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#### Cover

The SATORP export refinery in Jubail, Saudi Arabia processes 400 000 b/d  
Photo: TechnipFMC

The ExxonMobil logo is positioned in the top right corner. It features the word "ExxonMobil" in a bold, white, sans-serif font. The "X" is stylized with a double-stroke effect. The background of the entire advertisement is a dark, almost black, space filled with a complex, glowing molecular structure. This structure consists of a dense network of red and blue spheres connected by thin lines, representing a complex chemical or catalytic process. A bright, white, glowing point is visible within this network, from which a stream of light flows upwards, transitioning through colors like yellow, green, and blue, eventually forming a large, ethereal, purple and blue shape that resembles a flame or a large molecule. The overall aesthetic is scientific and futuristic.

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## Dual dilemma

There is no shortage of advice arriving daily in this Editor's inbox about the impact of coronavirus on the global oil industry, on crude prices in particular. They could fall to \$20/bbl, says one source; naphtha based ethylene production is back in favour, says another; and so on.

For its part, the International Energy Agency offers multiple scenarios, as it tends to do, for its analysis of the world picture. In its "pessimistic low case", the agency's global forecast is for a fall in daily oil demand by 730 000 barrels through 2020. The IEA's more optimistic outlook assumes that demand hardly falters, transport is closer to normal, and global demand grows.

Prices rise and fall; margins do the same; the world goes around; but the pandemic develops faster than expert analysis.

The simple argument goes like this: lower oil prices mean higher refining margins, meaning happiness for refiners but less so for producers. A reported instance in China during March saw its independent refiners doubling their margin in a week to a per-tonne \$80 as crude dipped to around \$35/bbl.

Notwithstanding that, the global oil industry has contrived its own mutation of the crisis, with resulting complex diagnostics. This raises the question of how much permanent damage may be done to the petroleum industry.

OPEC producers led by Saudi Arabia met in early March to cut their output by a little over 2 million b/d, in response to tanking global demand brought about by coronavirus and in an effort to shore up prices. Russia declined to follow suit and prices continued to fall.

The pivotal case in point to illustrate the complexity of this coronavirus-oil price nexus is to be found in the US fracking industry. At the time of writing, the developing impact of the virus on the people of the US was far from clear. The impact on tight oil producers arising from low prices is undoubtedly challenging.

It is fair to say that oil fracking has been the most extraordinary bonanza of recent years. As a result of its super-rapid development in the West Texas Permian and elsewhere, and a relaxing of export rules, the US has become the dominant new force in global oil supplies.

It is also true to say that it is a bonanza with plenty of potential for bust as well as boom. Such is the popularity of light, sweet supplies of crude among refiners and petrochemicals companies, fracking companies have not held back in a headlong rush to supply a seemingly bottomless pool of demand.

The problem is that fracking is an expensive business with relatively low potential for profit-making. Much of it is financed by debt, and when a previous instance of tumbling world prices came around many producers had to seek even more billion-dollar debt.

Right now, demand for oil products is down. Depending on the progress of the pandemic, demand could fall much further. Oil prices remain stubbornly low while Saudi-Russian arm-wrestling continues, too low for producing companies which in many cases struggle to turn a profit from a market level of \$50/bbl.

And those debts have to be called in some time.

CHRIS CUNNINGHAM



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**Q** We are processing high TAN crude and encountering corrosion problems in the vacuum heater of our VDU. Should we opt for pretreatment or new metallurgy?

**A** Collin Cross, Senior Product Analytics/Support Manager, SUEZ – Water Technologies & Solutions, collin.cross@suez.com

For many refiners who regularly process high acid crudes (HAC), the two options are not necessarily mutually exclusive. While the two tactics are both used for mitigating corrosion from HACs, the two options can sometimes be more complementary than competitive. This is because there are pros and cons to each method, the specific benefits of which depend upon a wide number of considerations. Some, but not all, of these considerations are:

1. Is HAC processing to be used as an ongoing strategy, or to take advantage of occasional opportunities?
2. What types of crude are being considered, what are their respective ranges of both TAN and sulphur?
3. What types of advanced metallurgy are already present in the crude unit?
4. How hard is the unit being pushed relative to the design case, and how will this affect velocity and shear stress throughout the unit?
5. What type of monitoring is being used to detect active corrosion and potential control chemical injection?

For a refinery that wants to occasionally take advantage of opportunity crudes, then chemical treatment is often considered as the best option. On the other hand, if the refinery wants to implement a strategy to consistently run HAC crudes, then metallurgy might be a better long term solution. Other important considerations are the relative ease and availability of capital budget vs operating budget as well as availability and lead time for both engineering design and maintenance resources, plus material availability and spot prices. Oftentimes, due to the timing and cost involved with the installation of advanced metallurgy, starting a chemical treatment programme can help to more quickly begin processing discounted HAC crudes. The improved profitability of many of these crudes can then be used to help fund and/or justify the more expensive and longer term option of advanced metallurgy.

Another important point is that many high temperature corrosion inhibitors work very effectively on stainless steel at even lower doses than needed for carbon steel, or both 5 and 9 chrome metals. Because the installation of advanced metallurgy is a costly long term plan, then using small dosages of chemical to help to preserve the lifetime of the investment after it is installed can also offer a good business case.

Finally, monitoring is extremely important to detect ongoing corrosion and its patterns and movements over time with changing crude slates. While advanced

metallurgy, especially 317L stainless steel, is extremely resistant to HAC, it is also quite expensive, especially if more than a single vacuum transfer line is desired to be protected. Many cases exist where no more than a TAN of 1.5 is planned to be used. In these cases, a well monitored system using multiple injection points and flexible control over the chemical can rival the results of metallurgy. If very high TAN levels are desired to be run for extended time periods, then metallurgy really becomes the best option over time.

**A** Chris Claesen, Director, Technical Consulting, NALCO Water, cclaesen@ecolab.com

As the TAN of the processed crude feed and associated streams increases, the areas particularly susceptible to naphthenic acid corrosion are those that encounter mixed phase flow, hot temperatures, and high vapour velocities; these are typically the radiant sections of furnaces, furnace outlet headers, and transfer lines.

With partial vaporisation of the feed occurring in the furnace, the remaining heavy liquids have in some cases the potential to be significantly higher in TAN. The residual liquid droplets in the high velocity mixed phase stream provide a good solvent for any metal naphthenates that are formed on the pipe walls, and where direct impingement occurs at elbows the shear forces will increase the removal of metal naphthenates. The net result can be very high rates of naphthenic acid corrosion in these susceptible areas.

The obvious solution to process higher TAN crudes in greater percentages is to employ construction materials that are resistant to naphthenic acid corrosion. Chromium containing steels, such as 5Cr, 9Cr, and AISI SS410, have commonly been used to provide increased resistance to sulphidic corrosion, and while some can provide some protection against naphthenic acid attack, the observed corrosion rates can still be unacceptably high.

For metallurgical protection against naphthenic acid attack, it is necessary to incorporate higher molybdenum into the alloy compositions. Of the materials commonly found in refinery environments, only AISI 316SS and AISI 317SS in the austenitic stainless steel family possess adequate resistance to naphthenic acid attack. The specification of these two alloys is for 2.0-3.0% molybdenum and 3.0-4.0% molybdenum respectively. For existing refineries built without these enhanced alloys, the option to fully upgrade the metallurgy is expensive, must be planned accordingly, and requires an extended shutdown period to upgrade the plant.

However, if the refinery is already committed to processing naphthenic acid crudes, the use of proven chemical corrosion inhibitors such as the Nalco Scorpion programme can be a very effective means of managing the threats. The protection philosophy for application of naphthenic corrosion inhibitors is based

on passivation of susceptible metal surfaces and this is achieved by converting the ferrous alloy surfaces from a typical sulphide film to one which is far more resistant to attack from naphthenic acids.

The key to maximising the chances of success of a chemical corrosion inhibition programme is to ensure that a proper and rigorous risk assessment is undertaken. The risk assessment procedure uses a number of important factors that affect the corrosion rates to provide an indication of the risk level associated with processing acidic crude blends across the whole process unit. The outcome of the risk assessment will outline the specific areas susceptible to naphthenic acid corrosion, identify suitable corrosion monitoring locations and which section requires chemical corrosion protection.

By applying this Scorpion risk based methodology coupled with appropriate corrosion monitoring strategies, chemical inhibitors can be very effective in mitigating naphthenic acid corrosion. For furnace tube corrosion, it is accepted that there are no viable methods to monitor corrosion inside the furnace tubes itself. However, placing appropriate non-intrusive corrosion monitoring on sections such as furnace outlet headers and transfer lines these can be used effectively to monitor corrosion and optimise the use of inhibitors to mitigate the threats posed by naphthenic acid corrosion.

**A** Sophia Xiaoxia Zhu, Materials & Corrosion Engineer, Shell Global Solutions US, Xiaoxia.Zhu@shell.com and Kaushik Majumder, Team Lead-Distillation and Thermal Conversion, Shell India Markets, Kaushik.Majumder@shell.com

High temperature corrosion in the crude distillation unit (CDU) and vacuum distillation unit (VDU) can be affected by many factors including total acid number (TAN); therefore the mitigation strategies should consider all the major contributing factors. The solution could be case specific, depending on the main mechanisms of the corrosion issue.

High temperature corrosion in refineries results from the sulphur and naphthenic acid compounds that are naturally present in crude oils. Sulphur compounds contained in crude oil or produced from thermal decomposition upon distillation range from gaseous ( $H_2S$ , for instance) to refractory (dibenzothiophenes, for instance) compounds. Accordingly, some form of sulphur distributes into all of the distillate fractions in the atmospheric and vacuum distillation units. Sulphur concentration increases with boiling and is most concentrated in the heavier fraction. Sulphidic corrosion tends to be self-limiting due to the development of a partially protective metal sulphide film. Naphthenic acid is a thermally stable, high boiling compound and tends to concentrate in 200-425°C TBP range products in atmospheric and vacuum columns. It produces low carbon organic acids such as formic, acetic, propionic, and butyric acid on thermal decomposition. Reaction with metal produces hydrocarbon soluble metal naphthenates, which dissolves away, leaving a lean, scale free metal surface. Naphthenic acid tends to proceed at a constant rate due to the oil soluble metal naphthenates that are formed.

For crude oil and distillate streams, sulphidic and naphthenic acid corrosion occurs simultaneously, and metal sulphide layers also protect from naphthenic acid corrosion. There is competition between the FeS layer formed and naphthenic acids breaking it down. Low  $H_2S$  partial pressure enhances corrosion due to insufficient FeS film formation. However, high  $H_2S$  partial pressure results in formation of pyrophoric iron sulphide. For this reason (and others), corrosion behaviour cannot be predicted based solely on TAN or sulphur concentrations. The corrosion rate depends on the concentration and type of sulphur and naphthenic acids present, the material of construction, velocity, temperature, and so on. These factors will affect the thickness, porosity, and protectiveness of the metal sulphide (and possibly metal oxide) films which determine the steady-state corrosion rate.

The traditional rule of thumb for identifying high TAN crudes has been a whole crude TAN of 1.0 or a sidecut TAN of 1.5. However, unexpected corrosion in high velocity areas of CDUs has been observed with whole crude TAN as low as 0.25-0.50. For crudes and crude blends where the sulphur concentration is low relative to the naphthenic acid concentration, formation of protective metal sulphides and metal oxides is more difficult. A ratio (S/TAN) of total S (wt%) to TAN (mg KOH/g) of less than about 1.5 is considered low, and this type of crude oil is named low sulphur/TAN ratio crude (LSTRC). When processing LSTRC, naphthenic acid may predominate even with relatively low TAN (<0.5) values, resulting in much higher corrosion rates than predicted by current corrosion prediction models.

It is generally recognised that API RP 581 is conservative in that it could overestimate the corrosion. Shell supplements understandings with its own high temperature corrosion model, SuLTAN, to predict corrosion rates for various pieces of piping and equipment in a crude distillation unit from sulphur and TAN in crude oil and side cuts. SuLTAN and API RP 581 are applicable for LSTRC, but not for LSTRC in unwetted areas like furnaces operating in mist flow, nor at the transition from annular to mist flow, and also not in high velocity areas where the protective sulphur layer is damaged.

It is recognised that naphthenic acid corrosion can be very aggressive for carbon steel or low alloy heater coils operating under mist flow conditions, or with superficial velocities greater than 150 ft/s where the sulphur to TAN ratio is less than 1.5. Corrosion prediction models (and API RP 581) are not applicable for these conditions because the FeS layer is not stable. The mist flow regime results in higher tube metal temperatures, locally more concentrated liquid phases, and intermittent wet/dry conditions (compared to annular flow) that can increase sulphidic and naphthenic acid corrosion rates.

In general, the mitigation strategies to the high temperature corrosion in the heater tubes and transfer lines of CDU and VDU could consider the following:

1. S/TAN ratio should be predicted for planned operations. Consider adapting schedule and operations (crude blend, heater intake rates) to avoid processing of low to very low sulphur crudes that contain low to

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moderate TAN values. The variation in S/TAN of crude and side cuts at a crude distillation unit can be monitored satisfactorily via offline analyses, depending on the frequency of changes in the crude slate that lead to swings in the level of TAN in the crude and side cuts.

2. If low S/TAN ratio cannot be avoided, heater modelling should be conducted to determine the operating window to avoid mist flow and high velocities in the atmospheric and vacuum heaters and transfer lines. By detailed modelling, inspection and alloy upgrade can be targeted for the specific length of the heater tube that can present mist flow conditions or high velocities, reducing downtime and capital costs of material upgrade.

3. If operating close to erosional velocity for CS or 5 Cr-0.5Mo tubes, it is advised to increase the inspection regime and inspect the furnace tubing by intelligent pigging.

4. If the above conditions cannot be met, establish a means to mitigate the issue by changing operation or alloy upgrades:

a. For low S/TAN ratio and low/medium TAN crude, 9Cr-1Mo is the recommended minimum alloy for heater tubes.

b. For high TAN crude, austenitic stainless steels (such as 316L, 316Ti, 317L, and so on) are recommended. 304 SS is not recommended due to low Mo content and therefore low corrosion resistance

c. For very high TAN crude, higher Mo-containing austenitic alloys should be considered (such as Alloy 904L, AL6XN, or 625)

In addition, to bridge the time period to the next turnaround, the use of high temperature corrosion inhibitors could be considered. These inhibitors are only effective for wetted areas and therefore not proven effective in furnaces and transfer lines. They have limitations and necessary requirements (one-phase flow, inspection coverage, injection facilities, effects on products and downstream units, and so on). So the long term solution is usually the use of appropriate and reliable metallurgy, especially for components subjected to high velocities inducing erosion or erosion-corrosion.

**Q** Can you suggest ways to reduce CO<sub>2</sub> emissions from our FCC regenerator?

**A** Chris Claesen, Director, Technical Consulting, NALCO Water, cclaesen@ecolab.com

FCC units are major contributors to refinery CO<sub>2</sub> emissions due to burning of the coke that is generated on the catalyst during the catalytic cracking process. If the refinery has an old FCC unit, it may get the most reduction in CO<sub>2</sub> production by revamping to a modern design with the focus on CO<sub>2</sub> minimisation. If the refinery is processing heavy residual feedstocks that contain contaminants such as Ni, V, Fe, Ca, Na and salts, these can to some degree contribute to the coke generated on the catalyst; this is called contaminant coke. To maintain the unit's heat balance, the operation will typically be adjusted and total coke make and CO<sub>2</sub> production may remain the same, but thermal cracking will increase

and the product mix that is produced may become less desirable. The most obvious is typically the increase in hydrogen content in the tail gas, but gasoline production may decline and LCO and slurry production increase. Looking from a CO<sub>2</sub> perspective, more CO<sub>2</sub> is produced for the same amount of desirable product.

Chemical programmes can help increase the FCC unit's efficiency and reduce the effects of contaminants in several ways. For example, the removal of Fe, Ca and salts in the desalters can be optimised by using the most suited demulsifier and solids removal programmes. Good desalting will also reduce or stop the need for NaOH injection downstream the desalters and reduce the Na content of the residual FCC unit feed. The effect of Ni and V and contaminant coke make can be reduced at the FCC unit by applying metal passivation programmes such as Nalco Passivation Plus and MVP. Doing so, these programmes can reduce the CO<sub>2</sub> produced per amount of desirable product.

**A** Alvin Chen, Global Technology Application Manager, BASF, alvin.chen@basf.com and Mark Schmalfeld Global Marketing Manager, BASF, mark.schmalfeld@basf.com

The FCC unit is by design created to be a heat balance unit which uses combustion of catalytic coke to efficiently provide the energy for the cracking reactions. This results in a significant amount of CO<sub>2</sub> generation from the FCC unit. Below are a couple of practical approaches to reduce the CO<sub>2</sub> emissions by a small amount. The approaches are use of a more coke selective catalyst and a second approach is to achieve an improved heat recovery/ integration of energy at the FCC unit:

a. A more coke selective catalyst would allow reducing the CO<sub>2</sub> per barrel of feed and with a catalyst activity optimisation often a higher activity produces improved coke yield vs conversion. This can allow some portion of coke selectivity to be captured with a simple activity optimisation.

b. Increasing the feed preheat will reduce CO<sub>2</sub> emissions if feed preheat is increased through heat recovery (more efficiency on feed/bottoms heat exchangers, increasing temperature direct from the crude unit).

A number of less practical or more challenging economic approaches are possible to consider such as reducing the feed rate to the FCC unit, installation of CO<sub>2</sub> capture system (justification is the challenge), and evaluation of options to utilise CO from units operating in partial combustion (may achieve blower energy savings with a CO boiler, maybe a potential feedstock in future processes). The extreme measures are not likely to be justified approaches today, but might be considered should the future environmental constraints change dramatically.

**Q** We are encountering worrying levels of phenolic contaminants in our wastewater. How can we best deal with this?

**A** Adrian Irimia, Energy and Utilities Engineer, Shell Global Solutions Canada, Adrian.Irimia@shell.com

Phenolics are a group of oxygenated aromatic organics

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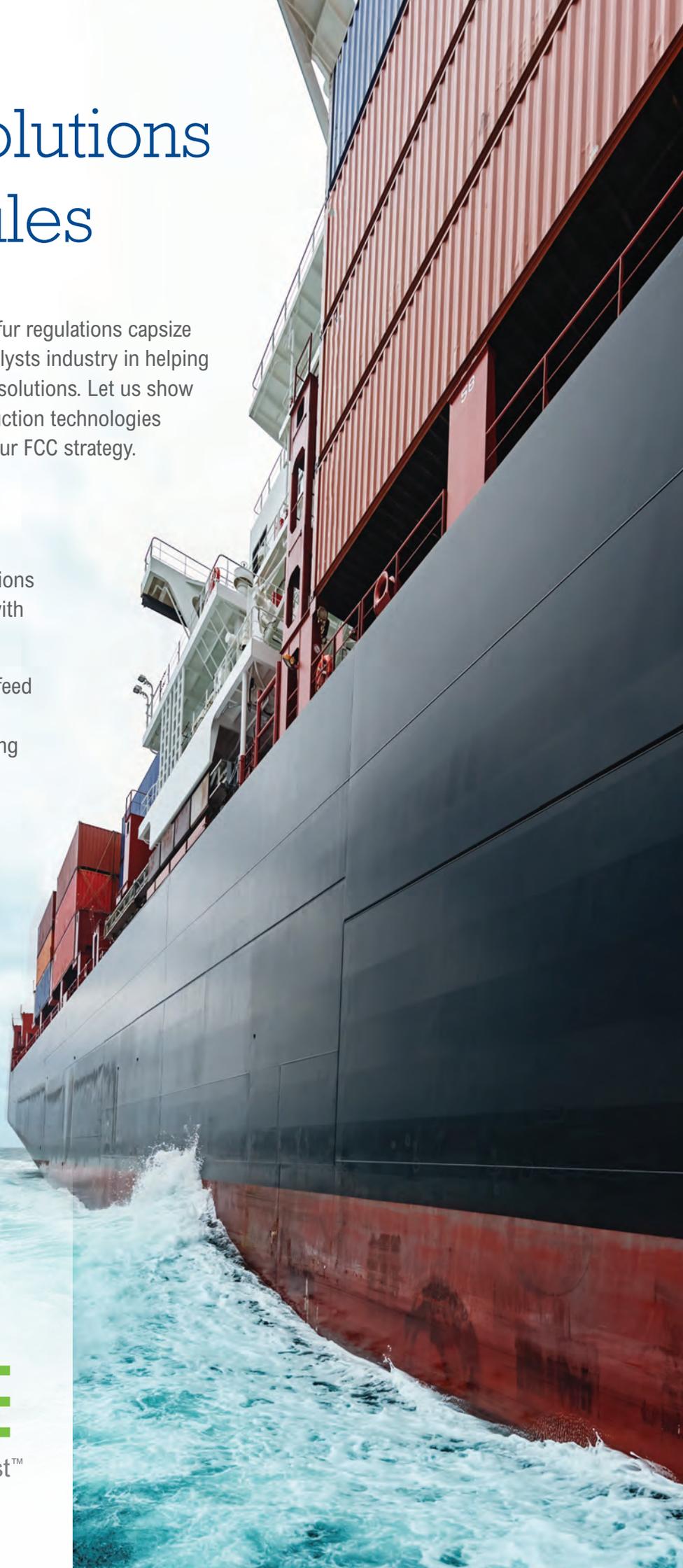
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that result from several crude processing units in refineries and from phenol plants in petrochemical works. The first issue to be clarified is the make-up of the 'phenolic contaminants' and the way in which they are measured. 'Total phenolics' as defined by the EPA Method 420.1 is a colorimetric test that captures not only phenol but also cresols and most derivatives of the phenol molecule, which pose different challenges to treatment. In general, the more complex the molecule, the more it is refractory to biological treatment.

### Sources

In chemical factories, phenol units are principal contributors, for obvious reasons.

In refineries, thermal cracking processes (mainly delayed coking) and catalytic ones (FCC units) are the main contributors of phenols in wastewater. The reaction products from the cat cracker contain steam, and the subsequent main fractionator uses stripping steam. Therefore, the main fractionator overhead reflux drum produces sour water containing phenols. Delayed coking units generate sour water and decoking wastewater, both containing phenols that need removal.

In addition, some refineries still use caustic treatment of cracked gasoline to remove S compounds (mercaptans, thiophenols) and phenolic compounds. Hence, spent caustics are another source of phenol.

Sometimes, depending on the slate of crude, desalter brines can contain phenols, which is mostly from stripped sour water that is reused for desalter wash water make-up.

Another minor source in the refinery is the reprocessed slop oil which carries phenol back to the desalter and phenolic spent caustic stream if treated in a biotreater.

The loads from these two types of industries vary greatly. Phenols (and acetone) concentration in wastewater from phenol units is orders of magnitude higher than from refineries, to the point that it is economical to recover these chemicals via distillation (after pH depression) in a dedicated stripper. These units are called dephenolation units and normally produce a dephenolated water at 100 mg/l. Wastewater from coking and FCC units (stripper) is already at about 100 mg/l or less.

### Source control

The best way to tackle phenolics (as any other contaminant) is at the sources: FCC/coker stripper, spent caustic tanks, desalter brine. This is because, for most of the end of pipe treatment options, high sulphide and ammonia concentrations will result in higher operational costs as they interfere with phenol removal.

A desalter can be a unit to remove a portion of phenol by lowering pH so phenol is preferentially coming into the crude phase. Desalter brines must have good solids and oil separation equipment, such that the final wastewater units are not overloaded and biological processes that are responsible for phenols degradation are not impeded. Shell has also developed a detoxification process for desalter brines aimed at naphthenic acids

removal via pH depression and adsorption in sponge oil. This process can also remove phenols, therefore can be used as a pretreatment step.

Phenolic caustics should be either shipped off site for disposal or phenols recovery by specialised companies or dosed into the refinery effluent treatment plant (ETP) under strict control.

### End-of-pipe mitigation

There are numerous physical and chemical processes that have been applied for phenols removal, from more conventional ones such as distillation, absorption, extraction, chemical oxidation, and electrochemical oxidation to more advanced treatments such as Fenton processes, ozonation, wet air oxidation, and photochemical treatment. Compared to physico-chemical treatment, biological treatment is environmentally friendly and energy saving, but it cannot treat high concentration pollutants. All the phenolics are biodegradable, and a well operated bio-treater should achieve a discharge total phenolics concentration of 5-15 wtppb or lower.

In conclusion, the answer to the question is highly dependent on the existing treatment processes the asset already employs. In Shell's experience, the most cost-effective final treatment is an activated sludge facility, where phenolic water is treated along with all the other wastewater streams generated in the refinery. If the phenols load in the ETP feed is escalating beyond the capacity of the bioreactor, the first step is to run a quick survey at sources, followed by the optimisation/debottlenecking of the local treatment unit (stripper, desalter detoxification), and finally a review of the processing capacity of the ETP. This last check might end up with a recommendation for debottlenecking or for adding of a tertiary treatment (granulated activated carbon), which can polish the ETP effluent to compliance specifications.

**A** Chris Claesen, Director, Technical Consulting, NALCO Water, [cclaesen@ecolab.com](mailto:cclaesen@ecolab.com)

Best to start with finding the root cause of the high phenol levels. Phenols typically come from FCC sour waters; a change in FCC feed quality or operation can have an impact on the sour water phenol content. Also the way the FCC sour waters are routed can have a significant impact on the phenol content in the feed to the effluent treatment plant. For example, desalters are great at removing phenols from sour water; letting phenolic sour water (partially) bypass a desalter or switching from a double to a single desalter operation can have a significant impact. You need to have a good understanding of how the phenolic sour waters are routed and how this impacts the feed to the effluent treatment plant.

While the bacteria in the effluent treatment plant should be capable of removing the dissolved phenols, the treatment plant layout may not enable removal to the low levels required for final effluent discharge. Revamping the effluent treatment plant to enable better biological removal may not prove practical. One solu-



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tion is to remove the root cause of the increased phenol levels, but if this is not possible there are, in general, two possible solutions to manage the contaminant levels at the wastewater plant: (partial) detoxification of the phenolic water source before effluent treatment or final effluent polishing. Chemical applications such as Purate have been successfully used to oxidise toxic contaminants in wastewaters. The chlorine dioxide reacts quickly and preferentially with phenols, showing good removal efficiency in high phenol containing waters like phenolic spent caustic and sour water stripper water. In order to lower the load to the effluent treatment plant, complete removal is typically not required. Partial oxidation of phenols and sulphides will significantly reduce the chemical oxygen demand and enable sufficient removal to lower the final effluent quality to permissible levels.

Depending on the plant set-up, final effluent phenol polishing may be preferred. Recent examples have shown that Purate applied to a large water stream containing only a few mg/l will remove it down to the required ppb level. The selective oxidation properties make it a better choice than, say, bleach which will not only have a high consumption due to side reactions with ammonia and organics, but tends to generate halogenated contaminants as well. On-site generation of chlorine dioxide also tends to be more flexible and much less capital and operationally intensive compared to wet air oxidation.

**Q** Fouling in the feed/effluent exchangers to our naphtha hydrotreater is at unacceptable levels. The feed is straight run naphtha. Solutions please.

**A** Chris Wajciechowski, Business Development Manager, Alfa Laval Energy Division, [chris.wajciechowski@alfalaval.com](mailto:chris.wajciechowski@alfalaval.com)

Fouling on the feed or effluent side of a naphtha hydrotreater combined feed/effluent (CFE) heat exchanger can occur for a multitude of reasons. It is most important to first understand the root cause of fouling and then select the proper mitigating measure. Most root causes are well documented and remedies are readily available. For example, oxygen or olefins in feed can cause gums to form as the feed is heated; and the remedy is to avoid oxygen ingress or install an oxygen stripper to remove the contaminants. Other contaminants like chlorides can create effluent salts or corrosion products that foul or plug heat exchanger surfaces. Water washing to remove effluent salts or upgrading metallurgy to stainless steel to avoid high corrosion rates are acceptable ways to mitigate the effects of chloride contaminants.

Alfa Laval has installed a number of Compabloc welded plate heat exchangers in naphtha CFE service and the performance has been impressive. In one particular straight run NHT unit, unusual feed fouling was detected which increased the pressure drop and gradually reduced thermal performance over time. Analysis of the fouling confirmed it was present only on the feed side and at the point where 100% vaporisation

occurred. A high-boiling component was being left behind when the naphtha flashed, and this component deposited on the plates. Gums were ruled out since the refinery was processing a low percentage of cracked feed, and feed composition or contaminant issues were ruled out as well. The root cause was an uncommon corrosion inhibitor additive that was being overdosed in the overhead system in the crude unit. This additive became the high boiling component of the NHT feed and coated the heat exchanger surface area at the dry point. Optimisation of the dosing rate reduced the fouling to an acceptable level to reach the typical unit cycle length.

**A** Chris Claesen, Director, Technical Consulting, NALCO Water, [cclaesen@ecolab.com](mailto:cclaesen@ecolab.com)

The first thing to find out is if the fouling is on the feed or the effluent side.

If possible a sample of the fouling material should be collected and analysed. The composition will help to understand what measures need to be taken to control the fouling. Fouling at the effluent side will typically be due to  $\text{NH}_4\text{Cl}$  formation and this can be controlled with a waterwash injection upstream of the fouled exchanger. Proper injection practices and water quality will have to be used to avoid corrosion problems. If fouling is at the feed side, it will need further investigation.

What does the deposit analysis tell you? Is the feed pure straight run or are other streams blended in? Is intermediate storage used? Is there significant corrosion taking place in the CDU overhead? Has something changed? Were new exchangers installed? The possible solutions will depend on answers to these questions but if things are not clear a detailed survey may be needed. In some cases, a chemical treatment based on an organics or inorganics dispersant such as Nalco Long Run can be the solution, in some other cases the fouling can be solved by changing upstream operations or corrosion control.

**A** Berthold Otzisk, Kurita Europe, [Berthold.otzisk@kurita.eu](mailto:Berthold.otzisk@kurita.eu)  
Straight-run (saturated) naphtha comes directly from the crude unit or vacuum unit. In general, cracked (unsaturated) feedstocks create more fouling problems than straight-run naphtha. It is important to define the fouling problem. What is the root cause of the fouling?

In the presence of oxygen, a polymerisation process can be initiated, creating a chain reaction resulting in hydrocarbons of high molecular weight gums. Naphtha storage tanks with floating roofs still allow oxygen contamination. A fixed-roof tank with proper gas blanketing is recommended and oxygen should be removed via a stripping process.

There is a direct correlation between oxygen and fouling severity, where polymers can be metal-catalysed or acid-catalysed. Please check the metals and oxygen concentration and acidity of the naphtha feedstock. Polymerisation reactions may begin at ambient temperature and be easily catalysed at higher temperatures (100-300°C).

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**A** **Xiomara Price, Senior Product Analytics/Support Manager, SUEZ – Water Technologies & Solutions, xiomara.price@suez.com**

Fouling in a straight-run naphtha hydrotreater can be caused by fouling precursors such as carboxylate salts, filterable solids, corrosion byproducts, amine salts or by other unwanted contaminants that can make it into the charge feed (leaking material into the feed). It is important to determine the root cause of the problem to be able to implement the right solution since mitigating the root cause should be the first solution (bypassing leaking exchangers, reducing corrosion rates, and so on). If eliminating the root cause is not plausible, chemical treatment is a viable option. Chemical suppliers should be equipped to provide a plan for root cause analysis and provide guidance on the best solution based on the results for your specific unit.

**Q** **We are processing tight oil and having issues with our diesel's cold flow properties. Would a catalytic dewaxing option help?**

**A** **Dean Parker, Technical Sales Manager – Downstream Catalysts, ExxonMobil Catalysts and Licensing, dean.e.parker@exxonmobil.com; Jonnie Verwoert, Hydrotreating Specialist, Albemarle Catalysts Company, jonnie.verwoert@albemarle.com; and Barbara Slettenhaar, Global Business Manager, Albemarle Catalysts Company, Barbara.Slettenhaar@albemarle.com**

Catalytic dewaxing of tight oil diesel could be considered an attractive option to achieve the desired cold flow properties.

Tight oil consists of liquid hydrocarbons that are obtained by hydraulic fracturing of shale formations, while shale oil is obtained by cooking the kerogen/stone in shale formations. Depending on the source, the tight oil (and shale oil) can be more paraffinic or more aromatic in nature.

For the dewaxing catalyst to work effectively, the nitrogen content needs to be largely reduced before the oil enters the dewaxing catalyst. Typically, the paraffinic tight oil requires extensive dewaxing and has a low nitrogen content, thus requiring less pretreat catalyst. For a more aromatic tight oil (or shale oil), the nitrogen content can be high, requiring extensive pretreatment and less dewaxing catalyst.

Depending on the tight oil properties and the available hydrotreating capacity, the Galexia platform of ExxonMobil and Albemarle offers an innovative and holistic solution.

This solution likely involves:

- ExxonMobil's MIDW iso-dewaxing catalysts, delivering target cold flow properties with minimum conversion to naphtha
- Albemarle's pretreat catalysts portfolio and/or application of ultra-high activity hydrotreating catalyst Celestia

Celestia is a joint development of Albemarle and ExxonMobil.

**A** **Fie Alice Hallkvist Wilbek, Senior Hydroprocessing Specialist, Haldor Topsoe, fahw@topsoe.com**

The short answer is yes, catalytic dewaxing will improve cold flow properties. Due to the very paraffinic nature of the tight oil, it is common to have issues with diesel cold flow properties, and many refineries are choosing catalytic dewaxing to solve this problem. When choosing catalytic dewaxing, you need to be aware of the different solutions available.

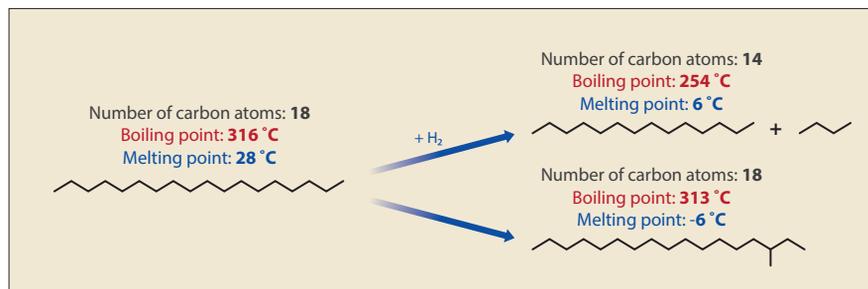
Catalytic dewaxing can be achieved by modifying the hydrocarbons in the feed by two reactions: either cracking or isomerisation, or a combination. The two reaction pathways are illustrated for C<sub>18</sub> n-paraffin in **Figure 1**.

Conventional (also referred to as traditional) dewaxing catalysts improve the cold flow properties of a feedstock by shortening the chain lengths of the paraffins – either by pure cracking of the n-paraffins or by a combination of cracking and isomerisation of both n- and iso-paraffins. Such catalysts usually have a high activity, and some can also contribute to lowering dist T90/T95 percentages ('back-end shift').

However, the big disadvantage of conventional DW catalysts is that their cracking functionality results in some of the hydrocarbons ending up as fractions with lower boiling points. This affects the diesel yield and can have big impacts on a refiner's revenue-earning potential.

Catalytic dewaxing by isomerisation, on the other hand, makes sure the number of carbon atoms in the hydrocarbon chain remains the same during the reaction processes inside the hydrotreating unit. As a result, there is hardly any loss of yield to lower boiling fractions. This significantly improves yields of higher value diesel with on-spec cold flow properties that make them commercially attractive.

Traditionally, the isomerisation dewaxing catalysts



**Figure 1** Reaction pathways in catalytic dewaxing

have been associated with very high costs, as these were promoted by noble metals, most often platinum. However, today isomerisation dewaxing catalysts are also available in a base metal versions, which have gained significant market share over the past three years due to the lower installations cost and higher tolerance towards sulphur and nitrogen in the feedstock.

**A** Amit Kelkar, Diesel Hydrotreating Application Specialist, Shell Catalysts & Technologies, Amit.Kelkar@shell.com

Oil extracted from low permeability rock using technologies such as hydraulic fracturing is referred to as tight oil. Compared to conventional crude oil, it is generally less dense and highly paraffinic with low sulphur content. Its paraffinic nature adversely impacts cold flow properties including cloud and pour point.

The linear, long-chain paraffins prevalent in tight oil are often referred to as 'wax'. Catalyst dewaxing is the removal of wax by conversion to isomerised (same molecular weight, different structure and properties) and/or cracked molecules (lower molecular weight). When processing highly paraffinic feeds, dewaxing may be needed to maintain the flow properties of diesel and meet the target winter cold flow specification. Economic benefits of dewaxing include the ability to upgrade heavier or renewable feeds which have higher cloud and pour points without use of cold flow additives (cost) or kero blending (high value in many places).

Shell Catalysts & Technologies has collaborated with several refiners to add dewax capability to existing hydrotreaters. The main items to be considered when applying dewax catalysts are:

- **HSSSE:** a thorough safety evaluation should be done to ensure adequate safety systems are in place for use of dewax catalysts
- **Delta cloud point target:** use of additives or kero blending can be economically favourable if target delta cloud point is low, i.e., only trim dewaxing is needed. A detailed analysis is needed to evaluate economic feasibility of catalytic dewaxing
- **First stage versus second stage dewax:** In first stage dewax, the dewax catalyst is in series with the hydrotreating catalyst and operates in a sour environment. second stage dewax is done in a separate reactor with inter-reactor separation. For revamp applications, first stage dewax has lower capex, while second stage dewax provides improved diesel yield at higher capex
- **Catalyst design:** replacing a portion of the hydrotreating catalyst with dewax catalyst will lead to higher LHSV and thus increased severity. A customised hydrotreating catalyst system is essential to achieve target HDS cycle life and ensure optimum feed quality to the dewax bed
- **Operability:** adequate heat input and quench gas capability is needed to ensure that target temperature profile is achievable in winter and summer months. Back end work-up section should be evaluated for increased naphtha yield



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**A** Ronald Gropp, Fuel Additives Center of Excellence Leader, SUEZ – Water Technologies & Solutions, ronald.gropp@suez.com

Some tight oils have high percentages of normal paraffins, and the normal paraffins range from  $\sim C_{10}$  to  $C_{28}$  in the diesel distillation range ( $\sim 390^\circ\text{F}$  to  $\sim 690^\circ\text{F}$ ). High concentrations of normal paraffins, and especially high concentrations of high C number normal paraffins, can negatively impact the low temperature handling properties of the diesel including cloud point, cold filter plugging point, and pour point. Catalytic dewaxing is a hydrotreating process that incorporates a catalyst to convert normal paraffins to isoparaffins that have much better low temperature flow properties. As a result, catalytic dewaxing may in fact be a good option if the economics fit the need. Other potential options include:

1. Lowering diesel distillation 90% and end point
2. Utilise cold flow improver additives (wax crystal modifiers) including cold filter plugging improvers and/or pour point depressants
3. Increasing the percentage of aromatic blend components like light cycle oil to improve responsiveness of the fuel to cold flow improvers
4. Combinations of options one, two and/or three

**Q** Which crude impurities stabilise oil-water emulsions and can we remove them before our desalter?

**A** Berthold Otzisk, Kurita Europe, Berthold.otzisk@kurita.eu

Crude oils are compositions with oleophobic and oleophilic components and vary from region to region. Crude oils contain many impurities, which should be removed to avoid fouling and corrosion in downstream processes. Typical oleophobic components are sulphates, sodium, calcium and magnesium salts, drilling muds, sand, soluble water, and emulsified water. Oleophilic components are organometallic compounds like iron, nickel, sulphur, nitrogen, naphthenic acids, waxes, and asphaltenes. These impurities can stabilise the formation of emulsions.

Mechanical methods such as settling, centrifuging or filtering can remove corrosion debris and solid particulates with 20-200  $\mu\text{m}$  particle size. Suspended solids  $<20 \mu\text{m}$  cannot be removed by centrifuging or standard filtration technologies.

When heavy crude oils with higher percentages of asphaltenes and paraffins or opportunity crudes with naphthenic acids are used, crude pretreatment is advisable. Chemical programmes can help to remove impurities, lowering the risk of stable emulsions. Complexing agents will bind calcium from naphthenates, which can be removed with the water phase at the tank farm later. Emulsion breakers (demulsifiers) should be applied at the beginning of the pipeline unloading line or tanker discharge before the crude oil is routed to the crude oil storage tanks. Proper mixing and coalescing of the crude oil, good residence time for settling, and removal of water are essential for a better treatment.

Electrical and thermal methods significantly support the emulsion breaking process, but cannot be realised

at the tank farm. For that reason, an electrical desalting step at the crude distillation unit is still required for the removal of salts and water, where a suitable emulsion breaker programme is applied into the crude oil before mixing valve.

**A** Chris Claesen, Director, Technical Consulting, NALCO Water, cclaesen@ecolab.com

Typical emulsion stabilising impurities can be:

1. Asphaltenes: asphaltenes have polynuclear aromatic cores and alkyl side chains that form inverse micelles (aggregates). Agglomerates can concentrate at an oil/water interface, forming an elastic film that can hinder the affinity of water droplets to coalesce. This parameter can be controlled via effective crude oil blending practices and through chemical treatment with an asphaltene dispersant.
2. Solids: fine particulates, such as inorganic iron and zinc complexes, sand, and clay, can become interfacially active via surface modification following adsorption on polar species contained in crude oil. The hydrophilicity of the particulate will directly affect the degree of emulsion stabilisation. This can be controlled with desalter acidification, solids removal agents, and effective crude oil blending and tank settling practices.
3. Amines: high amine content resulting from upstream treatment can raise the pH of the entrained water, retarding the rate of water droplet coalescence. Acidification can be used to manage high pH water.
4. Bacterial cells have been reported to stabilise oil-water emulsions. When present these biosurfactants can build up at an interface and behave in a manner similar to fine solid particulate.
5. Highly conductive crude: crude feeds containing high metals or naphthenate complexes may lead to more conductive crude oil blends, resulting in increased amps and a corresponding reduction in the voltage gradient within the desalters. Preemptive knowledge of the metals loading in a crude oil slate or proposed crude blend may help schedulers evaluate the risk of processing high metal containing blends, minimising emulsion stability concerns.
6. Upstream chemical additives such as surfactants, gelants, and  $\text{H}_2\text{S}$  scavengers can add surface active materials that exacerbate emulsion stability.
7. As refiners continue to adapt to IMO 2020 regulations, additional opportunity feeds will become more prevalent including re-run feedstocks that may contain elevated levels of coke fines and heavy metals. In addition, these opportunity feedstocks can cause increased blend stability concerns.
8. Many species in crude oil are acidic and can ionise at an oil-water interface to create a surface active moiety. This class includes high molecular weight carboxylic acids and naphthenic acids and soaps thereof.

**Mitigation of the above:**

1. Leverage assays and processing experience of different hydrocarbon feedstocks to provide insight into crude quality's impacts on refinery operations and emulsion stabilisation. The characteristics to monitor



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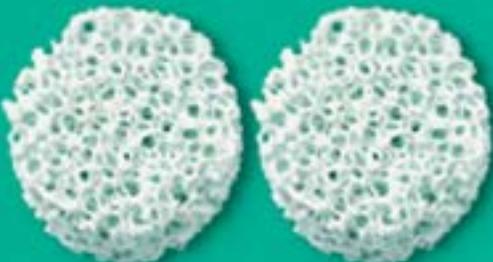
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include, but are not limited to, asphaltene stability, total acid number, naphthenic acid concentration, metals concentration, and solids loading

2. Treatment with an asphaltene dispersant to mitigate asphaltene instability

3. Treatment with a solids removal agent

4. Acidification can be employed for charge neutralisation and pH control

5. Addition of an emulsion breaker upstream of the desalters can precondition the crude oil for enhanced emulsion control

**A** Karl Kuklenz, Crude Unit Applications Specialist, SUEZ – Water Technologies & Solutions, karl.kuklenz@suez.com

There are many chemical species which stabilise crude oil and water emulsions, but the classification of which are impurities is sometimes an issue for debate. There are some who classify crude oil impurities simply by undesirable presence, whatever their natural occurrence: “If it’s bad, I don’t want it; it’s an impurity.” Others classify an impurity as an artificial addition, accidental or deliberate, that can be detected by its effect on its surroundings. In this answer, we will talk about the main stabilisers of crude oil/water emulsions and leave it to the reader to decide their label as an impurity. For the purposes of this response when we refer to oil, we intend to mean crude oil or a blend of crude oils.

The first and perhaps most significant stabiliser of crude oil emulsions is the asphaltenes. Asphaltenes are polycyclic hydrocarbon molecules of myriad shapes and sizes with polar functional groups containing oxygen, nitrogen, and sulphur. These functional groups and a network of conjugated double bonds tend to make these structures flat and polar, which gives rise to their surfactancy and interfacial viscosity, and more to the point, emulsion stability. Thus, it is not any one molecular structure that is responsible for the stabilisation, but all of them to one degree or another.

Generally, asphaltenes are not extracted from crude oil. In a few oil sand mining operations, paraffinic froth treatment is employed to precipitate asphaltenes and produce a high quality bitumen. And solvent deasphalting units are used to remove asphaltenic material from intermediate refined streams in preparation for further conversion. However, with the widely available hydrocracking, coking, and visbreaking processes to convert heavy hydrocarbons like asphaltenes into valuable products, extraction of asphaltenes from crude oil on a wide scale, while technically feasible, is economically unlikely. Asphaltenes in an unstable matrix or that are destined to become part of an unstable mixture (like when blended with light tight oils) can be treated with an asphaltene stabiliser upstream of the desalter to limit their precipitation and thus their interfacial activity which has the effect of improving desalter operations.

Metal naphthenates, especially calcium naphthenates, stabilise crude oil emulsions much like the asphaltenes. They are a class of crude oil surfactants with significant interfacial viscosity. And like the asphaltenes, they are

not typically removed from crude oil prior to desalting. They can be treated at the desalter with the judicious application of an acid programme to extract the metal from the naphthenates to reduce their polarity and lessen their affinity for the interface.

Solids are probably the best example of an undesirable species in crude oil. But, while some are naturally occurring, they provide no value to the refiner. And worse still, they tend to cause issues wherever they go. Solids come from many sources, some are produced from the geology of the well formation, and some are the result of corrosion in the production and transportation assets, like iron oxides and iron sulphides. In severe cases, the precipitation of asphaltenes can become a source of solids as well. Both organic and mineral solids are weakly interfacially active and, depending on their size, they can seek out interfaces like the oil-water interface of an emulsion. And they themselves are a solid-liquid interface, which attract surfactants like the asphaltenes. This asphaltene and particle matrix creates a rigid water droplet interface which is a significant barrier to droplet coalescence and therefore to emulsion resolution.

Fortunately, solids can often be dealt with in a variety of ways to reduce their impact on the desalter. One way to remove solids mechanically prior to the desalter would be with a process centrifuge. Centrifuges are sometimes utilised to remove solids from a slop-recycle stream before it is blended back into the crude charge. Process centrifuges can be augmented with various chemical treatments to further optimise their solids handling ability. Solids can be physically removed in tankage by lowering the viscosity with a dilution stock such as a light cycle oil or one of the less viscous crude oils in the blend. The solids settle more rapidly and are retained in the tank or removed as part of the dewatering process. Finally, solids can be removed at the desalter with appropriate chemical pretreatment. This chemical strategy de-oils (or water-wets) the solids to facilitate their transport into the water phase and leave the desalter with the effluent brine.

The last category of emulsion stabilisers are the additive chemical or polymer variety. These species are perhaps the best definition of an impurity in that they are not part of the native valuation of the crude oil commodity; they are added deliberately at or after production and their effect can most certainly be measured at the desalter. There are indeed many candidates, but two chemical families come to mind immediately: the drag reducing agents (DRA), and ‘water treatment polymers’. DRAs are additives that are added to oil pipelines to reduce turbulent flow and the associated energy losses over the pipeline network. Depending on their structure, they can inhibit emulsion resolution due to their high molecular weight and viscosity. Due to their high solubility and functional invisibility in crude oil (most DRAs are made only of hydrocarbons), there is not currently a method known to the author by which DRAs are effectively removed from crude oil prior to the desalter.

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Water treatment polymers are sometimes added in production to treat the produced water as it is co-produced with the oil from the well. Some desalters are treated directly with water treatment polymers to improve the quality of the effluent brine and overall emulsion resolution. When used appropriately for these purposes, the risk to emulsion resolution at the desalter is manageable and can even be a boon. However, the therapeutic dosage has a relatively narrow range and the consequences of over-treatment are severe emulsion stability. Water polymers of various functional species generally work in part because of their large molecular weight. In excess concentration, this turns the water in an emulsion to a viscous gel. Unfortunately, this stabiliser, once introduced at these levels, is very difficult to remediate. The best strategy is one of caution to avoid their overtreatment by following proper make-down procedures and not exceeding recommended dosage rates.

Weakly associated with this family of chemical emulsion stabilisers is anything that causes a deviation from optimal pH. Disrupting the optimal pH changes the solubilities and polarities of charged surfactants, including asphaltenes and the solids and can lead to poor emulsion resolution. These acidic or basic species can be naturally occurring or can be the result of contamination. An example of an acidic contaminant would be production flowback from a recently acidised well formation or acid treatments to inhibit metal naphthenate deposition in production. Examples of a basic contamination would be tramp amines from overhead neutraliser treatments, and triazines that come back to the desalter in the wash water or a misapplication of caustic in the crude feed. Often applied to the desalted crude, caustic can be mistakenly fed in concentration, amount, or location to end up back in the raw crude, or it can sometimes arrive in a purchased resid feed. In either case, acid or base cannot be removed as such but can be neutralised at or prior to the desalter with an appropriate pH adjustment programme.

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Crude oil received by refineries is comprised of the following main groups: hydrocarbons (PIANO and asphaltenes), hetero organic compounds (sulphur, nitrogen, oxygen, vanadium, nickel), basic sediment and water (BS&W) with its main components being clay, sand, inorganic scale, formation water from the production field, transportation ballast water, and soluble salts ( $\text{NaCl}$ ,  $\text{CaCl}_2$ ,  $\text{MgCl}_2$ , and so on).

In addition to the above components, which are innate to the reservoir (naturally occurring), synthetic chemicals-oil field chemicals (OFC) are dosed during crude production and transportation for various purposes.

Naturally occurring contaminants that contribute to oil-water emulsions are: asphaltenes/resins, water, solids, and naphthenates, and salts.

Asphaltenes and resin molecules contain heteroatoms, such as nitrogen, oxygen, and sulphur. This imparts acidic and basic traits to petroleum based fluids, and thus stabilises emulsions. Wax and asphaltene inhibitors (dispersants such as Shell Swim's) are used during the production of oil. They are used specifically to prevent asphaltene/resins/wax depositions in the production equipment. More is added to prevent asphaltene precipitation during transportation. Therefore, asphaltene/resins are substances that cannot be removed prior to desalting.

Water is produced together with the oil. An efficient way to minimise oil-water emulsions is to minimise the water reaching refineries; however, reduced water content does not necessarily mean reduced oil/water emulsion. Some high water content in crude may drop out easily in a crude storage tank. It really depends on whether the oil/water is already emulsified. The separation of water and oil at the production sites can be enhanced by heat and adding demulsifiers (for instance, ethylene oxide, propylene oxide based) on a continuous basis. The drawback of this method is the



relatively high dosage of demulsifiers that is needed to promote oil/water separation. Water separation is carried out again in refineries prior to desalting (in tanks). Shell survey results show that a tank residence time of at least one day has a positive impact on desalting operation.

A variety of different types of solid can be found in crude oil: sand, silt, clay, silicates, corrosion byproducts ( $Fe_2O_3$ , FeS), scaling salts from connate water (carbonates, sulphates), and asphaltenes (due to crude oil blend incompatibility). Solids are classified as basic sediment or filterable solids, based on the size of the solids. Basic sediment refers to solids that are 20-200 microns, while filterable solids defines solids that are less than 20 microns. Solids concentration is cargo specific as the solids load is dependent on the way the oil is handled. Therefore, Shell best practice is to measure the BS&W content of crude cargoes, to give warning of problematic shipping and suppliers. Some basic sediment will settle out and accumulate at the bottom of the storage tanks and can be removed either with the water or when the tank's cleaning is carried out.

Filterable solids cannot be removed before the desalter, and sometimes they are difficult to be rejected in the desalter as well due to mixed wettability. Filterable solids are hard to remove, the typical removal efficiency is 50-70%. The solids are trapped in rag layers, but we can remove the rag layer by withdrawing it so the solids are removed.

The term 'naphthenic acid' as commonly used in the crude oil industry refers collectively to all the organic acids present in crude oil. It is defined as total acid number (TAN) and it is expressed numerically as milligrams of KOH required to neutralise the acidity. Calcium and other metals can form soaps with naphthenic acid. These soaps will accumulate at the interface between the water and oil layers in desalters. Calcium and sodium naphthenate type soaps can promote the formation of stable emulsions either on their own or through interactions with other substances in the crude, such as asphaltenes and inorganic solids. The most common treatment for emulsions formed by calcium naphthenates in production facilities is to inject acetic acid. Reducing a system's pH converts the calcium naphthenate soap back to its naphthenic acid form and releases the calcium ion to the aqueous phase where it can be removed by the dewatering.

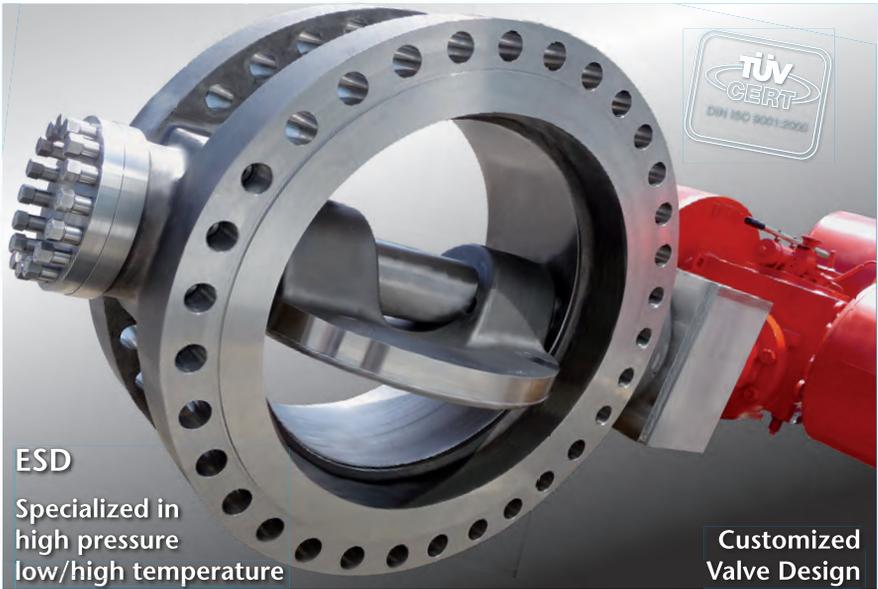
As above, cations from inorganic salts can promote the formation of soap like emulsions. Most of the salts present in the crude are in a dissolved form in water and therefore can be removed before desalting by proper tankage dewatering. Some inorganic solids are removed in a CDU preheat train prior to the desalter as crude is heating up. However, sometimes crystalline salts might be present in crude oils that have undergone processing that could lead to evaporation of the connate brine. The salt crystal may become oil coated and cannot be removed before desalting.



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Most of the OFC, particularly from fields using EOP techniques, dehydration surfactants, and so on promote emulsion formation in desalters. Parsimonious dosage in the production field coupled with occasional pH reduction for salts break up can help reduce the negative impact of these exogen chemicals on desalters' performance.

Some vendors are offering 'pretreatment programmes' that rely on chemical injection as crude enters the refinery (before tankage). The claim is that this treatment breaks residual emulsion in the incoming crude and 'preconditions' solids in the crude for quick settling. Recommended dosages are 12-24 ppm for heavy crudes and 5-10 ppm for light ones. The benefits include a reduction in raw crude BS&W, reduction of filterable solids and water settling time in tankage. All these claims must be verified by rigorous laboratory testing followed by field trials on various crude slates.

**Q** What is sintering in relation to catalyst deactivation and how can it be avoided?

**A** Maxim Ovchinnikov, Senior Research Scientist, Shell Catalysts & Technologies, Maxim.Ovchinnikov@shell.com and Karl Krueger, Senior Research Scientist, Shell Catalysts & Technologies, Karl.Krueger@shell.com

Hydrotreating catalysts during their commercial cycle application will undergo deactivation via three major mechanisms: coke formation and deposition, poisoning of active sites and/or pore blockage caused by several poison elements in the feedstocks, and sintering or agglomeration of active promoted metal sites.

Fresh catalyst is designed to provide the optimal dispersion, size distribution and promotion of MoS<sub>2</sub> sites. The agglomeration of active promoted metal sites during a commercial cycle caused by high temperatures will result in irreversible changes in the morphology of MoS<sub>2</sub> sites. This process will manifest itself in one or more of the following pathways: thermodynamically driven agglomeration of active sites to form larger, less active particles; loss of promotion of MoS<sub>2</sub> sites due to cobalt or nickel sulphide segregation; penetration of active metals to the catalyst support.

The same story is true for noble metal hydrofinishing catalysts. These catalysts rely on the high dispersion of Pt and Pd as very small particles for good activity. As these particles sinter and grow, the exposed metal surface area drops, reducing the number of active sites and therefore catalyst activity. Sintering in these catalysts is significantly accelerated by high temperature as well as exposure to oxygen and water.

In general, sintering cannot be avoided during a commercial cycle of hydrotreating catalysts, but it can be greatly reduced in low severity applications.

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Sintering as an FCC catalyst deactivation phenomenon

refers to the occurrences of liquid phases and surface melts on the surface of the particle that result in the formation of a densified layer, imposing a diffusional constraint for the penetration of large hydrocarbon molecules to the interior.

Typically, this is associated with the presence of certain 'fluxing' contaminant metals, namely Fe, Ca, Na, V, and K, which form eutectic mixtures with the oxides comprising the catalyst, lowering melting temperatures and resulting in vitrification.

Of this set, Fe is the most prominent and consequential; the other metals in the list serve mostly to exacerbate the effects of contaminant Fe (in addition to their attack on zeolite).

This surface vitrification will not be reflected in catalyst pore volume measurements by water saturation, nor the mesopore size distribution derived from the adsorption isotherm using nitrogen as the probe molecule. The interior pore structure is largely unaffected, and the occlusion of the surface pores is not sufficient to prevent entry of water under analysis conditions or N<sub>2</sub> molecules (small compared to heavy hydrocarbons), nor to alter the N<sub>2</sub> adsorption equilibrium. It will, however, almost invariably be accompanied by a characteristic morphology change, in which the catalyst particle will take on some degree of a corrugated, 'nodulated' and/or glazed appearance.

The imposition of these diffusional constraints on the surface of the particle results in catalyst performance degradation, particularly with regard to bottoms upgrade.

There are two general catalyst technology approaches to mitigating problematic deactivation by sintering:

1. A highly accessible catalyst with excess particle surface porosity such that a severe degree of sintering must occur before performance degrades
2. A catalyst whose composition and chemistry are designed to inhibit and minimise deleterious eutectic shifts and sintering reactions such that catalyst accessibility is more robust and degrades more slowly

These two approaches are complementary in nature, and the most Fe-resistant catalysts will feature both.

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Catalyst deactivation mechanisms are often very specific to the catalyst design and the specific application. A number of general deactivation mechanisms can be grouped into the following categories types: (1) sintering (2) catalyst fouling (3) catalyst poisoning (4) catalyst loss of active sites.

Sintering is generally classified as a type of thermal catalyst deactivation which often involves a catalyst on a metal support (precious metals on a metal support). The actual kinetics depend heavily on the catalyst material and the actual environment the catalyst is subjected to during the process. Sintering, with respect to zeolite materials, is most often referring to the thermal conditions resulting in the structural collapse and

transformation of the zeolite into an amorphous material or new structural form. Y-zeolite typically sinters at temperatures in the 1000°C range.

Poisons are impurities appearing in the feed streams which can build up on the catalyst and result in deactivation. In the FCC unit a typical catalyst poison is organic vanadium found in the feed stream. Catalyst fouling in a FCC has been caused by organic iron which can block the pores of the catalyst when the level becomes very high – this often depends on the specific catalyst design and production process.

Sintering of zeolites used in FCC catalyst is prevented through regenerator design which controls the heat transfer and the average temperatures in the regenerate unit to protect the catalyst from excessive temperatures. (Uniformity of the coke burn without creating hot spots is critical without excessive hydrocarbons to burn off.) Additionally, avoiding or minimising the use of flame impingement from torch oil use will reduce the risk of sintering of the FCC catalyst.

FCC catalyst sintering is normally not expected to be the first indication of a problem in the FCC unit. Surface area destruction occurs at temperatures well below sintering temperatures and the catalytic activity would rapidly decline. If you reach sintering temperatures in the order of 1000°C in the FCC, depending on contaminant level there are many other risks to equipment damage to focus on before catalyst sintering is a concern – like losing air grids, cyclones, vessel walls, and so on.

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Sintering can be caused by certain contaminant metals such as V, Fe, and Na. These can affect the structure of a catalyst particle by forming low melting point eutectics that can melt at regenerator temperatures and damage or destroy a catalyst particle. The catalyst activity will be reduced and conversion will drop.

If contaminant metals are confirmed to be the root cause, measures can be taken to either remove them from the feed or to use a special metal passivation additive that prevents the formation of eutectics.

The removal of Fe and Na (in salt) in the CDU desalters can be optimised by selecting the most suitable demulsifier and solids removal programmes. Good desalting will also reduce or stop the need for NaOH injection downstream the desalters and reduce the Na content of the residual FCC unit feed. Vanadium (V) is present in crude oil as a stable oil soluble complex and cannot be removed at the desalters. The effect of V, and V combined with Na, can be reduced at the FCC unit by applying a vanadium passivation programme such as Nalco MVP.



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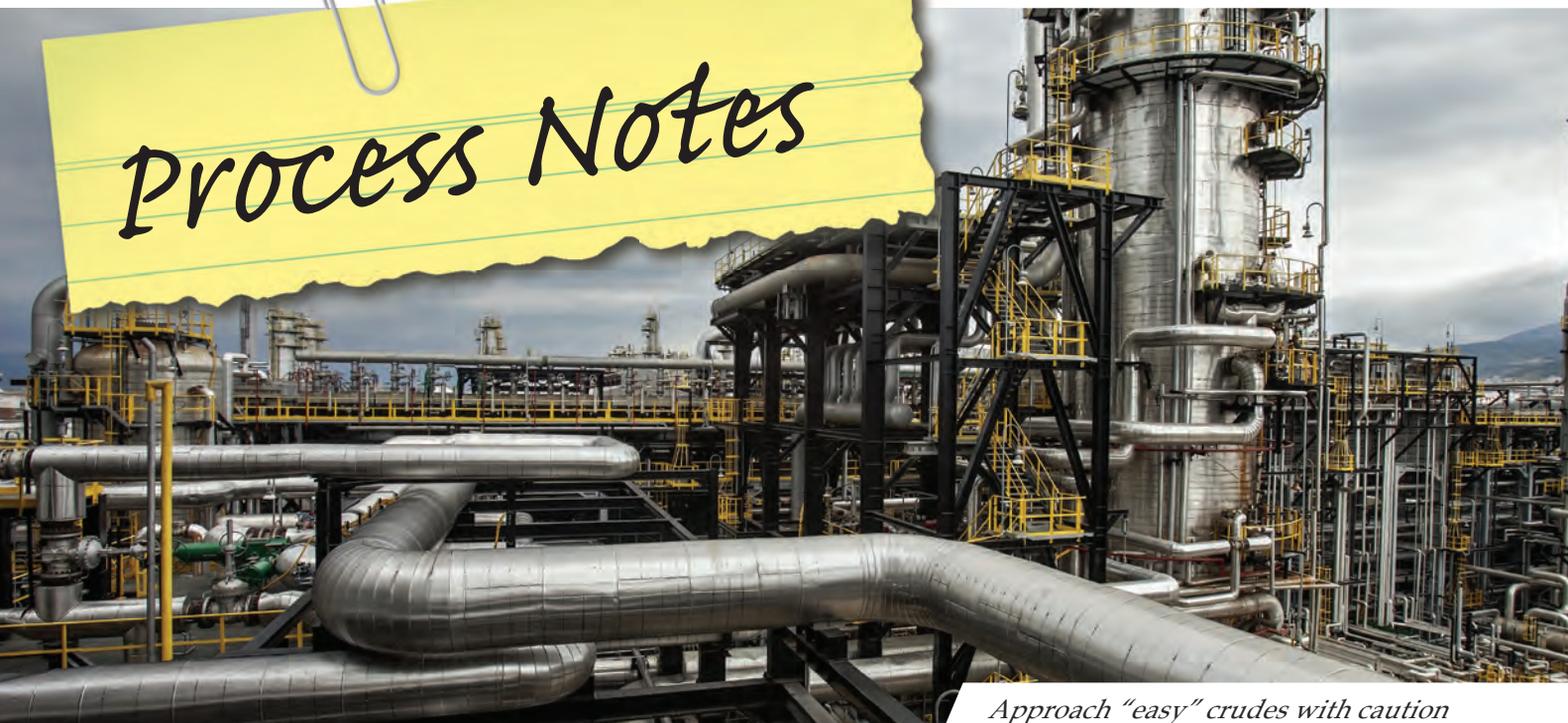
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# Process Notes



*Approach "easy" crudes with caution*

## Condensate is Crude

Ultra-light crudes and condensates are here to stay. These streams have flooded the market in recent years, and many of them are deeply discounted against reference crudes. Refiners have been processing increasing percentages of this light material through their Crude Distillation Units (CDUs) up against unit naphtha handling limits. On the surface, processing condensate and other ultra-light crudes with high API gravity and low sulfur should be easy. In reality, many refiners have experienced significant challenges, some of which are unique to ultra-light crudes and condensate.

Although their bulk properties signal that these crudes should be easy to process, new recovery techniques tend to leave undesirable compounds in the crudes that can adversely affect refinery CDUs or Condensate Splitters. Some of the bad actors are:

- High melt point waxes / high paraffin content
- Tramp amines from production  $H_2S$  scavengers
- Filterable solids
- Tramp phosphorous compounds



*Severe crude column tray fouling*

These undesirable compounds are the source of operating and reliability problems in CDUs and Condensate Splitters worldwide, and the onset and severity of certain problems can often be traced back to the introduction of new ultra-light crudes and condensates.

These supposedly "easy" crudes have been linked the following problems:

- Fouling in the cold preheat train
- Poor desalter performance
- Fouling in the warm and hot preheat trains
- Crude heater fouling and hot spots
- Accelerated overhead system corrosion
- Salting in the top of the crude column
- Plugging of kerosene section trays and exchangers
- Plugging of stripping trays

Despite the impression that new ultra-light crudes and condensates should all be easy to run, they are not. Condensates and ultra-light crudes are crudes, meaning that many of them can be difficult to process and can present unique refining challenges.

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# Evolution of a digital twin

## Part 2: Use of the digital twin

A two-part article provides a step-by-step introduction to the concept, construction and application of a digital twin for the lifecycle of an ethylene plant's steam cracker

OTMAR LORENZ, BERND-MARKUS PFEIFFER, CHRIS LEINGANG and MATHIAS OPPELT  
Siemens

Part 1 of this article (see *PTQ* Q1 2020) describes the development of a digital twin for application over the entire lifecycle of a process plant. Here, Part 2 of the article describes how, after the implementation phase is completed, it is possible to work with the digital twin to answer specified tasks in the engineering and operational area of the plant.

### Engineering

Typical segments of engineering where the digital twin can be used are in the design of basic engineering of process equipment and the automation system, virtual commissioning of the control system, and the training before start-up of a new plant or reconfiguration of an existing plant.

### Design

The goal of simulations in the design of a process engineering plant is the creation, verification, and refinement of the plant design. The focus is on considering the actual process. Controllers are only available in simplified form, if at all, as part of the process model. It is imperative that different process drafts can be compared with one another in order that the most suitable can be selected respectively. The accuracy of the simulation must be sufficiently good to be able to make process-related decisions correctly.

A static process simulation is sufficient for the design of plants in steady state continuous operation; a dynamic process simulation must be used for the simulation of start-up and shutdown processes and the transients between operat-

ing points. It may prove practical to combine models of different tools, either by exchanging models or via co-simulation.<sup>1</sup>

### Virtual commissioning

The aim of virtual commissioning is to achieve a fully tested automation system wherever possible.<sup>2</sup> The main focus is on testing the implemented PLC application software, developed uniquely for every system. For testing – for instance, signal routing, continuous function charts (CFC), sequential function charts (SFC), faceplate and pictures for operator station (OS),

## The aim of virtual commissioning is to achieve a fully tested automation system wherever possible

and alarms – a simulation model can be used, which operates the complete communication interface between automation and field and is connected to the real (hardware-in-the-loop) or emulated (software-in-the-loop) control hardware. It is imperative for both set-up scenarios that at least the communication behaviour of field devices (actuators and sensors) is replicated in the simulation model.

Replicating process behaviour (physical behaviour) will also prove practical for testing SFCs. This can be done, for example, with the simulation of a cold commissioning, in which the behaviour of the process

is observed as long as only water is pumped through the system as a medium and no chemical reactions take place. Extremely detailed process models are required wherever the controller is to be parameterised. The connection of existing process models via co-simulation can be exceptionally advantageous in this regard.

At least for the hardware-in-the-loop configuration, the simulation system must be capable of supplying and processing signals within stipulated real time. Simulation models are also implemented as part of the control program on the automation hardware in a special software-in-the-loop configuration, eliminating the need for additional simulation tools.<sup>3</sup> However, these advantages are offset by certain disadvantages. The control program is altered following testing, and simulation-specific functions such as a virtual time (faster or slower than real time), snapshots (saving model states) or even co-simulations may be difficult to attain with the resources of the automation system, if at all. Test cases which could be created automatically<sup>4</sup> and automatically executed would be beneficial in ensuring the most efficient test possible.

### Training (OTS)

The objective of training simulation is to prepare operators for their tasks as effectively as possible. This encompasses both interaction with the process control system (ideally on the basis of the original operating screens and programs), as well as familiarisation with the reaction of the process itself. Training for

interaction with the process control system can be realised in accordance with the selected modelling depth based on the model which was created for virtual commissioning.

For training related to the process itself, it is necessary to model this in detail. Such models are thus also ideal as training for limit situations, start and stop procedures, and emergency scenarios. It is therefore essential that training scenarios can be created and adapted. In addition, it must be possible to assess, compare, and verify the performance of trained personnel.<sup>5</sup> Moreover, particular attention must also be afforded to the didactic concept when devising the scenarios.<sup>6</sup>

### Plant operation

Typical segments of plant operation where the digital twin can be used are in the design of virtual sensors, advanced process control, optimisation, and maintenance systems. Various aspects of application are described in more detail in the following chapter.

### Soft sensor

Soft sensors represent an important application of a digital twin during the operation phase. A soft sensor estimates an unknown process variable based on a model of the process and other available measured variables. Common examples include the Luenberger state observer<sup>7</sup> or the Kalman filter,<sup>8</sup> which are based on dynamic process models in the form of differential or difference equations. As all variables are known in the simulation model, the variables being estimated can be obtained directly.

Reverting back to the digital twin of the system will ensure that a model based soft sensor does not have to be modelled anew for each application. A dynamic process simulation which is already available must be analysed and, where necessary, the sub-model separated for the process section for which a soft sensor is required. It is then only necessary to parameterise and to validate the soft sensor algorithm using process data.

The effort afforded for implementation of a soft sensor is worthwhile

if the estimated variable is essential for process control. Estimated variables can be applied for monitoring tasks in which the exothermic reaction is estimated, for example, and monitored for a maximum permissible value to avoid unfavourable or dangerous process states.<sup>9</sup> Direct control of estimated variables is also possible. Thus, in the example given, the yield can be measured, but only after several steps of the procedure have been executed. The resultant dead time, which is many orders of magnitude greater than the actual process dynamics, renders direct control of the measured yield impossible. The estimated yield at the output of the cracking furnace, however, is provided free of dead time via the soft sensor and may thus be used for direct control.

## A soft sensor estimates an unknown process variable based on a model of the process and other available measured variables

### APC

All higher level control procedures which go beyond standard single-loop PID controllers come under the APC keyword. In view of the task definition for a multivariable control on the steam cracker, model based predictive control (MPC) seems to be the most appealing option.<sup>10</sup> All predictive controllers are based on the basic principle of internal model control (IMC). A dynamic model of the controlled system is part of the controller and is used during runtime to predict future process behaviour in a defined prediction horizon. The model knowledge of the digital twin can be used as a basis for the process model of predictive controllers. Essentially, there are three procedures open to you in this regard:

1. For a non-linear predictive controller, a (sub)model is used from

the dynamic simulation model of the digital twin. However, this must generally be simplified considerably in light of the real time capability of the controller, as numerous simulations of the process model can be calculated throughout the entire prediction horizon in each scanning step of the MPC. Non-linear MPC concepts therefore present special challenges due to the process model, but also due to the dynamic online optimisation.

2. A model can be derived for a linear predictive controller through numeric linearisation around an operating point within the simulation software. The advantage of linear MPC concepts lies in the considerably reduced computing effort. It is thus possible to implement the MPC directly in the process-level component of a control system, with the respectively associated advantages regarding availability, operator control and monitoring, usability, and expenditure.

3. Step change attempts are performed with the simulator. The artificially generated training data is then used for the identification of linear models with the configuration tool of the MPC. This procedure has the advantage that it can be executed with the existing software infrastructure.

### MPC concept for a steam cracker

Precise apportionment of the functional scope in a multilevel solution concept with basic automation, soft sensors, MPC, and RTO is the result of intense discussions between the project partners at PSE and Siemens. The MPC concept comprises four controlled variables for one half of the furnace, all of which originate from the soft sensor. Any existing thermal couplings between both halves of the furnace are not represented in the simulation model and are thus not in the MPC.

The following are provided to the MPC as manipulated variables:

- Supply setpoint combustion gas
- Supply setpoint hydrocarbons (reactant)
- Ratio 'process steam to reactant'

The MPC specifies the setpoint for the gas-to-air ratio controller (GARC), and thus indirectly the

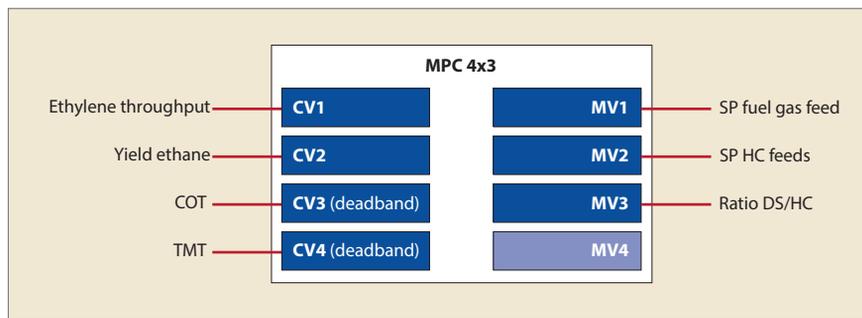
burner inflow made up of combustion gas and air. The ratio between combustion gas and air is regulated at a lower level by the GARC. The ratio of process steam to reactant supply can be influenced by the MPC as a third manipulated variable, however the permissible range for this ratio is very limited. This shows that the third MPC manipulated variable has minimal influence and the majority of the time is to be found in restriction of the manipulated variables. The MPC therefore has three degrees of freedom, of which only two are usually applicable.

The aim of closed-loop control on the one hand is to maintain the quantity of ethylene and propylene as high as possible, whilst at the same time attaining a high conversion rate of the reactants to ensure a minimum of wastage. Limit values must be respected at all times to ensure safety of the system. The throughput of the desired product is selected as the first controlled variable (see **Figure 1**). However, the ethylene throughput could only be measured with a greater dead time for the separation section of the overall cracker system.

The throughput is therefore calculated using the product from the supplied quantity of hydrocarbons and the ethylene yield, which is estimated by the soft sensor. The soft sensor variable conversion rate of the supplied ethane is applied as a second controlled variable. Since only two degrees of freedom are effectively available due to the restricted number of manipulated variables, additional control variables can no longer be regulated precisely to their setpoint. In order that a safe system state can be guaranteed, the coil outlet temperature (COT) and tube metal temperature (TMT) are therefore maintained by the MPC in tolerance bands as third and fourth controlled variables.

The following requirements are defined for plant operation:

- Maintain *Conversion Rate Ethane* at setpoint
- Run *Ethylene Throughput* to defined setpoint
- Maintain *COT* within specified range



**Figure 1** MPC 4x3 configuration with controlled variables (CV) and manipulated variables (MV)

- Maintain *TMT* below critical upper limit

Dead zones are specified for the controlled variables 3 and 4. Should either of these temperatures stray from the permissible range, the high weighting of the control deviation of these variables is brought to bear.

The MPC performs the following task from the viewpoint of the plant operator. Determine the suitable setpoints for supply of reactant, process steam, and combustion gas to achieve a defined production rate with the necessary conversion rate, and to ensure that the temperatures (COT and TMT) remain within the specified range. The concept can be expanded with measurable disturbance variables, for example composition of the reactant and the heat value of the combustion gas.

For the design of the MPC, the MPC engineering toll from Simatic PCS 7 was used.

The MPC concept can be combined with the lower level 'pass balancing'. A weighted COT mean value of all coils is calculated in this regard, where the respective feed is applied as a weighting factor. Controlled variables are then the deviation of individual COT values from this mean value; no dead zone is used here. Manipulated variables are the ratio factors with which the feed setpoints for individual coils are calculated from the overall feed.

This concept is compatible with previous, higher level MPC concepts for the overall cracker if the mean COT is influenced by the total reactant inflow as previously. The same amount of additional controlled and manipulated variables is added. In principle, pass balancing can be implemented with single-variable controllers, as long

as cross-influences between neighbouring pipe strings are negligible and the summary effect of pass balancing is neutral.

### Plant-wide optimisation

Many different system components are integrated in a large petrochemical plant for the production of ethylene and propylene. The requirement and the market environment of the individual reactants and products may therefore change from day to day. In order that an optimum operating profit can be achieved, the operating point of the plant must be adapted to the market environment. This problem can be solved as an optimisation problem based on the digital twin. To this end, the optimum setpoints are calculated for each individual system component under defined boundary conditions using a specified target function and the strict model of the complete plant. For example, a target function for maximisation of profit for each unit of time can be structured as follows:

$$Profit = Turnover - Costs = \sum_{i=1}^n (Proceeds_{p,i} Flow rate_{p,i}) - \sum_{i=1}^k (Costs_{e,i} Flow rate_{e,i})$$

Profit is calculated from the difference between the proceeds anticipated for the  $n$  various products  $p$  and the costs for the  $k$  various reactants  $e$ . This simple calculation could also incorporate additional boundary conditions, such as energy costs or maintenance planning.

The optimum values of the controlled variables calculated in this manner are applied to specify optimum setpoints for individual process units for a specific period of

time, referred to as stationary operating point optimisation. Optimum transition from one stationary state to another is a task for the lower level controller structures, for which a dynamic process optimisation may be applied.

### Maintenance

With increasing production time, coking occurs in cracking furnaces which crack long-chained hydrocarbon molecules. This causes the plant behaviour to change continuously throughout the production time, until coking has built up to such an extent that the corresponding cracking furnace must be shut down and cleaned. To minimise the downtime required for cleaning work, it makes sense to adapt the mode of operation to ensure that coking is kept to an absolute minimum.

The capital which has already been invested in the digital process twin during the plant design phase can also be used again here. The strict plant model is used to perform 'what if' experiments. These scenarios are helpful, for example, in maximising production time (remaining service life until cleaning is required) or for optimum planning of the maintenance time.

If the production conditions change due to bottlenecks in resources or due to volatile raw material prices, the economic balance can be improved by these kinds of experiments using a digital process twin.

### Re-engineering of the digital twin

In the course of plant operation, the plant undergoes constant changes, whether this be simple wear-related replacement of a component or optimisation of the process following a conversion. All constituents of the digital twin must be constantly updated in every scenario. For this, it is essential to provide automatically usable interfaces between the participating tools, as a manual adaptation is not only unrealistic in terms of cost, the error rate is also far too high with manual compensation using completely different tools.

Moreover, it is advisable to verify the behaviour of simulation models

regularly with real measuring data. On the one hand, this will allow unwanted changes in plant behaviour to be detected in accordance with the validity range and quality of the model. On the other hand, changes intentionally implemented in plant behaviour must be replicated in the simulation. Although such simulation adjustments can be supported by process identification and parameter estimation techniques, they should always be checked and validated by an employee.

### Conclusion

This article describes numerous facets of integrated utilisation of a digital twin for procedural systems. Concrete implementation of the various applications using the example of a steam cracker makes it possible to understand the interplay between the various components of digital twins and the various tools involved, and clearly shows the benefits of an overall consideration of the digital twin. This results in numerous benefits which are clearly visible in the application example, starting with plant configuration and the use of simulations for process design to virtual commissioning, all the way to process optimisation. So there is the hope that in the future no isolated cost-benefit estimate for the creation of a simulation model for tasks in the operating phase will be necessary any more because all the required information and models are not only available in the form of the digital twin, but are also always up-to-date and thus directly usable for the real twin.

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# HIGH STANDARD VALVES FOR NON-STANDARD CONDITIONS.



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# Process Notes



*Approach "easy" crudes with caution*

## Condensate is Crude

Ultra-light crudes and condensates are here to stay. These streams have flooded the market in recent years, and many of them are deeply discounted against reference crudes. Refiners have been processing increasing percentages of this light material through their Crude Distillation Units (CDUs) up against unit naphtha handling limits. On the surface, processing condensate and other ultra-light crudes with high API gravity and low sulfur should be easy. In reality, many refiners have experienced significant challenges, some of which are unique to ultra-light crudes and condensate.

Although their bulk properties signal that these crudes should be easy to process, new recovery techniques tend to leave undesirable compounds in the crudes that can adversely affect refinery CDUs or Condensate Splitters. Some of the bad actors are:

- High melt point waxes / high paraffin content
- Tramp amines from production H<sub>2</sub>S scavengers
- Filterable solids
- Tramp phosphorous compounds



*Severe crude column tray fouling*

These undesirable compounds are the source of operating and reliability problems in CDUs and Condensate Splitters worldwide, and the onset and severity of certain problems can often be traced back to the introduction of new ultra-light crudes and condensates.

These supposedly "easy" crudes have been linked the following problems:

- Fouling in the cold preheat train
- Poor desalter performance
- Fouling in the warm and hot preheat trains
- Crude heater fouling and hot spots
- Accelerated overhead system corrosion
- Salting in the top of the crude column
- Plugging of kerosene section trays and exchangers
- Plugging of stripping trays

Despite the impression that new ultra-light crudes and condensates should all be easy to run, they are not. Condensates and ultra-light crudes are crudes, meaning that many of them can be difficult to process and can present unique refining challenges.

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# Improving hydrotreater performance with welded plate heat exchangers

Pressure has never been higher on refiners to improve the efficiency of their energy-intensive processes

WIVIKA LAIKE and CHRIS WAJCIECHOWSKI  
Alfa Laval

Recent volatile crude oil prices have led to the growth of several alternative methods for crude oil extraction around the globe. These extraction methods can often produce sour crude oil with a high sulphur content. At the same time, the demand for high sulphur products is decreasing as the understanding of the environmental effects of burning high sulphur fuels grows. This necessitates that refiners are now required to reduce the sulphur content of their products before they can be sold. This is most commonly done by hydroprocessing, such as hydrotreating and hydrocracking.

Focusing on hydrotreating as a process for removing unwanted impurities such as sulphur, nitrogen, and metals, there are several alternative configurations. However, at the heart of hydrotreating there is always the reactor section, featuring a high pressure reaction vessel as well as reactor internal technology and catalyst, enabling the feed to react with hydrogen. This unit operation is common for hydrodesulphurisation, but also in other fuel upgrading technologies such as isomerisation and catalytic saturation. These processes are energy intensive and as such require a high degree of heat integration to lower

the energy operating expenditures (opex). Earlier, this heat integration was performed with shell-and-tube (S&T) heat exchangers. However, more recently refineries have been maximising energy efficiency by using welded plate heat exchangers by Alfa Laval in the main heat recovery positions. This article explains the advantages of using welded plate heat exchangers in key heat integration positions in hydrotreaters among several refiners around the world.

## Process layout

Figure 1 shows a general process layout of a hydrotreater. Hydrogen is

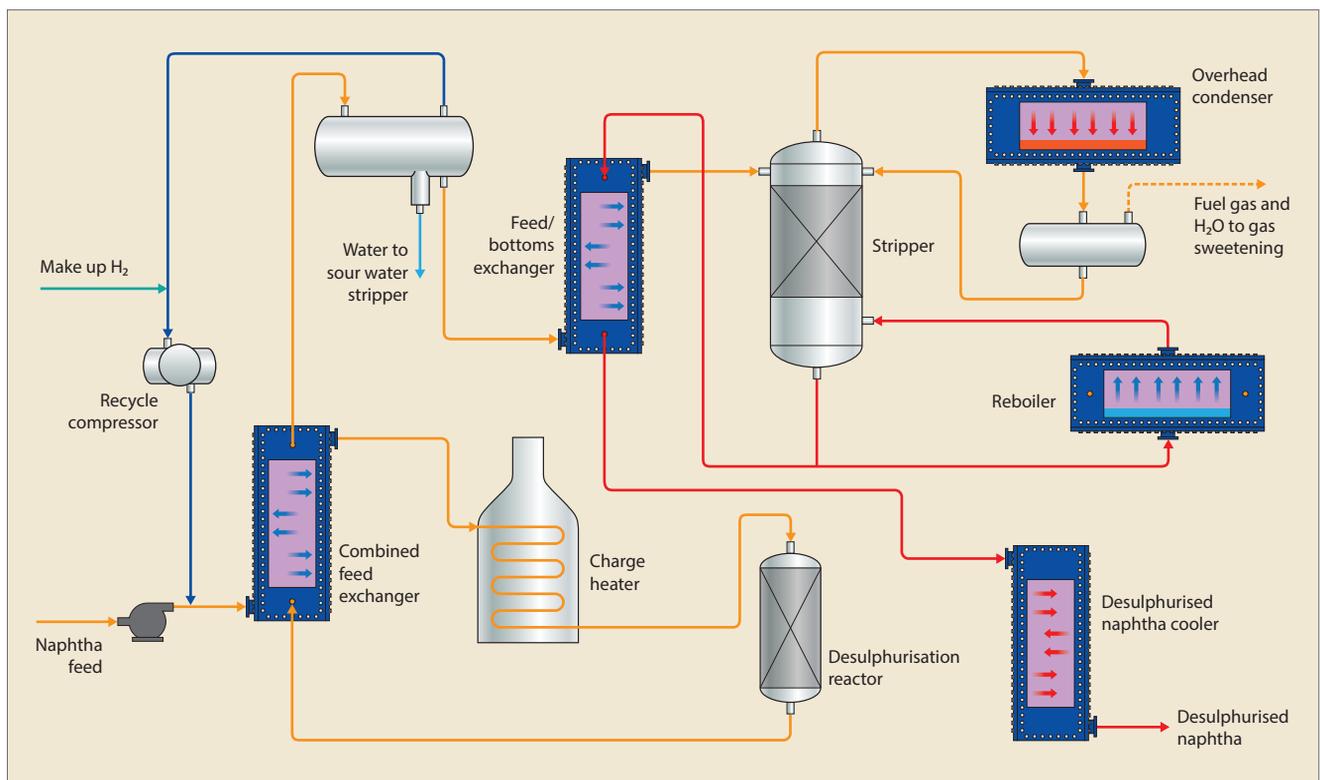


Figure 1 General process layout of a hydrotreater using welded plate heat exchangers in several positions: CFE, stripper feed/bottoms, product cooler, stripper reboiler, and stripper overhead condenser

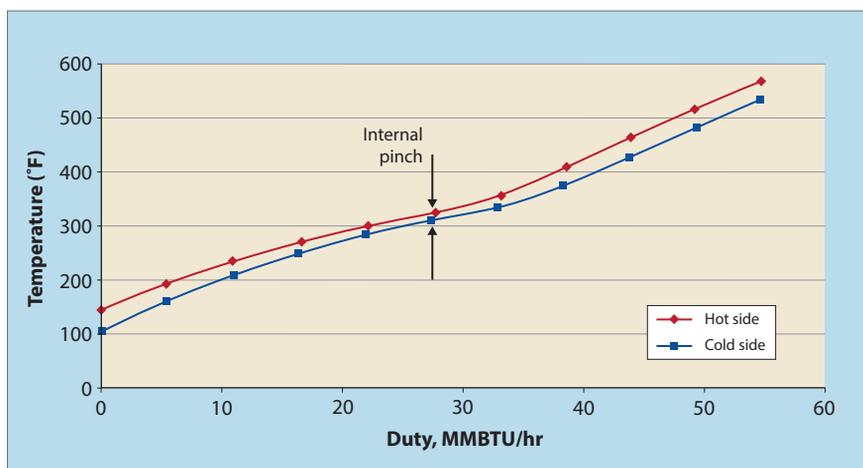


Figure 2 Generic enthalpy curves for a feed/effluent heat exchanger, welded PHE

added to the feed stream which is then vaporised and superheated in a heat exchanger. The reactor effluent is used as a heating medium; this in turn needs to be cooled and condensed before being separated into various products. The more you heat up the feed, the less energy has to be used in the furnace. And the more you cool down the effluent, the less energy has to be used in the subsequent (air) cooler. Therefore process designers usually focus on minimising the hot end approach temperature (HAT) and internal pinch (minimum delta T) of the combined feed/effluent (CFE) exchanger. Typical internal pinch temperature for this position when S&T is used is between 20°C and 40°C (36-72°F). However, with plate technology the temperature difference can easily be reduced to less than 6-10°C (11-18°F, see Figure 2). This means that by using a traditional technology, the number of S&T in series and the heat transfer area needed to do the same duty will be significantly higher, as will the cost of the heat exchangers.

Traditionally, a maximum of eight S&Ts in series have been used in such a service, as an optimum with respect to investment cost versus achieving more heat recovery. As the efficiency of these heat exchangers affects the surrounding process equipment, their performance should preferably be fixed at an early stage.

**Benefits of Compabloc/Compabloc+** Welded plate heat exchangers, such as the Compabloc, extend the prac-

tical performance limits of heat recovery by using engineered corrugated heat transfer surfaces to generate three to five times the heat transfer coefficient compared to traditional technologies. At the same time, fouling rate is minimised. The flow geometry also achieves very close to counter-current flow, perfect for small temperature differences and high heat recovery. Further on, as already explained, a single Compabloc can replace several S&T heat exchangers, significantly reducing the amount of space required for installation by up to 90% compared to traditional S&T heat exchangers.

The heat recovery capability of Compabloc means that less fossil fuel is consumed, and that emissions and carbon footprint are reduced. A compact design makes installation easier and more cost effective. By freeing up space, Compabloc resolves bottleneck issues, enabling new ways to increase production and heat recovery.

Compabloc operates with superior shear stress, minimises fouling, and allows operation with cleaning intervals substantially longer than traditional S&T heat exchangers. It is equipped with four removable panels to allow full access to the heat transfer area for cleaning or inspection. The fully cleanable design with cleaning lanes at each side of the plates means that the exchanger can be returned to 100% performance when cleaning the plate pack by hydrojetting, ensuring a long and highly efficient operational life cycle. This is especially important in duties like naphtha hydrotreater

CFE where gumming and salting can occur.

The classic range of Compabloc can go up to 38 or 42 bar depending on the model. The Compabloc+ range however allows operation up to 60 bar because of new features, and brings the benefits of Compabloc technology to a wider range of positions.

Compabloc+ is equipped with +Seal, a sealing concept that allows the panels to be reliably sealed at high pressure. In Compabloc+ design, the graphite gasket is fully contained in a groove and the compression is controlled with a metal-to-metal contact which prevents overtightening and related damage.

Several cases where Compabloc welded plate heat exchangers have been successfully utilised in hydrotreater systems will be presented in this article.

### Case 1

A major refinery in Europe sought to debottleneck an existing naphtha hydrotreater and identified the CFE heat exchanger heat recovery as a limiting factor to its project goals. Additional S&Ts in series with the existing S&T CFE train were evaluated, but the performance of the additional shells fell short of its goals and the new train consumed too much pressure drop. Finally, as is often the case, not enough space was available to install the additional S&T exchangers so the project was not feasible with this technology.

Concurrently, the refinery evaluated Compabloc heat exchangers to be placed on the hot end of the feed side, vaporising and superheating the feed with hot reactor effluent while still using the S&T exchangers on the cold end. The operating parameters were optimised by a team consisting of Alfa Laval specialists and the customer's process specialists who iterated the Compabloc and S&T performances at different temperature approaches. The optimum point in this case was to design the Compablocs for a 12°C (21°F) pinch temperature and a 22°C (39°F) hot end approach temperature, limited by the performance of the existing S&T equipment on the cold end. This performance met the



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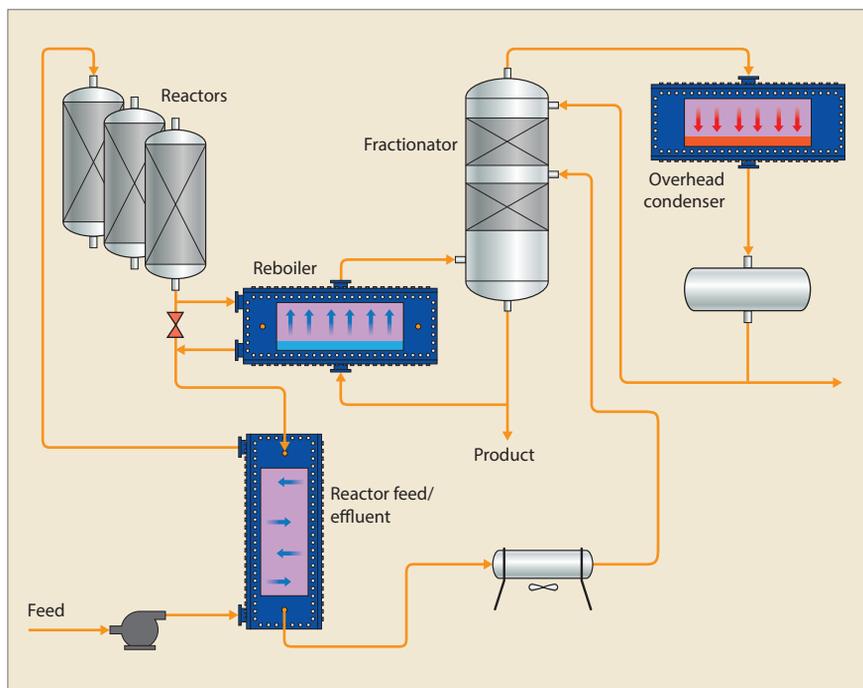
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**Figure 3** Isomerisation process employing Compabloc as CFE and process reboiler

project's goals and the solution was implemented in 2016.

The refinery started up the Compablocs in early 2017; initially, the thermal and hydraulic performance was exactly as expected. However, after several months of operation, increases in pressure drop were detected on the hot end of the feed side in the Compabloc, located at the dry point of the exchanger. The root cause of fouling was determined to be an excess of corrosion inhibitor being dosed in the naphtha upstream, causing the filming agent to be deposited at the dry point of the feed. The Compabloc was cleaned of the fouling material and a process adjustment was made, after which performance returned to typical unit cycle length. For more information, see page 12 in this issue's Q&A section where this case is discussed in more detail.

### Case 2

In 2014, a prominent US refinery was developing a project to build a grassroots naphtha hydrotreater as part of its Tier III gasoline sulphur reduction initiative. The refiner had the ambition to design the process as efficiently as possible, but also needed to be held to strict limits on total installed costs. For both of these reasons the refinery identified

Alfa Laval as a viable solution, and our specialists began working with the customer's process specialists to optimise a solution within cost constraints.

Similar to the European case, the first place to start was the CFE, and several iterations with different amounts of heat recovery were done to balance exchanger cost with the whole project cost. One counter-intuitive learning from the process was that increasing the Compabloc exchanger size actually lowered both the operating cost (better heat recovery) and the total capital cost (capex) by being able to downsize other equipment. The optimum design basis for this Compabloc CFE was to use a 7°C (12°F) internal pinch temperature and a 20°C (35°F) hot end approach (HAT) temperature. The economics also favoured employing Compabloc exchangers as the final reactor heater and as the stabiliser feed/bottoms heat recovery exchanger. All were optimised for efficiency and cost reduction and the three services were put into operation in 2016.

Since start-up, the performance of all exchangers has met expectations and no performance reduction has been detected. At this pace, the Compablocs will easily provide the required performance until the next scheduled turnaround. In

fact, because of the success of this installation, the same refining company is installing Compabloc as CFE exchangers in another naphtha hydrotreater unit (NHT) in 2020. This NHT is being debottlenecked to a greater degree by employing Compabloc with similar tight temperature approaches.

### Case 3

In this case, Compabloc was evaluated for debottlenecking an existing cracked naphtha hydrotreater (Axens Prime G+) in Canada as part of the refiner's Tier III sulphur reduction initiative. Seeking more heat recovery was interesting since the fired heater downstream of the CFE was limited both by size but also by environmental emissions limits. If this refinery wanted to expand, it needed to do so below the environmental limits. Recovering energy and saving emissions are ideal reasons to evaluate Compabloc, so a team of Alfa Laval and refinery process specialists optimised the heat recovery of the system and selected a Compabloc to be installed on the hot end of the gasoline desulphurisation (GDS) reactor in series with existing S&Ts. The HAT of this installation was optimised at 41°C (75°F), again limited by the S&T exchangers upstream on the cold end of the CFE.

In addition to the work at the process stage to optimise the process design, equally important is the installation and operation of Compabloc as CFE heat exchangers. Alfa Laval specialists worked with the refinery on developing the start-up and shutdown procedures, as well as design features such as water wash, sulphiding, and emergency shutdown procedures. This process unit was successfully started up in 2019 and is running smoothly.

### Case 4

Similar to hydrotreating processes, other refinery and petrochemical processes employ a reaction section with feed/effluent heat exchangers that could benefit from plate technology. In this example, Compabloc was evaluated by an independent US refiner to debottleneck a gasoline isomerisation process licensed

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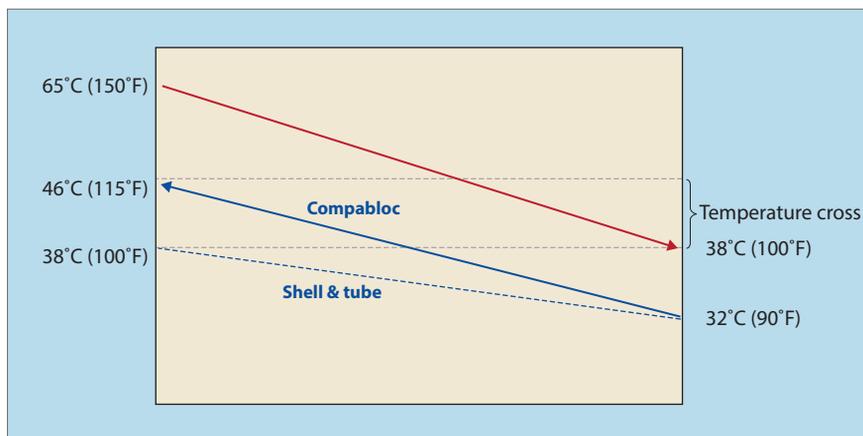
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**Aid to foster biofuel program**





**Figure 4** Typical product cooler temperature program

by UOP. The project was a revamp to increase throughput but also to replace a fired stabiliser reboiler with a heat exchanger using reactor effluent heat. The bottlenecks could be alleviated by recovering more heat in the CFE heat exchanger, so a Compabloc was a natural choice.

An evaluation was done by a team of Alfa Laval specialists and refinery site specialists to optimise the performance of the project. Quickly, it was discovered that if Compabloc was used both as the CFE heat exchanger and as the process reboiler, less reactor effluent needed to be diverted to the reboiler which reduced the overall project cost (see **Figure 3**). The refinery took full advantage of the performance benefits of Compabloc and the project moved forward. Both the CFE Compabloc and the reboiler Compabloc were installed and started up in 2017. Analysis of the operating data showed a 19°C (35°F) HAT on the CFE, exceeding design performance with no fouling detected in the first year of analysed data. The reboiler is performing with an average 3.3°C (6°F) temperature difference between the boiling fluid and the reactor effluent outlet temperature, which is world class performance for a reboiler. Both heat exchangers are expected to perform at this level well into the next turnaround schedule.

#### Case 5

Another process that employs reactor feed/effluent heat exchangers is the benzene saturation process. In this case study, a benzene saturation unit at a US refinery was revamped for additional capacity

in response to the MSAT II gasoline benzene reduction initiative of 2011. Alfa Laval specialists worked with the refiner's corporate engineering to develop the scope that produced the best possible project outcome by using Compabloc heat exchangers.

In this process the bottlenecks were identified as the CFE pressure drop and the stabiliser overheads and bottoms coolers. In the CFE, the additional capacity was thermally and hydraulically limiting the existing S&T preheat train, so a

### Recovering energy and saving emissions are ideal reasons to evaluate Compabloc

Compabloc was placed on the hot end of the CFE, thereby replacing some of the S&Ts. The Compabloc was able to make up the additional heat needed by the process, while staying within the allowable pressure drop. This heat exchanger was started up in 2009 and was recently inspected for the first time during a 2018 turnaround.

Of similar importance to this project was the cooling capacity of the overheads and bottoms products from the product stabiliser. S&T heat exchangers existed in this system and were designed as: (hot outlet temperature) – (cold outlet temperature) = 0. For example, if the products needed to be cooled to 38°C (100°F), the maximum cooling water outlet temperature would be 38°C (100°F) as well (see **Figure 4**). This is typically done for S&T heat

exchangers to avoid a temperature cross which forces shells-in-series, with the main disadvantages being that cooling water flow can be excessive or process outlet temperature is not as cold as desired. In this unit revamp, additional cooling capacity was needed but additional cooling water flow was not available. Either another S&T would need to be added in series, or a Compabloc could be installed in place of the S&T since this technology can easily perform temperature cross duties in a single heat exchanger. Compablocs were selected for both product coolers for their overall lower pressure drop, high thermal performance in a single shell, and dramatic space savings. These heat exchangers also started up in 2009 and have performed well throughout the turnaround cycle.

#### Conclusion

With increased focus on process sustainability and social responsibility, the pressure on refiners is high to improve energy efficiency. Energy-intensive processes such as hydro-treating are frequent targets for improvement; as evidenced by the cases in this article, many refiners are choosing to improve efficiency with Compabloc technology. This is because it is not enough any more to settle for the efficiency limitations of S&T technology to accomplish goals. Refiners must improve efficiency to stay in business, but to thrive in today's market they need to innovate in the area of energy efficiency. An increasing number of refiners are turning to Compabloc technology to improve in this area, even for their most critical and demanding processes such as hydrotreating.

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# Cross-unit APC boosts downstream performance

## Manipulating advanced process control in a crude unit reduced constraints in a downstream diesel hydrotreater

AZURA BINTI AZAHAR and SITI SARAH AHMAD NADZRI *Petronas*  
Y ZAK FRIEDMAN and SEUNGYUN NAM *Petrocontrol*

How difficult is it for advanced process control (APC) to manipulate CDU (crude unit) parameters in order to alleviate constraints of a downstream diesel hydrotreating unit (DHT)? Theoretically not too complicated, but practically difficult because these units are operated by two different operators. The CDU operator's first priority is to handle CDU constraints while maximising production of the more valuable products (see **Figure 1**). Indeed, there was an APC application in place to help accomplish such economic objectives. As it happens, at Petronas Melaka refinery such a strategy in isolation may cause problems in the DHT unit, forcing a throughput cut, costing the refinery dearly in lost premium diesel production. Where are the APC benefits then?

Management asked the Melaka APC team to mitigate this conflict, adding DHT feed constraints to the CDU APC application, and this article is about how such an order could be accomplished.

### Problem statement

Melaka's 180 000 b/d CDU2 is a high sulphur crude unit, feeding a downstream DHT, a hydrocracker, and a delayed coker. This modern high conversion complex (see **Figure 2**) produces mostly gasoline, jet-fuel and low sulphur diesel oil. It is important to keep all of these units working seamlessly, or conversion or throughput may suffer. **Figure 3** shows the connection of interest here. Light gasoil (LGO) goes to the DHT either directly or indirectly via an intermediate storage tank. Coming into the DHT, LGO is fil-

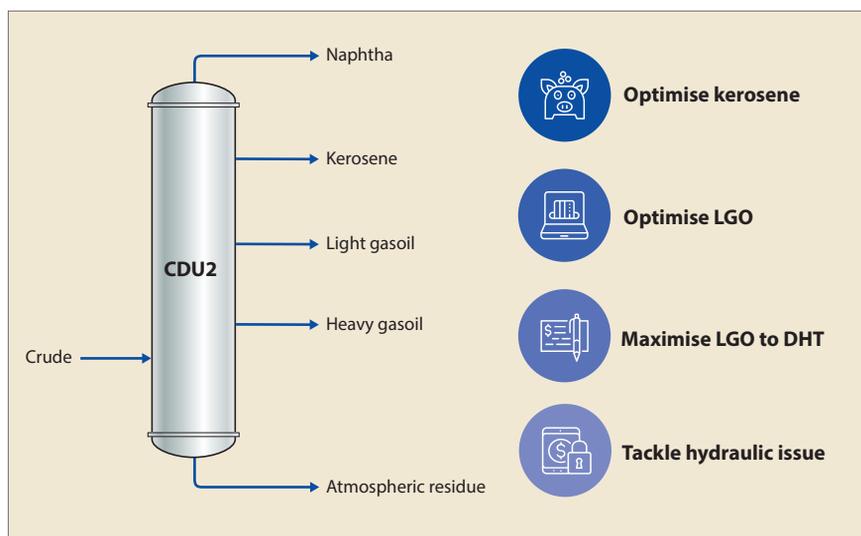


Figure 1 CDU2 products and control objectives

tered to protect the catalyst. There are two filters in parallel, one active while the other is being backwashed and then in standby. Filter switching should take place no more than once per day. However, feed flow from the tank is often contaminated by a slurry of rust, plugging up the filter, speeding up filter switching sometimes to three times a day, to the point that DHT throughput must be cut in order to reduce switching frequency. Not only is

frequent filter switching a major operating inconvenience, throughput reduction also costs dearly in lost premium diesel production.

Can we just simply increase the DHT hot feed draw and reduce tank rundown? It turns out that there are hydraulic constraints. The hot feed valve becomes saturated, and trying to maximise total LGO draw reduces the pump head, worsening the problem. The way to alleviate this situation is to reduce total LGO

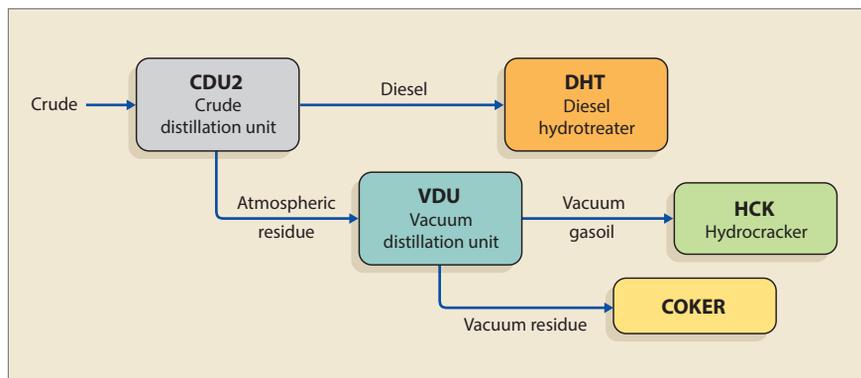


Figure 2 Melaka PSR2 complex

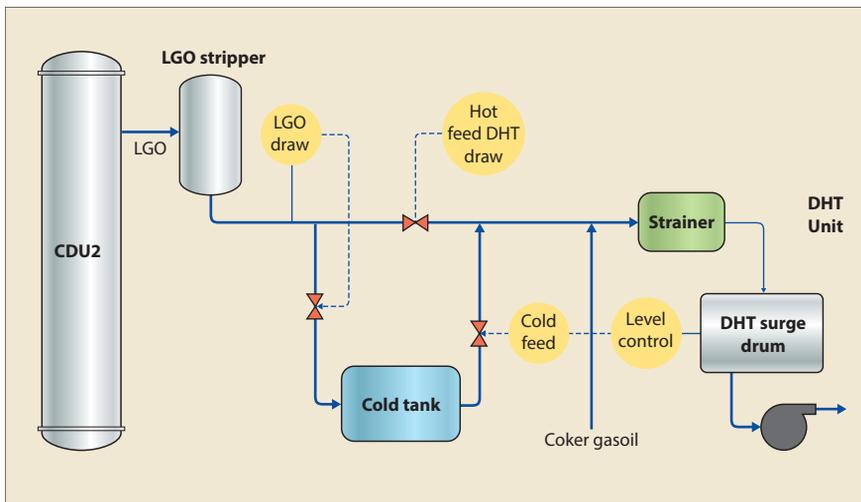


Figure 3 CDU2 LGO connection

draw, sending the excess diesel material either down to HGO and to the vacuum unit or to kerosene, provided kerosene is still on specification. Sometimes we are forced to increase the flow of overhead naphtha, an undesirable product, to reduce LGO yield. This is a complex multivariable constrained optimisation problem.

That was the driving force for configuring the CDU APC to consider DHT constraints. Ideally, the CDU APC should continue to maximise diesel production, but without running much of it down to the cold tank.

Adding to the complexity is light coker gasoil (LCGO), which is also fed to the DHT. LCGO is only 15% of DHT feed, but it fluctuates with coker drum switches, sometimes

dropping from 15% to 10% of feed for two hours or so, and that shortage is made up by dirty cold feed

## Ideally, the CDU APC should continue to maximise diesel production, but without running much of it down to the cold tank

from tank. This scenario begs for a coker APC drive to minimise LCGO fluctuations. Indeed, we have implemented such control logic,

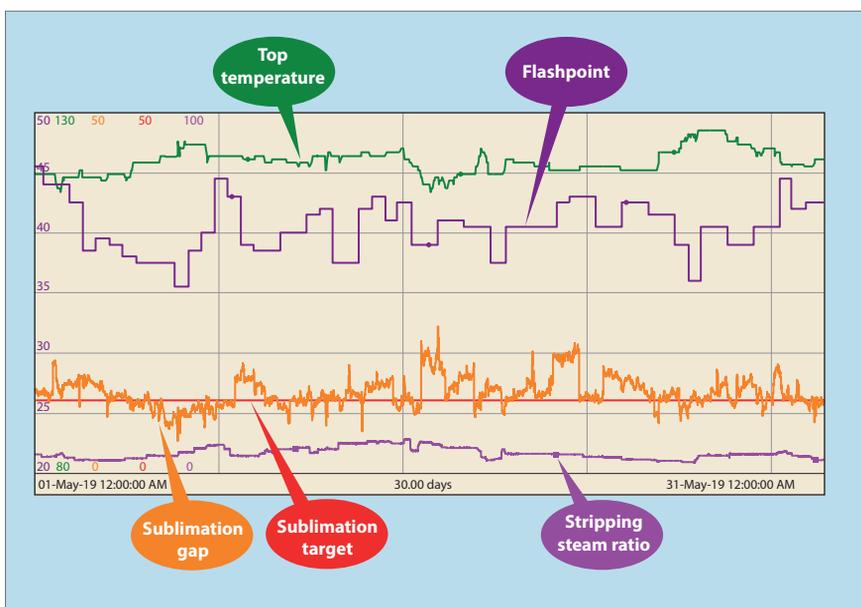


Figure 4a Kerosene flash 30-day trend before APC

and managed to reduce the fluctuations by letting coker unit levels rise and fall. Nonetheless, there is not enough LCGO inventory volume in the coker to eliminate the fluctuation completely. That coker APC scheme would be the subject of another article.

### APC history

Melaka has invested heavily in APC and presently all major units are under APC control. There is a sizable team in place to handle day to day maintenance as well as the occasional APC revamp or new applications. APC is much appreciated by operators and refinery management, and it typically runs at 90% service factor.

CDU2 APC was implemented in 2003 and has remained in closed loop almost continuously since then. It runs on RMPCT (Honeywell's multivariable controller) plus inferential control models based on a Petrocontrol CDU package called GCC. GCC works to identify the true boiling point (TBP) curve of the crude being run from column measurements, and from the crude TBP curve it estimates product properties. GCC is a reliable, well tested, first principles inferential package, and using this package permits the APC to continue working during crude switches. We would not describe GCC further here except to say that several papers have been published about its performance,<sup>1-9</sup> and one of these<sup>5</sup> describes our initial CDU2 APC implementation in Melaka. Over the years, CDU2 underwent revamps and process changes, and the APC has also been revamped to keep it current.

DHT APC was implemented in 2013, primarily in order to control diesel flash point, again a Petrocontrol inferential model. DHT APC was not the main carrier of this cross-unit optimisation drive and hence it will not be covered further in this article.

### Control and manipulated variables

The manipulated variables are typical of CDUs with one addition:

- Top temperature, controlling naphtha cut point

- Side draw flows, controlling side product cut points
- Stripping steam, controlling kerosene flash-point
- Other MVs because the actual application is more complicated than our current description
- The DHT hot feed flow is added; this can be manipulated below the point of valve saturation.

The control variables are also typical of CDUs with some additions:

- Product 95% points
- Kerosene freeze point
- Kerosene flash point
- Top temperature  $\text{NH}_4\text{C}_1$  sublimation point; top temperature must be kept above sublimation point to keep the column top trays clean. To some extent, controlling sublimation point causes kerosene flash point giveaway.
- Other CVs because the actual application is more complicated than this current description
- DHT hot feed valve position; saturation is to be avoided.
- LGO rundown to tank, to be minimised
- And most important, actual DHT cold feed flow, which directly affects filter switching frequency

### What this APC has achieved Sublimation and kerosene flash control

Figures 4a and 4b show our ability to control top temperature to sublimation limit while avoiding kerosene flash-point giveaway before and after implementation. Figure 4a is a 30-day trend before APC implementation, showing:

- Column top temperature (green)
- Kerosene flash-point lab test (purple)
- Stripping steam ratio (magenta)
- Gap between top temperature and sublimation limit (orange)
- Sublimation gap target, 10°C above sublimation model limit (red). The sublimation gap can be about 5-10°C off. Kerosene flash-point is 43-45°C, indicating giveaway of 4°C.

Figure 4b is a seven-day trend showing the same parameters plus kerosene flash-point inference (blue). APC controls the sublimation gap almost precisely to target,

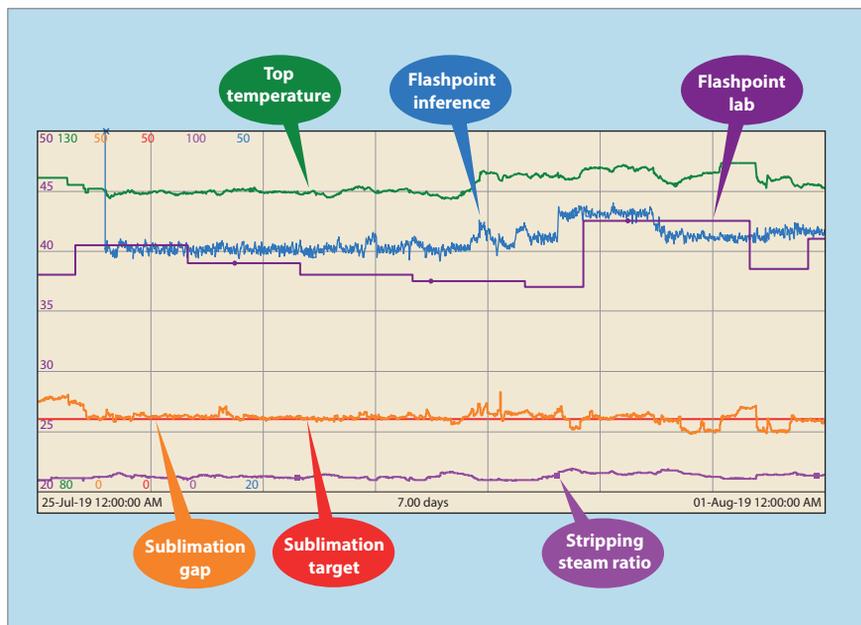


Figure 4b Kerosene flash 7-day trend after APC

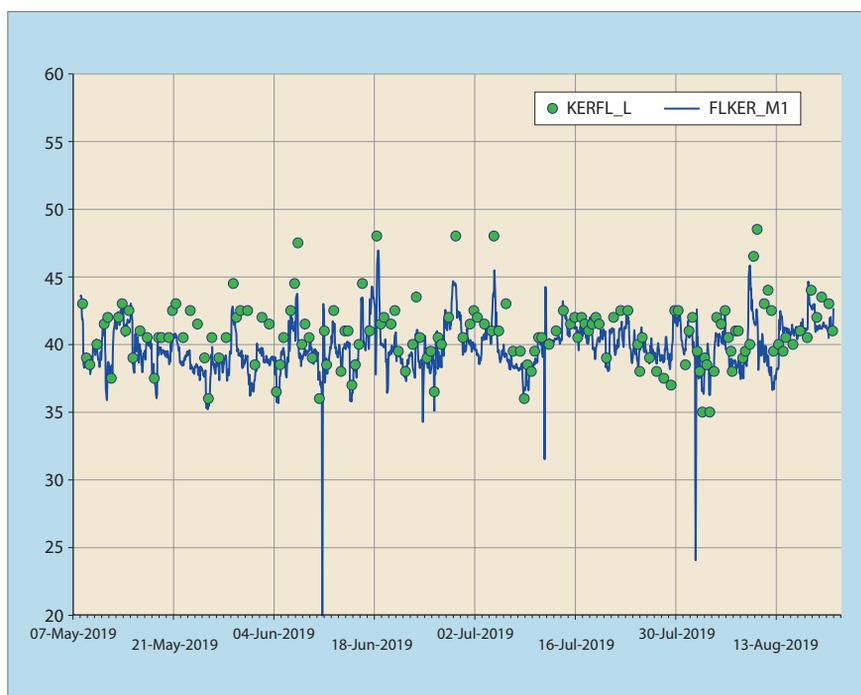


Figure 4c Kerosene flash inference performance trend

while maintaining kerosene flash-point around 41°C. At the time of this test, the flash inference had just been added to the historian and its reliability during those seven days can be questioned. Figure 4c is a four-month trend of the flash inference versus lab, showing they do track together well. This APC flash-point improvement indicates yield shift in the order of 1% from naphtha (low value) to kerosene (high value). Over a full year at current prices, that alone is worth several million dollars.

### Minimisation of LGO rundown to tank

The effects of those CDU2 actions on actual DHT filter switching frequency are illustrated in Figures 5a and 5b, which trend important DHT feed parameters for three days without APC and three days with APC:

- DHT total feed flow (magenta)
- DHT hot feed from CDU2 (green)
- CDU2 LGO production (orange)
- LCGO feed from the coker (red); coker drum switch disturbances, two per switch can be observed
- Cold feed from tank (purple); the response to LCGO shortage is obvious

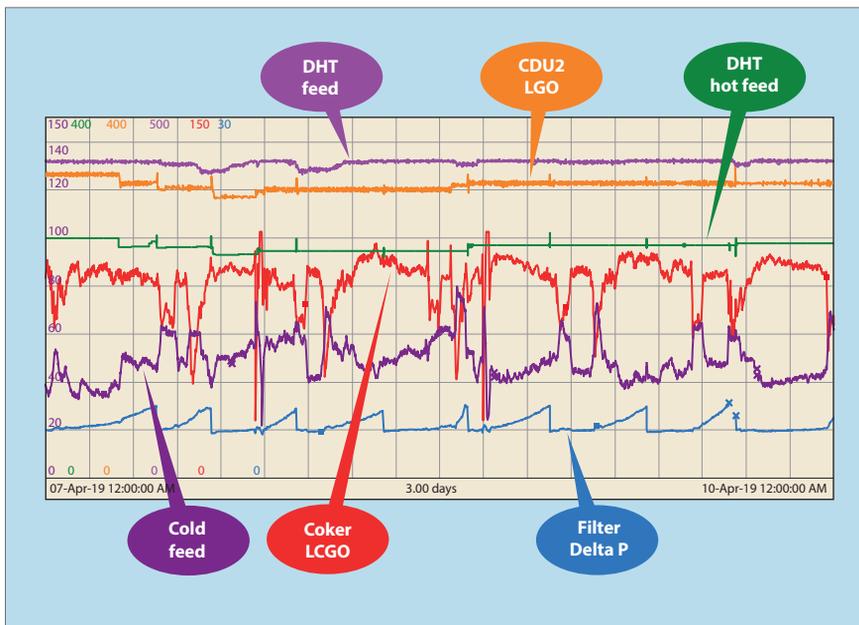


Figure 5a DHT feed filter three-day trend before APC

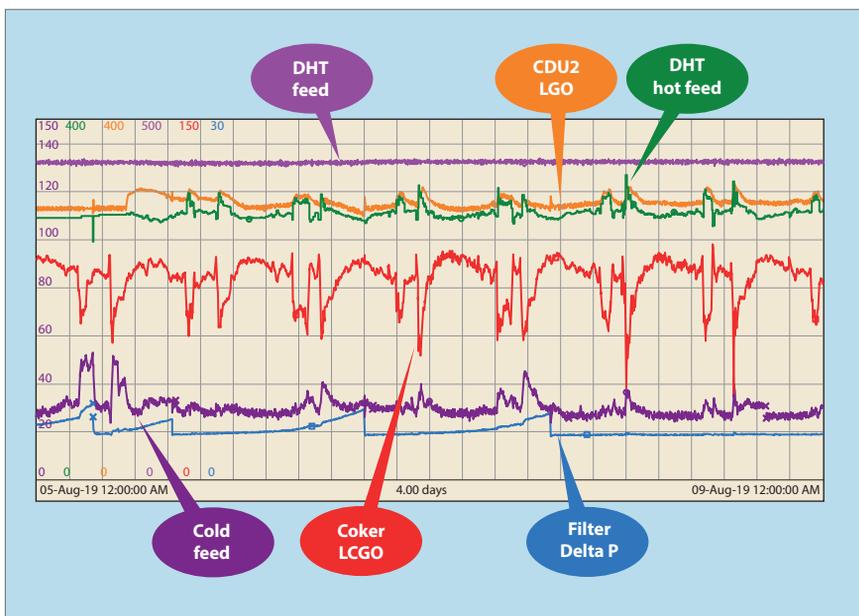


Figure 5b DHT feed filter three-day trend after APC

- Feed filter pressure difference (blue); abrupt drops in pressure difference indicate filter switches

Before APC, the DHT hot feed was manually set. Cold feed averaged at 50 m<sup>3</sup>/h, sometimes going up to 65 m<sup>3</sup>/h during LCGO disturbance. Filter switching frequency was 2-3 times a day, and sometimes the operator was forced to cut DHT feed (magenta trend in **Figure 5a**). A very different operation can be seen in **Figure 5b**. Under APC, filter switching frequency slowed down to once per day and DHT feed flow is steady. It is of interest to observe the response to coker drum switches.

CDU2 APC tries, among many constraints and targets, to minimise DHT cold feed. When cold feed increases, due to drum switch disturbance, CDU2 APC responds by temporarily increasing LGO production and DHT hot feed. With APC, the average cold feed dropped to 30 m<sup>3</sup>/h, with spikes up to 40 m<sup>3</sup>/h.

Without APC, in addition to the nuisance of frequent filter switching, there was an average 3 m<sup>3</sup>/h throughput reduction, valued as the difference between premium diesel and local diesel. Over a full year, even such a small reduction carries a penalty of several million dollars.

And during this time of CDU product yield manipulations, was the APC successfully controlling product qualities? **Figure 6** is a four-month trend of LGO 90% and 95% point inferences versus lab tests. During the initial period, May through mid-June, APC was off. Then from mid-June onwards, APC was active, and during that time LGO 95% point is stable, with minimal deviations from the 400°C target. In terms of inferential accuracy, there are certain laboratory outliers, though in general this is a high fidelity inference, tracking well against lab values.

### Conclusion

We are proud of this project on several levels. One is the elimination of a major refinery headache. No official benefit is associated with 'headache' but those of us involved with refinery operation know that operator stress may result in incidents. Secondly, while cross unit optimisation is difficult, local unit optimisers do not always improve refinery operation. Dynamic global optimisation of the entire refinery is beyond the ability of current control technology, but manipulating one unit to alleviate constraints in another is indeed within our capability, and such opportunities are the biggest APC money makers. We estimate the benefits of just tying these two units together in APC, keeping DHT throughput high, not at the expense of reducing CDU middle distillate yields, at \$10 million annually. That is on top of all other CDU APC control benefits.

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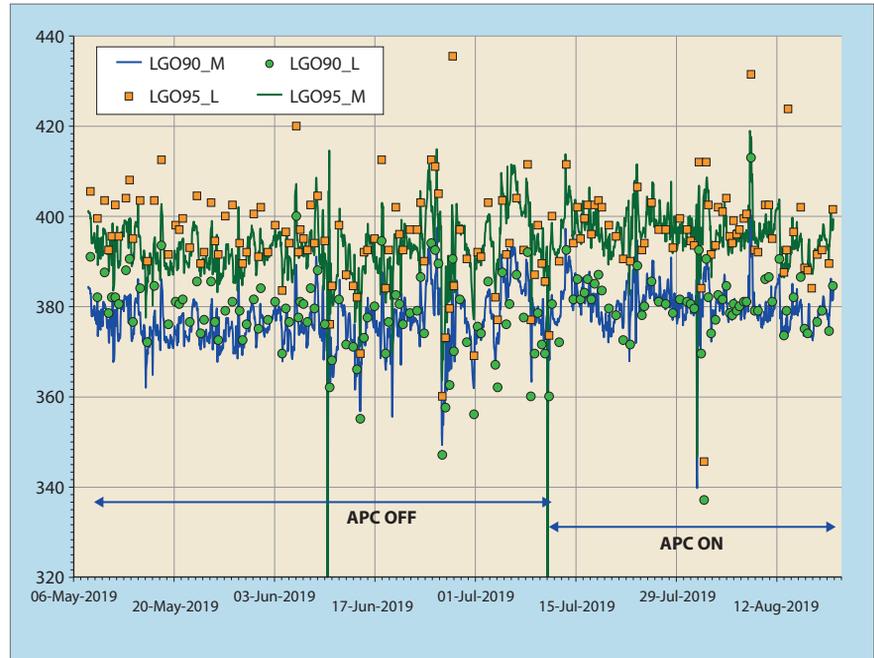


Figure 6 LGO 95% inference vs lab trend

CDU2 ,VDU2 reformer unit, lube complex and delayed coker.

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# Steam reforming catalysts raise production efficiency

The primary reformer is the most energy-intensive element in syngas production. New steam reforming catalysts reduce its energy costs and raise productivity

CHRISTIAN LIBRERA  
*Clariant Catalysts*

The production of synthesis gas (also known as syngas) involves the process of steam reforming, in which steam is used to convert natural gas containing hydrocarbons such as methane and naphtha into hydrogen and carbon based components (CO, CO<sub>2</sub>) in the presence of a nickel catalyst. The process takes place in steam methane reformers which are furnaces where catalysts are loaded in vertical tubes (each tube is considered to be a separate reactor), and externally fired to provide the required heat for the desired endothermic reforming reaction. Burners are positioned on the top, bottom or the sides of the reformer, depending on the plant's process design.

The steam reformer is the heart of a syngas plant – but also the element that consumes the most energy and costs. Its complex structure includes costly materials needed for the furnace and tubes, placing a considerable burden on the capital expenses of a project. Due to the high energy costs of the extreme temperatures required for the reaction the reformer also makes a major contribution to the operating expenses of running a plant.

## Steam reforming catalysts

One critical factor that greatly affects process costs and efficiency is the performance of the reforming catalyst used to boost the chemical reaction in the tubes. High operating temperatures, close to the limits of the reformer tube's material, combined with the endothermic steam reforming reaction, require highly active and stable catalysts to prevent the reformer tubes from overheat-

ing. In addition, the steam methane reformer typically has the highest pressure drop of installed catalyst beds, which can increase costs due to additional natural gas compression, and therefore can limit the throughput of a plant in some cases.

## Catalyst properties

As the catalyst is loaded in a multitude of tubes (up to several hundred), each tube needs to maintain the same pressure drop level during the entire lifetime of the catalyst charge in order to guarantee a satisfactory run for the whole reformer. Otherwise the gas would preferentially flow through tubes with lower pressure drop which can affect the performance of the reformer in terms of higher methane slip or increased tube wall temperatures, causing a restriction in the plant's throughput. Consequently, a catalyst which lowers pressure drop is preferred to reduce syngas compression energy.

The catalysts used for steam reforming are usually supplied in the oxide state (as nickel oxide on a suitable carrier), and then reduced during the plant's start-up since nickel is the desired active phase for the reforming reaction. During every shutdown and start-up process, the tubes undergo thermal cycling, meaning that they may potentially contract and expand to a larger extent than the loaded catalyst particles, applying very high localised forces on the catalysts. Hence the catalyst needs to be physically strong in both the oxidised and reduced state.

Optimal reforming catalysts must also enable a lower methane (CH<sub>4</sub>)

concentration at the tube outlets to maximise production and energy savings. This would not only ensure high hydrogen (H<sub>2</sub>) yield, but also reduce the inert level and purge in the ammonia synthesis loop, thereby reducing energy consumption. Furthermore, catalyst designs that promote lower tube wall temperatures are favoured as they extend tube lifetimes, thus reducing expenses.

## Performance criteria

There are several basic requirements that steam reforming catalysts must satisfy in order to ensure efficient syngas production. One essential feature is high catalytic activity – both intrinsic and apparent. A reforming catalyst's intrinsic activity is influenced by the interaction between its nickel content and carrier, on the nickel dispersion, as well as on the formulation of the carrier material. Apparent activity is determined by the catalyst's diffusion capacity, the volume and size distribution of its pores, and its geometric surface area. Other critical catalyst performance parameters include low pressure drop, efficient heat transfer, high physical strength and stability, as well as resistance to carbon formation.

## Importance of shape

A steam reforming catalyst's form is as important as its formulation. Every shape parameter – such as particle size, aspect ratio of catalyst height to diameter, void spaces in holes and external channels, or packing property between catalyst particles – influences both the geometric surface area per loaded

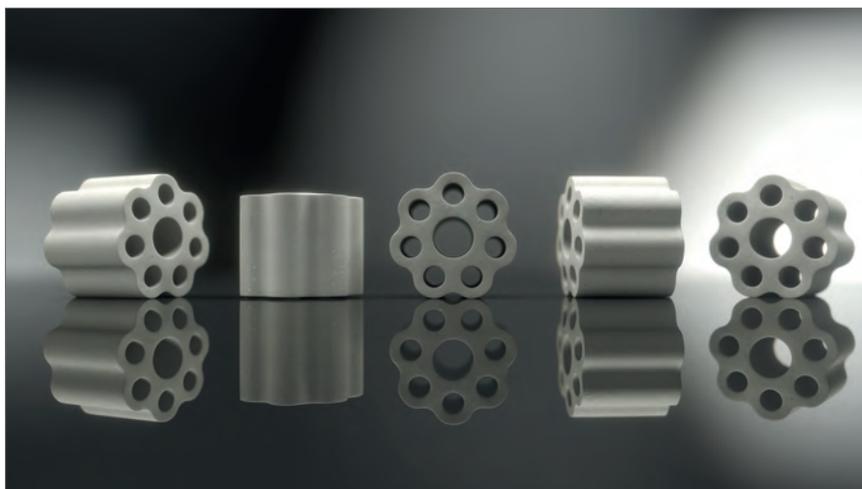


Figure 1 ReforMax 330 LDP Plus next-generation catalyst for steam reforming

volume, and the pressure drop across the catalyst bed. However, it is important that any optimisation during shape development must consider high crush strength as the guiding boundary condition.

### ReforMax LDP series

Nickel-based steam reforming catalysts have been available for several decades. One of the leading examples, focused on ensuring low differential pressure drop, is Clariant's ReforMax 330/210 LDP series. These catalysts are designed with a special 10-hole shape that ensures uniform radial crush strength, and an optimised geometric surface area. Those characteristics directly result in high activity, long catalyst life, and low tube wall temperatures, as well as a stable and minimum methane slip.

Each particle consists of a cal-

cium aluminate based carrier on which the active metal nickel oxide is applied. The carrier chemistry is based on a significant amount of irreversibly formed hiconite ( $\text{CaAl}_2\text{O}_9$ ), which plays a vital role in the mechanical strength of the catalyst particle and allows easy reduction of the impregnated nickel oxide to the active metal. The catalyst's robustness also allows its fast regeneration through steaming in case of catalyst poisoning (carbon formation). The catalysts are commercially proven to be highly efficient under various process feed or design conditions and are currently running in more than 130 ammonia, methanol, and hydrogen reformers worldwide.

### LDP Plus generation

The performance benefits of the 10-hole ReforMax LDP series have

now been further enhanced in a new generation of eight-hole, flower-shaped steam reforming catalysts (see Figure 1). The most striking feature of the new ReforMax LDP Plus series is that it combines high activity with a very low pressure drop, thus helping to further improve plant operations and efficiency. The series consists of ReforMax 330 LDP Plus (non-promoted) and ReforMax 210 LDP Plus (lightly alkalised). The size, aspect ratio and inner channels of these catalysts have been designed to optimise geometric surface area, pressure drop, and crush strength.

### Benefits of Plus series

Thanks to its high geometric surface area, the LDP Plus catalyst series maintains the high activity and longevity of its predecessor. The main difference, however, is that with the new eight-hole floral shape, the void fraction inside and between particles has been significantly increased (see Figure 2), thus reducing pressure drop in the reactor tubes by approximately 20% (see Figure 3). This gives plant operators the possibility of increasing gas throughput by up to 11%, provided there are no other limitations present. Another option for producers is to maintain current gas throughput levels and benefit from energy savings due to the reduced compressor load. Furthermore, the catalysts' large holes support highly efficient heat transfer in the reformer.

As with the former series, the new generation incorporates an uneven distribution of holes which leads to uniform stress distribution, a mechanically strong structure, and a crush strength almost independent of direction of force. Another fine but important design aspect of the catalyst is its domed shape, which helps to prevent the interlocking of catalyst particles, and promotes irregular stacking, both of which also benefit heat and mass transfer. Moreover, ReforMax LDP Plus efficiently suppresses carbon formation, particularly when the non-promoted (330) and lightly alkalised (210) forms are combined. Overall, the new catalysts can substantially reduce energy consumption in the production of synthesis gas.

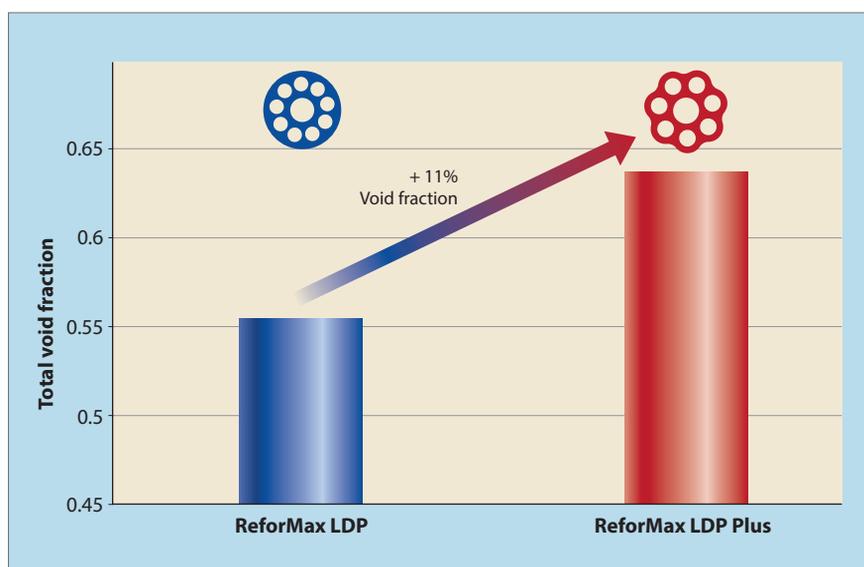


Figure 2 Enhanced void fraction



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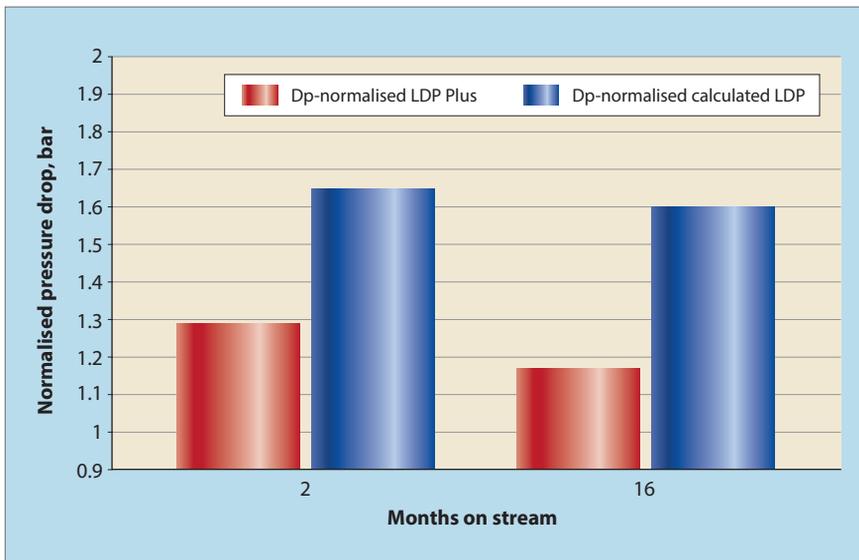


Figure 3 Approximately 20% decrease in pressure drop

### Monitoring tube wall temperature

Besides catalyst performance, the second aspect that greatly impacts process efficiency is internal heat distribution among the reformer tubes. The tube wall temperature must be carefully monitored to ensure even and consistent heat distribution. Overly cautious firing reduces the reaction rate and leads to lower production and/or higher energy consumption. Uneven firing results in substandard performance and can cause overheating of tubes (hot spots), which will drastically reduce tube lifetime if the maximum tube wall temperature is exceeded. Mechanical stress on the hot tubes increases exponentially, in worst cases causing tube ruptures. Another reason to observe the tube

wall temperature is that it provides information about the catalyst's performance.

Clariant offers various technologies for monitoring the tube wall temperature. The simplest of these is the pyrometer, which allows reliable single-point readings of the tube wall temperature. As it is designed for ease of use, pyrometer measurements are limited to tubes which can be accessed through the peepholes. In addition, readings should be corrected for background radiation.

A more advanced solution is the Thermal Scan, a high resolution thermal imaging borescope which can measure temperatures from 600-1100°C (1112-2012°F) at a short, accuracy-enhancing wavelength. The instrument presents a real-time,

three-dimensional heat profile of all tube rows in the furnace, allowing easy detection of uneven heat distribution or hot spots (see Figure 4). Plant operators benefit from a comprehensive temperature survey, which includes thermal images as well as recommendations for burner adjustments.

Alternatively, the Thermal Scan Plus Gold Cup has been developed for measurements using a golden reflector. This method provides superior precision due to its near-perfect reflectance. The data is used to calibrate the standard Thermal Scan, thus providing a very accurate temperature profile of the entire furnace ( $dT < \pm 5^{\circ}\text{C}$ ,  $41^{\circ}\text{F}$ ). These instruments enable more effective troubleshooting, and help operators achieve maximum reformer performance and lifespan, without unnecessary and costly outages.

### Case 1: eliminating pressure drop limitations

The first commercial reference for the ReforMax 330 LDP Plus catalyst is a leading European ammonia producer with a capacity of 1650 metric t/d at its world-scale plant. Prior to changing to the new catalyst, the plant was using ReforMax LDP catalysts in a side-fired furnace. Although performance was satisfactory, pressure drop over the front end was a crucial production limitation. The installation of the new ReforMax 330 LDP Plus catalyst and optimisation of catalyst volumes in other reactors have removed this limitation, significantly increasing the plant's energy and production efficiency.

Since its start-up in mid-2018, ReforMax 330 LDP Plus has demonstrated very stable operation, and provided approximately 17% reduction in pressure drop across the catalyst in the reformer tubes. This improvement will enable the plant to realise expected cost savings in excess of \$340 000 over the expected catalyst lifetime of eight years, compensating for the catalyst investment. It also provides the ammonia producer with the opportunity to increase throughput and hence its production capacity. The customer

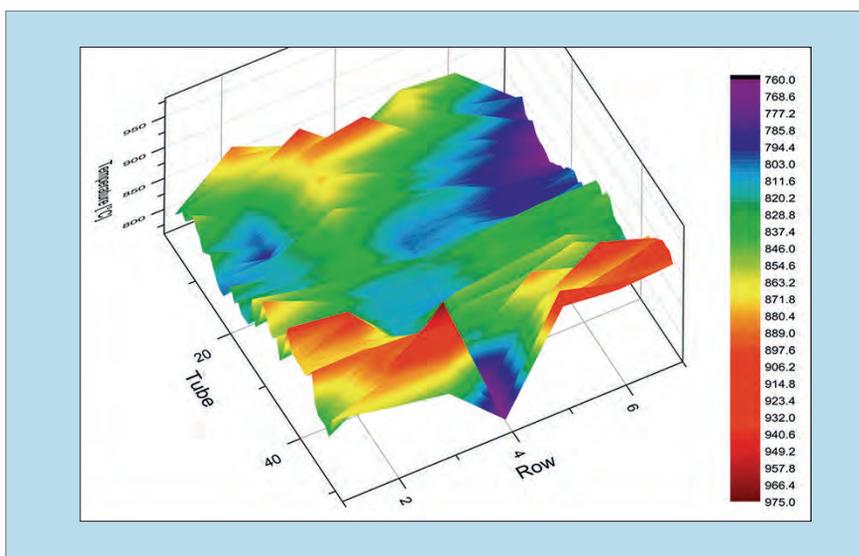


Figure 4 Real-time, three-dimensional heat profile of entire furnace

reports high satisfaction and recommends ReforMax 330 LDP Plus for all plants experiencing pressure drop limitations in the primary reformer.

### Case 2: successful recovery after severe coking

More recently, another major European ammonia producer, also facing pressure drop limitations in its primary reformer, elected to switch to ReforMax 330 LDP Plus to debottleneck its production, and increase ammonia yield. However, due to a malfunction of the flow meter during catalyst start-up, ReforMax 330 LDP Plus was operated with an exceedingly low steam-to-carbon ratio of 1.5 for several hours, which resulted in severe coking of the catalyst.

Once the ammonia producer reported the incident, Clariant advised steaming the catalyst for a minimum of 10 hours in order to remove the formed carbon as far as possible. Subsequently, a tube wall temperature survey of the reformer confirmed that the maximum per-

missible tube wall temperature of 940°C (1724°F) was not exceeded. Thanks to the corrective measures taken, the producer is now safely running the plant at full capacity and can continue to do so until the next scheduled turnaround. Furthermore, ReforMax 330 LDP Plus is operating at more than 16% lower pressure drop compared to the previous catalyst under the conditions prevailing in the producer's plant.

To further improve catalytic performance, another steaming of the catalyst was later executed to remove all remaining coke. Despite the severe coking, complete recovery of the catalyst was achieved, demonstrating the high robustness of the new catalyst. ReforMax 330 LDP Plus is expected to reach its guaranteed lifetime and to provide the producer with the desired pressure drop improvements.

### Conclusion

The steam reformer is the costliest link in the syngas production chain, thus it makes sense to focus on max-

imising its efficiency. ReforMax LDP Plus presents an opportunity to achieve this. Due to the pressure drop decrease of up to 20% enabled by the new catalyst, a typical ammonia plant with a capacity of 1500 t/d could save up to \$300 000 over the typical catalyst lifetime. Alternatively, a plant experiencing a bottleneck due to pressure drop can now benefit from up to 11% higher gas throughput, and considerably increase its production yield. As demonstrated in the cases presented, innovative catalyst technologies and thermal imaging possibilities can add significant value to ammonia, hydrogen, and methanol production processes.

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# Oil-to-chemicals: new approaches

## A review of developments and trends in the expanding business of oil-to-chemicals

JOHN J MURPHY and CLYDE F PAYN  
*The Catalyst Group*

Crude oil-to-chemicals (COTC) continues to be a powerful industry driver and a strong trend of high interest to all integrated refineries and chemicals producers in Asia/Pacific, China, the Middle East, and Eastern Europe. This is reinforced by many factors, most notably the forecasts which predict a slowing of transportation fuels growth approaching 2040 (with hybrids and electric vehicles), while growth in chemicals is expected to increase as populations and middle class wealth continue to rise, leading to increasing demand for packaging, consumer goods, and automobiles.

Are you aware that more than 12 corporations have committed over \$315 billion to date to reconfigure their assets to produce more petrochemicals than transportation fuels, as revamps as well as building new grassroots refineries during the next 5-6 years? Based on announcements to date, we anticipate in the next five years that another \$300+ billion, or more, will be announced as refiners and chemical companies all reassess their positions, knowing that the longer term outlook for transportation fuels from crude oil is expected to plateau and then decline. All players are taking this trend seriously and therefore you should also.

Considerable flexibility is being offered by petrochemical licensors, in particular petrochemical resid and VGO FCC upgrading units today. These are global changes including deep catalytic cracking (DCC) from Sinopec, as well as Western leaders such as Total's R2R modifications, and Axens' high-severity FCC (HSFCC) with Saudi Aramco. Technologies do not stand still. Advances in catalytic visbreaking may also be important in the future, when looking into advanced

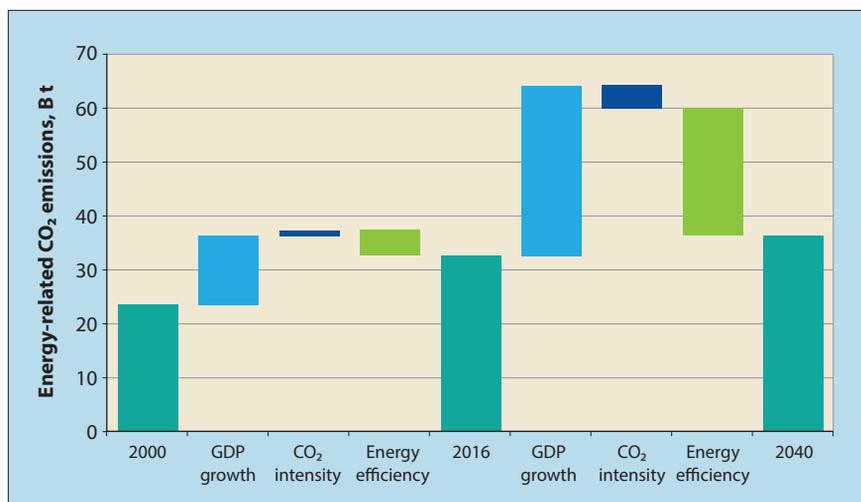


Figure 1 Energy efficiency gains are expected to nearly double by 2040, while carbon emissions are projected to increase by a modest 10%<sup>1</sup>

lower cost alternatives, and we have examined these R&D pipelines.

Already a large number of companies are closely examining their own responses and investments, bearing in mind each of these investment objectives will be site specific, influenced by feedstock choices, product slates/markets, energy/utility balances, capital/operating efficiencies, and health, safety and environmental (HSE) performance. It is clear from public domain information (such as the ongoing announcements by ADNOC, MOL and others) to see the progress in differentiation that is already under way.

Two main interests of producers are: to decrease capital intensity through scale, simplicity, and location; and to expand/maximise flexibility towards use of current (heavier) feedstocks in considering the oil-to-chemicals approach. The idea of better utilising assets from within an integrated refinery site means that most likely you are already dealing at 10x plus the size of a world-scale petrochemical plant. Although scale counts,

it is also only one of many factors. New advanced configurations will now start to incorporate the planning of improved efficiency gains and reduced CO<sub>2</sub> emissions. ExxonMobil forecasts that by 2040, while energy efficiency gains are expected to nearly double, carbon emissions are only projected to increase by a modest 10%.<sup>1</sup> BP statistics, along with Chevron forecasts, the IEA and the EIA, show similar trends (see Figure 1).

Regarding competitive crude oil-to-chemicals developments, in addition to Saudi Aramco/SABIC announcements, we are already seeing ongoing investments from others. In a more recent example, private chemical producers Hengli and Rongshengin in China are back-integrating their chemical plants to add over 9 million t/y of paraxylene capacity by 2021. This is expected to reduce imports by 4 million t/y, with plans to yield up to 45 wt% of chemicals processing heavy crudes, which will tighten medium to heavy crude markets while also adding a 40% surplus to distillates and gasoline markets.

One of the most difficult components has been to understand that all licensors need to prioritise their own businesses. Therefore, they will prefer greenfield investments to revamps – even if these can be accomplished at lower ISBL and OSBL costs. This is not a criticism but rather a statement of fact based on desired business focus. Moreover, one of the understandings is to appreciate how existing and new configurations can be tailored towards either aromatics or olefins – but this may not be the best measure if indeed your goal is towards more olefins. In this regard, assuming you have an existing steam cracker, your revamp approach may be quite different.

### Advances in heavy oil processes

In focusing on the processes by which the higher molecular weight constituents of petroleum (the heavy ends) can be converted to products that are suitable for use as feedstocks for the petrochemical section of the refinery, our assessments include carbon rejection and hydrogen addition approaches, along with process combinations and new configurations:

1. Carbon rejection
2. Hydrogen addition
3. Combining processes and treatment of intermediates
4. Configuration issues and advances
5. New processes likely to be deployed during the next five years

For decades, propane has been the mainstay in deasphalting heavy feedstocks, especially in the preparation of high quality lubricating oils and feedstocks for catalytic cracking units. Future units, which may well be derived from KBR's ROSE process, will use solvent systems that will allow operation at elevated temperatures relative to conventional propane deasphalting temperatures, thereby permitting easy heat exchange. This will require changes to the solvent composition and the inclusion of solvents not usually considered to be deasphalting solvents. Other areas of future process modification will be in extractor tower internals, studies with higher molecular weight solvent, accurate estimation of physical properties of mix stream, studies in combination

with other processes, and firming up design tools for supercritical solvent recovery configurations.

For heavy feedstocks, which will increase in amounts as hydrocracking feedstocks, reactor designs will continue to focus on online catalyst addition and withdrawal. Fixed bed designs have suffered from mechanical inadequacy when used for the heavier feedstocks, as well as short catalyst lives – six months or less – even though large catalyst volumes are used (LHSV typically of 0.5-1.5). Refiners will attempt to overcome these shortcomings through innovative designs, allowing better feedstock flow and catalyst utilisation, or online catalyst removal. For example, the OCR process, in which a lead moving bed reactor is used to demetallise the heavy feedstock ahead of the fixed bed hydrocracking reactors, has seen some success. But whether this will be adequate for continuous hydrocracking of heavy feedstocks remains a question.

Catalyst development will be key in the modification of processes and the development of new ones to make environmentally acceptable distillable liquids. Although crude oil conversion is expected to remain the principal future source of petrochemicals, natural gas reserves are emerging, and will continue to emerge, as a major hydrocarbon resource. This trend has already started to result in a shift toward use of natural gas (methane) as a significant feedstock for chemicals. As a result, deployment of technology for direct and indirect conversion of methane will probably displace much of the current production of liquefied natural gas.

The detrimental effect of coke on catalyst is a reduction of support porosity, leading to diffusional limitations, and finally blocked access to active sites. Nevertheless, moving bed or ebullated bed processes, alone or in combination with fixed bed reactor technology and/or also coupled with thermal processes employing suitable catalyst with metal retention capacity, represent the most efficient way of handling petroleum bottoms and other heavy hydrocarbons for upgrading. The features of the resulting process con-

figuration will be high liquid yields, high removal of contaminants, and reliable operation.

### Holistic economics and approaches to complexes

From a comprehensive or holistic perspective, the following approaches have been assessed as commercially viable or considered to become commercially viable in specific situations:

- New pipeline technology
- Advances in new configurations
- New catalyst approaches
- Economics of different catalysts and process improvements

### Competitive and strategic implications

In reviewing some of the key findings from our report (*The Catalyst Group Resources 2019*),<sup>2</sup> as well as the limits of current state-of-the-art based on the basket of crudes defined in the report, here are some key considerations:

- No study can take into account all possible site-specific issues and questions, as they may relate to existing configurations for revamp vs greenfield choices because they are highly dependent on each refinery's crude slates, availability/pricing, and the local/regional products desired. Given this situation, the study takes a 10 000ft view, looking into the hypothesis of a 50/50 fuels/petrochemicals refinery, and then discusses future technology options/changes in the pipeline in the direction for 40/60 fuels/petrochemicals.
- Today's resid FCCs (RFCC) can process feeds with up to 8 Concarbon, though 6-7 is more comfortable. Today's RFCCs are designed for catalyst metals levels of 10 000 wtppm. However, it is cheaper to take the metals out on an HDM pretreater catalyst which holds up to 50% of their weight in metals. A standard design is to include an extra riser for making olefins. A 100 000 b/d RFCC can make over 500 000 t/y of propylene, assuming a 10 wt% yield. Additional technologies can increase this to 30-40 wt%. For instance, VGO processing with an HDM/HDS unit can give around 29 wt% propylene. The FCC gasoline, which is about 50 wt% BTX, can also



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### Announced oil-to-chemicals investments 2019, \$billion

Zhejiang Petroleum and Chemical	Zhoushan, China	\$26	Greenfield	2019 (Phase 1)
Hengli Petrochemical	Changxin Island, China	\$11	Greenfield	2019
Shenghong Petrochemical	Lianyungang, China	\$11.84	Greenfield	2019
Ningbo Zhongjin Petrochemical (subs Rongsheng Petrochemical)	Ningbo, China	\$5 (est)	Revamp	2018
Saudi Aramco/NORINCO/Panjun Sincen (Huajin Aramco Petrochemical)	Liaoning Province, China	\$10+	Greenfield	2024
SABIC/Fuhaichuang Petrochemical	Zhangzhou, China	NA	Greenfield	NA
SINOPEC/SABIC (Tianjin Petrochemical)	Tianjin, China	\$45	Revamp	Operating, pre-2017
PetroChina	Dalian, China	combined	Revamp	Operating, pre-2017
PetroChina	Yunnan, China	(est)	Revamp	Operating, pre-2017
CNOOC	Huizhou, China		Revamp	Operating, pre-2017
SINOPEC	Lianyungang, China	\$2.80	Greenfield	NA
SINOPEC	Caofeidian, China	\$4.2	Greenfield	NA
SINOPEC	Gulei, China	\$4.26	Greenfield	2020
	<b>Total China</b>	<b>\$120.1</b>		
<b>Other Asia</b>				
Hengyi Group	Pulau Muara Besar, Brunei	\$20	Greenfield	2020
Saudi Aramco/ADNOC/India Consortium	Raigad, India	\$44	Greenfield	2025
Petronas/Saudi Aramco (RAPID)	Pengerang, Malaysia	\$2.7	Greenfield	2019
ExxonMobil (Singapore Chemical Plant)	Jurong Island, Singapore	<\$1	Revamp	2023
Pertamina/Rosneft	Tuban, East Java, Indonesia	\$15	Greenfield	2025
	<b>Total other Asia</b>	<b>\$82.7</b>		
<b>Middle East</b>				
ADNOC	Al Ruwais, UAE	\$45	Revamp	2025
Saudi Aramco/SABIC	Yanbu, Saudi Arabia	\$30	Greenfield	2025
Saudi Aramco/Total	Jubail, Saudi Arabia	\$5	Greenfield	2024
KNPC/KIPIC (Al-Zour Refinery)	Al Ahmadi, Kuwait	\$13	Greenfield	2019
Oman Oil Company/Kuwait Petroleum International (Duqm Refinery)	Oman	\$15	Greenfield	NA
	<b>Total Middle East</b>	<b>\$108</b>		
<b>Europe</b>				
MOL Group	Hungary, Croatia	\$4.5	Revamp	2030
	<b>Total Europe:</b>	<b>\$4.5</b>		
<b>Total Greenfield \$215</b>			<b>Total revamps \$100</b>	<b>Total global \$315</b>

**Table 1**

Source: TCGR 2019

be partially processed in the aromatics plant. Fine tuning in the RFCC for propylene is a lot less costly than propane dehydrogenation.

- When processing heavier feedstocks, the consensus is to have hydrogen-in revamps or greenfield designs.

- Increasing the severity of RDS/RFCC to produce more propylene decreases both gasoline and diesel yield. Forwarding heavy naphtha is required for reformate feed to aromatics. Improving liquid yields can be done to different degrees by upping VGO+DAO, while reducing coke to almost zero.

- Smaller (100 000 b/d) refineries will not be as likely to have the capital to integrate like >250 000 b/d and larger sites.

- All licensors, by their reprints, will try to sell complex greenfield site configurations based on their competitive advantages. Others have different levels of revamp expertise. When we use examples throughout our analyses, they are to highlight real world examples.

Critical to an assessment of the potential for oil-to-chemicals is the number and types of committed investments to date (mid-2019). This study documents those announced investments declared during the last five years as oil-to-chemicals projects, along with company, location, size of project, and investment. Where available and announced, we have also included the wt% fuel vs chemical targets (see **Table 1**). These all have been more closely researched, with sources and notes provided. What it does highlight is there is at least \$315 billion in already committed investment, of which \$100 billion is in revamps, \$120 billion in China, \$82.7 billion in Asia/Pacific, and \$108 billion in the Middle East.

There are project examples where these considerations have already been reviewed. For example, MOL Petrochemicals in Tiszaujvoros, Hungary, has decided to upgrade its 100 000 t/y to produce more polymer grade propylene from steam cracking and refinery feedstocks. In

doing so it has chosen two steps, utilising Lummus OCT and a CDHydro Deisobutenizer which will generate an isobutene-rich stream, whereas OCT will generate increased propylene production. These modifications are reportedly available for less than \$50 million. Also, the MOL revamp is interesting as the company intends to incorporate Innovacat swing fixed bed technology in the refinery.

Another example we highlight, which we think stands out with some interesting conclusions, is the revamp for the Polish refiner Grupa LOTOS when, in 2011, it installed and made operational a new generation of DAO hydrocracking technology as part of a major resid upgrading project called the 10+ Programme. In this case, it raised refining capacity by 75%, focused on higher margin diesel fuels to increase market share, and enhanced margins by \$5 per barrel.

In this case, the two units added by Shell Global Solutions were a 45 000 b/d DAO hydrocracker using 50/50 VGO/DAO straight off these units,

with the added DAO unit. What is interesting is that, using Urals feedstock, the hydrocrackers, inclusive of HDM, HDS and HDN, were able to increase conversion to 85% from 60% with a recycle mode.

Based on the information from these examples and assuming the VDU and ADU are already in-place investments, then an SDA unit (KBR ROSE, Axens Hyvahl or Selex-Asp), depending on the product slate chosen, is a considered first step at a lower approximate cost of \$250-280 million.

### Conclusion

In summary, there has been a long history (decades in fact) of incremental developments leading to what can be described as 'oil-to-chemicals'. For a long period, building larger and larger world scale and more complex refineries and steam cracking plants was the economic solution best suited to the fundamentals of medium to heavy crude oil conversion and, in some countries, this will still be the case. However, today we have entered a different era, where

the socio-economic as well as supply/demand trends are shifting, and the traditional business models of segregated refining vs chemicals production no longer hold true.

The ongoing drive for improved profitability profiles, derived by producing petrochemicals as opposed to fuels, has justified the increased pace of the oil-to-chemicals movement. Not only are demands for olefins and aromatics growing more quickly than gasoline and diesel, the profit margins for these petrochemicals are also higher, and even more so when made directly via oil-to-chemicals conversion routes.

Although moderate to date, the commitments to these plant configurations will require retrofits as well as new capex using skilled labourers and EPCs. As the former are limited and the latter is notoriously cyclical (as is the energy/fuels industry), it is important to assess how large and when these events will impact the availability of (and price for) skilled labour. To what degree will envisioned projects be delayed or their price increase as a result?

The two oil-to-chemicals approaches – carbon out and hydrogen in – have implications across related technologies. Coking will remain a 'go-to' for carbon out with any advances having outsized impacts (due to the breath of implementation). Hydrogen supplies will need to increase or become more flexible (without additional energy/CO<sub>2</sub> impacts) in order to address the range of upgrading requirements.

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1 ExxonMobil, *Outlook for Energy, 2018*, <https://corporate.exxonmobil.com/energy-and-environment/energy-resources/outlook-for-energy>

2 The Catalyst Group Resources (TCGR), *Oil-to-Chemicals II: New Approaches from Resid and VGOs*, 2019, [www.catalystgrp.com/multiclient\\_studies/oil-chemicals-ii-new-approaches-resid-vgos](http://www.catalystgrp.com/multiclient_studies/oil-chemicals-ii-new-approaches-resid-vgos)

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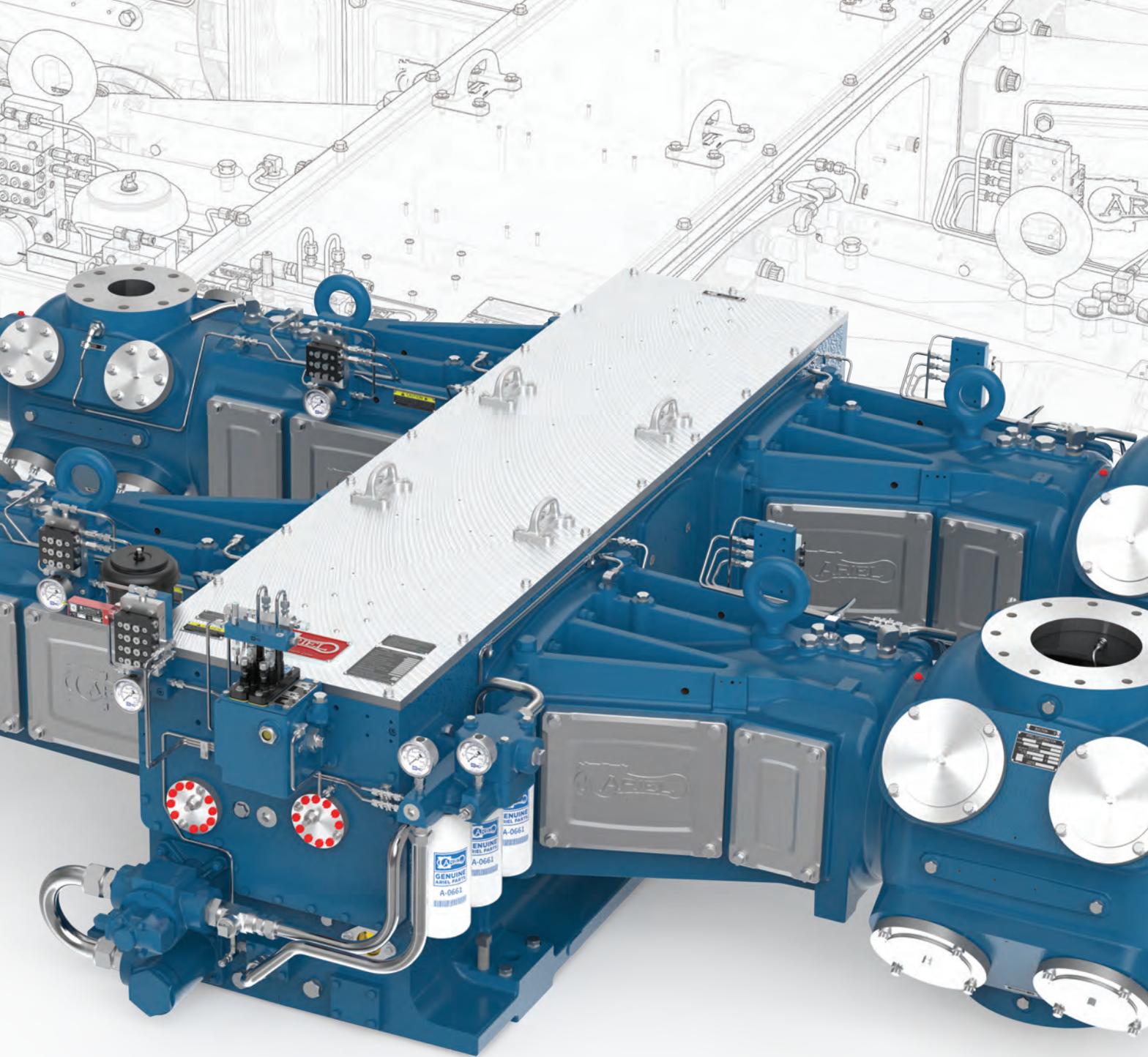
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# Model predictive control in a lube oil complex

## An application of model predictive control targeted optimum performance from a refinery's propane deasphalting unit

HANDAN ÇEVİK ŞANLI and BERKAY ER  
Tüpraş

**M**odel predictive control (MPC) refers to a class of computer control algorithms that utilise an explicit process model to predict the future response of a plant. At each control interval, an MPC algorithm attempts to optimise future plant behaviour over a time interval known as the prediction horizon by computing a sequence of future manipulated variable adjustments. It was originally developed to meet the specialised control needs of petroleum refineries and power plants. MPC technology can now be found in a wide variety of application areas including chemicals, automotive, food processing, and aerospace applications.<sup>1</sup>

Turkish Petroleum Refineries Corporation (Tupras) is the largest industrial corporation in Turkey with a refining capacity of 30 million t/y crude oil in four refineries. İzmir refinery is the largest refinery in terms of annual processing capacity, with 11.9 million tonnes of crude oil.

Advanced process control (APC) projects began at Tupras in 2006 with a third generation MPC technology, the Shell Multivariable Optimizing Controller (SMOC) algorithm, to create a sustainable and developable process control approach in the refinery. APC applications were commissioned on multiple units at İzmir refinery. With technological advances in MPC algorithms, Tupras decided to continue with fourth generation MPC technology, Robust Multivariable Predictive Control Technology (RMPCT), in 2017. RMPCT was supplied by Honeywell, and PCT was supplied by Profimatics. Both are third generation MPC technology. When Profimatics was purchased by Honeywell, the two algorithms were merged and RMPCT was born.

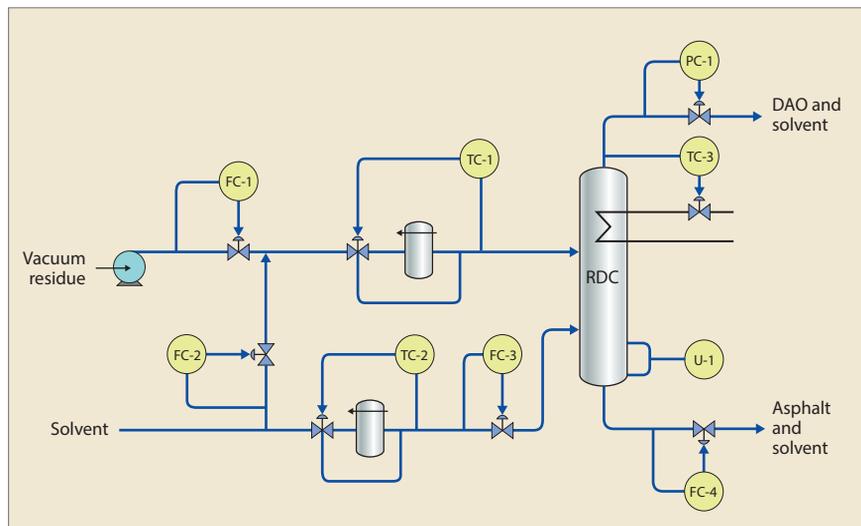


Figure 1 RDC section

The commercial name for RMPCT is Profit Controller. In this article, an application of Profit Controller on a propane deasphalting unit in the lube oil complex of İzmir refinery is presented. A process overview and base layer control strategies are presented, followed by discussion of controller scope, variables, inferentials, response testing, and dynamic modelling. The control algorithm, economic variables, commissioning and tuning, and results are discussed in the context of controller commissioning.

### Process overview

A propane deasphalting unit located in İzmir refinery's lube oil complex has the main task of producing high quality bright stock by removing asphaltic hydrocarbons from vacuum residue.

### Extraction column section

Vacuum residue from the lubes vacuum unit is pumped to the propane deasphalting unit as feed (see Figure 1). The unit input is controlled by FC-1. Prior to entering

the cooler exchanger, a small stream of propane is added to the feed. The propane is a viscosity cutter to permit effective heat transfer and avoid plugging of the exchanger. The mixed stream flows through the exchanger to the middle of the rotating disc contactor (RDC) column. The feed temperature is controlled by TC-1, which manipulates the opening of the exchanger bypass.

Propane solvent is pumped from the propane receiver drum. Some solvent is used to dilute the feed while the main part of it goes to the solvent cooler where it is brought to the desired temperature by TC-2. The flow controller FC-2 determines the amount of solvent for dilution, and solvent entering the bottom part of the RDC is determined by FC-3. Solvent entering the bottom part of the column flows through the bottom section of the RDC column. As this happens, the oil fractions of the residue are dissolved, leaving heavy asphalt to settle at the bottom while the oil is carried in solution through the top of the RDC column.

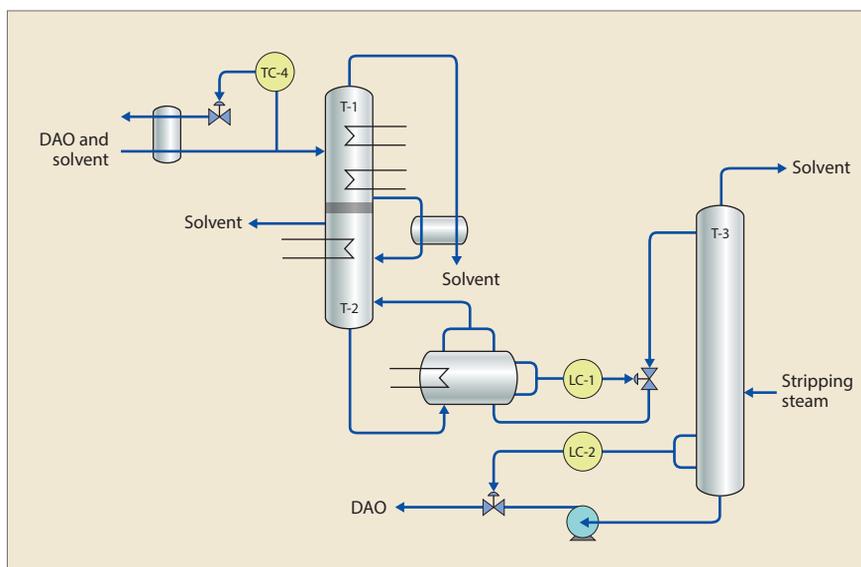


Figure 2 Deasphalted oil/solvent recovery section

Deasphalted oil leaves the top of the RDC column with the bulk of the solvent and flows to the deasphalted oil solvent recovery system. Asphalt mixture from the bottom of the RDC flows to the asphalt recovery section.

### Deasphalted oil recovery section

The deasphalted oil mix firstly flows to the low pressure steam evaporator, where part of the solvent is vaporised before flowing to the flash tower T-1 (see Figure 2). The inlet temperature of T-1 is controlled by TC-4. Hot vapour from the flash is condensed in the heat exchanger. The condensed solvent flows to the high pressure propane receiver drum. The deasphalted oil mix from T-1 flows to the second effect deasphalted oil flash tower T-2, where the vaporised solvent separates from the liquid. This liquid then flows to the steam reboiler, where still more solvent is evap-

orated. The deasphalted oil mix from T-2 flows to the deasphalted oil stripper T-3, where superheated steam removes the last traces of solvent from the oil. Deasphalted oil from the bottom of the stripper is pumped to the storage section.

### Asphalt recovery section

Asphalt mix from the bottom of the RDC column flows to the asphalt mix furnace, where most of the solvent is vaporised. The furnace outlet temperature is controlled by TC-5, which resets fuel pressure controller PC-2. The partially vaporised asphalt mix then flows to asphalt tower T-4. Asphalt mix from the bottom of the flash flows to the asphalt stripper T-5, where it is stripped of residual solvent with superheated steam. The stripped asphalt is pumped from the bottom of the stripper to the storage section.

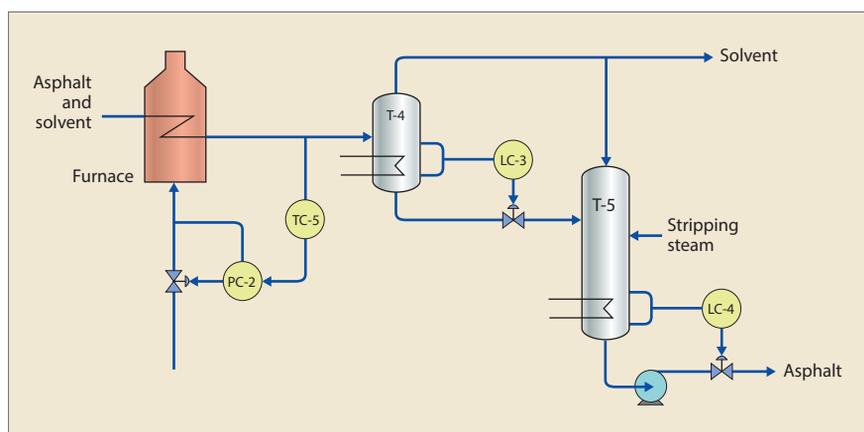


Figure 3 Asphalt/solvent recovery section

### Base layer control issues

Before implementation of the profit controller, a base layer control study including instrumentation checks, PID tuning optimisations, and required process control configuration changes was completed.

The base layer control scheme of the RDC column is shown in Figure 4. Critical control issues for the RDC column are discussed in the following.

### Dilution to feed ratio

A small stream of the propane solvent is added to the feed prior to entering feed exchanger 1. This propane is a viscosity cutter to permit effective heat transfer in the exchanger and to avoid plugging of the exchanger. If the feed throughput has been changed then the amount of this flow should proportionally change. Too small a value can lead to plugging and too large a value increases energy consumption. RIC-1 ratio controller is a solution for this control issue; it calculates the set point for dilution flow FC-2 such that desired dilution to feed ratio is achieved. The time filter constant for the ratio controller is configured in such a way that a change in the feed flow does not affect solvent flow too quickly. This also avoids disturbances in the propane vessel.

### RDC bottom interface level

According to the initial design of level control for the RDC column, the operator adjusts the level by manipulating bottom flow. This could lead to unsafe operation if the operator does not act on a level alarm; it also disturbs the furnace outlet temperature. A solution to this problem is a base layer control change; the level controller is configured as a master controller, and the bottom flow controller is configured as a manipulated slave controller. Then the level controller is tuned slowly.

### Solvent to feed ratio

If the feed quality is constant and the feed throughput has to be changed, one of the main manipulated variables for this adjustment is solvent flow. Too high a value could lead to off-spec quality and increased energy

consumption, while too low a value could also lead to off-spec quality. RIC-2 ratio controller is a solution to this control issue; it calculates the set point of total solvent flow FC-3 such that the desired solvent to feed ratio is achieved. The time filter constant for the ratio controller is configured in such a way that a change in the feed flow does not affect solvent flow too quickly, to avoid disturbances in the propane vessel.

### Controller design

The MPC controller design included the following steps:

- Feasibility study
- Understand the process and define the process control objectives
- Determine related controlled, manipulated, and disturbance variables to create a seed model matrix
- Apply a preliminary plant test
- Check base layer issues and optimise PID tunings
- Design an application and final test plan to obtain mathematical models
- Apply step tests systematically for all MVs and DVs, follow the CVs response, and store the real time data
- Obtain dynamic models by using plant test data and an identification package
- Configure the MPC controller by entering default initial tuning settings
- Test the MPC controller offline in closed-loop simulation mode, and monitor controller performance
- Load the configured controller to the APC server
- Test model predictions in open-loop mode
- Take the MPC controller online and finalise tuning settings
- Troubleshoot
- Operator and engineer training

The project's steps are summarised in Figure 5.

### Controller scope

Profit Controller operates in the RDC column section of the propane deasphalting unit. The process control objective for the RDC column is to steer manipulated variables in such a way that all process safety constraints are obeyed and operational targets are achieved. The general control objectives of the controller are:

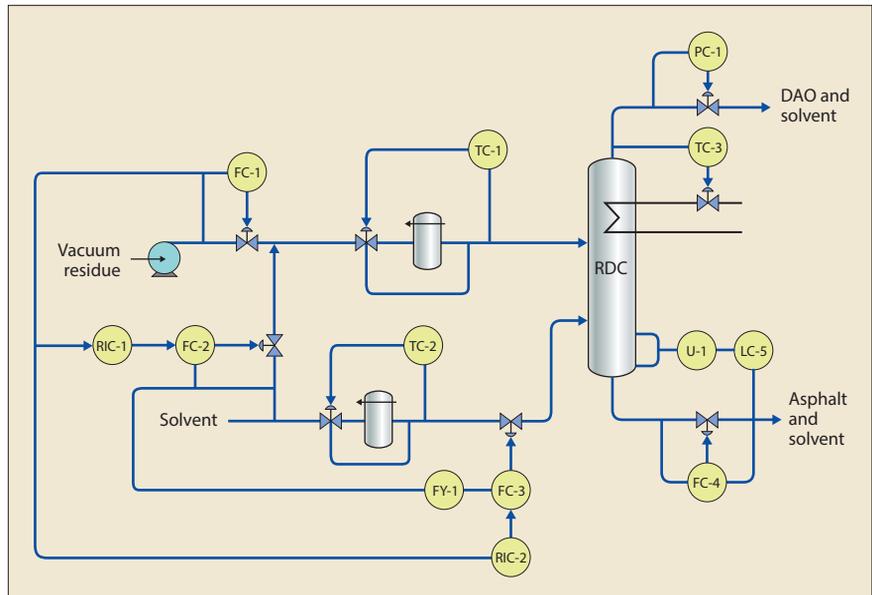


Figure 4 Base layer control scheme for the RDC column

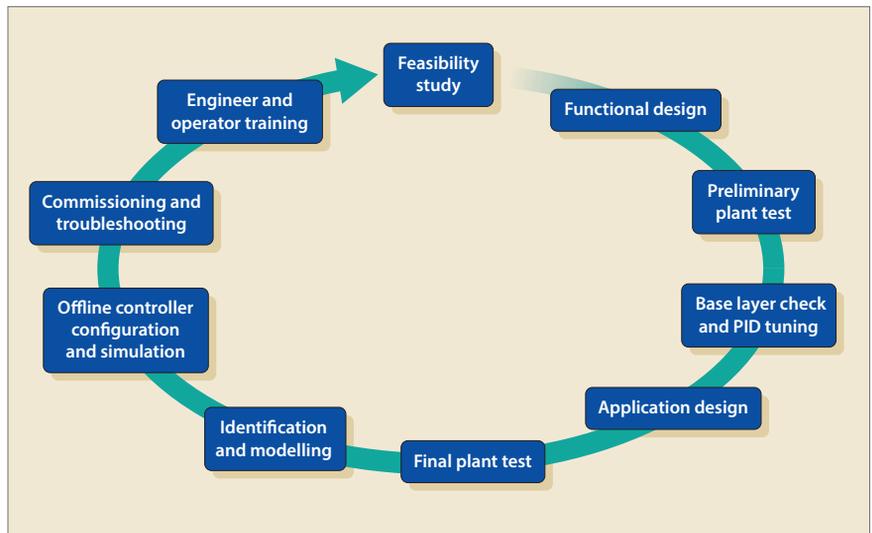


Figure 5 APC project stages

- RDC column top temperature
- Bright stock (DAO) viscosity
- Yield
- Solvent flow
- Feed flow
- Solvent/feed ratio
- Solvent drum pressure

### Variables

Variables are organised as categories in Profit Controller terminology:

Manipulated variables		
MV		Description
1	TC-3.SP	RDC top temperature
2	FC-2.SP	Dilution flow
3	FC-1.SP	Feed flow
4	FC-3.SP	Solvent flow

Table 1

manipulated variables (MVs), controlled variables (CVs) and disturbance variables (DVs).

Manipulated variables for the RDC column Profit Controller are shown in Table 1.

All MVs are configured as critical parameters which means that when one of the MVs is unavailable, the Profit Controller will turn off. The amount of feed is determined by a planning instruction and there is no need for feed pushing. However, the feed will be a manipulated variable in order to capture the transient effect caused by changes in feed. In the RDC column, while the feed is constant, the amount of solvent and the top temperature can be manipulated to change unit throughput. In the profit suite operator station view,

Controlled variables	
CV	Description
1 TC-3.PV	RDC top temperature
2 Bright viscosity	Bright viscosity
3 Yield	DAO yield
4 FC-1.PV	Feed flow
5 Total solvent	Total solvent flow
6 Solvent/feed ratio	Solvent to feed ratio
7 DAO viscosity	DAO viscosity
8 PC-2.PV	Solvent drum pressure
9 FC-1.OP	Feed flow valve
10 TC-3.OP	RDC top temp. valve

Table 2

the MV for which the CV is moved, and causes of the MV move, can be followed.

Controlled variables for the RDC column Profit Controller are shown in Table 2.

When establishing the control strategy, it is necessary to maintain column operation within safety limits as a priority, in which case DAO is produced within the planning viscosity instructions and yield maximisation will be achieved.

A disturbance variable for the RDC column Profit Controller is shown in Table 3.

### Inferential design

Two inferentials were developed for the viscosity of deasphalted oil: bright viscosity and DAO viscosity. The viscosity analyser is a model input for the bright viscosity estimator which is calibrated by laboratory sample. In the event of failure to use the analyser, the DAO viscosity estimator model, which is also calibrated by laboratory sample, can be used.

### Bright viscosity

Bright viscosity at 100°C is correlated to the viscosity analyser. The following relationship describes the estimator model:

Disturbance variable	
DV	Description
1 TC-2.PV	Solvent temperature

Table 3

$$V(\text{bright stock})cSt = 1.0668 * (\text{viscosity analyser}) + B$$

### DAO viscosity

The DAO viscosity at 100°C is correlated to the RDC column top temperature, the temperature difference between the RDC column top temperature and the mid-zone temperature, and the solvent to feed ratio. The following relationship describes this estimator model:

$$V(\text{DAO})cSt = -1.1894 * (\text{RDC Top Temp}) + 0.30193 * \text{DT} + 0.91977 * \text{S/F Ratio} + B$$

DT is calculated by:

$$\text{DT} = \text{RDC Top Temperature} - \text{RDC Mid Zone Temperature}$$

S/F, solvent to feed ratio, is calculated by:

$$\text{SF} = \frac{\text{Initial Dilution Solv} + \text{Solv to RDC}(1) + \text{Solv. to RDC}(2)}{\text{Feed Flow}}$$

### Response testing

Closed-loop identification methods require excitation signals to be applied in the process inputs such that they produce changes in the process outputs. The resulting additional variances in the outputs are subject to a trade-off. They must be big enough to produce persistent excitation for identification, but also to disturb the normal operation of the process as little as possible.<sup>2</sup>

For a propane deasphalting unit RMPCT application, the step size of manipulated variables is shown in Table 4.

MV step sizes	
Description	Step size
Column top temperature	4.5 °C
Solvent amount (in feed)	3 m <sup>3</sup> /h
Feed stream	3 m <sup>3</sup> /h
Solvent amount	10 m <sup>3</sup> /h

Table 4

To calculate the behaviour of the dependent variable, step input of an independent variable is made to the process. To calculate the top temperature effect on deasphalted oil viscosity, the column top temperature is changed over time and the response of the viscosity can be seen in Figure 6. Settling and dead time for the process are clearly seen after the step tests and can be compared with the final model to be sure to identify a correct behaviour for the system.

For each independent variable, pseudo-random binary sequence (PRBS) stepping is made at least 10-15 times to collect proper data and to eliminate unmeasured disturbance effects on modelling.

Two requirements are imposed during the test: no PID configuration or tuning changes are allowed. Operators may intervene during the test to avoid critical situations, but no synchronising or correlated moves are allowed.<sup>1</sup>

### Dynamic modelling

RMPCT adopts a three-step approach:

- Identify either a Box-Jenkins model using PEM or an FIR model using Cholesky decomposition.
- Fit the identified model to a low-order ARX model to smooth out large variances due to possible over-parametrisation in the FIR model. The output error approach is used to fit the ARX model via a Gauss-Newton method.

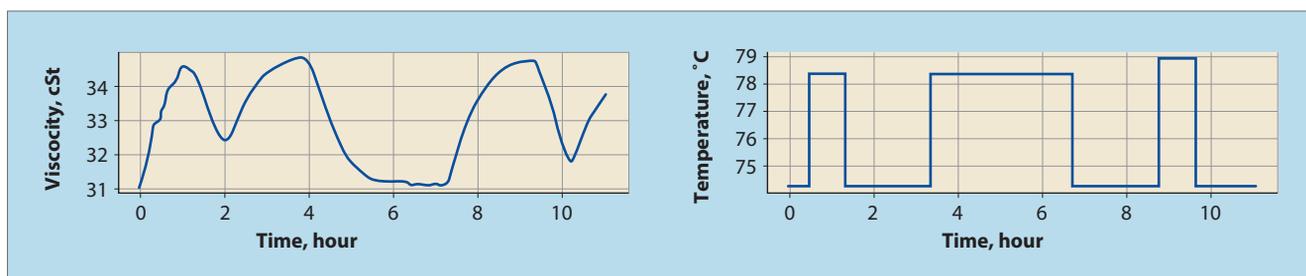


Figure 6 Response of viscosity

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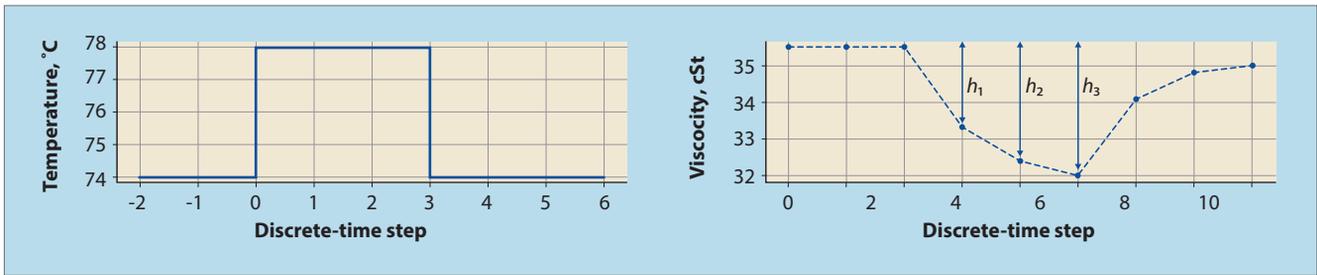


Figure 7 FIR structure

• Convert the ARX models as Laplace transfer functions.<sup>1</sup>

The general FIR structure used is given by:

$$y_K = \sum_{i=1}^{N_m} h_i u_{(K-i)}$$

in which  $h_i$  is impulse response coefficient and  $y_K$  and  $u_K$  are output and input variables. After a step test, the difference between the values of dependent variables at each discrete time step and input/output values is used in FIR to identify models.

As in Figure 7, FIR models are based on raw plant data which are obtained via step tests. FIR/PEM models are used to obtain proper initial guesses for the parametric models. The aim of parametric modelling is to fit FIR/PEM to a low order parametric model that reduces the variance.

The last step of a modelling procedure is converting models to Laplace transfer functions. Also, RMPCT

enables editing the final transfer function in the Laplace domain.

The general Laplace domain form is given by:

$$G(s) = \frac{k(\tau s + 1)e^{-ds}}{s(\tau_1 s + 1)(\tau_2 s + 1)}$$

After modelling, the top temperature effect on deasphalted oil viscosity is found by:

$$G(s) = \frac{-0.804(13.9s + 1)e^{-36s}}{649s^2 + 45.2s + 1}$$

### Controller algorithm

RMPCT manages its robustness with a range control algorithm. The controller determines how to move MVs coordinately, depending on the process model, by considering all of the effects of CVs and MVs. However, if the process model contains some error, the effects of MV moves on the process can differ from the effects of predictions, and the quality of control degrades to some extent. Unavoidably, all mod-

els contain some error typically due to a non-linear process and errors in the determination of the process model. One of the most important aspects of a multivariable controller is its ability to cope with this error.<sup>3</sup>

Four options are used by MPC algorithms to specify the future behaviour of CVs: setpoint, zone, reference trajectory and funnel. The RMPCT algorithm attempts to keep each CV within a user defined zone. When a CV goes outside the zone, the Profit Controller algorithm defines a CV funnel to bring the CV back within range.<sup>1</sup> Indeed, a Profit Controller algorithm defines a funnel and finds the optimal CV response and MV move trajectory. Only the violated part, which can be seen in Figure 8, is considered within the control problem.

$$\hat{y} = A u = A (u_{\text{future}} + u_{\text{past}})$$

Range control formulation:

$$\min_{u_{F,y}} \Pi W(Au - y)^2$$

$$ROC_l \leq \Delta u_F \leq ROC_h$$

( $\Delta$  MVs within ROC limits)

$\underline{MV} \leq u_F \leq \overline{MV}$  (MV's must lie within high and low bounds)

$\underline{CV} \leq y \leq \overline{CV}$  (CV's must lie within high and low bounds)

In the range control equation, " $u_F$ " represents the control moves, " $W$ " represents the weighting matrix and " $y$ " is used for optimal CV response trajectory. " $A$ " is the process model, " $\hat{y}$ " denotes the predicted CV and " $u$ " is used for a set of MV moves. The equation minimises the CV response error and finds the optimal response because of minimum MV movement. The RMPCT funnel approach uses two tuning parameters that can be used to adjust the desired speed of response for each output: performance ratio and decouple ratio.

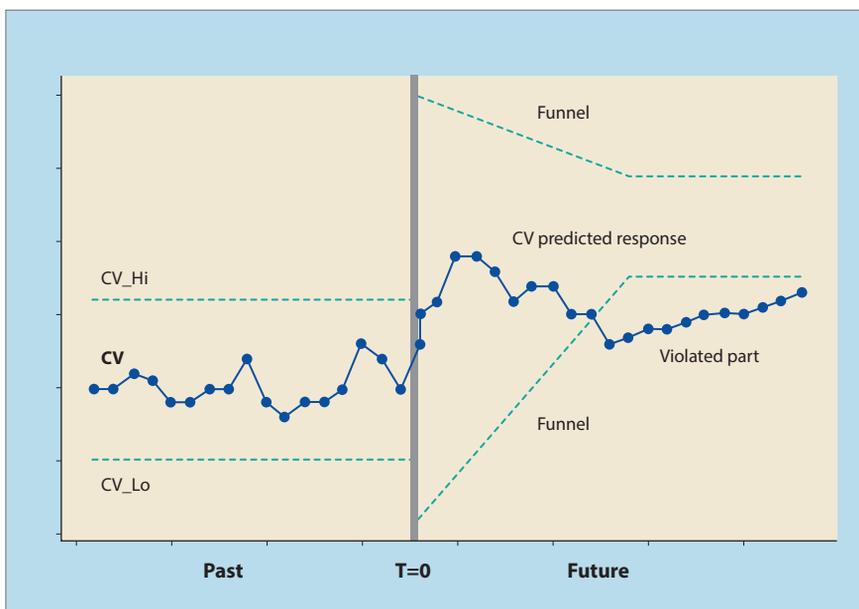


Figure 8 RMPCT funnel approach for CV response

## Economic variables

The economic objective of the controller is to maximise yield within the desired range. Put simply, yield calculation is the ratio of the amount of deasphalted oil produced to the total amount of feed. Optimisation settings which are used on the controller side are optimiser speed factor, linear objective coefficient and delta high/low soft limits. Optimiser speed factor is a tuning parameter relative to the average overall response time and controls how fast the optimiser is to drive the process toward an optimal economic solution.<sup>3</sup> Linear objective coefficient is another optimisation parameter; negative values drive optimisation to maximise the CV; positive values drive the optimisation to minimise the CV. Delta soft limits are offsets from the operating high and low limits which means high/low limits for optimisation. For delta low limit, the controller can allow disturbances to push the CV down to the operating low limit without taking immediate action. The optimiser attempts to keep the CV at or above the optimisation low limit:

$$\text{Control low limit} + \Delta \text{ Soft low limit} = \text{Optimisation low limit}$$

For the delta high limit, the controller can allow disturbances to push the CV up to the operating high limit without taking immediate action. The optimiser attempts to keep the CV at or below the optimisation high limit:

$$\text{Control High Limit} - \Delta \text{ Soft High Limit} = \text{Optimisation High Limit}$$

Delta soft limits are active for the bright viscosity and DAO viscosity; the linear objective coefficient is active for the DAO yield.

## Commissioning and tuning

In a propane deasphalting unit, feed amount and solvent amount in the feed stream are mostly constant or restricted. Therefore, total solvent amount and column top temperature can be manipulated to satisfy control objectives. Occasionally, the Profit Controller has to move away from the CV limits due to conflicts

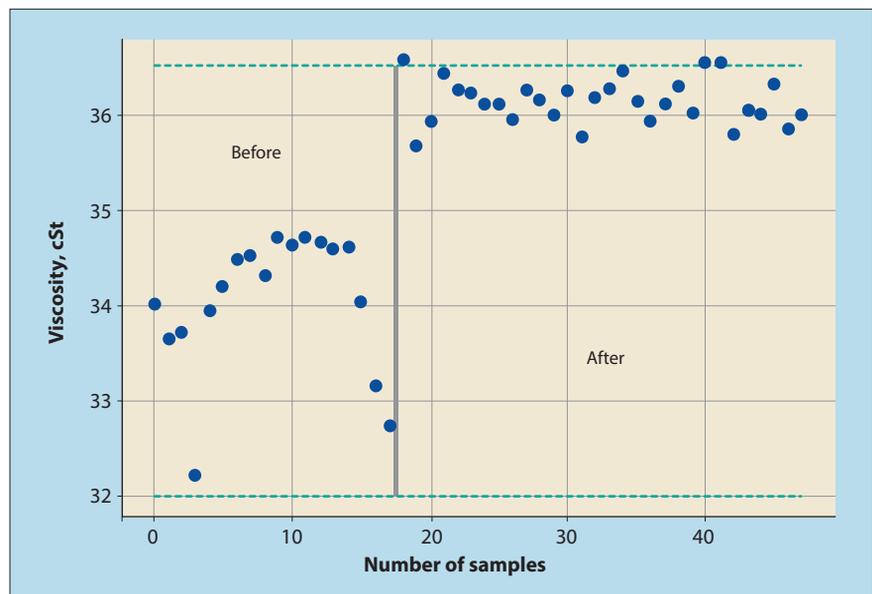


Figure 9 DAO viscosity lab results

between control objectives and MV limitations. The choice of which CV to waive is determined by considering the priority of the CVs. The solvent receiver drum pressure and column top temperature are major operational safety constraints, therefore their control should be considered as primary among other control objectives. Next, the reflux to feed ratio and valve openings should be considered, to avoid flooding and to sustain stable column operation. One of the most important tuning parameters of RMPCT for prioritisation of CVs within the control problem is EU high/low give up. Low EU give-up sets control priorities at low limits when it is physically impossible to keep all CVs within range. High EU give-up sets control priorities at high limits when it is physically impossible to keep all CVs within range. As the give-up value increases, priority decreases. Indeed, CV errors with small give-up would be corrected initially.

Tuning of the controller was configured in offline simulation mode, then checked after taking the controller online. When tuning settings were completed, optimisation settings were done. Optimising speed factor, delta high/low soft limits, and linear objective coefficient are used for the optimiser. During commissioning of the Profit Controller, observation of MV/CV responses according to CV limit changes and CV settling times is very important.

Engineer and operator training should be completed as a last stage of the APC Project, after commissioning and troubleshooting the Profit Controller.

## Results

The Profit Controller's performance was observed for about two months after commissioning. The amount of solvent given to the column and the temperature profile through the column were such that viscosity was more stable and remained close to the upper limit, and the DAO yield was optimised. There is a significant increase in bright stock yield after RMPCT.

After the controller was commissioned over 18 days, DAO viscosity increased significantly (see Figure 9). Average viscosity before the Profit Controller was commissioned was 34, but 36.1 after commissioning. This increase in viscosity draws attention to a portion of the product which was previously left as bottom asphalt product, but now was obtained as bright stock.

DAO viscosity mostly remains within quality limits and the controller pushes the system to increase viscosity and so increase DAO throughput (see Figure 10). From day 19 to day 25, viscosity decreased continuously because of the decreasing solubility of the solvent. After renewal of the solvent, the controller again approached the viscosity high limit to increase DAO throughput.

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When the DAO yield values were examined over the same time period, there was a 1.3% increase in yield. This increase corresponds to nearly 4700 tons of DAO and the gain in margin over a year is \$2.2 million. The profitability of the controller is \$450 000, based on the PDA and bright stock price difference.

Additionally, operator interventions were examined. While the average number of interventions made to the RDC column via DCS for control and optimisation purposes was eight per day before RMPCT, afterwards Profit Controller interventions to the column via DCS were never made. Operational interventions were managed entirely through APC and reduced to three per day. Average daily intervention decreased by 57%.

### Conclusion

This article describes a Profit Controller APC application for the propane deasphalting unit of the lube oil complex at Izmir refinery. APC projects began at Tupras in 2006 with a third generation MPC technology, SMOC. From 2017, it was decided to continue with fourth generation MPC technology, RMPCT. Initially, a quick base layer study was completed to check the control valves' PID performance. Next, step tests were performed to obtain dynamic process models between 4 MVs, 10 CVs and 1 DV (see **Figure 12**). Offline configuration settings and tuning checks were completed and model predictions were observed before taking the controller online. After commissioning of the Profit Controller, advanced tuning settings were finalised. As the last stage of the project, operator and engineer training was arranged. The Profit Controller's performance was monitored closely for about two months and significant improvements were observed.

The Profit Controller performs well in terms of uptime, maintenance and sustainability. With RMPCT, models can be analysed more easily and the need for revision can be quickly identified locally. Inferential model estimations also exhibit good performance when bias updates are taken into consideration.

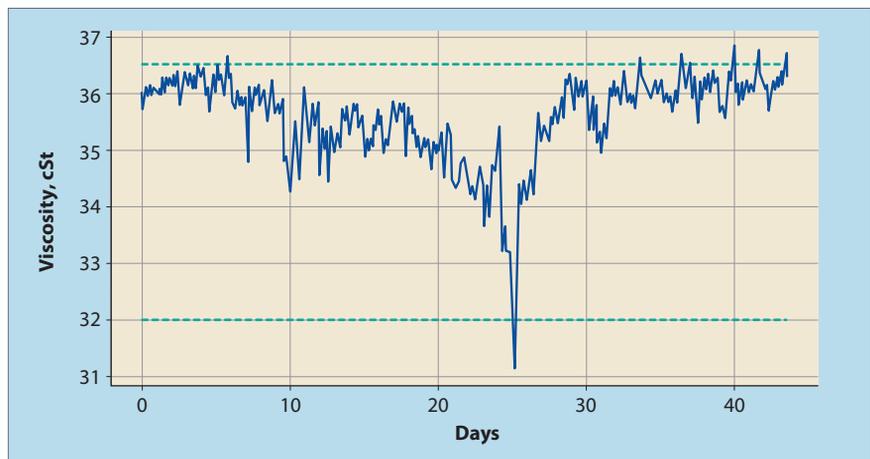


Figure 10 DAO viscosity after RMPCT



Figure 11 DAO yield

	MV1	MV2	MV3	MV4	DV1
	Top temp.	Initial solvent	Feed	Solvent to RDC	Solvent temp.
Top temp	✓			✓	✓
Bright viscosity	✓	✓	✓	✓	✓
Yield	✓	✓	✓	✓	✓
Feed			✓		
Total solvent		✓		✓	
Solvent/feed		✓	✓	✓	
DAO viscosity	✓	✓	✓	✓	✓
Solvent drum pressure			✓	✓	
Feed valve			✓		
Top temp. valve	✓				

Figure 12 Model matrix

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# When excess air becomes too much

## Excessive use of excess air impacts operating costs through fuel efficiency, furnace reliability, and stack emissions

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**S**aving fuel makes perfect sense; when you use less, you pay less. This applies to filling up your car and to fuel consumption in a process heater alike. How much money you save is easily calculated by multiplying the fuel savings by the fuel price per gallon. With combustion air, it is not so clear. Air is free, so why do you need to save on combustion air?

It is quite tempting for an operator to use a little extra air for the combustion process for several reasons. Oxygen requirements can vary because of fluctuations in the process, such as changing feed rates and feed quality. On top of that, the combustion side of the heater can be impacted by changes in fuel composition and ambient conditions. A notorious problem is that draft and air distribution inside natural draft heaters are impacted by wind gusts or rain. These and other variables may cause substantial variation in the firebox oxygen level. Any smart operator wants to keep that level well above zero, and if the fluctuations can be severe the cautious operator adds a good margin on top of the recommended level.

So, how much extra excess air is reasonable? To answer that question, we need to look at the cost of excess air. There is no simple gallon price but there are hidden costs that can be substantial.

### What is optimum?

From an efficiency point of view, the theoretical optimum excess air level is zero percent. Providing a flame the exact required amount of air for combustion ensures that all available heat is released from the

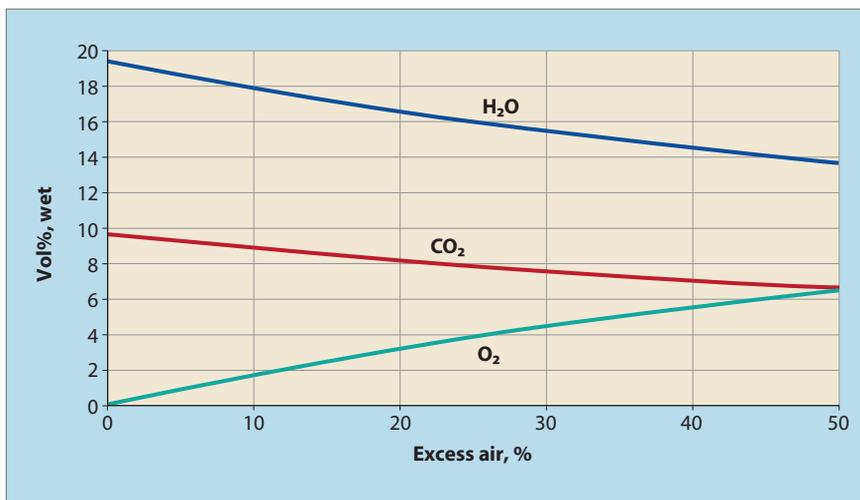


Figure 1 Flue gas composition: natural gas fuel, 60°F ambient air, 50% relative humidity

fuel. This ratio of air to fuel is called the stoichiometric ratio.

We certainly do not want to use less than the stoichiometric ratio because the combustion process would not receive enough air and risk filling the combustion chamber with unburned hydrocarbons. This is called firebox 'flooding' and the uncontrolled reaction of these hydrocarbons with any leakage air is a serious safety risk.

Keeping it at exactly zero is not feasible either. The typical heater instrumentation and control system is not able to keep up with any of the aforementioned fluctuations in the system due to response lag. Even if it could, it is very difficult to design a combustion process with perfectly mixed air and fuel. So, we need to provide some 'excess' air to the system to provide margin against fluctuations and ensure complete oxidation of the hydrocarbons.

The recommended excess air level for a gas-fired process furnace is

15% according to industry recommended practices like API 535. In certain process plants such as ethylene and hydrogen production, the furnaces operate very steadily and at high temperatures. In those cases, the industry norm is an excess air level of 8-10%. Combustion of liquid fuels, on the other hand, typically requires excess air levels of 20-25% to prevent soot formation.

The operator of the heater measures excess air indirectly by checking the firebox oxygen level. To convert from oxygen level to excess air percentage, use the following simple formula:

$$\text{Excess Air (\%)} = \frac{92 O_2}{21 - O_2}$$

with  $O_2$  expressed in vol% (dry). Using this equation, we see that 3%  $O_2$  translates to 15% excess air, and 5%  $O_2$  is equal to 35% excess air.

### The cost of excess air

Let us first discuss some firebox fundamentals that few people know

Flue gas properties and radiant efficiency as a function of excess air

Excess air %	Excess O <sub>2</sub> vol% dry	O <sub>2</sub> vol% wet	CO <sub>2</sub> vol% wet	H <sub>2</sub> O vol% wet	Emissivity	Flame temp. °F
0	0.0	0.0	9.6	19.4	0.284	3673
10	2.1	1.7	8.8	17.9	0.276	3430
15	3.0	2.5	8.5	17.2	0.273	3321
20	3.8	3.2	8.1	16.6	0.269	3219
25	4.6	3.8	7.8	16.0	0.266	3123
30	5.2	4.4	7.6	15.4	0.262	3034
40	6.4	5.5	7.1	14.5	0.256	2869
50	7.5	6.5	6.6	13.6	0.251	2722

Table 1

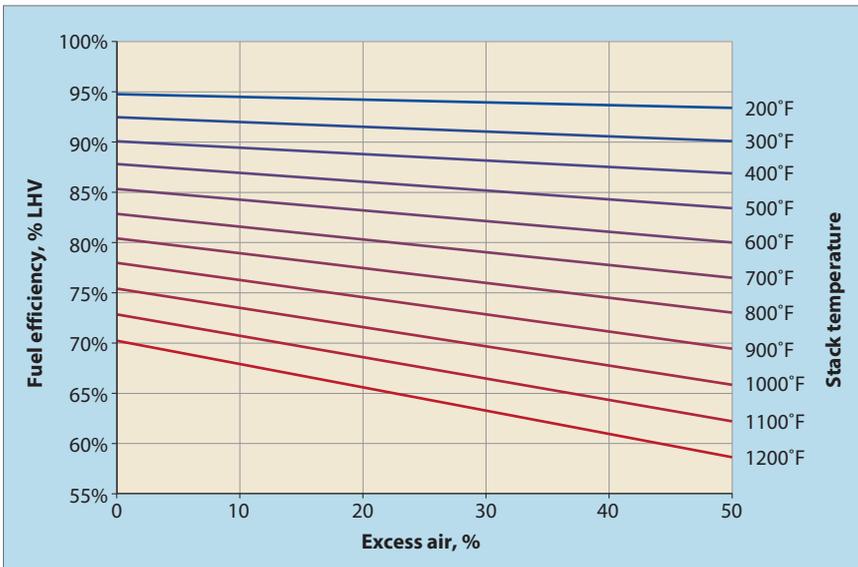


Figure 2 Heater efficiency vs excess air: natural gas fuel, 60°F ambient air, 50% relative humidity, 2% setting loss

or care about. Air consists almost exclusively of nitrogen and oxygen. Since they are diatomic, neither gas participates in the transportation of radiation energy. The only gases that cooperate in a meaningful manner are the water vapour

and carbon dioxide that form during combustion (see Figure 1). If the firebox operates at a high excess air level, the concentration of H<sub>2</sub>O and CO<sub>2</sub> is diluted, which lowers the effective emissivity of the flue gas. As the flue gas becomes a less effective

emitter of radiant energy, the firebox thermal efficiency drops.

The second problem is that every excess pound of air ‘steals’ heat from the combustion process. Each excess pound of air entering the heater is an extra pound that must be heated to the furnace temperature. It effectively lowers the equilibrium temperature, also known as the adiabatic flame temperature. Since radiation heat transfer is proportional with absolute temperature to the fourth power, the radiant efficiency of a firebox drops tremendously when its temperature drops because of all the extra air.

Table 1 lists the properties of flue gas from the combustion of natural gas with varying levels of excess air. The table clearly shows a strong dependence of emissivity and adiabatic flame temperature on flue gas composition. Between 15% and 25% excess air, the dry oxygen level only increases from 3.0 to 4.6 vol%. However, due to the drop in CO<sub>2</sub> and H<sub>2</sub>O concentration the flue gas emissivity drops 3% and the adiabatic flame temperature drops by an astounding 200°F (93°C). In a typical firebox, this combination of lower emissivity and lower adiabatic flame temperature reduces the radiant thermal efficiency by about 5%. The firebox needs to be fired proportionally harder to compensate and is less energy efficient.

The convection section is where the residual heat in the flue gas is used for feed preheating. The convection section will compensate for some of the loss of firebox radiant efficiency but not completely.

**The cost of ‘excess’ excess air**

One can use Figure 2 and Figure 3 to calculate the cost of too much excess air. Use Figure 2 to determine the fuel efficiency of a fired heater as a function of excess air and stack gas temperature and Figure 3 to find the cost of natural gas around the world, expressed in \$/MMBtu. An example calculation follows.

In Q3 of 2019, the US natural gas cost was approximately \$3 per MMBtu (see Figure 3). For a process heater operating at 100 MMBtu/h, the total fuel cost is then 100 MMBtu/h x 8760 h/year x \$3

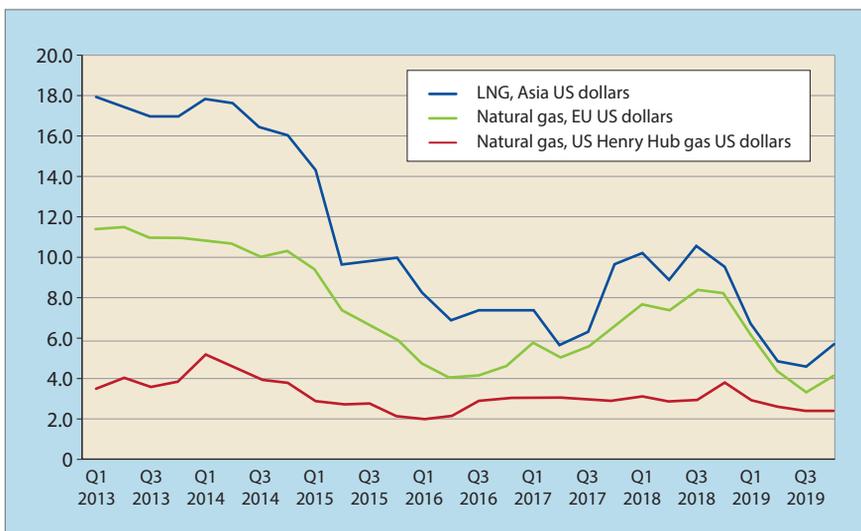


Figure 3 Natural gas price, \$/MMBtu

per MMBtu/h = \$2.63 million. That means each 1% reduction in fuel efficiency costs \$26 300/y. For a typical 300 000 b/d refinery each percent energy gain or loss represents around \$1 million.

### Case study

A train of four identical heaters runs at an average of 5.5 vol% O<sub>2</sub> (dry) at the arch, due to various design and operational issues. A change in the downstream process reduced the heat requirement from the heaters by 40%, which dropped the firebox temperature well below 1200°F (650°C). The floor-mounted burners are of the latest generation ultra low NOx design. Burners of this type reduce NOx emissions using internal flue gas recirculation. The dilution of the flame with inert gas causes a delay in combustion reactions and a reduction in thermal NOx. This approach works well at typical firebox temperatures of 1400-1600°F (760-870°C) but flame quality and stability deteriorate significantly when the firebox temperature becomes too cold. The only remedy available to the operator is to operate the heaters at higher oxygen levels.

Additional problems are caused by operating the burners at duties well below their optimal design point. In this case, the 40% reduction in heat liberation creates a soft and lazy flame with a tendency to roll into the coils. This is due to a lack of airside pressure drop used for fuel-air mixing. High wind speeds in the summer create large swings in air flow through the natural draft burners, occasionally producing high amounts of carbon monoxide. Again, the only remedy is to operate at higher excess air levels to increase the mixing rate and create a stiffer flame.

The heater absorbed duty is 94 MMBtu/h. The impact of running the heater at 30% excess air instead of 15% is shown in **Table 2**. The radiant efficiency has dropped from 58.4 to 53.9% which requires an increase in fired duty of 4.1 MMBtu/h.

The negative consequences of high excess air operation are compounding. Using a fuel cost of \$3/MMBtu, the operating cost

Effects of excess air						
Excess air %	Fired duty MMBtu/h	Stack temp. °F	Radiant eff. %	Fuel efficiency %	Fuel cost \$MM	CO <sub>2</sub> US ton/yr
0	102.7	435	63.1	91.4	2.61	48 266
10	105.0	490	59.9	89.3	2.76	49 368
15	106.2	516	58.4	88.3	2.79	49 955
20	107.5	540	56.9	87.3	2.83	50 567
25	108.9	563	55.4	86.2	2.86	51 207
30	110.3	585	53.9	85.1	2.90	51 871
40	113.2	626	50.9	82.8	2.97	53 291
50	116.5	663	48.1	80.5	3.06	54 840

**Table 2**

has increased by 4 x \$108 000 = \$432 000 annually. In addition to fuel costs, lower energy efficiency also increases greenhouse gas emissions. In the previous case, the total CO<sub>2</sub> emissions increase by 4 x (51 871 – 49 955) = 7664 t/y. In countries where CO<sub>2</sub> penalties/credits are considered at \$30/t, this would translate into an additional cost of \$230 000/y. For this site, the deci-

## While the fuel cost is obvious and easy to calculate, other hidden costs of running at high excess air may not be as readily visible and only truly manifest themselves over time

sion to run at 5.5% oxygen instead of 3% oxygen comes with a price tag of \$660 200/y – and this is assuming fuel costs remain low and CO<sub>2</sub> penalties do not increase.

### Other hidden costs

While the fuel cost is obvious and easy to calculate, other hidden costs of running at high excess air may not be as readily visible and only truly manifest themselves over time. Running at a higher excess air level changes the duty split between the radiant and convection section. The combination of a higher firing rate and lower radiant efficiency leads to an increase

in convection duty and temperature. In our example case, the stack temperature has increased by 70°F (20°C). Higher flue gas temperatures will increase the temperature of tubes, fins, tube supports, and stack, which could shorten the lives of each of these components.

There are also effects on coil longevity, electrical power consumption, and emissions. The higher firing density in the firebox and higher flue gas temperatures create hot spots on the radiant and convective tubes, which increases coking rates on the inner tube walls. The coking layers will gradually build over time, further increasing metal temperatures and creating additional pressure drop. High excess air results in high pressure drop and additional fan power when using a forced draft or induced draft fan and the loss of furnace capacity. Finally, running at a high excess air level can significantly increase emissions of nitrogen oxides. Running at 30% excess air compared to 15% could increase NOx emissions to 150-200% of the design values.

### How to fix high excess air – and sustain the gain

Using the methodology above it is straightforward to estimate the cost of high excess air and justify improvements for any fired heater. On the other hand, it can be more difficult to convince operators to run at low excess oxygen if there are fundamental issues with the heater that could jeopardise safety and/or reliability. It is important to measure the excess oxygen correctly, control the excess oxygen easily, and ensure that operators reliably

and safely operate the heater at the lower oxygen levels.

Measure the excess oxygen at the firebox arch to improve accuracy and analyser response time. In many fired heaters the analyser is located in the stack as mandated by environmental requirements. However, this location is not ideal for measuring excess oxygen. Tramp air that enters the heater through the various tube penetrations typically results in 1-2% more measured stack O<sub>2</sub> when compared to the actual firebox O<sub>2</sub> concentration. This is equivalent to an extra 5-15% excess air and makes it difficult for the operator to know the burner excess air and exercise proper control.

Measure carbon monoxide concentration at the firebox arch. Measuring excess oxygen is not a guarantee that combustibles are absent; flame issues may still occur due to poor air distribution among the burners, fouling of the burner tips, low firebox temperatures, and other combustion issues. A high concentration of combustibles can cause afterburning in the convection section and is a safety risk.

Non-contact measurement technology, like tunable diode lasers (TDL), improves reliability while simultaneously improving response time and accuracy. Zirconia based sensors sample from a single point. The laser (or even multiple lasers) used in TDL measurement systems average the oxygen and CO concentration over the length of the beam(s), thus providing a more representative value for larger fireboxes. Additionally, the lack of long flue gas sample lines makes the response time much faster than extractive methods.

Maintain arch draft at 0.1 inH<sub>2</sub>O (2.5 mmH<sub>2</sub>O), ensuring that the entire heater is under negative pressure. Use burner registers to control the firebox oxygen concentration. Often, heaters that operate at high excess air also operate with a high draft. Heaters are not designed to be perfectly airtight and the more draft in the heater, the more excess air is drawn in from locations apart from the burners. The burners are the only place where excess air is to

be introduced by design. Common locations for leaks include header box doors, tube seals around tube penetrations, seals around observation doors, viewports on burners, and joints in the heater casing.

The reason to operate at high draft can be purely mechanical, like a stack damper that is stuck and inoperable, but draft can also be notoriously difficult to control when everything is working correctly. The challenge is that changing the draft of a fired heater simultaneously changes the excess oxygen level. Operators must adjust the burner registers when the draft changes to keep the oxygen level at target. This can be a challenge for heaters with many burners.

Consider automatic DCS control of draft and oxygen. Operator safety and efficiency are reduced

## Besides having a direct impact on operating costs through fuel efficiency, excess air affects furnace reliability and stack emissions

by continual adjustment of the stack damper and burner registers in reaction to changing ambient conditions and firing rate. The hazard may become significant if the heater operates outside of its design parameters or there is a sudden change in load or weather. Instead, the DCS can control the draft by adjusting the stack damper or induced draft fan speed. Operators may control the excess oxygen by modulating combustion air to the burners through dampers on a common burner plenum or individually through actuators on burner registers. The required number of actuators may be reduced by connecting multiple burner registers to a single shaft controlled by a single actuator. Model-predictive control can be used to optimise the combustion control system by accounting for

factors such as excess oxygen, the concentration of combustibles, firebox temperature, heater firing rate, process flow, process outlet temperature, and ambient conditions.

Minimise tramp air. Ensure that tube and door seals are in good condition and keep sight ports and explosion doors closed when not in use. Consider glass-covered sight doors to improve operator safety and minimise tramp air.

Verify that a heater can safely and reliably operate at the target oxygen and draft level. A common reason for heater operation at high excess air is that flame problems occur at low excess air. Poor flame quality could result in flame impingement or even flame extinction. This could be caused by a whole range of issues: poor burner maintenance, ambient conditions like high wind gusts, operation outside the heater or burner design operating envelope, flame interactions caused by firebox aerodynamics, poor air distribution between burners, excessive air leakage, and poor instrumentation. Even this list is not exhaustive and a comprehensive root cause analysis typically consists of a combination of a detailed site survey, thermal modelling, burner testing, and a CFD analysis of the firebox and combustion air duct.

### Conclusion

Although air is free, running at high excess air is not. Besides having a direct impact on operating costs through fuel efficiency, excess air affects furnace reliability and stack emissions. Running at high excess air may afford improved resistance to fluctuations, but too much excess air can adversely impact profitability.

**Erwin Platvoet** is Partner and President of XRG Technologies. An innovator whose career spans more than three decades in the heat transfer and combustion industries, he holds eight patents in fired heat transfer and emissions control technology, has published numerous papers, and co-authored the *John Zink Combustion Handbook* and *Industrial Combustion Testing*. An active member of the API 560 and API 535 subcommittees, he is a graduate of Twente University in The Netherlands with an MS in chemical engineering.



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# Increase hydrogen production during a turnaround

## A foil-supported catalyst meets increased hydrogen demand within a normal turnaround

MATTHEW WILSON and KEN CHLAPIK  
*Johnson Matthey Catalysts Technologies*

The International Maritime Organization's (IMO) low sulphur marine fuel regulation is in place for 2020. There are still technology decisions being made and some wait and see approaches as to how the regulation will be managed. It is understood that a predominant portion of the low sulphur fuel required for meeting this regulation will be from increased low sulphur diesel production. Some complex refineries have the capacity to meet these increased hydroprocessing needs as sweeter tight oils have replaced sour crudes. This increased diesel production will require more hydrogen production from the site. Processing the tight oils has increased production of catalytic reformers, resulting in more hydrogen contribution to the site than in previous years. With this additional source of hydrogen on site, many hydrogen plants are running at 70-80% utilisation. The low sulphur marine fuel regulation, as well as refiners hydroprocessing more bio based feeds that are lean on hydrogen, is creating a step change in hydrogen demand. Even with the increase in hydrogen availability from the catalytic reformer, many refiners are needing about 20% more hydrogen than nameplate within their refineries.

Traditionally, the on-site choices that a refiner has for a 20% uprate are either building a grassroots hydrogen plant or uprating an existing hydrogen

plant by adding a heat exchange reformer. These solutions tend to pose challenges for the refiner in the following ways:

- Availability of capital
- Plot plan
- Turnaround scheduling
- Compression
- Steam production
- Increased asset wear

Johnson Matthey's Stackable Structured Reactor – Catacel SSR steam reforming catalyst technology addresses and helps minimise or remove these challenges while enabling a 20% uprate above nameplate hydrogen production. This article will address each of these challenges with respect to a 20% uprate using SSR technology.

### Availability of capital

Downstream margins are under tremendous pressure and compression, reducing the availability of capital. Improvements in capital utilisation and efficiency need to be apparent for expansion projects to come into consideration.

A Catacel SSR technology uprate offers one of the lowest capital usages for increased hydrogen production, thereby increasing overall return on investment. The uprate does require debottlenecking of the fired reformer to enable the process flow and increased heat load to be achieved with the fired reformer. Typically, the fired section of the steam methane reformer needs to be modified for this increased heat load, particularly the burners and ID fan. Depending on the plant, additional upgrades may be needed in the feed purification system, the outlet header of the fired reformer, and the pressure swing absorption system that purifies the hydrogen product. The level of capital needed will vary depending on the modifications required, but even with this a 20% uprate with SSR still offers lower capital for increased production by 20-50%.

### Plot plan/turnaround scheduling

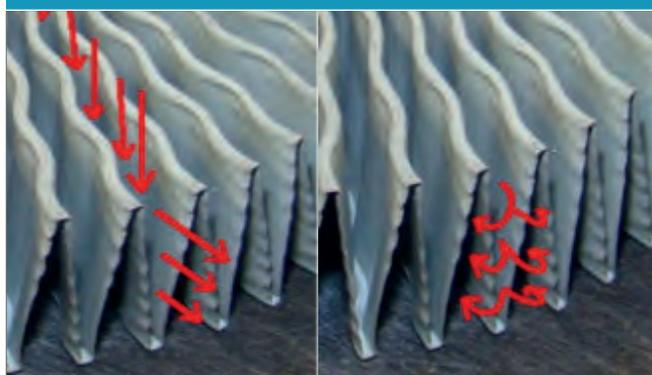
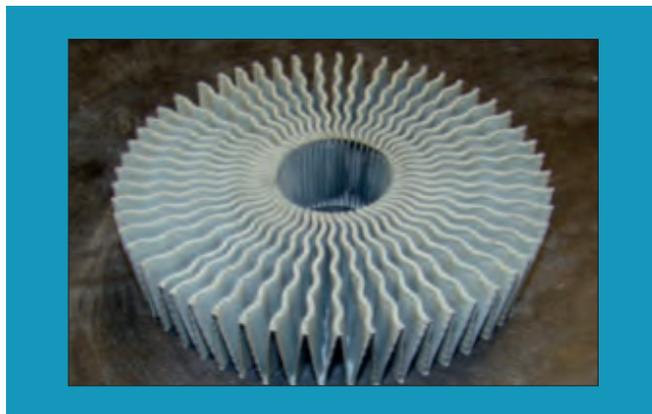
As an operating company considers increased capacity, some additional plot plan needs to be considered either for a grassroots plant or when adding additional equipment. Many refineries are in populated areas close to metropolitan cities, allowing limited space for expansion or addition. Most hydrogen plants have even less space available near the plant itself. Beyond having the space, extensive pre-work needs to be accomplished to make that space ready,



Figure 1 Typical refinery hydrogen plant with minimal plot plan availability



**Figure 2** SSR – Stackable Structured Reactor



**Figure 3** Directed flow pattern across the coated foil section of a SSR reactor

including environmental regulations, infrastructure, and preparing the site. These activities require site access during plant operation or additional work to be accomplished within the upgrade turnaround. For complex refineries that have the hydroprocessing capacity available for additional diesel production, extension to the typical three-week turnaround schedule quickly impacts the cost of the turnaround. The missed opportunity of production can quickly add 10-30% to the cost of the project. Obviously, a grassroots hydrogen plant will require substantial space and many approvals. Most grassroots hydrogen plants are taking 24-36 months from project approval to beneficial operation. This timing is close to a full hydroprocessing turnaround cycle to achieve production. Upgrades that add equipment to the plant are challenging to complete in a three-week turnaround. Delays in the turnaround schedule can remove millions of dollars in production from being achieved. A Catacel SSR technology uprate replaces the existing pelleted reformer catalyst in the same time that it would take to load pelleted catalyst. The typical

modifications to the steam methane reformer, burners, and ID fan are accomplished within the three-week turnaround.

Catacel SSR technology is a stackable structured reactor that has been designed for reformer duty. Using catalyst coated thin metal foils, the technology packs 40% more surface area per unit volume than pellets utilising the existing reformer tubes to drive the additional reaction needed for the increased flow (see **Figure 2**). Fired reformer performance in a hydrogen plant relies on getting the heat into the process gas. While developments have occurred through the years on pelleted reforming catalyst, the improvements are small and still rely heavily on the randomness of pelleted catalyst loading. The patented design of Catacel SSR technology directs the process gas to encounter all the catalyst surface available and then directs this flow against the inside tube wall to maximise heat transfer and process gas reheat while moving to the next layer of surface area in the structure. This results in a 20-30% increase in heat transfer into the process gas. The technology reactor reaction and

heat transfer intensification enable a tight steam methane approach to equilibrium to be maintained with the 20% uprate achievement using the existing reformer geometry.

### Compression

Many hydrogen plants are limited on feed gas compression. A 20% increased flow rate can put too much resistance to flow on the feed gas compression for a pellet loaded reformer, making the uprate unachievable. In a similar way, adding a heat exchange reformer adds additional resistance to flow and added pressure drop through the added piping and heat exchange reformer equipment. The patented design and flow pattern of Catacel SSR can reduce the reformer pressure drop by 20% (see **Figure 3**). The steam methane reformer pressure drop can be as much as 60-70% of the flow sheet pressure drop. A 20% uprate adds no additional pressure drop to the steam methane reformer, minimising the resistance to flow on the feed gas compressor.

### Steam production

Most hydrogen plants produce more steam through the heat recovery

system of the hydrogen plant flowsheet than can be used in the hydrogen plant, making it a net steam exporter. The steam is used within the overall steam system of the refinery for hydroprocessing units and for driving some of the major refinery equipment. While several refiners are moving from steam driven to electrically driven equipment, many still rely on and value the steam export from the hydrogen plant for the rest of the refinery. Adding a heat



**Figure 4** Common asset wear mechanism of overheated SMR tubes leading to reformer tube failures

exchange reformer depends on the fired steam methane reformer to produce the process steam, reducing or eliminating steam export from the hydrogen plant. With a SSR technology 20% uprate, additional export steam can be produced for the refinery. Additionally, if there are limits on the hydrogen plant process gas boiler/steam raising system, the technology is able to operate at lower steam to carbon ratios as it is more reactive, changing the conditions throughout a natural gas fired reformer to avoid carbon formation.

### Increased asset wear

When uprating an existing plant, operators are concerned that the uprated condition will create greater wear on the asset, adding to the cost of operation over the plant's lifecycle as well as increasing potential for mechanical failures within the flow sheet. Replacing the steam methane reformer tubes tends to be one of the higher cost asset replacements in the hydrogen plant (see **Figure 4**). Most of the major equipment is designed for the lifecycle of the plant but the reformer tubes are usually expected to be changed one to two times in the plant's lifecycle. With most hydrogen plant uprates, the additional firing and reforming heat load required can result in higher reformer tube wall temperatures, reducing the life of the tubes exponentially. Catacel SSR technology is effective in extracting

heat from the inside reformer tube wall. This increased effectiveness in heat transfer is resulting in hydrogen plants utilising SSR technology to experience a 20°C reduction in reformer tube wall temperatures. For a SSR technology 20% uprate, there is no increase in reformer tube wall temperatures from those experienced at nameplate production. In other words, the 20% uprate adds no increased asset wear on the steam methane reformer tubes and minimises the risk of reformer tube failures, thus improving the safety of the operation even with the uprate.

Catacel SSR technology continues to grow its application experience profile, demonstrating its capabilities while minimising the risk of its use for hydrogen applications. The technology has been designed and is operating successfully in over a dozen steam methane reformer applications, it has been loaded in over 650 reformer tubes, established two lifecycles of performance, and designed into five uprating applications. Johnson Matthey is optimising this technology and with growing demand it is doubling manufacturing capacity by 2021.

### Conclusion

With changing regulations on IMO marine fuel standards, refiners are reviewing their hydrogen management needs and realise more is needed for increased hydroprocessing demands including biofeed

processing. An SSR technology 20% uprate enables a timely increase in hydrogen plant production to be achieved within a typical turnaround. It provides a low capital solution with high return on investment without impacting barriers to hydrogen plant increases like feed gas compression, steam export, and steam methane reformer tube wear. It also minimises the time to beneficial operation compared to other solutions like a grassroots plant or addition of a heat exchange reformer. The 20% uprate requires fewer plant modifications, therefore reducing risk and increasing safety for a refiner in achieving these additional hydrogen needs.

**Matthew Wilson** is the Global Commercial Manager for Johnson Matthey Catalysts Technologies structured catalyst business, CATACEL SSR, based in Chilton, UK. He is responsible for growing the structured catalyst business globally to the ammonia, methanol and hydrogen markets. He holds BEng (Hons) in chemical engineering from Teesside University.

**Ken Chlapik** is the Global Market Manager – Hydrogen for Johnson Matthey Catalyst Technologies and is based in Oakbrook Terrace, Illinois, US. He has been with Johnson Matthey for 34 years and is responsible for advising Johnson Matthey's Refinery Business on the market dynamics and future growth of hydrogen as well as ensuring that hydrogen production catalysts and related technologies meet the global needs of the industry. He holds a BS degree in chemical engineering from Northwestern University, Evanston Illinois and has recently been awarded the AFPM Peter G. Andrews Lifetime Service Award.



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# Managing a control system migration

## Overcoming the inherent risks of a control system upgrade relies heavily on collaborating and sharing information transparently with a systematic, integrated approach

ANAND SRIVASTAVA, MANISHA DAS and MEGHNA BAHL  
*Fluor Daniel India*

Since the birth of control systems and their utilisation in processing industries, operators' interfaces have changed a great deal, but not as much compared to 'seen' hardware and 'unseen' software. There have been major technological advances which are not visible to non-instrumentation personnel.

For a production team, the priority is production numbers. The team will squeeze the last functional drop from a process control system until a shutdown or revamp takes place.

The past decade has been governed by considerations given to low sulphur containing fuels, hence the oil and gas industry has seen a shift towards clean fuels globally. The major contributors to this shift are new environmental regulations and revised fuel standards defining very low sulphur levels. To meet the regulations, the refining industry has invested heavily in the installation of new units and upgrades, or on the expansion of existing units. In today's oil prices scenario, the industry's focus is shifting towards brownfield projects which require relatively low capital expenditure and a high return on investment. This calls for effective migration of process control systems/emergency shutdown systems (PCS/ESD).

No two construction projects are exactly the same and vary in design, size, capacity, utilities, complexity, and so on. When it comes to a brownfield project execution, there are several unique design challenges that are not generally encountered in other brownfield or greenfield expansions.

With such upgrade projects comes a need to collaborate and share information across the entire value chain to drive down costs and

increase the accuracy of delivery dates. Owners, operators, contractors, and suppliers working in the petroleum industry must find innovative solutions to minimise complexity and risk in these massive undertakings.

This article outlines:

- The rationale behind control system migration
- Possible challenges
- Execution strategy
- Implementation issues

### Why is a control system upgrade needed?

The major reasons behind the decision to upgrade a control system include:

- System failures – unplanned outages and increased downtime, leading to production losses
- Lack of spare part availability
- Incompatibility of advanced interfacing applications with existing systems
- Inefficient operation
- Limited support from original equipment manufacturer
- Lack of skilled resources

### What if an upgrade is delayed?

The costs of maintaining an old system will eventually outgrow the cost of migration to a new system.

However, if an upgrade is delayed, the refiner has to bear certain costs associated with the risks of using old systems. These can be production losses, unplanned shutdowns, or higher maintenance costs.

### How to perform a control system upgrade

Once a decision is made to execute a revamp, the need for an execution plan arises. Multiple strategies can apply:

- Bring the plant to a shutdown, replace the system and take it back online. Because of extensive downtime, the economics do not work.
- Plan the migration, do the engineering, procurement and in planned shutdown execute the migration. Effective if planned in a pragmatic fashion.
- A piece-wise approach assigns the criticality of units; based on this, replace the set of I/O cards and processors followed by other related items like human machine interface.

A control system upgrade, being the critical part of any expansion project, is strategically scheduled during a planned shutdown. Shutdown means scheduled large scale maintenance activity in which an entire process unit is taken off stream for a period of time for revamp, debottlenecking, and replacement.

The main objectives of shutdown schedules are to make the plant safe to operate until the next outage, improve the efficiency and throughput of the plant by suitable modification, increase reliability/availability of equipment during operation, re-establish plant capacity, face minimum production loss and cost overruns, and complete corrective maintenance.

The baseline activities that are performed in an expansion or revamp project include:

- Site visit to analyse the current condition, physical locations, and surroundings to plan out revamp activities
- Collection of existing data and architecture to understand the existing system and formulate a plan for a revamp
- Validation of existing system design and installation documentation is carried out by the site engi-

neering team in close coordination with the operation and maintenance team prior to a planned shutdown

- Alignment with the system automation supplier over the plan for migration to the new system
- Preparation of a detailed shutdown schedule for complete revamp activities; scheduling of activities is of utmost importance for effective migration to take place without any slippage
- Alignment with owner and system supplier regarding the schedule and shutdown duration requirements
- Replacement and migration of the existing distributed control system (DCS) and ESD, with upgraded systems located in either the existing control room or prefabricated satellite rack room (SRR) by the automation vendor

The following are some activities to be performed by the system automation supplier:

- Site data collection including control room surveys for checking space requirements and making necessary backups
- Software migration and preparation of the database
- Detail engineering by system automation vendor and review of the same by the EPC contractor
- Factory acceptance test (FAT) in the presence of the EPC contractor, the owner, and the project management contractor (PMC) if applicable. FAT is of prime importance as it reduces surprises during construction and commissioning. It is suggested to involve operators in FAT so that logics and graphics can be verified to avoid changes during commissioning
- Site acceptance test in the presence of the EPC contractor, owner, and PMC if applicable to ensure all required hardware is available
- Actual system replaced during shutdown
- Commissioning of the new system
- Start-up assistance, if required by the owner
- Decommissioning of the old system

### Major challenges

Successful execution of a brownfield project is inherently complex due to constraints and unknowns in existing facilities. Shortfalls become

apparent only during execution, resulting in delays in project completion, leading to production losses to the owner and possible cost overruns to the contractor. This may lead to erosion of return on investment.

For any brownfield project, big or small, one major necessity is to avoid these unexpected changes or surprises that are often uncovered during detailed design or, worse still, during construction. It is therefore essential to identify issues early during design, and ensure alignment of all stakeholders, including the vendors.

Some major challenges are discussed in the following.

### Quality of existing asset documentation

For any existing refinery that has been running for decades, the available asset documentation and databases may not be up to date and might have vital missing information. This will have high potential for not capturing important information during detailed engineering which will be uncovered by the site construction team during execution. Such surprises for the site team will induce delays and will have a cascading impact on a successful turnaround.

Thus, current asset documentation that is provided by the owner's operation and maintenance team, along with the automation databases of the existing system, form the major building block during the detail engineering phase, which is rolled up into master documentation for execution during construction. Thus, close coordination, alignment, and cooperation from all parties involved is necessary for project success.

One of the direct impacts of incorrect documentation is the need for new or additional materials which can affect the sequence, duration, and schedule of work packages. Thus, certain materials like spare cables and spare junction boxes should be available at the construction site to avoid last minute constraints.

### Span of planned shutdown

Another critical and major constraint that can occur at the site is a predefined and short time frame

of the shutdown phase where all planned activities must be implemented without failure. Some activities that do not require shutdown should be completed in advance.

### Limited availability of labour

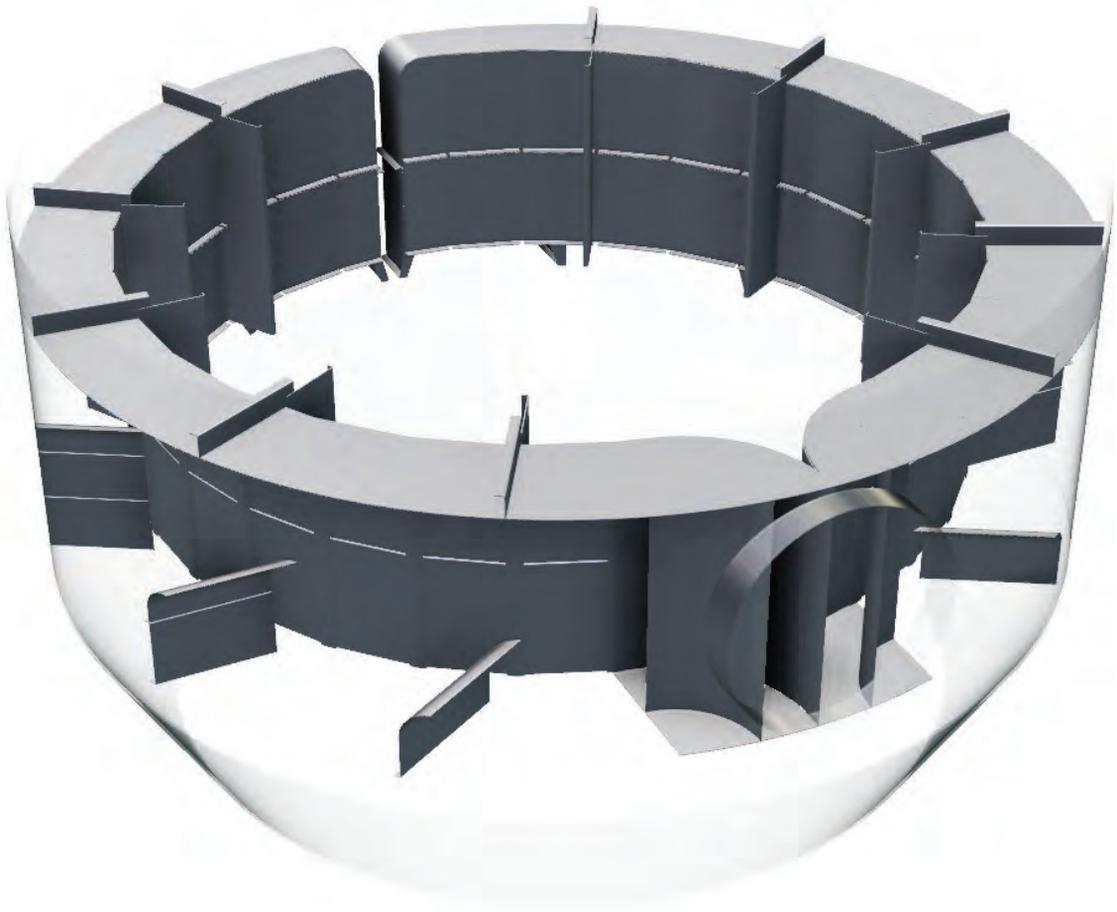
Due to limited shutdown time, availability of a large volume of skilled labour to perform the system turnover with accuracy is another challenge faced by the EPC contractor. The availability of skilled labour and its associated costs depends on geographic location.

Many times, unavailability of labour leads to having more than one construction contractor for a system turnover. In such a scenario, coordination and collaboration between various construction contractors is the key to achieving target within schedule.

### Pre-shutdown activities

As the shutdown period is limited, the majority of system turnover activities should be completed ahead of shutdown. Some of the major site activities are listed below:

- Collection of all inputs such as existing wiring reports, loop diagrams, junction box location plans, cable layouts, and so on
- Identification of existing multi-core cables running from field junction boxes to existing SRRs
- Field survey of cable routing to verify correctness of cable routing layouts and to ensure cable lengths are sufficient
- Field survey to collect missing information in documents provided by the owner
- Identification and verification of various tie-ins including third party system interfaces and communication with other control rooms
- Laying and termination of new multi-core cables in new marshalling cabinets placed in pre-fabricated SRRs
- Site acceptance test of new DCS and ESD systems supplied by the automation vendor. This is usually performed jointly by the EPC contractor, owner's maintenance team, and project management team
- Adequacy of existing UPS to be checked in order to support the new system



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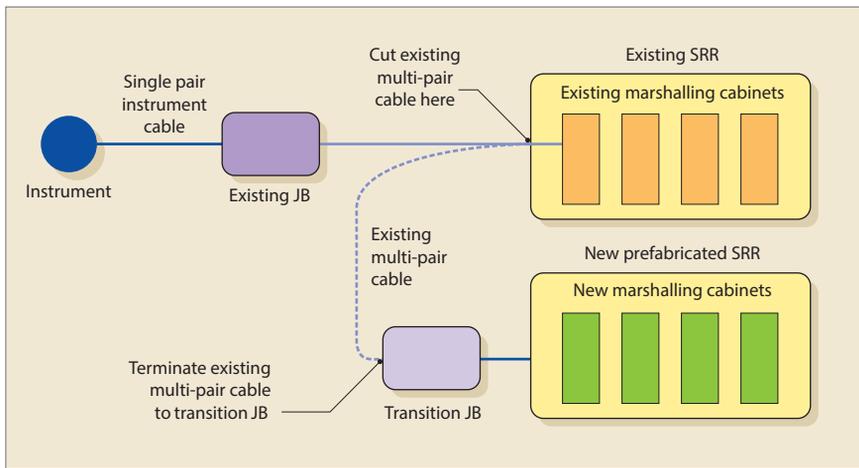


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**Figure 1** Existing cable termination to transition junction box

- Provision of power supply points at planned locations, for instance laying of additional cable, if any, for welding
- Erection of scaffolding if required in existing SRRs

### Execution strategy

The count of existing I/O signals may range from 1000 to 10 000 or even more, depending upon the process plant planned for upgrade. Shutdown periods have almost no option for extension, otherwise they may impact plant production, leading to huge losses for the owner.

Thus, the shutdown period plays the major role in deciding the execution strategy for a replacement project. System replacement can be executed in three ways, as follows:

### Introduction of a new junction box between existing junction box and new system

In such a case, a multicore cable between the new junction box and

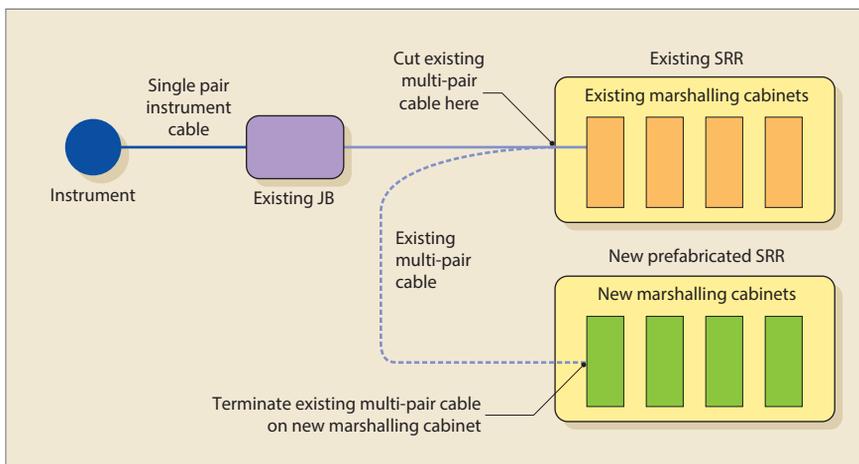
marshalling cabinet can be routed, connected, and loop tested before shutdown. This new junction box is referred to further as the transition junction box (see **Figure 1**).

The existing multicore cable can be cut at the existing system end during the shutdown and then connected to the transition junction box. This should be followed by loop testing, logic testing, and other commissioning activities.

For the success of this execution strategy, it is important to ensure availability of space for mounting transition junction boxes near an existing SRR. The length of existing multicore cable should be sufficient to be pulled and terminated in the transition junction box.

### Existing multicore cables to be terminated directly in new marshalling racks

In this strategy, existing multicore cables are cut at the existing system end in the SRR during shutdown



**Figure 2** Existing cable termination to marshalling cabinet

and terminated directly in the new marshalling cabinets located in a new prefabricated SRR (see **Figure 2**).

This execution strategy is short-listed when the cable lengths are sufficient to be routed and connected to new marshalling racks located in a SRR.

However, there is unlikely to be sufficient space available in an existing SRR to accommodate both new marshalling and system cabinets. Not all the cables being terminated in this SRR would have sufficient lengths to be pulled to a new installation. Hence, the former execution method is more prevalent in use, even though it is the more costly.

### New system cable to be laid between old marshalling cabinets and new system cabinets

In this execution strategy, old marshalling cabinets are retained and a new system cable is laid within the control room between the old marshalling panels and new system cabinets. This approach is used only if there is sufficient space available in the existing control room for new system cabinets to be placed. This approach is suitable when shutdown has a reasonable duration as an existing marshalling cabinet will undergo revamp during shutdown to cater for new system requirements.

### Major issues during shutdown

In any of the execution strategies outlined above, some challenges are faced during shutdown.

### Short cable length

The existing multicore cable length can fall short for termination into either the transition junction box or the marshalling cabinet.

To mitigate short cable length, another junction box and associated multicore is added before the transition junction box to compensate for the short length of existing multicore cable. This may lead to additional cost due to new materials and it may be difficult to arrange such materials in very short time.

As was mentioned earlier, it is advisable to buy some spare standard size junction boxes and cables to counter such problems on site.

### Loop test issues

In the case of failure of a loop test, detect and diagnose the reason for discontinuity in the signal flow. Some possible reasons for failure of signal flow include:

- Loose cable termination in the loop: the faulty point can be identified after checking for continuity at each cable termination.
- The cable pair of existing multicore cable has failed due to degradation of insulation. In such a case, a healthy spare of cable can be utilised, if available. If the healthy spare is not available in the same multicore cable, then a new cable can be added between the existing junction box and another nearby junction box which has a healthy spare available.
- Single pair cable terminated between instrument and existing junction box failed due to degradation of insulation. In such a case, the solution has to be sought in accordance and agreement with the operation and maintenance team. Usually in such a case, the instrument was

no longer required and was discontinued for use by the operators but was not updated on the system documentation.

### Functional test issues

After loop testing, the operations team along with the automation vendor initiates logic testing, graphic testing on new consoles, and monitoring and assessment of alarm configurations.

At times, it is found that some of the loops, especially the sequential logic loops, need to undergo a change in logic according to operator requirement and actual process scenario. Such issues can be mitigated by the automation supplier's site team during commissioning.

### Conclusion

The role of a control system upgrade is crucial in facilitating processes to ensure optimum efficiency, improve productivity, maintain quality, and conform to safety and environmental regulatory standards.

The inherent risk associated with these projects relies heavily upon collaborating and sharing information transparently in a systematic, integrated approach and breaking down barriers between all enterprises.

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**Meghna Bahl** works in Project Management with Fluor Daniel India. She has more than 13 years of experience in the oil and gas industry, in feasibility studies, basic engineering design and detail engineering, procurement and field engineering relating to major projects in refining, chemicals, and petrochemicals. She has led various projects in India, USA, Canada, Netherlands, Mexico, and Kuwait as a Lead Engineer in the control systems function.

**Manisha Das** is a Control Systems Engineer with Fluor Daniel India. She is an instrumentation engineer with five years of experience in the FEED, detail engineering, and construction stages of oil and gas projects.

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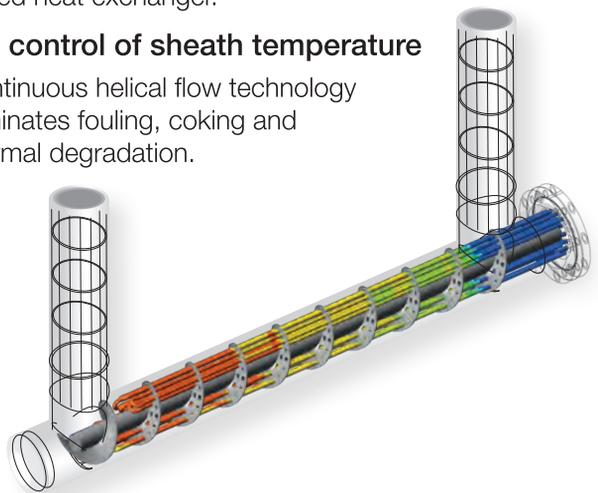
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# Increase duty in tube side condensers

A retrofit is shown to improve several aspects of tube side condensation processes

PETER DRÖGEMÜLLER  
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Tube side condensers are commonly used in refining. Allocation of the condensing fluid to the tube side can be driven by the requirement for low inventories, use of expensive corrosion resistant materials, or where air coolers are used. When condensing single component vapours, heat transfer coefficients are high. The situation changes when condensing multi-component mixtures or if condensation takes place with inert components. Under those conditions, the performance of condensers is often controlled by additional mass transport limitations between the liquid and vapour interface. To maintain the condensation process, the multi-component vapour must be cooled. This becomes a challenge since, particularly at low vapour velocities, the resulting sensible vapour cooling coefficient is low. In the case of horizontal in-tube condensation, as is found in air cooled condensers, at low vapour velocities towards the end of the condensation process stratified gravity-controlled flow is encountered. This can lead to thermodynamic non-equilibrium conditions which makes it difficult to correlate with standard heat exchanger design software.

Use of the hiTRAN thermal system improves several aspects of the tube side condensation process. The main benefit is the increased sensible vapour cooling coefficient. The wires also promote increased turbulence in the condensate film; in addition, film and vapour mixing contributes to a reduction in mass transport resistance. After explaining the underlying fundamentals of the use of hiTRAN thermal systems in multicomponent condensation, an industrial case study is presented where the systems were used to increase the performance of the condenser.

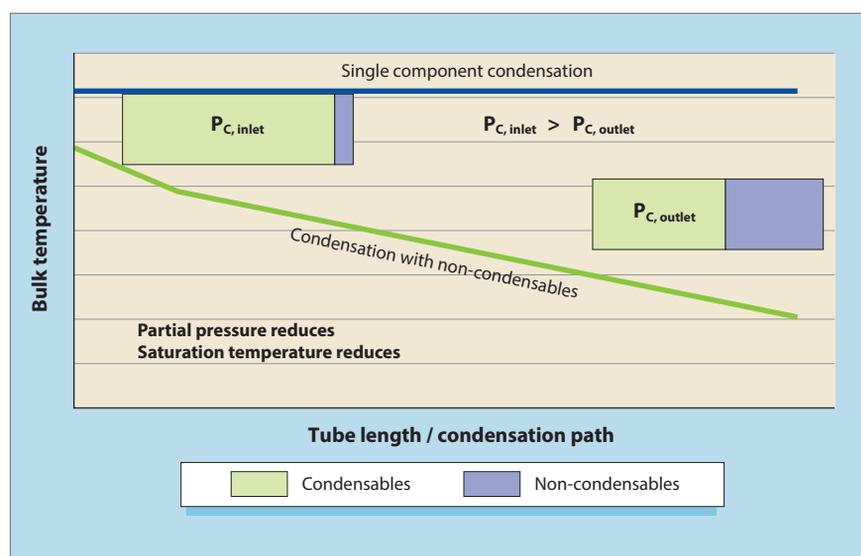


Figure 1 Temperatures along the condensation path for single component and condensation with non-condensables

When condensing multi-component mixtures, or if condensation takes place with inert components, there are two major differences compared to pure component condensation.

## Temperature profile along condensing path

When condensing a pure vapour, the condensation temperature is closely linked to the total pressure in the tube increment. It is therefore only influenced by frictional pressure losses and momentum losses or gains. In general, the condensation temperature remains almost constant along the tube length/condensation path (see Figure 1).

The situation is very different when condensing vapours containing non-condensable components in the applied temperature range. The condensing component depletes along the condensation path and, as a result, the partial pressure of this component and the associated saturation temperature reduces. To maintain the condensation process,

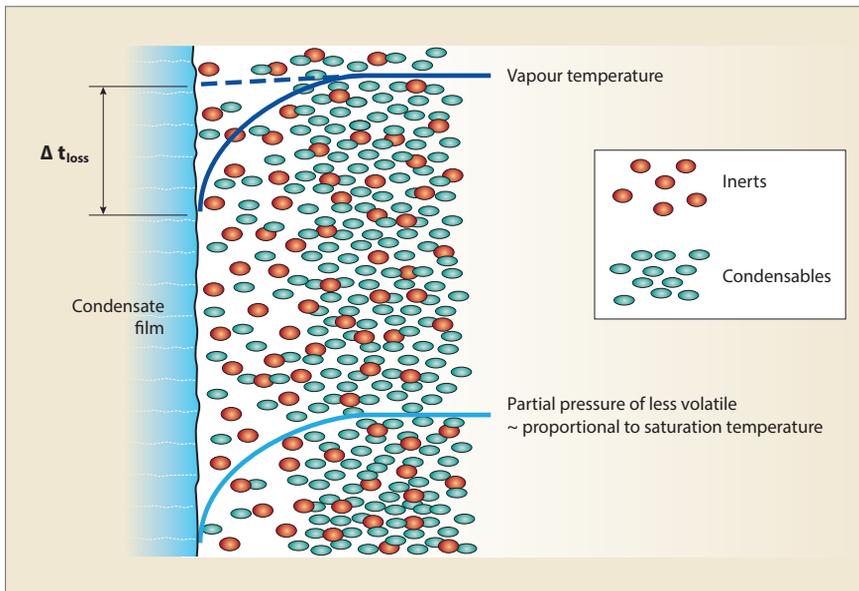
the multi-component vapour must be cooled accordingly. With constant cooling temperature this also equates to a loss in driving temperature difference between condensing vapour and cooling medium (see Figure 1).

Due to the reduction in volume, vapour velocities in condensers are lower towards the exchanger exit. With low resulting Reynolds numbers, the sensible vapour cooling coefficients will be low in this region. Therefore, heat transfer enhancement provides the greatest benefit towards the exit of the exchanger.

As the subsequent case study shows, hiTran enhancement can be installed partially. It can therefore target locations which benefit most from tube side enhancement.

## Temperature profile in a tube cross section

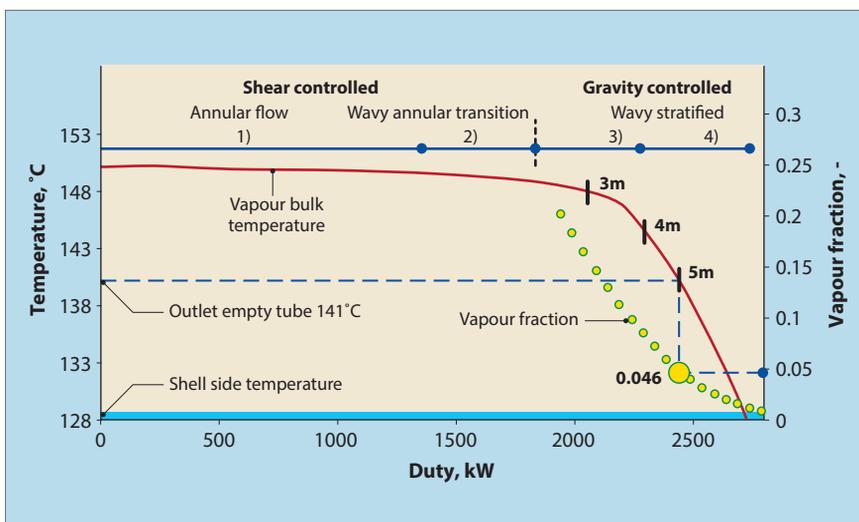
There is also a change in temperature profile over the tube cross section compared with pure component condensation.



**Figure 2** Temperature profile in the vapour-liquid interface with presence of non-condensables

The temperature profile in the plane along the condensate film and vapour flow in the presence of non-condensables is shown in **Figure 2**. Since condensable vapour components condense at the cold film, the mole fraction of inerts and more volatile vapour components increases at the interface between condensate film and vapour. As a result, the partial pressure of the condensing component reduces with reduced concentration nearer to the interface (light blue solid line). Since the saturation temperature of the condensing component is a direct reflection of local partial pressure, a characteristic temperature profile in the liquid

vapour interface is formed (dark blue solid line). Without a concentration gradient, the condensing temperature would be constant over the cross section of the tube and shifted to higher values (dark blue dotted line).<sup>1</sup> This indicates that, by ideal redistribution of the inert components to the bulk flow, the reduction in driving temperature difference ( $\Delta t_{\text{loss}}$ ) could be prevented. In highly turbulent vapour liquid interfaces, for instance with high vapour velocities, the concentration gradient is reduced. Again, at lower vapour velocities, towards the exit of the exchanger, concentration gradients are more noticeable.



**Figure 3** Condensing curve and vapour fraction along condensing path; in addition, flow regimes are indicated (numbering related to **Figures 6** and **7**)

## Case studies

### Retrofit of underperforming horizontal condenser (tube side)

The end user reported unwanted hydrocarbon vapour carry-over. Condensation took place on the tube side with evaporating water at a pressure of 2 bar (128.3°C) on the shell side.

The main condensing component, at over 98% mass, was aromatic hydrocarbons. The remaining vapour contained >1% carbon disulphide ( $\text{CS}_2$ ) and some water content. Operating pressure and temperature were reported as 120 kPa and 151°C inlet temperature respectively. At this temperature and pressure level,  $\text{CS}_2$  acted as a non-condensable. Design mass flow was given with 25 500 kg/hr. The exchanger was designed as a horizontal inclined BXM type, with 490 x 5m x 25mm x 2mm tubes.

The goal was to reduce hydrocarbon vapour carry-over by improving the cooling duty of the condenser.

Initial evaluations regarding the reported conditions were undertaken using HTRI Xchanger Suite<sup>2</sup> and Aspen Exchangers Design & Rating<sup>3</sup> software. The calculated results were similar and did also reasonably reflect the reported plant data.

In **Figure 3**, the integral condensation curve as a plot of temperature against cumulative heat removal rate is shown as a red solid line. The dotted line with the yellow/green markers shows the corresponding vapour fraction on the secondary y-axis. In addition, the constant shell side cooling temperature of the evaporating water is shown as a blue solid line. Over the first 3m, almost 85% of the vapour condenses. The depletion of condensing components causes a reduction in partial pressure (condensing temperature) which becomes very pronounced towards the end of the condensation process. This is reflected in a considerable reduction in bulk vapour temperature after 3m. The large yellow/green marker indicates that the calculated outlet vapour content was about 4.6%. The outlet temperature corresponds to about 141°C. Further reduction in vapour will require fur-

ther vapour cooling and associated loss in driving temperature difference to the cooling medium.

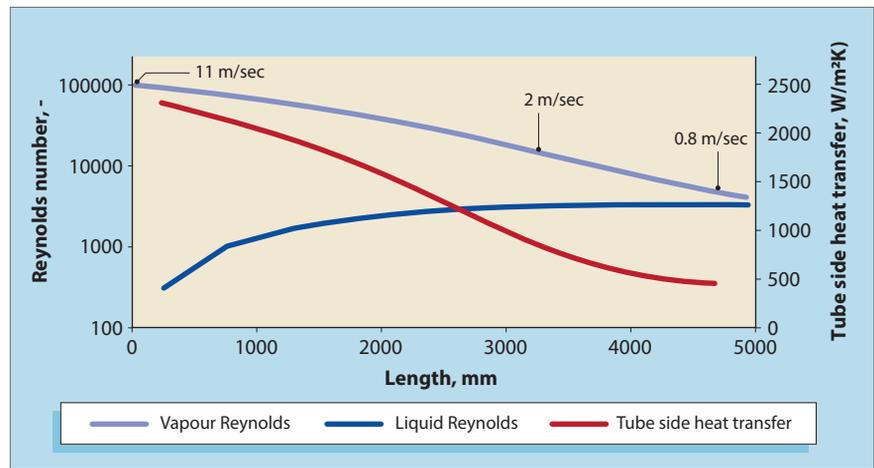
In addition, the capability to transfer heat is reduced by a change in flow regimes; this is also shown in the top section of the graph. Simulation results with both heat exchanger design packages indicate shear controlled annular and transitional flow regimes, induced by high vapour velocity, in the entrance section. After about 2.5m, this changes to a gravity controlled stratified wavy flow regime. For the last 500mm, Aspen reports even worse heat transfer conditions with stratified smooth flow.

In **Figure 4** the corresponding liquid and vapour Reynolds numbers are shown. It is evident that towards the end of the condensation process, the vapour velocity and associated Reynolds number are reduced considerably; in turn, the condensate flow is increased, leading to an increased liquid Reynolds number. However, since the density of the liquid is about 250 times higher compared to the vapour density, the condensate velocity and therefore the Reynolds number remain low. The liquid flow can be described as laminar to transitional. These conditions lead to very low heat transfer coefficients towards the end of the condenser.

### hiTRAN thermal system in condensing applications

The region towards the exit of the condenser was targeted in order to achieve higher condensing duty. In flow conditions characterised as above for vapour and liquid flow, hiTRAN can be used to increase the tube side heat transfer in this region. The technology is widely used in liquid and gas single phase applications. In these applications, enhancement levels up to 16-fold for laminar flows and up to four-fold for turbulent flows are possible.<sup>4</sup>

In condensing applications, the sensible heat transfer is of special importance when cooling of single phase vapour is required. Sensible liquid cooling is also required to cool the condensate in order to keep it as close as possible to thermal equilibrium with the vapour flow.



**Figure 4** Calculated flow Reynolds numbers and tube side heat transfer for case study

Since pressure drop equates to a loss of driving temperature difference between the condensing vapour and cooling medium, it is important when considering enhancement technology in condensing services to evaluate the impact of frictional and momentum pressure drop on overall performance. In horizontal two phase flow, the pressure along the flow pass is determined by frictional pressure drop and momentum pressure change.

For equal flow velocities, any enhancement device will generate a higher frictional pressure drop compared to an empty smooth tube. In contrast, the momentum or acceleration pressure change which is caused by phase change results in a pressure increase along the condensing path.

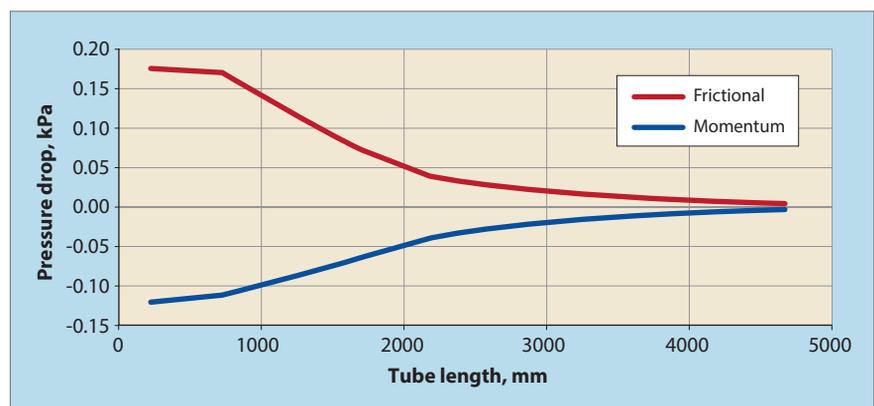
Simulated conditions in the hydrocarbon condenser are shown in **Figure 5**. It is evident that, as expected, the frictional pressure drop is highest in the inlet sec-

tion with high vapour velocities. Towards the condenser outlet, with low vapour velocities and stratified flow, the frictional pressure drop in the empty tube reduces to less than one tenth of the inlet pressure drop. This is very low and offset by momentum recovery. Since the frictional pressure drop is very low, the pressure drop penalty for using hiTRAN enhancement is also low.

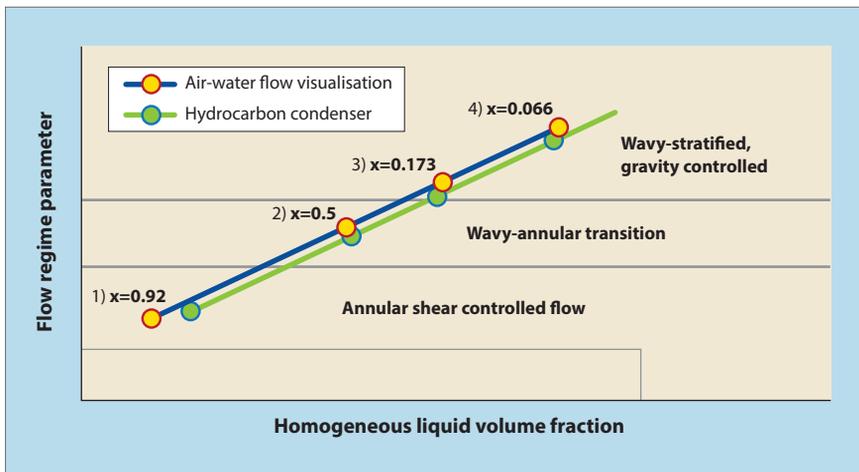
For this application, we also see considerable momentum recovery, highlighting the fact momentum recovery can offset the loss in frictional pressure drop.

### Flow patterns for this application

In order to understand the impact of hiTRAN technology for this kind of application, it is important to understand the changes in hydrodynamic for different flow regimes. Calgavin performed extensive in-house research into two phase pressure drop and flow visualisation in hiTRAN enhanced horizontal flow. **Figure 7** shows photographic images



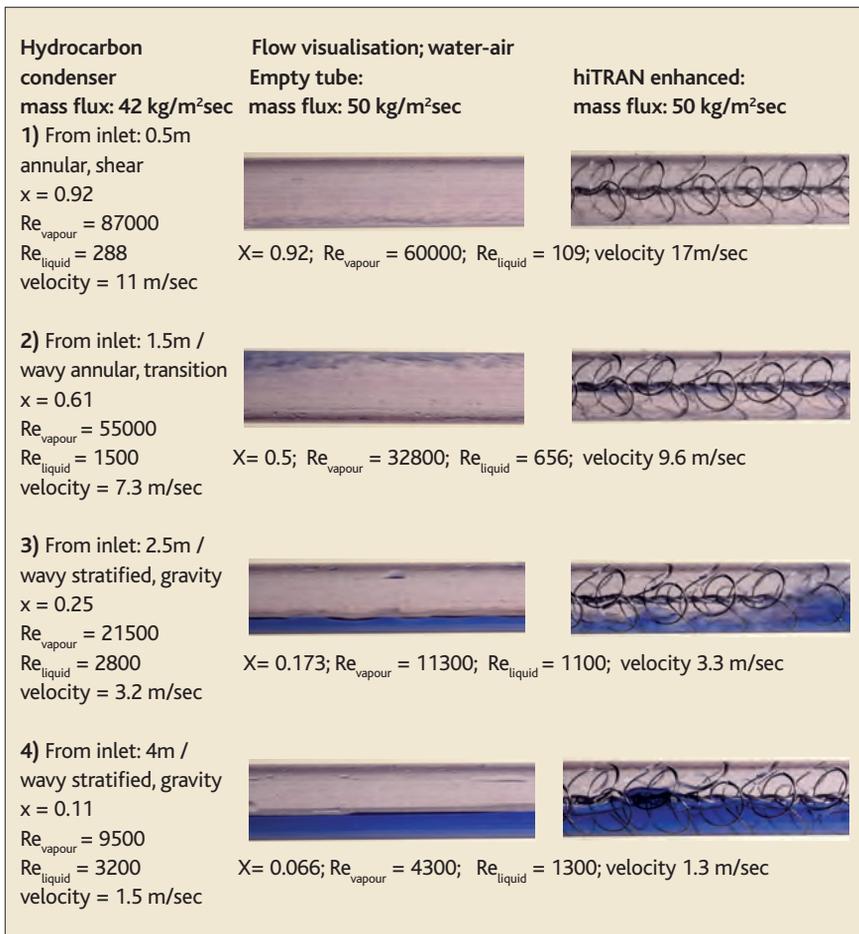
**Figure 5** Momentum and frictional pressure drop along flow path in hydrocarbon condenser (empty tube)



**Figure 6** Flow map comparing flow conditions in air-water experiments with hydrocarbon condenser

of the anticipated flow patterns in the hydrocarbon condenser with and without hiTRAN enhancement. Since the flow visualisation results were undertaken with air-water as the two phase systems, flow maps do provide a useful tool to identify like for like flow conditions compared to the hydrocarbon condenser.<sup>5,6</sup>

Simulations were done using the HTRI Xchanger suite, and a software integrated flow map<sup>2</sup> was used in order to map the simulation results with the flow map results for the air-water flow experiments. As **Figure 6** shows, Calgavin’s visualisation data for air-water and a mass flux of 50 kg/m<sup>2</sup>sec gave a very close match to the anticipated flow



**Figure 7** Photographed flow conditions in air-water experiments compared with process conditions in a hydrocarbon condenser at similar points in the flow map (see **Figure 6**)

conditions in the condenser. This is not unexpected since the density differences of vapour and condensate (260 times) and air and water (380 times) at experimental conditions are not too far apart.

The red/yellow markers for different air mass fractions in the air-water visualisation experiments are represented in the photographs in **Figure 7**. They are also referenced in **Figure 3**.

It is notable that the simulated flow regimes in the flow map do reflect very well the photographed conditions of the air-water flow shown in **Figure 7**, with flows ranging from shear controlled annular to gravity controlled wavy stratified flow.

For annular and wavy transitional flow, the photographs show, as expected for the empty tube, highly turbulent and well mixed flow conditions, suggesting high heat transfer and pressure drop. Installing hiTRAN in such flow conditions does not provide benefits. The situation changes notably once the flow regime switches to gravity controlled. This happens after about 2.5m and the flow becomes stratified. In item 3 in **Figure 7** we see little turbulence in the liquid flow; there is also a clearly defined interface between gas (vapour) and liquid with little mixing.

With hiTRAN installed, the flow condition changes notably. Much more complex velocity field and induced turbulence causes visible disturbance in the liquid flow. It is notable that intermittent slugs, which are also present in empty tube flow at this flow condition, are more frequent and have a higher amplitude. Those slugs play a vital role to reduce a thermodynamic non-equilibrium between the phases.

As expected, the situation worsens toward the end of the condensing process in item 4. Here the interaction between the phases reduces further.

As **Figure 3** shows, to achieve lower vapour outlet content, the vapour and condensate need to be cooled further. Since vapour and liquid Reynolds number indicate laminar and transitional flow respectively, this is difficult to



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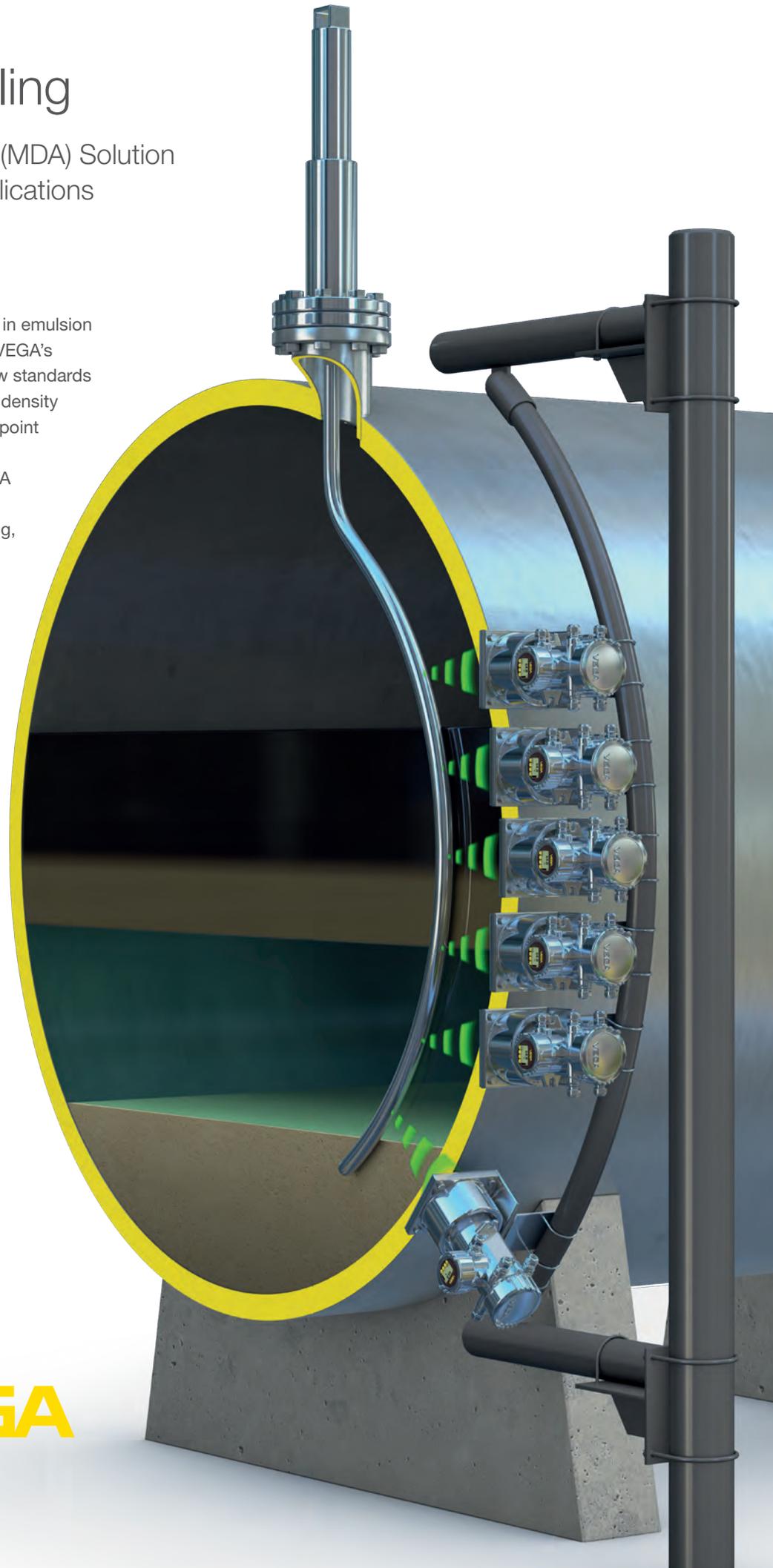


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achieve without enhancement. In addition, those flow conditions promote the build-up of non-condensables at the vapour liquid interface (see **Figure 2**). This leads to a loss in driving temperature difference. It should be noted that due to different properties between vapour and liquid, and difference in flow velocities, there will be a thermodynamic non-equilibrium between the phases. In fact, for flow conditions where vapour and condensate are separated, the integral condensation curve shown in **Figure 3** loses its validity; instead, a differential condensation curve would be more appropriate.<sup>7</sup> This analysis is complex and difficult to simulate with heat exchanger design software.

### Installation of hiTRAN thermal systems and feedback

Based on the above evaluation, it was decided to install hiTRAN thermal systems over the last 2m in the gravity-controlled flow area. The insert geometry, in terms of wire size and matrix density, was chosen in such a way that the pressure drop requirements were fulfilled.

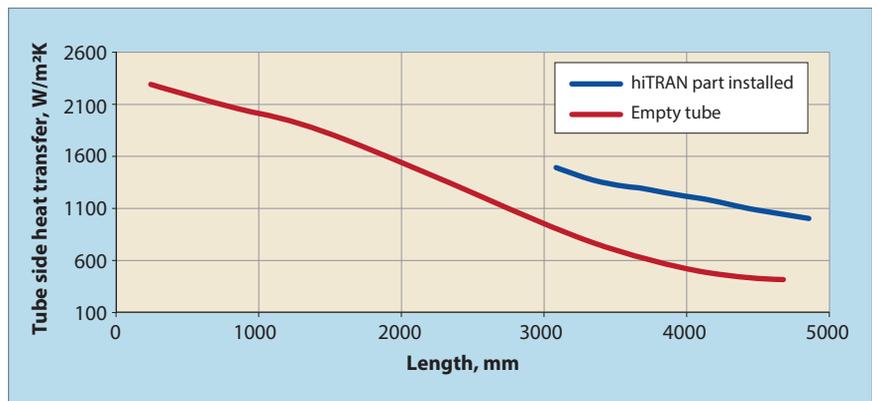
In-house simulations showed that heat transfer, with hiTRAN in the gravity-controlled section, more than doubled, with a calculated outlet temperature about 5°C lower compared to the empty tube (see **Figure 8**).

Installation was straightforward and undertaken on-site in two working days. The unit with removed headers is shown in **Figure 9**.

After start-up, measured plant data confirmed a reduced outlet temperature of about 5°C. According to the simulations, and from **Figure 3**, this corresponds to a reduction in vapour mass fraction at the outlet from 4.6% to about 2.5%.

With the design mass flow, this equates to a reduction in vapour carry-over of about 535 kg/hr.

To achieve a similar outlet temperature and vapour content in an empty tube design, simulations show that the tube length would need to be extended from 5m to 8m. Apart from the cost implications, this option was not available due to the limited plot space for the revamp.



**Figure 8** Heat transfer before and after installation of hiTRAN



**Figure 9** Hydrocarbon condenser with hiTRAN thermal system installed and secured

### Conclusion

It has been shown that hiTRAN thermal systems can be applied successfully in horizontal tube side condensation. It is important to determine whether partial installation, or installation over the whole tube length, is more beneficial. To do this, incremental evaluation along the flow path is required. Under stratified and wavy stratified flow conditions in the gravity controlled flow regime, hiTRAN thermal systems are beneficial. Sensible heat transfer is increased in the vapour flow and in the slow flowing condensate film. This increases the potential for heat removal under those flow conditions. Flow visualisation also shows that flow and vapour are homogenised, preventing thermodynamic non-equilibria under those flow conditions.

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### Glossary

Pc [kPa] Partial pressure of condensing component  
 $\Delta t$  [°C] Temperature difference

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# Downtime damages environmental performance too

Advanced technologies can reduce the environmental impact of unexpected shutdowns

JOHN HAGUE  
Aspen Technology

**W**hile unplanned downtime will always impact productivity and profitability in an oil refinery or petrochemical plant, the effects of unexpected stoppages often have implications that reach beyond financial. Safety is critically important, but so too is environmental efficiency. Today, as energy providers face increasing pressure to set and meet sustainability targets and reduce emissions, there is a growing focus on the environmental impact of unexpected shutdowns at oil and gas refineries.

Plant downtime is highly damaging in this context, with a single, unplanned shutdown lasting just hours leading to the release of a year's worth of toxins into the atmosphere for example. An emission event following a forced shutdown at a California refinery in 2017 resulted in 31 000 lbs of sulphur dioxide being released in one day – more than the refinery had released over 2015 and 2016 combined. This is just one of many such examples.

And that is in addition to the losses in profitability that we know these events result in, stemming from reduced productivity, higher maintenance costs, and the waste that comes from irregular operations. If there is a single plant process that illustrates this issue clearly, it is gas flaring, or the combustion of excess product that is typically released when a plant experiences over-pressuring operation, such as during an unplanned shutdown. Excessive flaring is a visual sign that something is outside of normal parameters in the facility, which means the safety risk is increased.

Flaring is also a significant source of greenhouse gas emissions. In fact, according to satellite data published by the World Bank's Global Gas Flaring Reduction (GGFR) programme, each year 145 billion cubic metres of gas is released into the atmosphere from gas flaring. That is equivalent to 270 million tonnes of CO<sub>2</sub> emissions per year.

These figures paint a grim picture, but there is good news from the industrial technology front. By tapping into the power of machine learning and predictive analytics, companies can begin to reduce unplanned upsets and capture all the benefits that come with that.

**What if it was possible to know which pieces of equipment are going to fail and when, so repairs could be performed as part of a managed shutdown?**

With technology that eliminates the surprise of unplanned downtime, companies can minimise the most dangerous conditions, reduce the amount of gases released into the environment, and realise significant financial gains by maximising uptime.

Without question, there is a lot at stake – financially and beyond – in avoiding unexpected shutdowns.

**Technology for decision agility**  
So, what if energy companies could

actually plan for downtime? What if it was possible to know which pieces of equipment are going to fail and when, so repairs could be performed as part of a managed shutdown? The benefits are significant, for both improvements in emissions reduction and profitability.

Today's asset performance management technology can deliver advanced warning of failures through a combination of predictive and prescriptive analytics, enabled by integrated software that incorporates artificial intelligence (AI) and machine learning. This type of solution provides a detailed view of all equipment, systems, facilities,, and networks, thereby enabling a capability we call 'decision agility'.

This means that, with the time to plan around predicted downtime and a holistic view of the operation, plant personnel can see exactly how a decision that changes any business process also affects the entire organisation. They will immediately know how it impacts planning and scheduling, how it determines which feedstocks are purchased, how it affects inventory, and even how it may impact the sales team and the potential for missed orders.

The right technology can simulate how any event will impact the system, the process, and the asset. When the outcome is known in advance, operators and engineers can collaborate to make the safest and most profitable decisions; they can work together to develop a plan. That plan becomes a clear roadmap of where to spend every pound to maximise the return on capital employed. The technology can even be scaled to cover multiple plants across a region to pro-

vide a look at how facilities are tied together and to better understand their co-dependencies.

So, when there is an issue in one location, the software can show how it will affect the pipeline coming in, the ships going out, and whether the facility is at risk of defaulting on any contracts.

By driving the best decisions, this technology also reduces risk across the entire operation, and there is a recognised value in doing that. Some providers in the insurance industry, which is also driven by data, have actually begun advising their customers about digital solutions for prescriptive maintenance and decision support. They are promoting these technologies as ways to reduce unplanned downtime and associated events — and also as an incentive to lower their insurance rates.

The ability to see wide and deep enables new ways of running the business. Digital transformation is knocking down the data silos and delivering the tools necessary to make sense of the data available at the enterprise scale.

### Putting it into practice

Achieving this level of technological integration starts with a ramping up of the organisation's digital capabilities. Companies in every sector now have access to technologies such as high performance computing, artificial intelligence, and advanced analytics to generate deeper insights from their operating data.

Fuelled by these data-driven insights, leading edge simulation programs enable operators to quantify the true value or cost of any renovation or improvement project, maintenance change, operations improvement, or supply chain constraint. This technology utilises statistical sampling techniques to predict the future performance of a system, analysing equipment behaviour patterns to derive a 'time to failure' estimate.

With the broad view of operations that simulation programs provide, plant personnel can be alerted to impending failures and understand the potential impacts to the wider

system. Operators can also model flow through the pipes and tank levels, as well as the utilised and available capacities of all units.

This is how it is possible to discover exactly which events are robbing an operation of money or negatively impacting performance in ways that can lead to environmental issues, for example. With a prioritised list of every single event in the business that is negatively impacting performance, the company can apportion budgets and put people where they are needed — and every decision is based on data.

If the software is in place at a refinery for example, it might alert to a failure of a fluid catalytic cracker or part of a cooling tower, likely to occur within the month, which would cause significant disruptions throughout the business. But with the advance notice provided by the software and time to

## Beyond the sustainability implications, companies also stand to gain financially as a result of the increased production that comes with more uptime

plan before the failure happens, personnel can then use scheduling models to find the best time to take that part of the plant offline, and even insert additional maintenance activities to make the most out of the planned downtime.

And if the alert comes even further in advance, perhaps six to eight weeks ahead of failure, this enables the staff to load the information into a longer term planning model that can account for impacts on sales or operations planning or integrative business planning.

Through those two models, not only is the organisation protecting itself from unplanned events, it is actually accounting for economic

impacts. Personnel are making informed decisions to take the best possible action in a multi-network supply chain with equivalent manufacturing facilities that can produce multiple goods. The greater the window of predictability over the planning horizon, the more powerful the business options are. This moves the conversation from "what's feasible?" or "what's going to get us by?" to "what's going to give us the best results as we're trying to deal with this issue?"

In short, management can always know when is the best time to take downtime, as well as what activities should be completed during that downtime, in order to preserve orders and maintain commitments to key customers as well as keeping emissions to a minimum.

### Scoping the results

An investment in the right advanced technologies not only delivers a significant return on investment by reducing unplanned downtime, but it also greatly improves a company's ability to maintain safe operations and meet environmental goals.

As we have already highlighted, unplanned downtime and transient conditions lead to flaring, which means product is released into the atmosphere. This is an area where predictive analytics integrated throughout the system can make a major impact. It seems clear that a large proportion of the overall emissions caused by gas flaring every year could be avoided by eliminating unplanned shutdowns.

What if we could see a pending problem on a piece of mining equipment before it starts degrading? What if we could be alerted to a failure on an offshore oil platform before it developed into a serious issue? All of this is possible, and it means that unplanned downtime can be turned into planned downtime.

In addition, this technology can be integrated with planning models that will provide specific recommendations and advice on what actions to take to maintain greenhouse gas emission limits and quotas.

## Maximise availability-to-plan for optimum operations

Beyond the sustainability implications, companies also stand to gain financially as a result of the increased production that comes with more uptime. Those that have optimised their maintenance processes to reduce unplanned shutdowns have realised incredible payback on their investment in predictive analytics technology.

For example, one refinery suffering from repeated hydrogen compressor failures was able to reduce shutdown time by eight days thanks to a 35-day time-to-failure prediction. In addition, the cost for planned maintenance was less than 30% of the cost of emergency repairs. Other oil and gas companies using predictive maintenance solutions have similar stories.

Unplanned shutdowns cost oil and gas companies significant sums every year. One source estimates that equipment failures causing unplanned downtime cost oil and gas companies an annual average of \$42 million and up to \$88 million in the worst case scenarios. The US Department of Energy reported 1700 shutdowns at refineries between 2006 and 2017; 46% were due to mechanical breakdown.

Just eliminating a portion of the abnormal events that rob an operation of productivity can add millions of dollars to the bottom line. And when companies are able to quantify exactly how much any particular event affects revenue, they know exactly where to target their technology strategy for maximum impact.

## Conclusion

Companies that implement this technology first can put themselves at a distinct competitive advantage, reaching new levels of profitability while also maintaining their 'social licence to operate' with improved safety and sustainability performance. Many are already putting the solutions in place to help them avoid the most dangerous conditions, reduce greenhouse gas emissions and maintain the most efficient operations.

As companies face growing pressures from shareholders, regulators, and consumers alike, the need for agility may well be greater than ever but so too is the need for sustainability and environmental efficiency. By reducing risk and uncertainty through the implementation of the advanced technology solutions available today, companies can put themselves in the best position to win in the marketplace of tomorrow.

**John Hague** is Executive Vice President, Operations, at Aspen Technology, Inc. He is responsible for the Global Sales, Customer Success, Product Marketing and Partners organisations, as well as the Asset Performance Management (APM) business unit. He joined AspenTech in 1995 and has previously held a number of senior management roles, including senior vice president, Middle East and head of global accounts. Prior to joining AspenTech, he worked for Conoco, Inc., Advanced Pipeline Technologies, Inc. and Scientific Software-Intercomp (SSI). He holds both BS and MBA degrees from Oklahoma State University.

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# Back-up for water treatment

## Mobile water services can provide support for refiners facing the challenge of ageing water treatment plants

MARK DYSON

*Veolia Mobile Water Services*

**A**cross Europe, in heavy industries like power, chemical, petrochemical and refining, operators are facing an increasing challenge from their ageing assets including water treatment plants. Health and safety guidance cites the potential degradation of plant and equipment due to age related mechanisms such as corrosion, erosion, and fatigue as a key issue for industry. In addition to safety concerns, ageing plants may also impact continuity of production. This can be seen in permanent on-site water treatment facilities. Water is a vital utility in many heavy industry processes and the demand for it is increasing with population and industry growth. Ageing plants are not always reliable or efficient and often emergency maintenance is required to avoid disruption to production which could result in costly downtime for operating sites.

Emergency provision of temporary water treatment is well established. However, more and more operators are recognising and seeking the need for a more 'permanent temporary' solution. Mobile water service providers can also bridge a gap for operators with a problematic plant until it can be replaced or repaired or even to bridge a time period if there are variations in feed water supply or quality.

Operators across several industries in Europe have worked with temporary water treatment service providers to provide a flexible longer term solution to tackle the reliability issue around the permanent ageing plant and even extend its potential life cycle. Preventative maintenance and refurbishment of a permanent water treatment plant needs to be carried out more fre-

quently, without interrupting the production schedule and not during a planned shutdown. A temporary water treatment system may be deployed to provide either full or partial replacement of the various processes during this interim need. Typical applications include resin replacement, pressure vessel maintenance, controls upgrades, reverse osmosis membrane cleaning or replacement, maintenance on chemical dosing equipment, and work on waste treatment plants.

Longer term asset rental is an attractive alternative to high upfront investment in permanent water treatment technology across many sectors. Tightening capex budgets, an emphasis on business continuity, and a desire for flexible, affordable water management have all generated a demand for mobile water services, which offer a cost-effective, alternative solution to procuring new installations for upgrading existing infrastructure, as well as providing emergency relief and fulfilling temporary water requirements.

### **An evolving solution**

A typical mobile plant may consist of two or three assets or skid-mounted systems, which can be assembled in a plug-and-play fashion. A typical set-up involves initial pretreatment by multi-media filtration or granular activated carbon adsorption, followed by reverse osmosis in a second trailer, and subsequent mixed bed ion exchange polishing in a third.

The modular design of many of today's mobile water services allows a variety of process configurations to be combined, making it possible to treat towns' mains,

borehole, river and reservoir water, and even wastewater resources. The containers' portability enables them to be positioned to make the best use of the available space, eliminating or reducing the need for building infrastructure to house the equipment. Any number of assets can be operated in parallel or in series to provide the required flow rate, and their modularity allows additional components or treatment steps to be added for extra functionality or increased throughput, even if it is only needed for a short period of time. Storage tanks and pumps can also be provided – together with interconnecting fixed pipework or flexible hoses, water meters, and fittings – and mobile generators support a completely standalone set-up. A final, and perhaps the most important, consideration is that these modular units can be easily exchanged over time for the latest, updated technology, ensuring that a company's water treatment systems remain at the cutting edge and benefit from the most cost-effective, available solution from its mobile water services supplier.

### **Capex considerations**

Plant operators face a number of current challenges. Environmental targets, changes in production demand, and tightening budgets have all made it more difficult to make a strong case for capital investment. The fall-out from this has been a reluctance to invest in projects where lifetime and return on investment are uncertain. In light of this, one of the attractions of mobile water services is their flexibility, as the rental payments can be covered by the operational budget,



Figure 1 Mobile water demineralisation plant

removing the need to raise capital. Mobile water service suppliers are often willing to enter into pay-as-you-go, multi-year contracts, which help to improve financial planning thanks to predictable, regular payments.

### Planned maintenance and turnarounds

A temporary water treatment system is a perfect solution in an emergency and can sustain a continuous supply of treated water for all unanticipated scenarios, such as coping with short-term demands. However, many facilities will also need to plan for maintenance of existing water systems, and mobile

water services can be brought in to cover equipment servicing, ensuring that production or business processes can continue and avoiding costly downtime. In some instances, a facility may need to cope with seasonal or unexpected changes to its raw water supply.

Scheduled turnarounds may involve a total suspension of operational activities. A turnaround that exceeds its timeline or budget can have serious financial consequences, so it is essential that an efficient, reliable water supply is available as needed. In these instances, mobile water services can be brought in to support all maintenance and cleaning activities, as demonstrated by

the experience of another multinational oil and gas company. A turnaround had been scheduled and, during the subsequent start-up phase, the refinery needed an extra back-up supply of demineralised water – 100 m<sup>3</sup>/h in operation and 100 m<sup>3</sup>/h in standby – running in parallel to its own demineralised water plant. Veolia Mobile Water Services provided a four-trailer configuration (see **Figure 1**) to guarantee the water supply and meet the water specifications, including a conductivity of <0.1 μS/cm and <10 ppm of SiO<sub>2</sub>. This short-term intervention offered a reliable and secure back-up and, as a plan and agreement was already in place, Veolia Mobile Water Services was able to offer fast deployment and commissioning.

### Awareness is key

There are numerous benefits that mobile water services can bring to refiners facing the challenge of ageing plants. Raising knowledge and awareness is key for mobile water services suppliers over the next couple of years. A distinct shift in outlook is necessary to help companies to transition from seeing temporary water services as only an emergency service provider, to understanding the value in a longer term solution to an ongoing issue. Mobile water services represent a sensible alternative to capital investment. As awareness grows, we can expect to see more mobile water services being implemented, supporting financial planning, ensuring business continuity, and helping to maintain resilient and effective water treatment plants.

**Mark Dyson** is the General Manager of Veolia Mobile Water Services. Experienced in both emerging and developed markets, he has been with Veolia for 11 years, leading the sales and marketing teams for Veolia's Industrial, Scientific and Healthcare process water division in the UK, leading the technical community for the development of new services and products across Europe and also managing the commercial and operations teams for the Mobile Water Services team across Europe. He holds a BSc in engineering and management from Brunel University and an MBA from the Open University.



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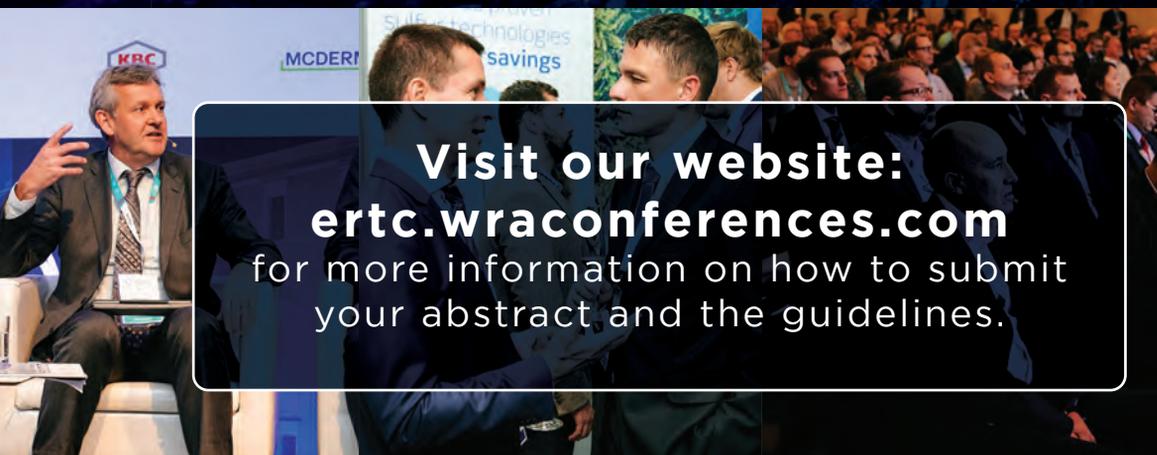
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# Petrochemicals from refinery intermediates – beyond polyolefins

Diversifying chemical synthesis routes based on simple olefins can add major value for an integrated refining-petrochemical complex

SUSHREE CHAUDHURI and ISHNEET KAUR NARANG  
Bechtel India

Economic growth curves indicate plateauing of fossil fuel demand and a significant increase in chemicals and petrochemicals in the coming decades. Oil to chemicals nonetheless directionally results in an increase in refining capacities to cater for increased demand.

Refining configurations are changing as bottoms upgrade and rigorous hydrotreating have become a necessity with a heavier and sourer crude diet coupled with stringent land and marine environmental emission specifications. The addition of bottoms upgrading by cracking, coking or catalytic conversion adds to the value chain but at the cost of significant capital investment. In this way, refining margins are becoming self-limiting if we only focus on producing fossil fuels for the market. The profit and pay-back values significantly increase if we start looking beyond conventional refining and petrochemicals operating models.

A world scale polymer complex downstream of a fuels refinery, fed by a petro-FCC or a steam cracker, adds significant value to the refining bottom line per barrel of crude. Bottom line can be improved further by valorisation of the refinery and steam cracker products and intermediates like off-gases,  $C_2/C_3$ , and aromatics via fuel to chemicals, deep chemical conversion, and diversion into specialty chemicals.

The petrochemical industry has seen more than 50% growth in this century with 100 million t/y of ethylene production in 2000, rising up to 150 million t/y of ethylene in 2016. The value curve also has been quite encouraging. Apart from

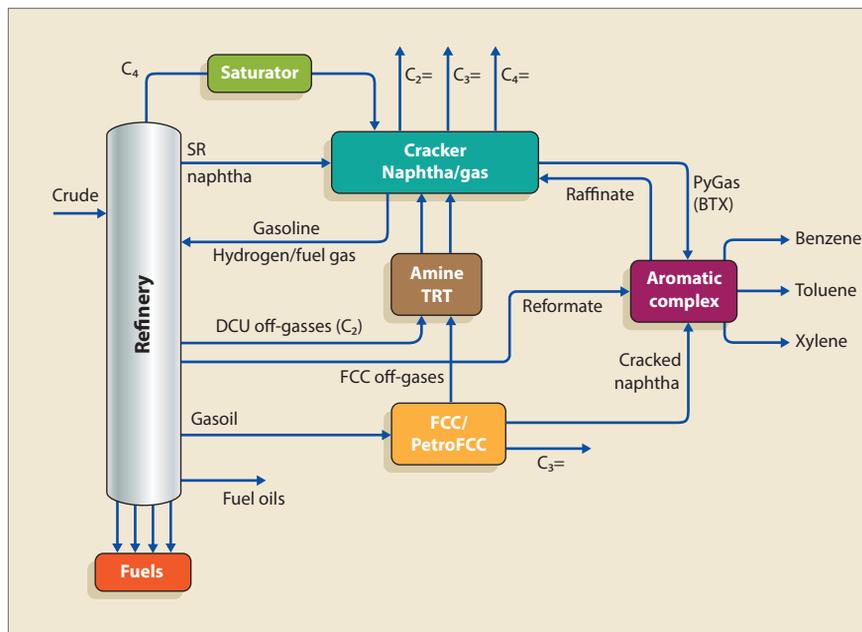


Figure 1 A conceptual hybrid refinery configuration

strong demand growth there have been other factors which have positively influenced the petrochemicals bottom line. Two major factors that can be attributed as positive economic influencers are feedstock diversity and pricing, and market geography and structure.

Feedstock pricing has been favourable to petrochemical manufacturers because of two major trends over the last few years. On the one hand, discovery of shale gases has supplied the petrochemical industry with an abundant alternative feedstock which has been extremely beneficial for the economics of  $C_1$  and  $C_2$  derivatives. On the other hand, with the sharp fall in crude oil prices in 2014, oil based feedstocks like naphtha have been preferentially diverted to petrochemicals to cash in on the more promising margins. This means that

more naphtha has been available as a petrochemical feedstock at a cheaper price, boosting both scale and margins.

Emerging economies like China and India have shown strong growth over the last decade. For the coming years, India's petrochemical industry features stronger growth forecast than China's. These emerging markets have been catered to by domestic growth of petrochemicals production and also have been the destination of export for Middle Eastern and American petrochemical producers, which already had a geographical advantage in accessing shale gases. However, the value created by  $C_2/C_3$  derivatives has also been subjected to margin erosion due to over-production by new chemical producers in the emerging markets, benzene being a good example. India has been a

Conventional FCC vs petro-FCC yield		
Component	Refinery with conventional FCC, wt%	Refinery with Petro-FCC, wt%
H <sub>2</sub> S, H <sub>2</sub> , C <sub>1</sub> & C <sub>2</sub>	2	3
C <sub>2</sub> =	1	6
C <sub>3</sub>	1.8	2
C <sub>3</sub> =	4.7	22
C <sub>4</sub>	4.5	5
C <sub>4</sub> =	6.5	14
Naphtha	53.5	28
Distillate	14	9.5
Fuel oil	7	5
Coke	5	5.5

Table 1

Steam cracking vs FCC yield		
Component	Steam cracking yield, wt% (typical)	FCC yield, wt% (typical)
H <sub>2</sub> , C <sub>1</sub> & C <sub>2</sub>	18	3
C <sub>2</sub> =	25	6
C <sub>3</sub> =	14	29
C <sub>4</sub> =	11	14
Gasoline	15	23
LCO & PO	17	25

Table 2

net importer of many chemicals and petrochemicals, except for benzene for which it is a net exporter. The export rate is likely to further increase in the coming decade.

### Configurations and intermediates: a bird's eye view

Refining and petrochemical producers have been strategising to keep up with the linear growth curve. On the refinery side, heavy end cracking and bottoms upgrade projects have been planned and undertaken to cater to demand for

lighter feedstocks downstream. Gas based crackers are being backed up by propane dehydrogenation to boost the C<sub>3</sub> value chain. At the midstream, many refiners are revamping their FCCs to petro-FCC to boost propylene production and invest in the C<sub>3</sub> value chain. Increasing interest is being demonstrated for diversification or deep chemical conversion. **Figure 1** shows a conceptual hybrid refinery which demonstrates the possibilities for integrating its intermediate streams with a petrochemical complex, all of

which is not necessarily practical for a single complex.

This configuration explores a petro-FCC alongside a cracker. Petro-FCC uses higher temperatures, less residence time and more catalyst to selectively produce more propylene instead of naphtha. A petro-FCC is a fairly small capital investment when compared to a cracker. **Table 1** demonstrates typical FCC vs petro-FCC yields for a light naphtha feedstock. Petro-FCC produces more propylene at a cost to naphtha when compared to conventional FCC yields. **Table 2** compares typical steam cracking vs FCC product yields for a light naphtha feedstock. From C<sub>3</sub> onwards, petro-FCC has a product slate comparable with that of a steam cracker. C<sub>1</sub> and C<sub>2</sub> from petro-FCC would be close to negligible to sustain any world scale downstream petrochemical unit.

The configuration also explores the integration of raffinates, reformates, off-gases, and fuel cuts such as naphtha, pygas, and gasoil. These intermediate products are converted to valuable petrochemical building blocks like ethylene, propylene, and C<sub>4</sub>s which can be further diversified into polyolefins, rubbers, resins, and specialty chemicals, enriching the product value chain but understandably with the premium of increased capital investment.

The primary petrochemical building blocks like ethylene, propylene, C<sub>4</sub>s, and benzene can be diversified in many ways to derive more value. A chemical product bouquet can be selected from the various routes available via a techno-economic evaluation. Internal rate of return (IRR) for a modern petrochemical complex in Asia can be as high as 15% with right investment strategies. However, the results are often driven by owner-partners' appetite for risk and their experience in the particular chemical or technology.

### Ethylene derivatives

Polyethylene dominates the ethylene derivatives market with a share of approximately 60%. The second largest derivative is ethylene oxide. **Figure 2** demonstrates opportunities

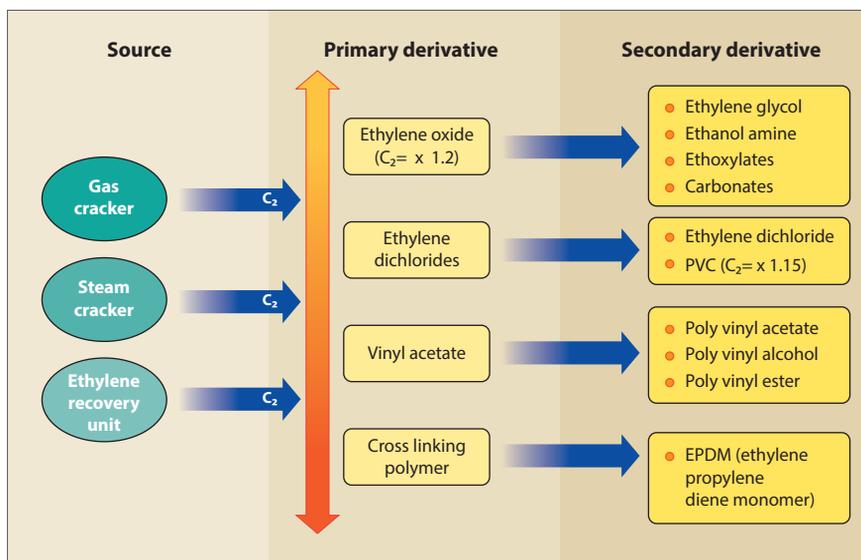


Figure 2 Ethylene diversification opportunities beyond polyethylene

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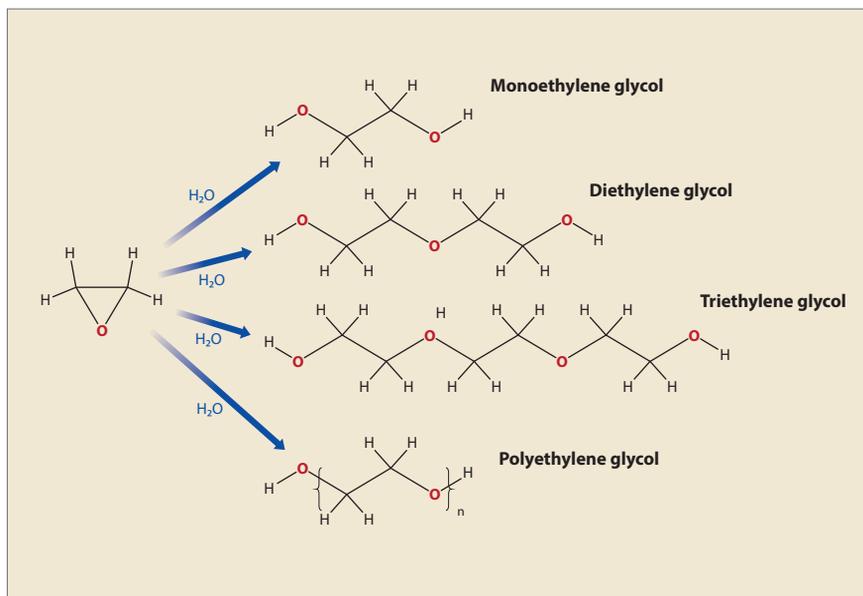


Figure 3 Ethylene oxide can be hydrolysed to ethylene glycols

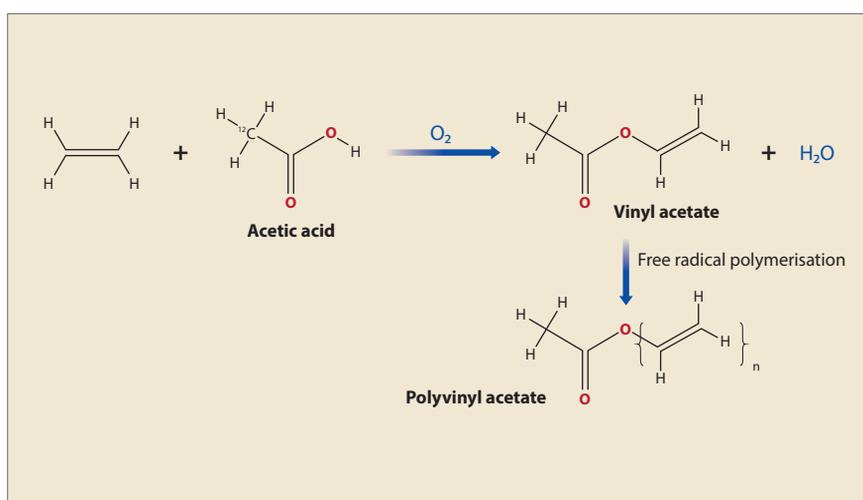


Figure 4 Vinyl acetate is produced by reaction of ethylene and acetic acid

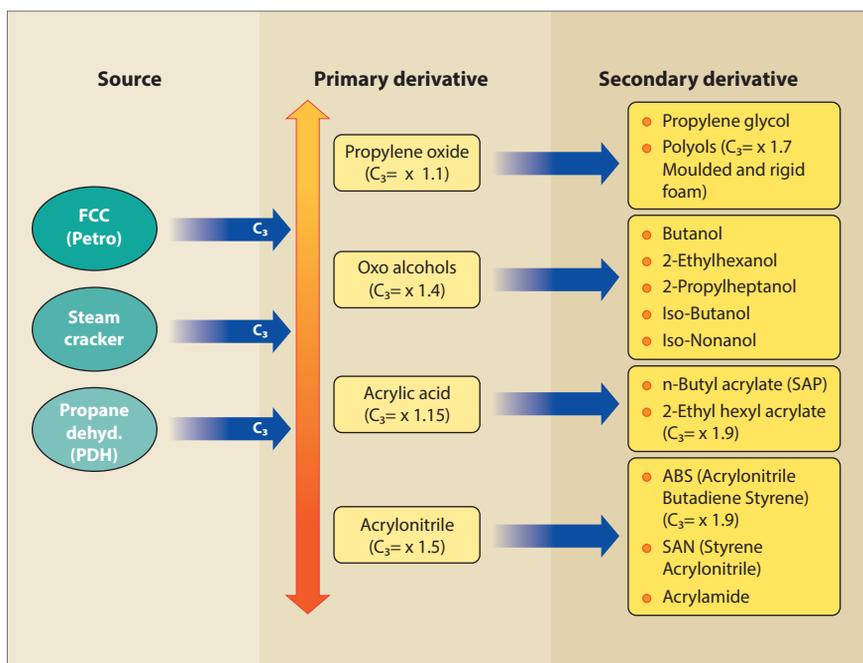


Figure 5 Routes for propylene derivatives

for ethylene diversification beyond polyethylene, and ideas regarding value improvement.

The following section discusses various ethylene derivatives and chemical reactions leading to end products. The reactions are indicative only.

Ethylene is often diversified via the ethylene oxide route. Ethylene oxide is produced by controlled oxidation of ethylene over silver catalyst. Ethylene oxide can be hydrolysed to form ethylene glycol which has an end use as antifreeze (see Figure 3).

Ethylene oxide can be also converted to ethanol amines by reaction with ammonia. Ethanol amines are used as detergents and agrochemicals.

Ethylene oxide can be converted to ethoxylates by direct reaction of higher alcohols, acids or amines in the presence of an alkaline catalyst. Ethoxylates find their end use in cosmetics, detergents, and intermediates for surfactants, to name but a few.

Ethylene can also be converted to linear alpha olefins which can be converted to fatty alcohols and linear alkyl benzene, both used as detergent components.

Another very popular route for ethylene diversification is via the chloride route. Vinyl chloride monomer (VCM) is produced via thermal decomposition of ethylene dichloride. VCM acts as the monomer of polyvinyl chloride (PVC). PVC finds its usage in packaging, construction, medical equipment, and piping.

Vinyl acetate is an ester of vinyl alcohol and acetic acid, and is industrially produced by reaction of ethylene and acetic acid with oxygen in the presence of palladium catalyst (see Figure 4). Polyvinyl acetate (PVA) finds uses in latex paints, adhesive lacquers, and cements.

Ethylene can also be used as a monomer for cross-linking polymers like EPDM which is widely used for thermoset roofing, membranes, and in the automotive industry. These products boost the value chain by 10-300% when compared to the unit value of feedstock ethylene.

## Propylene derivatives

Figure 5 shows opportunities for diversification of propylene beyond polypropylene.

The following section discusses different propylene derivatives and touches upon chemical reactions leading to the end product. The reactions are indicative only.

Propylene oxide is a chemical intermediate of propylene which can be converted into valuable polyols which are used in polyurethane elastomers and plasticisers.

A polyether polyol is the polymeric reaction product of an organic oxide and an initiator compound

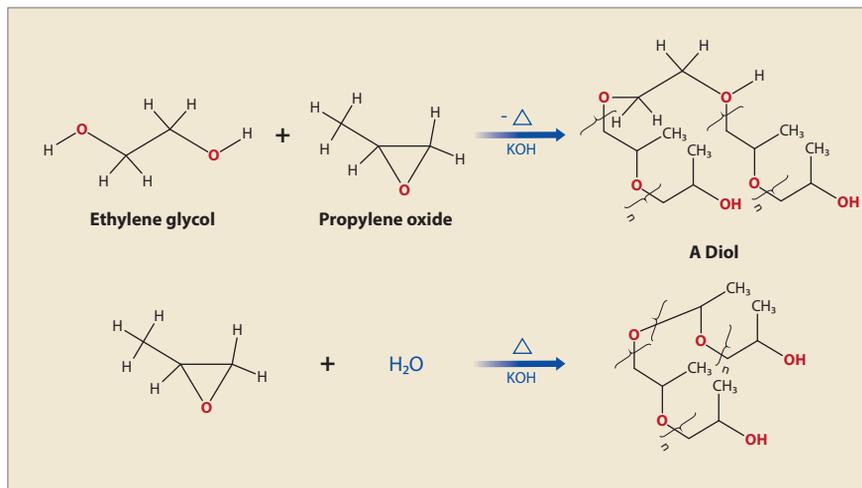


Figure 6 Production of a diol from propylene oxide

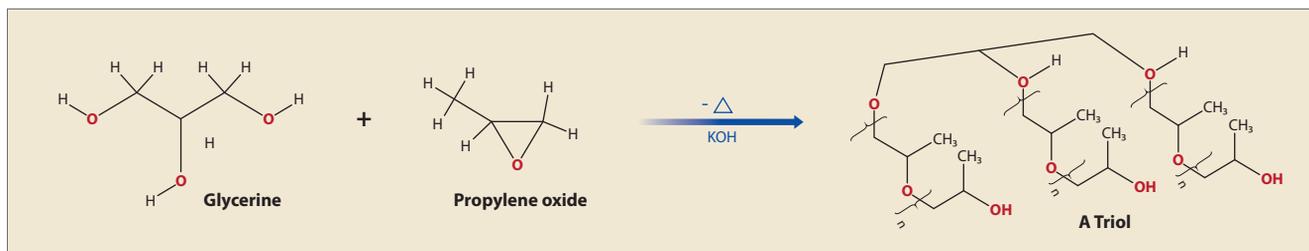


Figure 7 Production of a triol from propylene oxide

containing two or more active hydrogen atoms. The active hydrogen compound in the presence of a base catalyst initiates ring opening and oxide addition, which is continued until the desired molecular weight is obtained. If the initiator has two active hydrogens, a diol is produced (see Figure 6).

If a trifunctional initiator such as glycerin is used, oxide addition produces chain growth in three directions, and a triol is produced (see Figure 7).

Propylene can be converted to oxo-alcohols. The oxo process is used to produce aldehydes and alcohols from alpha-olefin feedstock using CO and H<sub>2</sub>. n-Butanol, isobutanol, and 2-ethylhexanol are among the best selling oxo alcohols which are used in plasticis-

ers, acrylates, air conditioning, and refrigeration.

n-Butanol is produced by the hydroformylation of propylene (oxo process) to form butyraldehyde. Typical catalysts are based on cobalt

and rhodium. Butyraldehyde is then hydrogenated to produce butanol (see Figure 8).

2-Ethylhexanol is produced industrially by the aldol condensation of n-butyraldehyde, followed

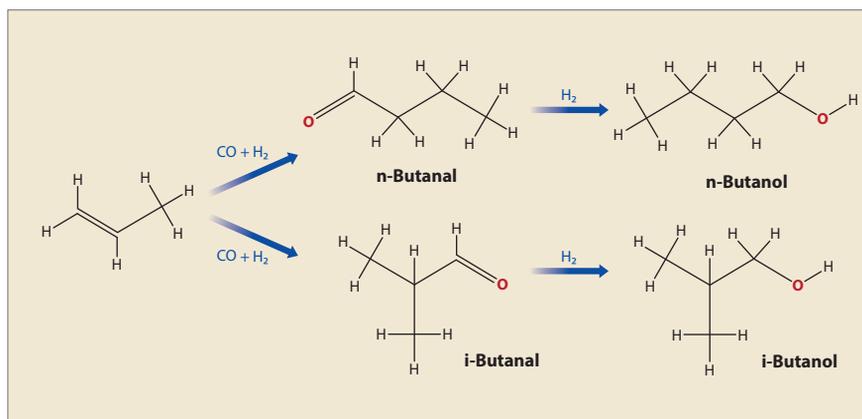


Figure 8 The propylene to butanol route

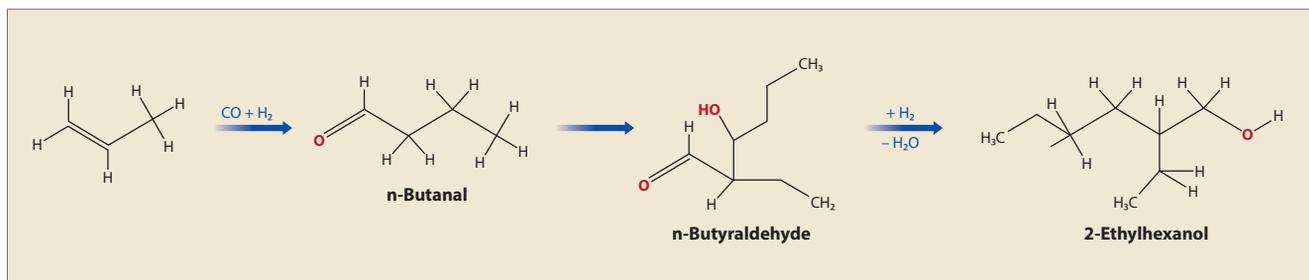


Figure 9 The propylene to ethylhexanol route

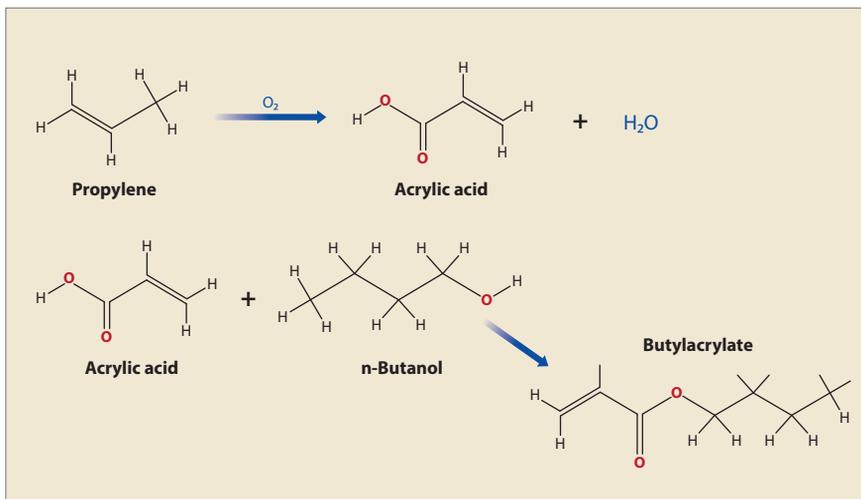


Figure 10 Propylene to acrylate route

by hydrogenation of the resulting hydroxyaldehyde. n-Butyraldehyde is made by hydroformylation of propylene. Butanol and butyraldehyde are used in coatings, varnishes, resins, gums, and vegetable oils.

Acrylic acids are produced traditionally by controlled oxidation of propylene. Esters of acrylic acid with oxo alcohols produce acrylates (see Figure 10). Acrylates are widely used in paints, surface coatings, adhesives, and textiles. Polyacrylic acids are used in super-absorbent polymer production.

Acrylonitrile is produced by catalytic ammoxidation of propylene.

Acrylonitrile is used in producing polyacrylonitrile which is used in sports and aerospace applications. All of these products represent a 15-200% gain in product value compared to the unit price of propylene.

#### C<sub>4</sub> diversification

Diversification based on C<sub>4</sub> molecules is not as widely explored as it is for C<sub>2</sub>/C<sub>3</sub> but can nonetheless offer a significant premium on operating margins with relatively low capital investment. Owners and operators in the Asia-Pacific region are opening up to new C<sub>4</sub> derivatives which are becoming increasingly crucial for the improved

profitability of a FCC or cracker based petrochemical or chemical complex (see Figure 11). Mixed C<sub>4</sub>s produced from a FCC or steam cracker contain varied percentages of 1,3-butadiene, isobutene, cis/trans 2-butene, 1-butene, isobutane, and n-butane. While FCC C<sub>4</sub> is rich in saturated C<sub>4</sub>s, steam cracker C<sub>4</sub> is rich in unsaturated C<sub>4</sub>s with comparable proportions of isobutene and 1-butene. Each C<sub>4</sub> component has a product value and can be further valorised by conversion to rubbers, polymers, and specialty chemicals.

The most common route for processing steam cracker C<sub>4</sub> uses a standard processing scheme involving butadiene extraction, 1-butene, and methyl t-butyl ether (MTBE). 1-3 Butadiene is used in synthetic rubbers whereas 1-butene is a comonomer of polyolefins. MTBE is still used as an octane booster in parts of the globe.

Other than the standard scheme, C<sub>4</sub>s can be diversified via various other processing routes by focusing on major chemical components present in the feedstock.

A C<sub>4</sub> mix from a steam cracker, when routed through a butadiene extraction unit to produce 1,3-butadiene, also produces raffinate, which contains n-butane, isobutane, isobutene, and 1- and 2-butene.

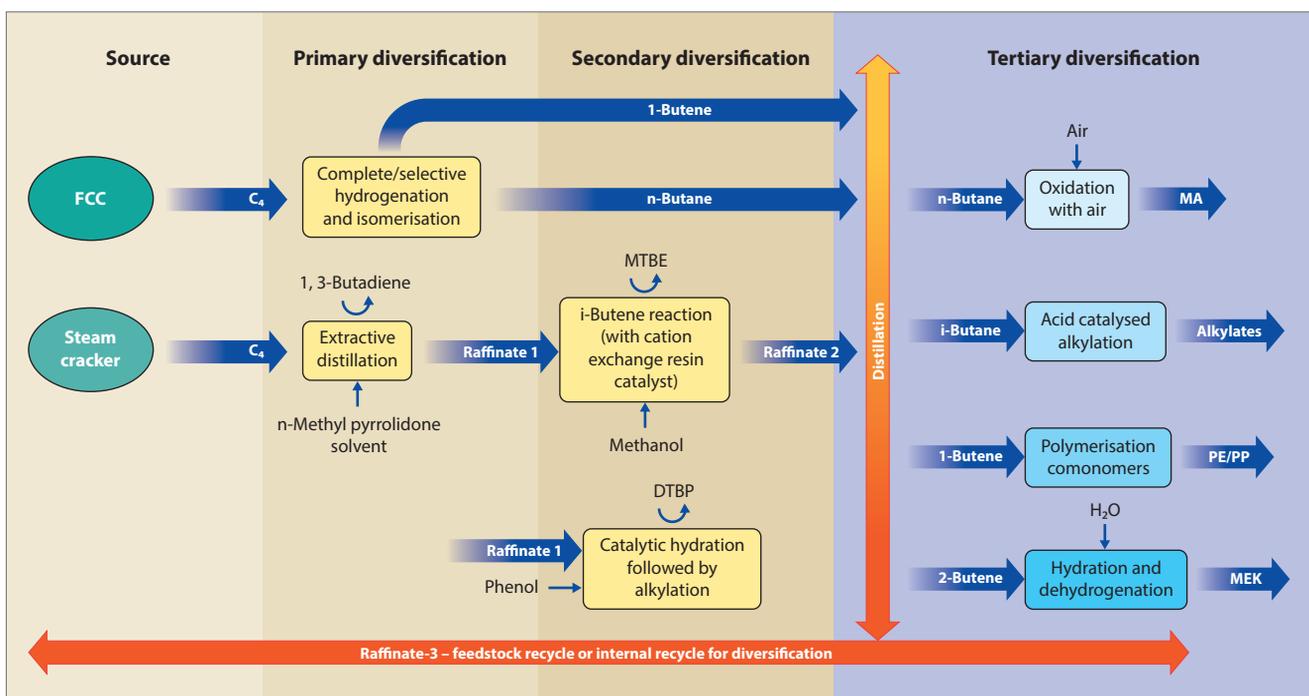


Figure 11 C<sub>4</sub> derivatives are becoming increasingly important for the improved profitability of a FCC or cracker based petrochemical or chemical complex

Raffinate is used as MTBE feedstock where the isobutene present in raffinate reacts with methanol to produce MTBE.

Raffinate can also be diversified to di-tertiary butyl phenol, a high value alkyl phenol derivative. The reaction involves catalytic hydration of isobutene to produce tertbutyl alcohol, followed by alkylation of phenol.

C<sub>4</sub> rejected from a MTBE plant is often called raffinate-2. This can be subjected to selective hydrogenation to maximise 1-butene which can be extracted as a product. Selective hydrogenation can also be combined with an isomerisation unit to control the ratio of 1-butene to 2-butene.

Alternatively, pure isobutene can be extracted from a C<sub>4</sub> mix and exported for production of butyl rubbers, polyisobutylene, methyl methacrylate, and polymethyl methacrylate. Technologies for extraction of isobutylene include MTBE cracking and cold acid extraction.

Raffinate-2 can also be diversified to produce methyl ethyl ketone (MEK). 2-Butene in raffinate-2 undergoes direct hydration with water in the presence of strongly acidic cation exchange resin catalyst to produce secondary butyl alcohol (SBA). Purified SBA undergoes dehydrogenation to produce MEK in the presence of a special catalyst.

MEK is an effective and common solvent and is used in processes involving gums, resins, cellulose acetate and nitrocellulose coatings, and in vinyl films. For this reason, it is used in the manufacture of plastics, textiles, in the production of paraffin wax, and in household products such as lacquer, varnishes, paint remover, as a denaturing agent for denatured alcohol, glues, and as a cleaning agent. 2-Butene can also be used in a metathesis unit with ethylene to form propylene.

After a secondary diversification block the raffinate is called raffinate-3. A n-C<sub>4</sub> rich raffinate-3 stream is conventionally recycled back to the cracker. Recycling the C<sub>4</sub> stream reduces the amount of naphtha required in the feedstock. In a refinery, raffinate-3 can be used in an alkylation unit.

Mixed feed butane (butane and i-butane) unclaimed so far from raffinate-3 can be distilled to gain the desired n-butane purity and can be further diversified. One such route is oxidation of n-butane with air to produce maleic anhydride. This is used primarily in the formation of unsaturated polyester resins for use in boats, cars, trucks, buildings, piping, and electrical goods, and its derivatives are used in pharmaceuticals, adhesives, engineered plastics, and coatings. All of these products represent 15-250% added product value when compared to the unit price of C<sub>4</sub> feedstock.

### What next for petrochemicals?

With the advantages gas based feedstock diminishing and the strong growth in demand in emerging markets stabilising, the likelihood of high intensity capital investment beyond 2030 might be in question. Where national oil companies are bound to make investments to cater to rising domestic demand, the petrochemical industry needs to find new ways to boost its bottom line. Deep chemical conversion is likely to be an effective strategy to increase the profitability of petrochemicals.

However, it is important to remember that valorisation of intermediates and downstream integrations are associated with significant capital investment. Niche chemicals, however profitable they are, do not contribute proportionally to the bottom line because they cannot match the economics of scale which act in favour of polyolefins. A balanced approach with polyolefins and specialty chemicals, with a flexibility to swing modes depending on market dynamics, will be the key to soaring profits in the future petrochemical industry.

**Sushree Chaudhuri** is a Process Engineer with Bechtel India. She has 18 years' experience in petroleum refining, petrochemicals, offshore oil and gas and flare systems, front end and detailed design. She holds bachelor's degrees in chemistry and chemical engineering from the University of Calcutta.

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## Coke drum repair restores capacity

The mechanical integrity of processing units is key for oil refineries to deliver smooth operations and maximum productivity. When structural damage was affecting the performance of a delayed coker unit in one of India's largest refineries, Sulzer's expertise proved fundamental to the repair of the equipment to full capacity.

Delayed coker units and their key components, coke drums, are exposed to harsh environmental conditions that take a considerable toll on these pieces of equipment. A typical coker cycle is characterised by high variations in temperature and mechanical load, as the coke drum is heated to separate coke from lighter oils and subsequently cooled to quench the solid masses of coke obtained.

Cyclic heating, cooling, and loading expose the delayed coker units to recurring, short-term increases in mechanical hoop and axial stress. As a result, common deformations in these vessels include radial bulges and corrugations, as well as cracks on the unit's shell, nozzle, and other areas.

When a major refinery and petrochemical complex in India noticed extensive bulging and cracking on one of its delayed coker units, it sought to restore its integrity. In this way, it could extend the equipment service life and minimise the risk of downtime while protecting the plant from costly equipment replacements and lengthy periods of reduced productivity. To repair the unit, the refinery turned to its preferred mass transfer specialist, Sulzer, which is also a provider of specialised field services for separation and mixing technologies.

### Weld overlay solution

The operators of the oil processing plant had already installed a number of Sulzer's components, such as column trays, as part of previous projects. Satisfied by the capabilities of these internals, the refinery asked for the support of Sulzer Tower Field Service (TFS) group.

Expert teams started by conducting extensive onsite inspections and non-destructive testing. These allowed them to get a comprehensive understanding and mapping of the damage in the unit. The damaged surface, which extended for approximately 160 m<sup>2</sup>, featured bulges, cracks, as well as traces of erosion and corrosion.

Based on Sulzer's findings, the two businesses agreed that weld overlay would be the fastest, most economical and long-lasting solution to extend the service life of the delayed coker unit. This repair technique consists of depositing a homogeneous layer of metal to fix losses of material on vessels. The weld overlay process would also be coupled with pre-welding activities, such as grit blasting and preheating, to maximise the effectiveness of the end result.

The in-situ investigations were also crucial to closely

simulate the physical entity and define a suitable course of action. In particular, by using finite element analysis (FEA), it was possible to get a computer-aided insight into how a particular weld overlay process would affect the thermal and mechanical behaviour of the structure when exposed to coker cycles.

These findings would then allow Sulzer's engineers to fine-tune the process parameters accordingly. As a result of these simulations, the engineering teams decided to utilise a temper-bead process in order to apply a 9 mm double layer of an upgraded metallurgy of Inconel 625 alloy. The material would provide the shell with high corrosion and oxidation resistance, even in the harsh operating conditions of the coker cycles.

### Automated operations

As it was necessary to prepare the damaged area by preheating it at 150°C before the actual welding process, it was impossible to conduct manual weld overlay operations.

Therefore, Sulzer's advanced automated technology, CladFuse, was selected to complete the repair. This uses a programmable logic controller (PLC) to define the movement of a carriage, which travels along a laser-levelled track system fixed to the wall that needs repairing. The PLC also controls a robotic index arm, located on the carriage, that moves the welding torch and the oscillator to create weld beads.

To cover the large worn area in a short amount of time, allowing the plant to reduce downtime and restore operations as soon as possible, Sulzer leveraged its global capabilities. The company was able to deploy teams from five different countries in order to repair various damaged areas simultaneously.

### Effective, fast repair

The repair of the delayed coker unit was completed in 25 days, during a planned shutdown of the plant. In this way, the refinery did not incur any unplanned downtime. As an additional benefit, the weld overlay has also reduced the potential impact of future thermal stresses and mechanical loads.

The delayed coker unit is now more robust and can run at peak performance again, processing 3.3 million t/y of crude oil.

Andrew Petticrew, Head of Global Sales for Automated Weld Overlay at Sulzer, concludes: "The feedback received from the customer was extremely positive. While delayed coker units and coke drums are regularly damaged by harsh operating conditions, advanced solutions, such as automated weld overlay, can extend service life while also improving their resistance to chemical, thermal and mechanical stress."

### Sulzer

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## FCC Ecat conversion to chemicals

Global refining catalyst is a multi-billion-dollar industry with an estimated 1-4% compound annual growth rate. In 2018 alone, 831 000 tons of refining catalyst was used. At the end of catalyst life, proper disposal of the waste catalyst is called for. Take FCC catalyst as an example. Some equilibrium catalyst (Ecat) with low metal content used for clean VGO feed can sometimes be traded to other units with much higher contaminant metals. For Ecat with high metal content, physical separation such as electromagnetic separation, and chemical rejuvenation was also practised since the 1980s. Physical separation does not recover the activity of the catalyst, while the activity and surface area recovered from chemical treatment are often short-lived. The final destinations of Ecat in the US have often been landfill or cement kilns. With the entrainment from groundwater and acid rain, the metals in the Ecat, particularly nickel, can eventually end up in the aquifer or watershed at a level impacting human health. Thus, Ecat is deemed a hazardous material in China. The conventional outlets like landfill and cement kiln are no longer options.

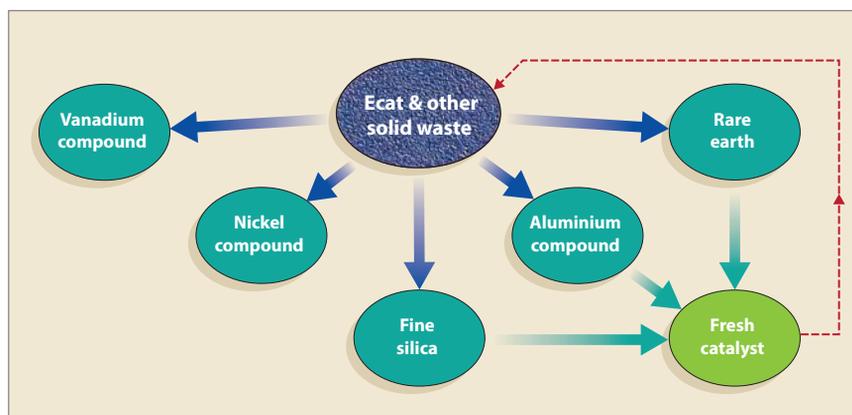
To resolve this issue, HCpect has developed a proprietary technology to break down Ecat and recover raw chemical materials. Some of the raw chemicals can be used in the synthesis and/or modification of zeolites, or the formulation of fresh catalysts. This circular design completes the life cycle of FCC catalyst and alleviates the environmental burden.

With a series of conversion and separation processes, the low or negative value Ecat is converted to commercial grade aluminium, silica, rare earth, nickel, and vanadium compounds (see **Figure 1**). Among them, the rare earth, nickel, and vanadium are of higher unit value. After the concept was proved in the lab, a 3000 t/y demonstration plant was commissioned in July 2017. This demo unit provided the design basis of the commercial plant. The first commercial plant, strategically located in Jiujiang, Jiangxi province along the Yangtze River, with a capacity of 10 000 t/y, was commissioned at the end of 2018. In 2019, the plant achieved the milestone of operating at capacity. Aluminium salt and other products are regularly produced and either sold to the market or used inside HCpect's business. So far, thousands of tons of aluminium salts have been sold to the commodity market.

HCpect's primary business has been production, marketing, sales, and technical service for fresh FCC catalyst and additives. Over the past year, HCpect manufactured and sold over 20 000 tons of FCC catalyst and additives in China and other parts of the world. Integration of the chemicals produced from the Ecat-to-chemicals plant back into its own value chain would be a benefit, but certainly not without challenges.

In zeolite synthesis, the aluminium salt can be used for synthesis of Y-zeolite. The rare earth can be used for treatment and stabilisation of Y-zeolite. Lab and pilot tests first proved the efficacy of those materials in the preparation of Y-zeolite for use in FCC catalyst or other applications. Commercial trials in the commercial zeolite synthesis facility were also successful. Zeolite activity and stability were maintained at the same level with the recovered aluminium salts in the synthesis process.

The functional silica material with high surface area can be sold to the market, or used as a component for the



**Figure 1** Scheme of completed catalyst life cycle, when chemicals from Ecat are reused/recycled in catalyst production

matrix of FCC catalyst. Lab tests proved that to a certain extent the matrix can be directly incorporated into the catalyst. If further adjusted, a higher percentage of the silica material can be incorporated. Following a pilot test, a commercial trial in the 20 000 t/y catalyst production facility has been successfully completed.

Incorporating waste-to-chemicals into the synthesis and production of FCC catalyst did not negatively impact the performance. Even better, extra benefits are gained when the high surface area silica is used in an already high-performing matrix in HCpect's catalyst. When the silica material is incorporated, measurable improvements were observed in coke selectivity and bottoms cracking, both critical for the profitability of today's FCC units.

In the foreseeable future, other used refining catalyst can also be used as feed to create useful chemicals. HCpect's plan is to roll out the 10 000 t/y module to multiple locations in China and then possibly elsewhere so that refinery solid waste can be minimised, recycled, and reused without compromised catalyst performance.

**HCpect**  
(Qingdao Huicheng Environmental Technology Co. Ltd.)

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## AI approach to equipment reliability

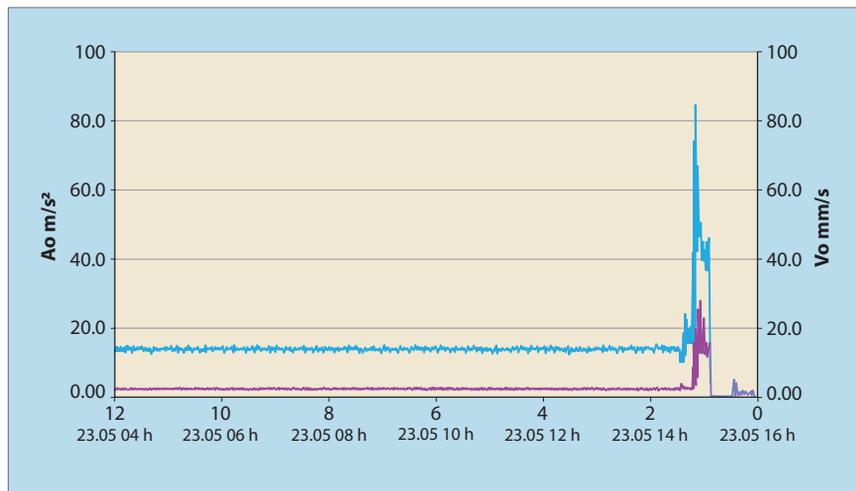
The primary costs and losses in petroleum refining happen due to sudden failures of equipment during nights and weekends when there is no expert on duty. For sustainable and reliable operation, it is necessary to provide

operators with timely, objective information on how different operating modes affect equipment health before a failure. Early diagnostics of defects allow people to eliminate future issues at an emerging stage. Using the real-time diagnostic Compacs system and assigning responsibility for equipment reliability to operators, petroleum refiners can both eliminate losses from accidents and downtime completely while reducing maintenance costs significantly. Because they are at the facility 24/7, operators should play a crucial role in equipment reliability. This is the only way to prevent not just a failure but even maintenance by promptly eliminating destructive forces as they emerge.

The Compacs system utilises physics based AI, developed to recognise the wide range of defects and malfunctions in various types of equipment. To do that, the system uses its own hardware to collect scientifically proven parameters from trustworthy sources which are physically related to machinery health. The system identifies over 95% of defects and malfunctions of rotating equipment using as few sensors as possible per machine. For example, an entire small cantilever pump or a motor with a capacity of less than 55 kW can be reliably monitored using just one piezoelectric accelerometer. The Compacs system measures around 80 different parameters of a vibration signal from every sensor including acceleration, velocity, and displacement; furthermore, the system measures current, temperature, pressure, and other parameters depending on machine type. The system has several interfaces such as Modbus and OPC UA, which are configured for data exchange with DCS, PI, SAP/3, and others. For security reasons, the software is not cloud based but could be integrated into a plant's diagnostic network.

With the Compacs system, an operator does not need to be an expert in vibration analysis to know how and when to react because the system identifies the wide range of defects and malfunctions of various types of equipment. The system warns personnel by texting prescriptions, colour-coding machines on a display in green/yellow/red colours, and speaking in a computer voice. The prescriptions are formulated as directives to personnel to carry out specific actions for improving reliability and extending a machine's lifespan. Thus, if operators remove destructive forces during an emergent stage of degradation, safe and reliable operation will become a reality.

An example of digital reliability in action is shown in **Figure 1**. Operators prevented a failure of the pump in a fluid catalytic cracking unit using prescriptions of the real-time diagnostic Compacs system. The figure shows 12-hour trends of the vibration acceleration and velocity measured on the front pump bearing. On 23 May 2017 at 15:02, the acceleration increased from 12 to 24 m/s<sup>2</sup> and the Compacs system revealed that the pump entered



**Figure 1** Twelve-hour trends of vibration acceleration (blue) and velocity (purple) of the pump P-323/1 of the FCC unit

an unacceptable condition and detected a front bearing defect, warning the operators by relevant prescriptions “Check bearing, lubrication” and “Check bearing cage”. Note that the velocity value remained low and in only 18 minutes, at 15:20, sharply rose 10 times, from 2.7 to 27 mm/s. Operators promptly took the corrective actions, switching to the standby pump and taking the damaged pump out of service, thereby preventing its breakdown. Analysis of the pump damage was done in the workshop where the pump was disassembled. The bearing suffered from overheating in the cage area.

Afterwards, a few more breakdowns of the pumps at the same unit, due to cage failures, were prevented because the operators were using the Compacs system and were deeply engaged in reliability. As a result, the refinery manager approved the decision to replace steel-cage bearings with bronze-cage bearings in all pumps, after which pump failures due to bearing problems significantly reduced. Today, the Compacs system monitors over 25 000 machines, more than 2400 types, at over 800 facilities, for almost 100 companies from 12 process industries.

#### USA Dynamics

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### Canned motor pump for diesel desulphurisation

Hermetic-Pumpen GmbH supplied a canned motor pump with a flow rate of 1000 m<sup>3</sup>/h and a delivery head of 65 m for a new diesel fuel desulphurisation plant in a Chinese refinery. The high system pressures and temperatures prevailing during the desulphurisation process require the highest safety and reliability of the equipment used. The hermetically sealed Hermetic pump, extensive measuring, and control equipment, as well as rigorous tests and inspections guarantee the plant operator a safe pump solution with high availability. The decisive factor for the refinery was Hermetic's expertise and experience with the complex challenges of large industrial oil and gas plants. In addition, a pump specialist in a

joint venture provides on-site support to manufacturing plants and service offices in China.

Under increasingly stringent global regulations on the reduction of pollutant emissions, China has a limit of 10 ppm for the sulphur content of diesel fuels. To comply with the requirements, refineries are required to invest in technological modernisation. The diesel desulphurisation process includes a hydrogenation process that requires system pressures of more than 100 bar and temperatures of more than 400°C. Since even the smallest leaks can cause fatal fires, plants and components require the highest level of reliability and safety. In the case of a reactor circulation pump requiring a delivery rate of 1000 m<sup>3</sup>/h and a delivery head of 65 m, a well-known refiner in China relies on the expertise and quality of Hermetic.

### Leakage-free pump solution

The core of the 10-ton pump system is a CNKfH+D 300-500 canned motor pump. The water-cooled motor of the pump according to API 685 has a shaft power of 185 kW. All pressure containing parts were made of forged solid materials. The housing is made of highly resistant SS 316 Ti stainless steel and has been tested at 310 bar. With a system pressure of 112 bar and an operating temperature of 405°C, the integrated, compact unit is designed for a nominal pressure of 130 bar. In addition to the rotor lining, the motor housing serves as a secondary containment, which provides a leak-free, safe pump solution.

### Low wear and maintenance design

The design-related omission of wear-prone components such as shaft seals, couplings, and bearings reduces the maintenance and repair costs and ensures a longer service life. The Zero Axial and Radial Thrust (ZART) technological part from Hermetic ensures contact- and wear-free running of the rotor unit. It is based on comprehensive axial thrust balancing in conjunction with hydrodynamic sleeve bearings. While relief mechanisms in the control unit continuously monitor the axial position of the pump shaft, a fine fluid film between the rotating and static sleeve bearing component allows full 'floating' of the rotor. To monitor axial thrust balancing, the pump is equipped with Hermetic MAP technology. The rotor position monitoring detects the slightest plays in the rotor movement and serves as a kind of early warning system to prevent undesired operating conditions and expensive damage. Due to this hydraulic power balance, the pump operates without radial and axial wear, thus offering plant operators significantly longer service life.

### Integrated measurement and control

To ensure high availability and durability, the canned motor pump is provided with additional measuring and control equipment. To prevent potential motor and bearing damage, this particular process avoids gas accumulations on the motor side, which can escape from the pumped medium. A dosing pump precisely tailored to the application generates a fine, uniform counterflow in the rotor space, which prevents this. A level indicator integrated in the system monitoring indicates whether there is no gas in the circuit. While a heat barrier prevents heat transfer from the pump to the motor part, an external heat exchanger ensures the dissipation of motor heat loss. To prevent overheating, temperature monitoring of the motor winding is integrated. Another sensor monitors the temperature of the motor cooling circuit, which ensures cooling performance and functional safety. Other components such as safety valves and extensive, rigorous testing and inspection procedures, including X-ray tests, vibration tests, and ultrasonic tests, as well as comprehensive documentation complete the scope of supply to a highly reliable system. Hermetic pumps demonstrate their safety and performance currently in 15 large refineries worldwide.

### HERMETIC-Pumpen GmbH

For more information:

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### Raising catalyst performance

Stringent constraints regarding automotive fuel specifications have pushed refiners to improve the efficiency of their gasoline and diesel production. Issues like octane number specifications, CO<sub>2</sub>, SO<sub>x</sub>, NO<sub>x</sub> and light particle emissions can be listed as the main points that require technological creativity and efforts.



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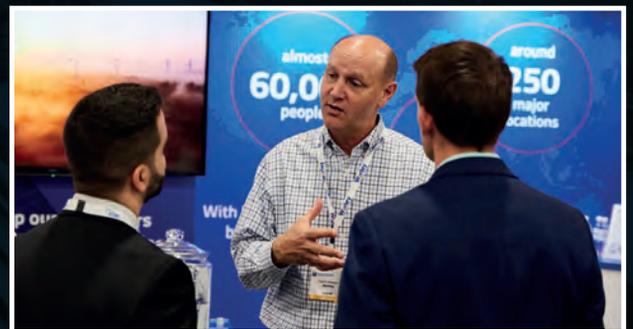
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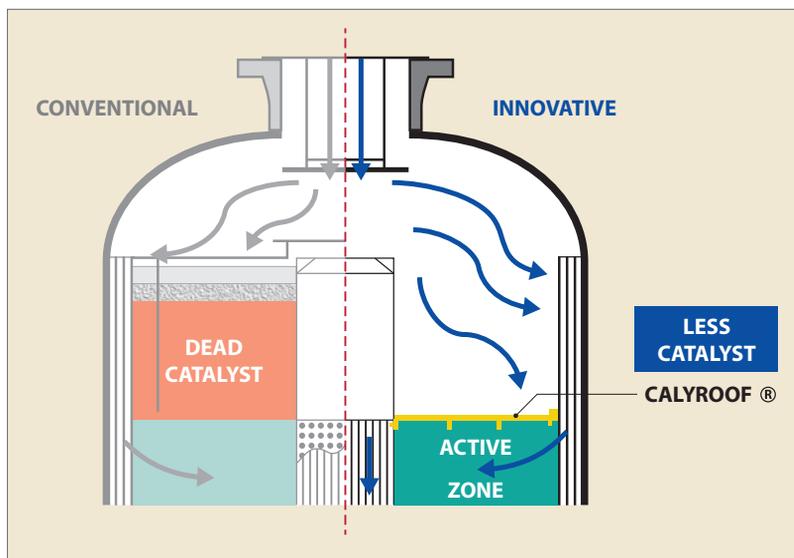


Figure 1 Less catalyst, same activity

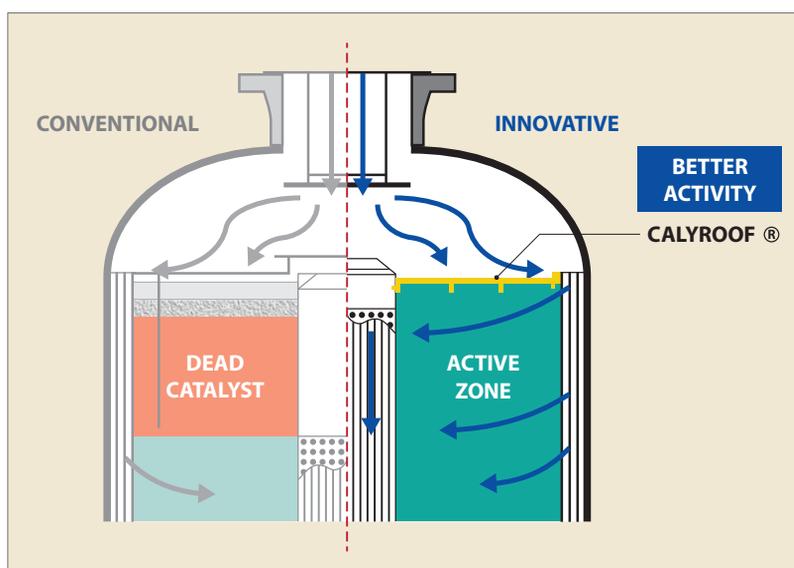


Figure 2 More active catalyst means higher performances

Such sacrifice should be accomplished at the same safety level for personnel and equipment while keeping financial balance as the upper standard.

One progressive way among others is improving the heart of the refinery production, the catalytic processes which deliver effluents to the highest possible commercial specifications.

Crealyst-Oil has the HDL Process and Calydens dense loading machine with the target of gaining catalyst weight within a given catalytic reactor, and enhancing bed homogeneity and yields.

In addition, Calyroof is used as a protective cover on the upper surface of the catalyst fixed bed of radial flow reactors, as in reformers.

The main advantage of Calyroof is the tightness obtained around the central pipe and its shroud, and close to scallops, avoiding any by-pass of the catalyst bed. In such conditions, the presence of a sacrificial zone composed of ceramic balls and catalyst (usually loaded at the top part of a reactor) is dispensable, even undesired. Less catalyst can be used, maintaining process

yields with the same quantity of catalyst (see Figure 1).

Moreover, the arrangement's weight and flexibility stabilise the surface against turbulence, in the case of compressor failure for instance.

Such a device does not affect human health during handling and installation, since it is free of carcinogenic elements. This flexible 'roof' is easily adapted to different sizes of reactor; it is composed of articulated tiles made of heat resistant stainless steel and laid on the catalyst bed. The system's material of construction, a refractory metal type SS 321, is reusable for several cycles of operation. Since the expected lifetime is 10 years, the maintenance programme consists of cleaning operations every four years, a gain in process operation both technical and financial.

Depending on reactor size, one can estimate an increase of around 15% of active catalyst quantity, leading to extra performance with an increase in overall stability. Such an improvement in processing capability means for refiners a higher return on investment due to increased productivity and lower maintenance by less frequent operations in the reactor top zone as well as limited dust deposition on the surface.

In the case of a unit revamp, a Calyroof cover allows for design modifications to the central pipe and scallops which can be extended almost to the lower part of the inlet distributor for a gain in stability (see Figure 2).

Increased feed rate can make an important difference, made possible because of the extra catalytic capacity.

The same advantage could be extended to increasing operation severity for higher reformat octane number at unchanged feed rate, assuming that existing process equipment can be adapted, in particular at the furnace stage.

Since gas distribution through the catalyst bed is improved (even more when combined with Homogeneous Dense Loading), there is no risk of hot spots and in consequence coke formation is reduced.

Calyroof follows the exact shape of the catalyst top bed surface during process operations and emergency shut-down, through occasional lift-up and set-down of the system.

Used in catalytic radial flow reactors, Calyroof is an efficient way to optimise overall performance, and raise process stability and operational security in existing as well as new units. The result is a gain in the yield and productivity of the final product and of internal hydrogen production used for downstream units.

**CREALYST-OIL**

For more information: [contact@crealyst.fr](mailto:contact@crealyst.fr)

## Enclosure heating leads to process heater failures

It has generally been assumed in energy process heaters that pins attaching heating elements to the power terminal do not generate heat on their own. That is why they are called cold pins. Engineers at Watlow believe cold pins actually generate heat creating an environment that is leading to heater failures.

When circulation heaters used in energy processing applications break, the failures predominantly involve the electrical termination in the device and not the heating element. They most often happen in the enclosure in the stand-off area between the flange and base plate. Circulation heaters for applications such as hydrocarbon refining, power generation, and liquefied natural gas processing usually consist of many tubular heating elements held in place by a flange creating a chamber where the liquid or gas flows around them. The heating happens through conduction.

On the other side of the flange away from where the heating is intended is the wiring of the heater connecting the power source to the heating elements. Engineers have traditionally calculated some heat leaking from the chamber to this enclosure. Models have predicted heating in this area only close to the flange, moderately increasing the temperature in the entire enclosure. The models did not take into account heat generated by the connecting pins.

For decades, engineers across the industry have relied on modelling to predict the terminal enclosure temperature for these kinds of heaters. Unfortunately, the modelling was too simple and flawed in that it did not take into account all of the heat generation points in the terminal. While the cold pins are not made of copper wire which would be found in a traditional heating element, if enough current is passed through the metal pin it in fact becomes a heater. It gets

warmer than most engineers anticipated or expected based on the old modelling.

"Our new predictive modeling shows much more of a curve where the temperature peaks in the centre of the enclosure where the cold pins are located," said Scott Boehmer, a principal engineer with Watlow. "We always assumed the heat in the enclosure was near the flange, but the temperature in the space was significantly higher away from that area because of the resistance with the pins."

As process heaters designs have gone to higher amperage levels, they have become physically larger and have more wattage. It has become more important to understand total heat generated in the 'non-heater' part of the assembly. If designers do not correctly predict the heat generation from all the electric components, the end user is going to have problems regardless of the fluid running through the heater.

"The terminal enclosure area is generally the weakest part of any process heater assembly, and that is true across the industry. But the heater is only as good as the entire system and it will fail because of a loose connection," Boehmer said. "That's where the problems occur. It may not be immediately but always before the heating elements fail."

Watlow engineers are looking at ways to improve enclosure designs to reduce temperatures. Another way to improve the lifespan of the termination area is to ensure that the proper wiring is used to install the process heaters. Using wiring not rated for the power load of the heater can also be a contributing factor.

"Better understanding of the temperature in the enclosure is leading us to multiple ways to deal with the issue and improve reliability for process heaters," Boehmer said.

### Watlow

**For more information:**  
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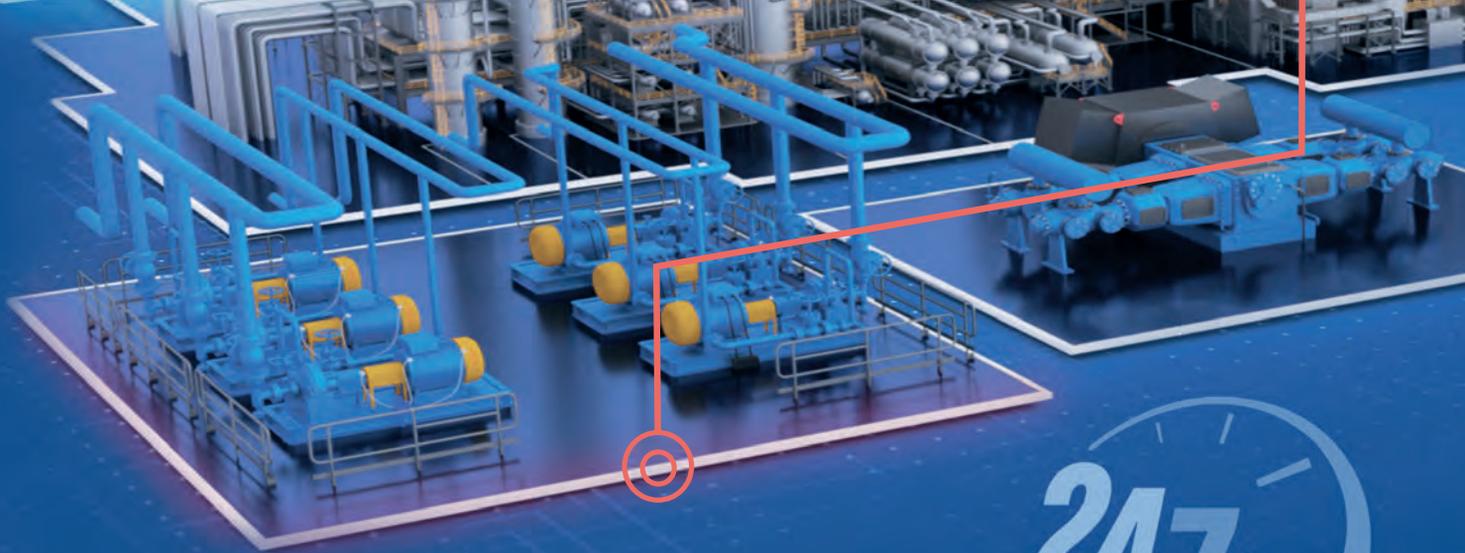
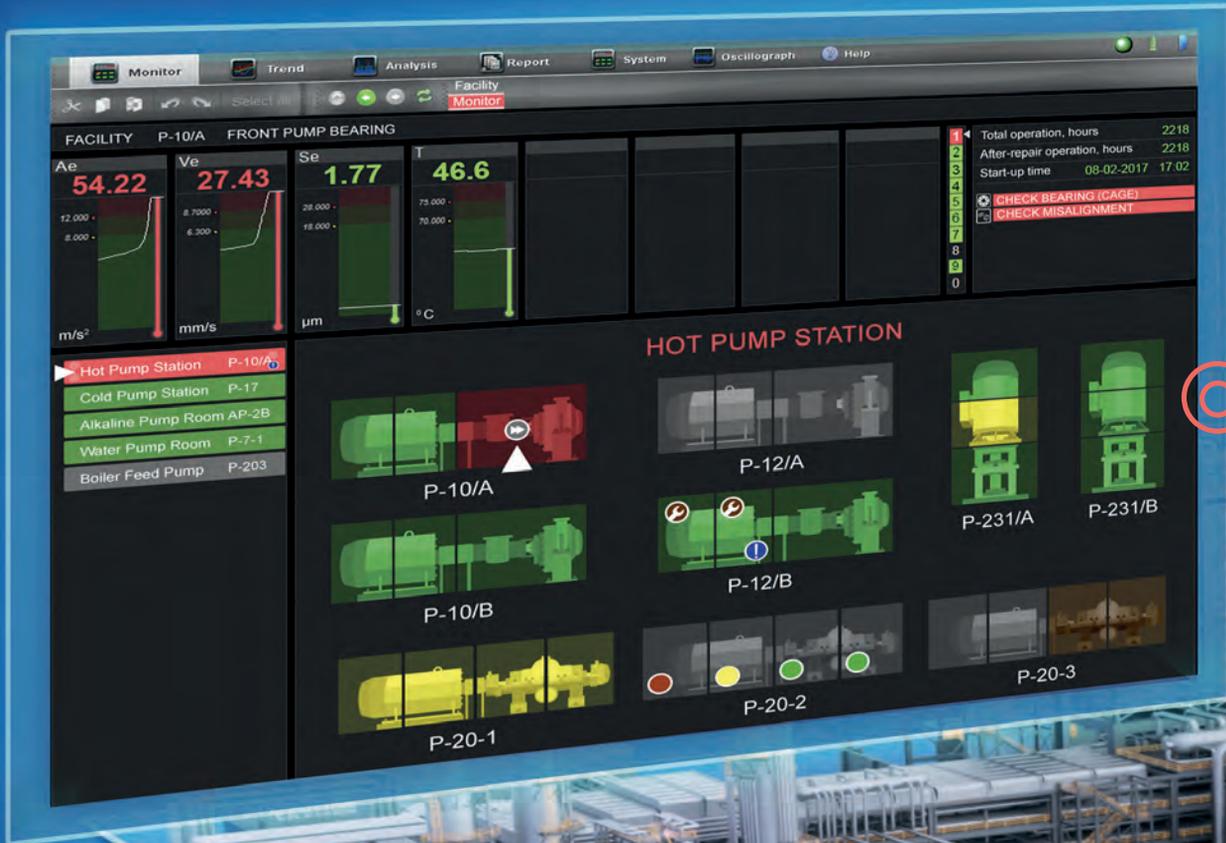
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