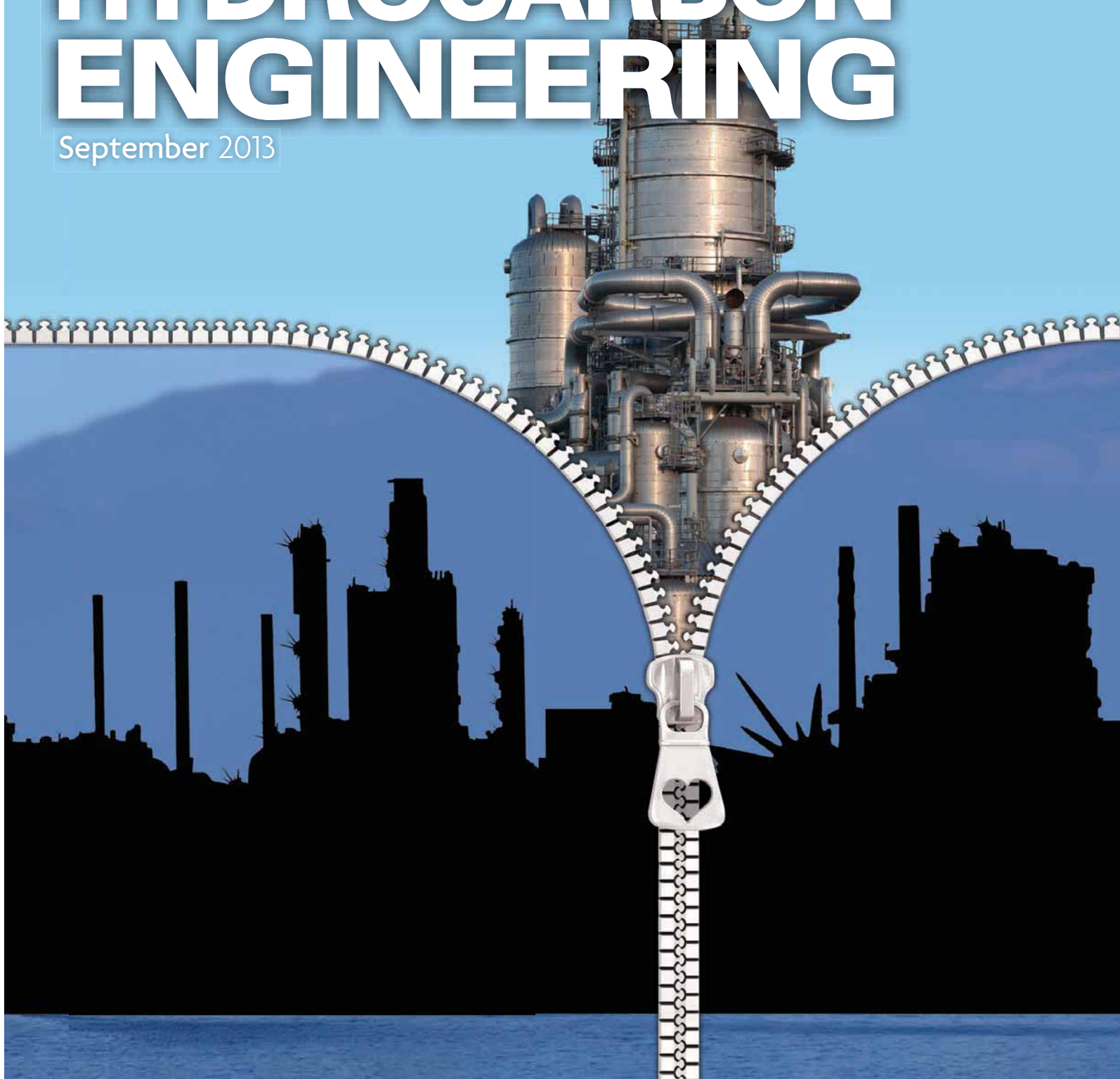


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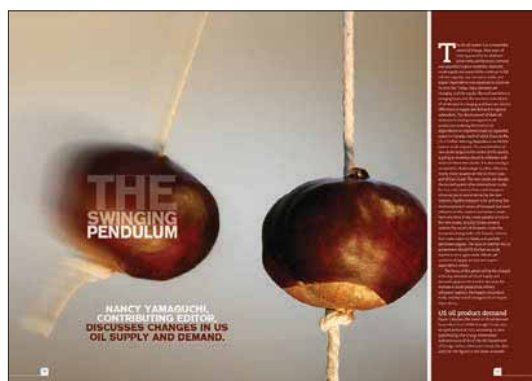
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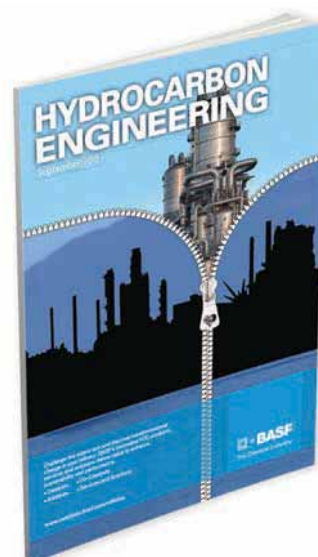
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comment

Claire Lloyd Editor

A new phrase has entered my oil and gas lexicon in recent weeks, 'the resource curse.' I have been aware of the concept that once the boom of a new finite resource has come to an end, economies suffer, however, the exact terminology is new to me. It was first brought to my attention when I read 'Oil in Uganda: International Lessons For Success' from Chatham House and was further highlighted by an article in the *Financial Times*, 'Entrepreneurs key to Papua New Guinea avoiding a resource curse'. What both these pieces emphasised is that the resource curse is essentially always a danger and there is no fixed set of rules that will help an economy avoid it.

First, let us look at Papua New Guinea (PNG). The island benefitted hugely in the early 2000s with the discovery of vast gold, copper and oil reserves. However, the near completion of the ExxonMobil LNG project is being cited as one of the main reasons why the country is seeing the imminent arrival of the resource curse. The *FT* has stated that the influx of expats to help construct the project has increased annual GDP to 8% over the last five years, and is likely to drop to 4% in the near future. The arrival of workers has also 'driven property and food prices to painfully high levels'. As these two fundamental things have reached steep prices and the expat exit is imminent, this doesn't bode well for the original inhabitants. Also, once construction has finished the number of jobs available will drop from approximately 20 000 to only 600. In a country that has a high level of violent crime and is divided into 'more than 800 tribes and languages' these dramatic price increases and the slump in employment opportunities is leaving the nation open to corruption, an increase in economic fluctuations and distortions.

With regards to saving PNG from crime, corruption and economic collapse, the *FT* article hails entrepreneurs as the nation's saviours. Mahesh Patel is crowned as such due to his development of a pharmacy chain on the island and the opening of the island's first commercial cinema. His entrepreneurial spirit and endeavours are hopefully going to aid the country by providing employment and help prices stabilise. However, when it comes to Uganda, there appears to be completely different ideas on how to help the country avoid the curse.

The Chatham House report states that 'an oft repeated mantra is that good governance is vital in avoiding the resource curse,' and from reading the rest of the piece, it appears that this is the approach Uganda is looking to take. The country is relatively new to the oil industry and there is a need for much development and investment before peak production can be achieved, so it is imperative that prevention measures are put in place now. The report comments that Uganda as a whole is attempting to avoid oil undermining governance, and is seeking a universally acceptable way for oil revenues to be allocated fairly. It is the inability to reach a fair and acceptable way that is portrayed as being Uganda's biggest challenge to overcome and therefore avoid the resource curse. As the report says, 'the biggest threat that oil poses to Uganda's harmony would come from allowing rumour and speculation to dominate, notably over how revenues are allocated.' How this will be achieved and justified is a different story but is one that many hope will be concluded successfully.

So, it seems, with reference to the above countries, that while the effect of the resource curse is pretty much universal, preventing it and then overcoming it does not have a one size fits all solution.

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world news



Italy | STRATEGIC PARTNERSHIP SIGNED

Neville Venture, a joint venture between GTC Technology and Neville Chemical Co., has signed a strategic partnership agreement with Versalis (Eni) for the production of hydrocarbon resins at Versalis' plant in Priolo, Italy. GTC also entered into a license agreement for producing the C5 feedstock which will supply the resin facilities. These new, high value products, synergetic in particular with

Versalis' elastomer business, are used in speciality applications such as adhesives, inks, coatints and rubber.

The project represents an important investment in Versalis' program for the overhaul of the Priolo plant site, and contributes to upgrading the products from the existing cracking facility. The products will be sold in Europe where Versalis has a widespread commercial presence.

Indonesia | WASTEWATER TREATMENT

Siemens Energy has received a contract from Indonesia to deliver a Zimpro wet air oxidation system. The technology will be part of the new Gundih gas processing plant where it will treat wastewater contaminated during the gas production process. The contract was awarded by a consortium consisting of PT Inti Karya Persada Teknik and PT Adhi Karya (Persero) Tbk. As general contractors, both companies are working together to build the gas processing plant for the end customer, the state owned Indonesian oil and gas PT Pertamina EP.

The Gundih gas processing plant is under construction in Blora in the country's Central Java province. The facility will have a daily gas production capacity of 50 million ft³ after start up at the end of this year. Siemens' Zimpro wet air oxidation system will treat a mixture of alkaline wastewater and wastewater generated during sulfur recovery by destroying odorous and high chemical oxygen demand pollutants and generating an effluent that meets the Indonesian environmental regulations for discharge.

Brazil | RECYCLING IN THE REFINERY

Veolia Water Solutions & Technologies Brazil and Veolia Water Americas have signed a contract with Petrobras to design, build, own and operate an oil bearing secondary material processing unit at its Alberto Pasqualini REFAP refinery located in Porto Alegre, South Brazil. It is the sixth largest of Petrobras' refineries with a capacity of 200 000 bpd. The new unit will use Veolia's proprietary processes to recycle the oil bearing material back into the refining production processes, which is anticipated to result in

approximately US\$ 1.5 million/y in cost savings.

The recycled oil bearing materials from the refinery's crude processing streams will be used in the refinery's coker unit. Veolia's proprietary processes will help Petrobras increase its refining utilisation, while at the same time minimising waste generation and thereby helping the environment.

The recycled oil bearing materials from the refinery's crude processing streams will be used in the refinery's coker unit.

UK | HI-FORCE OPENS NEW LOGISTICS CENTRE

Hi-Force, the UK based manufacturer of high pressure hydraulic tools, has taken a delivery of a new multi million pound purpose built facility, at the rear of its existing UK manufacturing plant. The new building is virtually self sufficient in regards to electricity, due to integral solar panelling within the roof area.

The new facility houses the entire UK logistical operation, from raw material through to finished product, plus a 1500 ft² Engineering Construction Industry Training Board (ECITB) approved Training School.

Alongside the relocation of the Logistics Centre, Business Development Manager, Pat Wright, and Training Manager, Steve Wakeline, have been equipping the new Training School to provide Mechanical Joint Integrity Training. Plans are already in place to expand the Hi-Force Training Centre, in order to enable the provision of training courses in other product and application areas.

Construction and equipping of the Training School has so far cost in excess of £ 200 000. Pat Wright has commented that the Hi-Force training facility is among the best in the industry.

Commenting more generally on the expansion, Group Managing Director, Kevin Brown, has indicated that the new Logistics Centre was primarily constructed in order that an increasing volume and range of products and components could be better handled. However, expansion has additionally freed up 7000 ft² in the main building. This space can be utilised to house additional manufacturing capability.



world news



INBRIEF

CHINA

Endress+Hauser is manufacturing more and more flowmeters in Suzhou, China, with the principal growth drivers being water and wastewater industry. Endress+Hauser Flowtec has reacted by enlarging its plant in the east of the People's Republic. The extension work was recently completed after getting underway in March 2012.

USA

Renewable Energy Group® (REG) has completed the purchase of a 30 million gal./y biodiesel plant in Mason City, Iowa, formerly owned by Soy Energy, LLC. REG acquired the biorefinery for US\$ 11 million in cash and the issuance of a US\$ 5.6 million promissory note.

CHINA

Technip was awarded by Shaanxi LNG Investment & Development Co Ltd an engineering, design and procurement contract, worth approximately € 35 million, for a mid scale LNG plant. The plant will be located in the Yangling Demonstration Area, Shaanxi Province, China.

USA

CB&I has announced the successful completion of the expansion and modernisation of its technology development and manufacturing centre in Pasadena, Texas. The new facilities will allow CB&I to more efficiently develop and advance innovative process technologies, and increase production capacity of proprietary structured packings manufactured at the site. The packings are essential components used in CB&I's catalytic distillation process technologies.

Germany | COMPRESSION CONTRACT

One of the four industrial gas production leaders worldwide and Burckhardt Compression signed a field service and valves maintenance framework supply agreement for the European territory. Under the agreement, Burckhardt Compression will provide on site field service and valve service on all types of

reciprocating compressors for more than 40 industrial plants in Europe.

This agreement builds upon the already excellent cooperation between the two companies. The industrial gas production company is a longstanding partner of Burckhardt Compression in the areas of new compressor systems and compressor components and services.

Foster Wheeler | THREE ANNOUNCEMENTS

Foster Wheeler AG has announced that a subsidiary of its Global Engineering and Construction Group has signed an equity purchase agreement with The Fourth Construction Company Ltd of China Electronics System Engineering (CEFOC). Under the agreement Foster Wheeler will purchase a 49% share in CEFOC's Chinese Design Institute, PECHDI.

A subsidiary of Foster Wheeler AG's Global Engineering and Construction (E&C) Group has also been awarded a contract by The Dow Chemical Company. The subsidiary will be providing detailed engineering, procurement and construction management (EPCm) services for the LA-3 Crak More Ethane (CME) project at Dow's Plaquemine petrochemical facility in Louisiana, USA.

The objective of this project is to improve the plant ethane flexibility to

take advantage of low cost feedstock. The scope will include brownfield additions and retrofit modifications to the plant. This project is part of Dow's comprehensive US Gulf Coast investment strategy.

Also, a subsidiary of the same Foster Wheeler Group has been awarded a project management and consultancy (PMC) services contract by the Nghi Son Refinery and Petrochemical Limited Liability Company for its Nghi Son Refinery and Petrochemical (NSRP) complex to be constructed in the Nghi Son Economic Zone, Thanh Hoa Province, Socialist Republic of Vietnam.

The NSRP Complex will consist of an integrated refinery with a processing capacity of 200 000 bpd and a petrochemical complex, with associated infrastructure, utilities and offsite facilities. The contract value was not disclosed.

USA | DOUBLE ANNOUNCEMENT

KBR has been awarded a front end engineering and design (FEED) contract by INEOS Olefins & Polymers USA (INEOS) for one of the largest grassroots high density polyethylene (HDPE) plants to be built in the Americas. The facility would produce 470 kilo tpy of bimodal HDPE using Innovene™ S process technology.

KBR has also secured a front end engineering and design (FEED) contract for a crude oil refinery revamp project for one of the oldest operating refineries in the US. During the feasibility phase, KBR's experience will help determine the optimum economic figuration to increase heavy crude processing capacity.

A large, multi-story industrial refinery structure is illuminated at night. The structure consists of numerous levels with platforms, ladders, and large cylindrical tanks. Bright lights are visible on each level, creating a strong contrast with the dark blue night sky. The overall scene conveys a sense of industrial activity and complexity.

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INBRIEF

USA

CB&I has announced that it has been awarded a contract valued in excess of US\$ 270 million by Sunoco Partners Marketing & Terminals, LP for the construction of a turnkey propane terminal and deethaniser facility in Marcus Hook, Pennsylvania. This is a third quarter 2013 award.

JAPAN

hte has announced that the Japan Petroleum Energy Center (JPEC) is now using the company's parallel reactor technology in order to develop new refining catalysts and kinetic models. hte has supplied an X4500 high throughput catalyst testing system for hydroprocessing applications to the JPEC Advanced Technology and Research Institute (ATRI) in Chiba, Japan.

RUSSIA

Siemens will deliver the key components for a power station that is going to supply the future Yamal LNG production plant in Northern Russia with electricity and heat. The Siemens scope of supply comprises the design, manufacture, factory testing, delivery, installation and commissioning of eight SGT-800 industrial gas turbines as well as nine step up transformers.

CHINA

Air Products has signed an agreement to supply its proprietary LNG process technology and equipment to Technip for a mid scale LNG project in the Yangling Demonstration Area, ShaanXi Province. Air Products' technology will be vital for a liquefaction train producing 500 000 tpy of LNG.

UK | UNDERGROUND GAS STORAGE

Yokogawa Electric Corporation has announced that its subsidiary, Yokogawa United Kingdom Ltd., has been awarded a multi million pound contract by SSE Hornsea Ltd for the upgrade of one of the UK's largest onshore underground gas storage facilities, located in Atwick, East Yorkshire. Yokogawa will supply an integrated control and safety system, together with a metering solution.

The order includes the supply and installation of Yokogawa's flagship

CENTRUM® VP integrated production control system to control the gas compressors. Also included in the project is the ProSafe®-RS integrated safety instrumented system, which will contribute to ensuring process safety at the facility.

The Atwick gas storage facility comprises nine manmade salt cavities that have been leached into a salt layer 1.8 km underground, creating approximately 328 million m³ of gas storage space.

The Netherlands | LICENSING AGREEMENT

Paqell B.V. and Frames Gas Processing B.V. have entered into a licensing agreement for THIOPAQ O&G desulfurisation technology. This agreement adds THIOPAQ O&G technology to Frames' gas treating portfolio, enabling Frames to offer sulfur recovery technology to their customers, thereby providing complete sour gas solutions.

THIOPAQ O&G technology offers an environmentally friendly way to remove H₂S from high or low pressure natural gas streams or associated gas streams and convert it into elemental sulfur. This cost effective system uses low energy and requires minimal effort to operate.

Canada | ENGINEERING AND PROCUREMENT

Jacobs Engineering Group Inc. has received a three year contract from Nexen Energy ULC for engineering and procurement services for the Long Lake oilsands facility. Long Lake is an integrated steam assisted gravity drainage (SAGD) and upgrading operation

that uses a proprietary OrCrude™ technology as well as hydrocracking and gasification to produce Premium Synthetic Crude (PSC™) oil.

The Long Lake site is located approximately 40 km northeast of Fort McMurray, Alberta, Canada.

Canada | RAIL LOADING TERMINAL

Keyera Corp. and Kinder Morgan Energy Partners LP, have announced a 50-50 joint venture to build a crude oil rail loading facility in Edmonton, Alberta. The facility will be called the Alberta Crude Terminal. When complete the terminal will be able to accept crude oil streams handled at Kinder Morgan's Edmonton Terminal for

loading and delivery via rail to refineries anywhere in North America.

The Alberta terminal will be constructed next to Keyera's Alberta Diluent Terminal on land recently acquired by a Keyera subsidiary. The facility will have 20 loading spots capable of loading approximately 40 000 bpd.

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EIA | BIODIESEL PRODUCTION RECORD

According to figures from the US Energy Information Administration (EIA), US biodiesel production in May 2013 reached a record level of 11 million gal. Production came from 116 active biodiesel plants with a total operable capacity of 2.2 billion gal./y.

During the first five months of 2013 production was 449 million gal., an increase of 17 million gal. from the same period in 2012. During the same period this year, 3427 million lbs of feedstocks were consumed for biodiesel production. This gives a ratio of 7.6 lbs of feedstock/gal. of biofuels produced.

The use of biofuel as a motor fuel is supported by federal mandates for

biodiesel blending under the Renewable Fuel Standard program. Under this program, the US Environmental Protection Agency sets annual targets for the use of biodiesel and other biofuel categories, such as advanced and total biofuels, which can also be satisfied through additional biodiesel use. Biodiesel blenders are also currently eligible for a federal tax credit of US\$ 1/gal. blended.

Most biodiesel fuel is blended with diesel fuel derived from petroleum and is used in the transportation sector. Biodiesel accounted for less than 2% of US distillate fuel oil product supplied of 3.9 million bpd during the first five months of 2013.

AFPM | CONSUMER FAVOUR VOTING

American Fuel & Petrochemical Manufacturers (AFPM) President Charles T. Drevna issued the following statement in response to the US House vote to approve the Energy Consumers Relief Act of 2013 (HR 1582):

'Proposed regulations on US refineries have grown at an alarming pace without due consideration of the adverse impact on consumers and the economy. This legislation injects important oversight to a rulemaking process that has been notable for its lack of transparency and system of

checks and balances. Today's vote by the House recognises the need for Congress, the Department of Energy, the Environmental Protection Agency and others to work in tandem to avoid the unintended consequences of an ill conceived energy agenda. AFPM commends the House for its actions in passing the Energy Consumers Relief Act and encourages the Senate to make the same commitment that will lead to more rigorous review of the most costly regulations that ultimately cause consumers great harm.'

IHS | HAZARDOUS AREA EQUIPMENT

According to a report from IMS Research, part of IHS Inc., three quarters of revenue in the Europe Middle East Africa (EMEA) region from sales of products suitable for use in hazardous areas were from the oil and gas industry in 2012.

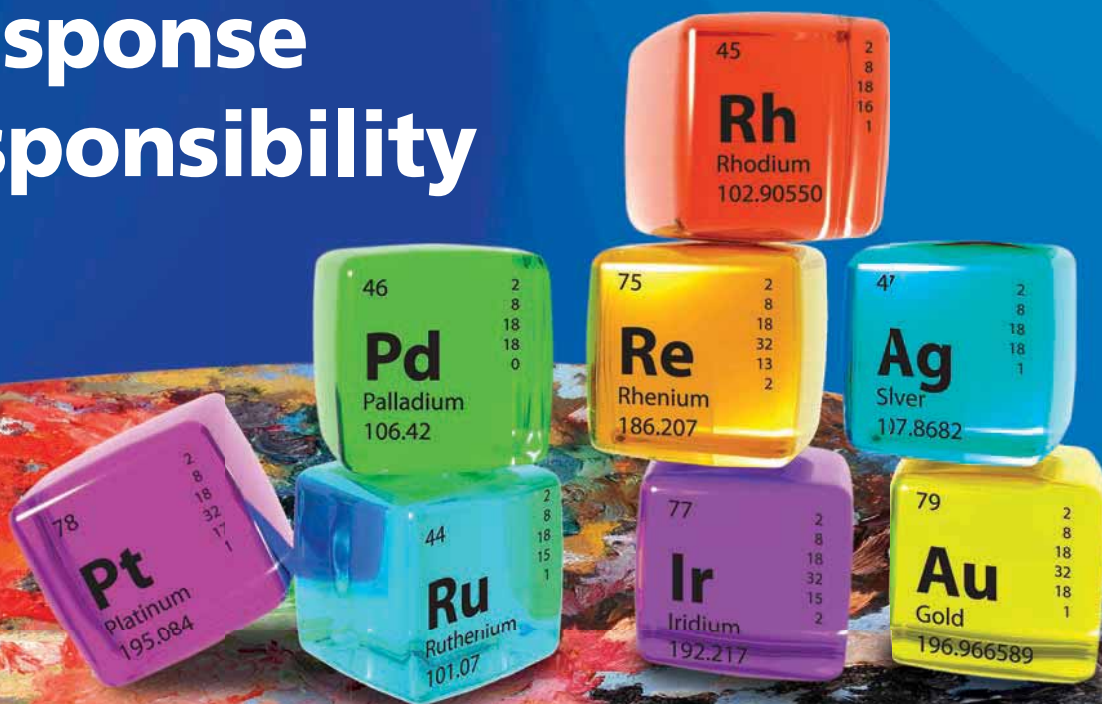
'Despite the well known fact that oil and gas is the most important

industry for hazardous area equipment, the very high market share it held last year came as a surprise', said John Morse, senior automation market analyst with IHS.

In particular, the results cited the growing volume of legislation being applied to oil and gas installations as one of the main increase factors.

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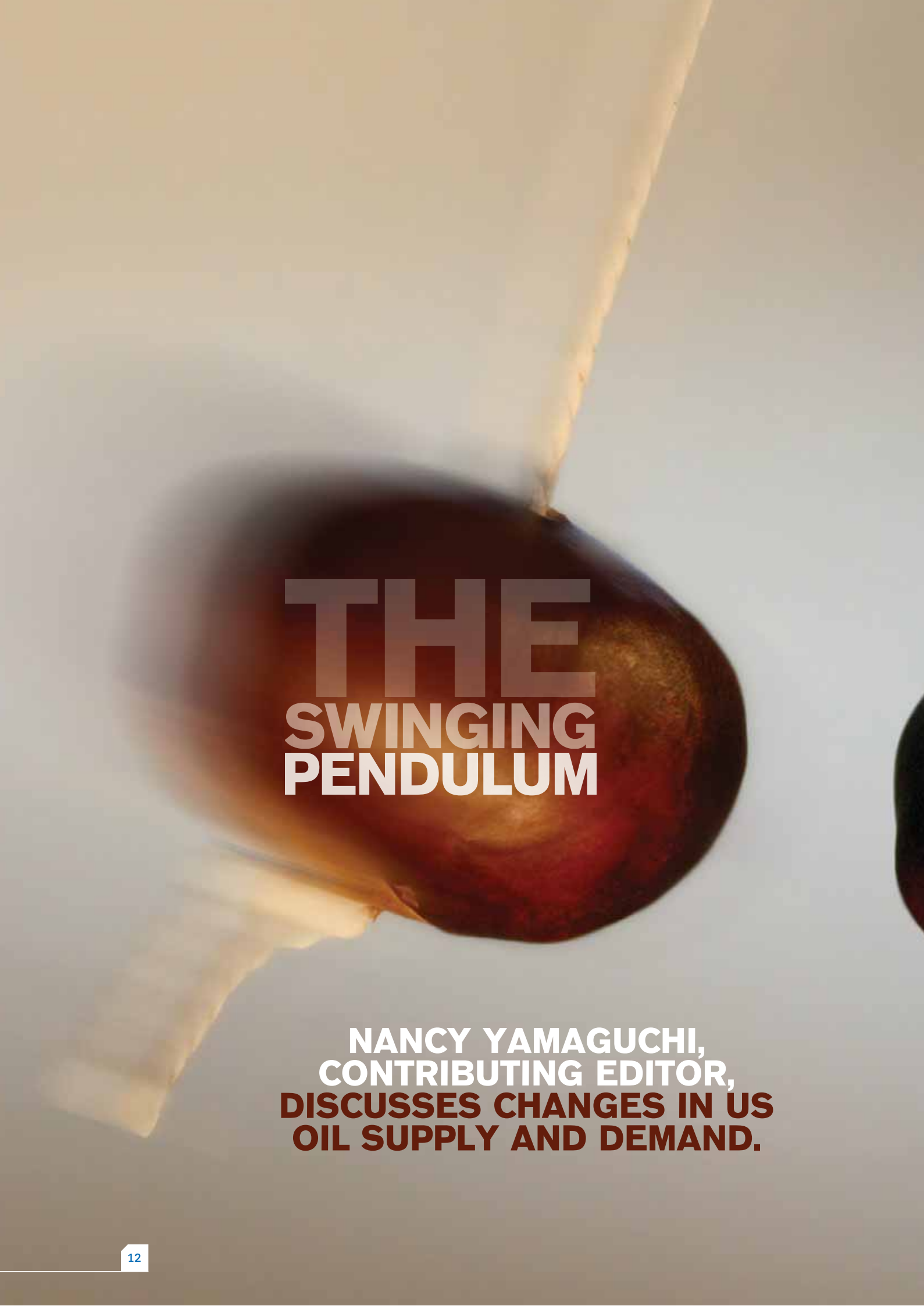


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THE SWINGING PENDULUM

**NANCY YAMAGUCHI,
CONTRIBUTING EDITOR,
DISCUSSES CHANGES IN US
OIL SUPPLY AND DEMAND.**



The US oil market is in a remarkable period of change, after years of what appeared to be staid and predictable performance. Demand was expected to grow modestly, domestic crude supply was expected to continue to fall, refinery capacity was viewed as stable, and import dependence was expected to continue its slow rise. Today, many elements are changing, and the supply/demand pendulum is swinging backward. The structure and volume of oil demand is changing, and there are distinct differences in supply and demand in regional submarkets. The development of shale oil resources is causing a resurgence in oil production, reducing the level of US dependence on imported crude oil. Expanded output in Canada, much of which flows to the US, is further reducing dependence on Middle Eastern crude imports. The concentration of new crude output in the centre of the country is giving an enormous boost to refineries with access to these new crudes. It is also causing a comparative disadvantage to other refineries, mainly those situated on the US West Coast and US East Coast. The new crudes are steeply discounted against other international crudes for two main reasons. First, crude transport infrastructure is overwhelmed by the new volumes. Pipeline transport is far and away the most economical means of transport, but most refineries on the western and eastern coasts have very little, if any, crude pipeline access to the new crudes. Second, US law severely restricts the export of domestic crude, the exceptions being trade with Canada, volumes from state waters in Alaska, and specially permitted cargoes. The issue of whether the US government should lift the ban on crude exports is once again under debate, yet concerns of supply security and import dependence remain.

The focus of this article will be the changes in the key elements of US oil supply and demand equation: the trend in demand, the increase in crude production, refinery utilisation, logistics, the impacts on product trade, and the overall change in US oil import dependency.

US oil product demand

Figure 1 displays the trend in US oil demand by product from 2000 through the January to April period of 2013, according to data published by the Energy Information Administration (EIA) of the US Department of Energy. Unless otherwise noted, the data used for the figures is the latest available

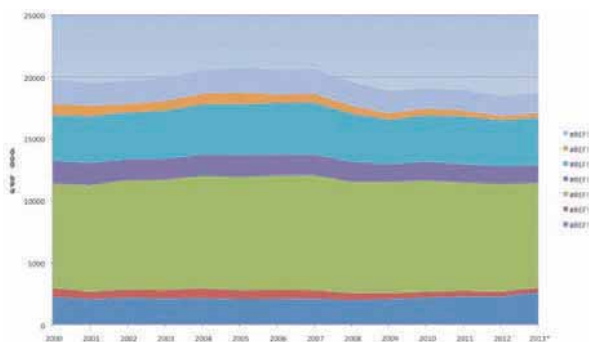


Figure 1. Trend in US product demand.

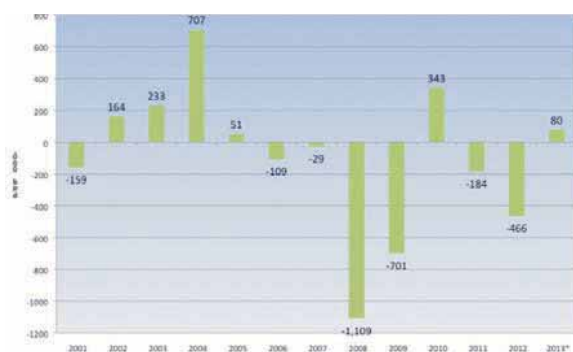


Figure 2. Incremental change in US oil product demand.



Figure 3. US gasoline demand.

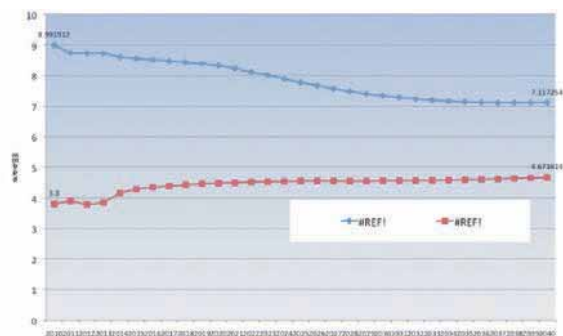


Figure 4. AEO long term forecast of US gasoline and diesel demand.

from EIA. At the beginning of the decade, most forecasts anticipated slow but steady product demand. Demand reached 20.7 million bpd in 2005, but it leveled off and began to stagnate. In 2008, a serious price shock hit the international market. The US economy had been languishing, and it fell into a serious recession. Oil demand fell in concert, despite a plethora of macroeconomic policies and programs designed to stimulate the economy.

Figure 2 presents a closer look at the changes in oil demand by calculating the incremental year on year changes in demand. Between 2007 and 2008, demand dropped by 1.109 million bpd. Between 2008 and 2009, demand dropped by another 701 000 bpd. Demand rebounded slightly by 343 000 bpd in 2010, then fell again by 184 000 bpd in 2011 and 466 000 bpd in 2012. Data for the first four months of 2013 shows an increase of 80 000 bpd, but few forecasts anticipate a serious long term rebound in US oil consumption.

Gasoline was the hardest hit product in the market. Figure 3 shows the steady growth in US gasoline consumption from 2000 to 2007, when demand rose from 8.472 million bpd to 9.286 million bpd, a growth rate of 1.3%/y. The oil price shock and recession caused demand to drop by nearly 300 000 bpd between 2007 and 2008. Gasoline demand has continued to decline, falling from 9.286 million bpd in 2007 to 8.503 million bpd during the January - April period of 2013. This equates to a downward trend of -1.5%/y. Moreover, the US gasoline market is expected to continue to shrink. Some of the recent drop in demand is attributed to poor economic performance and unemployment, which has reduced commuter miles and recreational driving. Some of the decline is attributed to progress made in Corporate Average Fuel Economy (CAFE) standards, vehicle fleet modernisation, and the spread of alternative fuel vehicles. Ethanol and other renewable sources are contributing close to one million bpd of non-oil liquid transport fuel supply. These changes are being wrought via public policy, and they should be expected to continue.

Note also, however, that diesel is one of the key fuels supplanting gasoline. Not only have diesel sulfur levels been brought to ultra low levels, enormous progress has been made in diesel engine technology. This has made diesel engine automobiles much more attractive to the American consumer. The improvements include quieter, lighter, lower compression engines that may achieve up to 30% better fuel economy than conventional gasoline engines. The USDOE/EIA recently forecast, in its Annual Energy Outlook, that distillate demand will grow at rates of 0.6%/y between 2010 and 2040, while gasoline demand will continue to shrink at an average rate of -0.7%/y. Figure 4 compares the AEO's long term forecast of gasoline and diesel demand. Starting from a baseline of 2010, gasoline demand is forecast to drop from 8.99 million bpd in 2010 to 7.12 million bpd in 2040, a drop of 1.87 million bpd in total. Distillate demand, in contrast, is forecast to grow from a 2010 baseline level of 3.8 million bpd to 4.67 million bpd in 2040, a rise of 0.87 million bpd. In other words, the US gasoline market was 5.19 million bpd larger than the diesel market at the start of the forecast horizon, but it is expected to be just 2.45 million bpd larger at the end of the forecast period. This is a significant shift toward dieselisation for the US market, which, unlike many markets in Europe and Asia, has long been a gasoline oriented market. The shift away from gasoline is a significant structural change, and it is having a major impact on refining and trade, as discussed further in the sections following.

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US GROWTH IN RENEWABLES TO DISPLACE GAS

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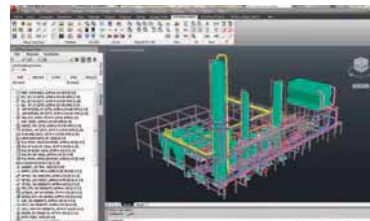
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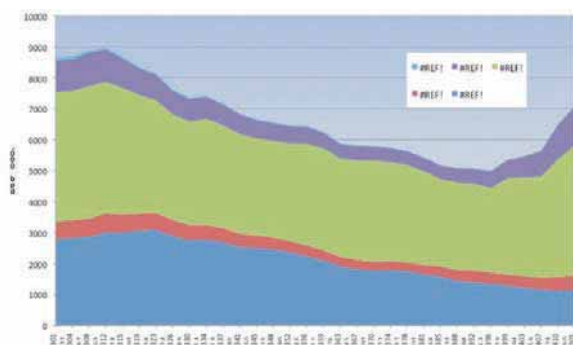


Figure 5. Rebounding US crude production.

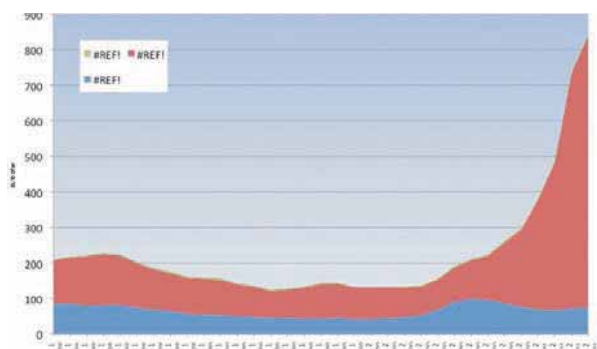


Figure 6. Crude production in the three key Williston production basin states.



Figure 7. Refinery utilisation rates by PADD.



Figure 8. PADD 2 domestic crude delivery modes: dominance of pipelines.

Rebounding crude supply and the shale boom

Without a doubt, one of the most remarkable turnarounds in the US oil market has been the rebound in crude oil production made possible through development of shale oil. Figure 5 displays US crude production by Petroleum Administration Defense District, or PADD. PADD 5 is the West Coast region. The strong crude output achieved in the late 1980s was chiefly the ramping up and peaking of the Alaska North Slope (ANS) oilfields, combined with solid output in California. But US West Coast crude production began to slide, falling from its peak of over 3 million bpd in 1988 - 1989 to approximately 1.1 million bpd at present. Production declines were in fact being experienced in all US PADDs. (PADD 1, the East Coast, is a minor producer only, barely visible on the chart). PADD 4, the Rocky Mountain states, saw crude production be cut in half, from 0.6 million bpd in the mid 1980s to approximately 0.3 million bpd by the year 2000. Production in PADD 2, the Midwest and Great Lakes region, also fell by more than half, dropping from 1.1 million bpd in 1984 to less than 0.5 million bpd by 1999. PADD 3, the Gulf Coast region, remained the largest producing region, led by Texas, but its crude output was also in decline, falling from approximately 4.3 million bpd in 1983 to approximately 2.7 million bpd in 2008. US crude production was viewed as being on a steady and unstoppable downward slide.

New exploration and development was limited, because many of the most promising oil prospective areas were in environmentally sensitive areas such as the Arctic National Wildlife Refuge (ANWR) and in offshore areas of Alaska, California, and the Gulf of Mexico. Allowing further exploration and development in these areas has been a contentious issue for decades. Public sentiment toward offshore drilling was hardened further by the Deepwater Horizon catastrophe in 2010, which caused the largest oil spill in US history.

The breakthrough in shale oil production via horizontal drilling and hydraulic fracturing, or 'fracking', pulled US production out of its slump. US crude production grew from 5.0 million bpd in 2008 to 7.1 million bpd in 2013 (January - April average). The two key shale deposits are the Williston Basin, which cuts across PADDs 2 and 4, and extends north into Canada, and the Eagle Ford shale play in PADD 3. While the Eagle Ford resource is relatively close to Texas refineries and the Gulf Coast, the Williston Basin resources are landlocked and far from centres of refining and population. As Figure 6 illustrates, crude production has soared in North Dakota, rising from less than 100 000 bpd in 2005 to 766 000 bpd during the first four months of 2013. North Dakota has only one refinery, Tesoro's 58 000 bpd gasoline oriented refinery at Mandan. Transporting Williston Basin crude to refiners is already posing a major challenge to transport infrastructure, and shale oil output is expected to reach 1.2 million bpd or more within the next six years.

Refinery utilisation and changes in trade flows

The new crude resources, particularly the Williston Basin output, have limited access to refineries. The booming production in the centre corridor of the US has suppressed crude prices, and refineries with access to the lower cost crudes are enjoying more favorable economics. Figure 7 compares refinery utilisation rates by PADD. PADD 1, the East Coast, is able to receive very little crude from the other PADDs, and East Coast refineries processes mainly foreign crude. Similarly, PADD 5, the West Coast, has very little

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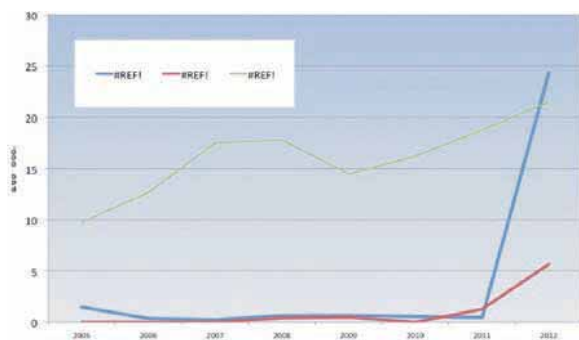


Figure 9. PADD 2 domestic crude deliveries, non-pipeline.

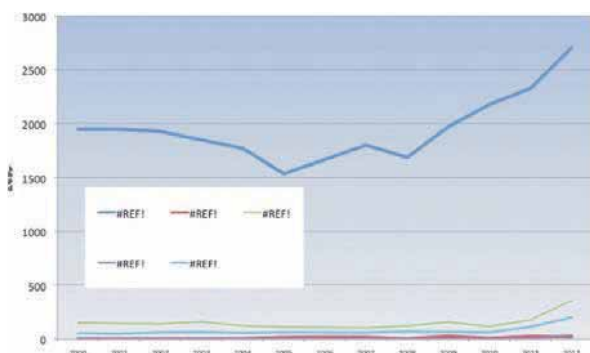


Figure 10. Crude oil delivery modes to PADD 3 refiners.



Figure 11. PADD 3 domestic crude deliveries, non-pipeline modes.



Figure 12. Reduction in pipeline transport of crude and product from PADD 3 to PADD 2.

access to crude from other PADDs. The West Coast market is more closely linked to the Asia Pacific market, while the East Coast market is tied to the Atlantic Basin market. This has had a marked impact on refinery competitiveness. As the Figure illustrates, refinery utilisation rates have fallen in PADD 1 and PADD 5, while they have strengthened in PADDs 2, 3 and 4. In 2012, PADD 1 refinery utilisation averaged 79.4%, and utilisation is averaging 81% for the first four months of 2013. PADD 5 refinery utilisation was 82.9% in 2012 and 79.4% more recently. In contrast, PADDs 2, 3 and 4 all enjoyed refinery utilisation rates of over 90% in 2012. During the January - April period of 2013, PADD 2 refinery utilisation averaged 86.4%, PADD 3 utilisation averaged 84.1%, and PADD 4 utilisation averaged 91.6%.

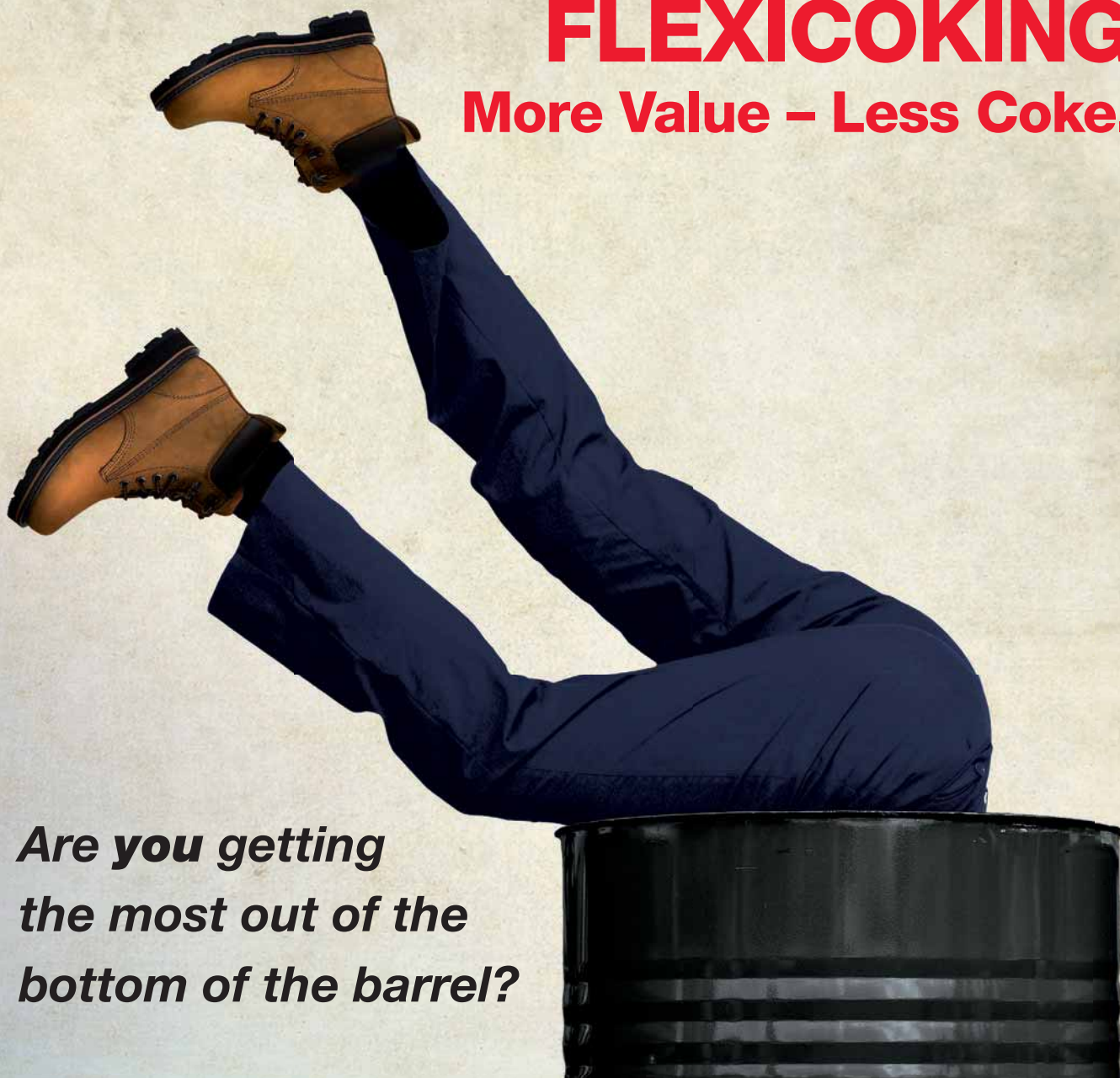
The country's oil transport infrastructure is straining to optimise trade flows. As noted, pipelines are typically the most economical means of transport, often by orders of magnitude. But the pipelines are running at capacity, despite line reversals and expansions, more of which are planned and underway. Figures 8 and 9 provide an illustration of how the new output is changing how domestic crude is delivered to refineries in PADD 2 (the Midwest and Great Lakes region.) Although the Great Lakes provide some capability for waterborne transport, pipelines are far and away the key mode of crude transport. In fact, pipeline transport is so dominant that in Figure 8, the other modes (tanker, barge, tank car and truck) are barely visible. But by removing the crude volumes transported by pipeline, Figure 9 is able to reveal how much growth there has been in crude deliveries via truck, barge and tank car. A number of pipeline expansions have been completed and others are planned and/or underway to cope with Williston Basin and other crudes, including the Enbridge Sweet expansion, the Enbridge Bakken expansion, the Butte Pipeline Expansion, the Butte Loop, Plains Bakken North, Tesoro's High Plains Pipeline System, and the Keystone XL interconnect. The Keystone XL project recently has gained a considerable amount of public approval, now that it has been rerouted. A number of rail transport options also are being expanded, including projects by EOG Rail, Dakota Transport Solutions, Hess Rail, and Rangeland.

This pattern holds true to the south also, in PADD 3, the Gulf Coast. Figure 10 shows the transport modes used to deliver domestic crude to refineries. Once again, pipeline transport is the main mode, and pipeline deliveries have grown strongly. But the pipelines are running at capacity, handling the new crudes and the new trade patterns. Figure 11 focuses on the non-pipeline modes, showing the upsurge in barge, truck, rail car and tanker transport. Even though these modes of transport are costly, it is still less costly to move most of the new domestic crude to PADD 2 and PADD 3 refineries rather than to the East Coast or West Coast. A variety of pipeline projects have been completed and are underway or planned, including plans by Kinder Morgan Energy Partners and Plains All American Pipeline. The Seaway Pipeline from Cushing Oklahoma to the Gulf Coast was recently expanded to 400 000 bpd, helping to ease some of the midcontinent pressure. One of the Kinder Morgan projects is the Freedom Pipeline, which if built would extend from the Permian Basin in Texas all the way to California.

The new supplies and high levels of refinery throughput in the centre of the country also have changed the flows of crude and product between PADDs. Figure 12 presents the trend in pipeline shipments of crude oil and refined product from PADD 3 (the Gulf Coast) to PADD 2 (the Midwest and Great Lakes.) As the Figure

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illustrates, upwards of 2 million bpd of crude was sent to PADD 2 refineries via pipeline from PADD 3. In recent years, however, this has been halved, and it averaged 0.94 million bpd in 2012. PADD 3 is also the country's largest refining centre, and refined product shipments from PADD 3 to PADD 2 were approximately 1 million bpd from 2004 through 2007. But these transfers fell to approximately 0.63 million bpd in 2012 as PADD 2 ramped up its own refinery activity.

Indeed, the rise in crude availability and the high refinery utilisation rates in PADD 2 have reversed the flow, and PADD 2 is now sending more crude and product south to the US Gulf Coast. As Figure 13 shows, PADD 2 crude shipments via pipeline were in the vicinity of 50 000 bpd for nearly two decades until they began to climb in 2009, and they quadrupled to 200 000 bpd in 2012. Pipeline shipments of refined product from PADD 2 to PADD 3 had

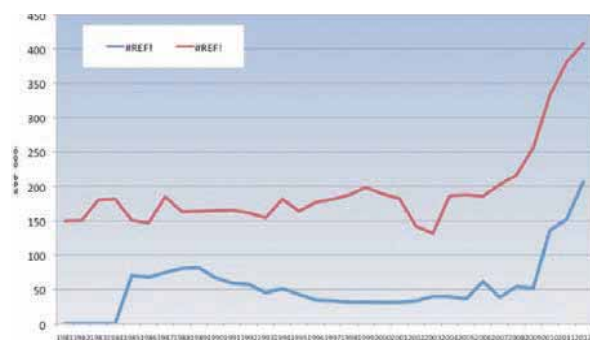


Figure 13. Increased pipeline transport of both crude and product from PADD 2 to PADD 3.



Figure 14. The US has become a net exporter of gasoline.



Figure 15. The US has become a major net exporter of diesel.

been in the range of 150 000 bpd to 200 000 bpd for decades, yet they soared to over 400 000 bpd in 2012.

With a localised surfeit of crude, and few options for export, crude prices have fallen, and the local refineries have kept throughput high. As noted, however, US oil demand has been weak. Demand for gasoline is falling. As Figure 14 shows, the US historically has been a net importer of gasoline, but this reversed by 2010. In 2005, the US imported 603 000 bpd of gasoline and exported 136 000 bpd. Imports then began to slide steeply, falling to 55 000 bpd in 2012 and averaging 38 000 bpd during the January - April 2013 period. Refinery production has remained high nonetheless, and exports climbed to 479 in 2011 before subsiding slightly to 423 000 bpd in 2012 and 414 000 bpd in January - April 2013.

The US has become an even larger exporter of diesel, as shown in Figure 15. From 2000 through 2006, the US was a net importer, typically importing 270 000 - 350 000 bpd of diesel while exporting 110 000 - 200 000 bpd. By 2008, the country became a net exporter, exporting 530 000 bpd of diesel while importing 214 000 bpd. Diesel imports have dipped below 200 000 bpd, while exports have soared, hitting a peak of 1.011 million bpd in 2012. Diesel exports averaged 810 000 bpd during the first four months of 2013. The US is now a net exporter of every major class of petroleum product.

Concerns over import dependence and crude trade

The changes in US crude and product flows have been enormous, and it is clear that some of the changes are being caused by the lack of sufficient transport infrastructure and the lack of international outlets. With most US crude restricted from export markets, and with a lion's share of the new crude production in landlocked states, the only practical option is to run the crude through existing refineries, whether or not they are geared for it. Many refineries are not geared toward the new light output, having been built and upgraded to take advantage of heavy sour crudes and bitumen based products from Canada. This has resulted in high refinery utilisation rates in the centre of the country, but the bulk of the population and the demand is located on the coasts. A surplus of both crude and product in the centre of the country is creating new trade patterns, as oil is pushing out along hitherto little used avenues. Naturally, some of these trades are occurring because crude exports are largely disallowed while exporting refined products is not.

There are many who espouse the idea of fully eliminating the restrictions on crude exports. It is inarguably a distortion of trade, but it has been a part of the market literally for decades, and politically it is difficult to undo. In the aftermath of the 1973 - 1974 Arab Oil Embargo, the US grew highly concerned over the issue of supply security. The Energy Policy and Conservation Act was enacted in 1975, followed by the Export Administration Act of 1979. The Iranian Revolution caused another serious price shock in 1979 - 1980, heightening public concern over oil import dependence. Although it is common to refer to the US policies that emerged as the 'US Crude Export Ban', the policies technically are not a ban. Some crude exports were to be allowed under certain definitions, and individual licenses could be granted if they were judged to be consistent with the national interest. Applications were to be reviewed and approved by the Bureau

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of Industry and Security, an agency housed within the US Department of Commerce. The licensing requirements are codified under the section 'Short Supply Controls'. President Ronald Reagan lifted the restrictions on exports to Canada in 1985, when Canada decontrolled its oil prices. It was considered that a significant oil trade would grow between the US and Canada, though the trade has been overwhelmingly from Canada to the US. The State of Alaska was also permitted to export some oil produced in state waters (called Cook Inlet crude or Drift River Blend). The State of California was allowed to export up to 25 000 bpd of heavy crude (20 degrees API and heavier). The restrictions on exports also have been relaxed during times of emergency, such as in the aftermath of a hurricane.

Figure 16 provides a look at US crude exports by destination. In the early 1990s, approximately 95 000 bpd of crude was shipped to the US Virgin Islands, site of Hess Oil's giant refinery. This trade vanished when the refinery began to operate in 1998 as the

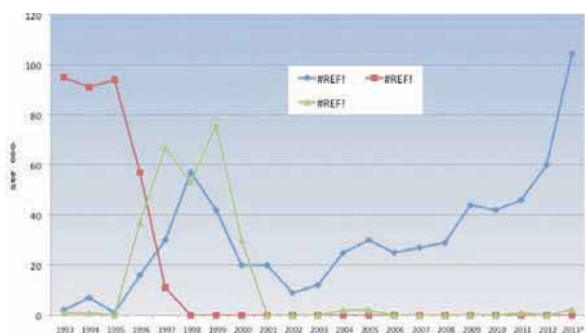


Figure 16. US crude exports by destination.

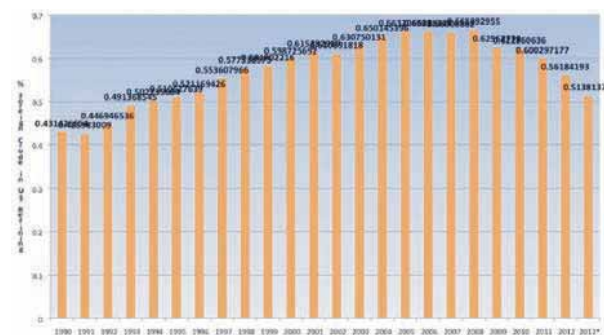


Figure 17. The falling percentage of US crude inputs into foreign refining.



Figure 18. US crude import dependency: Canada, Saudi Arabia, and OPEC.

Hovensa joint venture between Hess and the Venezuelan national oil company, PDVSA. (This refinery eventually closed entirely). In the mid to late 1990s, the US began to export significant volumes (50 000 - 100 000 bpd) of PADD 5 crudes to Canada, Japan, South Korea, China and Taiwan. But PADD 5 production waned, and these exports vanished. PADD 2 is now the main source of crude exports, trading with Canada, though every now and then a cargo has been sent elsewhere. Although the US exports very little crude, the increase in production has corresponded with an increase in exports. In 2002, the US exported just 9000 bpd of crude. This rose to 60 000 bpd in 2012, and is averaging nearly 105 000 bpd during the first four months of 2013.

The arguments for removing all restrictions on crude exports have resurfaced, and they have been given considerable impetus by the new developments in the US market. The US supply demand pendulum has been swinging back: demand has been falling, domestic crude production has been rebounding, ethanol production has risen, and Canadian supply is expanding. Oil prices have been high for several years now, and the US economy has been lacklustre. Many people remain concerned with the high cost of imported oil and with the idea of supply security. But it is not at all clear that the restrictions on US crude exports have improved supply security. Moreover, the terms 'supply security' and 'import dependence' mean different things to different people, and many do not yet realise how significantly the US oil balance has changed.

The US oil balance has improved to a remarkable degree. Figure 17 presents the percentage of foreign crude input to US refining from 1990 through the January - April 2013 period. The share of foreign crude was rising steadily, moving from 43% in 1990 to 67% in 2008. The price shock, recession, and shale oil boom have combined to reduce this to 51% currently, the lowest level seen since 1995.

Imports from 'unfriendly' or 'unstable' countries, typically defined as the OPEC countries, are also now much less than commonly thought. For example, the general press often cites crude imports from Saudi Arabia as a measure of US import dependence. In reality, imports of Saudi crude have declined significantly over the past decade, having averaged 1.7 million bpd in 2003 versus 1.36 million bpd in 2012 and 1.1 million bpd during the first four months of 2013. Without debating the relative 'friendliness' of countries exporting oil to the US, it easily can be said that Canada is a trusted and important trading partner with a long shared border. In 2003, Canada exported 1.55 million bpd of crude to the US, around 150 000 bpd less than Saudi Arabia did in that year. By 2012, however, Canadian crude exports to the US had grown to 2.4 million bpd, and they are averaging over 2.6 million bpd at the present time.


As Figure 18 establishes, of the foreign crudes imported by the US, OPEC's share has fallen from a peak of 86% in the mid 1970s to 45% at present. Of this, Saudi Arabia's share is only 14%. Canada's share is 35%. If import dependency is recalculated using the geographical regions of the Persian Gulf and the Western Hemisphere, rather than using OPEC, US dependence on Persian Gulf crude has fallen to 25%, while import dependence on other Western Hemisphere crude producers has risen to 65%.

Conclusion

The US oil market finds itself in a period of unusual change. After many years, decades even, of what seemed to be slow moving and predictable behaviour, several of the moving parts in the oil supply

and demand equation are now heading in different directions. Oil demand, which had been expected to grow slowly but more or less steadily, fell sharply after the price spike in 2008 and the economic recession. Current demand is approximately 2.5 million bpd below its level in 2007. Demand for gasoline, in particular, has dropped, and there is little reason to expect a resurgence of growth in gasoline demand. US crude production has pulled out of its long downward slide, growing from 5.0 million bpd in 2008 to 7.1 million bpd in 2013 (January - April average). Foreign crudes now account for 51% of the refinery slate, down from a peak of 67% in 2008. Western Hemisphere crude exporters, led by Canada, provide 65% of this, while exporters in the Persian Gulf provide only 25%.

The US oil transport infrastructure is labouring to cope with these changes. The new shale oil is being produced in the country's central corridor, and pipelines have been running at full capacity. A number of pipeline expansions have been completed and more are underway and/or planned. Additionally, rail transport options have been expanding. Even crude movements by truck have expanded, though truck transport is usually the highest cost mode on a t/mile basis. With restrictions on crude export, refineries with access to the new low cost crudes have been running at high utilisation rates, while refinery utilisation rates have fallen on the East and West Coasts. The US has become a net exporter of all classes of product. In fact, when net product exports are factored into the equation, US oil import dependency drops significantly. When net crude imports are stacked next to total oil product supplied, import dependence amounts to 47%. However, when net crude plus net product imports are considered, US oil import dependence drops to 41%. Essentially, the US is exporting some of its extra crude in the form of refined product.

The US oil market is complex, and peopled by companies, groups and individuals with strong views, often working at cross purposes. Free market economists tend to favour the elimination of policies such as restrictions on energy exploration and development, crude oil exports, transport modes, and product trade. But the US energy market has evolved with a number of political, economic and environmental regulations that are so deeply ingrained that changes come slowly. The pendulum is swinging back, however; domestic energy demand has fallen, and domestic supplies are expanding. US oil import dependence has fallen significantly. Some geologists even believe that it is possible for the US to become fully self sufficient in oil supply, though many urge caution about such forecasts for a variety of environmental, political and technical reasons. Moreover, most large petroleum markets operate most efficiently only when free trade is allowed, so it is unlikely that shale oil and other developments would continue to climb without access to international markets. Fully satisfying every market demand in every corner of the continent using only domestic resources is inherently inefficient. There is no shortage of energy modeling and analysis identifying energy market reforms and strategies that would increase efficiency and maximise profits. But the public inevitably asks, maximise whose profits? And thus the politics of energy in the US remain complicated. But the current relaxing of the oil supply and demand balance may indeed continue, and insofar as it visibly improves the country's economic situation without damaging the environment, many new directions are now becoming possible in the US. 



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GORDON COPE, CONTRIBUTING EDITOR, DISCUSSES THE SIGNIFICANT CHALLENGES FACING RENEWABLE FUELS.

Renewable fuels, also known as clean fuels, have several advantages over fossil fuels. They are considered good for the environment, and one can produce them wherever there is plant life, making them immune to Organisation of the Petroleum Exporting Countries (OPEC) fuel embargoes. Various governments have established mandated usage of renewable fuels at significant levels. In 2003, the EU set a directive to obtain 5.75% of transportation fuel needs from renewable fuels by 2010. The EU now has a target of 10% by 2020. The US Congress enacted various renewable fuels standards (RFS) mandates, culminating in the 2007 Energy Independence and Security Act (EISA), that guarantees a US market of renewable biofuels amounting to 36 billion gal./y by 2022.

But the economic, fuel production and consumption circumstances that fueled the boom in renewables legislation in the 2000s have largely been reversed. Are clean fuels ultimately doomed?

Biodiesel

Biodiesel use is mandated under EISA. The Environmental Protection Agency (EPA) set a target of 1.28 billion gal. for US markets in 2013. The most common production method, known as base catalysed transesterification, is to react vegetable oil with methanol in the presence of lye. This process removes glycerine, and the resulting mono alkyl esters of long chain fatty acids perform similarly to petroleum diesel. Although most biodiesel is made from soybean byproducts, anything from recycled oils from chip vats to coffee grounds can be used. The fuel is environmentally benign, biodegradable when spilled, and reduces the amount of exhaust particulates.

But biodiesel has been known to clog fuel filters, form gels in cold weather and absorb water (which reduces the heat of combustion and promotes corrosion and microbe growth). Its energy content is approximately 11% less than petroleum diesel, making it less fuel efficient by volume. Its high oxygen content also increases nitrogen oxide emissions.

Finally, biodiesel is more costly than fossil fuel diesel. When the EPA announced its 2013 target, industry response was negative. 'Given the exorbitant cost of biodiesel, its poor performance qualities, significant fraud in the biodiesel industry, and the drought facing our nation's farmers and ranchers, this is a bad decision at the wrong time,' said American Fuel & Petrochemical Manufacturers (AFPM) president, Charles Drevna.

Ethanol

Ethanol is the main biofuel being produced under the aegis of the EISA. An ethanol facility extracts sugars from plants (corn in the US, sugarcane in Brazil), then ferments the sugar into alcohol using yeast. The alcohol, at 8% volume, is then repeatedly distilled until it reaches over 99% purity.

Like gasoline and diesel, ethanol emits greenhouse gases (GHGs), in the form of CO₂. But because ethanol is made from living organisms that trap carbon during photosynthesis, net lifecycle analyses generally conclude that its impact on GHG is neutral, or beneficial. Ethanol is also an efficient replacement of methyl tertiary butyl ether (MTBE), a gasoline additive that was phased out due to a propensity to contaminate surface and groundwater.

Ethanol's energy content is significantly lower, however; it takes approximately 1.5 ltrs of ethanol to drive as far as 1 ltr of gasoline. Ethanol is very corrosive, and high concentrations can dissolve seals and gaskets in pipelines, containers and engines. Production is also costly; capital costs range up to US\$ 33 000/bpd of capacity. Price surges in staple commodities such as corn have been linked to its usage for fuel.

Thanks to regulatory support, however, ethanol production has run into the 'blend wall'. In 2010, the US consumed 13.2 billion gal. of ethanol, which meets the 10% blend (E10) requirement of all gasoline sold in the country. Fuel sales have been flat and decreasing due to the global recession, however, and producers applauded the EPA's decision to allow 15% ethanol usage for all cars and trucks built after 2000.

The largest challenge to ethanol, however, is the Renewable Fuel Standard (RFS) mandate within EISA which states that, of the 36 billion gal. of renewables, no more than 15 billion gal. can be met by ethanol produced from corn; the next 21 billion gal. must come from non-food sources. Various techniques have been used to convert cellulosic (non-food) organic matter to ethanol. In 2008, BP announced it would open a 36 million gal./y cellulose ethanol facility in Florida. That same year, POET, which produces over 6 billion ltrs/y of corn based ethanol, teamed up with Netherlands based Royal DSM to develop a plant in Emmetsburg, Iowa, to make ethanol from corn crop residue. In 2010, Iogen Energy, of Canada, allied with Royal Dutch Shell to develop commercial cellulosic ethanol technology and open a C\$ 400 million plant in Saskatchewan to produce 70 million ltrs/y from corn husks.

Commercial scale cellulosic ethanol production has proven to be far harder than food based facilities, however. Of the three projects listed above, only the POET facility is expected to open, in 2014. This has led to a situation where the EPA is mandating the inclusion of cellulosic ethanol at production levels that do not exist. The result has been a contentious series of court challenges between industry and government that has created widespread uncertainty in the future of renewable fuel standards.

Equally as damaging is the growing realisation that the drivers for EISA have been reversed, making its goals and aspirations questionable, if not irrelevant. Back in 2007, crude prices had been

rising steadily from a low of US\$ 20, to an eventual high of US\$ 145 in 2008. Crude oil production in the US had been declining for over two decades, and the Energy Information Administration (EIA), was predicting gasoline production to climb from 142 billion gal./y in 2007 to 168 billion gal./y in 2020. Ethanol was seen as a cheap and reliable means of dealing with these challenges.

Fast forward to 2013: While oil prices are hovering in the US\$ 100 range, domestic production of both natural gas and crude are on the rise, thanks to the shale gas and oil revolution in the sector. Gas consumption was gutted by the recession, and now stands at 133 billion gal./y. The price of corn, a food staple in many poor countries, has doubled from US\$ 3/bushel to US\$ 6.

Currently, several bills, including the RFS Reform Act, seek to eliminate corn based ethanol requirements, cap the amount of ethanol that can be blended into conventional gasoline at 10%, and require the EPA to set cellulosic biofuels levels at production levels.

But North America is still a long way off from energy independence, and the issue of GHGs causing climate change is still not settled; the recent announcement that CO₂ had reached 400 ppm reminded everyone that humanity is changing the chemistry of the atmosphere with very little understanding of the consequences. Are there alternatives to gasoline and diesel that do not increase GHGs, do not put pressure on food prices, but do promote energy independence?

Viable alternatives

Compressed natural gas

In North America, compressed natural gas (CNG) is currently used to replace gasoline in some taxi and delivery vehicle fleets, but it accounts for only a tiny fraction of the fuel use. CNG has various advantages that make it a candidate to control a sizable portion of the fuel market in the near future. Life cycle emission studies conducted by the Argonne National Laboratory show that CNG emits up to 11% lower GHGs than gasoline, as well as dramatically lowering the amount of carbon monoxide and nitrogen oxides coming out of the tailpipe. In addition, CNG is non-toxic, and will not harm soil or water.

But the greater beneficial factor is cost. Thanks to the combination of horizontal drilling and hydraulic fracturing that has opened up shale gas as a commercial success, basins throughout the Lower 48 are now producing large volumes of natural gas. US production has risen and prices have plummeted. When you compare energy equivalency, CNG is one fifth the price of gasoline.

Small wonder, then, that CNG is being heavily lobbied in Congress. T. Boone Pickens, a Texas oil billionaire, has been touting a plan to convert the continent's approximately 7 million transportation trucks from diesel burning to natural gas. In addition to reducing the US dependency on imported oil by 2.5 million bpd, it would also promote the use of surplus gas. Vehicle conversion is expensive, however, because transportation vehicles have to meet stringent emission controls. Most participants agree some form of subsidy program is needed to accelerate the switch.

Another concern is a lack of fueling infrastructure, with most urban centers having less than a handful of stations. An innovative and inexpensive alternative may be to employ portable fueling stations. The Gas Technology Institute recently fielded the FuelMule, a mobile fast fill CNG fueling station that can be positioned anywhere. It has enough fuel onboard to fill up to 50 heavy duty trucks and can dispense the fuel quickly.

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Hydrogen

Hydrogen is one of the most common elements on earth. As a fuel, it reduces urban pollution, cuts down dramatically on GHGs, and enhances energy security for oil importing nations. Hydrogen can be obtained from water through the electrolysis process; a strong electrical current is passed through water to generate bubbles of hydrogen at the cathode.

Electrolysis requires large amounts of energy per unit of hydrogen produced, however. And most hydrogen is produced through the steam methane reformer (SMR) process. Natural gas is mixed with high temperature steam under pressure in the presence of a catalyst; hydrogen and carbon monoxide (CO) are produced. Approximately 50 billion ft³/d of hydrogen is already produced globally; the majority is used in refineries to upgrade less valuable fractions of crude into gasoline.

Hydrogen possesses such great potential that carmakers are planning to manufacture hydrogen fuel cell vehicles (FCVs). In January, Toyota and BMW announced they had forged an alliance to develop FCVs, about the same time that Nissan said it would team up with Daimler AG and Ford Motor Co. All FCVs rely on a proton exchange membrane fuel cell. Hydrogen from a storage tank and oxygen from the air are injected into the fuel cell to generate an electrical current. The current is sufficient to power an electric motor, which in turn drives the vehicle wheels. The main exhaust to the vehicle is water vapour.

Unfortunately, hydrogen faces a long list of challenges. Most hydrogen production also consumes fossil fuels directly (natural gas), or indirectly (coal fired plants). Although it costs around US\$ 1 to produce 1 kg of hydrogen (415 ft³), delivering it to retail outlets raises the price into the US\$ 14 - 15/kg range. Hydrogen fueling stations are currently very rare, and the Department of Energy (DOE) estimates it would cost US\$ 300 billion spread over 30 years to build a nationwide distribution system.

But none of these challenges are insurmountable. Wind and solar power can be used as electricity sources. A mobile device similar to the FuelMule would greatly reduce infrastructure costs. Cella Energy, a spin off from Britain's Rutherford Appleton Laboratory, is devising a technology that allows far larger volumes of hydrogen to be loaded and stored in a vehicle's fuel tank. The process uses hydride nano fibres, which are capable of adsorbing large amounts hydrogen. If roadblocks are removed, industry experts foresee the potential of five to 10 million fuel cell cars on the road in 2020, 50 million by 2030, and 100 million by 2040.

Algae

Algae and other microbes hold great promise. They thrive in non-commercial environments with little more than basic nutrients and sunshine. They grow far more rapidly than conventional crops, and generate a much higher fraction of their biomass as oil (up to 60%, versus 2 - 3% for soybeans).

The most common processes for algae fuel production start by pumping nutrient rich water into open pond systems or plastic tubes containing specially chosen species. The algae, in turn, use photosynthesis to produce high quantities of rich lipids which can then be converted into biodiesel and related fuels. Sapphire Energy currently has a 1.5 million gal./y Green Crude Farm operating in Columbus, New Mexico.

Creating biofuels from microbes has many advantages. Algae can grow in low lying areas unsuitable for conventional crops. It

can yield 8000 ltrs of fuel per acre per year, compared to 2600 ltrs for palm oil and 200 ltrs for soy. Algae can use brackish water or wastewater as a growing medium, eliminating the freshwater needs of ethanol production. Importantly, algae production does not compete with food crops such as corn or soy for acreage, nutrients or fresh water. Furthermore, biofuels are similar enough to gasoline and diesel that they do not require special treatment during transportation and mixing at the refinery.


Making biofuels from microbes is costly, however. Industry estimates place the production costs anywhere between US\$ 3.50 and US\$ 15.00/ltr, depending on the type of algae and the processing technique used. One of the most significant costs consists in separating the biomass from solution. Typically, centrifuges are used to raise the concentration from approximately 1 mg/ltr to 200 mg/ltr (where it forms an easily handled paste). Recent studies conclude that this algae dewatering process costs over US\$ 3000 in energy alone to produce one tonne of dry weight biomass equivalent, making algae an uneconomic source of fuel when compared to fossil fuels.

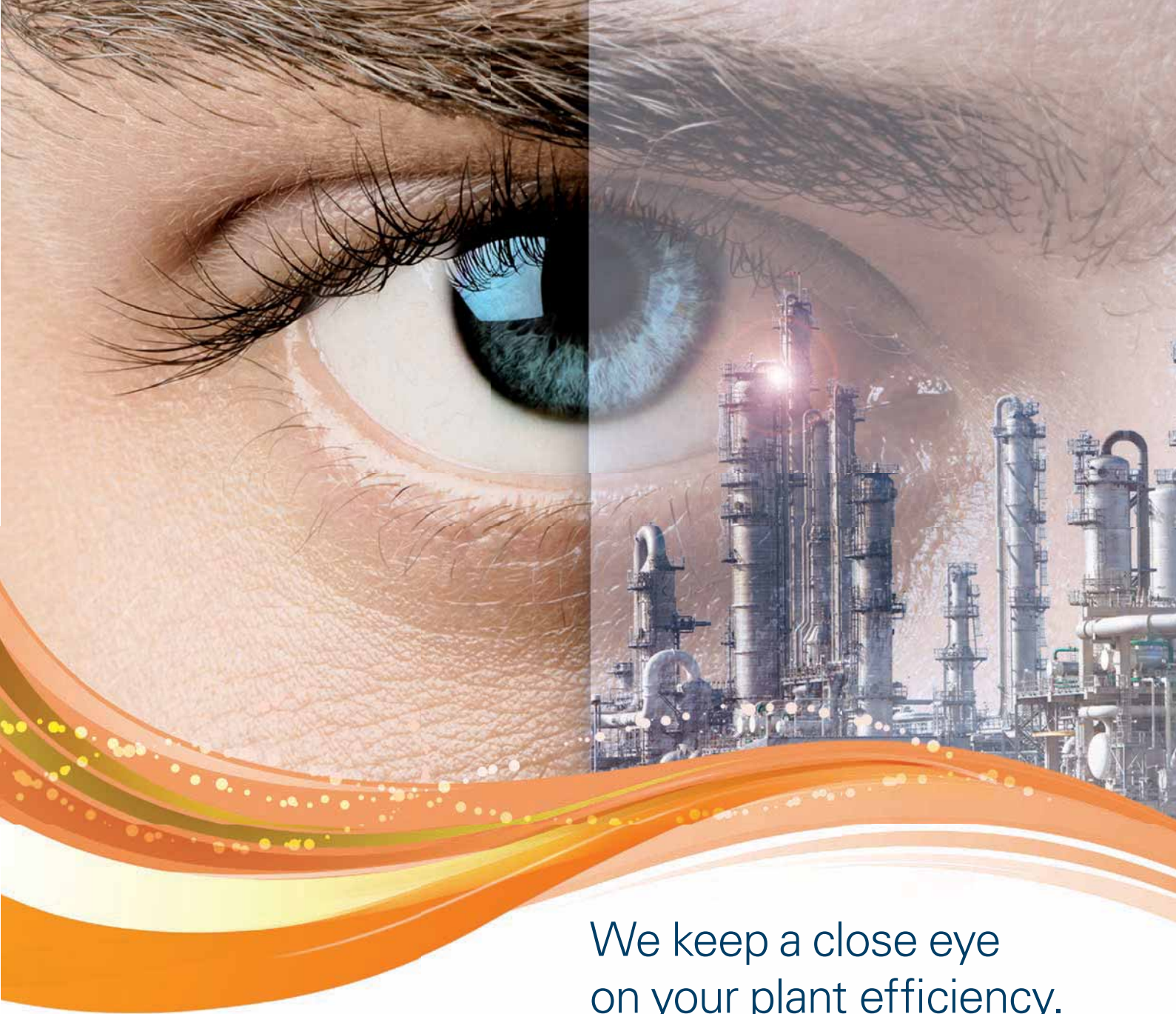
Biofuel mandates for vehicles and airplanes are spurring R&D, however. Synthetic Genomics Inc (SGI) formed a joint venture with ExxonMobil to engineer more productive oil producing microbes. Other companies are looking to reduce the expensive concentration process;

- Pall Corporation has developed the Algae Separation & Concentration Filter (ASCF), which utilises a robust, hollow fibre filter technology. As the algae and water flows through, 100% of the algae biomass is captured and concentrated in the middle of the hollow fibre. The process can increase concentration from a 0.5 g/ltr harvest to a 200 g/ltr slurry, using a fraction of a centrifuge's energy consumption.
- California based OriginOil has developed the Solids Out of Solution™ (SOS) process, which employs two stages: extraction and then concentration. In the first stage of the harvesting process, electrodes induce the algae cells to expel ions that alter the cells' surface charges, inducing the algae to flocculate. In the second stage, an electromagnetic pulse creates gas microbubbles that lift the flocculated algae cells to the surface of the water column. The algae biomass is removed, and the water is recycled back to the bioreactor. The system is low cost, low energy, and chemical free.

The future of algae biomass looks bright, regardless. Several major petroleum companies have invested heavily in algae research. BP and Martek Biosciences Corp, for instance, have teamed up to convert sugars into biodiesel using microbes. The new dewatering processes which are being developed promise to lower operating costs significantly. Finally, algae biofuels are very environmentally friendly when compared to fossil fuels.

Conclusion

In conclusion, it is relevant to note that renewables have been around in significant amounts for less than a decade, and that growing pains are inevitable. Yet much has been accomplished in terms of meaningful production, distribution and acceptance by the public. A great deal of regulatory refinement and legislation needs to be done to work out the wrinkles, and R&D promises a host of new clean fuel sources. Clearly, there is a market for clean fuel, but equally as clearly, the market has to evolve for it to prosper. 



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A HIGHER TIER

DAVID WALL, TRINITY CONSULTANTS,
USA, DISCUSSES TIER 3 LOW SULFUR
GASOLINE REGULATIONS AND THEIR
EFFECTS.



On May 21st, 2013, the US Environmental Protection Agency (EPA) proposed new 'Tier 3' motor vehicle emission and fuel standards to further control air pollution from motor vehicles across the US.¹ If finalised, this proposal would establish more stringent vehicle tailpipe emissions standards and reduce the allowable sulfur content of gasoline beginning in 2017. The proposal follows the approach of previous standards; coupling new requirements for motor vehicle manufacturers with requirements for fuel producers. The logic is that a lower gasoline sulfur standard would make emission control systems more effective for both existing and new vehicles, and would enable more stringent vehicle emissions standards.

Low sulfur levels in gasoline would reduce sulfur dioxide (SO₂) emissions from gasoline combustion. SO₂ contributes to acid rain, smog, and particulate matter. Sulfur in gasoline also decreases the lifetime and effectiveness of a vehicle's selective catalytic converters used to control tailpipe emissions. Thus, lowering gasoline sulfur levels helps motor vehicle manufacturers meet the increased control technology effectiveness requirements also included in the proposed rule.

The EPA expects these new standards to significantly reduce pollutants such as ozone, particulate matter, and air

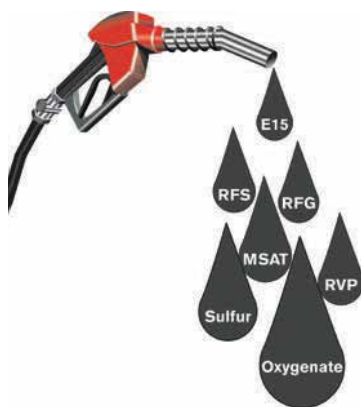


Figure 1. US clean fuel regulations for gasoline.

toxics nationwide and will assist state and local agencies in attaining and maintaining the health based National Ambient Air Quality Standards (NAAQS). Mobile sources, such as motor vehicles,

continue to be significant contributors to air pollution in areas not achieving the NAAQS.

The US refining industry; however, has voiced concerns regarding the timing, cost, and benefit of implementing these new requirements.

History of US gasoline standards

Empowered by the 1990 Clean Air Act amendments, the EPA has promulgated several increasingly stringent and complex fuel standards over the last 30 years. The numerous fuel standards vary greatly from state to state, and different sets of standards apply during different times of the year. This amalgam of requirements poses an extreme challenge for the refining industry, and creates barriers to rapidly addressing regional fuel shortages (e.g. resulting from unexpected refinery downtimes caused by extended maintenance, accidents, and natural disasters) as specific refineries are not positioned to produce compliant gasoline for all of the various regions of the country. As a further obstacle, meeting these fuels standards often runs contrary to producing fuels that meet adequate performance specifications (e.g., octane).

Figure 1 illustrates various gasoline standards that refineries are currently challenged to meet. These standards are summarised in the following sections.

Gasoline sulfur

Previously, gasoline sulfur has been regulated under 'Tier 1' and 'Tier 2' standards. The Tier 2 standards were phased in beginning in 2004 as the Tier 1 standards phased out. These requirements jointly limited the sulfur content in gasoline while placing new requirements on motor vehicle manufacturers. Tier 2 standards set gasoline sulfur content limits for refineries of 30 ppm on an annual average basis, with no batches allowed to be more than 80 ppm sulfur. This represented a reduction of approximately 90% from previously uncontrolled levels.

Mobile source air toxics

The mobile source air toxics (MSAT) rules were promulgated in 2007 to reduce hazardous air pollutants (i.e., air toxics) emitted by motor vehicles. Air toxics of concern from motor vehicles include benzene and other hydrocarbons such as 1,3-butadiene, formaldehyde, acetaldehyde, acrolein, and naphthalene. In addition to further requirements for vehicle manufacturers, the MSAT rules set requirements for refineries to meet benzene composition limits beginning in 2011. Specifically, the rule required that refineries meet an annual average gasoline benzene content

standard of 0.62 vol% on all gasoline produced (down from a typical 1.0 vol% benzene content).

Reformulated gasoline

Reformulated gasoline (RFG) is gasoline blended to burn more cleanly than conventional gasoline in order to reduce smog forming and toxic pollutants in ambient air. The first phase of the EPA RFG program began in 1995 and the second (and current) phase began in 2000. To demonstrate compliance with RFG standards, refiners utilise computer simulations known as the 'complex model' to assess how specific fuel compositions affect emissions. EPA requires the use of RFG in cities with high ground level ozone (i.e. smog) levels. RFG is currently used in 17 states and Washington D.C., and accounts for approximately 30% of the gasoline sold in the US.

Volatility

The EPA regulates the volatility of conventional gasoline sold at retail stations during the summer ozone season (June 1st - September 15th) to reduce evaporative emissions that contribute to smog and thereby lessen the effects of ozone related health problems such as asthma. EPA set volatility standards (measured based on the Reid vapour pressure (RVP) of gasoline) that vary by state (and the severity of its smog problems) and time of year. Phase I was applicable to calendar years 1989 through 1991 and included RVP limits of 10.5 psi, 9.5 psi, or 9.0 psi. Phase II became applicable in 1992 and sets limits of 9.0 psi or 7.8 psi. EPA provides an additional 1.0 psi RVP allowance for gasoline containing ethanol at 9 - 10 vol%, recognising that the use of ethanol has environmental benefits, but poses problems for meeting volatility standards.

Oxygenate

Oxygenates are fuel additives that contain oxygen, usually in the form of alcohol or ether. Oxygenates can enhance complete fuel combustion and thereby reduce exhaust emissions. The EPA requires the use of oxygenated gasoline in areas where winter season (generally October - February/March) carbon monoxide levels exceed federal air quality standards. Without oxygenated gasoline, carbon monoxide emissions from gasoline fueled vehicles tend to increase in cold weather. Historically, refiners made extensive use of methyl tertiary butyl ether (MTBE) as an oxygenate, however due to concerns regarding groundwater contamination, the use of MTBE has been largely discontinued. Ethanol has replaced MTBE as the oxygenate of choice in most cases.

Renewable fuels standards

As concerns regarding climate change and greenhouse gas (GHG) emissions increase, so does pressure to find alternatives to traditional fossil fuels. One approach would require refiners to produce motor fuels, at least partially, from renewable resources via the introduction of Renewable Fuels Standards (RFS). The EPA has not yet finalised any formal RFS requirements; however, states like California are proceeding with state specific requirements. One approach would increase the amount of ethanol (a renewable resource currently produced in the US predominantly from corn) in gasoline blends. Current blends contain 10% ethanol (E10). As an interim measure, a requirement for gasoline to contain 15% ethanol (E15) is being considered.

Process Insight:

Ideal Stage or Mass Transfer... Which Model Should Be Used?

The design and optimization of separation processes is carried out using process simulators, which utilize various calculation approaches. Two techniques that are widely used for modeling distillation are the ideal stage model and the mass transfer model.

IDEAL STAGE MODELS

The ideal stage model requires a minimum amount of data—only equilibrium relationships and enthalpy data for the heat balance. The assumptions of the ideal stage approach are: 1) that the vapor and liquid are both perfectly mixed so that the vapor and liquid leaving a stage are at the same composition as the material on the stage and 2) that thermodynamic equilibrium is obtained on each stage. The equilibrium assumption also means liquid and vapor leaving a stage are at the same temperature. Ideal stage models can also account for non-ideal column performance through the use of reaction kinetics as is done for amine sweetening columns.

Obviously, the main disadvantage of the ideal stage approach is just that—the use of ideal stages to model real trays or packing depths. However, for most processes encountered in gas processing and other industries, the overall efficiencies are well established for proper operating conditions of the column. For systems that are unavailable, similar systems often exist to allow for efficiency estimation. If not, the mass transfer approach is available as an option.

MASS TRANSFER MODELS

For the end user, the notable feature made available via the mass transfer approach is the ability to model a column with the actual number of trays in the unit or the actual depth of packing. However, there are still several assumptions that are made in this approach that can have a significant impact on results. Two that are worth mentioning include the mixing model for trayed columns and the discretization of the packing depth for packed towers.

Application of the mass transfer model to random or structured packing requires the column height to be discretized into vertical segments or stages. For trayed columns, various mixing models can be used for the liquid and vapor phases. The most basic assumption is that of complete mixing in both the liquid and vapor phases. However, the concentration gradients that develop on a tray can significantly impact the predictions made by this model since this gradient is the driving force for mass transfer. As the column diameter becomes larger, the perfectly mixed flow model is less applicable.

For modeling both liquid phase chemical reaction and mass transfer, the use of the enhancement factor technique may be considered. The enhancement factor describes the increased rate of absorption due to the effect of a chemical reaction. The material

balance requires kinetic rate expressions for all chemical reactions occurring in the system. As with equations for a non-reacting system, an appropriate model for interface behavior must be used.

Mass transfer models require data necessary to calculate interphase mass and heat transfer coefficients and interfacial area based on correlations of the following transport and thermal properties: diffusivities, viscosities, densities, heat capacities, thermal conductivities, etc. Furthermore, mass transfer models require detailed information on the column internals. For trays, this includes information such as weir heights and fraction active area. For packing, this includes surface area per unit volume and void fraction.

If the simulator allows the user to select from various alternatives for these parameters, knowing the correct selection may be problematic. Further, the prediction of multicomponent mass transfer coefficients is of questionable accuracy. These facts prompt the recommendation that columns modeled with the mass transfer approach be checked against an ideal stage model with an expected efficiency until sufficient experience with the particular application is achieved.

CONCLUSIONS

When performed properly, both the ideal stage and mass transfer approach as implemented in ProMax 4.0 can calculate accurate results for a variety of separation processes with and without reactions. The ideal stage approach can be used initially to determine appropriate equipment sizes and operating conditions. More detailed studies can be performed using the ideal stage approach, the mass transfer approach, or both. Although significant operating experience provides reasonable efficiency estimates for most processes, the empiricism in scaling up from ideal to real stages or ideal stages to real bed lengths can be a disadvantage when accurate overall efficiencies or HETP's are unavailable.

The mass transfer approach requires more equipment design details and does not make use of overall efficiencies or HETP's. More detailed composition and temperature profiles are produced by this method at the expense of longer calculation time. The mass transfer approach may appear more predictive in nature, but is not necessarily more accurate. It relies on more parameters that must be estimated, as both require thermodynamic data to model equilibrium—for the tray composition in the ideal stage approach and for the interface composition in the mass transfer approach. Many of these mass transfer parameters are of limited accuracy but also may be of limited sensitivity in some systems. Both techniques are useful tools in process simulation.

For more information about this study, see the full article at www.bre.com/support/technical-articles.aspx.

In response to a request by Growth Energy and 54 ethanol manufacturers, the US EPA has granted two partial waivers that allow, but do not require, the introduction into commerce of gasoline that contains greater than 10 vol% ethanol and up to 15 vol% ethanol for use in model year 2001 and newer light duty motor vehicles, subject to certain conditions. However, both refiners and vehicle manufacturers are concerned that increased amounts of ethanol in gasoline may compromise engine life and performance.

And now for Tier 3

Tier 3 low sulfur requirements are the latest in the expanding list of US gasoline standards. Under this proposal, the EPA would require that gasoline contain no more than 10 ppm sulfur on an annual average basis. This limit would become effective January 1st, 2017. In addition, the EPA is proposing to either maintain the current 80 ppm refinery gate and 95 ppm downstream caps or lower them to 50 and 65 ppm, respectively.

US EPA is also proposing an averaging, banking, and trading (ABT) program that would allow refiners and importers to spread out their investments through an early credit program and rely on ongoing nationwide averaging to meet the 10 ppm sulfur standard. The proposal includes a three year delay for small volume refineries processing 75 000 bpd of crude oil or less.

How will refineries meet Tier 3?

If the proposed Tier 3 standards are finalised, US refiners will be faced with yet another, more stringent, gasoline standard. As US EPA's proposed rule is being reviewed, refiners are studying the requirements and evaluating options for modifying their facilities to produce compliant gasoline by the 2017 deadline. In proposing the rule, US EPA has assumed that refineries will be able to comply with Tier 3 standards via projects quite similar, but smaller in scope, to the projects that refineries completed to meet the Tier 2 sulfur standards.²

The common practice for removing sulfur from refinery process streams is the use of hydrodesulfurisation (or hydrotreating) reactors. In a hydrotreating process, hydrogen (H_2) is reacted with the process stream in the presence of a catalyst, forming hydrogen disulfide (H_2S), which is then separated from the process stream and further processed, typically into an elemental sulfur byproduct. Hydrotreating is an energy intensive process, and requires large amounts of high purity H_2 reactant. This hydrogen can be produced as a byproduct of catalytic reforming units (utilised to upgrade gasoline blendstock components), but reformer hydrogen is often not of sufficient purity for direct use in hydrotreating. Instead, refineries typically turn to dedicated hydrogen production units. These hydrogen units often utilise the steam methane reforming (SMR) process whereby methane (from refinery fuel gas or purchased natural gas) is reacted with steam to form high purity H_2 and, as a byproduct, significant amounts of CO_2 .

In many cases, the largest contributor to sulfur in a refinery's gasoline blending pool is naphtha, specifically naphtha produced by the fluidised catalytic cracking (FCC) units. To comply with the Tier 2 standards, many refineries constructed new hydrotreating equipment in order to treat gas oil feed to the FCC and/or to treat the naphtha streams produced by the FCC. Hydrotreating FCC feed can entail significant expense, but can improve the overall performance of the FCC. To meet the Tier 3 standards, the

EPA expects that refineries will need to either install new hydrotreaters or modify/expand existing hydrotreaters (e.g., increasing hydrotreating severity, adding additional reactors). In some cases, treating FCC naphtha may not be sufficient to meet the overall Tier 3 standards. Additional gasoline streams that may require further desulfurisation include light straight run naphtha (from crude units), natural gas liquids, and butane. A refinery's current level of hydrotreating as well as the sulfur content and density of current and anticipated future crude slates may dictate which approach is most economical for meeting these standards.

Impact to industry

The EPA claims, based on modelling conducted as part of rule development, that 29 refineries would not need to make any capital changes, 66 would need to revamp their existing FCC naphtha hydrotreaters, and 16 would need to construct new hydrotreaters (either for FCC naphtha or FCC feed).³ The EPA believes refiners will need two years or less to revamp existing FCC hydrotreaters and that new hydrotreaters will require up to three years to install and start up⁴. With that anticipated schedule, it believes that US refiners can meet the proposed 2017 deadline.

The refining industry, however, has expressed significant concerns regarding this deadline. Consideration must be given to the time required for engineering and design, lead time for ordering equipment, and the time required to construct, 'shake down', and prepare to operate the equipment. Implicit in this timeline is the required construction permitting to allow for the increase in air emissions from the new and modified equipment. Often this permitting is the critical path influencing a capital project timeline. The EPA estimates that only 19 refineries would need permits for projects that would be considered 'major modifications' under the requirements of the Clean Air Act. Even for these larger projects, it asserts that the permits can be issued in 9 - 12 months, allowing for the 2017 deadline to be met. Past experience indicates, however, that permits for these major modifications often take closer to 18 - 36 months, raising serious concern about the 2017 deadline.

The refining industry has petitioned for at least five years between the promulgation of a Tier 3 rule and its effective date, claiming that refiners need this much lead time to schedule equipment modifications or install new equipment around predetermined turnaround maintenance schedules. Barring this allowance, refineries may be forced to perform unplanned shutdowns which could create fuel supply shortages and price increases. The industry notes that these Tier 3 standards are not specifically required by the Clean Air Act and therefore, the selection of a 2017 deadline is arbitrary. The industry proposes, instead, a 2019 deadline.⁵

The industry also disputes the magnitude of the environmental benefit resulting from the incremental reduction in gasoline sulfur from Tier 3 standards, particularly given that it comes at the price of significant emissions increases from the refining industry. Costing as much to implement as the Tier 2 standards, refiners believe the benefit of Tier 3 standards will be far smaller. In terms of ambient ozone levels, this reduction is estimated to be no more than 0.5 ppb by 2022 compared to a 12 ppb reduction achieved from Tier 2 (or one twentieth of the reduction). Basically, Tier 2 will help reduce ozone emissions 10% below 2008 levels by 2022. Moving to Tier III would only achieve another 0.7% reduction. This incremental reduction will be at the



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expense of a 1 - 2.3% increase in GHG emissions from refineries, due to the installation of energy intensive hydrotreating equipment and the resulting increased hydrogen production demand.⁶

There is also disagreement over the expected cost to industry. EPA estimates that compliance with Tier 3 standards will cost the refining industry approximately US\$ 2.1 billion, compared to an estimate cost of US\$ 6.1 billion for the Tier 2 standards.⁷ Refiners believe that the cost will be closer to US\$ 10 billion in capital costs and US\$ 2.4 billion/y in operating costs, increasing the cost of producing gasoline by up to US\$ 0.09/gal.⁸ In a time of a slowly recovering economy, the auspice of higher fuel costs is not being well received.

The future

Given the controversy over the Tier 3 proposal, EPA extended the public comment period until July 1st, 2013 to allow for additional debate and exchange of information. The industry, however, is not optimistic that EPA will make significant changes to the rule requirements or timeline.

Once the Tier 3 dust has settled, other new fuel standards loom on the horizon. The EPA is evaluating further volatility reductions as well as formalised renewable fuels standards. Refiners are concerned that these rules may pose even greater challenges, estimating that further RVP reductions could raise

gasoline prices by US\$ 0.25/gal and that renewable fuels standards could raise gasoline costs by up to 30% in the next few years.⁹ Despite the debate over the efficacy and prudence of additional fuel standards, all sides agree that it will prove to be an ongoing challenge to balance the environmental impact of motor vehicle emissions with the cost and environmental impact of modifying refineries to meet these increasingly stringent fuel standards. 

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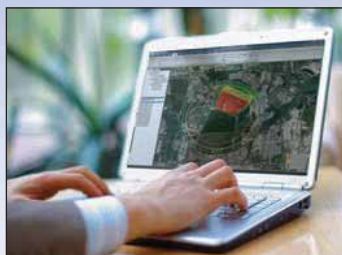
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HEAT RECOVERY



KIRK S. WESSELOWSKI, ZEECO, USA, DISCUSSES THE PROPER APPLICATION OF PREHEATERS AND OIL HEATERS FOR THERMAL OXIDISERS IN GAS PROCESSING FACILITIES.

In order for gas plant operators to keep up with stringent environmental emissions regulations, operators are using thermal oxidisers to maintain environmental compliance as well as maximise plant efficiency through heat recovery. The majority of gas plants utilise regenerative chemical solvents in an absorption column to remove CO₂ and H₂S from the process gas. These solvents are typically regenerated in a stripper with a reboiler, a process that requires an external source of heat. The off gas from the regenerator must often be burned in a thermal oxidiser to ensure environmental compliance, so recovering heat from the thermal oxidiser flue gas with a convection oil heater and using it in the reboilers is a beneficial approach for many applications.

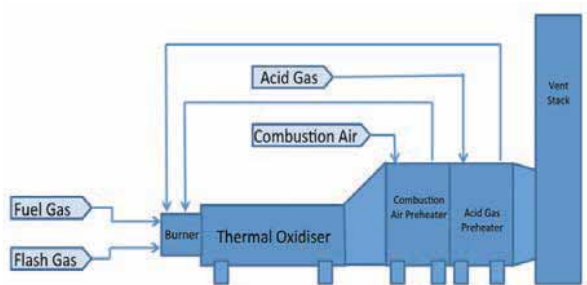


Figure 1. Acid gas thermal oxidiser system with preheaters.

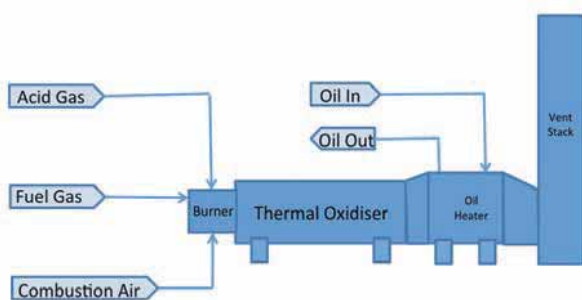


Figure 2. Acid gas thermal oxidiser system with hot oil heater.

However, in some plants sufficient heat for the reboilers and other consumers of process heating is available from other sources. In these cases, using the thermal oxidiser flue gas to preheat the streams fed to the unit can be an attractive design option. This article will illustrate the types of applications that can benefit from preheaters and oil heaters, and will give some advantages and potential pitfalls for each. With these insights, thermal oxidiser waste heat recovery systems in gas processing plants can be integrated into the plant design, significantly reducing operating costs while maintaining excellent reliability and long operating lifetimes.

Thermal oxidisers

The purpose of a thermal oxidiser is simply to destroy combustibles in a waste stream that would be harmful to the environment and/or health and safety of people near the source of the waste stream. To completely oxidise the waste stream, it must be heated to a sufficient temperature, and held at that temperature for an appropriate time in an oxygen rich environment. By controlling the flow paths within the thermal oxidiser so that every part of the waste stream is exposed to these conditions, experienced combustion engineers can ensure that the combustibles will be completely oxidised to their relatively benign combustion products.

In a typical gas processing plant, the inlet gas may be 'sweetened' by removing the H_2S and CO_2 using an amine unit, and/or dehydrated using a glycol contactor. The amine solution and the glycol solution each require a regenerator (a stripper with a reboiler) so they can be reused. These reboilers require heat input, usually by using heat transfer oil or

sometimes steam. The vent stream generated by the amine (the 'acid gas') and the glycol reboiler off gas can each contain hydrogen sulfide, BTEX (benzene, toluene, ethylbenzene, and xylene), and hydrocarbons, which must all be oxidised in order to maintain environmental compliance for the plant. While flash gases and glycol reboiler off gasses often have significant hydrocarbon content, the acid gas is generally more than 95% CO_2 and water vapour, and requires significant external energy input in order for the low levels of combustibles in the acid gas to burn.

Another way to remove CO_2 from the processed gas is by using a permeable membrane separator. It may be difficult to achieve the required efficiency with a permeable membrane separator alone, so they are often used in conjunction with an amine unit. The membrane is more permeable to CO_2 than methane, leaving most of the methane in the residue gas. However, the membrane is not a perfect separator, so the permeate gas contains a significant amount of hydrocarbons, in addition to the CO_2 . This makes the permeate gas a good source of energy, provided it can be burned effectively with a waste heat unit in place to recover the energy from the flue gas. The permeate gas is generally burned in a thermal oxidiser to meet the required emissions levels and to recover that energy.

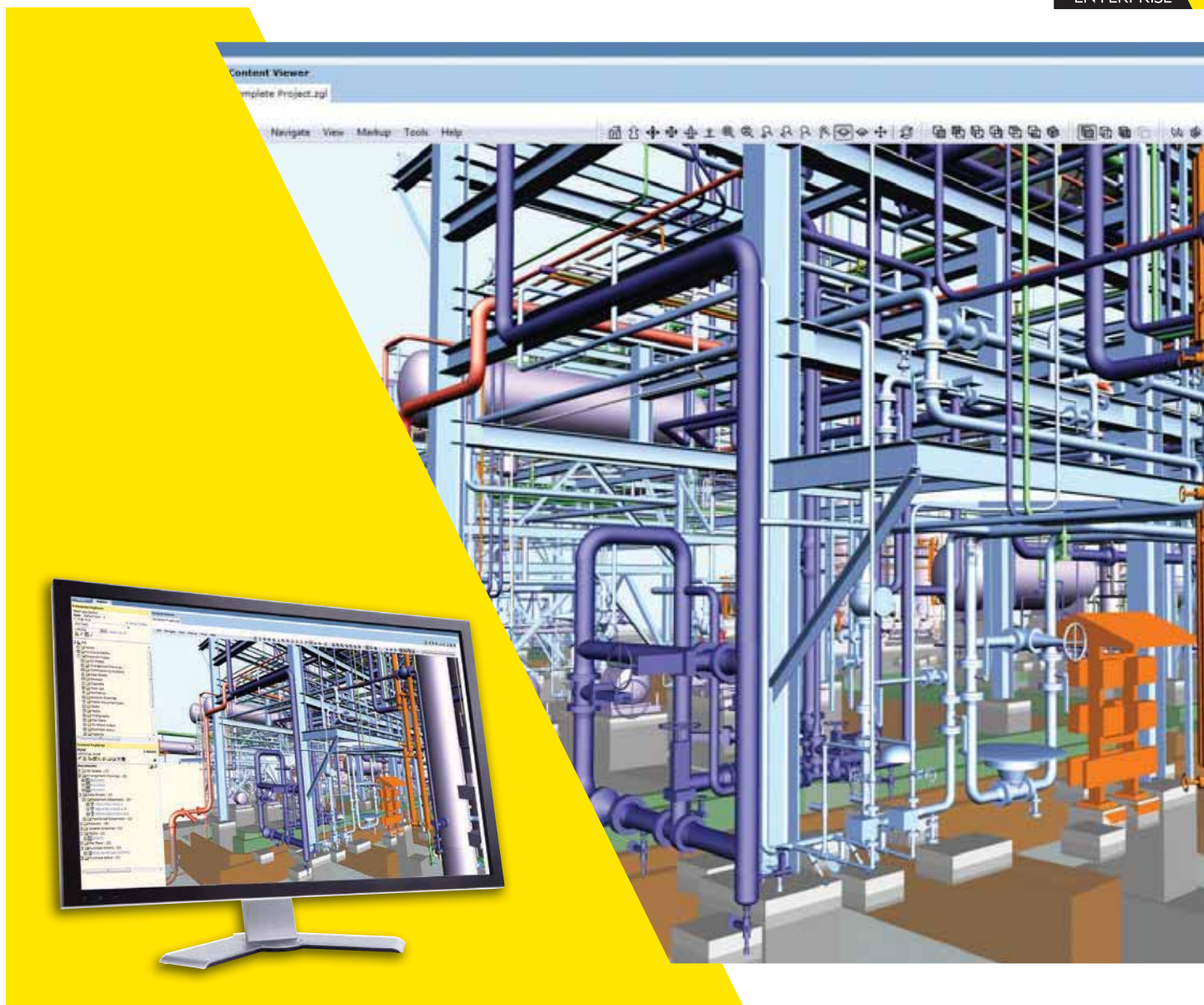
When considering options for waste heat recovery from a thermal oxidiser, system designers must consider how much of the combustion energy will originate in the waste gas and how much will originate in any supplemental fuel that may be required. In thermal oxidiser design there is an important distinction between a waste gas that provides a net positive release of heat when burned at the required combustion temperature (an 'exothermic' waste) and one that requires external energy input in order to burn at the required temperature (an 'endothermic' waste).

In a gas processing plant, the amine acid gas is typically an endothermic waste gas, meaning that the thermal oxidiser requires additional fuel gas in order to achieve the combustion temperature. On the other hand, exothermic streams such as flash gases and permeate gases have sufficient hydrocarbon content such that they release more than enough energy when burned to achieve the required operating temperature.

Heat recovery

There are two ways to approach improving the overall thermal efficiency of the plant. The first is by preheating the waste gas and/or combustion air to directly reduce fuel consumption of the thermal oxidiser. The second is by capturing heat generated by the destruction of waste gases in the thermal oxidiser that can then be used elsewhere in the plant. The principle is to take heat out of the thermal oxidiser flue gas and put it back into the process where it is needed. In the case of an endothermic waste stream, the flue gas heat can be used in a preheater, or captured by a steam boiler or oil heater. For an exothermic waste, there is no benefit to preheating the streams, so a waste heat steam boiler or oil heater would typically be used.

Preheaters can be used in conjunction with hot oil heaters if a gas processing plant generates an endothermic acid gas stream but only a fraction of the enthalpy in the flue gas



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Figure 3. A thermal oxidiser with acid gas and combustion air preheaters.

needs to be captured for use elsewhere in the plant. In this manner, the thermal oxidiser heat recovery system can be tailored to reduce the overall plant fuel consumption by decreasing the thermal oxidiser fuel consumption while reducing (or perhaps eliminating) fuel consumption in a direct fired heater.

Acid gas thermal oxidisers

A typical system will incorporate both a combustion air preheater and a waste gas preheater, as shown in Figure 1.

In previous decades, such a configuration was avoided because of the difficulty in designing gas-gas heat exchangers that would stand up to the extreme environment. The thermal oxidiser flue gas temperatures will typically be 800 °C - 950 °C. In order to maximise thermal efficiency, acid gas and combustion air preheater outlet temperatures often exceed 450 °C, resulting in heat transfer surfaces that can often surpass 600 °C, with the possibility of much higher temperatures during turndown conditions. While many metals have been proven sufficient at these temperatures, early iterations of waste gas preheaters were prone to failure because of uneven distribution of flue gas in the preheaters, especially at turndown flow conditions, and because of differential expansion between components of the preheaters.

These problems can be avoided by good engineering design of the thermal oxidiser system in general and preheaters in particular. Working with an experienced combustion company such as Zeeco can eliminate these concerns. Some specific design considerations include:

Co current preheater design

Arranging the inlet cold combustion air or acid gas stream at the same end of the preheater as the inlet hot flue gas will reduce the maximum heat transfer surface temperature when compared to a counter current design. Although this results in a somewhat larger heat transfer area for a given duty, the increase in reliability and equipment lifetime makes this a beneficial trade off.

Preheater arrangement

If both a combustion air preheater and an acid gas preheater are used, then the combustion air preheater is typically situated upstream of the acid gas preheater. Since combustion air will be flowing any time the thermal oxidiser is operating, the heat transfer surfaces will always be cooled by the combustion air, and surface temperatures will be less than the inlet flue gas temperature. However, the thermal oxidiser must operate without flowing acid gas at least part of the time, typically during the plant's startup. If acid gas is not flowing, then the heat transfer surfaces in the acid gas preheater will be at the flue gas temperatures, since there is almost no heat being removed by the cold side of the heat transfer surface. Therefore, the acid gas preheater is situated downstream of the combustion air preheater. The flue gases exiting the combustion air heater will be cooler than the thermal oxidiser operating temperature, mitigating the severity of the no acid gas flow condition.

Heat transfer surface design

Preheaters must be carefully designed to allow for differential expansion of elements comprising the preheaters. Differences in thermal expansion can generate strong forces that will rip apart a heat exchanger that does not account for thermal growth. Tubular surfaces are preferred over plate heat exchangers because the design is typically more robust. Appropriate tube expansion bellows, prebent tubes, or U-tube designs will allow each row of tubes to grow individually, preventing large expansion forces from developing.

Thermal oxidisers with oil heaters

If the waste heat captured from the thermal oxidiser flue gas can be used elsewhere in the plant, then a thermal oxidiser with an oil heater design may be a good choice. A typical arrangement of an acid gas thermal oxidiser with an oil heater is shown in Figure 2.

The oil heater is placed at the discharge of the thermal oxidiser, allowing a limited amount of control of the oil heater duty in this system. The thermal oxidiser operates at a specified furnace temperature in order to ensure good destruction efficiency, so the heat transfer rate in the oil heater is determined largely by the mass of flue gas. The mass of flue gas from the oxidiser is predominantly dictated by the acid gas rate, since sufficient air and fuel for combustion must be injected into the thermal oxidiser for a given acid gas flow rate. It is not possible to reduce the oil flow rate arbitrarily, since sufficient flow must be maintained in order to keep the oil below the maximum allowable film temperature specified by the heat transfer oil manufacturer. Consequently, for a given acid gas rate it will not be possible to reduce the oil heater duty below a specified minimum amount. The heat transfer oil system must have sufficient heat consumers, such as reboilers or air coolers, to remove the specified minimum amount of heat generated by the oil heater. On the other hand, the thermal oxidiser and oil heater system can be designed to fire additional fuel gas and combustion/quench air to enable higher flue gas flows than would be required simply to burn the acid gas. This overfiring of the thermal oxidiser will give the operators the flexibility to increase the oil heater duty when required.



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Figure 4. A permeate and acid gas thermal oxidiser with a hot oil heater.

Case studies

Case 1: 80 million ft³/d gas processing plant

A gas processing facility processes gas from an offshore facility and produces condensate, liquefied petroleum gas (LPG), and sales gas. The amine unit, which removes CO₂ and H₂S from the natural gas, produces two waste gas streams, a large flow of endothermic acid gas stream that is 90% CO₂, and a smaller flow of exothermic flash gas. An ethylene glycol (MEG) dehydrator removes water from the natural gas, and the MEG regenerator off gas is a high BTEX stream that is exothermic at the thermal oxidiser operating conditions. The largest flow by mass is the acid gas stream, which requires a significant amount of energy to reach the thermal oxidiser operating temperature. The energy supplied by the flash gas and MEG off gas is not enough to eliminate fuel gas consumption at the thermal oxidiser operating temperature of 850 °C. Therefore, the thermal oxidiser shown in Figure 3 incorporates a combustion air preheater and acid gas preheater in order to reduce the fuel consumption by more than 75% during normal operation. The payout period for incorporating combustion air and acid gas preheaters into the design for this case and similar cases is generally less than two years.

By incorporating quench flow systems into the process design, the system is configured to handle a wide range of waste gas flow rates and compositions. The system can handle fully exothermic conditions when the flash gas and MEG off gas flows are high and the acid gas flow rate is low, as well as very endothermic conditions at the maximum acid gas rate and minimum flash gas rate. This is accomplished without using dampers or other moving parts in the hot flue gas stream, maximising reliability and reducing maintenance costs for long term operation.

Case 2 : 100 million ft³/d gas processing plant

This gas processing plant reduces the CO₂ in gas produced by several fields to less than 5% by volume, and maximises recovery of natural gas liquids. It uses both a permeable


membrane separator and an amine unit to separate carbon dioxide from the inlet gas. Since the membrane is not a perfect separator, the permeate gas contains a significant amount of hydrocarbons in addition to the CO₂.

Although the permeate gas burned in the thermal oxidiser is 89% CO₂, it contains 9% methane, resulting in a lower heating value of approximately 21 500 kcal/kmol. This places the permeate gas in an intermediate range of heating values, much lower than methane at 191 700 kcal/kmol, but still a significant source of energy. The permeate gas lower heating value is too low for it to sustain combustion on its own, requiring a specialised thermal oxidiser design. This design uses assist fuel and staged air injection in order to enable the permeate stream to realise its full heat release with a minimum of assist fuel gas. The acid gas contains 91% CO₂ and 8% water vapour, resulting in a lower heating value of approximately 3000 kcal/kmol, so it is endothermic at the thermal oxidiser operating temperature of 815 °C.

When burned in the thermal oxidiser, the permeate gas releases 75 million kcal/hr and the acid gas achieves a heat release of 5 million kcal/hr. The total duty of the amine stripper reboiler and other preheaters and reboilers in the plant is 41 million kcal/hr. The thermal oxidiser has a hot oil waste heat recovery unit sized for 45 million kcal/hr, making it sufficient to provide all the heat necessary in the gas processing plant. The flue gas stack temperature is sufficiently above the sulfur dew point of the flue gas to avoid any corrosion problems.

The thermal oxidiser system is shown in Figure 4. Three fuel gas burners provide fuel gas as required to initiate and stabilise combustion, depending on the operating case. The thermal oxidiser is a horizontal cylindrical refractory lined vessel. Primary, secondary, and tertiary injectors add air into the thermal oxidiser at staging intervals that ensure proper control of the combustion process and complete combustion of the permeate and acid gas streams. Heat is recovered from the flue gas in the hot oil heater tubes before flue gas is discharged to the atmosphere via the vent stack.

Conclusion

Thermal oxidisers play an important role in emissions control for gas processing plants. Direct oxidation of the waste gas streams is a simple and effective way to meet emissions requirements, but it often comes at the expense of fuel consumption in the form of assist gas. By capturing a portion of the heat energy used for waste gas destruction and using this to meet some, or all solvent regeneration needs, processing plants can enhance their efficiency and minimise their overall operating costs. If this energy is not needed elsewhere in the plant, then a waste gas and/or combustion air preheaters can be used to reduce the fuel consumption of the thermal oxidiser, thereby reducing the plant's operating costs. Through assessment of the total duty requirements of the plant, and proper understanding of the requirements of the thermal oxidiser process, heat recovery can be implemented in a thermal oxidiser system to positively impact plant economics, operational flexibility, and the environment. 



Triple play, triple profits

Douglas C. White, Emerson Process Management, USA, discusses how refining companies can make a 'triple play' on their investment; modernising plant operations in order to simultaneously reduce energy use, reduce carbon emissions, and increase profitability.

Refining is an energy intensive industry responsible for over 15% of the total world industrial energy usage. The average primary energy usage in refineries ranges from 300 000 - 600 000 BTUs/bbl (0.265 - 0.530 GJ/m³) with the usage highly dependent on the complexity of the refinery, the quality of the crude feed, and the average age of the equipment, i.e. newer refineries typically have less energy usage than older ones of the same complexity level with similar crudes. For most refineries, energy is the second largest cost after crude purchase, comprising 40 - 60% of the non-crude costs. A typical 200 000 bpd (10 million tpy) refinery may spend US\$ 100 - 400 million/y on energy depending on the refinery efficiency and market pricing for incremental fuel. An energy efficient smart refinery may be able to save a significant portion of these costs.

Due to limited functionality and higher cost, older measurement and automation systems were typically applied only to control critical plant assets. Minimal measurements were installed to allow the plant to operate safely, but they were not designed for optimal

energy efficiency and productivity. The legacy systems provided basic controls for critical assets but use of advanced automation functionality was limited as it was complex and expensive.

Today, there remain many opportunities for saving energy in refining. In fact, the theoretical minimum energy usage for refineries is approximately 25% of current average energy usage and potential savings from application of best available technology¹ (with no consideration of cost or financial payback requirements) could be 50% of the energy usage. The financial savings may be even greater if carbon trading incentives are factored in. Almost every refining company's annual report lists reduction in energy usage as one of the top corporate priorities.

Energy use and greenhouse gas emissions

Regulatory reductions in greenhouse gas emissions (or reduced carbon footprint) are being mandated in many regions of the

world and reducing energy usage is an attractive way to address these regulations.

The most significant refinery GHG emissions come from the combustion of fossil fuels for energy. Every t of carbon based fuel saved avoids generating more than 3.5 t of CO₂ emissions. Therefore, management of GHGs is best achieved through management of energy usage.

The changing cost of fuel

Since natural gas is normally the fuel purchased to supplement the internally generated refinery fuel gas (RFG), natural gas pricing has a significant impact on the return on investment for energy saving projects. There has been substantial volatility in natural gas pricing around the world over the last few years which has complicated evaluation of potential investments in this area. At US\$ 10/million BTU natural gas pricing, a 5% saving is worth US\$ 0.2/bbl or US\$ 14 million/y for a typical 200 000 bpd (10 million tpy) refinery.

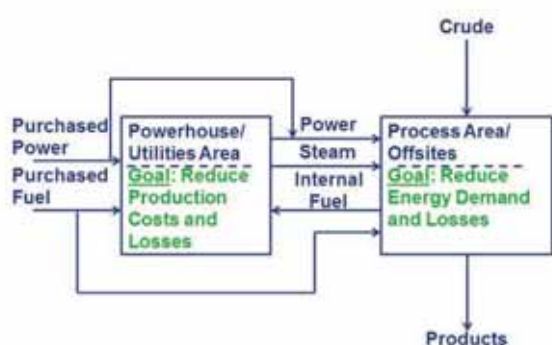


Figure 1. Refinery energy use includes two major areas.

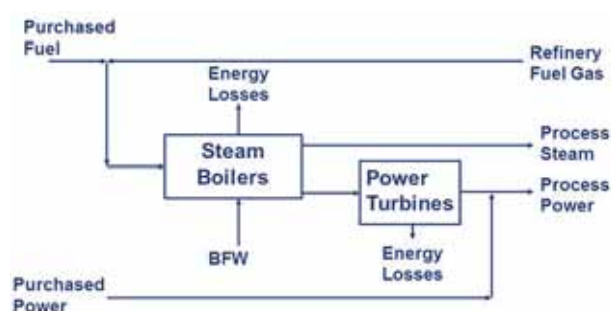


Figure 2. Refinery utilities overview.

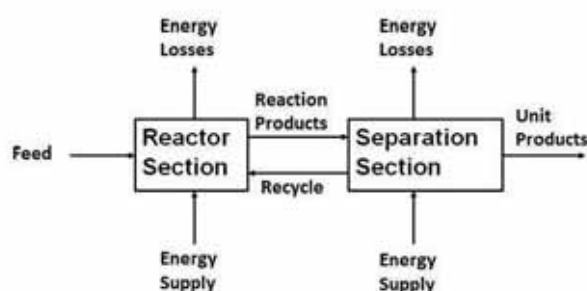


Figure 3. Typical refinery process unit.

Emergence of smart refineries

A key factor that has enabled the emergence of the smart refinery^{2,3,4,5,6} is the phenomenal advances in computer electronics, communications and software technologies over the past 20 plus years. These advances are best illustrated by the evolution of cell phone technology from the 1990s to the 'smart phones' available today. Today's ultra compact and powerful 4G smart phones offer personalised 'apps' at an affordable cost. In like fashion, smart applications are now available to make refineries much easier to operate and maintain at an affordable cost.

The foundation of a smart plant is built on smart field devices such as multivariable transmitters, valves with intelligent positioners, and pumps with built in diagnostics. These devices are configured to measure variables at a high frequency, perform control actions based on the measured data, and analyse data over a historical time period to predict a future operating state of the plant asset. Next, the smart plant is based on smart communications networks which enable these smart field devices to communicate with one another. The smart networks may be designed to automatically reconfigure to establish alternative communication paths in case of a failure or multiple failures on a path.

Finally, the smart plant architecture offers decision support tools, solutions, and interfaces (referred to as smart plant 'apps' or functional modules). These 'apps' are preengineered solutions that integrate measurement technology, process knowhow, and intricate knowledge of the plant asset to enable a user to make 'smart' and 'proactive' decisions based on predictive technology. The 'apps' may be customised for the plant personnel (e.g. for the operator, engineer, maintenance, quality control, or business office) to easily analyse plant assets and refinery production systems.

Saving on energy costs

As shown in Figure 1, energy use in the refinery may be broadly segmented into two areas. The first is the powerhouse/utility area which consumes purchased and internally generated fuel to produce power and steam. The second is the process area and offsites which consumes the utilities and fuel to convert crude oil into higher value products such as refined gasoline and diesel. During the conversion process, the process area produces refinery fuel gas (RFG) and coke as byproducts. RFG and coke byproducts may be used as fuel in the powerhouse/utility area or the process area itself or sold.

Since the energy demands from the process area and offsites are continually varying based on consumer demand, quality of the crude feed, and other external factors, the powerhouse/utility area is required to proactively and predictively produce energy to meet the dynamically changing demand at minimum cost while minimising losses. The powerhouse/utility area can increase energy efficiency by reducing externally purchased power and fuel, increasing use of internally generated fuel, and minimising losses in power and steam generation and distribution. The objective of the process area and offsites is to minimise energy demand and losses while meeting targets for production rate and product quality.

Extensive monitoring of process variables in a smart refinery is typically essential but not sufficient. That is, in addition to the expanded monitoring of energy related measurements, the

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Figure 4. Steam trap acoustic monitoring.

measurement and automation system in a smart refinery must also provide the infrastructure (e.g. communication networks) and application software ('apps') for seamlessly sharing the measurement data, performing data analysis of the collected data, predicting occurrence of future events, and issuing alerts and displays via a real time refinery operations dashboard.

A 'systems' approach to reduced energy usage within a refinery starts at the device/plant asset level, by continuously optimising performance of the assets and predicting occurrence of events that may degrade their normal operation. Next, the systems approach focuses on the major physical equipment itself, by optimising control and energy use strategies for larger process units such as boilers, steam headers, burner systems, fired heaters, distillation trains, and turbines/generators. Finally, the systems approach focuses on developing control strategies to reduce energy use for each process area.

Energy savings strategies in the real world may vary from low cost programs such as reducing steam losses to capital intensive, complex investment programs such as installing combined heat and power (CHP) plants. This article will focus primarily on achieving energy savings in a smart refinery through use of energy saving 'apps' at the device/plant asset level, at the process unit level, and at a refinery wide level.

Saving in the plant utilities area

Figure 2 gives a very simplified schematic of the utilities section. There may be multiple boilers producing steam, some steam possibly directed to power turbines to generate power, and the rest sent to the process area for heating. The percent of refinery power usage produced locally varies widely. Some refineries purchase all the power from the grid, while others produce 100% locally. In addition, many refineries have combined heat and power (CHP) or cogen facilities where gas turbines produce power directly with large amounts of steam produced from exhaust gas heat recovery. Some CHP facilities may sell excess power to the commercial grid.

The smart refinery saves energy in the utilities area by:

- Operating individual boilers at target efficiencies at their current load.
- Maximising use of low cost fuel when available.
- Coordinating multiple boilers to meet the current and future predicted process steam demand at minimum cost.
- Controlling the steam headers and the equipment connected to the headers to minimise unnecessary letdown steam while simultaneously meeting demand.
- Meeting power demand at minimum cost (or maximising profit if surplus power is available and can be exported).

The referenced strategies for improving area wide energy efficiencies may be implemented using apps based on multivariable, model predictive control working in conjunction with smart field instrumentation. These apps are most useful to continuously control highly interactive multivariable processes that operate under constraints or exhibit long delay and lag times. For example, a boiler app may be used for controlling multiple boilers having varying loads and varying fuel compositions.

Saving energy in the process area and offsites

Many of the process units have the structure shown in simplified form in Figure 3. The process unit includes a reaction section where the molecular weight and molecular structure of the feed is modified. The reaction products mixture is then processed in a separation section to produce individual products. Energy in the form of heat is provided to the reaction section for endothermic reactions and often for vapourisation of the feed. Separation section energy is used to produce the final product. It is estimated that close to 90% of onsite fuel use in refining is applied toward process energy requirements with 65% direct firing and an additional 23% consumed in the generation of steam used.

As discussed in Reference 4, the largest net energy users in the refinery are the crude and vacuum units that are just separation units without a reaction section. Next are the 'reaction' or 'conversion' processes that include cracking of high molecular weight hydrocarbon molecules into lower MW molecules. Examples include process units such as the fluid catalytic cracking (FCC), hydrocrackers/hydrotreaters, and naphtha reformers. As stated previously, the objective for the process section is to use minimum energy while meeting production rate and quality targets and ensuring reliable operation.

The smart refinery saves energy in the process area by:

- Minimising energy oversupply.
- Reducing losses.
- Maximising heat recovery.
- Optimising/minimising recycle.

Similar to the utility area, the referenced strategies for improving process area energy efficiencies may be implemented using 'apps' based on multivariable, model predictive control and use of smart measurements. For example, a distillation column app may be used for controlling both overhead and bottoms product purity, minimising energy use, and maximising feed, while operating the column within equipment constraints. An embedded neural block in the app may be used to predict product compositions.

Monitoring energy use for essential plant assets

Approximately 40% of the energy usage in refineries is steam.⁶ Therefore, its efficient use is a key to reducing overall energy savings. Steam is difficult to manage because use and individual equipment controls are distributed widely around the refinery as are the losses.

One of the essential plant assets that may be continuously monitored for performance and energy savings is steam traps. Steam traps are typically valves used for controlling steam distribution that trap or filter out condensate (e.g. condensed steam) and non-condensable gases (e.g. air) while allowing steam to pass through. The condensate can be recovered in the



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condensate system and recirculated to the boiler as boiler feed water (BFW), thus recovering some heat and reducing the need for fresh feed water and its treatment.

Traps can fail in two modes, a fail open mode in which steam is lost to the atmosphere, or a fail shut mode that can be a safety hazard if undetected. There may be 5000 to 10 000 steam traps in a medium size refinery. Losses from any individual trap may be small but the cumulative effect of small losses may be quite large.

An example of new smart refinery technology is a real time app for acoustic monitoring of a steam trap to detect failures as shown in Figure 4. This technology can be installed with a clamp on a pipe, and no cutting or welding required.

It has been a challenge previously to determine actual losses from traps due to their wide geographic distribution and relatively small scale. Studies have shown that steam losses from traps are proportional to the frequency of checking. Many refineries are on a five year inspection cycle where only 20% of the traps are checked every year. Measurement and accounting of steam losses may be significantly improved by real time acoustic monitoring, particularly for high pressure traps. The expected payback on the investment is typically less than one year, often just a few months.

Barriers to implementing energy efficiency

Of course, reducing energy consumption has been a goal of refineries for many years. Why has there not been more progress? While implementing many energy studies and projects some common barriers have been identified.

One typical obstacle is a lack of detailed knowledge of the actual energy usage of process units and major equipment and whether or not the current usage is higher than expected. Plants in the past were commonly designed and constructed with the minimum measurements required for safe operation and stable control. Energy was cheap and extra measurements to monitor performance were not justified. In addition, normal operating variability in feed rate, feed quality and product target specifications can make it difficult to compare usage from one time period to another. Fortunately these issues can be overcome. Missing measurements can be installed quickly, at relatively low cost. Wireless communication for transmitters avoids wiring costs and the complexity of installing wiring in an operating plant. Models of energy usage are available to convert the current usage to expected usage at standard conditions which is required for effective analysis.


In addition, saving energy generally requires that customary operating practices be changed. Operational buffers against disturbances will be reduced, for example the current operation of a distillation column at a much higher reflux ratio than necessary. More action is then required to change the plant conditions when disturbances do occur and this action should preferentially be automated through the use of enhanced control algorithms. One observation is that operator training for these new procedures and algorithms is often inadequate and not sustained over the long term. This issue can be overcome but requires planning and commitment. Poor dynamic performance of the regulatory control loops is sometimes a limit on reducing the operational buffers. Again, this can be overcome with systematic performance analysis and correction of the underlying causes.

After the potential projects have been identified, obtaining energy saving investment approvals can be challenging. Although energy is a significant part of the operating costs, it is not always given a high enough priority as a business driver, even when the calculated return on investment is high. Projects that increase capacity or yield may take precedence. Part of the issue is that there is risk and uncertainty in fuel costs. Wide fluctuations in natural gas costs have occurred in the recent past making forecasting future costs uncertain. In addition, it can be difficult to measure results. Measuring and validating energy savings with normal plant variability is harder than documenting production increases, sometimes reducing management confidence in projections. Moreover, purchasing energy efficient equipment may cost more than standard equipment, creating a challenge for the manager involved in the justification.

How to overcome these barriers

Many companies are challenged due to budgeting procedures that discourage assigning funds to energy savings projects. The first step, as discussed in Reference 6, is to develop a plan that prioritises projects. Energy savings projects that may provide high return with little or no investment, and quick payback are implemented first. Examples include proper tuning of all regulatory control loops, training to change internal operating procedures to make employees more accountable for energy use along with production, acoustic monitoring of the steam traps, and others. This can generate incremental revenue to fund additional projects. Often the initial step is to add more measurements so the specific process asset and its energy usage can be calculated and brought to standard conditions. The goal is then to implement real time energy dashboards for refinery wide monitoring and individual process areas that allows drill down selection to the unit level as well as to individual equipment level. The smart refinery technology enables a change in energy management from a reactive to a proactive and predictive mode.

Conclusion

The smart refinery's technologies can yield substantive reductions in energy consumptions and emissions at very attractive returns on investment. These technologies are affordable and often self funding after the initial investment has been made and equipment is operational. However, technology alone is never sufficient. Sustaining energy savings requires a disciplined plan, training in the use of new technologies, and changes in plant operational and maintenance procedures to make the triple play a reality. 

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LIMITING LOSSES

The upward trend in crude oil prices makes it a valuable commodity, hence small changes in price can affect refinery margins considerably. Being in such a delicate profit balance, the control of losses through good hydrocarbon management seems to be vital for a refinery operator, assisting in the reduction of costs and the limitation of environmental effects.

The information and notes in this article are mainly derived from some common cases experienced during industrial loss prevention programs developed in Turkish Petroleum Refineries Corporation (TUPRAS) refineries and specific on site trainings for loss management between 2006 - 2012.

The classical refinery material balance is key to loss accounting and can be summarised as follows:

Refinery loss (or gain) = Receipts + opening stocks – deliveries - closing stocks - internal consumption

Although the fact is not illustrated in refinery balance, balancing actually has two items to be taken into account: physical (or real) and unidentified losses. Real losses can be measured, or at least estimated, with some accuracy. The higher the measured or estimated part of loss, the less is the unidentified part and it shows the effectiveness of instrumentation and a well defined accounting system.

Physical (real) losses

Flaring

A refinery's flaring is the largest single source of physical loss, although it is intended to be an integral part of refinery safety. It provides an easy option for operations when there is a need for burning excess gases or depressurising units. But during normal operations, flare may be a leak source if it is not managed effectively.

Osman Kubilay Karan, Turkish Petroleum Refineries Corporation (TUPRAS), discusses the importance of good hydrocarbon management in the face of upward trending crude oil prices.

Possible sources of flaring :

Refinery fuel gas system set pressure

Contributes to flaring, especially during summer season night/day variation of temperatures, by opening of pressure control valve (PCV) due to fluctuations in pressure. A close observation of PCV set pressures can prevent releasing excess refinery fuel gases to flare.

Tankage area

Tank breathing and leakage of pressure safety valves (PSV) and pressure relief valves (PRV).

Process units

Refinery process plants contain many PSVs/PRVs that are used for process safety. A comparison of their set points against operational pressures that they resist may give many clues to the likelihood of opening PSV/PRV. In general, the difference between the set pressure and operation pressure is accepted to be wide enough when operation pressure is 5% less than the set pressure (Figure 2).

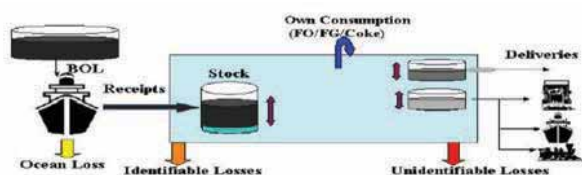


Figure 1. Refinery mass balance.

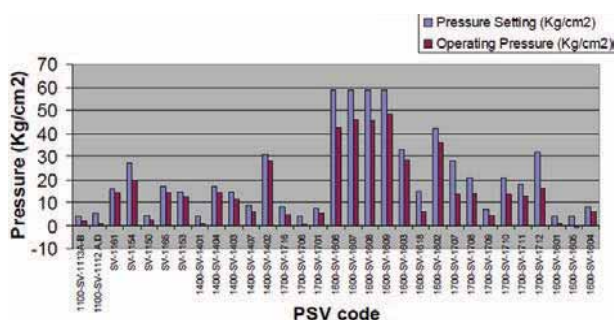


Figure 2. Comparison of setting versus operation pressure.

Some specific actions can be made to prevent/identify flare sources:

Frequent monitoring of PCV, PRV and PSVs

These checks can be made by a routine program with devices detecting the sounds resulting from gas released to flare. Once a leaking safety or control valve is located, its root cause has to be analysed (i.e. process fluctuation, process design limitation, equipment limits/malfunctioning etc). All of these data will lead one to classify flaring events as routine/non-routine to formulate the solution. An engineering approach for leaking PCVs to flare is to monitor their openings with a data historian system like RIMS or PHD. A map of PCVs to flare can be placed to an excel sheet and PCVs that should not be opened can be followed throughout the day on the refinery's intranet.

Process parameters adjustment and maintenance

After the leaking PCV, PRV and PSV's can be identified, some corrective actions should be taken like lowering pressure (process or equipment permits) or maintenance of the leaking valve depending on situation, tuning of process control parameters with equipment reliability improvement and procedure improvement. Among typical maintenance solutions, it would be proper to carry out maintenance and cleaning of air coolers in front of the drums; their control valves open to flare during periods of high ambient temperatures.

Routine analysis

Routine analysis of the flare gases can also help to identify the source(s) of flaring by comparing the flare gas composition with the suspected flaring points (i.e. content of H_2 , C_3 and C_4 etc.).

Quantifying of flare gases

This can be done by placing ultrasonic flow meters on the main flare lines. The amounts measured here should be monitored at least once a year to decide on the capacity of a flare recovery

system project. Measuring of flare losses can also promote refinery management by showing the cost of flaring on a time basis. After taking all the preventive measures, reducing and quantifying for flaring from every single source, a further step for flare gas recovery system (FGRS) can be taken. Without taking all corrective measures, a FGRS can only be effective for circulating flaring gas within the system.

Establishing a FGRS is beyond the scope of this article, but deciding on the capacity required, an economic analysis of routine and non-routine flaring records of at least one year for various refinery capacities and operational scenarios.

Fires, fugitives and ventings

Fires for a refinery are unfortunate events and nobody can think about loss of material at the time. But at least some estimate of hydrocarbon quantity can be made since measuring such a loss is not always possible.

Fugitives are loss of light hydrocarbons to the atmosphere from flanges, pump seals, glands, drainage systems and uncapped drain points. Fugitives can be detected by portable gas detectors or by highly developed IR cameras that enable one to pin point leaks even in distances that can not be reached. A continuous inspection program can be applied and each leak can be fixed during operation or during unit shut downs.

Industrial experience shows that only 2 - 4 % of valves account for 70% of total leaks in a refinery. So, an average loss per source can be determined and can be multiplied by the number of sources to estimate fugitive leaks for loss accounting with the help of EPA protocols (EPA 453/R-95-017). The basic estimation for fugitive estimation is done by taking 0.05% of intake too.

Venting are small hydrocarbon losses from places like compressor seal vents, drain vessels and light product vents. Such vents are small compared to other losses but they should be added by sampling and calculation. The ventings can also be estimated like fugitives and can be included in the identified losses.

Evaporation losses from tanks

Evaporation losses from tanks are related with products stored (vapour pressure, tank temperature), tank design (floating or fixed roof, tank colour) and meteorological conditions. An EPA program named Tanks 4 (available at www.epa.gov.tr, free of charge) and API-Chapter 19 (evaporative loss measurement) gives a comprehensive calculation method for tanks that can be used to quantify these losses.

For the minimisation of losses from floating roof tanks, a regular maintenance program works well for roof seals and roof legs. Periodical checks for the tank seals with gas detectors helps seals to be fixed on time. Sometimes a secondary roof seal or a dome over the tank may also be necessary.

For fixed roof tanks, PV valves maintenance is important and a proven practice is to have double PV valves on a fixed roof tank to be a spare PV for refurbishment. Installing an internal footing inside the fixed roof tank is also a common experience and works well. The vapour space of fixed roof tanks may also be connected to a vapour recovery system to prevent evaporation losses.

Tank filling (displacing vapours) and tank emptying (clingage on walls) also causes vapour losses due to movement of the tank levels. For estimating the quantity of these losses API MPMS Chapter 19 can be utilised.

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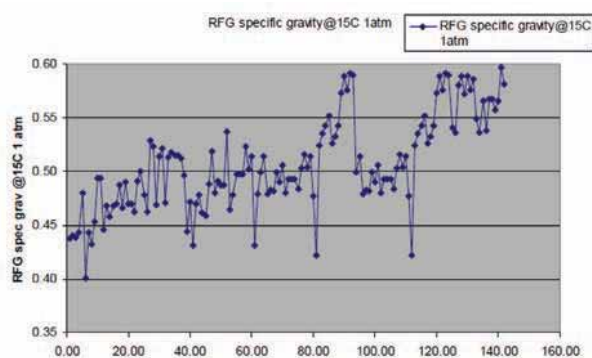


Figure 3. Variation of refinery fuel gas density through operation days.

Chemical losses

Chemical reactions in refinery process units are also a main source for losses. The main chemical losses in refinery processes are:

Claus process

For sulfur recovery where conversion of hydrogen sulfide to sulfur and water occurs. As a consequence loss of H_2S happens due to inefficiency of process and the removal of H_2 to form water and recover sulfur. A loss figure is estimated from the following stoichiometric equation by considering unit molar efficiency (Eff).



Hydrogen plants

One of main sources of chemical losses. The formation of CO_2 with the reaction of an alkane and water molecule represents the chemical loss. The composition of the hydrocarbon feed can be related to the lower heating value (LHV) of the gas and hence the following formula can be used for calculating chemical loss in the hydrogen manufacturing unit (HMU).

Loss in HMU = $(3.68 - 2.162 \cdot LHV/40\,500) \cdot H_2$ produced (LHV is the lower heating value of H_2 plant feed)

Feeds to HMUs can be different from light naphtha, liquefied petroleum gas (LPG) to natural gas or some blends of them. The changes in LHVs in the feed can also change chemical losses due to hydrogen production. Maximisation of hydrogen plant efficiency lessens H_2 losses to refinery fuel gas or flare and hence decrease chemical losses.

Chemical losses through wastewater

Wastewater that contains sulfur and nitrogen in crude oil or NH_3 formed during hydrodenitrification reactions via H_2S and NH_3 also contributes to refinery loss figure. Their amounts can be estimated from the waste water analysis.

If a refinery has a bitumen blowing process, the hydrocarbons in bitumen off gases should also be shown as chemical losses.

Coke formed and burned in CCR and FCC units

This can also be accounted and estimated from periodical catalyst analyses. The limit for coke catalysts is approximately 50 tpy to add to the loss accounting. When coke is a product from a refinery, its weight is normally quantified by weightbridge and added to loss table.

Unidentified losses

Unidentified losses are losses sourcing from measurement or calculation errors like instrument inaccuracies, uncertainties in tank calibrations, correction errors etc. The magnitude of the unidentified part of loss accounting actually shows how effectively losses are being managed. Unidentified losses can be either due to measurement with random error (measurement errors due to device) or bias error in one direction (procedural or spurious errors). Errors which lead to unidentified losses are perceived as normal practice after a period of time so such abnormalities should be properly investigated. Here is a list of common errors:

Tankage and oil movements

Tank levels, stocks, crude and other receipts, temperature corrections for tank level measurements, and meter readings, blending, intertank transfers etc. are all constituent elements of hydrocarbon accounting and accuracy of the static and dynamic measurements defines the error sources in refinery mass balance. The following items can be used as a check list:

Calibration of tanks

Changing of levels exerts pressure on tank walls and causes tank deformations and hence according to best practice, a regular tank inspection and calibration programme using methods like strapping, Opti-strap and electro optical distance ranging (EODR) should have regular intervals of 10 years. A recalibration should follow a maintenance that requires fixing inside a tank.

Level measurement

Level measurement is also a source of error in refinery mass balance. Levels can be recorded as a log manually and/or automatically by a tank gauging system. Today, modern systems mainly use radar gauges, which have no moving parts to increase their sensitivity. Measuring by radar is also useful for viscous products like asphalts since there is no part that touches the surface of the level. But there should also be a regular calibration program for radar gauges with a period of approximately three months. A level measuring log for a gauging system should contain initial tank dip, initial tank temperature, final tank dip and final tank temperature. Corrections against $15^\circ C$ should also be completed, when the data is transferred to a data historian system of the refinery. Level measurements for dynamic tanks is also a source of error since the tank level is moving.

When there is a failure in the automatic measurement oil movement people should be checking these systems and aware of failure on time to do the level measurements manually to track tank movements for oil accounting.

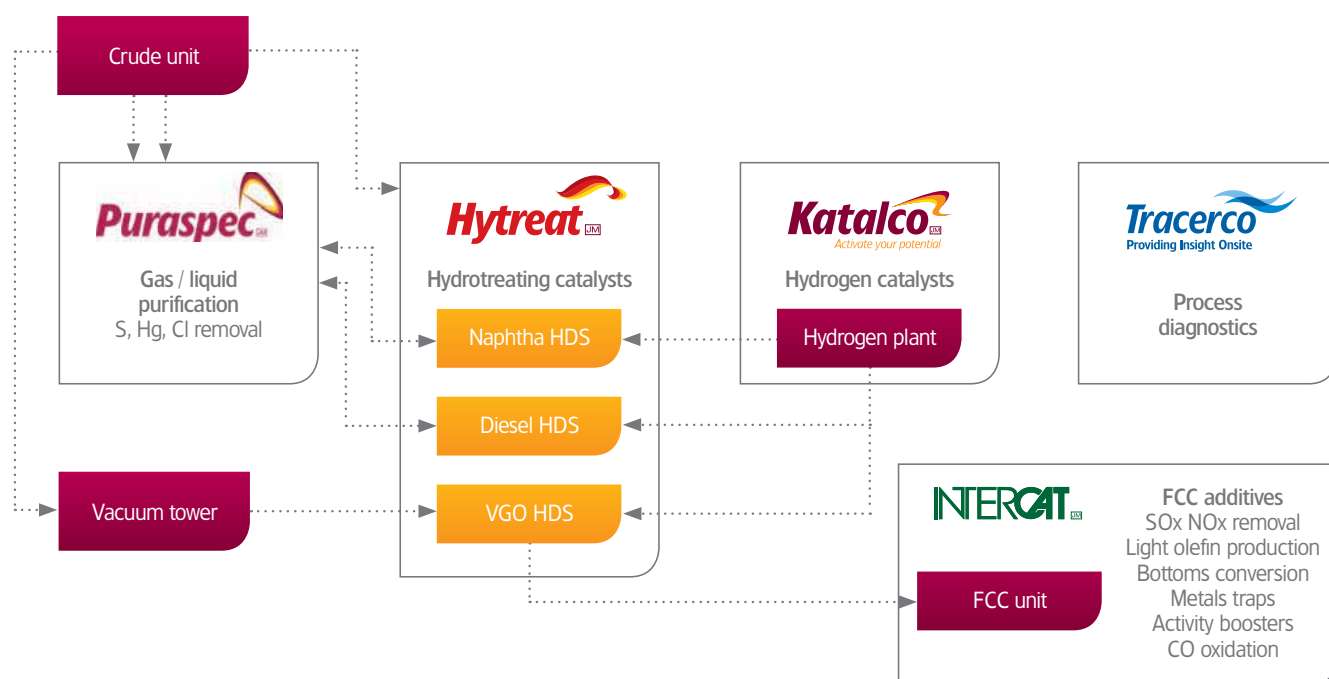
Sampling practices

Sampling practices should be arranged to ensure the samples taken will be representative of the bulk fluid. Tank samples should be composites of different levels or alternatively spot samples may be analysed separately to calculate a mean value. It also gives one an idea of how well that tank was mixed.

Water measurement for crude

This is also a main concern in loss management and refinery mass balance. It can mask the real losses if a proper analysing method is not used. There are two main methods; centrifuge (ASTM D4007) and Karl-Fischer methods (ASTM D 4377 or ASTM D 4928) where

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Table 1. Refinery loss calculation and accounting example for a complex refinery

	Mass balance and loss calculation	Tonnes		Loss accounts	Tonnes
Processed inputs	Crude	6 543 698	Flaring	Hydrocarbon flaring (measured)	10 813
	Other liquid feedstocks (import high sulfur diesel)	254 326		Pilot gas	110
	Total natural gas import (for fuel gas and H ₂ production)	62 719		All fuel gas pumps	1783
	Other inputs (oil slops from jetties)	12 000		H ₂ S flaring	
	Total	6 872 724			12 706
Non-processed inputs	Blending feedstocks	3 091 998	Measurement	Dry oil assumption in lines from jetty and SBM (1% H ₂ O)	708
	Blending additives	2.000		Measuring losses	
	Other inputs	1 897 447		COSTALD correction (assumed 0.2% increase in density)	296
	Any product returns outside custody transfer point			Crude sediment allowance (included in import accounts)	0
	Total	4 991 445			1004
Inventory	Opening stocks	376 543	Hydrocarbon emission to atmosphere	Vents to atmosphere	0
	Closing stocks	567 249		Fugitives: process and offsite areas	2 058
	Inventory change	-190 706		Tank evaporation (US EPA AP42 method)	924
Outputs	Measured dispatches	11 151 510		Process sewer system evaporation	10 457
	Other outputs (CO ₂ & sulfur sales)	18 300		LPG venting on disconnection	0
	Total exports	11 169 810			13 439
Site produced fuel	Fuel gas	139 513	Environmental discharges and waste	Oil/sulphides measured in effluent water (upstream of bio plant)	286
	Fuel oil/gasoil	213 540		Oil in sludges	286
	CCR coke	52 200	Process losses (controllable)	H ₂ S in tail gas from sulfur unit	4273
	Total site produced fuel	405 253		CO ₂ /COS/CS ₂ /hydrocarbons to sulfur unit	21
Purchased fuel	Natural gas	52 057.6		Asphalt blower	52
	Total purchased fuel	52 057.6		Solid waste (coke on catalyst, visbreaker coke .etc)	
				SWS overheads	
				SO ₂ stacks	435
				VDU ejector offgas (assumed)	2 340
			Process losses (uncontrollable)	Salt in crude	176
				Hydrogen diffusion from hydrocracker (assumed)	198
				CO ₂ from hydrogen unit (C portion)	7946
				Hydrogen ex SRU feed	1823
					10 143
Loss		46 361	Identified real loss		44 699
Loss as % of processed inputs		0.68%	Identified real loss as % of mass balance loss		96.42%
Reported loss (% loss of processed inputs of uncontrollable loss)		0.53%	Unidentified real loss as % of mass balance loss		3.58%

the former mostly measures free water and the latter gives total amount (free + dissolved) of water. Centrifuge method generally underestimates water content and does not recover water dissolved at 60 °C. Industrial practice shows that there may be considerable difference between the two and hence Karl-Fischer should be preferred for water measurements.

Some refiners also prefer to make a water balance for tracking of water entered to refinery in imported crude to allow confidence in crude receipt figures.

Especially in crude imports, amount of water may be a point of conflict between the parties. This is due to determination density and basic sediment and water (BS&W) content of crude that affects cost of crude to refinery and hence has critical importance for the measure of net crude receipt.

Static measurement uncertainties

There are always uncertainties during static measurements like tank levels, temperature, density, water etc. These uncertainties affect loss calculations between $-/+ 0.05\%$ to $-/+ 0.1\%$, depending on method used.

Refinery fuel consumption may also have a considerable masking effect on real losses. The main fuels consumed in a refinery are:

- Fuel oil.
- Refinery fuel gas (including PSA off gas from HMU and natural gas).

Fuel oil

This can either be measured by meter or by tank dip. The accurate fuel oil meters like Coriolis types for individual plants are more costly. In many refineries, a master meter is used to validate individual plant/furnace or boiler orifice meters and tank dip measurements. Individual furnace meters should be compensated against temperature and pressure although the quality and operating conditions are mainly stable. It should also be noted that mainly quarter edged orifices are preferred for fuel oil measurements for individual meters. Mass flowmeters for high viscosity fuel oils also have some applications in refineries. But they should be kept away from any vibration source like pumps or compressors. They are also limited by a line size of 6 in. max.

Refinery fuel gas (including PSA off gas and natural gas)

The measurements for fuel gas should also be made by a master meter (or sometimes more than one depending on fuel gas sources number) and individual plant and furnace/boiler individual meters. For fuel gas measurements, ultrasonic flow meters are the most suitable type and are widely used. They have no pressure loss, large flow range and good reusability.

The analysis of fuel gas is very important for density, hydrocarbon and inerts composition. The composition of fuel gas changes and hence also affects density, causing it to change considerably even in a day, depending on operation conditions of units (amine units, hydrogen plants, CCR etc.) feeding the system (Figure 3).

This variation of density for fuel gas also causes refinery material balance to oscillate. Hence, many refiners use fuel gas flowmeters (at least for master metering purposes), measuring flow

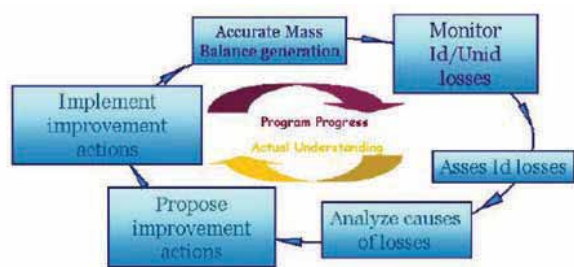


Figure 4. Loss improvement cycle.

and density together and compensating the flow automatically. For an overall measurement of loss/energy performance this is a valuable tool that can indicate areas for improvement.

Inerts in fuel gas should also be reviewed carefully to include or not include in refinery balance depending on their origin either

from crude or natural gas.

After the principal items for refinery mass balances and loss accounting, one will see a sample loss calculation with an accounting table from each one of them for a complex modern refinery. The figures in the tables were arranged to represent typical industry ranges (Table 1).


It should be noted that an average loss management means to identify at least 80% of total losses and if it reaches more than 90%, it is an indication of good control over losses. So, such a loss management also means how well that refinery is managed besides paying less money for the feed received.

Implementation of loss control

For loss control, every site may have a different culture and motivations but a recommended approach may be:

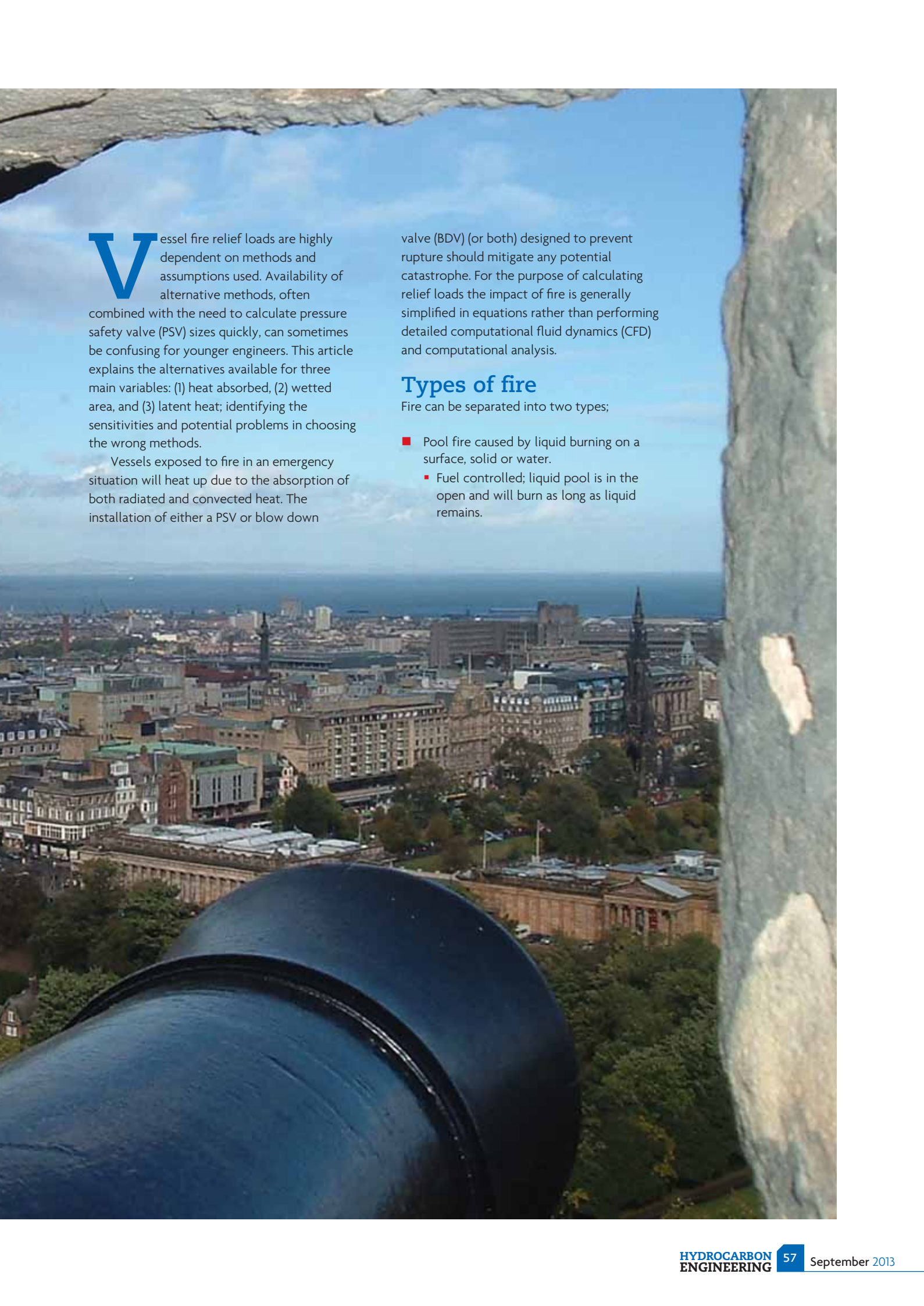
- A loss survey with an experienced outside or internal agency to identify potential areas to improve.
- Appointing a loss coordination team with clear individual responsibilities from different departments like production, finance, operation planning etc. under a wide respect coordinator that will stimulate communication between departments.
- Having training on site for measurement techniques and common errors.
- Putting objectives and targets for the refinery management for their support with a loss policy document.
- Rewarding achievements in public for loss control by monitoring progresses.
- Organising loss meetings like safety meetings at least per quarters with management team and refinery staff.
- Internal/external audits for loss control to review a better understanding of current situation and indicating any areas of weakness.
- Constituting a loss control system by procedures and written quality documents.

One of the biggest items in the operational cost list for an average refinery is the crude oil feed and from an economic point of view, refinery loss is a real financial penalty for not having an operational excellency. A loss figure found by a good oil accounting and loss prevention system is valuable information for real improvement and profitability. For European refineries, losses between 0.5 - 1.0 % of intakes are an industrially accepted pace setter.

Forming a loss improvement cycle with the steps of accurate mass balance, analysis of losses and causes, proposing improvement actions and implementing them will show that there is always room for better operations, with fewer losses and environmental effects, by using modern technology to locate and prevent them. 

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Vessel fire relief loads are highly dependent on methods and assumptions used. Availability of alternative methods, often combined with the need to calculate pressure safety valve (PSV) sizes quickly, can sometimes be confusing for younger engineers. This article explains the alternatives available for three main variables: (1) heat absorbed, (2) wetted area, and (3) latent heat; identifying the sensitivities and potential problems in choosing the wrong methods.

Vessels exposed to fire in an emergency situation will heat up due to the absorption of both radiated and convected heat. The installation of either a PSV or blow down

valve (BDV) (or both) designed to prevent rupture should mitigate any potential catastrophe. For the purpose of calculating relief loads the impact of fire is generally simplified in equations rather than performing detailed computational fluid dynamics (CFD) and computational analysis.

Types of fire

Fire can be separated into two types;

- Pool fire caused by liquid burning on a surface, solid or water.
- Fuel controlled; liquid pool is in the open and will burn as long as liquid remains.

Table 1. Equations for absorbed heat for a specified area

	Equation	Formula		Notes
		Q is heat absorbed by the vessel (kW)		
e1	API 521/ISO 23251 Pressure vessels	$Q = C F A^{0.82}$	C = 43.3 F = environmental factor	Drainage and fire fighting equipment available
e2		$Q = C F A^{0.82}$	C = 70.9 F = environmental factor	Drainage and fire fighting equipment not available
e3	API 2000 Low pressure tanks	$Q = C A A$ $Q = C A^{0.566}$ $Q = C A^{0.338}$ $Q = C F A^{0.82}$	C = 63 C = 224 C = 630 C = 42.3	2 <A> 20 m ² 20 <A> 95 m ² 95 <A> 260 m ² A > 260 m ²
e4	NFPA 58 Liquefied petroleum gas code	$Q = C F A^{0.82}$	C = 70.9 F = environmental factor	
e5	NFPA 30 2012 Flammable and combustible liquids code	$Q = C A A$ $Q = C A^{0.566}$ $Q = C A^{0.338}$ $Q = C F A^{0.82}$	C = 63 C = 224 C = 630 C = 43.2	2 <A> 20 m ² 20 <A> 95 m ² 95 <A> 260 m ² A > 260 m ²

Table 2. Equations for heat flux

	Equation	Formula		Notes
		Q is Wm ⁻²		
e6	Fire equation Combination of radiated and convected heat absorbed	$Q_{abs} = \sigma (\alpha_s \cdot \epsilon_f T_f^4 - \epsilon_s T_s^4) + h(T_g - T_s)$		Assumes vessel is fully engulfed in fire
e7	Fire heat flux No reradiation	$Q_f = \sigma (1 \cdot \epsilon_f T_f^4) + h(T_g - T_s)$		Heat from fire surface to a cold surface
e8	Fire heat flux Vessel is at a distance from fire	$Q_f = \sigma (1 \cdot \epsilon_f T_f^4)$		Vessel not fully engulfed
		Global fire (average load)	Local fire (max. point load)	
e9	Pool fire Enclosed area ventilation controlled	130	200	Initial flux against cold vessel

- Ventilation controlled; liquid pool is enclosed and burning is dependent on air supply.
- Global fire; engulfs a large part of the process units.
- Local fire; exposes only a small part to fire.
- Jet fire ignited pressurised gas or liquid (e.g. a pipe rupture or vessel flange failure).

Calculating Q

There are several equations used at present in the industry for calculating the heat input to vessels in the incidence of fire. Q can either be calculated based on kW absorbed for a specified area or a heat flux value in kWm⁻².

Heat absorbed

(e1) API 521 is probably the most used both historically and currently, but is it the best?

It is evident that these equations, shown in Table 1, will result in different answers depending on the area specified. Figure 1 shows the values obtained for areas ranging from 10 m² to 300 m².

The standard API521 equation calculates the absorbed heat and maintains the heat flux constant throughout the fire scenario with no credit given for the changing wall

temperature of the vessel nor the falling liquid level/wetted area. This may under or over predict the actual heat absorption resulting in an incorrectly sized PSV. Note that the NFPA30 equation predicts higher heat absorption for areas less than 260 m².

Heat flux equations

Other available equations are heat flux based. e6 will be recognised by many engineers as the Stefan Boltzmann equation as used in Hysys, e7 and e8 are variations on this equation.

This equation provides a more accurate estimation of the heat flux to a surface and is fully outlined in the forthcoming API 521 (ISO 23251) due out in 2013.

Valid for pool and jet fires the equations can be used to calculate both surface (e6) and local peak (e7) average [RHS1] fire heat fluxes. Equation e6 assumes a surface area fully engulfed in fire. If this is not so then the surface is exposed to flame radiation only and the convective term should be disregarded (e8). The radiation term λ_f will be adjusted with a scaling or view factor between 0 and 1. The scaling factor, or the view factor, has to be determined in each case.

The effect of fire on vessels is both complex and dynamic in behaviour. If one stands next to a large bonfire one will see flames swirling and varying in size and feel changes in temperature as fire induced air movements occur. Scale this up to an industrial size fire and one realises the difficulty in trying to model these effects in simple equation form.

Table 3 shows that the basic API521 equation e1 $Q=43.2FA^{0.82}$ in assuming a constant flux and wetted area actually predicts the lowest of the values for heat absorbed.

Several site fire trials have shown that actual heat flux rates on vessels exceed those calculated by API521 and consequently could result in undersizing of PSVs.

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Table 3. Comparison of values Q absorbed by a 100 m² wetted surface area vessel with an initial well temperature of 50 °C

			kW	%e1
e1	API 521/ISO 23251	Pressure vessels	1885	100%
e3	API 2000	Low pressure tanks	2987	158%
e4	NFPA 58	Liquefied petroleum gas code	3095	164%
e5	NFPA 30 2012	Flammable and combustible liquids code	2987	158%
e6	Fire equation		6997	371%
e7	Fire heat flux	No reradiation	9325	495%
e8	Fire heat flux	Vessel is at a distance from fire	7617	404%
			Global fire (average load) kW	Local fire (max point load) kW
e9	Pool fire	Enclosed area ventilation controlled	13 000	20 000
e10	Pool fire	Open area fuel controlled	10 000	15 000
e11	Jet fire		25 000	25 000

In reality the vessel wall temperatures, wetted areas and latent heat of vaporisation all change with time during the course of the fire. Several computer programs are commonly used nowadays to model fire scenarios dynamically providing much more accurate relief data.

Wetted area calculations

The engineer now faces the problem of calculating a wetted area of a partially filled vertical or horizontal vessel with flat, hemispherical or 2:1 elliptical heads.

The equations for these are fairly complicated and tedious for repetitive calculations if you do not have Excel at hand.

Horizontal vessels

$$A_{Wet,Cyl} = 2LR \times \cos^{-1}\left(\frac{R-H}{R}\right)$$

Detailed results indicate:

- Hemispherical heads result in the largest calculated wetted area.
- The absolute difference increases with diameter but the relative difference is independent of diameter.
- % contribution of heads relative to the body is constant at 16.7% with flat ends, 23% for 2:1 elliptical and 33.3% for hemispherical.

Hence, one can define total wetted area, 50% full vessel:

$$A_{Wet,Head} = \frac{\pi R^2}{2} \left\{ (F-0.5) \times \sqrt{1+12 \times (F-0.5)^2} + 1 + \frac{1}{4e} \ln \left[\frac{4e(F-0.5) + \sqrt{1+12 \times (F-0.5)^2}}{2-\sqrt{3}} \right] \right\}$$

- Flat ends: $1.167 \times \pi D/2 \times L$
- 2:1 elliptical: $1.23 \times \pi D/2 \times L$
- Hemispherical: $1.33 \times \pi D/2 \times L$

Although useful, this is restricted to a 50% full vessel with L/D = 3 only. It would be very useful if a simplified equation could be generated that covered varying liquid levels. Analysis of sensitivity of the wetted area to liquid height using a fixed diameter and L/D = 3 was made, liquid height is expressed as a % of diameter.

Unfortunately, a curve fitting these results does not produce a simple equation that can be used.

Vertical vessels

Similar analysis of a vertical vessel including only the bottom head and a liquid height of 50% from bottom tan reveals results identical to those of the horizontal vessel.

Total wetted area, 50% full vessel, L/D = 3:

- Flat ends: $1.167 \times \pi D/2 \times L$
- 2:1 elliptical: $1.23 \times \pi D/2 \times L$
- Hemispherical: $1.33 \times \pi D/2 \times L$

Vertical and horizontal vessels with identical dimensions and heads present equal wetted area if both are 50% full.

It was also observed that for all head types:

- For liquid heights < 50% the heads contribution is greater in a vertical vessel.
- For liquid heights > 50% the heads contribution is greater in a horizontal vessel.

This comparison is shown in Figure 2 for a horizontal and a vertical vessel, both with 2:1 elliptical heads.

Latent heat considerations

This is the final part of the jigsaw required in order to calculate the relief load to flare.

Latent heat of vaporisation for pure fluids

Calculation of latent heat for pure (single component) fluids is easy since the composition is the same for both liquid and vapour phases.

It can be quickly evaluated by performing a bubble point flash (or a flash with a very small percentage vaporised) at the required pressure and using the specific enthalpies of liquid and vapour (h_{Liq} , h_{Vap}) $\lambda = h_{liq} - h_{vap}$

Normally for single component systems the peak relief load occurs at the start of the fire when the wetted area and hence Q are at a maximum.

Latent heat of vaporisation for multi component mixtures

This procedure cannot be followed for multi component systems since the enthalpy is an intensive property and the

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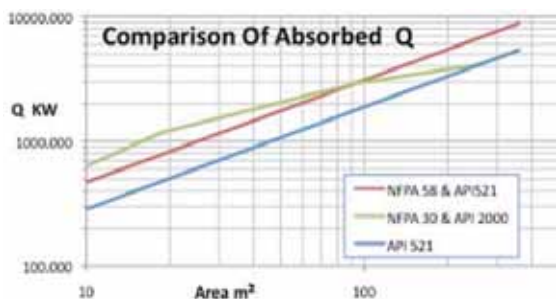


Figure 1. Values obtained for areas ranging from 10 m² - 300 m².

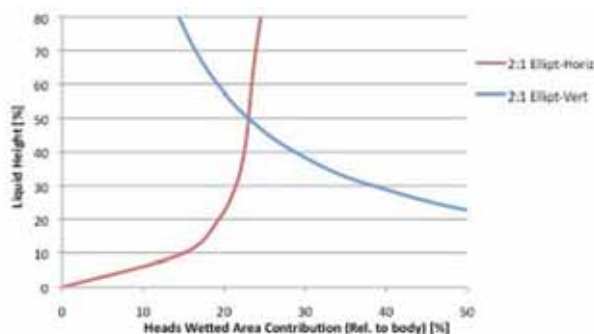


Figure 2. Relative heads wetted area comparison between horizontal and vertical vessels.

above equation does not hold when the liquid and vapour compositions are different.

In addition the liquid and vapour phase properties change during the vaporisation. The lighter components are first to leave the liquid phase and so the remaining liquid composition increases in molecular weight as the vaporisation continues. Generally for a hydrocarbon mixture the boiling point temperature will rise with time but latent heat decreases (lighter components have higher latent heat). The vaporisation procedure is consequently a batch type process and in order to estimate the latent heat for a multi component system, a commercial process simulator can be used to generate the flowsheet below:

The latent heat of the mixture will then be calculated as:

$$\lambda = (H_{\text{stream2}} - H_{\text{stream1}} - H_{\text{sensible}}) / W_{\text{vap}}$$

$$H_{\text{sensible}} = H_{\text{Liq,stream2}} - H_{\text{stream3}}$$

Where H is the heat flow and W_{vap} the vaporised stream mass flow.

It should be noted that at the beginning of relief the latent heat is higher, leading to lower relieving mass flowrate, whereas at the end a lower latent heat can lead to a higher mass flowrate. For complex mixtures the engineer should investigate the relieving flow rate resulting from different vaporisation %. The peak flow rate may occur at any time during fire relief depending on the fluid properties at that time. Detailed analysis is required to avoid improper design of safety devices and piping. This can be done using the above flow sheet scheme repeated at different vaporisation % increments in order to plot flow rate curve versus vaporisation % and define the sizing condition.

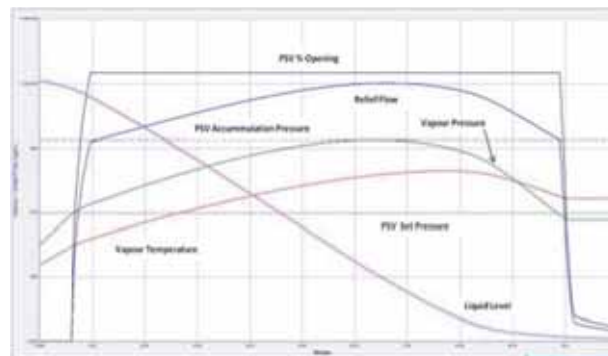



Figure 3. Information that can be obtained for a multi component fire relief case using dynamics.

Despite the complexity of the above process API521 recommends that in the absence of any idea of the value of latent heat a value of 116 kJ/kg (50 btu/lb) be used!

So finally if its not simple how does one handle it?

Dynamic simulation using a commercial process simulator is obviously the most accurate way to model fire relief systems. Figure 3 shows the type of information that can be obtained by this method and reveals some interesting points that bring into question the assumptions used for so many years in relief system sizing.

- PSVs do not lift at the start of the fire nor simultaneously with each other.
- The peak flow into the header system can occur many minutes later.
- Maximum accumulated pressure actually reached is below the 120% Pset limit as the installed PSV size is always larger than the calculated required size, so fluid parameters need to be referenced at these conditions.
- Vessels may rupture due to high wall temperatures occurring while still under pressure. 

Acknowledgements

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COVER FEATURE



In the first instalment of a two part series, **Vasileios Komvokis, Carl Keeley and Stephen D. Challis, BASF**, discuss the use of catalyst fundamentals in catalyst selection.



BOTTOMS UP!

PART ONE

Globally, approximately 35% of fluid catalytic cracking (FCC) units process predominantly vacuum gasoil (VGO) feedstock and 65% process feedstock containing resid; in some regions, the tendency to process resid is even greater. To improve the profitability of the FCC unit, increased flexibility to process resid is desired. This can be achieved through the use of advanced FCC catalysts, supported by value added technical service. Part 1 of this series will explain how resid quality impacts the FCC operation, and what key catalyst features are required to process resid. Part 2 will explore options for conducting resid catalyst selection studies and use case studies to demonstrate how innovative FCC catalysts, supported by advanced technical services and tools, can be used to increase refinery profitability.¹⁻⁵

Impact of H₂ and coke on FCC operation

Additional production of H₂ and coke from processing resid has a very significant impact on the FCC operation, as most FCC's operate to coke and gas handling limits.

Coke is any carbonaceous, high molecular weight non-volatile residue formed from cracking, and has low hydrogen to carbon ratio, i.e. has low hydrogen content. There are four types of FCC coke.

- Catalytic coke is coke associated with acid site catalysed cracking. This coke usually increases linearly with the second order activity of the catalyst.
- Additive coke is related to feedstock properties, such as, basic nitrogen content, average molecular weight and Conradson carbon. It is independent of catalyst activity.
- Contaminant coke is derived from feed contaminants that are dehydrogenation catalysts, e.g. nickel (Ni), vanadium (V), iron (Fe) etc. that remove H₂ from hydrocarbon molecules, thereby increasing the tendency to form coke.

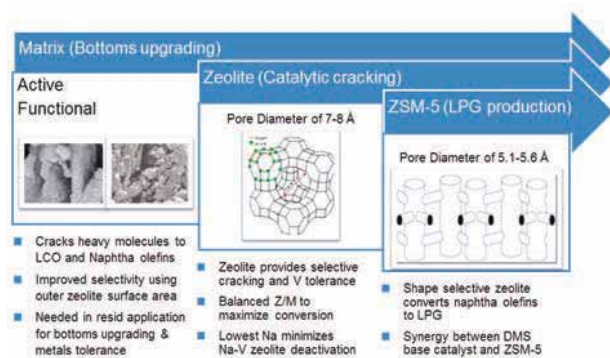


Figure 1. Catalyst fundamentals for resid processing.

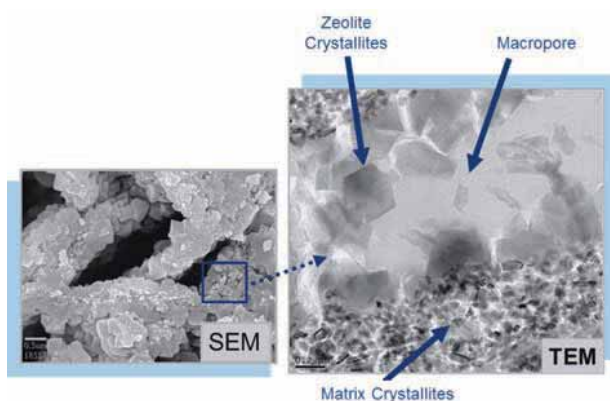


Figure 2. DMS selective precracking by exposed outer surface of zeolite.

- Cat to oil coke results from unstripped hydrocarbons remaining on the catalyst when it leaves the spent catalyst stripper.

Excessive production of H_2 and coke will constrain the amount of residue feedstock that can be processed due to wet gas compressor (WGC) and regenerator operating limits, respectively.

H_2 has a low molecular weight and density and is difficult to compress in the WGC. Increasing content of H_2 in the WGC inlet reduces the hydraulic capacity of the WGC and also the downstream equipment. Thus, it is desirable to minimise the amount of H_2 produced in the FCC.

The formation of coke on the catalyst results in FCC catalyst deactivation due to the blocking of active acid sites. Thus, the coke must be burned off the catalyst in the regenerator to restore catalyst activity. The coke burning requires O_2 from air supplied from the main air blower (MAB) and the reaction generates a significant heat release.

The MAB capacity will limit the coke burning capability. The heat release is useful and necessary to close the FCC unit heat balance as feed needs to be heated to the reactor riser temperature, and cracking reactions in the riser are endothermic (heat absorbing).

However, any heat generated in excess of unit heat balance requirements can lead to higher regenerator temperatures, which can contribute to higher catalyst deactivation and consumption, and reduced equipment operating life. Thus, it is very important to control the

production of coke, and this is particularly challenging when processing resid feedstock.

The additional H_2 and coke associated with resid feedstock can be accommodated and controlled through equipment improvements, and/or by the use of advanced FCC catalysts supported by value added technical service, or a combination of these.

FCC catalyst features required to process resid feedstock

FCC catalysts with the following features are required to increase the flexibility to process resid:

- Coke selective matrix for bottoms upgrading and metals tolerance.
- Optimised zeolite and matrix content, to maximise either conversion, or distillate yield.
- Moderate to high zeolite content to provide selective cracking.
- Low fresh catalyst sodium (Na) content to minimise Na-V zeolite deactivation.⁵

To maximise refinery profitability, FCC catalysts need to be tailored to meet the specific requirements of each refinery. The customised catalyst should have a carefully designed pore structure, matrix type and surface area, zeolite surface area, rare earth on zeolite (RE/Z), metals traps, additives etc.

These FCC catalyst fundamentals are summarised in Figure 1 and are discussed in the following sections.

Zeolite and matrix

Zeolite is an ordered tetrahedral structure of silica and alumina atoms. Several forms of zeolite exist; zeolite Y is the type presently used in most FCC catalysts. It has a regular repeating pattern giving the characteristic measured as the unit cell size. The ratio of alumina to silica in a zeolite gives it its cracking characteristics. Zeolites are deactivated by the loss of alumina from the structure in a hot atmosphere in the presence of steam (i.e. the conditions found in the FCC regenerator). A controlled dealumination by the manufacturer can be used to give the catalyst good performance characteristics. These are known as ultra stable Y (USY) or reduced unit cell size catalysts. Zeolites are responsible for most of the FCC cracking activity. Diffusion of oil through a zeolite crystal has a marked effect on product selectivities.

Matrix is defined as the non-zeolite part of the cracking catalyst. It serves both physical and catalytic functions. Physical functions include providing support for the zeolite, providing particle integrity and attrition resistance, acting as a heat transfer medium and allowing free flow of feed and products into and out of the catalyst. Catalytic functions include heavy oil upgrading and resistance to feedstock contaminant metals. There are different matrix materials available to provide for these different functions, e.g. active matrix for bottoms upgrading.

Coke selectivity

To improve coke selectivity, BASF developed the unique, award winning, distributed matrix structures (DMS)



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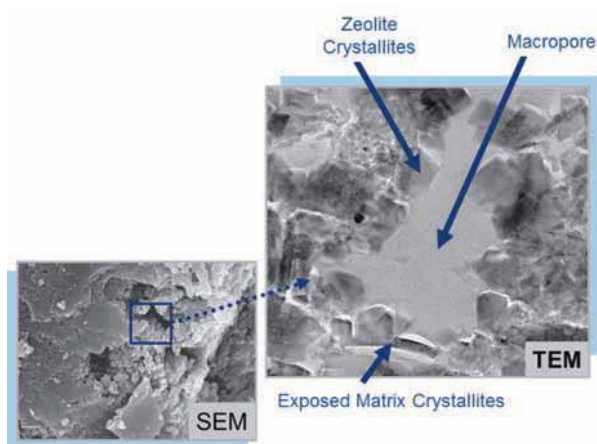


Figure 3. Close proximity of ultra low Na Zeolite and matrix.

technology platform.⁹ In the case of DMS, the matrix is designed to provide enhanced diffusion of the feed molecules to precracking sites located on the external, exposed surface of highly dispersed zeolite crystals (Figure 2).

The feed initially cracks on the zeolite surface itself, rather than on the active amorphous matrix material as is the case with other FCC catalyst technologies. This difference results in improved selectivities with the reduced coke formation characteristic of zeolite cracking.

The secondary diffusion pathway of the cracked products to the internal crystalline zeolite surface is also minimised, resulting in less over cracking (undesirable conversion of gasoline to LPG, dry gas and coke). The net result is, higher bottoms upgrading, with lower delta coke, leading to increased yield of valuable products.

A low delta coke FCC catalyst will improve conversion, and to understand this, it is first necessary to explain the relationship between delta coke, the FCC unit heat balance and operation. Delta coke is defined below:

$$\text{Delta coke} = \text{coke yield (wt\% of feed)} / \text{cat to oil ratio}$$

According to the definition, a low delta coke catalyst must either decrease the coke yield or increase the cat to oil ratio (C/O). The coke yield is essentially set by the reactor heat requirements, therefore, a low delta coke catalyst will increase the C/O in the FCC unit through increased catalyst circulation. The increased catalyst circulation delivers the following benefits: lower dry gas make, higher conversion, and higher liquid volume yield (both distillate and gasoline).

Bottoms upgrading and distillate yields

Conventionally, low zeolite to matrix surface area (Z/M) catalysts are used to maximise the distillate yield by both increasing the matrix surface area and lowering the zeolite surface area. Matrix cracking also directionally improves gasoline octane. Hydrogen transfer is reduced using matrix cracking compared to zeolitic cracking, leading to higher olefinicity.

However, without further optimisation, a low Z/M catalyst results in exaggerated matrix cracking with poor coke and gas

selectivities.¹ Thus, to maximise the bottoms upgrading, without this penalty, the correct amount of matrix and zeolite and an optimised synergy between both are required. Due to these factors, the optimal catalyst for residue feeds is one with a coke selective active matrix surface area and moderate to high zeolite surface area.

To respond to these needs, and to deliver improved bottoms upgrading and higher distillate yields, BASF developed the unique proximal stable matrix and zeolite (Prox-SMZ) technology platform; each of these functionalities is required to maximise bottoms upgrading.

Prox-SMZ was the result of the development of a unique manufacturing process. It is built on two main features, one being the presence of an ultra stable and coke selective matrix, and the other being the close proximity of an ultra low Na content zeolite and the matrix (Figure 3), which are innovatively created in a single manufacturing step.

It is the close proximity of the matrix and zeolite that leads to rapid transfer of reactants from the matrix acid sites to the zeolitic acid sites. This enhanced transfer helps to stabilise coke precursors produced by the matrix cracking, leading to higher light cycle oil (LCO) production without excessive amounts of dry gas and coke.

In Part 2 of this article, case studies will be discussed to demonstrate the benefits of the DMS and Prox-SMZ technology platforms.

Nickel tolerance

It is generally accepted that the dehydrogenation activity of metals can be expressed in terms of equivalent Ni as: $\text{Ni} + \text{V}/4 + \text{Fe}/10 + 5 \text{ Cu} - 4/3 \text{ Sb}$; typically, resid feeds contain larger amounts of Ni, V and Fe. Thus, it is especially important to mitigate the dehydrogenation effect of feedstock contaminant Ni. This is done within FCC catalysts by incorporating an active speciality alumina into the matrix to trap the Ni (this technology is available from all suppliers).

By examining spent FCC catalyst from refiners processing resid feeds with electron microscopy, it was generally observed that while V is distributed throughout the particle, Ni mainly deposits and accumulates on the outer surface of the catalyst as shown in Figure 4.

Conventional manufacturing techniques, practiced by all FCC catalyst suppliers, result in the active speciality alumina being more or less evenly dispersed throughout the particle resulting in a large portion of the alumina being located in the interior of the particle unavailable to react with Ni and in essence wasted. By using BASF's breakthrough novel and unique multi stage reaction catalyst (MSRC) manufacturing platform (Figure 4), the spatial distribution of the speciality alumina within the particle is optimised to maximise its efficiency in Ni trapping and this leads to improved catalyst performance.⁴

Fortress™ catalyst is a two stage catalyst based on the MSRC platform. The inner stage of the catalyst has the DMS structure (Figure 2) to allow enhanced diffusion of heavy molecules and selective precracking on the exposed zeolite surface. The outer stage is also based on the DMS technology platform, but is enriched with speciality alumina to trap Ni where it enters and deposits on the catalyst surface (Figure 4).

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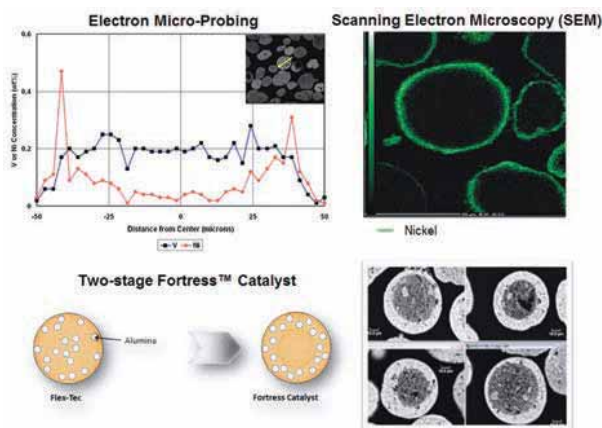


Figure 4. Multi stage reaction catalyst and Fortress™ catalyst.

When Ni on Ecat is >1000 ppm, Antimony (Sb) passivator is also injected into the feed, and the effectiveness can be monitored using the absorber off gas H_2 to methane ratio. Sb forms an alloy with Ni mitigating the dehydrogenation activity of the Ni on catalyst.⁶

Vanadium tolerance

In addition to high levels of Ni, residue feed tends to have high levels of V as well. The V deactivates the catalyst (whereas Ni does not significantly impact activity). The following summarises the various deactivation steps:⁵

- V is deposited on to the catalyst in the FCC reactor and is oxidised in the FCC regenerator.
- The oxidised form undergoes further reactions to form several highly mobile types of vanadic acids.
- These vanadic acids facilitate the removal of Na^+ from the zeolite exchange site via an acid base reaction.
- The sodium vanadate or meta vanadate hydrolyses to sodium hydroxide (Na^+OH^-).
- The hydroxyl group (OH^-) then attacks the silica oxygen zeolite framework leading to zeolite collapse/destruction.

There is residual Na^+ on the fresh catalyst from the manufacturing process. In BASF FCC catalysts, the amount of residual Na^+ on zeolite is reduced to ultra low levels by a unique combination of calcination and ion exchange steps during the catalyst manufacture. As a result, BASF's unique in situ manufacturing process produces the lowest Na^+ on zeolite FCC catalyst in the industry. This minimises the destructive effect of Na-V zeolite deactivation when processing residue feedstock containing high amounts of V.

Increasing zeolite surface area can also mitigate the destructive effect of Na-V zeolite deactivation, but this is limited by the manufacturing process.

FCC catalysts can be manufactured using either the incorporated or in situ manufacturing processes.

In the incorporated manufacturing process, zeolite and matrix are made separately and then combined together with inert fillers using a binder ('glue') to form catalyst particles. A certain amount of binder is required to sustain the physical integrity of the catalyst particle. This

constrains the flexibility to produce catalyst with high matrix and moderate to high zeolite levels.


BASF practices the in situ manufacturing process. In this process, a catalyst microsphere is manufactured and then zeolite is grown directly onto the microsphere from nutrients provided both by portions of the microsphere and supplemented by separate nutrient addition. The chemical bond between the microsphere and zeolite eliminates the need for a separate binder (used in the incorporated manufacturing process); therefore, compared to the incorporated manufacturing process, higher zeolite levels are possible, without compromising the catalyst physical integrity.

Conclusion

There is a trend towards processing heavier crudes and heavier FCC unit feeds. To process these feeds, advanced FCC catalysts are needed.

It is generally recognised that selecting the optimum catalyst for the FCC unit can lead to significant financial and reliability benefits. In a refining environment, where FCC investment opportunities are limited, refiners' are increasingly challenging the catalyst supply status quo. Drivers for changing catalyst are numerous and may include, but aren't limited to, a feed quality change, a change in operating objectives or constraints, a novel catalyst development, catalyst cost pressures etc.

To improve the flexibility to process residue feedstock, the FCC catalyst design should include: a coke selective matrix for bottoms upgrading and metals tolerance; optimised matrix content to maximise either conversion, or distillate yield; moderate to high zeolite content to provide selective cracking; and a low fresh catalyst Na content to minimise Na-V zeolite deactivation.

In Part 2 of the series, options for conducting residue catalyst selection studies will be explored with case studies to demonstrate how innovative FCC catalysts, supported by advanced technical services and tools, can be used to increase refinery profitability. 

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MAXIMISING PYGAS OPERATIONS PRODUCTIVITY

**YUZOH SATOH, CLARIANT, JAPAN,
STEFAN BREJC, AND PAIGÉ MARIE
MORSE, CLARIANT, GERMANY, AND
IAN BUTTRIDGE AND JOSEPH GENTRY,
GTC TECHNOLOGY, USA, OUTLINE THE
RECENT CATALYST AND EQUIPMENT
DEVELOPMENTS CAPABLE
OF MAXIMISING PYGAS
OPERATIONS PRODUCTIVITY.**

Pyrolysis gasoline is a common byproduct in the production of ethylene, particularly when heavier feeds are used. This mixture, typically referred to as pygas, has many reactive components and impurities that make it unsuitable for gasoline blending or other purposes without further treatment. After full hydro treatment and desulfurisation, the pygas can be upgraded into valuable petrochemical feedstocks such as benzene, toluene, and xylenes.

The raw pygas from the cracker has high potential for gum formation, in addition to existing gums, which creates a significant challenge in processing. The hydrotreatment process is highly exothermic so careful selection of equipment and catalyst can have a major impact on the reliability of operations. Recent commercial operations using OleMax® 600 catalyst from Clariant with a newly designed distributor, developed by GTC Technology, has demonstrated very stable operations for first stage diene hydrogenation with a much more uniform temperature profile and longer operating lifetime than other systems.

Pyrolysis gasoline contains hydrocarbons chains of C5 and higher and with varying levels of saturation. Processing proceeds in two stages: diolefin removal, followed by olefin and sulfur removal.

The first stage converts diolefins (or dienes) primarily, but also removes acetylenes and styrene by selective hydrogenation. Effective conversion in this stage is critical to downstream operations, enabling such processing benefits as:

- Improved oxygen stability and colour.
- Reduced gum content.
- Reduced fouling in the downstream hydro desulfurisation unit.

OleMax 600 is a high performance palladium based catalyst for this first stage hydrogenation. It uses a spherical alumina carrier and exhibits very high activity and excellent physical durability. Clariant has a long history of expertise in selective hydrogenation technology, building from predecessor company Süd-Chemie.

OleMax 600 is particularly effective at reducing styrene levels and gum formation in first stage processing. The formation of insoluble gums is a particular concern, as it is the primary reason for catalyst deactivation. This catalyst can be used in single or double bed systems.

In the single bed system, typical operations combine heavy gasoline from the quench tower and light gasoline from the debutaniser bottoms. The feed is pumped to reactor pressure, mixed with recycle and then introduced to the reactor. Hydrogen is added to the reactor to maintain system

pressure. Part of the reactor effluent is cooled and recycled to the feed, to limit the temperature rise and vaporisation across the catalyst bed.

The vapour fraction of the reactor effluent is cooled and condensed. The vent containing the excess hydrogen can be used as a hydrogen source for the second stage hydrotreater, or may be

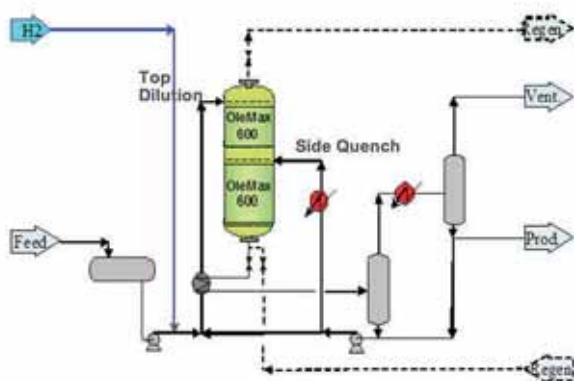


Figure 1. Pyrolysis gasoline 1st stage (two bed system).

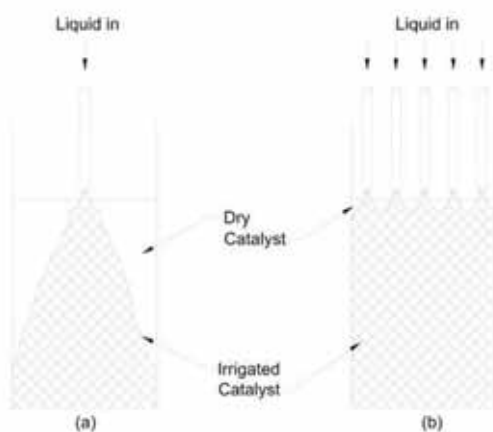


Figure 2. Increased drip points in the distributor provide improved liquid spreading and much better catalyst coverage.

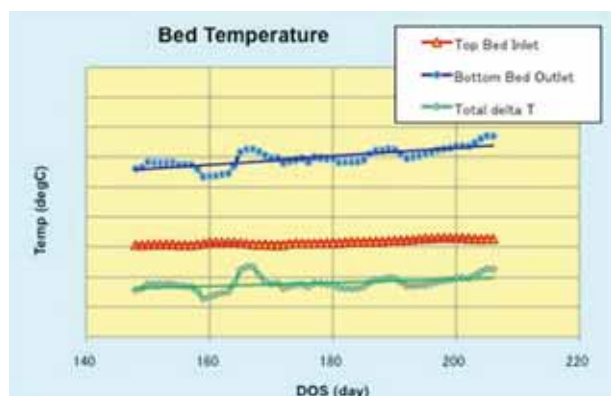


Figure 3. Stable temperatures achieved across both reactor beds using only top recycle.

directed to a Claus unit for sulfur recovery. The condensed vapour product and the balance of cooled liquid reactor effluent are fed to a depentaniser or directly to the second stage pygas hydrotreater unit. It is challenging to maintain styrene conversion above 90% per pass in the single bed system.

The two bed design is used for large scale reactors when higher diene and styrene conversion are required. Two configurations are common in the dual bed system. One process configuration recycles most of the reactor effluent back to the feed as a top dilution, which limits the temperature rise and the vaporisation across the catalyst bed.

The other configuration option uses an additional side quench that recycles product to the second bed also, as shown in Figure 1. As the top bed is more exposed to feed contaminants, it can be replaced more frequently than the bottom catalyst bed.

GTC Technology has invented a much improved distributor design that enables even and efficient mixing of the feed to the catalyst beds, and thus much more stable and efficient operations. The distributor is designed to distribute the liquid evenly across the tower cross section, while maximising the number of drip points. Figure 2 shows a simplified diagram of distributor operations, and illustrates the benefits of a drip point multiplying effect. It is important to note that physical constraints often do not allow a large amount of drip points or risers in conventional arrangement.

The GTC distributor combines unique drip point design and placement to achieve minimal cross sectional temperature variance, and greater operational stability.¹ Operations with this distributor design have four to seven times higher effective drip point density, without excessively small holes, than typical designs and can achieve an effective drip point density of 200 - 300/m².

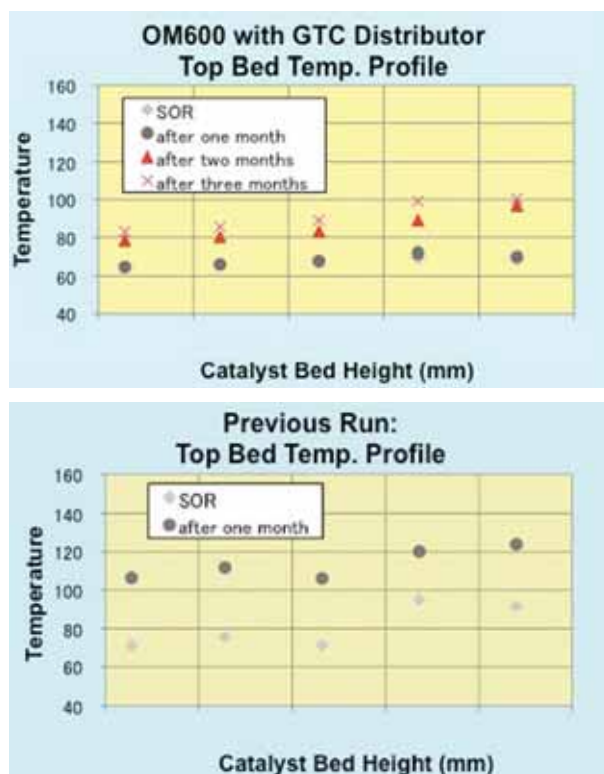


Figure 4. Top bed temperature profile for first few months of operation, with the GTC distributor and with alternate equipment. NOTE: Same temperature and height scales used for both charts.

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GTC uses a modified Moore-Rikoven method to evaluate the geometry of each distributor. This method has been used in distillation columns to calculate the distribution quality (DQ) of any given design as a fractional percentage, with 100% reflecting the ideal design. The primary drip point will be completed using an equilateral triangular pitch. However some issues prevent a perfect flow pattern, as noted below:

- The reactor is round, thus at the periphery the pitch will be slightly distorted, and spacing to the reactor wall can cause dead areas.

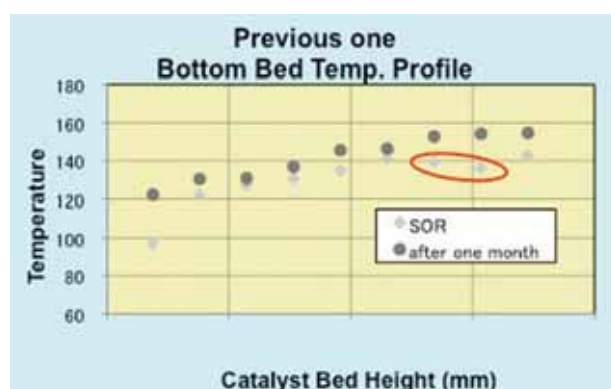


Figure 5. Bottom bed temperature profile for first few months of operation with the GTC distributor and with alternate equipment. NOTE: same temperature and height scales used for both charts.



Figure 6. Projected performance using GTC distributor with OleMax 600 catalyst.

- The distributor must be assembled in parts with a bolted removable construction.
- In some large reactors, the required support beams also distort the ideal pattern.

An optimised design will minimise the effect of these issues as much as possible, and the DQ calculation can be used to evaluate each design. An optimised layout enables distribution quality approaching 90%. The efficiency of operation is further enhanced by the use of a mixer between the reactor beds. This mixer collects and redistributes the liquid from the upper bed and mixes it with any quench feed. The liquid from the top reactor bed and the quench is brought into a central distribution point, mixed, and then sent to the distributor below. This process ensures that the lower bed achieves maximum performance, and also that any issues with the upper bed caused by catalyst deactivation or hot spots do not propagate down to the second bed. This is achieved with minimal reactor volume.


The OleMax 600 catalyst has excellent activity and stability and well balanced physical properties, which combine to increase cycle length without any process modifications. With the addition of the GTC distributor to the equipment, even longer cycle length is possible with further optimised temperature control.

The advantages of this catalyst system and distributor design are illustrated by results from commercial operations at an existing processing unit. The feedstock for this unit is raw pyrolysis gasoline and the unit was operated with a gradual inlet temperature increase to prevent leakage of outlet diene values and styrene levels. Diene values are maintained well below the target and styrene levels are less than 0.2 wt%.

The reactor demonstrated excellent temperature stability with minimal increase required after more than 200 days of service, as shown in Figure 3. The performance shows notable improvement over the previous catalyst and distributor specified by the original licensor of this unit, confirming superior catalyst stability and improved feed distribution.

This reactor stability is maintained throughout the run, as shown in Figures 4 and 5. The temperature profiles across the top and bottom beds, respectively, at one, two and three months of operation reflect only a gradual increase across the period. This contrasts with the previous system for this unit that exhibits a much less stable temperature profile.

One key performance advantage from the combined operation of OleMax 600 with the GT-599 distributor is the significant resistance to gum content in the feed. Based on current performance, the operating lifetime of the catalyst is approximately 25% longer when compared to previous operations without the GTC distributor. The expected performance improvement is illustrated in Figure 6, showing estimated cycle length improvement correlated with gum content in the feed.

Overall, the system utilising the new distributor from GTC with the OleMax 600 catalyst enables more stable operations with uniform temperature profile and longer operating lifetime. The value of this performance has been further verified with the recent selection of this system by two new customers. 

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FLEXIBILITY THROUGH CATALYSIS

C. W. Olsen, Woody Shiflett, Dan Torchia, Advanced Refining Technologies, USA, and Dave Brossard, Chevron Lummus Global, USA, discuss the best ways of optimising hydroprocessing catalyst systems for hydrocracking and diesel hydrotreating applications.

Growth in refined products is strongly driven by the demand for clean diesel and new regulations on gasoils between now and 2020. Notably, cumulative annualised global growth in diesel/gas oil demand is predicted to be about 2%, outpacing gasoline at 1.3% with the estimated gasoline/diesel gas oil ratio dropping from 0.85 in 2012 to 0.81 in 2020.¹ In addition to continued growth in Asia and continuing recovery elsewhere, motor vehicle mileage standards and ethanol mandates in North America and emerging regulatory restrictions on marine fuels add further momentum to this trend. Bunker fuel regulations becoming effective in 2015 requiring 0.1 wt% sulfur limits in Emission Control Areas (ECAs) in North America and northern Europe will underlie a demand shift to diesel/gas oil products with an expected concurrent boost in diesel price primarily due to quality requirements. Meanwhile, refining capacity additions outstripping global demand in the 2015 – 2020 period will continue to pressure refining margins. In short, it will be a period of opportunity for those refiners that have the flexibility of having hydrocracking capabilities, especially those coupled with a robust ultra low sulfur diesel (ULSD) hydrotreater, who can marry their catalyst system needs and operational responses to changing economic scenarios.

This article addresses catalyst technology approaches and processing tactics to optimise hydrocracker and ULSD unit operation in order to meet the needs for diesel and gas oil production in the future.

Molecular management

Of all diesel boiling range materials, fluid catalytic cracking (FCC) light cycle oil (LCO) stands out as one of the lowest value feedstock materials. It is usually the most difficult to manage operationally in the unit largely due to the combination of olefins and the refractory nature of the feed. It has the highest demand for hydrogen to produce a clean diesel or even 0.1 wt% sulfur marine gas oil, and offers heat release management challenges when processed at higher fractions in a unit feed. Provided there is adequate hydrogen supply,

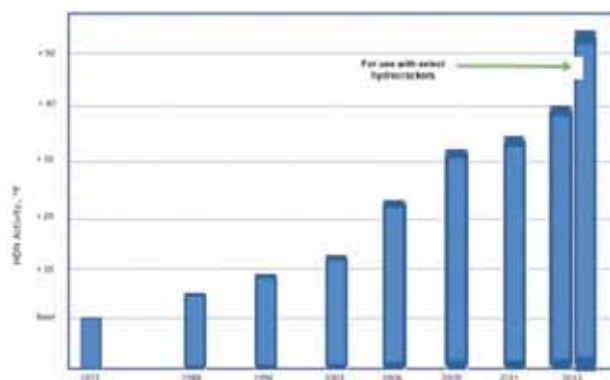


Figure 1. Innovation and delivery timeline for ART/CLG hydrocracking pretreatment catalysts.

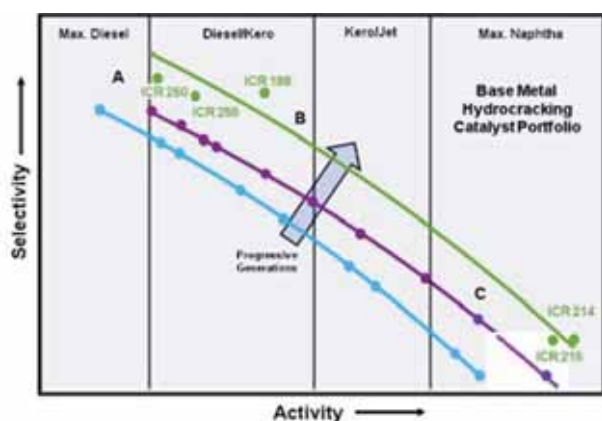


Figure 2. Performance progress of ART/CLG hydrocracking catalysts.

LCO is sometimes best processed in the hydrocracker along with other feeds such as atmospheric gas oils (AGOs) or vacuum gas oils (VGOs). Adding to the equation is that all LCOs are not created equal. Depending upon the refinery configuration and the FCC, the LCO may be produced from an FCC with a feed pretreater and consequently contain fairly modest levels of sulfur and nitrogen. Although they appear to be 'easier' feeds due to their lower levels of contaminants, the remaining impurities are also the toughest to treat.

Coker gasoils can be processed in either the hydrocracker or ULSD unit subject to individual unit capacities and infrastructure limitations such as hydrogen availability, pressure, end point, and impurities. Heavy coker gas oils (boiling well above the diesel range) are best sent to the hydrocracker although they can present challenges for catalyst life cycles depending upon design and limitations. A hydrocracker originally designed or revamped for VGO service is the most suitable outlet. This offers higher potential to maximise diesel yields, especially in recycle flow configurations and higher pressures. Light coker gasoils, on the other hand, are readily processed to ULSD in a diesel hydrotreater provided there is adequate hydrogen partial pressure and the unit has the appropriately tailored catalyst system to remove contaminants and provide sulfur conversion.

Straight run (SR) gas oils present the least challenging processing constraints, and can be fed to either the hydrocracker or ULSD unit although the ULSD unit is typically the preferred outlet. Exceptions include cases where the SR feeds are needed as 'diluent' components to aid in managing hydrogen

consumption limitations and heat release issues in hydrocrackers designed for more paraffinic and naphthenic feeds.

Processing tactics are balanced between these feedstock molecular management considerations and the designs, limitations and strategic intent of the unit in the refining scheme. Hydrocrackers have traditionally been designed with intent to pump hydrogen into the feedstock to convert heavier, higher boiling materials into more valuable products while capitalising upon aromatics saturation to increase volume swell as well as product value parameters (density, cetane, smoke point, etc.). Until recently, ULSD has been a secondary priority and generally not even a consideration in the original design of the bulk of hydrocracking units operating today. With a robust ULSD unit in the refinery, this prioritisation need not be overridden, but can be augmented by the utilisation of the latest catalyst systems for hydrocracking that have been designed for maximum hydrodesulfurisation (HDS) activity as well as the fundamental hydrodenitrogenation (HDN), hydrocracking, and saturation needs.

The versatile hydrocracker

While several hydrocracker configurations are in current usage, two dominate the landscape, especially when addressing clean fuels production: single stage, once through (SSOT) configurations and two stage, recycle (TSREC) configurations.²

The SSOT configuration is both simple and versatile and represents the simplest configuration when unconverted oil (UCO) has high value as either a lube plant feed or FCC feed. This configuration dominates the market for conversion below 70%.

Catalyst system optimisation for the SSOT is often influenced strongly by the desired outlet for the UCO it produces: lube plant feed will favor higher viscosity index (VI), aromatics saturation and desulfurisation (HDS) while FCC feed will favour nitrogen removal (HDN), poly nuclear aromatics removal and desulfurisation (HDS). Balanced with these needs are the light product drivers: ULSD or the less demanding 0.1wt% sulfur marine fuel. If ULSD production is a target and cannot be produced within the SSOT unit constraints, it is critical to factor in and model the effect of this preprocessed component as feed to the ULSD unit. It will clearly include more difficult, sterically hindered sulfur compounds for HDS in the ULSD unit.

Catalyst system selection and optimisation is controlled by a number of constraints that must be accommodated in a single stage:

- Hydrodemetalisation (HDM) needs driven by heavy vacuum gas oil (HVGO) and/or deasphalted oil (DAO) components as well as by crude source (arsenic and other contaminants) and coker products in the feed (silicon contaminations).
- HDN needs for the hydrocracking function lower in the catalyst system.
- Hydroconversion targets for removal of heavy components.
- HDS needs for products such as ULSD, marine gas oil, and UCO.
- Aromatics removal (lube or FCC applications).
- Isomerisation (for lube needs).

SSOT catalyst system optimisation is further challenging when the application involves a unit converted from former service such as FCC pretreatment or, in less common cases, diesel

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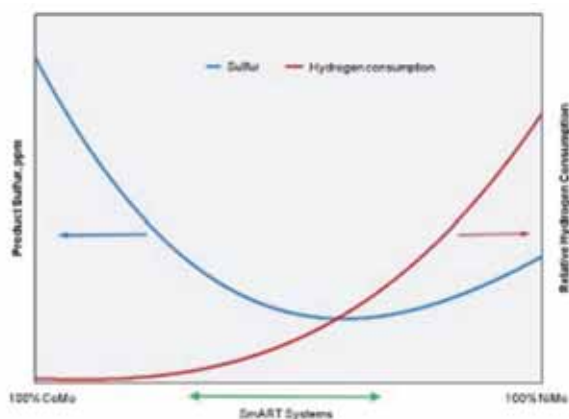
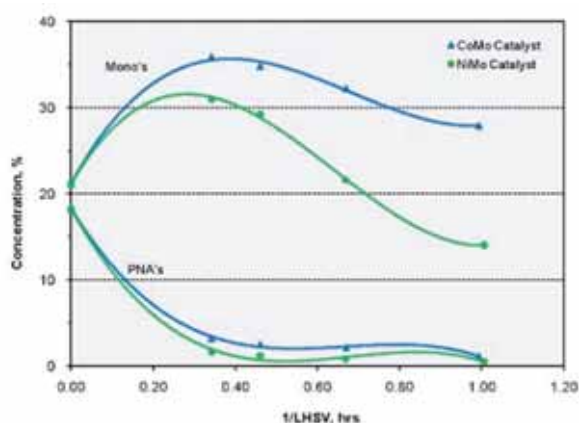
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Table 1. Comparison of hydrocracker feed and operation

	1960s	2000s
Average unit size, bpd	10 000 - 20 000	50 000 - 60 000
Feed S, wt%	0.5 - 1.0	2.0 - 3.0
Feed N, ppmw	500 - 1000	2000+
Feed % cracked	10 -20	25 - 70+
Primary product	Naptha/jet	Jet/diesel
Naptha S, ppmw	0.5 - 1.0	0.5 - 5.0
Diesel S, ppmw	100 - 500	5 - 10

**Figure 3.** Optimal hydrogen use by SmART catalyst system.**Figure 4.** Comparison concentration profiles for poly and mono ringed aromatics as a function of residence time for a CoMo and NiMo catalyst (under typical ULSD unit condition).

treating. In such cases, heat release and hydrogen consumption come into play as these units typically contain only a few, deep beds. Semantics can sometimes obscure the proper application of catalyst technology. A 'mild hydrocracker' is merely a low conversion SSOT (<40%) and is most effectively evaluated as a part of the SSOT catalyst system continuum.

SSOT systems typically demand the highest activity catalyst components for hydrotreating needs for HDN and HDS. ART and CLG have met this challenge over time as can be noted from the rapid innovation and delivery of hydrocracking pretreatment catalysts (Figure 1). ART's latest pretreat catalyst ICR 512, has successfully met the feed severity challenges to enable obtaining the most from the downstream hydrocracking catalysts. In

addition, it has allowed product properties, such as ULSD, to be maintained and even improved with more difficult feedstock.

Lacking the flexibility of recycle and adjustment of the recycle cut point (RCP), product selectivity in the SSOT is controlled by the choice of the catalyst system, and largely the choice of the hydrocracking catalyst component(s), and the operating temperature regime and span. Hydrocracking catalyst product

families typically exhibit a trade off between selectivity to distillates and activity (temperature required for a target conversion level). Premier catalyst performance is defined by innovations that increase both selectivity and activity. Figure 2 shows the progression of such performance for the hydrocracking catalysts provided by ART.

Catalyst system design in an SSOT can involve more than a single solution. While a single hydrocracking catalyst from the 'B' range might seem an obvious solution for a refiner desiring 'A' selectivity but lacking the infrastructure to compensate for the lesser activity, synergies in multicatalyst combinations might instead point to a system of 'A' and 'C' catalyst and can actually achieve a better result than pure 'B' alone.

TSREC configurations offer a high level of flexibility in addition to providing the more favorable means to achieve conversion levels of 90% plus. TSREC configurations also are the preferred means to achieve full naphtha/gasoline selectivity.

TSREC units offer the refiner the ability to operate the two stages differently in order to simultaneously meet separate goals for each stage. This configuration also offers the flexibility to 'balance' the stages to optimise the desired product selectivity and qualities. Note that although this unit is shown with two reactors, they are often built with multiple reactors, generally as part of the first stage providing even greater ability to process poorer value stocks. As an example, the first stage could be targeted to both provide a diesel draw suitable for marine fuel blending as well as pretreatment for the second stage which could be targeted to produce ULSD. Contrasted to the SSOT, the TSREC has added operational flexibility provided by the ability to adjust RCP and per pass conversion in each stage plus a second catalyst system that allows optimisation of an additional catalytic component. In addition, TSREC will enable production of other very high quality products such as JET-A which can be recovered as a single product or recovered with the ULSD, depending upon economics of the day. In addition, feedstocks can be shifted with the ULSD unit to further add operating space. This integration and flexibility permits the refinery to take full advantage of seasonal or frequent economics.

Catalyst system selection and optimisation in the first stage is often influenced by feed quality and contaminants; nitrogen, sulfur, metals, silicon, and arsenic being the typical suspects. This is especially challenging for older, existing units but equally even for new units. A comparison from the early days of hydrocracking to the present is shown in Table 1.

Following removal of feed contaminants, the remaining catalyst volume can be used to achieve the conversion and selectivity goals. Often the first stage is required to achieve 50 - 60% conversion after removing feed contaminants for both



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stages. Depending upon unit objectives, first stage cracking catalyst can be chosen from any of the 'A', 'B', or 'C' groups and will include catalysts such as ICR 183, ICR 180, or ICR 162. These catalysts have exhibited very stable and robust performance to achieve these levels of conversion to the desired products and with a long operating cycle.

Second stage catalyst selection will largely be driven by performance of the first stage to achieve the desired overall unit goals. Second stage catalyst will contribute significantly to product quality improvement and to the ability to achieve very high levels of conversion to the desired product, typically diesel or total distillate. ICR 240 and ICR 250 have provided many two stage recycle units with a very robust operation and very high product quality such as jet smoke point and diesel cetane.

Combined with a robust first stage pretreat and hydrocracking operation described earlier the second stage can provide unmatched selectivity and product qualities while processing even the most difficult feeds.

Catalyst system optimisation

Catalyst system selection and optimisation is also a critical aspect for ULSD production in a diesel hydrotreater. As discussed above for hydrocrackers, catalyst selection for ULSD is dependent upon feed quality and contaminants. In addition, operating constraints such as H₂ availability exert significant influence on the catalyst system selection. ART introduced the SmART Catalyst System® in 2001 to help refiners deal with these complex interactions between feed and operating constraints while at the same time meeting the unit operational goals. The system utilises the latest catalyst technology, which is layered in the proper proportions to provide the best performance while at the same time meeting individual refiner requirements. The catalyst staging is designed to take advantage of the different reaction mechanisms for sulfur removal; a high activity CoMo catalyst such as 420DX or 425DX efficiently removes the unhindered, easy sulfur via the direct abstraction route, and a high activity NiMo catalyst like NDXi or 545DX then attacks the remaining sterically hindered, hard sulfur. Experience has proven that the properly configured SmART Catalyst System provides better performance than either the CoMo or NiMo catalyst alone.

One of the key advantages of the SmART Catalyst System is the efficient use of hydrogen. Figure 3 illustrates how the system can be tailored to provide the best balance of high HDS activity while minimising H₂ consumption. The figure shows that as NiMo catalyst is added to the system there is a significant increase in HDS activity relative to the all CoMo reference, and eventually, a minimum in the product sulfur curve is reached (i.e. maximum HDS activity). The position and magnitude of this minimum varies with feed properties and operating conditions.

The figure also shows the relative H₂ consumption, and again, as the percentage of the NiMo component increases, the H₂ consumption relative to the base CoMo system increases. In the region where the system shows the best activity the hydrogen consumption is only slightly greater than that for the all CoMo system, and well below that for the all NiMo catalyst. This is a direct result of the different kinetics for sulfur and aromatics removal and is a critical consideration when customising a SmART Catalyst System.

For units that have hydrogen constraints the key to designing the proper catalyst system is increasing the


hydrogenation selectivity to provide highest HDS activity while at the same time minimising hydrogen consumption (i.e. minimising excess aromatics saturation). How this is accomplished is shown in Figure 4 which compares the concentration profiles for poly and mono ringed aromatics as a function of residence time for a CoMo and NiMo catalyst under typical ULSD unit conditions.

The figure shows a rapid decrease in poly aromatics concentration and a corresponding increase in mono ringed aromatics for both catalysts as the residence time is increased. Clearly, however, the NiMo catalyst is much more efficient at hydrogenating the final aromatic ring as evidenced by the lower mono ringed aromatic concentration with increasing residence time compared to the CoMo catalyst. At the longest residence time (lowest LHSV) on the chart, the NiMo catalyst has about 15 numbers (absolute) lower mono ringed aromatics concentration than the CoMo catalyst, and that corresponds to about 50 Nm³/m³ higher hydrogen consumption for the NiMo catalyst.

These data demonstrate that the hydrogenation activity of the system can be tuned by adjusting the relative volumes (i.e. residence times) of the CoMo and NiMo beds in the reactor. Of course, not all units have a H₂ constraint, and in those cases the incremental increase in aromatics saturation and the correspondingly higher hydrogen consumption obtained by the NiMo catalyst like 545DX offers benefits such as cetane improvement and the ability to process more cracked stocks. Commercial experience with the SmART catalyst system has demonstrated that a properly designed ULSD unit combined with the right catalyst system can process up to 100% cracked stocks to produce <10 ppm sulfur and providing significant cetane uplift and volume swell.

A hallmark of the SmART system is flexibility. In applications where there is sufficient H₂ availability and partial pressure, a catalyst like 545DX is likely the most active system for HDS. However, it will consume significantly more hydrogen due to its efficiency at catalysing hydrogenation reactions. If the incremental hydrogen consumption cannot be tolerated, a SmART system can be designed which will deliver high HDS activity and minimise hydrogen consumption. In cases where the hydrogen pressure is lower, the SmART system is often more active than either component alone without increasing the H₂ consumption significantly over the all CoMo system. For low pressure diesel units the best option is a catalyst like 425DX which has been specifically designed for superior performance at low pressure.

Conclusion

State of the art technical support is required to get the most from today's advanced catalysts and catalyst systems. Assessing and recognising unit limits is key to achieving the refiner's operating goals, especially before a hard limit is reached that will result in a shortened cycle, reduced throughput, or both. 

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ENHANCING CATALYST PERFORMANCE



LARRY S. KRAUS AND JOHN A. SMEGAL AND KARL M. KRUEGER,
CRITERION CATALYSTS & TECHNOLOGIES, USA, DISCUSS
METHODS OF IMPROVING ULTRA LOW SULFUR DIESEL CATALYST
PERFORMANCE AND FEED FLEXIBILITY IN LOW PRESSURE UNITS.

Improving ultra low sulfur diesel (ULSD) catalyst performance and feed flexibility in low pressure units and when processing difficult high endpoint/high nitrogen feeds are ongoing challenges in refining. Difficult feeds offer the opportunity to increase unit flexibility and profitability, but make achieving product specifications and process objectives, such as cycle length, more challenging, especially in a H_2 constrained environment. Criterion has used improved catalyst formulation and manufacturing technology to provide an improved CENTERA® CoMo catalyst to address these needs.

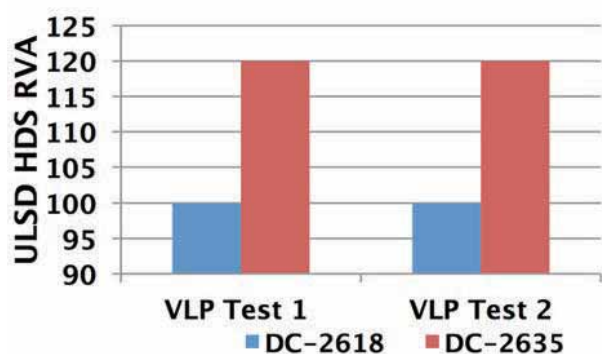


Figure 1. ULSD HDS RVA: VLP Test 1 and VLP Test 2.

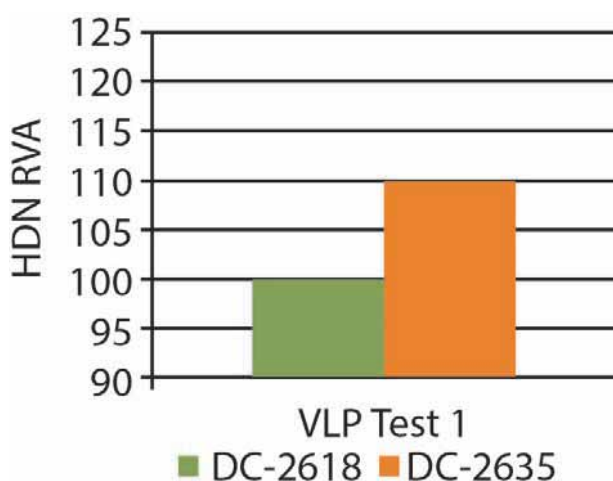


Figure 2. HDN RVA: VLP Test 1.

Table 1. VLP pilot plant test feeds.

	VLP Feed 1	VLP Feed 2
Feed type	SRGO	SRGO
Density @60F (g/cc)	0.8483	0.8413
API Gr @60F	35.3	36.9
Sulfur (wt%)	0.378	1.14
Nitrogen (wppm)	20	52
UV aromatics (wt%)		
Mono	6.03	5.25
Di	4.30	3.90
Tri	0.56	0.82
Tetra	0.44	0.52
Poly	5.3	5.24
Total	11.33	10.49
D-2887 distillation (wt%)	°F/°C	°F/°C
IBP	252/122	269/132
10%	446/230	454/234
20%	489/254	505/263
30%	512/267	531/277
50%	549/287	572/300
70%	582/306	602/317
90%	618/326	649/343
95%	631/333	666/352
EP	658/348	707/375

CENTERA® Type II CoMo and NiMo catalysts have shown improved activity compared to previous Criterion catalyst families in ULSD¹, fluid catalytic cracking pretreatment (FCC-PT)², and hydrocracking pretreatment (HC-PT) application areas. The improved active site architecture responsible for the high activity of CENTERA® catalysts is documented and discussed elsewhere.¹ Recent proprietary improvements in catalyst formulation chemistry/technology³ and manufacturing technology have resulted in major improvements in CENTERA® catalyst performance. Pilot plant testing data is presented demonstrating the ULSD, hydrodesulphurisation (HDS) and hydrodenitrogenation (HDN) activity improvements of DC-2635, Criterion's latest generation CoMo CENTERA® catalyst, compared with DC-2618, the first generation CoMo CENTERA® catalyst, over a wide range of process conditions and feed properties.

Activity testing results

Results from several pilot plant studies comparing the activities of DC-2635 and DC-2618 CoMo CENTERA® catalysts are reported. These catalysts were compared at very low (290 - 340 psig/20 - 23 barg) to moderate/high (910 psig/63 barg) pressure levels with light, low organic N (N = 20 - 52 wppm, low/moderate T - 95 = 631 - 666 °F/333 - 352 °C) and higher sulfur, nitrogen and endpoint (S = 1.4 wt%, N = 210 wppm, T - 95 = 795 °F/424 °C) feeds.

Very low pressure (VLP) testing

Pilot plant tests were performed comparing DC-2635 and DC-2618 ULSD HDS and HDN activities at very low pressure (VLP - 290 - 340 psig/20 - 23 barg). The feeds tested were light straight run gas oil (SRGO) materials. The properties of the test feeds are shown in Table 1.

These process conditions and feed properties are representative of typical commercial VLP ULSD operations. The ULSD HDS results obtained in VLP Test 1 and VLP Test 2 are shown in Figure 1. HDN results for VLP Test 1 are shown in Figure 2. These plots show the relative volume activity (RVA) of DC-2635 and DC-2618 for ULSD HDS (Product S = 10 wppm) and deep HDN (Product N = 5 wppm).

In both VLP tests, DC-2635 provides a 20% improvement in ULSD HDS activity. In VLP Test 1 DC-2635 showed 10% higher HDN activity than DC-2618. These significant increases in activity can allow processing of more difficult feeds or higher feed throughputs in the challenging VLP environment. In VLP Test 2 essentially identical product nitrogen levels were observed with both DC-2618 and DC-2635. This indicates that an 'HDN floor' had been reached with both catalysts.

The H₂ consumption was the same with both catalysts in VLP Test 1. Under the VLP Test 1 conditions, DC-2635

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provided substantial ULSD HDS and HDN improvements without increasing H₂ consumption.

Low/moderate pressure testing

In previous Criterion publications, synergistic performance of CoMo/Al₂O₃ and NiMo/Al₂O₃ catalysts in stacked systems has been discussed. Stacked systems have been used to increase catalyst system activity while maintaining/managing

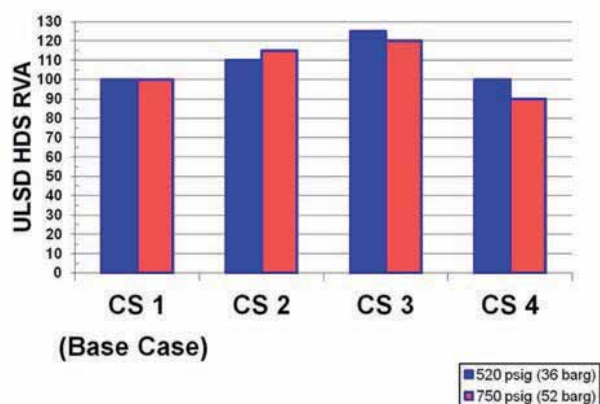


Figure 3. ULSD HDS RVA: Stacked catalyst system tests.

Feed type	SRGO/LCO (820/20 vol. ratio)
Density @ 60F (g/cc)	0.8697
API @60F	31.20
Carbon (wt%)	86.09
Hydrogen (wt%)	12.47
Sulfur (wt%)	1.310
Nitrogen (wt%)	206
UV aromatics (wt%)	
Mono	6.44
Di	8.35
Tri	2.48
Tetra	0.97
Poly	11.80
Total	16.24
SFC aromatics (wt%) (D-5186)	
Mono	17.3
Poly	21.3
Total	38.6
D-2887 distillation (wt%)	°F/°C
IBP	228/109
10%	409/209
30%	484/251
50%	537/281
70%	594/312
90%	667/353
95%	695/368
FBP	747/397

H₂ consumption at acceptable levels. This has been an effective means to improve the performance of low pressure and H₂ supply constrained units and has been demonstrated commercially.^{4,5}

This pilot plant test series evaluates CoMo/NiMo/CoMo catalyst systems utilising first generation CENTERA® CoMo (DC-2618) and NiMo (DN-3630) catalysts as well as the latest generation CENTERA® CoMo (DC-2635) and NiMo (DN-3636) catalysts. DN-3636 is the latest generation NiMo catalyst and utilises similar proprietary improvements in catalyst formulation chemistry/technology³ and manufacturing technology as are used with DC-2635. The ULSD HDS RVA of DN-3636 is 115 - 120 compared to DN-3630 (ULSD HDS RVA = 100).

All CoMo/NiMo/CoMo catalysts systems were used in a 15/30/55 volumetric ratio. Catalyst System 1 is the incumbent system used in a commercial ULSD unit and is the base case for the comparison/evaluation. Catalyst System 2 replaces the DC-2618 in Catalyst System 1 with DC-2635. Catalyst System 3 utilises new generation DC-2635 and DN-3636 in the same CoMo/NiMo/CoMo proportions as Catalyst System 1. Catalyst System 4 is DC-2635.

The feed used in the stacked catalyst system testing was an 80/20 blend (volumetric basis) of SRGO/light cycle oil (LCO). The properties of the feed used in these pilot plant tests are shown in Table 2.

The process conditions used in this series of tests are representative of the commercial operation processing the feed described in Table 2. The actual commercial operation pressure is 520 psig/36 barg.

A primary motivation for using a stacked catalyst system in this commercial unit is maximising ULSD HDS activity while controlling/managing H₂ consumption. ULSD HDS and relative H₂ consumption (RHC) data obtained with the catalyst systems tested are shown in Figures 3 and 4.

From Figures 3 and 4 it can be seen that at 520 psig/36 barg, the DC-2635/DN-3630/DC-2635 catalyst system (CS2) ULSD HDS RVA was 110 compared to the DC-2618/DN-3630/DC-2618 base system with no additional H₂ consumption. At this pressure level the DC-2635/DN-3636/DC-2635 catalyst system (CS3) ULSD HDS RVA was 125 compared to the base case catalyst system while incurring only ~2% higher H₂ consumption. At 520 psig/36 barg DC-2635 (CS4) showed the same ULSD HDS activity as the previous generation stacked system at about 4% lower H₂ consumption.

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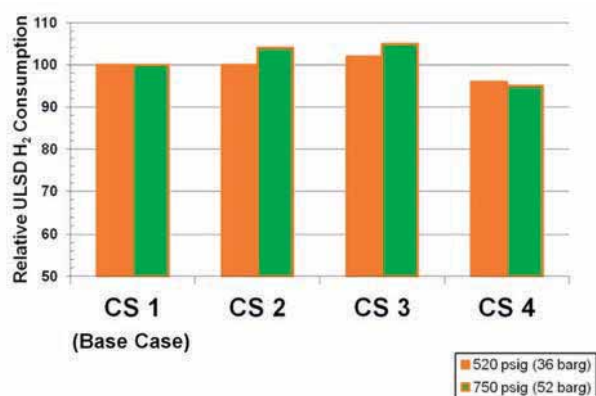


Figure 4. ULSD RHC: Stacked catalyst system tests.

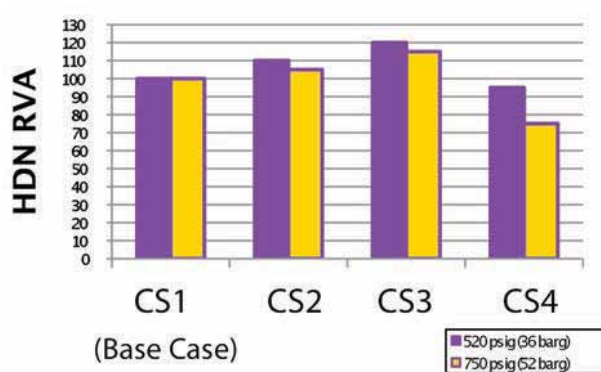


Figure 5. HDN RVA: Stacked catalyst system tests.

Table 3. High endpoint SRGO feed properties	
Feed type	High EP SRGO
Density @60F (g/cc)	0.8680
API Gr @60F	31.5
Sulfur (wt%)	1.41
Nitrogen (wppm)	210
UV aromatics (wt%)	
Mono	5.10
Di	3.81
Tri	1.87
Tetra	1.29
Poly	6.97
Total	12.07
D-2887 distillation (wt%)	°F/°C
IBP	305/152
5%	443/228
10%	488/253
30%	568/298
50%	619/326
70%	676/358
90%	760/404
95%	795/424
EP	861/461

At 750 psig/52 barg pressure, the DC-2635/DN-3630/DC-2635 and DC-2635/DN-3636/DC-2635 catalyst systems had ULSD HDS RVA values of 115 and 120, respectively, compared to the base case stacked system. The corresponding RHC values were 104 and 105, respectively. At 750psig/52barg, the DC-2618/DN-3630/DC-2618 system had a ULSD HDS RVA of 100 and an RHC of 100 compared to respective values of 90 and 95 for DC-2635. The difference in the relative performance of these two systems at the 520 psig/36 barg and 750 psig/52 barg pressure levels is likely better utilisation of the DN-3630 in the base case system at the higher pressure level.

The HDN RVA observed with the four catalyst systems tested are shown in Figure 5. In general, the NiMo containing systems show higher HDN activity than DC-2635 at both pressure levels tested. The higher HDN RVA observed with DC-2635/DN-3630/DC-2635 when compared with the HDN RVA of DC-2618/DN-3630/DC-2618 indicates that DC-2635 enhances the HDN capability of the CoMo/NiMo catalyst system. This is consistent with the results observed with direct DC-2635/DC-2618 comparisons. The increased HDN activity of the DC-2635 catalyst systems make them more robust and flexible to feed changes. Incorporating DN-3636 into the stacked system with DC-2635 results in the highest catalyst system HDN activity. Combined CoMo/NiMo catalyst system ULSD HDS activity has been shown to correlate with catalyst system HDN activity in previous development work.⁴

Moderate/high pressure testing

In distillate hydrotreating, high endpoint (EP) feeds are more difficult to process than lighter feeds. It is well known that the feed total S concentration increases, and, even more importantly, the concentration of difficult to remove, 4,6-dimethyl-dibenzothiophene-derived species increases as feed EP increases. In addition to this, the nitrogen content of a distillate feed and the concentration of difficult to remove N species increase sharply as EP increases. Organic N is a strong inhibitor of ULSD HDS reactions and must be reduced to low levels for ULSD HDS to be effective.⁶ The ability of a catalyst to process high EP feeds to produce ULSD strongly depends on its ability to effectively remove feed nitrogen.

The pilot plant testing discussed in this section evaluates the performance of DC-2635 and DC-2618 when processing a very high EP (T - 95 = 795 °F/424 °C) SRGO feed. The properties of this feed are shown in Table 3.

Process conditions correspond to the process conditions employed in the commercial operations that process this feed. The results obtained with DC-2635 and DC-2618 when processing the feed described in Table 3 at the conditions described in Table 4 are shown in Figures 6 and 7.

In Figure 6 it is seen that DC-2635 has ULSD HDS activity that is 17 – 19 °F (9 – 11 °C) more active than DC-2618. This is approximately equal to a 135 - 140 ULSD HDS RVA for DC-2635 compared to DC-2618 (ULSD HDS RVA = 100). DC-2635 showed a 9 – 13 °F (5 – 7 °C) HDN activity advantage over DC-2618 in the high EP feed

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Some may describe Innovative R&D as mysterious and obscure. Karl Krueger describes it as an opportunity to put inventive minds to work. With the goal of gaining a more fundamental understanding in surface science in order to advance the performance of hydrotreating catalysts to the next level, Karl and his colleagues are given the opportunity to pursue cutting-edge research and experimentation using less traditional methods. It's freedom like this that's led CRITERION to deliver game-changing technologies like CENTERA® which has shown superior catalyst performance. After all, don't most breakthroughs come from novel and forward-looking science?

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Table 4. High feed endpoint pilot plant test process conditions


	Condition set 1	Condition set 2	Condition set 3
Pressure (psig/barg)	655/45	655/45	910/63
LHSV (hr ⁻¹)	0.64	0.61	0.90
H ₂ /oil (SCFB/Nm ³ /m ³)	2030/340	1805/300	2085/350
Target S (wppm)	10	10	10

Commercial success

CENTERA® DC-2635 provides strong ULSD HDS and HDN performance gains over the first generation CENTERA® CoMo

catalyst. This allows it to provide value improvements to the refiner over a wide range of operations. Since its release in early 2013, DC-2635 has been selected for use in three commercial catalyst loads and has demonstrated top tier performance in industrial pilot plant testing programs.

Conclusions

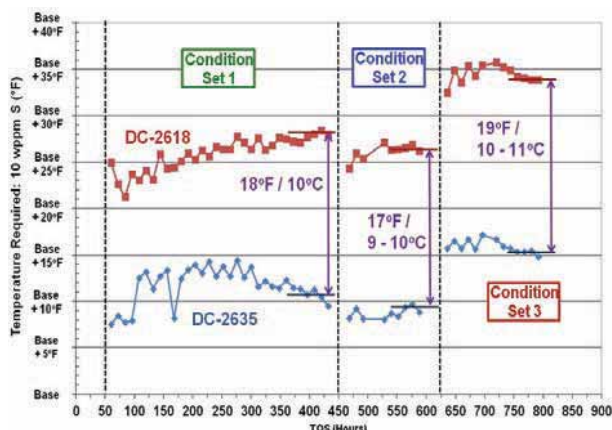
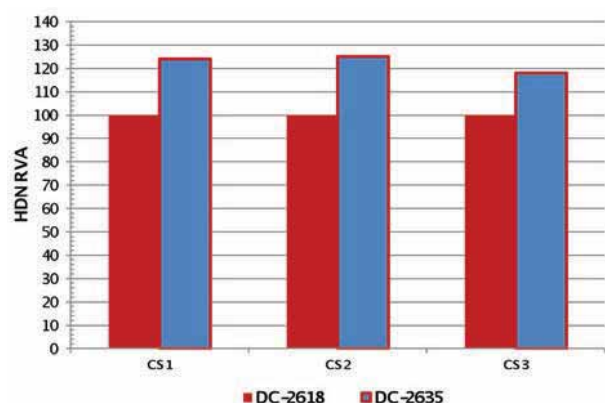
The improved catalyst formulation chemistry/technology and manufacturing technology of DC-2635 result in it having substantial performance improvements over the previous generation of CENTERA® CoMo catalyst (DC-2618). DC-2635 ULSD HDS RVA = 120 - 140 and HDN RVA = 110 - 125 compared to DC-2618 (ULSD HDS RVA = 100, HDN RVA = 100) over a wide range of operating pressures (290 - 910 psig/20 - 63 barg) with essentially no increases in H₂ consumption. These activity improvements enable DC-2635 to process more difficult feeds and make it robust to feed property variations. The high HDN activity of DC-2635 make it a means to implement product upgrading technology requiring low organic nitrogen levels, such as shape selective cracking dewaxing⁷, in units that are H₂ constrained and, therefore, are difficult applications for traditional HDN (NiMo) catalysts. DC-2635 shows the same high stability of other Criterion CENTERA® and Ascent products. 

Acknowledgements.

The authors would like to acknowledge the contributions of Theo Meuris and David Grisafe in the development of DC-2635.

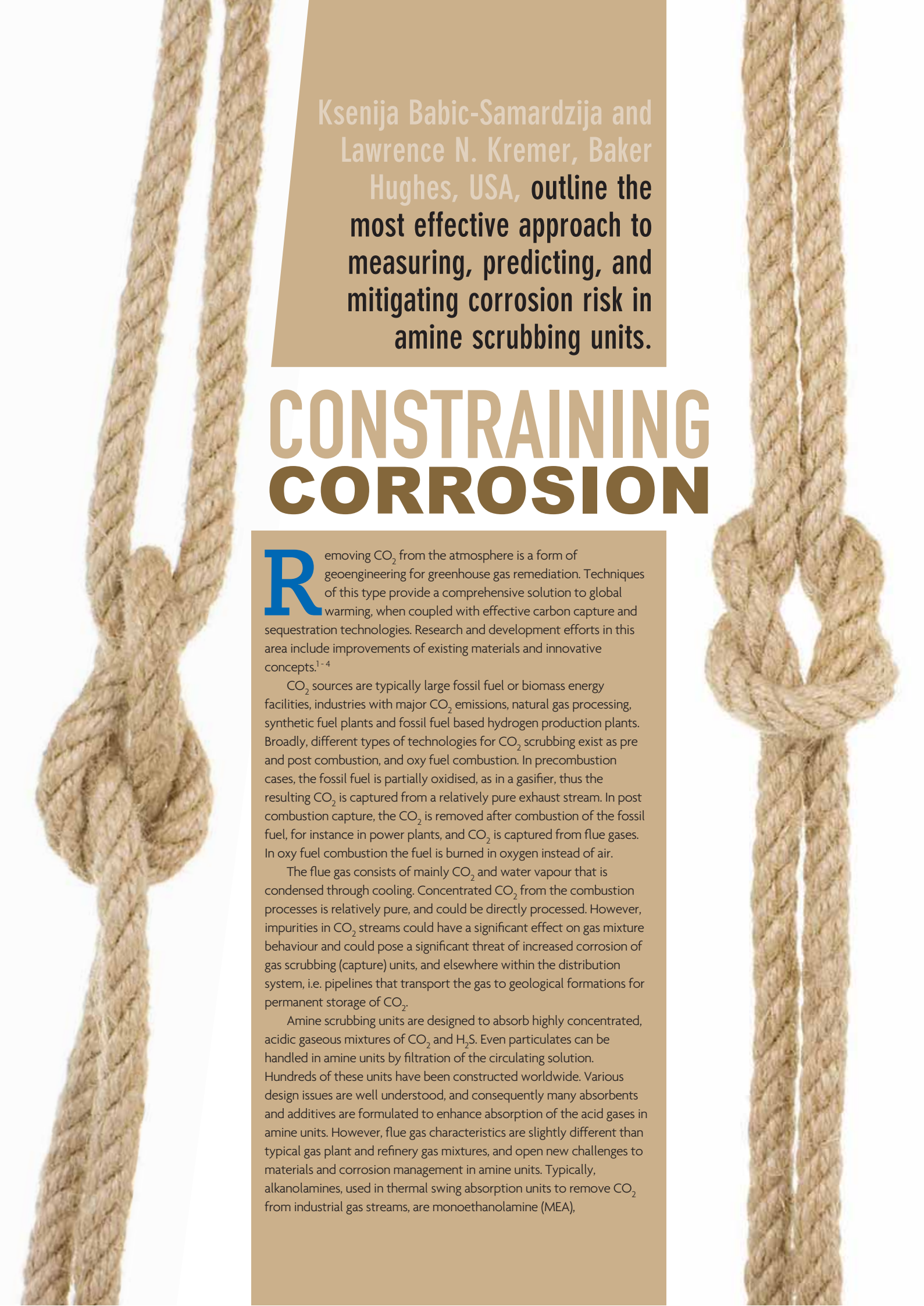
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**Figure 6.** High endpoint feed processing: ULSD HDS temperature requirements (10 wppm S).**Figure 7.** HDN RVA: High endpoint feed processing.

testing. This corresponds to an HDN RVA = 120 - 125 for DC-2635 as compared with DC-2618 (HDN RVA = 100) as shown in Figure 7. The improved ULSD HDS performance of DC-2635 can be in part attributed to its superior HDN activity. The ULSD HDS activity stability of DC-2635 was equivalent to that of DC-2618.

Managing H₂ consumption in commercial hydrotreating units is an important issue. Many units that apply CoMo catalysts are H₂ constrained and H₂ consumption management is critical. The H₂ consumption data obtained with the high EP feed indicate that at start of run (SOR) conditions and equivalent product S levels, the H₂ consumption with DC-2635 was the same as that observed with DC-2618. Equivalent H₂ consumption at SOR with DC-2635 is due to the large reduction in SOR temperature requirements (17 - 19 °F/9 - 11 °C) to meet the target S level with this catalyst. This results in an SOR operating temperature requirement that is in a temperature region where the rate of aromatics saturation (ASAT) is reduced.



Ksenija Babic-Samardzija and Lawrence N. Kremer, Baker Hughes, USA, outline the most effective approach to measuring, predicting, and mitigating corrosion risk in amine scrubbing units.

CONSTRAINING CORROSION

Removing CO₂ from the atmosphere is a form of geoengineering for greenhouse gas remediation. Techniques of this type provide a comprehensive solution to global warming, when coupled with effective carbon capture and sequestration technologies. Research and development efforts in this area include improvements of existing materials and innovative concepts.¹⁻⁴

CO₂ sources are typically large fossil fuel or biomass energy facilities, industries with major CO₂ emissions, natural gas processing, synthetic fuel plants and fossil fuel based hydrogen production plants. Broadly, different types of technologies for CO₂ scrubbing exist as pre and post combustion, and oxy fuel combustion. In precombustion cases, the fossil fuel is partially oxidised, as in a gasifier, thus the resulting CO₂ is captured from a relatively pure exhaust stream. In post combustion capture, the CO₂ is removed after combustion of the fossil fuel, for instance in power plants, and CO₂ is captured from flue gases. In oxy fuel combustion the fuel is burned in oxygen instead of air.

The flue gas consists of mainly CO₂ and water vapour that is condensed through cooling. Concentrated CO₂ from the combustion processes is relatively pure, and could be directly processed. However, impurities in CO₂ streams could have a significant effect on gas mixture behaviour and could pose a significant threat of increased corrosion of gas scrubbing (capture) units, and elsewhere within the distribution system, i.e. pipelines that transport the gas to geological formations for permanent storage of CO₂.

Amine scrubbing units are designed to absorb highly concentrated, acidic gaseous mixtures of CO₂ and H₂S. Even particulates can be handled in amine units by filtration of the circulating solution. Hundreds of these units have been constructed worldwide. Various design issues are well understood, and consequently many absorbents and additives are formulated to enhance absorption of the acid gases in amine units. However, flue gas characteristics are slightly different than typical gas plant and refinery gas mixtures, and open new challenges to materials and corrosion management in amine units. Typically, alkanolamines, used in thermal swing absorption units to remove CO₂ from industrial gas streams, are monoethanolamine (MEA),

Table 1. Lean MEA average solution analysis for 3 years prior to use of new corrosion inhibitor

	Lean amine appearance	Amine conc. (wt%)	H ₂ S conc. (ppm)	CO ₂ conc. (ppm)	HSAS (wt%)
3 year average	Yellow to brown	11	22.2	2856	7.4

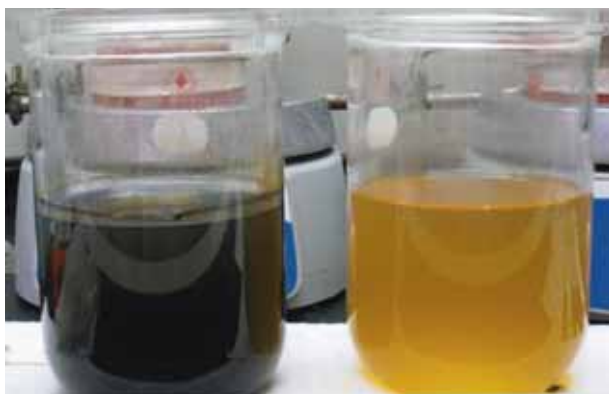


Figure 1. Appearance of the field lean MEA solution after laboratory evaluation, under gas mixture of 14% CO₂, 3% O₂ in N₂ containing approximately 4% of HSAS and 1100 ppm of soluble iron; left sample is untreated versus right sample treated with OPTIMA corrosion inhibitor.

diethanolamine (DEA), dimethylethanolamine (DMEA), and methyldiethanolamine (MDEA).

Standard amine unit solutions are sensitive to oxygen contamination. When oxygen levels are more than a few ppm in the scrubbed gas, formation of heat stable amine salts (HSAS) increases and the corrosion potential increases dramatically. Flue gas contains a 2 - 6% level of oxygen and a high CO₂ concentration (10 - 15%). Consequently, the absorption of flue gas increases the potential for severe corrosion and fouling. The addition of oxygen scavengers has been used successfully in the past to counter the oxygen in amine units in gas plants and refineries⁵, in cases where oxygen was present in ppm levels. In coke oven amine units, oxygen scavengers have not been used successfully because the level of oxygen is too high.

Coke plants, for example, convert coal into coke that is used to smelt iron ore in blast furnaces. Heated coal produces coke and coke oven gas (COG). The COG contains many valuable gaseous components that are recovered in the byproducts plant (naphthalenes, phenols, as well as benzene, toluene, and xylene) and contaminants that must be removed so the resulting gas can be sold or used internally, such as when firing the coke battery. The remaining COG contains methane, carbon monoxide, and hydrogen that can be burned to provide heat for the coke ovens or used for other industrial purposes. Unfortunately, the COG still contains H₂S, CO₂ and oxygen. The H₂S must be removed because, when burned, it forms sulfur oxides (SO_x) that are detrimental to other uses. The CO₂ must be removed because it degrades the heat value (Btu) of the COG. Coke plants use temperature swing adsorption units containing MEA to remove the acid gasses. The coke oven gas contains 0.5 - 1.5% oxygen, so it is similar to the gases that might be expected to be scrubbed from coal fired power plant flue gas. The high levels of oxygen can degrade the amine solutions and greatly increase the corrosion of steel in the system. Many COG amine units experience high corrosion rates because of the high levels of oxygen, as shown in our laboratory study and field application.⁶

To support flue gas scrubbing units, new corrosion inhibitors have been developed that are effective in oxygenated amine solutions. Hence, in the light of industry challenges with carbon capture it is important to provide support and develop new technologies to implement effective corrosion control methodologies. This article covers a new technology development for acid gas mixture scrubbing and its application in the field. Advancements in corrosion inhibition using an additive program were followed over a number of years and supported from a field application perspective to obtain optimum unit performance.

Experimental

For in house corrosion measurements, lean amine field samples were obtained from US operating plants, where amine units are used as scrubbers for gas mixtures containing mostly CO₂, H₂S, and in some cases a small amount of oxygen. Other gas components might be present in <1% levels. Field lean amine samples were investigated, taking into account unit processing conditions, and evaluated using adequate metallurgy.

Typically, carbon steel (CS1018) and stainless steel (SS316L) metallurgy were used for corrosion evaluation, relative to the in place metallurgy at the field location. Weight loss methods and electrochemical measurements were used for corrosion rate evaluation. Experiments were conducted with a continuous purge of a gas mixture containing 14% CO₂ and 3% O₂ in nitrogen in the glass kettles at a temperature of 180 °F (82.2 °C). In experiments where weight loss was not used as the primary source of data, the three electrode setup was designed for an electrochemical cell. Electrochemical measurements were performed specifically for observation of corrosion rate vs. time and changes occurring on metal surfaces.

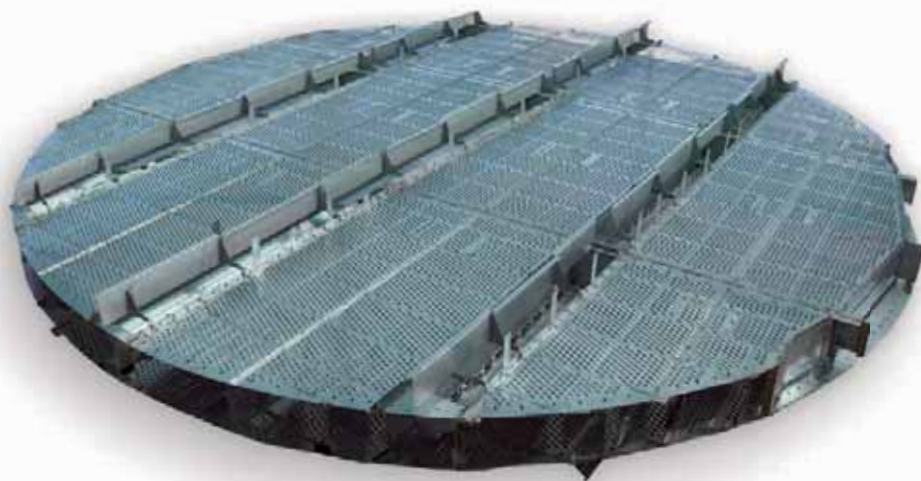
Field corrosion techniques include soluble metals (specifically iron) measurements by inductively coupled plasma (ICP), corrosion coupon measurements by weight loss, and measurement of other control parameters, such as heat stable amine salts and amine absorbent concentrations.

Results and discussion

Traditional corrosion inhibitors become weak surface protectors when used in highly oxygenated systems.⁴ These inhibitors are typically attracted to the metal surface and form a protective layer. The oxygen in the flue gas, however, causes the layer to become unstable and incapable of providing full protection. This situation has been verified in the laboratory and field operations in amine units scrubbing gas mixtures with oxygen contamination.⁶

Synthetic flue gas, composed of 14% CO₂ and 3% O₂ in nitrogen, was used in laboratory tests to measure the difference in corrosion inhibitor performance. At the conditions where flue gas is purged into an amine absorbing system, the corrosion rate increases significantly and becomes uncontrollable in the presence of traditional corrosion inhibitors.⁴ In addition, an accelerated corrosion mechanism is typically followed by formation of heat stable amine salts i.e. bicine, formates, oxalates, acetates, etc. Consequently, development of corrosion protective additives that are used in oxidising gas environments

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during the carbon capture processes becomes essential for deployment of new technologies.

Newly OPTIMA™ corrosion inhibitors show substantial decrease in corrosion and fast kinetics, followed by strong and continuous absorption on the surface. It is obvious that uncontrollable corrosion in these systems leads to high metal dissolution and high soluble iron content, as seen in a laboratory experiment showing the untreated solution with a black appearance, Figure 1. The chemically treated solution, even highly contaminated with heat amine stable salts and particulates from

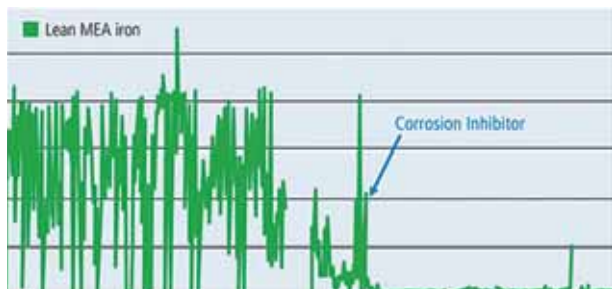


Figure 2. Corrosion measured by soluble iron concentration (g/L): past and current data.



Figure 3. Stainless steel reboiler on left (before new inhibitor applied) shows severe corrosion while stainless steel reboiler on right (after using new OPTIMA corrosion inhibitor for one year) shows no signs of corrosion.

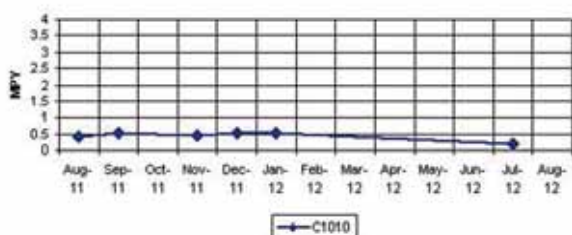


Figure 4. Corrosion rate (mpy) in the amine unit after monthly or six month inspections; carbon steel metallurgy.

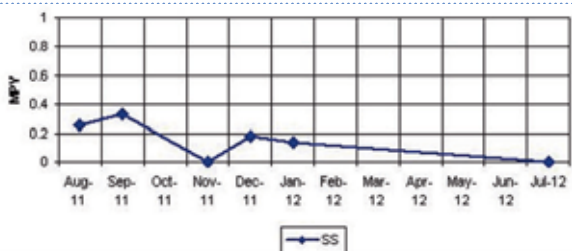


Figure 5. Corrosion rate (mpy) in the amine unit after monthly or six month inspections; stainless steel metallurgy.

the field lean amine solution, show no traces of discoloration and no elements of corrosion in the presence of the newly developed OPTIMA corrosion inhibitor. Fresh lean amine solutions are taken from the field location, containing a significant amount of soluble iron caused by corrosion and other contaminants as HSAS. Further experiment was performed with carbon steel electrode in the presence of flue gas (CO_2 and oxygen) thus affecting metallurgy by increasing corrosion in untreated sample and suppressing corrosion in inhibitor treated solution, Figure 1.

Field application of the newly developed corrosion inhibitor was observed over a one year period in a system where the acid gas mixture was captured in a circulating MEA system. To understand the chemical balance and challenges in the system, three years of monitoring data were evaluated. The amine shows variations in concentrations during the time period, followed by system upsets in the range of 4 - 15% by weight, Table 1. Unstable amine concentration potentially influences other system changes such as heat stable amine salts formation, corrosion, and amine strength.

Within the circulating amine system, regenerated lean amine is returned to the system carrying heat stable amine salts. Observations of the field amine unit revealed that within the three year period there were significant amounts of chemical changes in the system that caused high contamination and instability, changing the integrity of the system. Table 1 shows the average acid gas mixture addition to the amine system, the concentration of lean MEA in a range below typical concentration and high level of HSAS, 7.4%. Of these, mostly formate, thiocyanate and acetate were extremely high. Within the equilibrium of the lean rich amine, rich amine was highly loaded with CO_2 and H_2S gases, Table 2.

Corrosion was so severe that frequent repair and replacement was required for carbon steel and stainless steel trays, transfer lines, and reboiler tube bundles. Over the period of three years, system upsets and uncontrollable operations led to higher amounts of HSAS, and the corrosion process became more pronounced. Typically, HSAS should be in the range of 2 wt% or below. Increase in HSAS significantly increases potential for corrosion, affects pH, and brings contaminants and corrosive substances to the system. Consequently, the soluble iron concentration in lean amine was enriched as a result of metal dissolution, i.e. vessels, pipes, exchangers and reboiler parts, and increased rapidly to very high levels, in average of >4400 ppm.

Table 3 shows soluble metals average for the three year period, measured by ICP analysis from lean amine solution. Very high levels of soluble iron brought attention to a high and continuous corrosion process that obviously was not under proper operational control. Equipment leaks and failure were the response to this uncontrollable corrosion mechanism in the unit. Table 3 shows high levels of iron in the lean MEA as well as chromium and nickel from corrosion of stainless steel. To bring the system under control, it was necessary to control the corrosion.

Understanding process challenges and investment in new technology development can deliver solutions for even the most difficult systems. The application of a new OPTIMA corrosion inhibitor in lean amine MEA solution in this system resulted in significant improvements and system stabilisation within a short period of time, Figure 2. Before the inhibitor injection, lean MEA amine shows elevated levels of HSAS (approximately 5 wt% and

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Table 2. Rich MEA solution acid gas load average

	H ₂ S rich loading (ppm)	CO ₂ rich loading (ppm)
3 year average	1402	9523

Table 3. Lean MEA solution selected metal ICP analysis average

	Iron (ppm)	Chromium (ppm)	Nickel (ppm)
Avergae 3 years	4407	13	8.4

more) and >1500 ppm of soluble iron (period recorded from 1 - 20 months from the graph). This condition was caused by application of a traditional corrosion inhibitor, which in this case where traces of oxygen were present, did not perform at expected efficiency. Before new inhibitor application, the amine unit was shut down, cleaned up and fresh amine, with the addition of make up lean amine was filled in to meet 100 ppm of soluble iron concentration. Thus, following new OPTIMA inhibitor injection, starting from month 21 on the graph, the corrosion rate in circulating amine solution stayed continuously low e.g. iron levels did not increase over period of application. Figure 2 shows that iron levels, since beginning the new inhibitor treatment, have remained below past concentrations resulting in a very consistent unit operation.

The presence of the newly developed corrosion inhibitor in the lean amine solution supported continuous operation; the amine unit was managed more efficiently. Even HSAS concentration was stabilised to about 2 wt% either by the presence of corrosion inhibitor and/or manageable operation condition. The resulting soluble iron was approximately 24 ppm, continuously and consistently resulting in improvement of unit corrosion. The success of the new inhibitor treatment program can also be attributed to a much improved reclaimer operation, with consistent operating temperatures, while maximising the cycle runs with effective boil out and dumping procedures.

The coke plant Sulfiban unit, during the annual turnaround inspection, showed successful operation and reduced system maintenance with minimal diversion due to equipment failure. Since the use of corrosion inhibitors at startup, iron levels have been controlled below 30 ppm and HSAS concentrations between 1.5 - 2%. Inspection of the contactor (absorption tower), and top of the acid gas, showed no corrosion pitting or localised corrosion attack. Inspection of the reboiler (made of stainless steel metallurgy) was of special interest because it was a problematic area in the past. The reboiler was inspected and cleaned up consistently every three months because it was generally very corroded and with limited use, Figure 3. After the OPTIMA inhibitor application, visual and detailed inspection confirmed that there was no evidence of corrosion; unlike its highly noticeable presence in the past.

Figure 3 shows the stainless steel reboiler tube bundle from the 2006 outage after three months of exposure compared to the same metallurgy in 2012 (after a one year experience with the new OPTIMA corrosion inhibitor).

Corrosion in the Sulfiban is also monitored monthly with the use of corrosion coupons installed in the reboiler line. These coupons are changed and weighed each month, to determine monthly metal loss due to corrosion. Taking into account an extremely high corrosion rate that occurred previously in the system, corrosion coupon measurements


appeared to be an unsuccessful technique because coupons were fully dissolved after one month of exposure. Soluble iron was the main source of corrosion monitoring; and as shown in Figure 2 and Table 3, the corrosion levels were very high.

During the corrosion inhibitor application, metal coupons showed low corrosion rates in ranges <0.5 mpy on metallurgies, carbon steel, and stainless steel. For the past six month period coupons were unable to change due to the leakage of the isolation valve. Due to safety reasons, the period was extended to the unit turnaround date. Figure 5 displays the corrosion coupons monthly rate after 196 days of exposure, resulting in an excellent year for the Sulfiban unit. The plant has operated in a very efficient manner, maintaining compliance, minimising downtime, reducing equipment failure, and maximising MEA usage.

Figures 4 and 5 show the corrosion coupon results in the presence of the OPTIMA corrosion inhibitor, on carbon steel, and stainless steel metallurgy over the one year period.

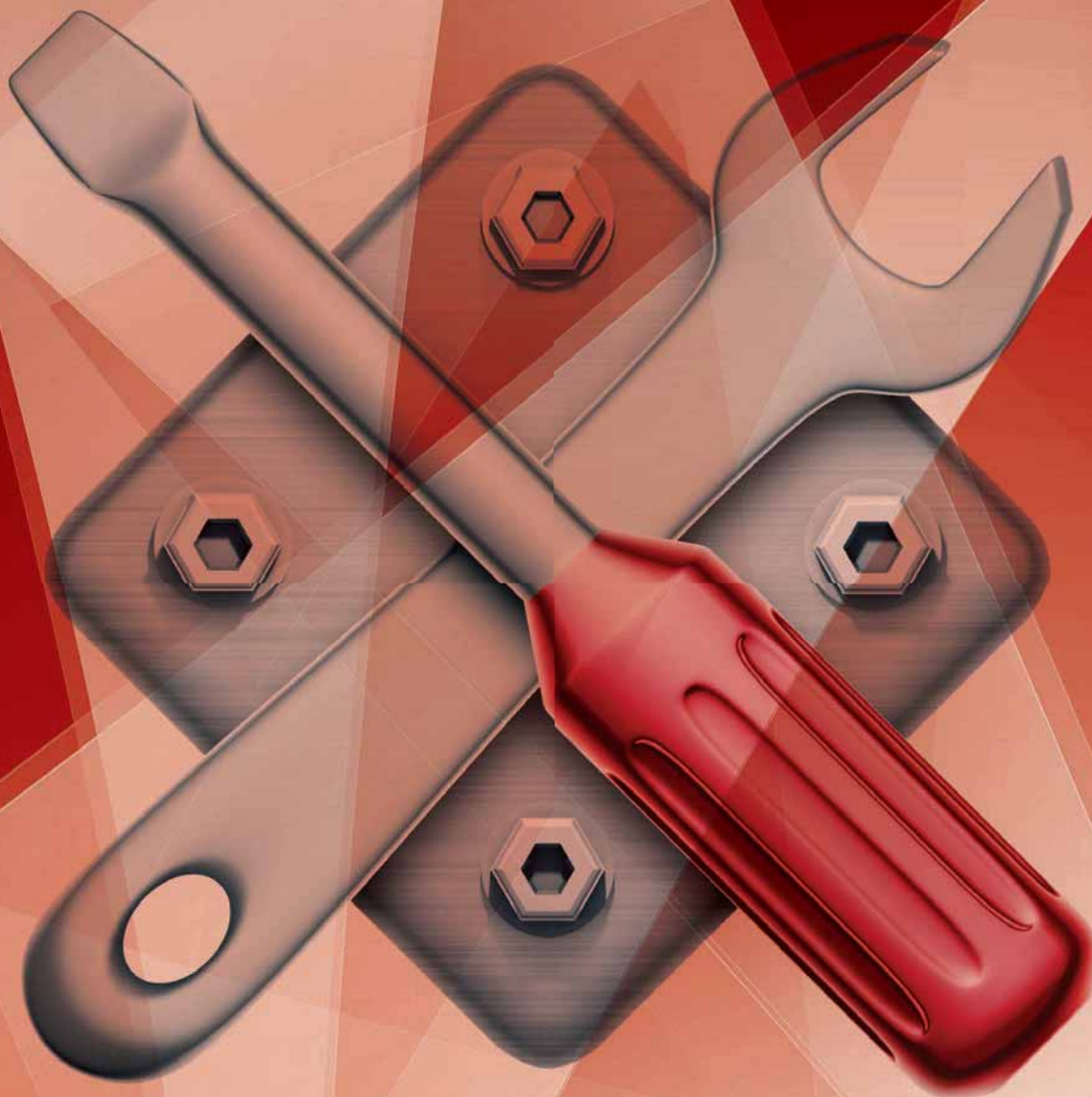
Conclusion

The key to the reliable and safe performance of amine units designed for acid gas scrubbing is efficient corrosion control. Development of a new corrosion inhibitor that can efficiently protect metal surfaces is one of the critical factors for carbon capture operation and control. The design and selection of the proper corrosion inhibitor and right mitigation strategy are important for use in these systems.

The OPTIMA corrosion inhibitor described and characterised in this article, together with real time field application, has been developed to provide new solutions, and helps to improve the efficiency, reduce the cost, and meet the challenges of the carbon capture process. 

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CORROSION REPAIR

THOMAS KLINE, STRUCTURAL, USA, DISCUSSES THE BEST STRATEGIES FOR ANCHOR BOLT REPAIR IN PETROCHEMICAL ENVIRONMENTS.

An anchor bolt, by definition, is 'a metal bolt or stud, headed or threaded, either cast in place, grouted in place, or drilled into finished concrete, used to hold various structural members or embedments in the concrete, and to resist shear, tension, and vibration loadings from various sources, such as wind, machine vibration, etc'.¹ Much of the 'working' development length of the anchorage is hidden from view due to embedment into building construction materials, such as brick, stone, mortar, wood, and concrete. In industrial facilities, anchorage of process stream lines, vessels, and rotating/reciprocating equipment onto supporting foundations is handled by threaded metal rods embedded into conventionally

reinforced concrete. Unfortunately, the condition of the 'business end' of the anchor bolt is unknown until an owner/operator experiences an unfortunate event in the form of:

- Bolt tensile failure.
- Pullout (concrete tensile) failure.
- Lateral bursting (blowout) failure.
- Localised bearing failure.
- Concrete splitting failure.

It is important from a maintenance standpoint that unplanned maintenance events be monitored and mitigated, if possible. Industry best practices suggest that 'anchorage design should be approached as a global structural design



Figure 1. Deteriorated anchor bolt conditions including metal corrosion, lateral movement, and loss of protective cementitious grout.



Figure 2. Ultrasonic testing of anchor bolts for a tall industrial pressure vessel.

issue, focusing more on the development of ductile load resisting paths as opposed to the ductility of a single element.²

In petrochemical facilities, anchor bolts are used to fasten process vessels or machines to a foundation structure, thus allowing the loads to be transmitted to the foundation, and ultimately to the earth. As with concrete

infrastructure, all too often anchor bolts are taken for granted and the thought of providing a maintenance budget for these critical fasteners is more often an after thought, or a direct result of a failure. Anchor bolt failures, in some instances, can generate a process outage or a more serious accident involving personnel injuries, such as when conditions precipitate a 'failure without notice'. Corrosion under insulation (CUI) is often the culprit in anchor bolt failures, allowing insidious corrosion processes to accelerate unnoticed when the anchor bolts are hidden from view by insulation materials. Recognising CUI telltale signs, such as rust staining through insulation systems, increases in vibratory movement, and misalignment of equipment, can provide owners/operators with risk reduction opportunities when coupled with a thoughtful assessment strategy.

While there are many types of anchorages, this article will focus on anchor bolt types that are cast in place during original construction and deteriorate in service, thereby requiring an assessment prior to repair and restoration. Structure/process designers will vary in their cast in place anchor bolt preferences, but generally they will follow along two paths of thought, an 'L' or 'J' (i.e. hook) bar or a welded plate/fastened nut and threaded rod. The original prefabricated design can vary significantly between available alternatives. The real difference comes from ease in product

availability, especially in rapid turnaround situations where 'fabrication lead time' is at a premium or non-existent.

Understanding which of the anchor bolt types was installed is a critical feature in providing an accurate assessment while the anchor bolt is in service. Essentially, when tell tale signs generate concern by an owner/operator (i.e. surface corrosion, loose surface nuts, lateral/vertical thread displacement, severed bolts; Figure 1), means and methods are available to accurately evaluate in situ anchor bolt conditions. However, non-destructive and semi destructive testing requires some archival design/as built construction data for comparison purposes.

Once concerns have been raised, the implementation of a proactive evaluation program should be initiated that treats the anchor bolts as critical components of an important asset, requiring assessment. Several types of evaluation programs can be assembled which are generally a mix of non-destructive (NDT) and semidestructive testing (SDT) techniques that work concurrently to quickly assess the type, quality, and quantity of the tested anchor bolts. Essentially the testing techniques available include:

Acoustic impact

Each anchor bolt is struck with a 3 lb maul hammer and the audible resultant tonal variations are noted. For example, a sharp 'metallic ring' typically identifies a sound condition relative to the impacted anchor bolt, while a dull 'drummy' sound is indicative of an unsound anchor bolt, probably separated/broken along its length.

Ultrasonic metal flaw detection (ultrasonic)

The top of each anchor bolt is ground smooth and flush using an abrasive grinder, perpendicular to thread orientation (Figure 2). Using a piezoelectric transducer, applied to the ground portion of the anchor bolts' top surface, a mechanical stress wave is propagated (500 KHz - 10 MHz) whereby these waves will travel through a given metallic medium at a specific speed or velocity, in a predictable direction. When they encounter a boundary with a different medium (metal/concrete interface), they will be reflected or transmitted according to predetermined wave patterns applying simple rules of wave propagation. When compared to a stock anchor bolt of similar metallurgy, size, and configuration, detected wave propagation anomalies can be assessed, and further action taken, to provide 'truth and verification' and validation of in situ conditions.

Ground penetrating radar (GPR)

Depending on the geometry of embedded anchor bolts within the structural concrete member, imagery can be obtained using GPR. GPR emits an electromagnetic wave radiating from a transmitting antenna (900 - 1500 MHz), traveling through the testing material at a velocity that is determined primarily by 'permittivity' (permittivity relates to a material's ability to transmit, or 'permit', an electric field) of the material. The wave spreads out and travels inward until it hits an object that has different electrical properties from the surrounding medium, is scattered from the object, and is detected by a receiving antenna. The output is a graphical display that, with

Sulzer Technologies for Biofuel Applications



Background

In recent years, there have been many advancements in the development of biofuels for the global transportation markets. First generation fuel ethanol plants performed fermentation of grains into high purity ethanol. Further R&D led to the development of biofuels from renewable feed stocks: grains, bagasse, switch grass, sugar cane, beets, turnips and wood pulp. These second generation biofuel processes produce high purity methanol, propanol, butanol and other oxygenated hydrocarbons for this emerging market. The development of biofuels has included new technologies for mechanical and thermal processing of biomass, fermentation, and chemical separation processes.

Process Technology

Sulzer offers several technologies for second generation biofuels. Our process design experience includes pressure swing, azeotropic and extractive distillation to separate alcohols from aqueous feed streams. Membranes can be used to perform dehydration and overcome azeotropes to produce high purity alcohol products. The combination of these technologies leads to significantly lower energy demands. Our test center can verify the capabilities of novel processes with pilot plant studies that can then be applied to scale-up to demonstration and industrial plants. Process plant equipment can be delivered as skid-based solutions that include distillation columns and other process equipment, piping and controls.



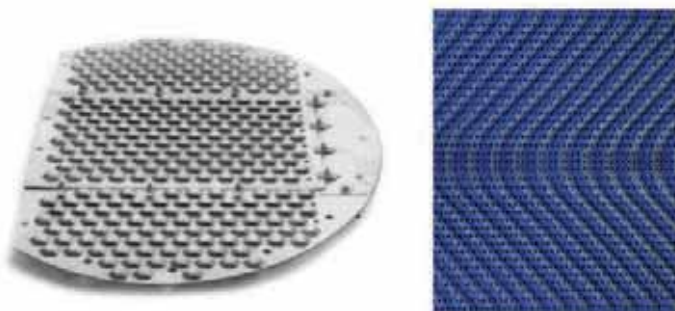
Membrane Module and Skid Mounted Column

Mass Transfer Equipment

Sulzer's product portfolio has been successfully used in biofuel applications for many years. In first generation fuel ethanol plants, VG AFT™ trays have a proven track record of resisting fouling by biomash solids and extending the plants' operating time. MellapakPlus™ structured packing offers low pressure drop and high capacity for distillation and absorption columns. The Kühni extraction column with its rotating agitation chambers that can be applied in liquid-liquid extraction applications. Vapor permeation with zeolite membranes is most suitable for bioethanol processes. These membranes can be incorporated into hybrid distillation systems used to dehydrate biofuels.

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Figure 3. Repair of corroded anchor bolts using coupling and welding techniques.



Figure 4. Excavation and inspection of corroded anchor bolts and anchor bolt chairs covered by a masonry brick and mortar veneer.

appropriate software, can provide an apparent 3D image of the anchor bolt embedment within the concrete.

X-ray

Using radiographic techniques, gamma radiation emitted through a medium from a radioactive source, such as a cobalt isotope, can provide anchor bolt imagery using a single side film attached to boundary surfaces. Due to the hazards associated with radiographic processes, this technique is rarely used in routine inspections, and large areas need to be cordoned off as a precautionary safety measure to eliminate accidental radiological exposures.

Vernier calipers

Exposing sub surface regions of embedded anchor bolts, the bolts are cleaned of laminar deposits of corrosion (i.e. rust pack) and a vernier caliper is used to determine the existing diameter of the exposed anchor bolt. The caliper operates as a graduated rule with one sliding 'jaw' and one that is stationary. The vernier is attached to the caliper and is used to provide fine measurements up to 0.0001 in.

Once an anchor bolt opportunity has arisen and the owner/operator has engaged the necessary expertise with testing results indicating the appropriateness of anchor bolt repair and/or replacement, several strategies are available to reestablish anchor bolt integrity. Anchor bolt restoration strategies can be as follows:

Do nothing

Numerically it may have been shown that the equipment/vessel requires only a portion of the hold down capacity; therefore, monitoring of the status of the remaining 'sound' anchor bolts is necessary until conditions change requiring a corrective action program.

Repair existing anchor bolts

Near surface separated anchor bolts can be repaired using mechanical couplings or welding stud extension. Both techniques remove and replace anchor bolt threaded regions

damaged by corrosion, lateral shear, or vertical 'cup and cone' failures. Properly installed and incorporating appropriate metallurgical detailing, repairs as described can add significant service life to the existing anchor bolt system, once the 'root cause' of the failure has been determined. Repairing a failed anchor bolt, without recognising the root cause failure mechanism, places owners/operators in the dubious position of 'repairing the repair' should the anchor bolt fail again in a similar manner.

Relocation of anchor bolts

When changes occur in original design and equipment/vessel loadings require new 'fastener' configurations, then anchor bolt chair assemblies can be designed and incorporated into existing anchorage 'schemes' that reposition fastening points. These new anchor bolt locations can be along the same circumference (as in curved/circular foundation support bearing systems) or along the perimeter (as in rectilinear configurations). Should following this type of placement be inapplicable, positions inward or outward may be considered as a viable alternative. Anchor bolt chairs are welded or bolted/riveted assemblies that are solidly affixed to the foundation support member/skirt just above the concrete foundation interface as shown in Figures 4 and 5. One of the major disadvantages in anchor bolt relocation in reinforced concrete members is the probable obstructions caused by cast in place rebar that may be present at the new proposed location. To reduce the possibility of cutting embedded reinforcing steel, the use of pachometers (i.e. Eddy-Current Rebar locators) or GPR is recommended to locate steel up to 19 in. in depth below the concrete's top surface. Once located, the embedded rebar can be marked and concrete coring operations can be started with the prospect of cutting embedded steel greatly minimised.

Full depth anchor bolt extraction and reinstallation

Many times the prospect of cutting embedded reinforcing steel bars is unacceptable, especially in those structures that are being upgraded by increased vessel loadings or compromised by 'optimised design' (i.e. slightly overstressed intentionally by design for cost savings). The very idea of compromising the structural integrity of the reinforcing systems can be a rehabilitation project 'non-starter'. The best alternative in many situations is the complete removal and replacement of the anchor bolts. In elevated structures, large equipment/vessels can have the anchor bolts extend through the entire reinforced concrete member thickness, as shown in Figure 6. The existing corroded/broken anchor bolts were forced through the grout sleeve annulus using modified jack hammers fitted with drill steel tips, and the existing anchor bolts were essentially driven/hammered out of position, either from the top down or from the bottom up. Subsequent to removal, anchor bolt hole sidewalls were cleaned, new threaded bolt studs were installed, washer/nuts were assembled, and the vessel was tightened to original equipment manufacturer (OEM) compliance. The resulting new anchor bolt annulus was grouted 'tight' for corrosion protection with a flowable, cementitious grout that typically incorporates corrosion inhibiting admixtures.



Figure 5. Corroded anchor bolt chair assemblies with bearing plates were only visible after insulation was removed from the vessel bearing skirt.

Unfortunately, many times, especially in very large massive structures, the anchor bolts terminate within the heavily reinforced concrete structural section. In situations such as these, the process known as 'over coring' is employed. This process places new anchor bolts at the same location as the original anchorages, but the new anchorages can be longer with a larger diameter and, therefore, can have a higher 'hold down' capacity than the original anchorage system.


Conclusion

Retrofit anchor bolt selection should only be performed by engineering professionals with the assistance of the manufacturer's technical service staff, both working toward providing a project with an appropriate system that can address the service environment and required loading. Installation of these retrofit anchor bolt systems should only be performed by contracting personnel knowledgeable and



Figure 6. Full depth corroded anchor bolts removed and new anchor bolts installed extending through a thick reinforced concrete structural slab.

specifically trained in the 'means and methods' for successful installation of these sophisticated anchorage systems.

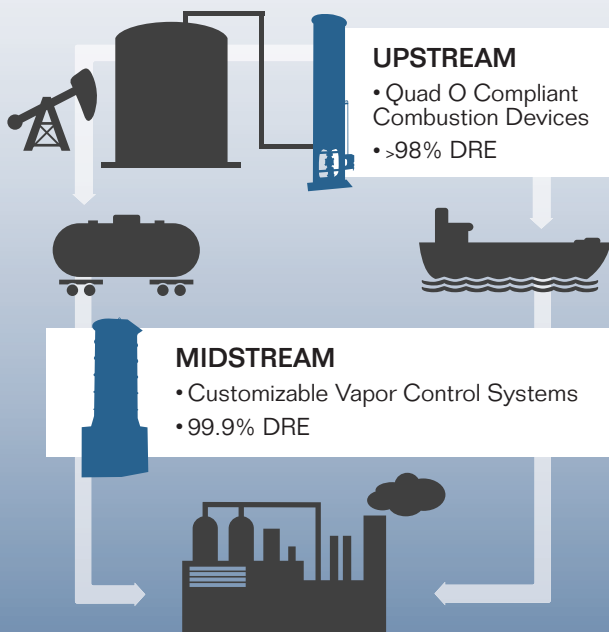
Understanding the current condition of a civil asset's anchorage is an important aspect for continued reliable service. Being proactive prior to failure assures unplanned 'events' are kept to a minimum, and allows for maintenance costs to be planned, controlled, and funded without placing owners/operators into a 'crisis' mode for restoration/repair. 

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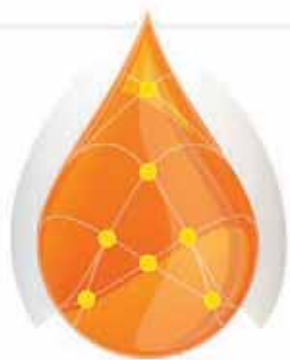
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Crude oil refinement is a very energy

intensive process. Steam and hot water are of great concern for processing. Their contamination leads to corrosion in equipment such as pipes, boilers and heat exchangers. The monitoring of these provides enormous cost savings. Therefore, an oil in water analyser is required, and in this case a total organic carbon (TOC) analyser is an excellent choice.

The production of steam and hot water is vital for the operation of many types of installations and facilities in crude oil refinement. High pressure steam is used for generating electrical power, for driving turbines, utility pumps and for process heating. Medium pressure steam is

Werner Arts, LAR Process Analysers AG, Germany, explores the potential of TOC analysers for cost saving and money making in refinery operations.

typically used for heat transfer to refinery products via heat exchangers and the low pressure steam is used for a variety of purposes e.g. pipe tracing, process heat exchange and deaerators. Although the temperature of the steam continues to decrease and it becomes hot condensate, it still contains a lot of energy and remains very valuable. Significant cost savings can be attained through the reuse of the so called return condensate or steam condensate. However, the purity of this return condensate is of vital importance and should be monitored by an online TOC analyser that measures any oil type in a range from 10 ppb to over 100 ppm together with organics that do not belong to oil groups. At refineries, the



Figure 1. Heat exchanger with plugged, corroded heat pipes.



Figure 2. Commissioning and startup of a QuickTOCpurity TOC analyser.

alarm thresholds are approximately 40 times cleaner than drinking water (100 to 1000 ppb range). Therefore, the analytical performance of the TOC analyser must be outstanding and tough demands are made by the industry regarding accuracy, response times and traceability.

Condensates' purity is vital

A specific oil refinery, with a site covering an area of approximately 4 miles², has many petrochemical facilities like crude oil distillation columns, reformers, crackers, hydrogen generation units, desulfurisation units, fuel oil and gasoline blending units in operation. Furthermore, five huge cogeneration plants are in operation.

Typically, a cogeneration plant consists of a demineralisation water plant and a steam boiler. Even traces of CO₂ can harm the boiler installation. Inside the boiler at the elevated temperature and pressure, CO₂ will be transformed to carbonic acid, a corrosive substance that causes the pit corrosion of metals. Dissolved oxygen, especially when it increases rapidly to a high temperature, will cause pit corrosion in the boiler system too. Oxygen pitting is a costly problem and it can be minimised when even traces of dissolved oxygen are removed from the makeup water using a deaerator. The oxygen solubility in water decreases tremendously at increasing water temperatures. The makeup

water flows through a deaerator where it is heated using steam, consequently purging out the oxygen. Any remaining traces of oxygen are removed by a chemical reaction through the introduction of oxygen scavengers. Additionally, a variety of other chemicals are dosed. Examples include amines to build up a protective film over the metal surfaces, antiscaling chemicals to maintain the heat distribution and antifoam chemicals to prevent the forming of foam in the boiler.

The costs involved in the production of pure makeup water are very high, and savings can be made by reusing the steam condensate. Every ton of steam that has passed numerous heat exchangers and is returned to a feed water holding tank at the boiler facility as return condensate, saves the production of one ton of demineralised water. Another lucrative aspect to the recycling of return condensate is that normally it is still hot, e.g. 90 °C to 120 °C, thus representing another large energy saving. However, the drawback of this well thought out cost saving system is the risk of importing contaminated steam condensate.

The steam's energy is transferred to process fluids e.g. crude, mineral oils and other products by use of heat exchangers. A common type is the pipe bundle heat exchanger and, in a refinery, hundreds of these heat exchangers are used. Due to the high temperature of the steam and the presence of traces of oxygen that have penetrated into the plant's steam piping system, these heat exchangers are also subject to pit corrosion. Despite the associated inspections and maintenance intervals, there is always a risk of a breakthrough, leading to pure products e.g. hydrocarbons, crude oil, petrol, kerosene and lube oils leaking into the return condensate. This contaminated return condensate can flow into the holding tank, finally entering the boiler, where it could lead to disastrous metal corrosion. However, it is possible to monitor the quality of the return condensate using a quick online TOC analyser and to drain it off as soon as a trip alarm TOC level is exceeded. Here, it is important to note that all organics will harm the boiler, and that non-oily types of organics, just as any other type of organic, the smallest amounts of hydrocarbons or oil traces are measured by the TOC analyser.

A TOC analyser system should respond quickly to a breakthrough of an oily matter, triggering an alarm as fast as possible, to allow the prevention of oil from entering and fouling the installations. This early warning system activates a process valve with any abnormal TOC increase, allowing the contaminated condensate to be drained off and the leaking heat exchanger to be located and repaired.

Requirements of a TOC analyser within a refinery

Generally, a TOC analyser monitoring system should measure accurately, giving precise TOC measurements at a ppb level, respond quickly, and present the result within minutes. The TOC levels to be monitored are in the low ppm or even ppb region. The hydrocarbon molecules have a tendency to stick to the wetted parts of the sample transfer pipes and analyser parts (adsorption). After a while, saturation takes place and a more or less stable equilibration level is reached, however, minutes later these absorbed molecules can become free, so called desorption. Not only are the analytical TOC results influenced by this cross contamination, but also the memory

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effects can cause the entire analyser system to operate very slowly. The extent of these phenomena depends on the type of material of the wetted parts, the flushing velocity, the flushing time, the temperature of the fluid and, more specifically, the type of organic matter. Oily matter in particular is subject to adsorption and desorption. This all becomes especially challenging when a single TOC analyser is used to monitor several sample streams. The multistream TOC analyser should rapidly and sequentially measure samples with different TOC levels and different compositions. However, cross contamination and carry over effects are then hard to avoid. Nevertheless, the TOC analyser must be able to rapidly return to accurate TOC values, even after having been exposed for hours to a highly oily mixture. Many TOC analysers



Figure 3. Cylinder with calibration gas.



Figure 4. QuickTOCpurity in a NEMA 4X housing.

continue to suffer from severe memory effects because of the pollution of the wetted parts by the sticky oil. As a consequence, these analysers show very high measurements over long periods, during which plant operators keep flushing expensive pure condensate down the drain.

Another potential problem to be faced is the fact that the TOC analyser is placed in the centre of a refinery. The CO₂ and volatile organic carbon (VOC) content of the ambient air here is often at a higher level than normal and these gases should not be able to influence, penetrate, permeate or leak into the analyser system or the samples.

Besides all this, there is a need to be able to calibrate or validate the TOC analyser quickly and easily. Usually, this is performed by the use of wet chemical standards, using an expensive water purifier producing 5 ltrs./d of ultra pure water with a TOC purity of 10 ppb. Under stringent circumstances and using a considered time frame, these standards can be prepared. But as soon as they are ready, they are subject to degradation and contamination, as they can only be stored at ideally cooled conditions in totally filled, sealed bottles for a few days at a time. When a TOC analyser measures a TOC peak, it becomes suspect to the plant operator, as they are not happy to drain their expensive return condensate. In this case, the analyser engineer must prove immediately that his TOC analyser is taking correct measurements. An ad hoc certified standard is needed. However, it is not possible to prepare such a standard in an instant.

Outstanding TOC analyser

A TOC analyser consists mainly of sample transfer and injection system, reactor, carrier gas circuit and gas detector and calibration system. At the heart of the QuickTOCpurity TOC analyser is the well designed multi loop injection system as the preeminent methodology for sample transference and injection of oily samples. During a programmed determined time period, all parts, including the sample injection loop, are flushed with a fresh sample. Then a precise volume of the sample is blocked inside the injection tube. Seconds later, the content of the sample injection loop is pushed by carrier gas via the injection port into the 1200 °C reactor. Normally, one single loop injection is sufficient to achieve a high resolution detector signal. However, for the lower ppb TOC range, the sample loop can be successively injected multiple times. This loop injection system is constructed of inert material to avoid absorption and memory effects and can resist temperatures of up to 90 °C.

The second most important analyser building block is the high temperature reactor. In principle, it is based on non-catalytic thermal oxidation at 1200 °C. The sample, containing hydrocarbon molecules, is injected into this reactor tube. At 1200 °C, it is physically impossible for compounds to survive. Experience demonstrates a full recovery rate even for very hard to oxidise compounds such as fluorocarbons, chlorinated hydrocarbons and amines. Without exception, all hydrocarbon compounds are completely oxidised, thereby forming CO₂ gas. A carrier gas flows through the reactor and transfers this CO₂ gas to a non-dispersive infrared CO₂ gas detector (NDIR CO₂ detector). The amount of carbon dioxide that is measured is linearly proportional to the amount of organic carbon molecules (TOC) in the sample. This shows that


LAR's ultra high oxidation technique is free from memory effects, giving excellent low TC and TOC level performances in comparison to other oxidation techniques, such as 800 – 900 °C catalytic oxidation systems, which do suffer from such phenomena. Another disadvantage of catalytic combustion systems is the 'blind' value phenomenon which means that beyond a certain point in the lower ppb range values cannot be measured due to the catalytic memory effect, peak tailing and a shifting of the CO₂ baseline.

The QuickTOCpurity uses a smartly designed carrier gas system. Its flow circuit is kept at a stable overpressure of 0.4 bar. This design eliminates the risk of any ambient air interferences, which occur frequently at refineries.

The third most important analyser building block is the calibration and validation system. The LAR QuickTOC analysers offer a unique patented validation feature. Whenever needed, the operator can easily and immediately calibrate or validate the analyser by using a cylinder filled with a certified gas mixture. This calibration gas, such as a methane and nitrogen mixture, is very stable and can be stored over a period of several years, making it readily available and ensuring that it is always ready for use. The injection loop is filled with the gas and injected into the reactor. There, the gas is oxidised to CO₂ and the NDIR CO₂ detector then measures the amount. Within 10 mins the calibration and validation routine is complete.

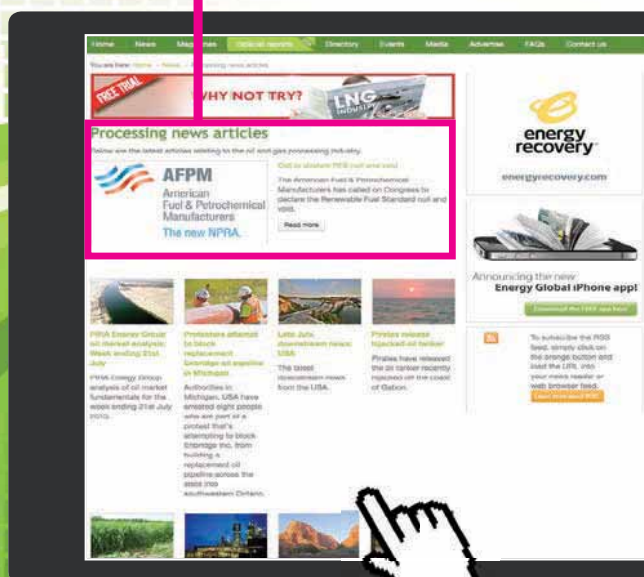
Another important feature of the LAR QuickTOC for pure water is the possibility to analyse grab samples from individual heat exchangers. This can save a lot of time in identifying which specific heat exchanger is leaking.

Cash savings

Money is to be saved not just through cutting energy costs in reusing return condensate alone, but also by the effective analysis of its purity. The TOC analyser used needs to be fast and reliable and the QuickTOCpurity analyser is just that. Its patented thermal oxidation technique guarantees that all hydrocarbon compounds are completely oxidised without exception, and offers the added saving of this technique negating the need to use expensive catalysts, as with other catalytic oxidation analysers. The totally sealed system, maintained at an overpressure of 0.4 bar, means that only the sample is measured without any environmental interferences influencing the measurements. Furthermore, the well designed multi loop injection system has reduced internal dimensions and minimised wetted surfaces, as well as no dead volumes, meaning that memory effects are avoided. The quick and easy calibration and validation of the LAR analysers offers operators the assurance that their analysers are giving accurate results and that they are not pouring expensive condensate down the drain. 



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TANK CLEANING COST STRUCTURES

**SØREN HUBERT PETERSEN AND ALEKSANDRS VDOVINS, ORECO A/S, DENMARK,
DISCUSS COST STRUCTURES IN STORAGE TANK CLEANING AND IDENTIFY THE
MOST COST EFFICIENT WAY OF PERFORMING TANK CLEANING JOBS.**

The operation of a tank farm, whether inside a refinery or at a terminal, is ultimately driven by the economy and profitability of a business. Crude oil or refined products must move as fast as customers demand.

Delays in product availability or delivery can mean losing a contract to a competitor. Equally important, product quality targets must be met in order to avoid claims or penalties. And the entire operation must not compromise health, safety and environment (HSE) standards.

Storage tank cleaning is an integral part of the operation in the oil and petrochemical industries. The cleaning of oil and petrochemical tanks is required prior to periodic inspection of and subsequent maintenance work in the tanks. Particularly in the case of larger tanks containing crude oil or other 'black oil' types such as residue and heavy gas oils, the cleaning of these tanks is a major undertaking that requires a great deal of resources in terms of planning, manpower, time and money. Moreover, the increasingly stringent legislation and regulations regarding the environment, workers' health and safety have placed additional pressures on the oil and petrochemical industries to utilise safer and more environmentally friendly tank cleaning methods. The main known tank cleaning methods can be classified into two groups: manual

and fully automated approaches. Semi automated methods are included in the manual group.

Tank cleaning costs structure

Efficient and profitable plant or tank farm operation depends on a cost effective, timely, fast, efficient and safe maintenance practice. Therefore, the economic aspects in oil tank cleaning are of significant importance for any tank owner and operator. Typically, costs or investments can reach hundreds of thousands of US dollars or euros. The tank cleaning project contains multiple tasks, including desludging, water wash, hydrocarbons recovery from bottom residuals and waste disposal. Tank cleaning costs can be classified into two overall groups: direct and indirect costs as depicted in Figure 1.

Manual tank cleaning is normally perceived as having the lowest cost compared to automatic non-man entry methods. However, this is only true if you look at the immediately 'stripped naked' costs of a cleaning job, which is the contractor cost and cost of utilities. The fact is that there are a number of other indirect costs associated with tank cleaning that are rarely accounted for. If one instead takes a 360° view on tank cleaning costs and include indirect costs as well, the picture changes

Table 1. Selected tank's specification

Parameter	Value
Tank type	Above ground storage tank
Product type	Crude oil
Diameter	60 m
Roof type	Single floating roof
Volume	50 000 m ³
Sludge volume	1000 m ³ /6290 bbls
Region	Europe

**Figure 1.** The tank cleaning cost structure.

dramatically. The question is how much. A detailed review of the costs structure is performed in the next section and the conclusions will undoubtedly challenge conventional thinking.

Tank cleaning cost comparison

For the purpose of making a comparative analysis between manual and automated tank cleaning costs the author has selected a typical crude oil tank (Table 1) and examined each of the direct and indirect cost drivers.

Direct costs

Cleaning cost from contractor

Service providers that offer manual tank cleaning services propose a price that is approximately 2 - 3 times lower than companies that are operating automated systems. The price difference may even exceed € 200 000 - 300 000.

Utilities cost

For manual tank cleaning methods a limited amount of consumables is required. In some cases only steam is used. When it comes to automated methods, a tank with the specified dimensions and content will typically require approximately 15 000 m³ of nitrogen, 1000 t of steam and 110 000 kW/h of electricity. The price for utilities depends on the source of supply (central networks, refinery produced, or portable generators), but usually automated tank cleaning is more expensive than manual and the difference ranges from € 20 000 - 90 000.

The examination of direct costs clearly shows that an automated approach is far more costly than a manual approach. However, when one considers all the indirect aspects the picture changes considerably.

Indirect costs

Tank outage time

To clean a large crude oil tank manually is much more time consuming compared to automated methods. A 50 000 m³ tank can be cleaned manually in a period of 3 - 6 months. This is 3 - 6

times longer than required for automated processes.

A rule of thumb of how to calculate the cost of taking a tank out of service was made by a Dutch consultancy firm back in the 1990s. They estimated that it costs US \$ 0.25/bbl tank capacity per month. Their method is still valid today, and when considering an annual inflation rate of 3 %, the present

value of tank downtime is US\$ 0.32/bbl tank capacity per month. A rough estimation of the cost of having a 50 000 m³ (315 000 bbl) tank out of service is € 100 000 (2013 figures). This clearly demonstrates that the tank cleaning time is a crucial cost driver in favour of automated methods.

Loss of hydrocarbons

Accumulated hydrocarbons from the oily sludge represent a significant value of resaleable hydrocarbons. Sludge composition for typical crude oil tanks is summarised in Table 2.

The value of recovered hydrocarbons from the sludge can amount to approximately 30 - 60% of the oil market price. During manual tank cleaning the sludge is mechanically removed to centralised facilities or to municipal reprocessing plants. Recovered product volume can be between 5 - 30%, so 70 - 95% of hydrocarbons values are lost. Application of an automated tank cleaning method allows for recovery rates of up to 95% of the hydrocarbons. For evaluating loss/recovery economic value one can present a formula where the sludge amount is multiplied with the oil, recovery rates and market value.

Value of recovered oil for automated tank cleaning:

$$\begin{aligned} \text{ROV} &= V_s \cdot o_c \cdot r \cdot \text{OP} \cdot p_c \cdot c_c \\ &= 6290 \cdot 0.9 \cdot 0.95 \cdot 102.43 \cdot 0.4 \cdot 0.771 \\ &= \text{€ } 169\,888.28 \end{aligned}$$

Value of recovered oil for manual tank cleaning:

$$\begin{aligned} \text{ROV} &= V_s \cdot o_c \cdot r \cdot \text{OP} \cdot p_c \cdot c_c \\ &= 6290 \cdot 0.9 \cdot 0.15 \cdot 102.43 \cdot 0.4 \cdot 0.771 \\ &= \text{€ } 26\,824.15 \end{aligned}$$

- ROV: Recovered oil value for automated tank cleaning (€).
- V_s : Accumulated sludge volume, bbls.
- o_c : Oil contained in the sludge, %.
- r : Oil recovery rate (for automated tank cleaning 95% and for manual tank cleaning 15%).
- OP: Oil market price/bbl, US\$ 102.43 Brent (30/05/2013).
- p_c : Recovered oil value in relation with oil market price.
- c_c : US\$ to € currency transfer coefficient, 0.771 (30/05/2013).

If one assumes that the recovery rate for automated tank cleaning equals 95%, and for manual methods 15%, the result is 170 000 and € 27 000 worth of resaleable hydrocarbons respectively. Despite the rather academic approach, the results speak for themselves.

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Table 2. Tank bottom sludge, typical composition

Parameter	Value
Oil	50 - 95%
Water	1 - 15%
Solids	2 - 25%

cleaning methods will be more than 10 times less compared to manual tank cleaning methods.

Water disposal

The required amount of water needed to be disposed for automated tank cleaning and manual tank cleaning are approximately the same.

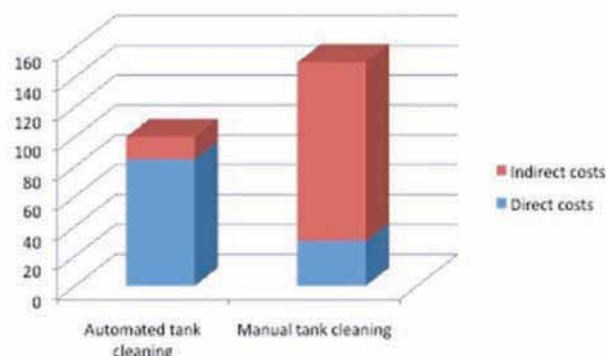


Figure 2. Direct and indirect cost structure for tank specified in Table 1.

Solid waste handling

Waste disposal becomes an increasingly expensive procedure because of stricter legislation and increased control from environmental authorities. The price for reprocessing 1 t of sludge ranges between € 20 – 350 depending on region and composition. Hence, price for sludge reprocessing after automated tank cleaning application will be significantly lower than after manual due to smaller amount, higher dryness and lower oil content. For sludge reprocessing cost calculation, one multiplies sludge volume by density factor, remaining percentage of solids, and price for utilisation/ t.

Solids disposal price for automated tank cleaning:

$$\begin{aligned}
 \text{SDP} &= V_s \cdot D \cdot S \cdot P \\
 &= 1000 \cdot 2 \cdot 0.1 \cdot 60 \\
 &= \text{€ } 12\,000
 \end{aligned}$$

Sludge disposal price for manual tank cleaning:

$$\begin{aligned}
 \text{SDP} &= V_s \cdot D \cdot S \cdot P \\
 &= 1000 \cdot 2 \cdot 0.85 \cdot 80 \\
 &= \text{€ } 136\,000
 \end{aligned}$$

- SDP: Solids disposal price (€).
- V_s : Sludge volume, m³
- D : Solids density, 2000 g/ltr, typically in range between 1500 - 3000 g/ltr.
- S : Coefficient of solids for disposal from total sludge amount after automated tank cleaning, 10% for automated tank cleaning and 85% for manual tank cleaning.
- P : Price for solids disposal, 60 €/m³.

Consequently, the price difference in solids between automated and manual tank cleaning methods is € 58 - 280. In fact, the cost of sludge reprocessing using automated tank

HSE fines, workers' safety, process safety

According to worldwide accident statistics human failure is a cause of approximately 80% of accidents in the oil and gas industry. Manual tank cleaning requires human presence inside the crude oil tank. First of all it is associated with high risks for personal health, including diseases, injuries, high risk of development of cancer, and even death. For manual tank cleaning jobs low qualified labour is usually used, which may increase the risk of failure or accidents. High hydrocarbons emissions into the atmosphere and soil pollution with hydrocarbons are common for manual tank cleaning and are often subject to fines and penalties. Automated non-man entry closed loop system ensures safe and environmentally friendly tank cleaning procedures with significant reduction of environmental payments and improved image of the company.


Summary

Direct and indirect costs structure for the tank specified in Table 1 for manual and automated methods is summarised graphically (Figure 2). Automated tank cleaning expenses are considered as 100%.

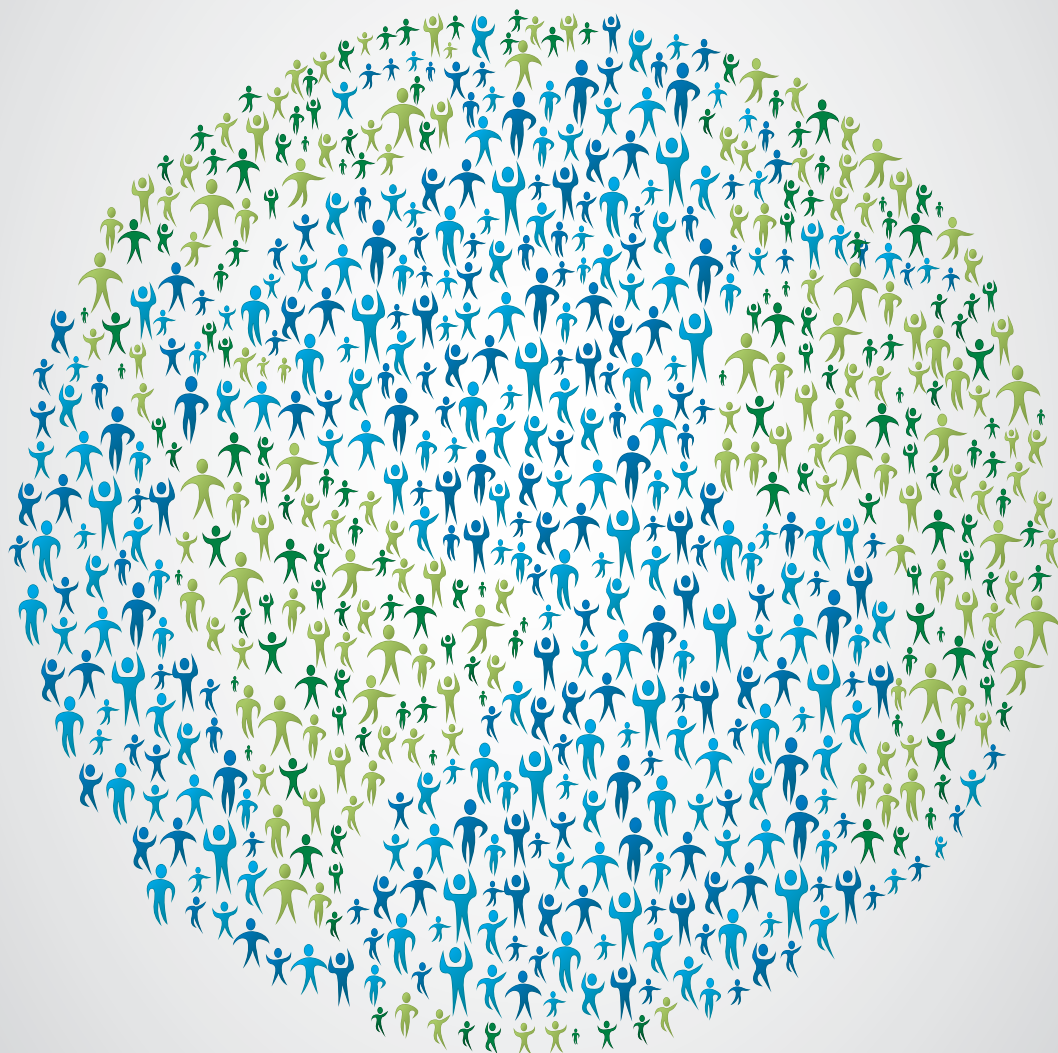
The tank cleaning time and hydrocarbons recovery rate form the biggest part of indirect expenses and they change according to the tank size and accumulated sludge volume. Accumulated sludge volume can seriously affect tank cleaning time and amount of recovered hydrocarbons. Crude oil tanks can contain several metres of bottom sludge accumulations. In general, automated tank cleaning becomes more financially beneficial the larger the tank is, and the more accumulated sludge volume it contains. Also, the market price of oil can seriously affect tank cleaning costs. When increases in oil prices occur, automated tank cleaning methods become more profitable. Growing environmental payments influence costs structure as well. Due to the above factors, cost benefit analyses should be made individually for specific cases and constantly updated with actual market prices.

Conclusion

For large crude oil and heavy fuel oil tanks automated tank cleaning methods with simultaneous hydrocarbons recovery is evidently the most efficient and financially beneficial option. For smaller tanks the cost of automated tank cleaning methods are more or less equal to those of manual methods, but because of additional value in accordance to HSE and total process safety should be recommended as a standard.

So what is the message from this exercise? Despite the theoretical nature of the analysis, the outcome is unmistakable. Contrary to conventional thinking, automated tank cleaning is by far the most cost efficient. One cannot look at the contractor costs in isolation as tank cleaning influences, and is influenced by, several indirect cost drivers. 

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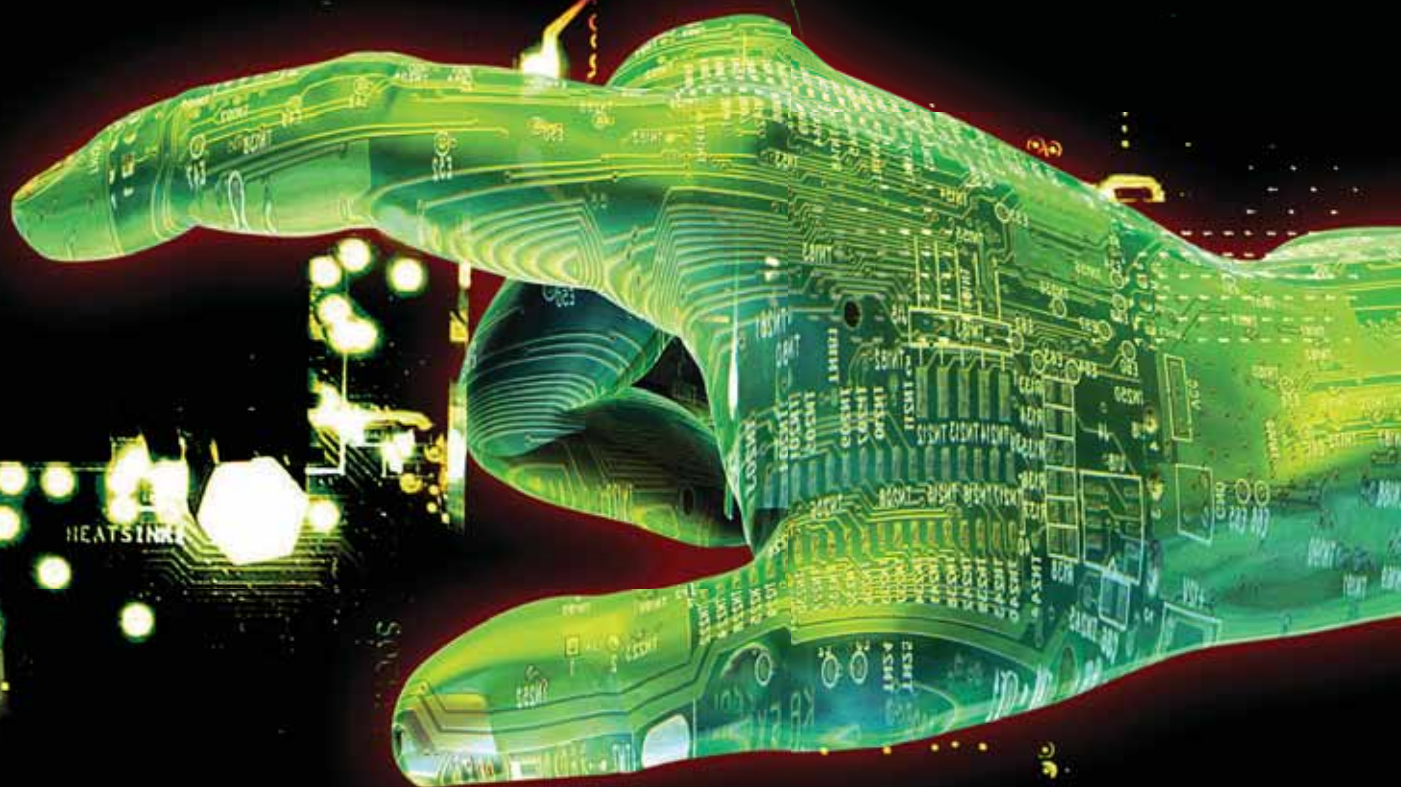


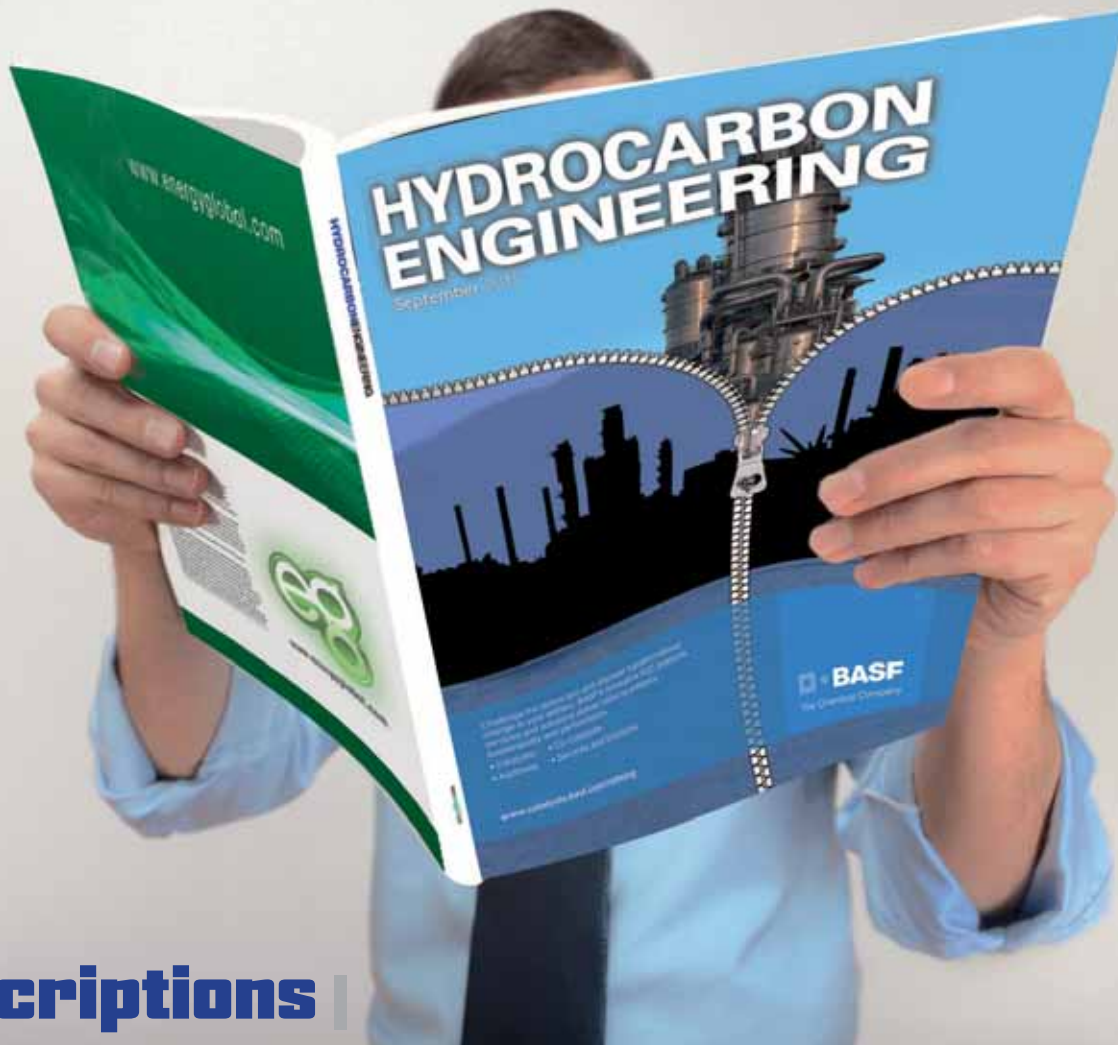
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THE BEST OF BOTH WORLDS

As business needs evolve, the increasing demand to integrate data from real time process control and business systems (such as production planning, yields, quality, energy consumption and emissions) has resulted in more 'dynamic business production'. While this integration of the corporate network domains and process control domains better enables asset owners to adjust output to meet both demand and capitalise on market change, it also increases the risk of cross contamination.

If a process control network is still separate from a corporate network and the plant operator expects it to stay that way, then risk of cross contamination is relatively low. Yet if the operator seeks greater



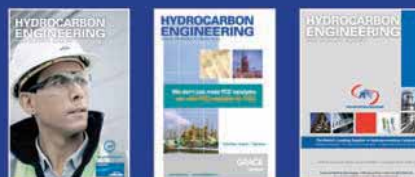


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connectivity between the corporate and process control domains then this calls for greater collaboration among those that protect the process control and engineering domains. Those that protect the business domains and IT infrastructure should also be of great interest.

Now more than ever, with a number of high profile incidents of cyber security breaches focusing on the automation layer, the modern refinery needs to establish a closed loop when it comes to safety. Bringing functional safety and cyber security together is integral to creating that closed loop.

The human factor

Risk assessment is still reliant on humans. When examining the rise in recent attacks, the Stuxnet virus is probably the most well known due to publicity, in depth reporting and the fact that it specifically targeted process automation systems. Crucially, it also incurred human weakness and error. The virus was introduced via a USB device that was left where employees would find it. The persons finding such a device most likely did not consider the risk, impact and consequence of using the device in the process control domain.

This threat was created outside the plant and designed to cause disruption to the routine running of processes. The virus was designed with the specific knowledge of the protocols used on the process control network, enabling it to wreak havoc. It would be interesting to know if any pre HAZOP or risk and threat assessment had considered this type of attack when the systems were designed and implemented.

Another recent attack, this one by Shamoon malware, consisted of three elements:

- Reporting.
- Over writing drivers, programs or libraries.
- Over writing the master boot record and instructing the machine to reboot, rendering it useless.

Had cyber security been considered during the functional safety assessment then the consequences and impact may have been assessed and understood, driving the relevant risk reduction measures to be implemented. Potential safety measures that could have been taken include disabling USB ports, as well as ensuring that policies and procedures are implemented and personnel are trained on the risks of using rogue USB devices.

Converging protection models

Closing the safety loop requires departments to share responsibility for the safety, security or risk and threat assessment. For years, the IT industry has been meticulous in carrying out risk and threat assessments from inception and design, through to deployment and implementation of IT networks. These assessments have been carried out in accordance with a plethora of standards, including BS7799, BS17799, ISO27001-27005 and ISA99.

In parallel, the process industries have been carrying out similar activities to identify and evaluate potential risks to personnel, equipment and the environment, including structured and systematic safety and risk and threat assessments; HAZOP analysis; layer of protection analysis (LOPAs); safety integrity level (SIL) determination and validation, etc. These functional safety activities are conducted in accordance with their

pertinent industry standards, such as IEC61511, ANSI/ISA S84.00.001, etc.

As today's threats become more and more malicious, with sights set on automation systems, their continued evolution suggests that plants should battle them with the combined force of both the IT cyber security approach and engineering functional safety approach.

It is no longer adequate to consider that each possible hazard could come just from equipment failures, fires, floods or other events within the plant or refinery. Hazards can now be initiated from outside of the plant, and some of these would never even be considered if viewed from just an engineering perspective. Even if they had been considered, the analysis could already be out of date because the threat to any system is not a constant; it is a continuous evolution.

The functional safety assessments typically focus on the failure of a piece of equipment, addressing the probability of failure, the potential consequences and safety, environmental and commercial impacts. The IT assessments are very similar, but the consequences of a system being compromised would more likely be the massive economic impact of a production interruption, rather than loss of life.

Following corporate IT practices, tools have been created for use on process control engineering networks that scan the system for details such as asset identification, protocols in use, operating system status, version levels, patch levels and service pack installation. If not applied with cyber security concerns in mind, however, these same tools may provide potential hackers with intelligence that was not previously accessible to them, by the very nature of the network design. For this reason, a combination of skill sets or even departments should be employed in all aspects of safety from the field device to the corporate firewall that connects to the internet.

Assessing vulnerability

Assessments should be carried out on the use of protocols down to the field device level, and should ask which are vulnerable to attacks over an internet protocol network. Assessment should continue to the fieldbus module (FBM) or any equivalent device that converts field device output for compatibility with the distributed control system (DCS). A multi skilled assessment team must determine, for example, whether a compromised FBM effectively controls or interrupts the field devices. Making such determinations requires an understanding of many different technologies, including the protocols, the network on which they reside, the switches, the cable traces and the power supply.

If a network switch connecting field infrastructure to the process control domain were to lose power or be otherwise compromised, for example, it is important to question what the consequences would be. Redundancy is not always the answer, since the design could make redundant switches available to a virus attacking one switch.

Another common attack is when a hacker ports forward process control commands from one network port/route to another. This is used to exploit data from switches. If this attacker is able to reroute command and control data from one port or network to another, then this poses an additional problem.



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After assessing field protocols, network components and ports, the next in line would be the process control domain (PCD) or DCS. The days of DCS network infrastructures being managed solely by the instrumentation and controls engineers are waning fast. These networks are now subject to malware, viruses and common trends through threat analysis. Therefore it is the IT department that has the necessary skill set and knowledge to ascertain the damage that can be done to the network should miscreant programs enter it. Of course the engineering department still has to assess the damage caused should an IT based attack take down a particular device.

Further, the change in the process control domain is traditionally slow compared with that of a traditional IT system. IT systems have been responding to outside interference since the 1970s and have evolved with a tactical mentality in their approach to security, whereas process control took a more strategic approach. Only now are they becoming more tactical, making it essential that both IT and engineering skills and practices be combined in the assessment of today's plant and refinery risk.


Modernising plants

Today's plant and refinery operators are facing new, more complicated challenges as their equipment and workforce continue to age. In order to comply with increasingly complex regulatory guidelines, minimise downtime and maintain safe, environmentally friendly operations, many companies are seeking to modernise their plants with a holistic view of their business requirements instead of making like for like equipment replacements.

A holistic approach to modernisation can provide the clearest, most cost effective pathway to improving plant efficiencies, increasing production rates and sustaining profitability. The Invensys modernisation programme delivers full scope consulting, project management, engineering, installation and maintenance services. It also covers products and solutions that address all operational areas of the plant, including instrumentation, input/output and human machine interface (HMI), safety and critical control systems, turbomachinery assets, process safety lifecycle components, cyber security systems and other cross enterprise assets.

Integrating practices and procedures

In a modern refinery, simply having IT and engineering groups in communication is not enough. Effective collaboration requires a close analysis of the practices and procedures for both departments to see if there are any contradictions. Synergy is good, but any contradictions could be a potential weakness in the system.

The time has come for refineries to combine the best of the IT world and the functional safety world. The next time that a HAZOP is performed, operators should consider not just the process hazards, but also the IT hazards, consequences and impact. The sooner the common divide that exists between the safety and cyber security protection functions can be eradicated, the better the plant environment will be, from both a welfare and financial perspective. The best way to do that is to take, learn and use the best from both worlds. 

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questions with...



Keith Thomas, Vice President Project and Business Development Gas, E.ON New Build and Technology GmbH; Chairman, Gas Processors Association Europe.

Having completed a degree in chemical engineering, Keith has worked for 33 years in the gas industry from package equipment manufacturers through being an independent consultant to heading up E.ON's project development for its natural gas activities. Keith was made Chairman of GPA Europe in November 2012 and has been a die hard Crystal Palace fan for 40 years (therefore, an eternal optimist).

1|Describe your GPAE position in 6 words? Chairman, keeping order and giving direction.

2|What is your favourite thing about the world of gas? Continuous change.

3|Why? The market is always driving technology in new directions. Shale gas and LNG are just the last two major trends. And the value chain is so diverse the variation in applications is enormous. If you are an engineer there is no better or more challenging place to be at the moment.

4|What's your favourite book? Catch 22

5|How has GPAE changed since you became involved? In the late 1980s the major oil companies were far more dominant. Now there is a mix of oil companies, specialised gas companies and utilities as well as suppliers. Ultimately the GPAE has mirrored the market and become more diversified.

6|What was your first pet? A goldfish - won at a fun fair.

7|If you could go back in time, what period would you go to? As a rule I prefer not to look backwards, but the Renaissance was culturally a very interesting period.

8|What was your first job? Doing a paper round at age 13.

9|What advice would you give people looking to enter the oil and gas industry? Be flexible and be forward thinking. The industry will only survive by adjusting to the challenges, of

which there have been and will still be many. But challenges are also opportunities, which makes being in the gas business exciting.

10|What industry events (other than GPAE's obviously) do you attend? Gastech and possibly events linked to the geographical area we are working in at any time.

11|Do you use social media as part of your working life? No, I still prefer face to face contact.

12|What is your golden rule? Think first, then act.

13|What is your biggest achievement? Professionally the next project we finish. Closing out every project is an achievement in itself.

14|What would you say is the highlight of the upcoming GPAE Annual Meeting? This year the Scottish Minister for Energy will be making a presentation, and we will be having a discussion on how natural gas fits into the European energy mix, from a political, academic and industrial perspective. It is an interesting excursion from our traditional technically based meeting but highly topical at the moment.

15|It's GPAE's 30th Annual Meeting, what do you see happening to the GPAE by the 60th? I wish I knew. I would not even like to hazard a guess to the 40th meeting. The pace of change is much faster now but I would expect gas to play a more dominant role in the energy mix with renewable energy worldwide and ultimately replace coal as the primary fossil fuel. But we are a long way from that at the moment, and as long as the real price of coal is not reflected through the carbon certificate scheme it may never happen.

The word "Impulse" is written in a large, white, sans-serif font, tilted diagonally upwards from left to right. The letters have a subtle blue glow. Several stylized white lightning bolts with blue outlines are positioned around the text, some appearing to strike the letters. The background is a dark blue grid with diagonal lines of varying shades of blue, creating a sense of depth and energy. A thick red curved line separates the top graphic area from the bottom text area.

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