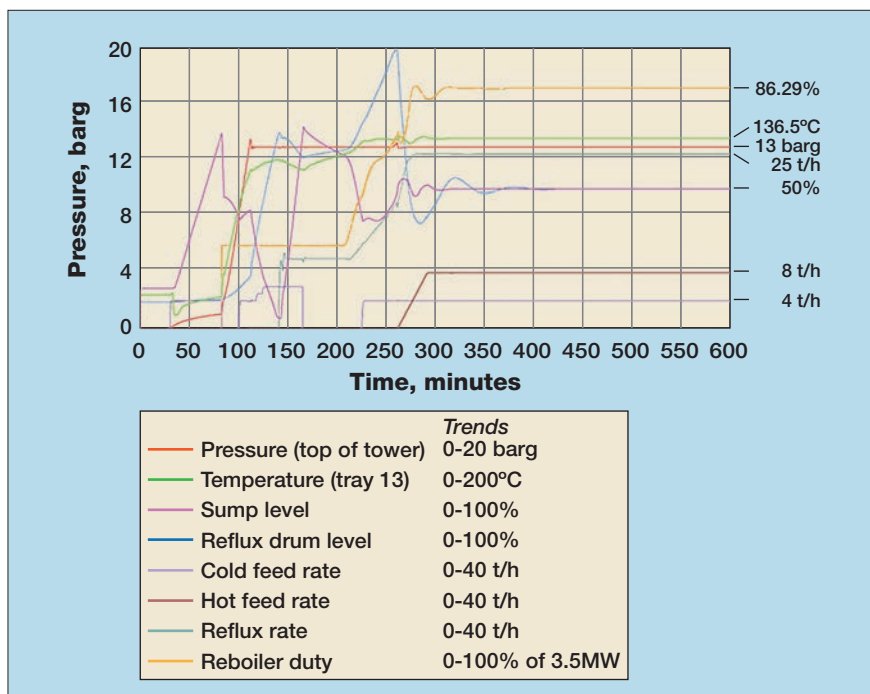


Figure 3 Automated start-up simulation results

2. Heat removed by the condenser (Qc)
3. Reflux rate (L)
4. Distillate product rate (D)
5. Bottoms product rate (B)

Of these, three must be used to maintain the column inventories: pressure, reflux drum level and reboiler sump level. The remaining two variables are available to control the product compositions. More precisely, one of the remaining variables will be used to control the column mass balance (split of feed into products) and the other to set the column energy balance or degree of fractionation. The two composition variables are often used to describe the control scheme. For example, a LV control scheme will utilise the reflux rate and reboiler duty to control the compositions. The distillate rate will be used to control the reflux drum level; the bottoms rate will be used to control the reboiler sump level; and the condenser duty will be used to control the column pressure. This is a very common configuration but other combinations are also widely used, depending on the process characteristics.

A significant consideration for start-up is that not all of the final control elements are available at all times. For example, distillate and

bottoms product flows start at zero and might not be available to manipulate until column conditions approach the steady state. This means that (automatic) inventory control using the product flows is not necessarily available during start-up. Interestingly, the rarely used DB control scheme (rarely used because it lacks flexibility because D and B are not strictly independent at steady state) offers the advantage that the steady state

A significant consideration for start-up is that not all of the final control elements are available at all times

inventory control scheme will work continuously during start-up. This significantly simplifies any start-up procedure.

Returning to the example case and simulation, it is necessary to assess the specific plant configuration and start-up challenges, including:

- The largest inventory in the system is in the reflux drum but the

control system is only effective in controlling this when the overheads product is flowing (not during the initial start-up phase)

- The second largest inventory is the reboiler sump but the control system is only effective in controlling this when the bottoms product is flowing (not during the initial start-up phase)
- The main (hot) feed requires the column to be hot before it can be introduced.

It is also prudent to do an analysis of things that can go wrong. This list will be specific to each case but there are some general principles that should be considered, including:

- Too much liquid added too soon so that the level increases beyond the measurable range (one of the primary causal factors at Texas City)
- Not enough liquid so that the reboiler runs dry, with potential for high metal temperatures
- Heat applied too quickly, leading to pressure overshoot and possibly unintended pressure relief, or excessive thermal stresses.

Finally, the physical parameters of the system need to be considered to quantify the individual steps (for instance, 'How much heat?' and 'How long for each step?').

All of these factors need to be addressed efficiently and robustly to produce an acceptable start-up (either manual or automated). The sequence shown in Figure 2 was developed after several iterations using the simulation test-bed. The simulation results are shown in Figure 3 and discussed below.

Discontinuous phase (two hours)

- The column is initially fed at 4 t/hr, which is the normal cold feed rate. It takes nearly two hours to build an inventory in the reboiler sump. This period is dependent on the column dimensions but is longer than might be expected because some liquid accumulates on the trays and in downcomers and some of the feed (volatile at low pressures) will flash as it enters the column.
- Heat is not applied until there is sufficient inventory. It then takes

one hour to build temperature/pressure in the column. The application of heat through the reboiler boils off much of the liquid inventory so that more fresh (cold) feed needs to be added.

Semi-continuous phase (two hours)

- Pressure control is established quickly.
- Reflux is started over 30 minutes to avoid an overly large disturbance to pressures and levels.
- The reboiler heat input needs to increase in line with the reflux rate. This is achieved by putting the reboiler temperature controller into Auto.
- Feed continues to be supplied to the column to replace the boil-off, although there is no direct calculation of the required rate of addition.

Continuous phase (three hours)

- The progression from semi-continuous to continuous phase is made when the column temperature profile starts to approach the expected steady state values.
- Forward flow is reached after approximately four hours.
- Steady operation is attained after a few more hours, although the product compositions change very slowly and concentrations have not yet fully stabilised after 10 hours.

There are a number of specific issues that are worthy of deeper analysis:

- Level control of the reboiler sump (LIC-101) and reflux drum (LIC-100) are sub-optimal (see **Figure 4**). In both cases, the normally operating PID loop is unavailable. The start-up sequence relies on on/off control using the cold feed and reflux rate, respectively. More advanced control is possible using a dedicated start-up PID loop.
- The pressure controllers are sufficiently well-tuned to be nearly decoupled from the temperature controller, except for the case when the condenser duty is approaching its operational limit (see **Figure 5**).
- There is an initial cooling effect during the discontinuous phase as the incoming feed flashes from its supply pressure (see **Figure 6**). The discontinuous phase could poten-

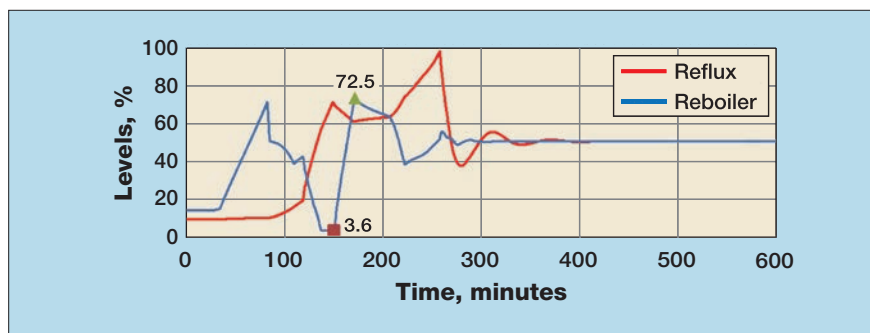


Figure 4 Sub-optimal control of reboiler sump and reflux drum

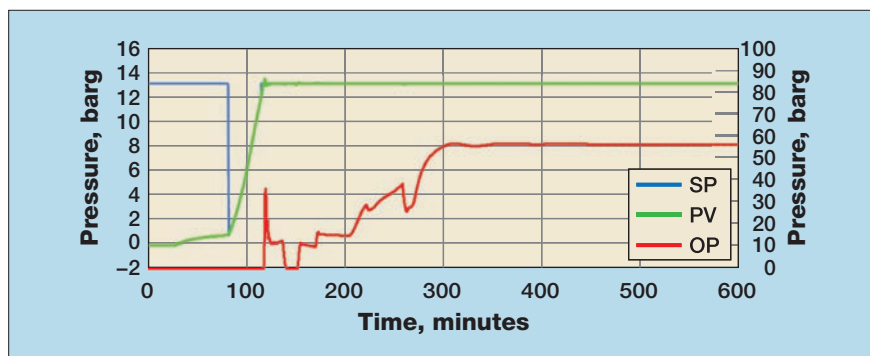


Figure 5 Pressure controllers are sufficiently well-tuned to be nearly decoupled from the temperature controller

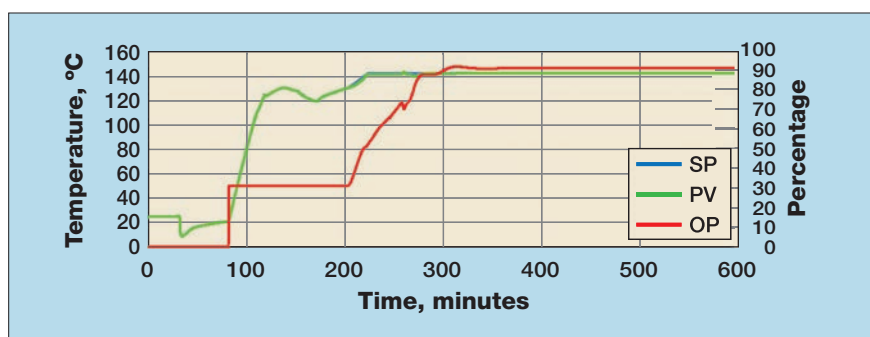


Figure 6 There is an initial cooling effect during the discontinuous phase

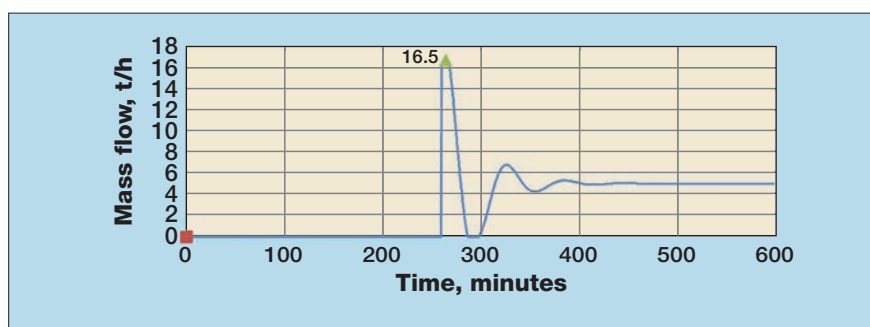


Figure 7 The transition from temporary level control to PID control results in some peaks of flow

tially be shortened by applying more heat but that might risk overshoot on temperature and/or pressure.

- The transition from temporary level control to PID control is relatively smooth but results in some

peaks of flow (see **Figure 7**). Compositions continue to change for 3-4 hours after flow is started (see **Figure 8**).

This start-up is unlikely to be considered perfect but might well be judged acceptable. Further simu-

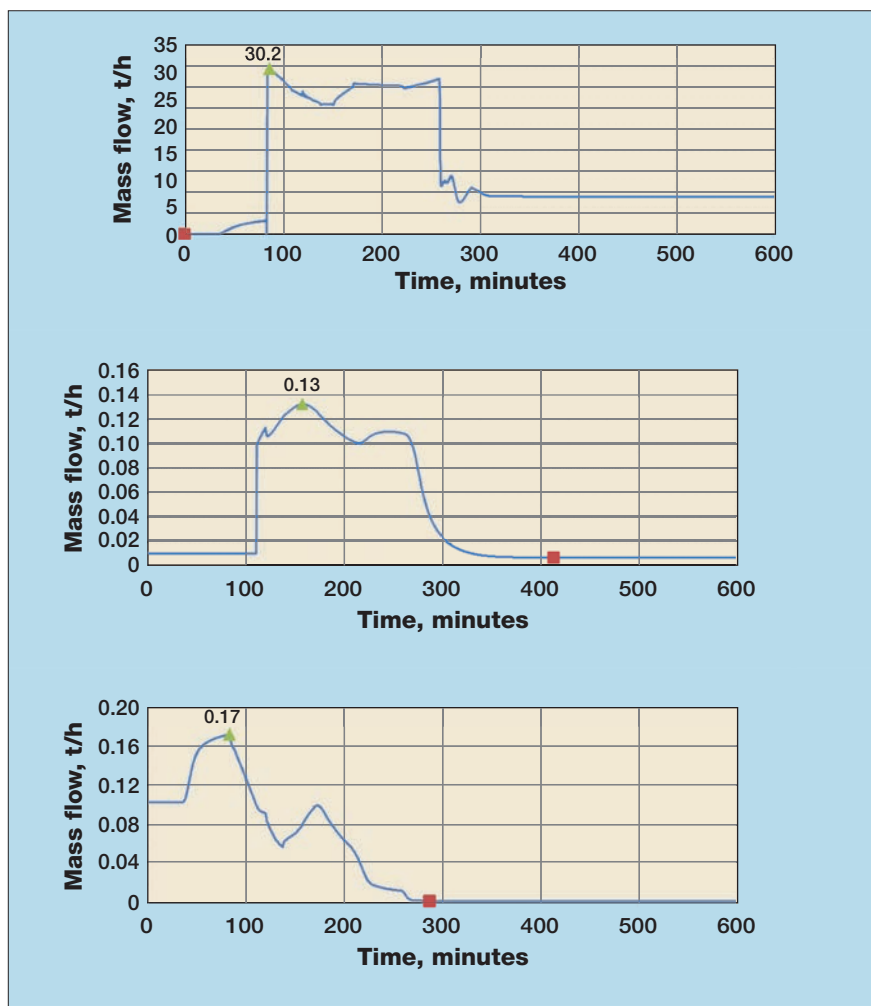


Figure 8 Compositions continue to change for 3-4 hours after flow starts

lations using aggressive operating techniques (a limiting case that is unlikely to be employed on an active plant) suggest that it is possible to reach the continuous phase more quickly. However, the composition dynamics are relatively slow and very substantial additional time savings are not available. Nevertheless, further optimisation (and increased complexity) of the existing procedure is clearly possible to improve on the preliminary procedure described below. Robustness should be the primary goal before increased time savings. This can also be investigated using the simulation by re-running cases with variations in equipment specifications and boundary conditions. Speed improvements, if sought, are realisable by accelerating the progress through the various start-up phases (fill more quickly; heat more quickly; and so on). This needs to be done whilst considering mechanical constraints and a

good safety analysis to correctly balance the risks and rewards.

These (preliminary) results are promising and suggest automation is desirable if it can be accomplished with a modest amount of engineering effort. The actual project execution yields its own additional rewards: improved understanding of the column in its transitory state and greater appreciation of the risks involved during the process transitions. At worst, an automated start-up is something worth exploring; at best, it might improve average start-up times and deliver a strong safety case and the opportunity to reduce the load on the process operators who are always likely to be stretched during start-up.

Conclusion

Automated column start-ups might not be considered 'low hanging fruit' but they are something that could easily be considered 'world-

class'. There are some potential economic benefits but improved process safety is likely to be the main driver for an automation project. Dynamic simulation is the perfect tool to explore this type of opportunity. However, it is important to recognise that the simulation itself is just a tool, which must be used with intelligence, insight, understanding and experience. A small, mixed team is usually best – an experienced operator, an experienced process engineer, a simulation expert and a control system engineer. Get these people together and focused, and they will deliver value. Expert solutions using dedicated multi-variable control systems are possible but not necessarily required or justified. Instead, consider what is possible with your local team, enhanced by, maybe, one or two external consultants.

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Martin Sneesby is an independent dynamic simulation consultant with Advanced Process Engineering & Simulation Services (APESS) in Cambridge, UK. He has more than 20 years of experience in process simulation and modelling, including many operator training simulators and detailed engineering studies utilising a wide range of dynamic simulation tools and software in many of the major process engineering sectors.

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Upgrading FCC catalyst performance

A US refiner processing crude from the Permian Basin, including both sweet (WTI) and sour (WTS) crude oil, underwent a major FCC turnaround in 2014. Following this major revamp and re-baseline of the unit, the refinery sought to increase profitability via a catalyst change. Unit objectives were: increase LPG olefin yield, preferably C₄ olefins; reduce dry gas make; improve slurry upgrading; increase LCO yield; improve coke selectivity; and improve conversion, all at constant fresh catalyst addition rates. An additional operating objective of the refinery is to keep the downstream alkylation unit full year-round, using a combination of FCC operating parameters, an optimised catalyst formulation, and traditional ZSM-5 additive as needed. The FCC feed is comprised primarily of a VGO feed-stock, plus a combination of various side streams.

Achieve 400 FCC catalyst

Refiners are continually challenged to respond to changing market dynamics. For example, a common challenge for refiners operating on unconventional feeds, such as shale or tight oil, is a loss of gasoline pool octane, caused by reduced volume of alkylation feedstock. Grace launched the first product in the Achieve series, Achieve 400, to address these octane debits. This breakthrough technology has delivered an economic uplift on the order of \$3-7 million/y in multiple commercial applications due to the boost in slurry conversion, FCC naphtha octane, and butylene yield.

During the Achieve R&D programme, five key catalytic functionalities were developed to provide the yield flexibility desired by refiners:

- Increase distillate yields with high diffusivity matrices
- Reduce dry gas with advanced metals traps
- Drive conversion with ultra-high activity zeolites
- Maximise resid processing with leading coke selectivity
- Boost refinery octane with dual-zeolite technology.

Achieve catalyst is formulated to include each of

Reactant	Selectivity	Relative reactivity	C ₄ = / C ₃ =	Achieve 400
C ₆ →	2C ₄ = 44% C ₃ = + C ₅ = 56%	100	1.57	<div style="display: flex; align-items: center;"> <div style="width: 10px; height: 10px; background-color: green; margin-right: 5px;"></div> <div style="writing-mode: vertical-rl; transform: rotate(180deg);">ZSM-5 additive</div> </div>
C ₇ →	C ₃ = + C ₄ = 95% C ₂ = + C ₅ = 2%	12	1.00	
C ₈ →	2C ₃ = 83% C ₂ = + C ₄ = 16%	1.5	0.10	

Figure 1 Selectivity comparison between Achieve 400 catalyst and traditional ZSM-5 additive

these five critical functionalities. High diffusivity matrices maximise distillate yield from the bottom of the barrel and provide resistance to poisoning from unconventional metals. The catalyst can be formulated over a range of activity, rare earth exchange, and isomerisation activities. Grace's capability to modulate hydrogen transfer activity in self-manufactured zeolites affords each refiner the optimal LPG to gasoline ratio for their operation. Achieve 400 also features multiple zeolites, specifically traditional Y-zeolite and pentasil-type zeolite, to selectively enhance LPG olefinicity by preferentially cracking gasoline olefins at C₇ and above into butylenes.

Incorporation of isomerisation activity into the cata-

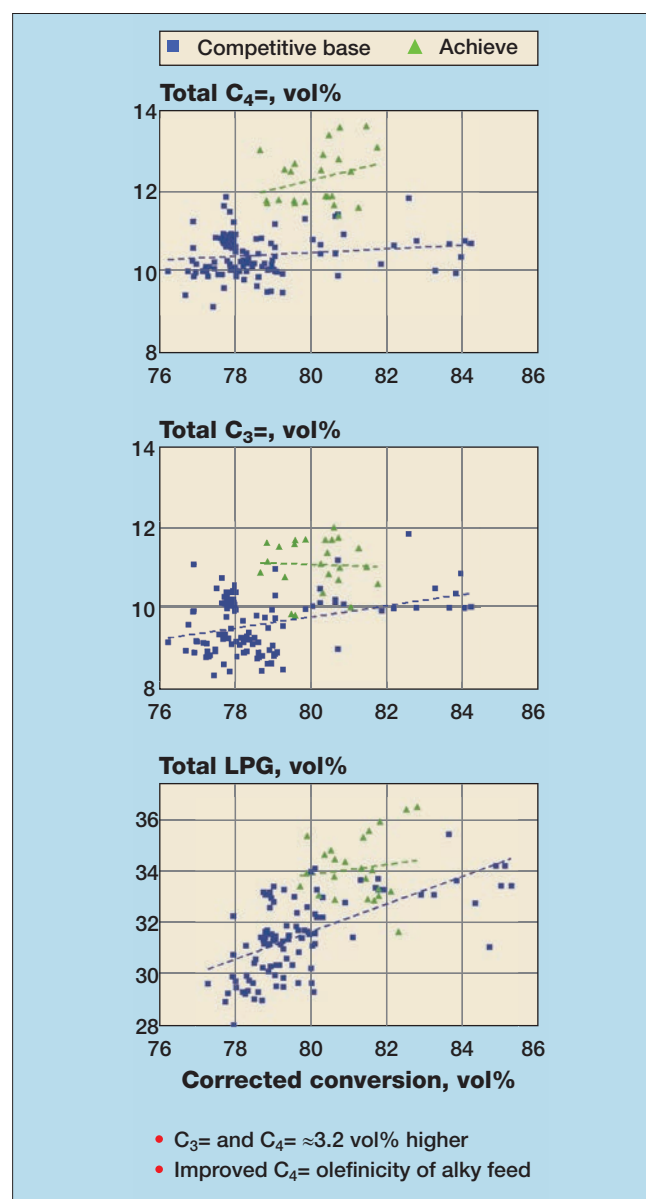


Figure 2 Total LPG, C₄ olefins and propylene yield at constant conversion

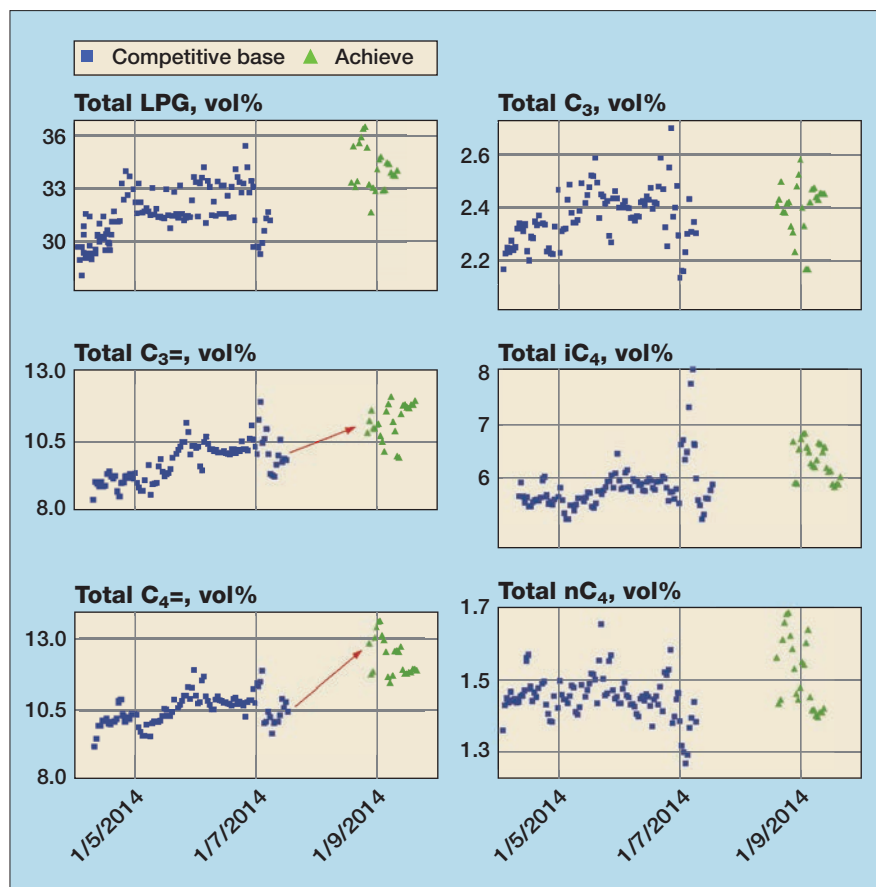


Figure 3 LPG yield breakdown suggests minimal increase in saturate production

lyst particle itself results in a more desirable yield pattern than would be realised by use of a traditional octane boosting FCC additive. Achieve 400 catalyst yields higher octane, with an improved butylenes to propylene ratio when compared to ZSM-5 additives. The zeolite isomerisation activity in the catalyst increases the yield of FCC butylene and iso-butane, keeping the alkylation unit full and maintaining refinery pool octane.

With traditional ZSM-5 technology, cracking of gasoline range olefins continues on into the C_6 range, generating a disproportionate amount of propylene relative to butylenes (see Figure 1). Achieve 400 catalyst, however, works to selectively crack gasoline olefins at C_7 and above into preferentially more butylene, thus generating a higher ratio of C_4 to C_3 olefins than separate light olefins additives.

US refinery trial

In mid-summer of 2014, the refiner began a trial of Achieve 400 against a traditionally formulated high activity base catalyst. The refinery previously utilised approximately 1.5-3.5 wt% ZSM-5 additive to attain the desired olefinicity and octane shift.

Throughout the catalyst trial, fresh catalyst addition rates were held constant, and Ecat activity remained at 74 wt% on average. Total LPG olefin yield increased by ~3 vol% (see Figure 2). The alkylation unit remained at full capacity, despite a gradual reduction in additions of a separate particle ZSM-5 additive. The site eventually stopped adding ZSM-5 additive and was still able to

Figure 4 Liquid yield shifts, consistent with refinery trial objectives

keep the alkylation unit full.

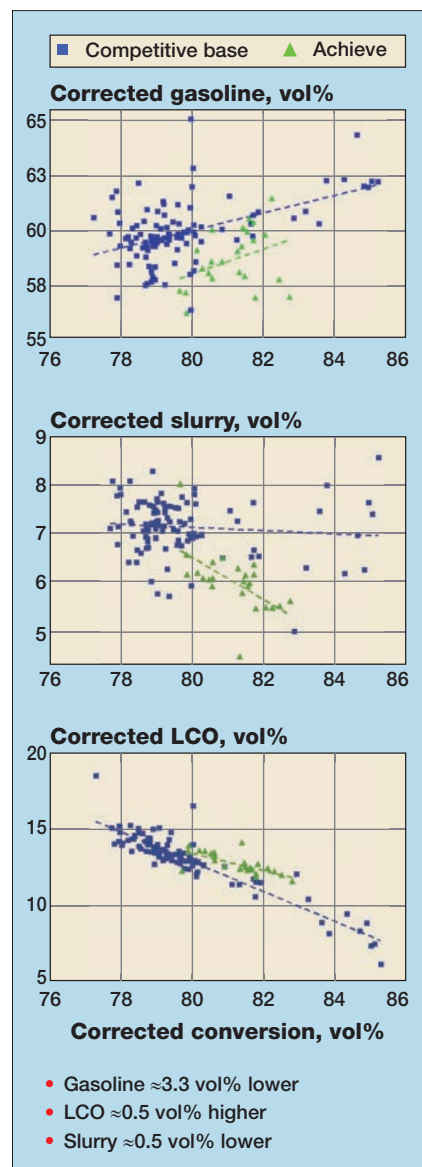
The constrained downstream alkylation unit benefited from the increased loading of C_4 olefins in the feed. Moreover, there were minimal increases in C_3 and C_4 saturates commensurate with these increases in olefin yield. This can be attributed to the tailored acidity and optimal hydrogen transfer activity of the Achieve catalyst system. LPG saturate yields are shown in Figure 3.

Moving to a low Z/M, high activity matrix catalyst also helped to improve slurry upgrading and increase LCO yield. Slurry yield dropped by ~0.5 vol%, while LCO yield increased by ~0.5 vol% at constant conversion. Plots of these shifts are shown in Figure 4.

Additional benefits included reduced dry gas at comparable metals levels on Ecat, as well as improved overall conversion and volume swell.

Results and conclusions

A review of the unit data shows that the refiner was able to significantly improve profitability during the trial. Stated objectives were met, and the refiner obtained an additional benefit of lowering opex by reducing and eventually stopping addition of ZSM-5 additive at full FCC charge rates. Due to the success of



the trial, the refiner is still using Achieve technology at this FCC today.

The favourable performance of Achieve 400 has generated an economic uplift of approximately \$0.70/bbl.

As refiners' needs continue to become more complex and diverse, FCC catalyst suppliers are continually challenged to respond with new technology to meet these evolving needs. Catalyst technologies must be flexible and adaptable, and continue to fulfill short term refinery objectives. The Achieve series (100, 200, 300, 400, 800) comprises a suite of state-of-the-art catalyst technologies, optimised to meet specific refinery opportunities while not exceeding operating constraints. These FCC catalysts have been applied in 30 commercial applications to date, and have delivered between \$0.40-0.95/bbl of incremental value.

Gary Cheng, Kevin Burton, Kent Turner, Chip Schaeffer and Rosann Schiller
Grace Catalysts Technologies

For more information: Rosann.Schiller@Grace.com

Resolving piping corrosion and replacement

Amec Foster Wheeler provided engineering analysis for a fine tailing treatment project in the Athabasca Oil Sands region in Alberta, Canada. The project handles 100 000 b/d capturing the fines from flotation tailings through a thickener. The project equipment included the flotation thickener, the recovered bitumen tank, the overflow water tank and gravity distributor, slurry pumps, the gland water pump and bitumen sump pump.

The 42 inch diameter pipe for overflow water would operate at a temperature of 95°C and a pressure of 2500 KPag. The pipe runs 1000 metres in a common trench with four other pipelines. Normal engineering practice calls for carbon steel (CS) for deep buried piping, but for overflow water service, the service life of CS is only five to eight years due to corrosion. It would be difficult to replace this individual pipe in a common trench with other pipes, entailing huge costs for soil excavation and backfill. To achieve a lifetime operation, fibreglass pipe (FRP) material would be needed.

Calculating shallow-buried pipe stress

Amec Foster Wheeler used Intergraph Caesar II for 56 stress calculations. The FRP overflow line required the UK Offshore Operation Association (UKOOA) design code, one of the many standards included with Caesar II.

Being shallow-buried, the FRP pipe would experience higher movement and greater chance of bend failure plus high thermal movement with all changes in direction and at the buried terminal points. The engineers input the soil data and FRP configuration into Caesar II to compute the soil stiffness in axial, lateral and vertical directions.



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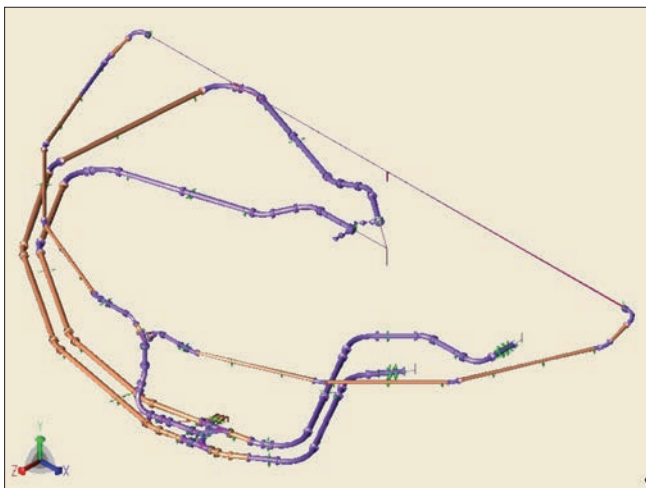
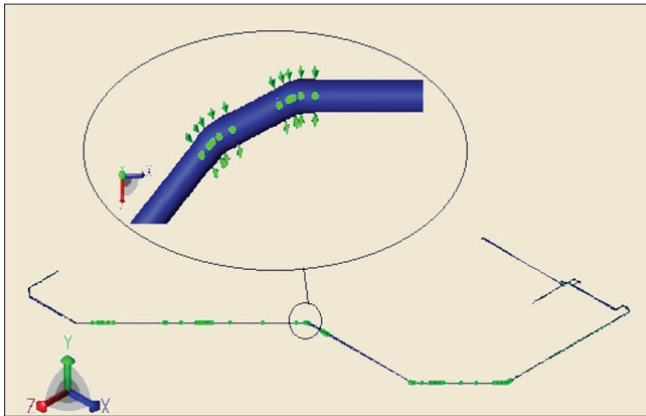
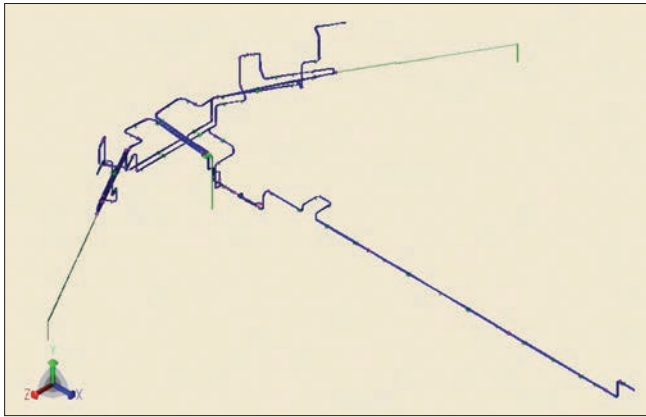


Figure 1 For the design of a 42 inch FRP pipeline in the Athabasca Oil Sands, Caesar II's graphic modelling capabilities avoided a loss of accuracy and design quality that would otherwise be a problem

To prevent bend over-stress, the engineers recommended installing resilient ethafoam material customised to the bend profile. Using the Caesar II buried pipe modeller, the team input the ethafoam spring rate to calculate the foam stiffness, length and number of layers to make sure the FRP pipe would be safe.

Ensuring data accuracy

With Caesar II's graphic modelling capabilities, the company avoided a loss of accuracy and design quality that would otherwise pose a problem under such a tight schedule. The team also avoided many wasted

hours on analysis that would have been required to address code requirements on such unique and challenging design conditions.

"This would typically require numerous tries, and we would have missed the project deadline without Caesar II," explained Bhaskar Shitole, piping discipline lead at Amec Foster Wheeler.

"With Caesar II, we saved a lot of time and were able to adjust the foam length to prevent bend over-stress. It helped us optimise the design and ensure quality, which in turn saved us in project costs."

Amec Foster Wheeler received the 2015 Caesar II Runner-Up Drivers of Success Award for its use of the software. The annual competition recognises innovative applications of Intergraph products, impressive project results and significant benefits from collaboration among disciplines and the integration of the products.

Intergraph Corporation

For more information: icas.marketing@intergraph.com

Getting to grips with noise

A particular challenge when designing valves for compressible media and in particular steam is the noise calculation.

For many years, Arca Regler GmbH has been designing, building, testing and delivering all kinds of valves for the circulation of steam, condensate and water in industrial power plants and in particular in refuse incineration plants, or thermal waste recycling plants as they are called today.

In addition to the plant-specific operating data and performance features, growing environmental consciousness means that great importance is increasingly being attached to noise emissions. There are clear legal directives for this as well as arithmetic methods for performing a preliminary calculation. Arca makes contact with plant planners and plant manufacturers at a very early stage in order to harmonise the plant needs and technical requirements. This is very important for noise emissions in particular, because several factors are incorporated into the preliminary calculation here, such as the inlet and outlet sections, the dimensioning of the nominal sizes, the pipe wall thicknesses and the structure of the decoupled thermal and noise insulation. Only then can already very precise noise calculations be carried out, only then can the control valves that have to relax the fresh steam from a high primary pressure down to a vacuum be precisely designed and only then can the combination of regulated and unregulated reduction steps be defined. Apart from the standardised calculation principles, a great deal of experience is required in order to be able to design as safely as possible.

Nevertheless, it is occasionally the case that theory and empirical values do not suffice in practice. For instance, Arca measured considerable development of noise in the case of a generously sized, multi-stage, subcritical design in the turbine bypass station that

relaxed down to vacuum. Since this was noticed during the commissioning phase, fast action was absolutely necessary. As the boiler had to be operated for planned tests and the turbine was not yet in operation, the fresh steam had to be discharged via the bypass station. Immediate measures had to be initiated for the employees who were still busy with the construction and Arca looked for a solution together with the plant manufacturer.

First of all a noise spectrum was recorded for various load cases and the load cases were compared with the design data. Deviations were found, but these lay within the tolerance field. As a side effect, it could clearly be seen how accurately Arca had performed the flow calculation for the very complex multi-stage valve. In parallel, the isometrics and the structural implementation of the thermal and noise insulation were checked. Here, too, no striking deviations were found that could have explained this noise phenomenon.

Hence, the only thing that remained was to check secondary measures. To this end the flow design with the various regulating and fixed stages was recalculated with the current operating data to find possibilities to reduce the noise. An additional throttle package was developed as a solution. The particular challenge here was that the throttle package had to be retrofitted into the existing DN 900 pipeline and calibrated with the existing regulating and fixed stages.

In close cooperation with the customer, the solution was presented and the decision was taken immediately to implement it. The exact positioning of the throttle package in the existing pipeline was also intensively discussed. Thus, only a few days remained before the next planned shutdown to precisely dimension and manufacture the interface for the throttle package in the pipeline. Accuracy took precedence over speed here. Thanks to special measures and particularly intensive chasing by the project team, the throttle package arrived at the building site in time for the planned plant shutdown, which lasted only two days.

The thermal part of the refuse incineration plant was started up via the turbine bypass stations and put back into operation. The result of the research and recalculation could clearly be heard – or rather, to put it better, could no longer be heard.

Afterwards an attempt was made to comprehend the noise phenomenon arithmetically with the retrofitted implementation and the operating cases concerned. These efforts and also the involvement of noise experts produced no conclusive explanation. This phenomenon could only be solved thanks to Arca's experience and knowledge, without knowing the exact cause. All the richer for the experience, Arca has proven once again that it can react quickly and competently to such cases and is always there to help its customers.

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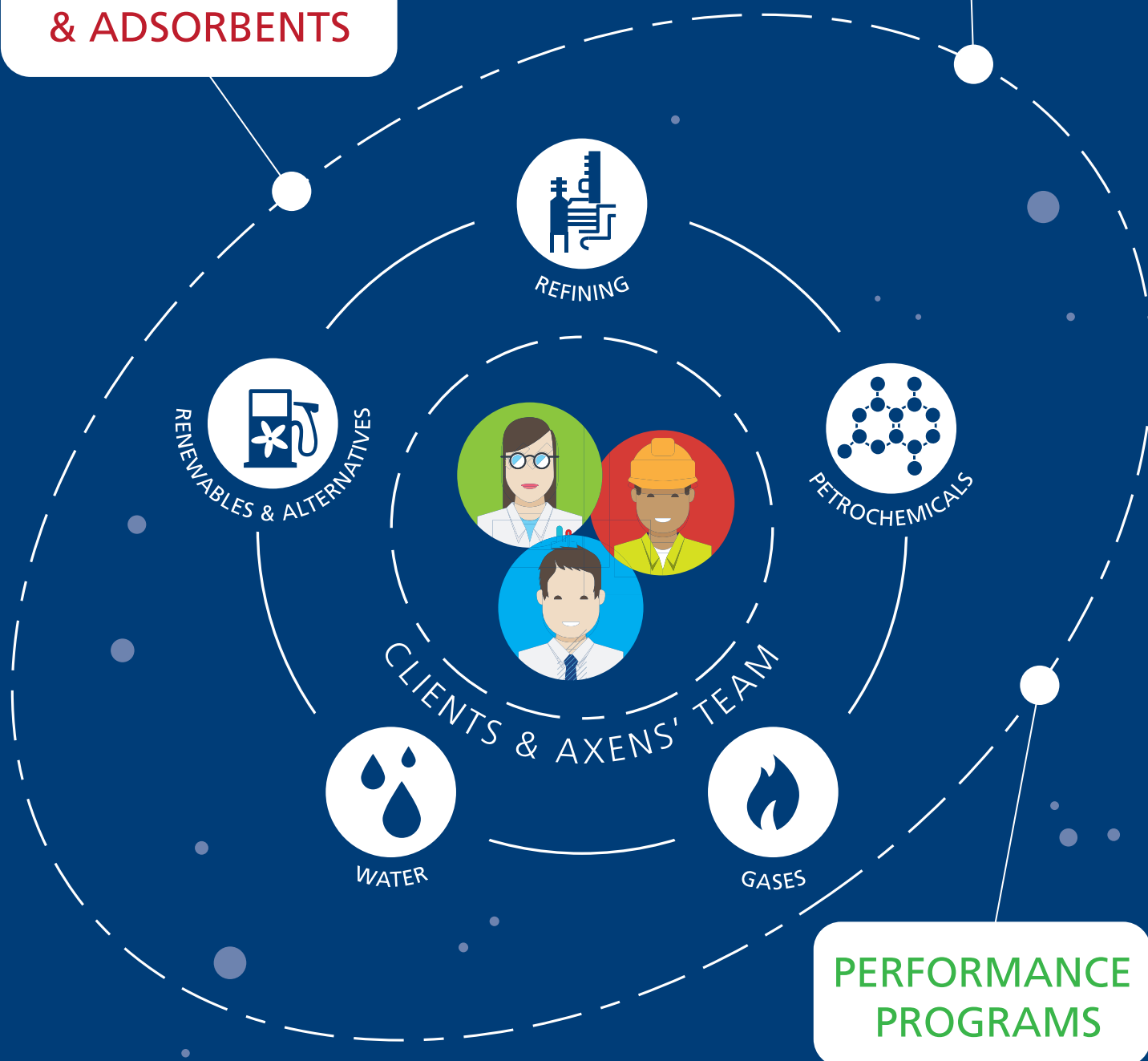
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