HYDROCARBON PROCESSING®

HydrocarbonProcessing.com | SEPTEMBER 2013

REFINING DEVELOPMENTS
Shale oils and heavy feedstocks need new processing schemes to make 'clean' transportation fuels

VIEWPOINTS
Outlook for renewable fuels and the refining industry's future, and how GTL will fit into the HPI

PROCESS CONTROL
Innovative, advanced control system radically improves operations and yields

HYDROCARBON PROCESSING®

SEPTEMBER 2013 | Volume 92 Number 9
HydrocarbonProcessing.com

SPECIAL REPORT: REFINING DEVELOPMENTS

ENGINEERING AND CONSTRUCTION—SUPPLEMENT

PROCESS CONTROL

DEPARTMENTS

COLUMNS

Cover Image: BP is nearing completion of its Whiting refinery modernization project. In July 2013, the refiner announced the commissioning of the upgraded distillation unit. This refinery is located outside of Whiting, Indiana. For the full story, see Construction Profile, p. 33. Photo courtesy of BP Products North America.

HP Industry Perspectives

A sad turn of events

Many pages in *Hydrocarbon Processing* have and continue to discuss the technical solutions required by the US Renewable Fuel Standard (RFS) and similar legislation. At a time when developed nations struggle to sustain economic recovery, government and regulatory agencies continue to build roadblocks, crippling major industries. For the US refining industry, the task to remain viable is a struggle every day. In short, petroleum and refined products are under fire constantly.

On July 23, representatives of various stakeholders for the US refining, automobile, biofuels and transportation industries testified before the US House of Representatives and Commerce Committee-Subcommittee on Energy and Power regarding RFS. Testifying before the committee were the American Petroleum Institute (API), American Fuels and Petrochemical Manufacturers (AFPM), National Biodiesel Board and Renewable Fuels Association (RFA). All parties have significant interests on the interpretation of RFS and RFS2 going forward. In reviewing the testimony, several key points stand out:

1. The RFS and later versions are designed to reduce oil consumption. This was a key factor in the passage of these rules. For some time, the US and other nations were highly dependent on oil imports, and tremendous fears existed over the possibility of oil supply disruptions by "unfriendly" parties. Such events would devastate national economies. Fear levels were elevated following the 911 attacks on the US in 2001. The emphasis at that time was on energy security and independence from imports. However, conditions have changed since the passage of the 2005 law and RFS. The largest reshaping factor is the drastic decline in US gasoline consumption resulting from the 2008 economic recession (depression). The US and other developed nations continue to consume less gasoline. US gasoline consumption has declined from 9.3 million bpd (MMbpd) in 2007 to under 8.5 in 2012. With changes in CAFE requirements for later model automobiles and light-duty trucks, gasoline demand will decrease even further. The US is likely to consume less crude going forward.

2. RFS was geared to increase usage of biofuels in the transportation fuel mix. The initial RFS plan was to start with food-based materials for biofuel production and then progress to nonfood-based biofuels such as cellulosic and biomass materials. These are great goals; however, implementation of nonfood-based biofuels has its share of problems. Large-scale commercialization of cellulosic fuel production is much slower than anticipated. Equally important, as gasoline demand declines, refiners have fewer options to comply with the mandates on renewable fuel blending. API and AFPM testified at the hearing that the US refining industry is under great pressure to comply with RFS and needs more flexibility.

3. RFS is intended to reduce greenhouse gas (GHG) emissions. API president, Jack Gerald testified that the US is making substantial reductions in GHG emissions with new technology and innovation. Energy efficiency measures return great benefits, especially in lower energy bills and reduced GHG emissions.

HYDROCARBON PROCESSING®
www.HydrocarbonProcessing.com

P. O. Box 2608
Houston, Texas 77252-2608, USA
Phone: +1 (713) 529-4301
Fax: +1 (713) 520-4433
Editorial@HydrocarbonProcessing.com

PUBLISHER
Bret Ronk
Bret.Ronk@GulfPub.com

EDITORIAL

Editor	Stephany Romanow
Managing Editor	Adrienne Blume
Reliability/Equipment Editor	Heinz P. Bloch
Technical Editor	Billy Thinnes
Online Editor	Ben DuBose
Associate Editor	Helen Meche
Director, Data Division	Lee Nichols
Contributing Editor	Loraine A. Huchler
Contributing Editor	William M. Goble
Contributing Editor	ARC Advisory Group

MAGAZINE PRODUCTION

Vice President, Production	Sheryl Stone
Manager, Editorial Production	Angela Bathe
Artist/Illustrator	David Weeks
Graphic Designer	Amanda McLendon-Bass
Manager, Advertising Production	Cheryl Willis

ADVERTISING SALES
See Sales Offices, page 160.

CIRCULATION
Director, Circulation — Suzanne McGehee
+1 (713) 520-4440
Circulation@GulfPub.com

SUBSCRIPTIONS
Subscription price (includes both print and digital versions): Print—One year $239, two years $419, three years $539. Digital format—One year $239. Airmail rate outside North America $175 additional a year. Single copies $35, prepaid.

Because *Hydrocarbon Processing* is edited specifically to be of greatest value to people working in this specialized business, subscriptions are restricted to those engaged in the hydrocarbon processing industry, or service and supply company personnel connected thereto.

Hydrocarbon Processing is indexed by Applied Science & Technology Index, by Chemical Abstracts and by Engineering Index Inc. Microfilm copies available through University Microfilms, International, Ann Arbor, Mich. The full text of *Hydrocarbon Processing* is also available in electronic versions of the Business Periodicals Index.

ARTICLE REPRINTS
If you would like to have a recent article reprinted for an upcoming conference or for use as a marketing tool, contact Foster Printing Company for a price quote. Articles are reprinted on quality stock with advertisements removed; options are available for covers and turnaround times. Our minimum order is a quantity of 100.

For more information about article reprints, call Rhonda Brown with Foster Printing Company at +1 (866) 879-9144 ext. 194 or e-mail rhondab@FosterPrinting.com.

Hydrocarbon Processing (ISSN 0018-8190) is published monthly by Gulf Publishing Company, 2 Greenway Plaza, Suite 1020, Houston, Texas 77046. Periodicals postage paid at Houston, Texas, and at additional mailing office. POSTMASTER: Send address changes to Hydrocarbon Processing, P.O. Box 2608, Houston, Texas 77252.

GULF PUBLISHING COMPANY

BPA WORLDWIDE

President/CEO	John Royall
Vice President	Ron Higgins
Vice President, Production	Sheryl Stone
Business Finance Manager	Pamela Harvey

Part of Euromoney Institutional Investor PLC. Other energy group titles include: *World Oil* and *Petroleum Economist*

Publication Agreement Number 40034765

Printed in USA

4. Consumers lack confidence in driving flexible fuel vehicles (FFVs). Testimony by The Alliance of Automobile Manufactures highlighted that even if you can construct a vehicle that operates on a particular fuel, that is not enough for a successful energy policy. The case here is the development of FFVs. Yes, the automakers have successfully provided FFVs for consumers. Yet, consumers are not using 85% ethanol (E85) fuels in significant quantities. FFVs are part of the alternative fuel strategy—but not the single solution. There are 15 million FFVs on US roads; most are located in the Midwest, which has the highest density of E85 fueling stations. However, market penetration by FFVs has not led to a meaningful pickup in renewable fuel usage.

Flawed policies. The promotion of flawed policies to fix the ills of RFS will negatively impact consumers. That is the case involving the 15% ethanol (E15) waiver by the US Environmental Protection Agency (EPA). Automakers believe that EPA's E15 waiver for vehicles model year 2001 and later is the poster child for bad policies adopted to fix the shortcomings of RFS2. The automakers hold that EPA made this decision without critical testing and research. The very likely outcome from E15 waiver will be misfueling and vehicle damage. For the record, all vehicle models before 2001 were designed, certified and warranted to handle up to 10% ethanol (E10). At present, only two original equipment manufacturers produce E15 capable vehicles, and all FFVs are able to use stated fuel blends. Automakers believe that misfueling is unavoidable, and consumers will launch a backlash against renewable fuels, the automakers and the refiners—not the government and EPA.

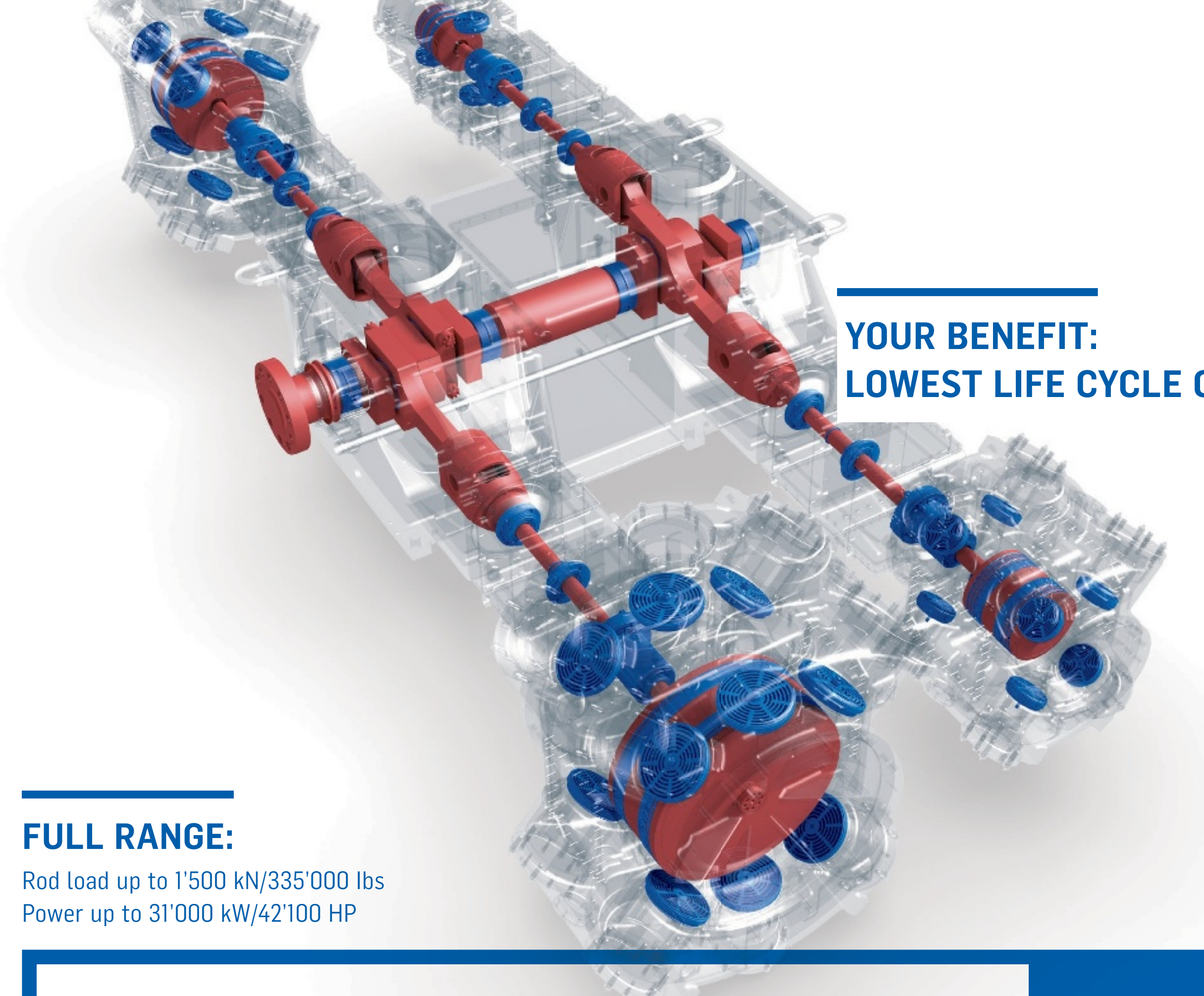
In addition, there is little discussion and provisions on legacy fuels for older vehicles. The US has the largest vehicle population with over 240 million units. Vehicle populations do not turn over quickly; what provisions and funding are being developed to support these consumers? We do hear not anything on how to fund fueling for these vehicles.

Unintended consequences. The unfortunate part of RFS and other policies is that the consumers have little protection against flawed policies and the "fix up" measures. Consumers are paying, and will pay, the full price for these policies that have not been updated to comply with changing conditions.

Note: As always, *HP* encourages our readers to get the facts so that prudent business decisions can be made. The full testimonies and other materials on the July meeting can be found at: http://energycommerce.house.gov/hearing/overview-renewable-fuel-standard-stakeholder-perspectives. HP

EDITOR'S NOTE

During the production of this editorial, EPA issued a reduction in the 2014 blending requirements. The reductions address 2014 targets for cellulosic ethanol from 14 million gal in to 6 million gal. The changes are in response to stakeholder fears that insufficient cellulosic ethanol would be available and the continued decline in gasoline consumption. The action is an attempt to avoid the "E10 blend wall" in finished gasoline with mandated ethanol targets. Both API and AFPM believe that EPA missed an opportunity to fix the ethanol problem for US consumers.



Metso ethylene process valves offer a minimum 7 to 9 years operational uptime.

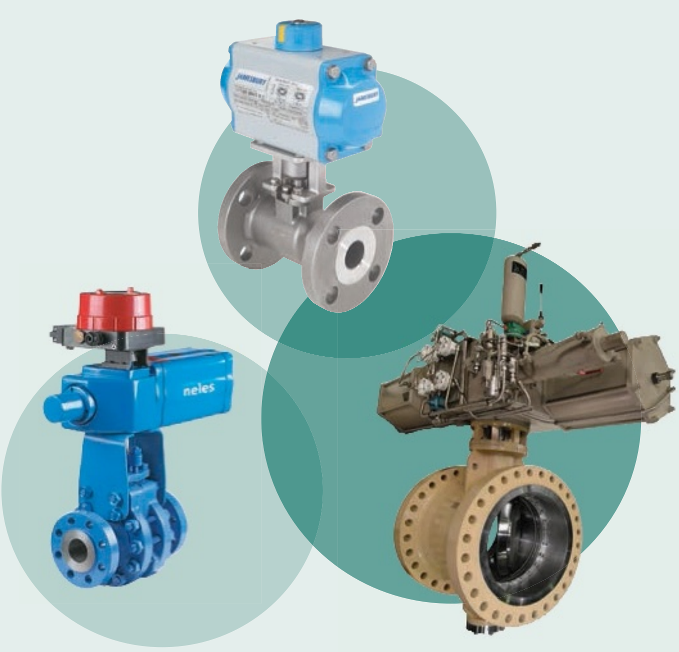
How is that possible? Seat design born from deep industry experience combined with the right coating technology.

Not all process valves are created equal. The right valve will ensure longer performance and save millions in unscheduled shutdowns. This is especially true for ethylene steam cracking and propane dehydrogenation processes that pose extreme temperature, corrosion, abrasion and erosion challenges. Metso's experts understand these variables and have engineered their valves for every critical olefin plant line of control. Not all process valves are created equal. The right valve will ensure longer performance and save millions in unscheduled shutdowns. This is especially true for ethylene steam cracking and propane dehydrogenation processes that pose extreme temperature, corrosion, abrasion and erosion challenges. Metso's experts understand these variables and have engineered their valves for every critical olefin plant application.

These valves are proven to operate 7 to 9 years of operational uptime – compared to competitive valves that operate at roughly half that. The reason Metso valves far outperform others is because of smart seat design combined with the right coating technology. Metso's solutions eliminate leaks, prevent material buildup and provide superior upstream and downstream sealing.

Material Combinations for Specific Ethylene Applications

Coating	Description	Thickness, mm	Hardness, HRC/HV	Application Process	Suitable Substrates
HCr	Hard Chromium	0.05 - 0.1	70 / 1000	Electroplated	Stainless steels, nickel base alloys
ENP	Electroless Nickel Plating	0.06 min	68 / 950	Electroplated	Carbon and stainless steels, nickel base alloys
NiBo	Nickel Boron	0.5 - 1.0	55 / 600	S&F	CF8M, AISI 316 Stainless Steel
WC-Co	Tungsten Carbide	0.1 - 0.15	70 / 1000	HVOF	Stainless steels, nickel base alloys
(W/Cr)C	Tungsten Chromium Carbide	0.1 - 0.15	70 / 1000	HVOF	Stainless steels, nickel base alloys
CrC	Chromium Carbide	0.1 - 0.15	65 / 800	HVOF	Stainless steels, nickel base alloys
Cobalt Based Alloy		1.0 - 3.0	46 / 480	PTA	Stainless steels, nickel base alloys



Metso offers a complete olefin industry solution which includes:

- Comprehensive ethylene-focused products
- Experienced technical application support
- Sophisticated tools to aid sizing and selection
- Global after-sale services throughout the entire product life cycle

Find out how to optimize your olefin processing. Download Metso's special checklist, *"The 5 Keys to Avoiding Ethylene Valve Failures"* at metso.sysmkt.com/l/5Keys/

HP | Brief

Light crude supply for US Gulf Coast refiners

The supply of light crude and condensate will begin to overwhelm simple refining capacity in the US Gulf Coast (USGC) and narrow spreads between Light Louisiana Sweet (LLS) and Mars after 2014, according to ESAI Energy's five-year refining outlook. So far, the region has managed the mismatch between refinery configuration and domestic crude production by backing out light crude import volumes. However, USGC refiners will likely eliminate light sweet imports next year. With throughput growth also limited by weaker exports of products to Latin America, USGC refiners will struggle to process additional domestic light crude.

The addition of topping units and condensate splitters are helping refiners process lighter crude oil, but to entice more complex refineries to substitute lighter domestic volumes for preferred medium sour crude, prices will need to adjust. ESAI Energy's analysis of the refineries on the USGC indicates that for the refineries with more significant residual upgrading capacity and crude slates that are mainly medium to heavy crudes, the LLS-Mars price differential will have to move close to parity.

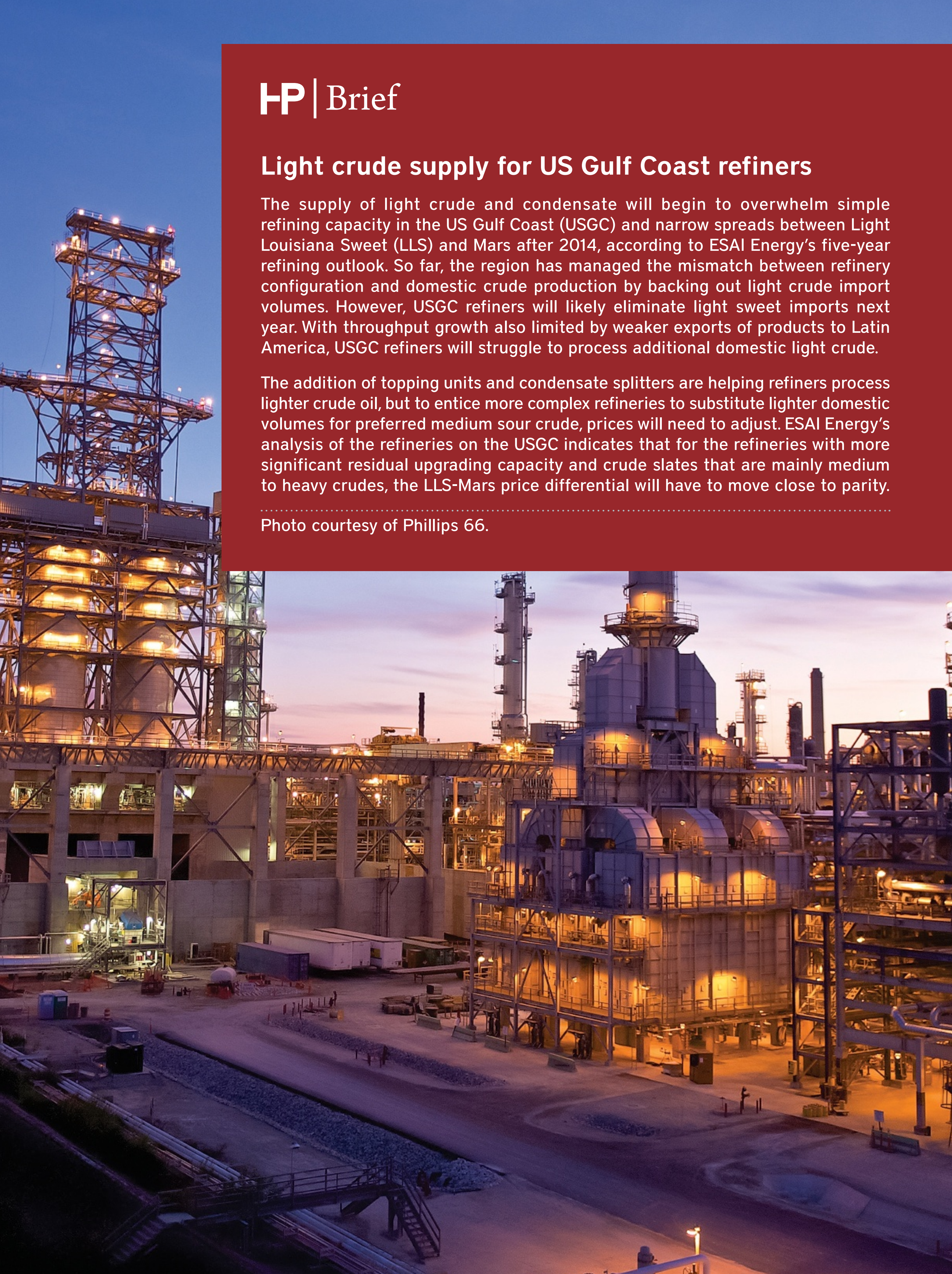


Photo courtesy of Phillips 66.

BILLY THINNES, TECHNICAL EDITOR / Billy.Thinnes@HydrocarbonProcessing.com

HP Brief

Hess has agreed to sell its energy marketing business to Direct Energy, a North American subsidiary of Centrica plc. The sale price was reported as $1.025 billion (B). Hess' energy marketing business supplies natural gas and electricity to 23,000 commercial, industrial and small business customers in the eastern half of the US. The transaction is part of the previously announced plan for Hess to exit the the downstream business as it transforms itself into a pure play E&P company. The sale of energy marketing, along with the sales of four producing assets earlier this year, brings total year-to-date divestitures to $4.5 B.

The US Environmental Protection Agency (EPA) has finalized the 2013 percentage standards for four fuel categories that are part of the renewable fuel standard (RFS) program established by the US Congress. The final 2013 overall volumes and standards require 16.55 billion gallons (B gal) of renewable fuels to be blended into the US fuel supply (a 9.74% blend). This standard specifically requires: biomass-based diesel (1.28 B gal, 1.13%); advanced biofuels (2.75 B gal, 1.62%); and cellulosic biofuels [6 million (MM) gal; 0.004%]. During this rulemaking, the EPA received comments from a number of stakeholders concerning the "E10 blend wall." Projected to occur in 2014, the "E10 blend wall" refers to the difficulty in incorporating ethanol into the fuel supply at volumes exceeding those achieved by the sale of nearly all gasoline as E10. In the issued rule, the EPA proposes to use flexibilities in the RFS statute to reduce both the advanced biofuel and total renewable volumes in the forthcoming 2014 RFS volume requirement proposal. The EPA is also providing greater lead time and flexibility in complying with the 2013 volume requirements by extending the deadline to comply with the 2013 standards by four months, to June 30, 2014.

The US National Cyber Security Center of Excellence (NCCoE) at the National Institute of Standards and Technology is inviting industry to help address two information technology challenges faced by the energy sector. The center is seeking feedback on two proposed "use cases" whose solution would provide centralized control of access to structures and systems and reduce security blind spots in their operations. The first proposed use case is focused on energy companies' need to control physical and logical access to their resources, including buildings, equipment, information technology and industrial control systems. This requires the ability to authenticate identity with a high degree of certainty and to enforce access controls consistently, uniformly and quickly.

The second use case solution would allow security analysts to see operational and information technologies as a cohesive whole, making it easier for them to detect issues that could disrupt services. Energy companies rely on two distinct types of IT systems. Business enterprise systems run their billing, personnel and other enterprises functions while operational systems, which rely heavily on so-called cyber-physical systems, allow them to generate, distribute and meter power.

Gas stations in Shanghai, China, will sell cleaner fuel starting in December. Local officials recently announced the timetable for implementing the Shanghai V (5) standard for fuel. This includes the Shanghai V standard for gasoline and "fifth-phase" standard for diesel. The cleaner fuel will be available at two or three fuel stations as a trial starting in September. It then goes on sale citywide in December. The present cap on sulfur content is 50 parts per million (ppm) and the new regulations will bring that number down to below 10 ppm. The local standard is similar to the European V automobile diesel standard and the Beijing standard V for automobile gasoline, according to the deputy director of the Shanghai Quality and Technical Supervision Bureau. The new regulations are not expected to affect fuel prices in the area.

India's federal cabinet has approved a proposal to sell a 10% stake in Indian Oil. The Indian government owns 78.92% of the country's largest fuel retailer and is expected to raise at least 47.5 B rupees ($786 MM) from the offering. The share sale is part of the government's program to raise 400 B rupees in the fiscal year through March 2014 by selling shares in state-run companies.

Gunvor Group and GE Capital have launched a €225 MM factoring program for Gunvor's receivables portfolio to finance part of Gunvor's German refining business. The deal is a part of Gunvor's broader strategy to diversify its financing base while reducing liquidity risk throughout its businesses. Gunvor is one of the first trading houses to establish a factoring program as a way to diversify how it supports its operations. This deal represents the largest single receivable finance program in Germany.

TransCanada is moving forward with the 1.1-MMbpd Energy East pipeline project based on binding, long-term contracts received from producers and refiners. The conclusion of the successful open season confirmed strong market support for a pipeline with approximately 900,000 bpd of firm, long-term contracts to transport crude oil from Western Canada to Eastern Canadian refineries and export terminals. Eastern Canada imports approximately 700,000 bpd of oil. The project is expected to cost approximately $12 B, excluding the transfer value of Canadian Mainline natural gas assets. The Energy East pipeline will have a capacity of approximately 1.1 MMbpd and is anticipated to be in service by late 2017 for deliveries to Québec and 2018 for deliveries to New Brunswick. HP

Egypt's role in international energy markets

A country analysis from the US Energy Information Administration (EIA) underlines the key role that Egypt plays in international energy markets. Egypt's influence is felt mainly through the operation of the Suez Canal and Suez-Mediterranean (Sumed) pipeline. In 2012, about 7% of all seaborne-traded oil and 13% of liquefied natural gas (LNG) traded worldwide transited through the Suez Canal. Egypt's 2011 revolution and the unrest that has followed have not had any noticeable effect on oil and LNG transit flows through the Suez Canal or Sumed pipeline.

The Suez Canal is an important transit route for oil and LNG shipments traveling northbound from the Persian Gulf to Europe and North America, and southbound from North Africa and countries along the Mediterranean Sea to Asia. Changes to total oil and LNG flows through the Suez Canal in 2012 mainly occurred because of events outside of Egypt, particularly the restart of Libyan oil production and changing dynamics in LNG markets. Further, the Sumed pipeline is the only alternative route nearby to transport crude oil northbound if loaded tankers are too large or ride too low in the water to navigate the Suez Canal.

Oil flows. In 2012, southbound oil flows reached a record high as Libyan oil shipments through the Suez Canal quadrupled in 2012 compared with the previous year, reflecting the ramp-up of Libyan oil production after its civil war. In 2012, about 2.97 million bpd (MMbpd) of crude oil and refined products transited the Suez Canal in both directions. This is the highest amount ever shipped through the Suez Canal, and made up about 7% of total seaborne-traded oil. Crude oil flows through the Sumed pipeline decreased in 2012 to 1.54 MMbpd, as more crude oil was shipped through the Suez Canal via tankers.

LNG flows. Total Suez LNG flows, as a percentage of total LNG traded worldwide, fell to 13%, or 1.5 trillion cubic feet, in 2012, compared with 18% in 2011, for two main reasons. First, northbound LNG flows through the Suez Canal decreased by nearly one-third in 2012, largely because of decreased LNG exports from Qatar to the US and Europe. Second, northbound flows also fell because of reduced LNG exports from Yemen as a result of sabotage attacks on a gas pipeline.

Supply chain. Although external factors have, to this point, played a larger role in altering hydrocarbon flows through Egypt's transit points, unrest in Egypt still presents a risk, and the Egyptian army continues to guard the Suez Canal. Closure of the Suez Canal and the Sumed pipeline would necessitate diverting oil tankers around the southern tip of Africa, the Cape of Good Hope. This would add 2,700 miles to ship oil from Saudi Arabia to the US, increasing both costs and shipping time, according to the US Department of Transportation. Moreover, shipping around Africa would add 15 days of transit to Europe and 10 days to the US, according to the International Energy Agency.

Oil and natural gas production. Egypt's oil and gas production has largely been unaffected by the social unrest. The most visible effect of the revolution on Egypt's energy sector has been a series of attacks on the Arab gas pipeline that contributed to a significant drop in the country's pipeline gas exports. In addition, growing local demand for oil and gas amid stagnant production has led to energy shortages, contributing to continued protests and sporadic unrest in the country.

Divestitures drive US oil and gas deal activity

Divestitures drove US oil and gas merger and acquisition (M&A) activity in the second quarter of 2013, according to PwC US. While second quarter deal volume and value decreased 26% and 43%, respectively, compared to the second quarter of 2012, interest in energy M&A transactions remained robust.

For the three month period ended June 30, 2013, there were a total of 39 oil and gas deals with values greater than $50 MM, accounting for $17.2 billion (B) in deal value, a decrease from the 53 deals worth $30.4 B in the second quarter of 2012. On a sequential basis, deal volume in the second quarter dropped by 5% compared to the first quarter of 2013, with deal value falling by 37% during the same time period.

"There were two main factors that caused a drop in announced deals during the second quarter: companies remained focused on the critical task of integrating assets they acquired during 2012 and sellers bringing deals to market that are fully priced," said Doug Meier, an executive with PwC. "However, interest from potential buyers in acquiring quality assets continues. We are seeing dealmakers go deeper and broader in their diligence to assess whether current deal valuations can deliver long term value. Well-positioned buyers have the right deal strategies, integration plans, and controls in place to execute quickly on opportunities, while successful sellers are providing a clearer and more transparent picture of their assets to minimize transaction timing."

There were 35 total asset transactions, representing 90% of total deal volume, which contributed $11.4 B in total deal value. Divestiture activity is expected to continue as oil and gas companies continue to rebalance portfolios in rapidly changing markets, according to PwC. There were four corporate transactions totaling $5.9 B in the second quarter of 2013, a drop from the 15 corporate deals that totaled $18 B during the same period in 2012.

For deals valued at over $50 MM in the second quarter, upstream deals accounted for 22 transactions, representing 56% of total deal volume and totaling $6.4 B. Additionally, there were 10 midstream deals, accounting for 26% of total deal volume in the quarter worth a total of $6.2 B. According to PwC, demand remains strong for gathering assets as companies look to build out the infrastructure in shale plays. Three downstream deals added $1.1 B, while oilfield services contributed four deals worth $3.6 B during the second quarter of 2013.

Shale remained a key driver for deal activity in the second quarter, with 15 shale agreements with values greater than $50 MM that contributed $7.5 B, or 44% of total deal value. In the upstream sector, shale deals represented nine transactions and accounted for $3.1 B, while six midstream sector shale deals contributed $4.4 B in the second quarter of 2013.

The most active shale plays for M&A with values greater than $50 MM during the second quarter of 2013 include the Eagle Ford in Texas with three total transactions contributing $1.5 B; the Marcellus shale, with three deals totaling $416 MM; and the Bakken in North Dakota, with two deals totaling $910 MM. The Utica shale experienced no deal activity for the first time in seven quarters, according to PwC.

Financial investors' deal activity, including private equity (PE), in the oil and gas industry continued its trend from the first quarter, with volume remaining low with just two transactions involving values greater than $50 MM in the second quarter. Total deal value was $686 MM, a slight increase from the first quarter of 2013, but a 90% decrease compared to the same period in 2012. PE investors opted to stay on the sidelines, partly due to the uptick in valuations.

PwC notes that, in the second quarter of 2013, master limited partnerships (MLPs) were involved in seven transactions, representing 18% of total deal activity. Additionally, in the first half of the year, there were 17 total MLP deals, or 21% of total deal activity. MLPs help satisfy certain investors' thirst for yield, and there is an ongoing interest in MLP "drop downs" to support increased distributions. PwC expects accelerating deal activity in the upstream MLP space, as these businesses supplement depleting assets to sustain and grow distributions.

In the second quarter of 2013, there were no announced deals from foreign buyers, representing a significant change from previous quarters. "Strategically, foreign buyers remain interested in US oil and gas assets," said Mr. Meier. "Similar to our perspective on PE buyers, we expect foreign buyers to continue to be active players in the US oil and gas sector."

PwC's Oil and Gas M&A analysis is a quarterly report of announced US transactions with value greater than $50 MM, analyzed by PwC using transaction data from IHS Herold.

US energy sector vulnerabilities exposed

The US Department of Energy (DOE) has released a report that assesses how critical US energy and electricity infrastructure is vulnerable to the impacts of climate trends. Historically high temperatures in recent years have been ac-

companied by droughts and extreme heat waves, more wildfires than usual, and several intense storms that caused power and fuel disruptions for millions of people. These trends are expected to continue, which could further impact energy systems critical to the US economy.

The "US Energy Sector Vulnerabilities to Climate Change and Extreme Weather" report notes that annual temperatures across the US have increased by about 1.5°F over the last century. In fact, 2012 was the warmest year on record in the contiguous US, and it also included the hottest month since the country started keeping records in 1895. Implications for the US energy infrastructure are explored in this report.

Climate conditions have increased the risk of temporary partial or full shutdowns at thermoelectric power plants because of decreased water availability for cooling and higher ambient air and water temperatures. A study of coal plants, for example, found that roughly 60% of the existing fleet is located in areas of water stress.

Energy infrastructure located along the coast is now vulnerable to damage from rising sea levels, increasing intensity of storms and higher storm surge and flooding. Such weather patterns can disrupt oil and gas production, refining and distribution, along with electricity generation and distribution.

Energy companies and consumers also now face increased risks of disruption and delay to fuel transport by rail and barge during more frequent periods of drought and flooding that affect water levels in rivers and ports. Another potential problem pertains to higher air conditioning costs and risks of blackouts and brownouts in some regions if the capacity of existing power plants does not keep pace with the growth in peak electricity demand due to increasing temperatures and heat waves.

In addition to identifying critical areas at risk from climate change and extreme weather, the report identifies activities already underway to address these challenges, and discusses potential opportunities to make the energy sector more resilient.

Potential future opportunities for federal, state and local governments could include innovative policies that broaden the suite of available, climate-resilient energy technologies and encourage their deployment; improved data collection and models to better inform researchers and lawmakers of energy sector vulnerabilities and response opportunities; and enhanced stakeholder engagement.

The report says these activities will increase the resilience of the US energy infrastructure by "hardening" existing facilities and structures to better withstand severe droughts, floods, storms or wildfires, and by contributing to smarter development of new facilities.

Hurricane Sandy is one recent example of energy sector vulnerability. Sandy's storm surge caused eight million customers in the US Northeast to lose power while fuel pumps at gas stations ceased functioning across the region. Six refineries with a cumulative capacity of 862,000 bpd were forced to shut down or severely reduce output. HP

ADRIENNE BLUME, MANAGING EDITOR
Adrienne.Blume@HydrocarbonProcessing.com

Noncorrosive bearings have high load capacity

EDT Corp. has added Type E Solution bearings (**FIG. 1**) to its Solution line of severe service bearings. Type E Solution bearings have high load capacity and the same dimensional footprint of industry-standard Type E tapered roller bearings, along with the advantages of being fully noncorrosive and completely greaseless.

Type E Solution bearings are constructed with a 304 stainless steel housing, a Polysphere plane bearing made from a high-performance polymer, and a 316 stainless steel locking sleeve. Plane bearings work without rolling elements, so grease is unnecessary and seals are eliminated. Plane bearing design also makes it possible to offer completely split inserts in addition to split housings.

EDT Type E Solution bearings are available in one piece or in split styles of pillow blocks, four bolts and piloted housings for shaft sizes from 1³⁄₁₆ in. to 5 in. Split units are drop-in replacements for popular spherical roller bearings, with the added options of split bearings and split sleeves.

In locations where the environment compromises the grease of standard roller bearings or where more corrosion resistance is needed, EDT Type E Solution bearings can provide a superior alternative when used in appropriate locations. EDT can assist in identifying applications where Type E Solution bearings will last longer, reduce maintenance and save significant operating costs over time.

Select 1 at www.HydrocarbonProcessing.com/RS

Oil wiper improves compressor reliability

The Oil Film Dynamic (OFD) wiper ring (**FIG. 2**) from HOERBIGER improves compressor reliability and cuts operating costs by eliminating crankcase oil losses and oil contamination.

The wiper ring's innovative design and operating principle eliminates the leakage paths, oil pumping action and sharp edges that characterize conventional oil wipers. As a result, it pushes oil back into the crankcase instead of outward, as do conventional oil wipers. The OFD ring is made from hard-wearing polymer instead of metal, ensuring long life without damaging the piston rod or impairing lubrication.

Oil wiper rings are used on the piston rods of reciprocating compressors. They stop crankcase oil from contaminating the gas being compressed in the cylinders, without compromising the lubrication of the machine.

Operators of reciprocating compressors typically waste oil and suffer reduced reliability because of the poor design of conventional metal oil wiper packings. Outward leakage leads to crankcase oil losses as high as 5 liters (1 gallon) per day per compressor. At the same time, oil carried back into the crankcase can be contaminated with process gases, such as corrosive hydrogen sulfide, which can damage the compressor and require more frequent oil changes.

Extensive testing, both in-house and on 25 field applications, has shown a large reduction in oil leakage with almost no wear. Four factors characterize the design and performance of the OFD oil wiper ring. The first is geometry. The OFD wiper ring features a two-piece design based on two concentric rings of L-shaped cross-sections. Each ring covers the gaps in the other, closing off oil leakage paths. The inner ring has a carefully engineered profile for best oil control.

The second factor is a new spring plate that applies a side (axial) load to the OFD rings. This stops the rings from moving, eliminating the "shuttling" effect that plagues conventional wiper rings.

The third factor is a change in material. The OFD ring is made from a high-performance polymer that eliminates leaks because it follows the shape of the piston rod accurately. The polymer is durable yet soft, so there is no danger of rod damage.

These three design aspects combine to create the fourth key differentiator of the OFD wiper ring: their innovative working principle. The OFD ring uses elasto-hydrodynamic effects to pump oil back into the crankcase without losing the vital lubricating film, while their soft material avoids rod damage.

For process gases and natural gas, the ability to combine oil wipers and pressure packing in the same housing gives reliable performance at high speeds, in vertical or horizontal configurations, and with rod diameters up to 130 mm (5 in.). Where

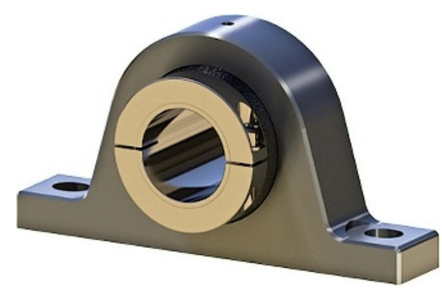


FIG. 1. Type E Solution bearings have high load capacity and are greaseless and non-corrosive.

FIG. 2. The Oil Film Dynamic wiper ring improves compressor reliability and cuts operating costs.

condensation is an issue, a double seal arrangement—with oil wipers mounted on each side of the pressure packing—keeps liquids out of the crankcase.

Select 2 at www.HydrocarbonProcessing.com/RS

Controller gives precise petrochemical blending

Honeywell's Enraf recently introduced the Fusion4 MSC-A (**FIG. 3**), an advanced and easy-to-operate controller that uses intuitive technology similar to today's mobile phone applications to ensure efficient, safe petrochemical handling.

The controller is operated using intuitive, onscreen icons that reduce training time and improve accuracy. The interface delivers detailed process data, resulting in superior operational monitoring.

Key features of the controller's advanced function include:

- Safe, secure, two-way data communication with the handheld Fusion4 local access device allows rapid transfer of transaction data, configuration files and calibration records
- Advanced process monitoring and control for up to 12 streams simultaneously at the load rack
- Detailed transactional data, alarms, diagnostics and calibration event logging for a secure audit trail for business and regulatory auditing
- Fast, efficient management of multiple systems and sites
- Compatibility with Enraf's safe area device monitoring software package Fusion4 Portal for remote monitoring and configuration of hazardous area devices, as well as Measuring Instruments Directive-approved bill of lading printing, scheduled summary printing, alarm management and transaction audit trails to boost efficiency and compliance.

The Fusion4 MSC-A controller can be supplied as a standalone device or coupled with additive injection hardware for a complete off-the-shelf additive solution. Allowing any downstream oil and gas operation to add standalone additive injection capability, it integrates with industry standard terminal automation systems and load computers, and avoids the need for expensive alterations to existing equipment and infrastructure.

Select 3 at www.HydrocarbonProcessing.com/RS

Enhanced CENTUM software released

Yokogawa Electric Corp. recently released an enhanced version of the company's flagship production control system platform, CENTUM VP R5.03. Yokogawa is continuously developing CENTUM VP as the foundation of its IA

FIG. 3. The Fusion4 MSC-A controller uses intuitive technology to ensure efficient, safe petrochemical handling.

business' VigilantPlant vision for operational excellence.

New features in the CENTUM VP R5.03 update are integrative and intuitive:

- A wide-area communication router enables a CENTUM VP system to remotely monitor and control equipment over a wide-area network. The router enables reliable, secure and inexpensive monitoring and control of widely distributed facilities, such as the production platforms in deepwater offshore oil and gas production; pumping and compressor station controls along pipelines; and wellhead, gathering treatment and separation facilities in shale oil and gas onshore fields.
- The CENTUM VP R5.03 batch package is based on ISA-88 and has enhanced functions that allow greater flexibility when accommodating changes in production procedures. The number of operating procedures has been increased, and the recipe procedure has been provided with additional levels of granularity, simplifying the process of making recipe changes. CENTUM VP R5.03 is ideal for companies that employ complex batch production processes and are engaged in such industries as specialty chemicals, pharmaceuticals and foods.
- The platform supports the IEC 61850 communication protocol, enabling it to integrate data from intelligent electrical devices. This allows better integration with the electrical equipment in process plants, eliminates the need for multiple systems in the plant, and enables the monitoring of electrical energy in process plants, enabling better plant energy management.
- A new keyboard has been released with sets of dedicated function keys that can be used to simultaneously adjust eight control loops. This keyboard has been designed to make it easier for plant operators to perform tasks.

Select 4 at www.HydrocarbonProcessing.com/RS

Device protects equipment against ground faults

ABB has introduced the F200 Ground Fault Equipment Protector (GFEP) (FIG. 4), designed to protect equipment against damaging line-to-ground currents by disconnecting all ungrounded conductors of the faulted circuit. A ground fault is caused by an insulation loss between a live conductor and an exposed conductive part that creates a flow of current to the ground.

Most short circuits initially manifest as ground faults, which, if undetected, can cause serious and costly damage to electrical systems and equipment. An F200 GFEP added to an electrical system trips when the system leaks a significant current to the ground, immediately preventing the potential short circuit before such damage can occur.

FIG. 4. The F200 Ground Fault Equipment Protector prevents damage to electrical equipment.

The ABB F200 GFEP line consists of the 2-pole F202 for single-phase networks, and the 4-pole F204 for split-phase Delta and Wye networks. All variations are compliant to the UL 1053 standard for ground fault sensing and relaying equipment, and in accordance with ANSI/NFPA 70, for use in ordinary locations. The F200 GFEP device can be applied to most electrical networks up to 480VAC, and it can be installed in distribution panels and on critical equipment.

Select 5 at www.HydrocarbonProcessing.com/RS

Wireless GWR transmitter released

Emerson Process Management has launched the industry's first truly wireless guided wave radar (GWR) transmitter (**FIG. 5**) for continuous level and interface monitoring. The Rosemount 3308 wireless GWR transmitter has been introduced to meet the need for accurate level monitoring in remote or difficult-to-reach locations where installing new cabling would be costly or impractical.

GWR transmitters are widely used for a broad range of measurement and control applications. These include process-level measurement in vessels and storage tanks in refineries, oil fields, offshore platforms, chemical plants and industrial plants.

Emerson's Smart Wireless technology significantly reduces installation and configuration time. Since there is no need for cabling, trenching, conduit runs and cable trays, costs are typically reduced by 30% or more compared with a wired solution. Once a wireless network has been established, any WirelessHART-enabled device can join the network and take advantage of the benefits of an existing Smart Wireless infrastructure. This makes it easy to make changes or move devices to meet specific requests.

The transmitter features enhanced configuration and diagnostics. It provides an easy and cost-effective way to add visibility across a wide range of industries and applications, ensuring plant and operator safety and improving process efficiency. The transmitter is virtually unaffected by changing process conditions such as density, conductivity, temperature and pressure; and because it does not have moving parts, no calibration is required and maintenance is minimized.

Select 6 at www.HydrocarbonProcessing.com/RS

Emissions-monitoring system meets EPA standards

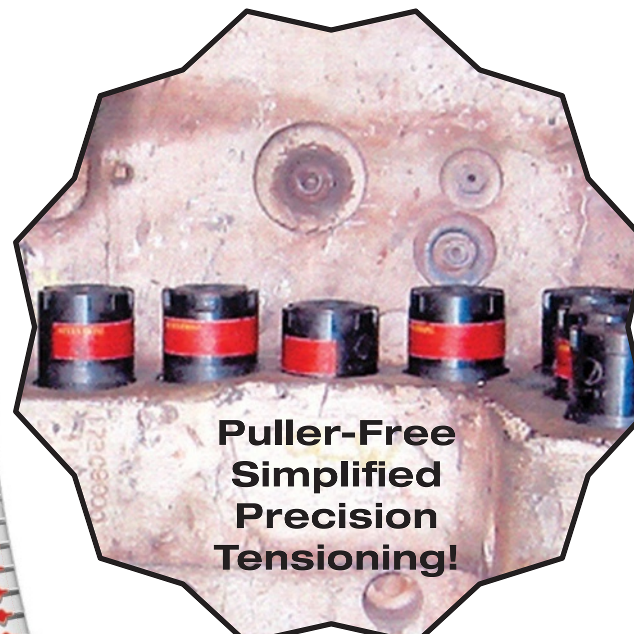
The Thermo Scientific Omni Fourier transform infrared (FTIR) multi-gas

FIG. 5. The Rosemount 3308 wireless guided wave radar transmitter extends continuous level monitoring to difficult-to-reach locations.

HYTORC®
Since 1968

7FA CASING Next Time - ZIP IT!

Make Safety a Habit Reliability a Fact!



HYTORC Casing-Bolt TOOL KIT

Everything you need for the Horiztals and Verticals of the 7FA in one single box!

Switch from hydraulic tensioning or hydraulic torque to Stretch-to-Load without replacing the bolts. Just click the tools on the reusable nuts and stretch all bolts to the desired load within 5%, hands-free from a safe distance. There is no puller to fail, no reaction arm to pinch, no bolt relaxation, no torsion or side load, no measuring, just straight axial bolt load to compress the joint evenly amongst all fasteners or to loosen it up – as if it had a ***ZIPPER!***

TIME STUDY

Number of Fasteners stretched simultaneously to the desired load: Horizontal Joint: 2 on each side, Vertical Joint: 2 on each side.

Total Time: 1 hour, 27 Minutes!

HYTORC, Division UNEX Corporation +1 201 512-9500 Sales@HYTORC.com www.HYTORC.com continuous emissions-monitoring system (CEMS) uses FTIR technology, enabling a user to simultaneously analyze up to 10 stack gases, including hydrogen chloride (HCl).

The system's rugged design makes it possible for facilities to accurately measure emissions in harsh industrial environments, in accordance with US Environmental Protection Agency (EPA) 40CFR Part 60/63 standards and HCl performance specifications.

The CEMS uses an advanced software platform to achieve precise and stable analysis of complex gas spectra found in demanding industrial applications. It features:

- High resolution with multiple scans and strong compensation algorithms
- Minimal drift with self-calibration
- Real-time dynamic alignment for long-term stability
- Continuous, clean operation using a unique flowback mode.

Select 7 at www.HydrocarbonProcessing.com/RS

Touchscreen viewable in direct sunlight

The SRMTR-10.4V Sunlight Readable touchscreen (**FIG. 6**) from TRU-Vu Monitors features a 1,000-nits brightness screen. This is 4–5 times brighter than standard monitors, and enables the viewing of clear, sharp, vivid color images, even with bright sunlight falling directly onto the screen.

The monitor offers a 5-wire resistive touchscreen, 800 × 600 resolution, and video graphics array input in a rugged, powder-coated steel enclosure. It also features exclusive TRU-Tuff treatment for maximum shock and vibration resistance.

TRU-Tuff Series monitors have been deployed in a wide range of industrial, military and commercial applications including inspection systems, monitoring systems, navigation and guidance systems, vehicles, helicopters, offshore drilling rigs and more.

Select 8 at www.HydrocarbonProcessing.com/RS

FIG. 6. The SRMTR-10.4V touchscreen monitor enables the viewing of clear, sharp color images in bright sunlight.

FIG. 7. The Clariant sunliquid pilot plant now produces cellulosic ethanol from a variety of agricultural residues.

Clariant expands sunliquid biofuels technology

Clariant is demonstrating the international potential of its process for manufacturing sustainable biofuel, expanding its sunliquid technology to include additional agricultural residue feedstocks.

Following the opening of Germany's largest demonstration plant (**FIG. 7**) for manufacturing ethanol from agricultural residues in July 2012, Clariant has discovered that corn stover (the most valuable raw material in North America) and sugarcane bagasse (a cellulose-rich byproduct of sugar and ethanol production in Latin America and Asia) can also be converted efficiently with the process. European wheat straw has primarily been used to date, and has delivered good results.

The high degree of efficiency of the sunliquid process, independent of the raw material used, is achieved largely through the use of raw material-specific, highly optimized enzymes. Clariant intends to market the sunliquid process internationally, making it vital that different raw materials can be flexibly and effectively used in the process.

Corn stover is primarily important in North America, where 570 million metric tons per year (tpy) of this plant residue are generated from the corn harvest. In Brazil, ethanol production could be increased by some 50%, compared to current levels, by using excess bagasse. Using these plant residues means that there is no competition with food production or for agricultural land.

Cellulosic ethanol produced using the sunliquid process reduces greenhouse gas emissions by around 95% compared to fossil fuels. Domestic production of this liquid fuel can also reduce dependence on oil imports and generate economic growth in local markets.

The next step in commercializing the sunliquid technology is the construction of the first commercial production facility, with a capacity of 50,000 metric tpy–150,000 metric tpy, compared with around 1,000 metric tpy of capacity at the existing pilot plant.

Select 9 at www.HydrocarbonProcessing.com/RS

Biobutanediol helps cut carbon, costs

Myriant Corp. and Johnson Matthey–Davy Technologies (JM Davy) announced the successful production of biobutanediol (BDO) and tetrahydrofuran (THF) made from Myriant's biosuccinic acid.

The qualification work was conducted at JM Davy's facility in Teesside, England, using biosuccinic acid supplied by Myriant and the JM Davy BDO/THF process. Combining the efficiencies of Myriant's biosuccinic acid process and the JM Davy BDO/THF process, the biobutanediol and biotetrahydrofuran have an overall carbon efficiency of 87%, believed to be substantially better than the carbon efficiency achieved in the direct fermentation route to biobutanediol.

The JM Davy BDO/THF technology has undergone significant improvements over the past 10 years. It can now be offered with process and performance guarantees to produce commercial-grade biobutanediol, THF and gamma-butyrolactone, equivalent to the petrochemical material that is produced in commercial plants from Myriant's biosuccinic acid, at a competitive cost level.

JM Davy has licensed 800,000 metric tons per year of capacity for the production of butanediol, THF and gamma-butyrolactone, representing approximately 25% of global installed capacity. Myriant's flagship biosuccinic acid plant in Lake Providence, Louisiana—the largest biosuccinic acid plant in the world—is presently being commissioned.

Select 10 at www.HydrocarbonProcessing.com/RS

ADRIENNE BLUME, MANAGING EDITOR
Adrienne.Blume@HydrocarbonProcessing.com

Inaugural GTL Technology Forum delivers insights, opportunities

Gulf Publishing Company's inaugural Gas-to-Liquids (GTL) Technology Forum and exhibit took place in Houston, Texas from July 30–31. Speakers and attendees shared knowledge on gas processing technology developments, project economics and business challenges, with a focus on GTL processing technologies. The event, which featured five technical sessions and two keynote speakers, was sponsored by Honeywell and drew over 160 attendees representing 96 companies from 10 countries.

Networking lunches and refreshment breaks in the Forum's exhibit space allowed delegates to discuss business strategies over coffee and desserts, and to learn more about the technology and data management solutions offered by conference exhibitors Pentair, Forum Energy Technologies, AMACS and *Hydrocarbon Processing's* Construction Boxscore Database.

Day 1. The Forum opened on Tuesday, July 30 with a keynote speech by Mark Schnell (**FIG. 1**), the general manager of marketing, strategy and new business development for Sasol, on the role of GTL in the new North American energy landscape. Mr. Schnell called it an "exciting time to be in the North American gas business" for those on the demand side of the equation. He addressed three major topics, including Sasol's progress on its GTL plant in Louisiana, the company's experience on its GTL journey, and where GTL might fit into the energy landscape going forward.

Sasol's proposed GTL complex in Lake Charles, Louisiana would have a capacity of 96 thousand barrels per day (Mbpd) and be co-located next to an existing ethylene cracker. GTL startup is presently targeted for 2019, although the final investment decision in 2014 will be based largely on economics. During a conference break, Mr. Schnell told *Hydrocarbon Processing* that he expects as many as 6 or 7, and at least 4 to 5, of the proposed ethylene cracker projects in the US to move forward. The Sasol GTL plant would produce GTL diesel, naphtha, paraffins, waxes and lubricant base oils—making it the first facility in the US to manufacture GTL transportation fuels and other products.

Mr. Schnell also addressed some of the challenges of the evolving GTL sector, noting that, at present, "Commercial capacity is in the hands of a few companies." The addition of more players would offer improved security of supply, greater advocacy for alternative fuel policies, and more security for original equipment manufacturers. "To be truly taken seriously, [GTL] will have to become an industry, rather than a handful of clients or players," Mr. Schnell acknowledged. "It's up to us as business and technology providers to step up and provide solutions."

Session 1. The first session opened with a presentation by Srinivasan Ambatipati of R3 Sciences on the development of modular technology for gas-to-methanol conversion. Mr. Ambatipati emphasized the need to use flared gas, which totaled 5 trillion cubic feet (Tcf) worldwide in 2011. R3 Sciences' gas-to-methanol (G2M) technology uses a three-step process involving autothermal reforming and synthesis gas (syngas) conditioning to produce methanol.

Next, Dr. Ronald Sills of the XTL & DME Institute presented his view on the use of dimethyl ether (DME) as a transportation fuel in North America. Dr. Sills named the three major fuel applications of DME as LPG blendstock, power generation and transportation fuel. On this last point, he noted that Volvo is the first vehicle manufacturer to announce plans to commercialize DME-powered heavy-duty trucks in North America, which will happen as early as 2015. "DME is safe, inexpensive to store, simple to transport and does not require cryogenic or high-pressure storage," Dr. Sills acknowledged. Although DME is cheaper than diesel, offering compelling economics, the large-scale production of DME will be needed to encourage the availability of DME-compatible vehicles, he said.

Dr. Carl Hahn from Pentair next spoke about reducing capital and operating expenditures (CAPEX and OPEX, respectively) through more effective separation technologies. Dr. Hahn outlined the major challenges to the commercial viability of GTL as high capital intensity, high investment risk and cashflow con-

FIG. 1. Sasol's Mark Schnell opened the GTL Technology Forum on Day 1 with a keynote speech on the role of GTL in North America.

straints. Conventional separation technology can cause operational problems in GTL plants, although Pentair offers its Polarex technology as an extractive separation technology that can be used in lieu of a conventional contactor.

Nearing the end of the first session was a presentation on small-scale GTL as an economic solution for distributed gas by Dr. Paul Schubert, CEO of Velocys Inc. Dr. Schubert acknowledged a growing set of small- to medium-scale prospects, with five small-scale GTL technology clients progressing to the engineering stage since the start of 2013, along with 19 companies in the pre-engineering stage. Potential applications for small-scale GTL in North America include gas flaring, diversification for midstream companies and refinery integration. From there, Anindita Moitra of Indian Oil Corp. finished off the session with a presentation on off-balance sheet project risk management.

Session 2. After a lunch and dessert break that generated numerous networking opportunities for attendees, the second session kicked off with an insightful talk by Tara Fatima of Bechtel Hydrocarbon Technology Solutions Inc. on low-cost opportunities for methanol-to-olefins (MTO) and methanol-to-propylene (MTP). Ms. Fatima acknowledged that MTO technologies are competitive with traditional steam cracking only when there is a reasonable price spread between gas and oil (i.e., \$4/MMBtu or less for gas, and \$80–\$90/bbl or more for crude oil). This makes MTO and MTP technologies largely uneconomical in the US, but viable in the Middle East.

Next, V. K. Arora of Kinetics Process Improvements Inc. gave a talk on the advances and challenges of syngas preparation. Mr. Arora reviewed the advantages and disadvantages of major syngas technologies, including steam methane reforming, autothermal reforming, partial oxidation, convective reforming, two-step reforming and prereforming. In addition, he discussed the major hurdles for GTL operations in North America, specifically those concerning process complexity, process efficiency, economic challenges and longer financing requirements.

Session 3. ExxonMobil Research and Engineering Co.'s Mitch Hindman led off the third session with a presentation on ExxonMobil's methanol-to-gasoline (MTG) technology as an alternative for liquid fuel production. The technology, which is a fixed-bed process that produces sulfur-free gasoline with 92-RON octane, was first commercialized in 1985 in New Zealand. There are seven licenses for MTG technology around the world, according to Mr. Hindman.

A panel discussion followed that included Honeywell's Randy Miller, Invensys' Bill Poe, Emerson Process Management's Chuck Miller and Aspen Technology's Michael Harmse. The panelists discussed various aspects of process control challenges for GTL, including the link between process control and operator competency, online modeling for GTL applications, process and safety requirements with regard to automation systems, cyber security concerns, the benefits of standardization of small-scale plants and the need for dynamic simulation and remote monitoring technologies to ensure proper operation of control systems.

Day 2. The second day of the GTL Technology Forum began with a keynote address by Iain Baxter, the director of business development at CompactGTL in the UK. Mr. Baxter spoke about transformational gas solutions for the upstream industry and the problems inherent in conventional GTL operations. He noted that there are only a handful of companies with the experience and technologi-

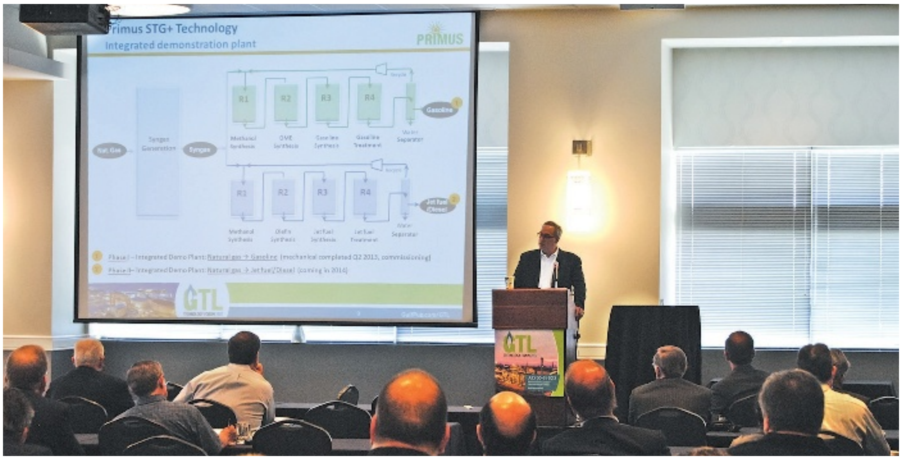


FIG. 2. Dr. George Boyajian of Primus Green Energy discussed the cost-effectiveness of Primus' STG+ small-scale GTL technology on Day 2.

FIG. 3. John Royall (left), CEO of Gulf Publishing Company, and Mark Schnell (right), general manager of marketing, strategy and new business development for Sasol, were among the speakers at the GTL Technology Forum.


Think globally. Act responsibly.
Sabin's processes for recovering and refining PGMs from spent catalysts raise environmental protection to an advanced state for safety and regulatory compliance. Your advantage: assured peace of mind, and our promise of maximum value returns from your precious metal-bearing materials.
Learn more at sabinmetal.com
sabin
METAL CORPORATION
Platinum • Palladium • Rhodium • Ruthenium • Rhenium • Gold • Silver
Processing facilities and technical service offices around the world
Select 68 at www.HydrocarbonProcessing.com/RS

cal expertise to design large-scale GTL plants. These companies tend to be large, integrated energy firms and are often protective of their intellectual property.

However, CompactGTL's compact reforming technology for syngas generation provides fully modular GTL production in a range of design capacities, from 10 MMscfd to 150 MMscfd. The company is presently working with Petrobras, Total, Gazprom and 17 other companies around the world for small-scale GTL projects, both onshore and offshore.

Session 4. The fourth session kicked off with a discussion by Dr. Uday Turaga of ADI Analytics LLC on benchmarking gas monetization opportunities. Dr. Turaga explained that, in North America, LNG export projects offer the most attractive returns, followed by MTG and GTL projects. He also emphasized the need for creative thinking during the commercial structuring of a GTL project to reduce CAPEX and OPEX.

Lee Nichols, the director of *Hydrocarbon Processing's* Construction Boxscore database, next offered a recap of the global construction outlook for GTL projects. Mr. Nichols noted that Construction Boxscore is tracking over 800 gas processing projects globally.

Afterward, Daniel Barnett of BD Energy Systems LLC discussed reformer furnace outlet systems and needed improvements to conventional steam methane reformer furnaces. Mr. Barnett noted that conventional practices apply a design temperature for outlook system components that is not adequate for GTL or ammonia plant reforming. The final speaker of the session was Dr. Dave Sams of Albemarle Corp., who explained how Albemarle's MA-15 catalyst aids in the thermochemical conversion of syngas to ethanol.

Session 5. The fifth and final session of the conference opened with a discussion by Dr. George Boyajian of Primus Green Energy (FIG. 2) on the cost-effectiveness of Primus' STG+ GTL technology, which enables the conversion of natural gas to drop-in liquid fuels on a small scale. Robert Herrmann of Robert P. Herrmann LP next discussed the use of a gas lift apparatus for a Fischer-Tropsch production riser.

Following these presentations, John Oyen of ABB Inc. spoke about next-generation facilities, with emphases on trends in automation and improvements in control room technology, and how these developments can enhance operations at GTL facilities.

Lastly, Steve Worley of Worley Engineers Inc. discussed the design requirements for floating vessels intended for offshore GTL production. Mr. Worley advised the use of a proven floating production, storage and offloading (FPSO)-style vessel, or an alternative proven design, to minimize sea motion at floating GTL (FGTL) projects.

John Royall, CEO and chairman of Gulf Publishing Company (FIG. 3), concluded the conference by thanking attendees for their participation and input. "We anticipate some major market changes by the time [the conference is held again next year], especially in compact and small-scale GTL technologies," Mr. Royall said.

Gulf Publishing Company's second annual GTL Technology Forum will be held in Houston in 2014. HP

STEPHANY ROMANOW, EDITOR
Stephany.Romanow@HydrocarbonProcessing.com

US Midwest refinery nears completion of major modernization project

In early July 2013, BP announced that it has completed the commissioning and startup of a new, 250,000-bpd crude distillation unit at the Whiting refinery, marking a major achievement in the multi-billion-dollar upgrade of the facility in Northwest Indiana. Commissioning of the state-of-the-art distillation unit is a major milestone in the refinery upgrading project designed to unlock $1 billion in future operating cashflow.

The Whiting refinery has a long history; the original plant was constructed in May 1889 on 235 acres. This former Standard Oil (of Indiana) refinery shipped refined products to nearby Chicago and other Midwestern cities in the early days. Later, it was part of Amoco Corp. until 1998, when it was purchased by BP. The Whiting refinery is major part of the Petroleum Administration for Defense District (PADD) 2 refining district.

Need to modernize. In September 2006, BP announced a massive modernization project for the Whiting refinery. The goal was to update this refinery with flexibility to process heavier crude oils. In May 2008, after necessary permits were issued, BP began construction.

FIG. 1. View of the upgraded distillation unit and new coker at BP's Whiting refinery.

FIG. 2. BP's Whiting refinery sits outside the city limits of Whiting on Lake Michigan.

Some project highlights are:

- Environmental improvements for the complex will exceed $1 billion.
- A modern gasoil hydrotreater will remove/treat more sulfur and nitrogen from refinery gasoil.
- The sulfur recovery complex will be upgraded to a more efficient system to remove more sulfur from fuel product streams.
- A new, state-of-the-art coker will replace the existing one. The new coker will enhance process safety through increased automation, along with higher coke and naphtha production.
- The refinery's largest crude distillation unit is being reconfigured to process heavier crudes, including Canadian oil sands.

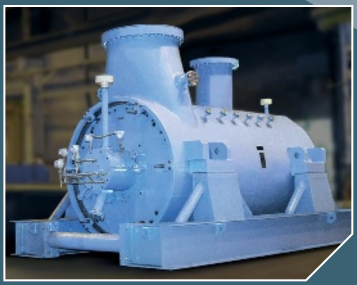
"The safe startup of this large, sophisticated crude processing unit at the Whiting refinery has returned the refinery to its nameplate processing capability of 413,000 bpd—initially of mostly light, sweet crude—and paved the way for the remaining upgrades to the plant to be brought online," said Iain Conn, chief executive of BP's refining and marketing segment. "When the new coking and hydrotreating units are commissioned and operating at full rates in the second half of this year, the reconfigured refinery will have the flexibility to greatly increase heavy, sour crude processing, delivering an expected incremental $1 billion of operating cashflow per year, depending on market conditions."

Construction of the Whiting refinery upgrade project is more than 95% complete. BP expects to commission a new, 105,000-bpd gasoil hydrotreater, a 102,000-bpd coker and other associated units in the second half of 2013. When all of the new equipment is in full operation, the refinery will have the ability to significantly increase heavy, sour crude processing to roughly 80% of its overall crude run, up from 20%. The Whiting refinery is the seventh-largest US refinery and the largest refinery in its PADD. It is the largest, most complex refining project undertaken in BP's recent history.

"The Whiting refinery project is at the heart of our US fuels strategy to operate sophisticated, feedstock-advantaged refineries tied to strong logistics and fuels markets," Conn added. "This world-class refinery is in the right location and will soon be running the right equipment to process growing supplies of North American crude oil, including oil from Canada."

The multi-billion-dollar investment in the refinery is the largest private-sector investment in the state of Indiana history, and it also includes several hundred million dollars in state-of-the-art environmental controls for water treatment and air emissions, according to Whiting refinery manager Nick Spencer. "Our investment in Whiting's future shows BP's commitment to creating jobs in America and safely providing energy," Spencer said.

Other notables. The modernization project required over 10,000 skilled craftspeople at different stages of construction. It involved installing over 380 miles of pipe, 1,200 pieces of major equipment, and 600 shop-fabricated modules and 50,000 tons of steel. More importantly, it logged 40 million work-hours without an injury resulting in a day away from work. HP



HEINZ P. BLOCH, RELIABILITY/EQUIPMENT EDITOR
Heinz.Bloch@HydrocarbonProcessing.com

Leaders must allocate time for technology updates

Just as a good medical doctor is expected to have wide knowledge on modern treatment options, we would expect reliability professionals to be equally informed on technology advances. There are numerous training and technology update opportunities, and this column has highlighted these for well over two decades. Conscientious reliability professionals should take advantage of books, articles, conference proceedings, vendor-organized training, and, of course, and "opportune times."

Never stop learning. Opportune times present themselves in many different ways. Two hours between job assignments or a half-day postponement of a scheduled meeting are opportunities for the reliability professionals to update their understanding of numerous topics and issues. If you have checklists in your computer, why not update them based on an article that you filed away a few months ago, or a conference paper that you read just last week? Here are three examples that we consider relevant and pertinent because each can be found in recent publications. The very fact that you're reading this column greatly increases the probability that these examples apply to your plant. Anyway, remember that you are seeking for an opportunity to investigate these and similar matters.

Example. Start with recent illustrations and documentation referring to mechanical seals with bi-directional tapered pumping impellers that effectively pump the flush liquid through a small heat exchanger, as shown in **FIG. 1.** Which of the hundreds of process pumps at your facility would qualify for, and benefit from, this arrangement? Have you asked your seal supplier? Could you learn from the supplier's answers?

Next, consider using opportunistic time to review the latest illustrations and documentation dealing with oil rings, also known as "slinger rings." These components pick up oil from a bearing housing sump and direct the lubricant into rolling element bearings. A current article strongly recommended measuring the as-installed width of an oil ring and to again measure the as-found ring at the next repair event. The difference in the measured widths relates to a volume of abrasive dust which, with virtual certainty, contaminated the oil and shortened the bearing's service life. So, then, could you benefit from allocating time to study the shaft-drive principles of **FIG. 2?** Could a similar drive arrangement power a small oil pump, and could pressurized oil be routed to a spray nozzle mounted a few millimeters from the rolling elements? Would that eliminate using vulnerable oil rings and constant level lubricators?

Finally, **FIG. 3** shows an example of the numerous feasible anti-swirl labyrinth configurations. These are needed to improve the rotordynamic stability of process gas compressors. Is it possible that choosing a particular configuration could extend the safe operating speed range of an existing compressor at your plant?

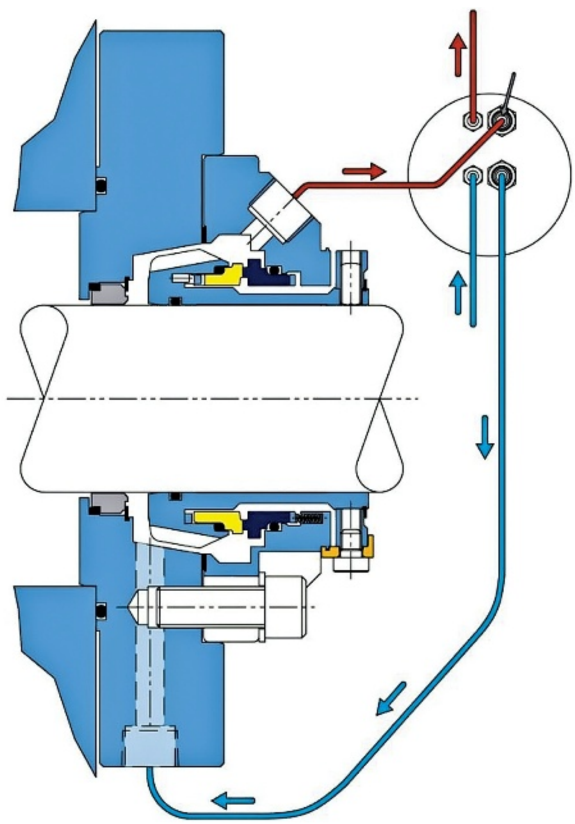


FIG. 1. Mechanical seal with bi-directional tapered pumping device.[1]

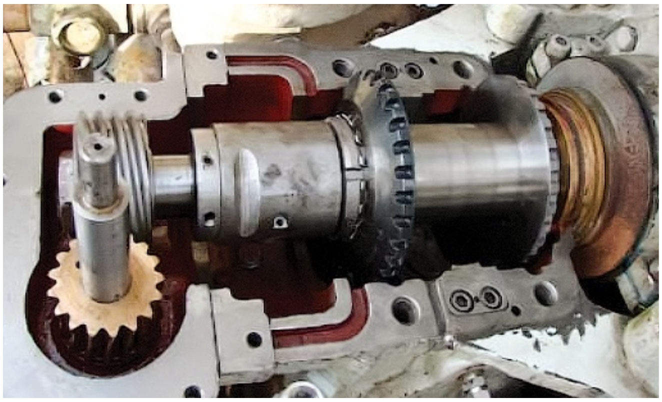


FIG. 2. View of the journal region of a small turbine. The worm gear arrangement on left—a similar layout could be used to power a spray-producing lube-oil pump for reliable lubrication of process pumps.

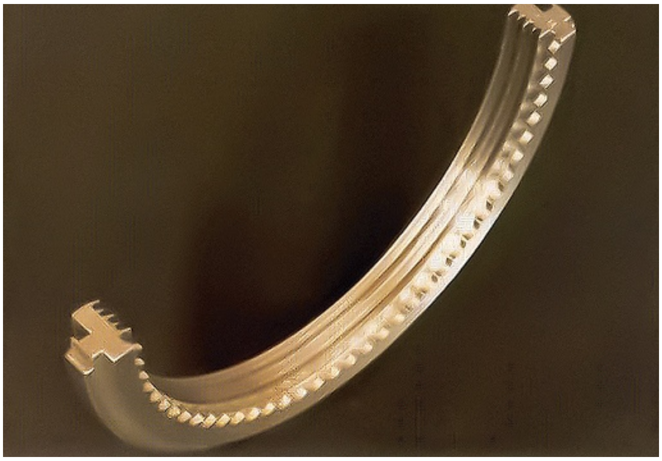


FIG. 3. Example of an anti-swirl labyrinth for a high-pressure centrifugal compressor.[2]

All three situations are just examples on how to make good use of "opportune times." Self-motivated professionals are already educating themselves and do not need our reminders. But others, and especially managers endeavoring to guide or groom their successors, may need some nudging.

Good managers know that checklists, sound work processes and well-written procedures are of great value to technical support staff. Checklists merit periodic updating and should be reviewed by personnel representing the three job functions that come into almost daily contact with process equipment: process unit operators, mechanical/maintenance work force members, and engineering/project/technical personnel.

Well-trained teams. All job functions must periodically update their knowledge, and reliability professionals must pay attention to details. They may start with an insistence on recording all types of measurements. To the appropriate extent, reliability professionals should be on the lookout for applicable new technologies. Concentrating on opportunities to stay informed on bearing lubrication and sealing issues is of great importance, and it provides quick returns on time invested. HP

LITERATURE CITED

1 Bloch, H. P., *Pump Wisdom*, John Wiley & Sons, Hoboken, New Jersey, 2011.

2 Bloch, H. P., and F. K. Geitner, *Compressors: How to achieve high reliability and availability*, McGraw-Hill, New York, New York, 2012.

HEINZ P. BLOCH resides in Westminster, Colorado. His professional career commenced in 1962 and included long-term assignments as Exxon Chemical's regional machinery specialist for the US. He has authored well over 520 publications, among them 18 comprehensive books on practical machinery management, failure analysis, failure avoidance, compressors, steam turbines, pumps, oil-mist lubrication and practical lubrication for industry. Mr. Bloch holds BS and MS degrees in mechanical engineering. He is an ASME Life Fellow and maintains registration as a Professional Engineer in New Jersey and Texas.

Evergreen Safety Lifecyle: Is your process safety data sustainable?

MIKE SCOTT / Mike.Scott@aesolns.com

Safety lifecycle documentation can be cumbersome and costly to maintain. Many organizations have identified a goal of evergreen documentation to reduce these maintenance efforts. Part of developing a sustainable evergreen safety lifecycle is to establish mechanisms for executing and tracking Management of Change (MOC). aeShield improves the MOC process by making changes easy to track and share across multiple projects with multiple users. In contrast to traditional databases which require a separate copy of the database for each project, aeShield utilizes a single database across multiple projects, maintains one master record, and tracks the status of lifecycle documentation automatically.

Project Management

Unlike traditional databases which only support one version of data, aeShield functions more like a document control system (**Fig. 1**). When project changes need to be made, a copy of the AsBuilt record is checked out. This allows changes to be applied without affecting the AsBuilt record and supports effective project management by allowing for ongoing concurrent project changes.

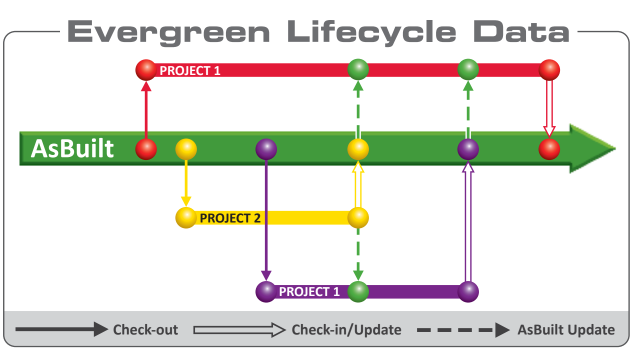


FIG. 1. Easily run concurrent projects from one AsBuilt record.

Document Management

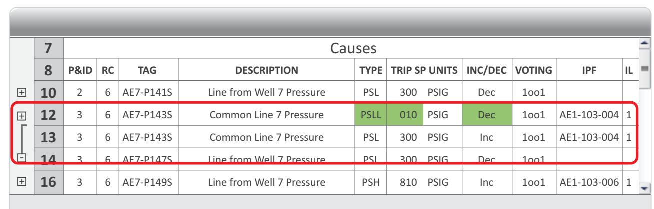
Document control allows project managers to identify competent personnel who will be tasked with the check and approve process per safety lifecycle documentation type. This helps ensure that personnel with the correct competencies are identified and help support the safety management plan. Individual users identified for these document control activities will receive dashboard alerts when the documents require action from them. Users tasked with checking and approving can view documents requiring action across multiple projects in the same dashboard. This helps manage effectiveness of high value resources by ensuring that the most competent people can be used over multiple projects and for gatekeepers to effectively monitor multiple projects.

Lifecycle Management

The revision shading, impact reporting, and status reporting in aeShield help facilitate MOCs by identifying and tracking changes and their impacts on individual documents and projects.

Revision Shading

Revision shading highlights project changes in a record while still providing a view of the original AsBuilt record. This facilitates the review and approval process by simplifying the comparison between the project changes and the original values (**Fig. 2**).



7	Causes										
8	P&ID	RC	TAG	DESCRIPTION	TYPE	TRIP SP	UNITS	INC/DEC	VOTING	IPF	IL
10	2	6	AE7-P141S	Line from Well 7 Pressure	PSL	300	PSIG	Dec	1oo1		
12	3	6	AE7-P143S	Common Line 7 Pressure	PSLL	010	PSIG	Dec	1oo1	AE1-103-004	1
13	3	6	AE7-P143S	Common Line 7 Pressure	PSL	300	PSIG	Inc	1oo1	AE1-103-004	1
14	3	6	AE7-P147S	Line from Well 7 Pressure	PSL	300	PSIG	Dec	1oo1		
16	3	6	AE7-P149S	Line from Well 7 Pressure	PSH	810	PSIG	Inc	1oo1	AE1-103-006	1

FIG. 2. Revision shading reduces checking and approval time.

Impact Reporting

Impact reporting facilitates cooperation between multiple projects affecting the same process safety documentation. This communication allows users to see changes from other projects and the status of those planned changes. Impact reporting also helps project teams reduce unnecessary reporting and identify potential conflicts that may not have been identified prior to project completion.

Status Reporting

Status reporting allows project managers to track the progress of ongoing projects. They can view the checking and approving processes on a record by record basis, as well as view what work is remaining for individual projects (**Fig. 3**).

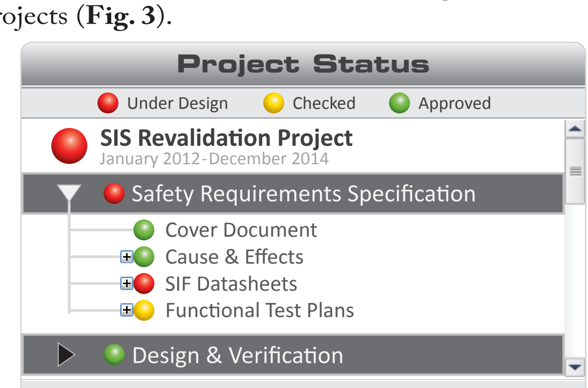


FIG. 3. Dashboard reporting provides high level status updates.

By allowing users to maintain and alter records in a single database with connectivity to original AsBuilt data, aeShield enables evergreen safety performance.

GREG GORBACH, CONTRIBUTING EDITOR
ggorbach@ARCweb.com

The information-driven operating company

New information technologies—such as cloud computing, mobility, social technologies, big data/analytics and 3D visualization—have been receiving a lot of attention in the industrial community. Why? Because each of these technologies has the potential to disrupt and radically change the way companies conduct business. Still, many owner-operators in the hydrocarbon processing industry (HPI) remain conservative and slow to embrace new information technologies. When they do, they often demand concrete references from industry pioneers before even considering any investment on information technology.

Pitfalls for the ultra-conservatives. However, this go-slow strategy can be far riskier than anticipated. New information technologies may disrupt not only existing practices in refineries and petrochemical plants, but also entire business processes throughout the supply chain and across the value network, and they may do so in fairly short order. The upstream and downstream environments are increasingly dynamic and volatile; new business models are emerging. In addition to a host of potentially disruptive technologies entering the marketplace, HPI operating companies must also manage rapid changes in government regulations, market dynamics, competition, feedstocks and even fluctuating weather patterns. ARC Advisory Group believes that, by deploying leading-edge information technologies, today's companies can overcome these obstacles and thrive.

New downstream competitors, unencumbered by legacy systems, are already leapfrogging to the latest technologies. Based on the new capabilities provided, these systems have the potential to serve global customers in completely new ways.

Operating companies that choose to stay at the back of the information technology pack will find it increasingly more difficult to reverse that decision. Not only is technology changing at an ever-faster pace, but the resulting hurdles to be cleared in the catch-up process will become even greater. Those companies with better software skills will be better positioned to succeed. Those that tend to run the latest software versions will be able to react quicker in response to changes. So, what strategies can owner-operators adopt to help better position them to take advantage of the latest round of technology changes?

21st century approach to technology adoption. On the whole, HPI companies have a well-known reputation for being conservative, especially when it comes to new information technologies. Energy companies often have a 20th-century mindset of "run it forever." This mindset avoids the cost of technology additions, replacements, updates and associated disruptions. It represented a winning strategy when conditions were fairly static, and the environment did not change very much or quickly.

In the past, this approach often paid off. With a "go-slow" approach to technology adoption, HPI companies could still obtain significant benefits, while avoiding the risk of reaching too far with technology and failing. But, in the last decade, that mindset began changing. Leading companies adopted a "fast follower" strategy to stay abreast of technology changes to ensure that they did not get blindsided by competitors who discover a way to use a new technology to their advantage.

At present, changes come at an accelerating pace. Technology is not only flattening, but it is also shrinking the world; engendering more and better competitors, more volatility and faster innovation. We are seeing increasing governmental regulations, unpredictable energy costs and raw materials shortages that drive up costs. In addition, the rapid introduction rate for new information technologies promises to enable dramatic, yet difficult-to-discern disruptions to the business processes, value networks and people of industrial companies. In this dynamic environment, companies that hold on to the 20th century technology adoption mindset actually face more risk from technology disruptions within their competitive ecosystem.

Of course, this risk does not apply immediately and it is equally distributed across all industries and all companies. It does not mean that owner-operators should rush out and invest in new technologies just for technology's sake without appropriate business justification. But the trend is clear. All operating companies should, at minimum, review and evaluate their technology adoption strategies to ensure that they are appropriate at this time and for the foreseeable future.

Information-driven operating companies. Information-driven owner-operators use core solutions extensively. They actively seek opportunities to leverage disruptive technologies to their competitive advantage. Information-driven operating companies have a bias to using the latest technology and implementing the latest software updates. These companies avoid building their own custom systems. Instead, they work in close partnership with core solution providers, use more of the latest available technology from those providers, and actively drive the providers to introduce new technologies to solve specific problems. HP

GREG GORBACH is vice president ARC Advisory Group. His focus areas at ARC Advisory Group include operations management, business process management, manufacturing business intelligence, plant-to-business connectivity, real-time performance management and operational excellence solutions. Mr. Gorbach has been with ARC since 1998. Prior to ARC, he was a product manager at an MES software company. He brings over 25 years of hands-on experience to ARC, including experience with both technology end-user and supplier organizations. His education includes a BSEE degree and an MBA.

LEE NICHOLS, DIRECTOR, DATA DIVISION
Lee.Nichols@GulfPub.com

Brazil plans major refinery expansions

Brazil is the largest producer of liquid fuels in South America, and this nation is the ninth-largest energy consumer in the world. The majority of Brazil's crude oil reserves are contained in the offshore pre-salt fields of the Campos, Espírito Santo and Santos basins. Located off the country's southeast coast, the pre-salt fields hold estimated reserves of approximately 100 billion barrels of oil equivalent. These pre-salt oil deposits could possibly transform Brazil into one of the world's largest oil producers.

Despite abundant hydrocarbon reserves located just offshore, Brazil still lacks the installed refining capacity to meet its surging domestic demand for transportation fuels. At present, the Brazilian refining industry operates 13 refineries with a throughput capacity of just over 2 million bpd (MMbpd). Even though Petrobras, a state-owned oil company, posted record oil production of 2.2 MMbpd from June 29–30, 2013, this nation still suffers from a shortfall of refined transportation fuels and must import refined products to cover the gap between supply and demand.

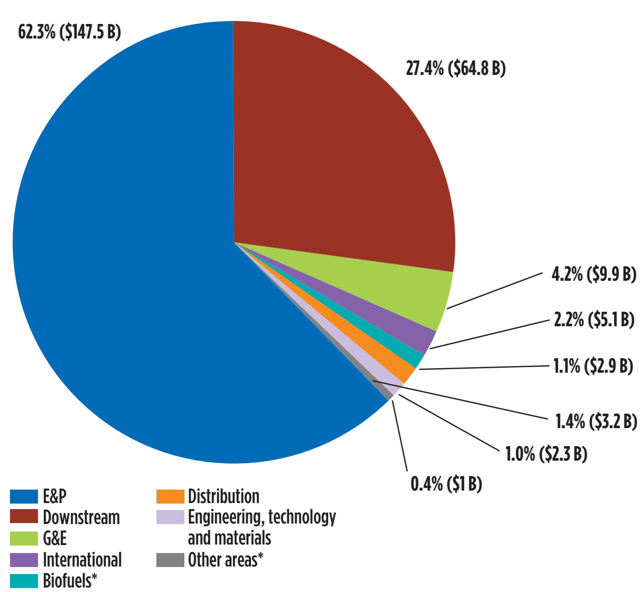
Investments. In its 2013–2017 investment plan, Petrobras has allotted $237 billion (B) for capital expenditures (CAPEX) to boost crude oil production and the development of ultra-deepwater offshore fields, as shown in **FIG. 1.** Almost $65 B is allocated for the downstream sector. The downstream investments are focused on refining capacity additions, as shown in **FIG. 2.** Petrobras plans to nearly double domestic oil production by 2020. In addition, the company will reduce the energy intensity at its refining operations by 10% and lower refinery CO_2 emissions by 8%.

To meet these new capacity expansion goals, Brazil is planning an aggressive wave of construction projects. The centers involve four major projects: the Abreu e Lima, Comperj, and Premium 1 and 2 refineries (**FIG. 2**). Domestic demand for transportation fuels is forecast to reach 3.25 MMbpd by 2018. With the new refining capacity additions announced, Brazilian refining capacity will rise to 3.6 MMbpd. This surplus will enable Brazil to export petroleum products.

Abreu e Lima. It has been 32 years since Brazil constructed a grassroots refinery. The Abreu e Lima complex breaks that trend. This refinery is under construction at the Suape Industrial Port Complex in the state of Pernambuco, Brazil. The project is a joint venture (JV) between Petrobras and Venezuelan state-owned oil company PdVSA; the JV agreement was signed in 2008. Petrobras holds a 60% stake in the project, with PdVSA owning the remainder. Under the JV agreement, PdVSA is required to pay 40% of the construction costs in cash. At present, Petrobras has yet to receive any payments from PdVSA.

The $17 B, 230,000-bpd (Mbpd) refinery will process heavy crude oils from Venezuela and Brazil. Construction will be completed in five phases:

- **Phase 1**—Identify the scope for the refinery project



* Pbio = Petrobras Biofuel | ETM = Engineering, Technology and Materials | Other areas = Financial, Strategy and Corporate
Phase I: Opportunity Identification; Phase II: Conceptual Project; Phase III: Basic Project ; Phase IV: Execution

FIG. 1. Petrobras' investment plans for 2013–2017.

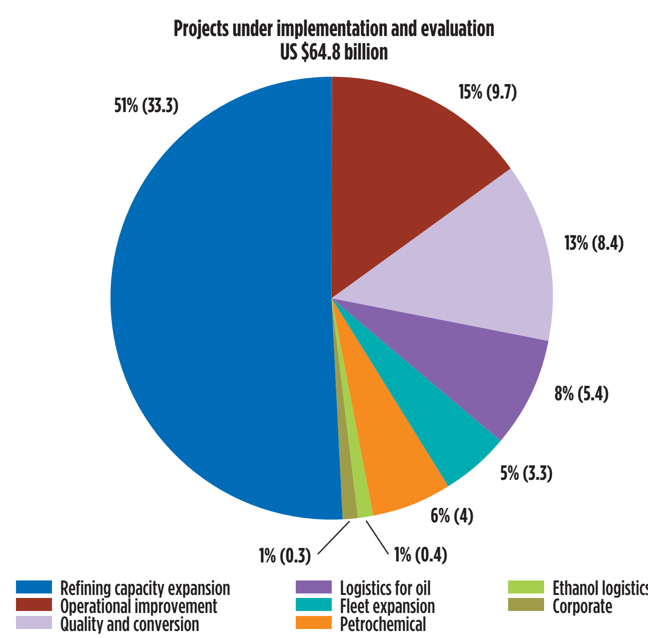


FIG. 2. Downstream investment breakout by sector.

- **Phase 2**—Develop the concept phase in which investment and complex capacity are set (completed in late 2006)
- **Phase 3**—Conduct the feasibility study (involved an adjustment in the proposed refinery's throughput capacity to 230 Mbpd and was completed in 2009)
- **Phase 4**—Site execution (included the completion of earthworks and the installation of infrastructure equipment and systems)
- **Phase 5**—Commence construction of the refinery (a completion date is set for late 2014).

The processing units for the new refinery will include atmospheric distillation, diesel hydrotreating, hydrogen generation, naphtha hydrotreating and regenerative caustic treatment, as well as two delayed cokers, control rooms and substations, and a water-treatment plant. Once completed, the refinery will provide transportation fuels for the northeastern region of Brazil.

Comperj. The $8.4 B Rio de Janeiro Petrochemical Complex (Comperj) is an integrated refining and petrochemical complex located in the city of Itaboraí outside of Rio de Janeiro. The greenfield project will process an initial capacity of 165 Mbpd of heavy oil from the Marlim field in the Campos Basin. Due to environmental license issues and changes to the refinery's configuration, completion of this new refinery has been delayed to 3Q 2016. A second phase is planned; it will double the complex's throughput by 2018.

The facility will also include basic and second-generation petrochemical units. The basic units will produce ethane, propane, benzene, paraxylene and butadiene. The second-generation units will upgrade basic-unit product streams into styrene, ethylene glycol, polyethylene, polypropylene, purified terephthalic acid and polyethylene terephthalate. **TABLE 1** summarizes the major project/licensors for the Comperj project. Upon completion, this complex is forecast to save Petrobras over $2 B per year in oil derivatives and petrochemical product imports.

Premium 1. Petrobras and China Petroleum & Chemical Corp. (Sinopec) are conducting a feasibility study to construct Brazil's largest refinery. The Premium 1 refinery will have an initial processing capacity of 300 Mbpd. Expansion plans will eventually double the site's capacity to 600 Mbpd by 2020. Both Petrobras and Sinopec have signed a letter of intent to study a possible JV.

The two-train project will be located in the northeastern state of Maranhao. Major contracts awarded include:

- UOP LLC is the front-end engineering and design (FEED) contractor and is a technology licensor. Premium 1 will utilize UOP's Unicracking hydrocracking process and Unionfining hydrotreating technology.
- Foster Wheeler (FW) is subcontracted to perform project FEED and is a technology licensor. FW will supply its Selective Yield Delayed Coking (SYDEC) technology.

Other major processing units include atmospheric distillation, amine generation, a hydrocracker, a diesel desulfurization unit, a delayed coker, a hydrogen

unit, a sour-water stripper unit and a sulfur recovery unit. Completion for Phase 1 is scheduled for 2017, and Phase 2 will be finished by 2020.

Premium 2. The Premium 2 refinery project is almost an exact mirror of the Premium 1 project. Completion of the 300-Mbpd refinery is scheduled for 2018. Petrobras is studying a possible JV with South Korean energy company GS Energy for this project. The refinery will be constructed in the northeastern state of Ceara.

Both UOP and FW were awarded contracts for FEED and technology licensing. FW will perform the basic design and FEED for the main process units, and will supply its SYDEC technology. UOP will be the FEED contractor, along with providing its Unicracking hydrocracking and Unionfining hydrotreating technologies. Design of the crude and vacuum systems will be provided by Process Consulting Services.

Construction of the Premium 1 and 2 refineries is imperative for Brazil to curb diesel and gasoline imports. Without these plants, Brazil's demand for refined products would outpace supply by roughly 1 MMbpd by 2020. HP

TABLE 1. Major technology awards for Comperj

Technip Stone & Webster Process Technology	Provide technology for the ethylene plant and petrochemical fluid catalytic cracking (FCC) unit
WorleyParsons	Awarded $110 MM service contract for integration and project management services, FEED for utilities and offsites, technical assistance during detailed design, commissioning, pre-operation, startup and performance testing
Skanska AB	Design, detailed engineering and construction of refining units
Toyo Engineering Corp.	Construct utilities and hydrogen production facilities, including detailed design, procurement of equipment and materials, installation and commissioning support
Axens	Supply its ParamaX technologies (paraxylene technology suite), as well as middle distillate, kerosine, pyrolysis gasoline and heavy-naphtha hydrotreating units

LEE NICHOLS is director of Gulf Publishing Company's Data Division. He has five years of experience in the downstream industry and is responsible for market research and trends analysis for the global downstream construction sector.

BILL KLESSE
Chairman of the Board and Chief Executive Officer,
Valero Energy Corp., San Antonio, Texas

Renewables from the refiners' viewpoint

BILL KLESSE serves as chairman of the board and chief executive officer of Valero Energy Corporation. Mr. Klesse joined Diamond Shamrock, now Valero, in 1969 as a junior process engineer and has worked in various managerial positions in engineering, petrochemical feedstocks, planning and development, and marketing. Mr. Klesse received his undergraduate degree in chemical engineering from the University of Dayton, and an MBA degree with an emphasis in finance from West Texas A&M University.

Recently, I had the opportunity to testify before the US Senate Committee on Energy and Natural Resources, representing our company and our industry that employs approximately 108,000 American workers and four times that many in support industries. In particular, I addressed the federal policy on renewable fuels, which affects consumer prices and Valero's operations.

Valero, based in San Antonio, Texas is the world's largest independent petroleum refiner. Valero does not explore for, or produce, crude oil or natural gas. Rather, we purchase these and related feedstocks to manufacture refined products such as gasoline, jet fuel and diesel. Valero is also a leading ethanol producer, with 10 plants that together can produce 1.2 billion gallons of corn ethanol per year. Because we are in both the petroleum fuels and the ethanol business, we believe Valero approaches energy policy from a uniquely pragmatic, rather than strictly ideological, point of view.

Although the price of crude oil represents by far the largest component of gasoline prices, other factors influence retail gasoline prices, some of which can be affected by government policy. Policies that make it more difficult to make transportation fuels in the US are contrary to the public's interest.

One of the most challenging factors facing the fuels marketplace is the implementation of the Renewable Fuels Standard (RFS) and a 10% ethanol (E-10) blend wall. As the third largest corn ethanol producer in the US, and now a producer of renewable diesel fuel, Valero supports alternative fuels and believes that they are an important part of the nation's transportation fuels. But whether or not one supports alternative fuel production, it's obvious that the RFS is broken and needs to be fixed.

Under the RFS, refiners and importers—not blenders—are obligated to make sure that the mandated amount of ethanol is blended into gasoline. This is done through the submission of renewable identification numbers, or RINs. US refiners that are merchant spot-market sellers do not generate RINs; instead, they must buy them. The US Environmental Protection Agency (EPA) originally said that RIN prices would be negligible, and compliance was not an issue with the original 2005 law (Energy Policy Act of 2005).

However, obligations were increased in the 2007 (Energy Independence and Security Act of 2007) law at the same time gasoline demand was beginning to decline. The amount of ethanol required to be blended into gasoline under the RFS keeps rising; but domestic gasoline consumption has

FIG. 1. Valero's Corpus Christi petroleum refinery was commissioned in 1983. It is one of the world's most complex refineries.


status quo loves change
Challenge the status quo and discover transformational change in your refinery. BASF's innovative FCC products, services and solutions deliver value to enhance sustainability and performance.
At BASF, we create chemistry for a sustainable future.
www.catalysts.basf.com/refining
BASF
The Chemical Company
Select 70 at www.HydrocarbonProcessing.com/RS

significantly declined. As a result, obligated parties have to purchase additional RINs to make up shortfalls against their renewable volume obligation. In response to increasing demand, RIN prices have shot up from a few cents to a recent record high over \$1.50, or 15¢/gal of E10 gasoline. Earlier this year, we estimated that Valero's costs due to skyrocketing RIN prices would be between \$500 million and \$750 million this year alone. Now, we are at the high end of that estimate, and it could go even higher. These costs are eventually passed onto consumers in higher prices. Also, the playing field is very unfair and uneven among companies in the same market segment.

In addition, the RFS requires the use of cellulosic ethanol, a product that does not yet exist outside of laboratories. At present, cellulosic ethanol is totally uneconomic, and to meet the Advanced Biofuels Mandate, you have to import sugarcane ethanol.

The EPA's solution to the blend wall is to blend higher levels of corn ethanol into gasoline, going from 10% ethanol to 15% ethanol (E15). However, the idea of going to E15 is not practical because its use has not been approved in *all* automobiles or in boat motors, motorcycles, etc. Several automakers will void warranties if drivers use E15, and the US American Automobile Association has recommended against it. This does not even mention the impact on global food prices from the increased use of corn for fuel.

It is time to eliminate the RFS and start over. Valero supports alternative fuels—and what makes sense to Valero, from a pragmatic point of view, is to link ethanol blending to 10% of actual gasoline usage, and to make those who do the blending of ethanol the obligated parties under the RFS, instead of putting that burden on refiners. Also, a new RFS should not require the importing of sugarcane ethanol.

Revolutionary things are happening in North American crude oil and natural gas production, creating huge economic opportunities. Manufacturing jobs can come back to the US with the correct policies. We welcome Congress's attention in helping to make this happen. HP

FIG. 2. Valero is the third largest US corn-based ethanol producer with a capacity of 1.2 billion gal/year; the refiner operates several ethanol plants such as the plant in Albert City, Iowa.

HP | Special Report

REFINING DEVELOPMENTS

Demand for crude oil and refined products is increasing. However, the new demand centers for transportation fuels are shifting to developing nations. Sustainability and pending environmental rules will require refiners to make adjustments to operations and processing schemes. In addition, the definition for "cleaner" fuels is evolving. To remain profitable, the refining industry must be flexible to adapt to changing economic conditions, availability of different feedstocks, mandated transportation fuel specifications and more.

Photo courtesy of BP Products North America.

N. V. KARPOV, Lukoil Nizhegorodnefteorgsintez Refinery, Russia; and **C. KEELEY, J. MAYOL, V. BOZUKOV, S. RIVA, V. KOMVOKIS** and **S. CHALLIS,** BASF Corp., Iselin, New Jersey

Case history: Optimization of FCCU with multivariate statistical modeling

Process models and simulation methods can be used to simulate the fluidized-bed catalytic cracking unit (FCCU) process.[a,b] Likewise, kinetic or multivariate statistical models have been used. Rigorous non-linear reactor kinetic models have typically been applied to develop project design, support refinery planning, etc. However, kinetic models have had limited applications in maintaining FCCU operations and in optimization projects for many reasons. Kinetic models tend to be time-consuming and expensive to build and run. They are also difficult for the refinery engineers to maintain due to their complexity.

Solutions. Multivariate statistical models based on operating data and using standard software can provide a suitable alternative to support process optimization. These models can be readily and cost-effectively developed. Such models can be used for FCCU troubleshooting, along with optimization and training purposes. They are also suitable for real-time process monitoring and are suitable for evaluating changes in feed and operating variables.

Several generic examples will demonstrate the power of multivariate statistical process models. For example, a case history will illustrate the development of accurate models for the Lukoil's Nizhegorodnefteorgsintez (NNOS) refinery.[c] The Lukoil NNOS models can accurately estimate online FCCU product yields, gasoline research octane number (RON) and regenerator bed temperature (RBT). With this information, Lukoil was very successful in meeting the new market demand for premium gasoline by optimizing the catalyst, feed and FCC operating conditions.[1]

Background. In August 2005, Lukoil announced a major investment to upgrade the NNOS refinery.[c] This investment was in response to increased gasoline demand by passenger vehicles.[2] For Russia and the CIS, demand for regular gasoline was replaced in favor of premium gasoline with the migration to Euro 5 standards.[3]

In 2005, a licensor was selected to provide the design for a new FCCU. The new unit was successfully commissioned and operational in 2011.[4] The unit is a modern, short contact time riser design FCCU.[d,1] The FCC catalyst uses a distributed matrix structures (DMS) technology platform.[e] DMS technology enables high bottoms conversion with low coke make; all contributing to higher yields of gasoline and light olefin products.[5] In addition to catalyst, a technical service team assisted the NNOS refinery and was requested to build accurate FCC process models and provide simulation capability.[c] Application of multivariate statistical modeling was selected as the best approach to meet the refiner's needs.

Benefits from statistically modeling an FCCU. Rigorous non-linear reactor kinetic models have been used to develop project design yields and to support refinery planning, especially in defining linear programming sub-models and evaluating new FCC feeds. However, applications of kinetic models to support FCCU operations and optimization projects are less common due to the significant effort to produce accurate results for each operating scenario.[6] Kinetic models, even if more accurate, require more specialized knowledge to build and calibrate. **Result:** Kinetic models are time-consuming and costly to run; in addition, they are difficult for the refinery engineers to maintain.

An attractive alternative to kinetic models is using multivariate statistical models based on unit operating data. These models can be readily and cost-effectively developed by refinery engineers with the support of their catalyst supplier. Benefits of multivariate statistical models include:

- Easy to build using standard software[f]
- Does not require a detailed knowledge of the FCC hardware design to develop
- Easy to maintain and update, e.g., by updating models if conditions change
- Enables a detailed understanding of the impact of process variables on unit operations in a transparent and user-friendly

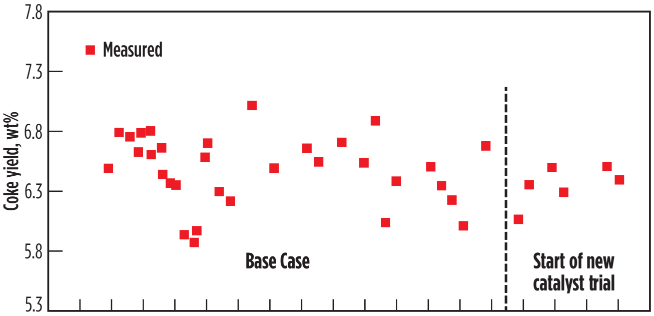


FIG. 1. Time series of coke yield (Customer A).

way; they can be used for unit troubleshooting, optimization and training purposes.

- Suitable for real-time process monitoring to detect deviation from expected base operation behavior.
- Easy to use in the prediction mode; they can be used to examine the impact of different feed qualities, change in reactor temperature, etc.

POWER OF STATISTICAL MODELS

There are many good reasons why a refiner should consider application of statistical modeling to an FCCU, as illustrated in several examples.

Example 1: Real-time monitoring of catalyst and SO_x. In this example, an FCCU was experiencing coke-yield limit issues. Following the change to a new catalyst, initially the operating data trended as the previous condition, as shown in **FIG. 1.** Later, coke yield began to increase (**FIG. 2**). At this point, the operations and management team became concerned. The issue was explaining why the coke yield was increasing and how to reverse the trend.

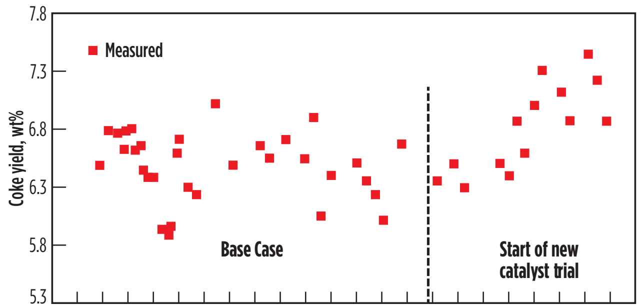


FIG. 2. Time series of coke yield (Customer A).

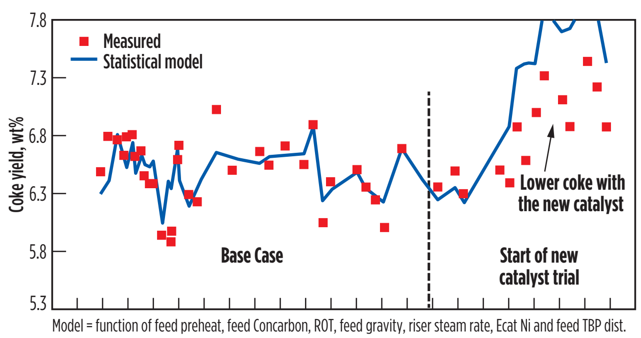


FIG. 3. Time series of coke yield (Customer A), process model estimate.

One approach would be to use a valid process model to set the baseline before the catalyst changeover. Then the team could evaluate the process change due to the new catalyst. It would be possible to estimate the expected coke yield for the present operating conditions. By comparing the estimate with the present data for the new catalyst, operations could decide whether the present coke yield is satisfactory (**FIG. 3**). Based on **FIG. 3,** the coke yield for the new catalyst is less than expected for the base catalyst at the present operating conditions. But why? Engineers could use the knowledge captured in the process model to identify which process variables are contributing the most to the higher coke production.

The model in this example is coke yield as a function of feed preheat, feed Conradson carbon residue, riser operating temperature, feed gravity, riser steam rate, equilibrium catalyst nickel content and feed distillation. In this specific example, Ecat activity and feed rate have not been included in the correlation since these variables did not vary significantly during the trial and, therefore, statistically weren't relevant. However, in general, coke yield is influenced by these variables and would be included in the correlation when Ecat activity and feed rate vary significantly. **FIG. 3** confirms that the new catalyst is more coke selective than the base catalyst for the same processing conditions. Also, the previous model now needs recalibrating to accurately forecast performance with the new catalyst.

In another example, multivariate statistical modeling was used for real-time monitoring of FCCU regenerator flue-gas sulfur dioxide (SO_2) emissions during a SO_2 reduction additive trial, as shown in **FIG. 4.** In the past, refiners applied a simple relationship between SO_2 and FCCU slurry-oil sulfur to monitor a SO_x additive trial. Unfortunately, this practice often had poor correlation, i.e., a more sophisticated approach is justified. As shown in **FIG. 4,** the multivariate statistical modeling can provide an accurate estimate of regenerator flue-gas SO_2 emissions. It can be used in real-time monitoring for the performance of a new additive trial. In this specific example, the new SO_x reduction additive had an improved, higher pick-up factor.

Example 2: Estimation of flue-gas emissions. This example shows how real-time monitoring and estimation of coke yield was possible even during a failure of the online flue-gas analyzer (**FIG. 5**). It demonstrates how a multivariate statistical model can be used, *with care,* for a short period to continue the operation

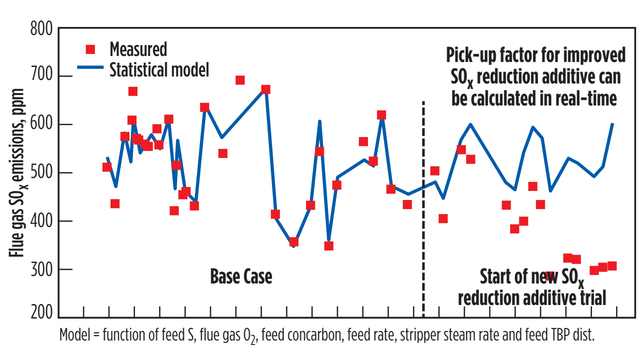


FIG. 4. Real-time flue-gas SO_x monitoring (Customer B).

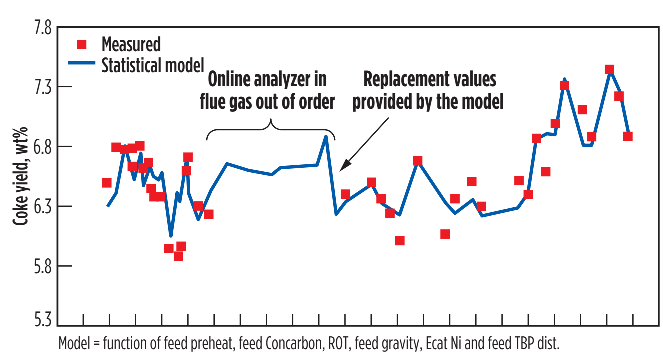


FIG. 5. Real-time monitoring and estimation of coke yield (Customer C).

of a unit. Applying the model avoided other consequences that could have caused a unit shutdown, while repairs were made to the analyzer. Furthermore, statistical modeling can be considered for estimating, in real time, FCCU flue-gas emissions—e.g., SO_2, nitrogen oxide and carbon dioxide—to enhance emissions monitoring and reporting to regulators. Even when the flue-gas analyzer is working, many FCCU reports indicate that the analyzer is not always providing reliable measurements. Thus, flue-gas emission models can also be used to highlight issues with the analyzer.

Procedure for multivariate statistical model. The main process steps to develop a multivariate statistical model are summarized in **FIG. 6.** The steps include:

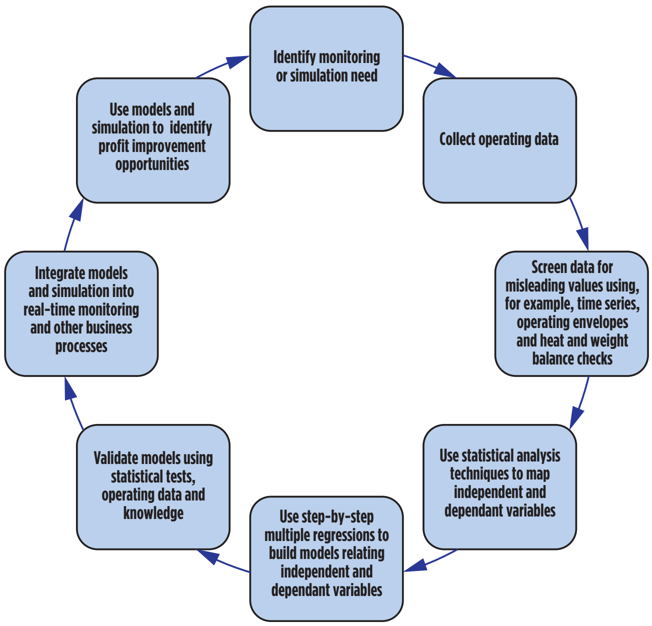


FIG. 6. FCC operating data analysis and statistical modeling process.

1. Unit operating data is conveniently stored in MS Excel (or similar software).[f]
2. These data are initially screened using basic graphical techniques to identify misleading values. At this stage, it is helpful to plot time series and compare them to known unit operating envelopes and constraints. These same graphs can also be used to identify poorly behaving instrumentation.
3. Once the obvious misleading values are excluded, then unit heat and weight balances are built. This is used to screen out datasets with poor heat and weight balance closure.
4. The good data are transferred to statistical analysis software.[f] The FCCU engineer often has a good feeling for which independent variables impact dependent variables. It is important that the regressions developed have a basis supported by a theoretical understanding of the FCC process. Using standard statistical techniques, the relationship between independent and dependent variables can be confirmed.
5. The next step is step-by-step multiple regressions to build models relating to independent and dependent variables.
6. To validate these models, the results are compared to real operating data. The statistical tests and operating knowledge are used to fine-tune the models.
7. The result is a set of models relating independent to dependent variables that are in a format that can be easily entered into a spreadsheet or the control system. If required, these models can be combined with a graphical user interface (see **FIG. 7**) and be used in a purely prediction mode to examine new feed, operating scenarios, etc. If this is done, then it is wise to document clearly the limit of validity for each model.

This work process was used to build models for the Lukoil's NNOS refinery FCCU.[c]

Example 3: NNOS refinery FCCU models.[g] In response to the growing demand for premium gasoline, Lukoil's NNOS refinery choose the DMS catalyst technology.[c, e] This catalyst technology enables high bottoms conversion

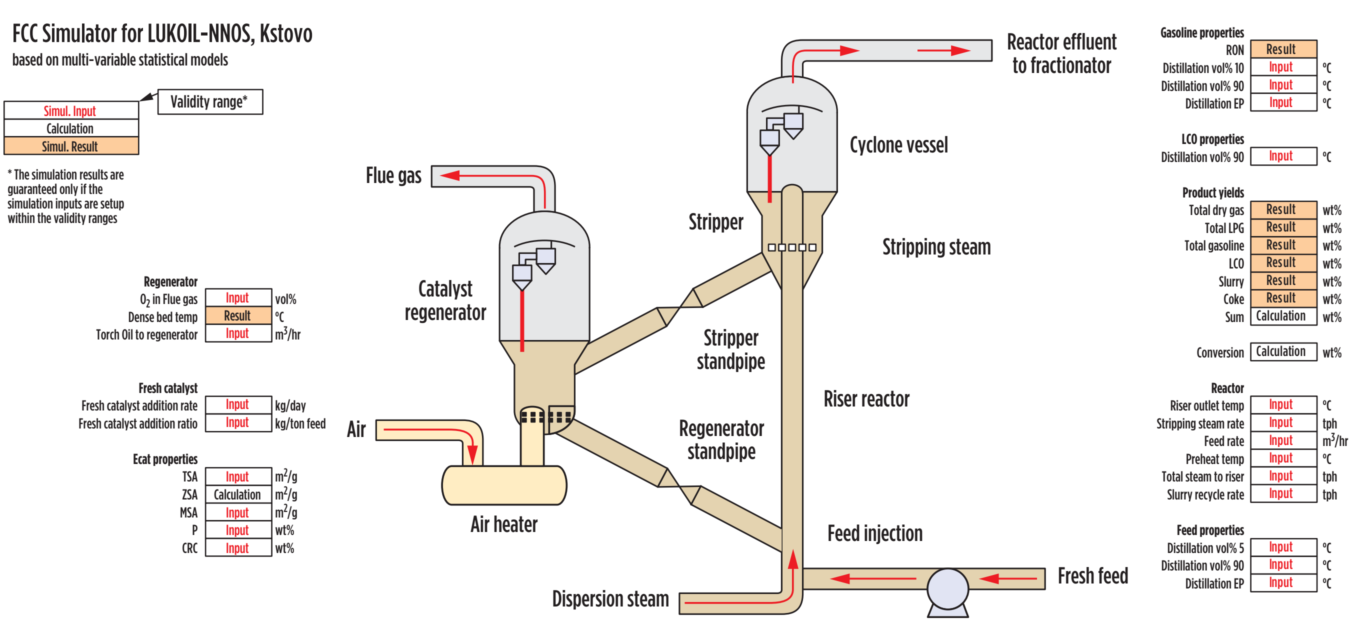


FIG. 7. FCCU process simulator interface.

with low coke for higher yields of gasoline and light olefin products. In addition to catalyst, accurate process models were built for the Lukoil NNOS refinery.[c] When faced with changing feed and operating objectives, these models, used in real-time, help detect deviations from the production plan. Thus, these models assisted the Lukoil NNOS refinery to maintain operating conditions close to the optimum and to improve profitability.[c]

FIGS. 8–10 show the commercial unit operating data and the real-time multivariate statistical model estimate for liquefied petroleum gas (LPG) yield, gasoline yield and gasoline RON, respectively. From these figures, the estimate values provided by the models trended very close to actual conditions. Thus, the models provided *good estimates with high accuracy as processing conditions change.*

For example, with reference to **FIG. 8,** the model's coefficient values and the form of the equation will depend on the specific FCCU design, catalyst and operating conditions. Therefore, the models will be different for each FCCU and, they will require recalibration each time there is a significant equipment or catalyst change.

The Lukoil NNOS refinery FCCU feed is a severely hydrotreated vacuum gasoil. Therefore, it has very low levels of coke precursors.[c] Furthermore, the unit operating objectives

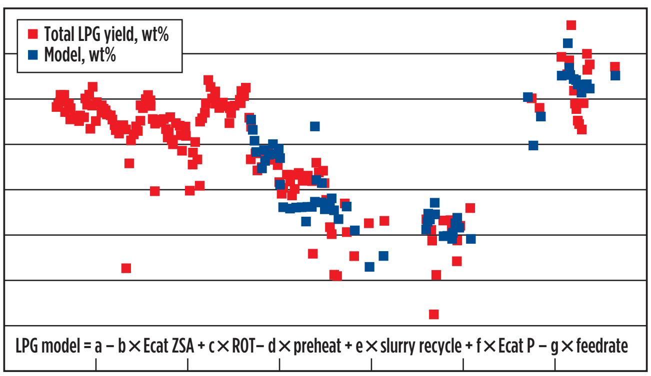


FIG. 8. Time series of LPG yield.

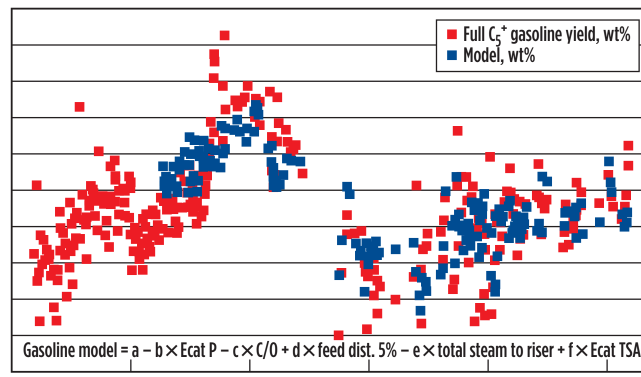


FIG. 9. Times series of full-gasoline yield.

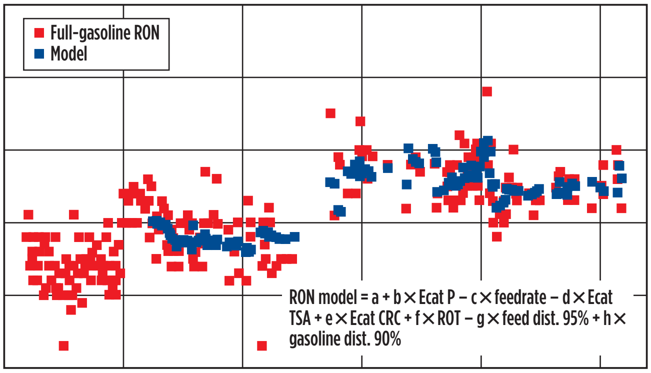


FIG. 10. Times series of full-gasoline RON.

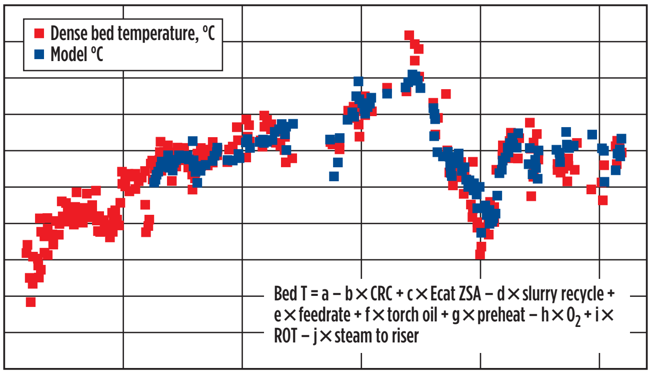


FIG. 11. Time series of RBT.

are to maximize bottoms conversion with low coke, and to deliver high yields of gasoline and light olefin products. This is achieved using a catalyst that is customized to the unit's short-contact time design. Combining a low-coke make feed with a coke-selective catalyst leads to a low unit coke production and a low RBT. To manage this low RBT during feed and operating condition changes, an accurate model to estimate the RBT was built, as shown in **FIG. 11.** The model accurately predicted the RBT for a large range of operating conditions.

In addition to real-time monitoring of product yields, gasoline RON and RBT, Lukoil's NNOS refinery needed an FCCU simulator to examine new feed and/or operating scenarios.[c] A cost-effective simulator was built in MS Excel by combining the unit-specific models with a graphical user interface. Conditional formatting and comments on specific cells were used to clearly document the validity of the models. **FIG. 7** shows an example of the graphical user interface. This interface provides a convenient way to enter user inputs and also summarize the results. This page can be saved or printed to create a record of the case. Alternatively, the results can be archived if a study is being done to estimate the response of the unit to step-wise changes in the feed or operating conditions.

Observations. Multivariate statistical models, based on unit operating data and using standard software, can be successfully used for support of FCC unit operation and optimization. These models can be readily and cost-effectively developed by refinery engineers and catalyst suppliers, as demonstrated by Lukoil at the NNOS refinery.[1] The success of this project, combined with favorable market conditions, has resulted in Lukoil's decision to duplicate the design for a second complex.[7] HP

NOTES

[a] The model represents the key characteristics or behaviors of the process.
[b] Simulation uses models to imitate the operation, enabling the examination of new feed and/or operating scenarios.
[c] Lukoil Nizhegorodnefteorgsintez (NNOS) refinery is located in the Nizhny Novgorod area of Russia.
[d] The basic engineering design for the process, technology and equipment is provided by UOP. The unit features side-by-side vessel layout, optimix feed distribution system, vortex separation system riser termination, advanced fluidization reactor stripper technology, RxCat riser technology and combustor style high-efficiency regenerator.[1] The licensor's process technology and equipment, in combination with the appropriate catalyst, enabled the Lukoil's NNOS refinery to maximize profitability by achieving best-in-class conversion, total liquid product yields and olefin selectivity.[7]
[e] The catalyst used by the Lukoil NNOS refinery FCC unit is a customized version of BASF's NaphthaMax catalyst.
[f] Easy to build using standard software such as MS Excel and Statsoft Statistica.
[g] Specific numbers have been removed to protect commercially sensitive data.
[h] The customer is a refinery in Southern Europe.

LITERATURE CITED

1. M. Griffiths, "Highly Successful Implementation of Lukoil's Nizhny Novgorod Refinery," 12th Russia and CIS Refining Technology Conference and Exhibition, Sept. 20–21, 2012, Moscow.
2. Lukoil Presentation, London, UK, March 2012.
3. Hart Energy IFQC Flash Alert, Sept. 9, 2011.
4. *Nizhegorodskaya Delovaya Gazeta*, #12 (111), Dec. 21, 2012.
5. McLean, J. B., W.A. Weber, and D.H. Harris, "Distributed matrix structures—A technology platform for advanced FCC catalytic solutions," 2003 NPRA Annual Meeting, San Antonio, Texas, March 2003.
6. Llanes, J. M., M. Miranda, And S. Mullick, "Use modeling to fine-tune cracking operations," *Hydrocarbon Processing*, September 2008.
7. Lukoil/Honeywell UOP press announcement.

NIKOLAY V. KARPOV is the chief engineer and first deputy general director of Lukoil's Nizhegorodnefteorgsintes refinery.

CARL KEELEY is BASF's marketing manager, refining catalysts for EMEA. He earned his MEng in chemical engineering and applied chemistry from Aston University. Mr. Keeley is a professional engineer with over 12 years of experience. Previously, he worked for UOP, BP and Dow.

JEREMY MAYOL is a BASF technical account manager, refining catalysts for EMEA. With over 15 years of experience, he is a specialist in FCCU operation and process simulation. Mr. Mayol has worked for BASF for four years. Prior to this, he spent 12 years working for INEOS and BP.

VASIL BOZUKOV is Lukoil's technical account manager, refining catalysts for CIS and Russia for BASF. He joined BASF from Lukoil's Bourgas refinery, where he worked for eight years in refinery operations, including FCCU. Mr. Bozukov has a diploma in chemical technology from Bourgas University.

STEFANO RIVA is a technical service manager, refining catalysts for EMEA. He has over 20 years of FCC experience—10 years with Engelhard and BASF in technical sales and service; two years with Tamoil, and eight years working in Exxon Mobil's technical support group.

DR. VASILEIOS KOMVOKIS is BASF's technology manager for refining catalysts for EMEA. He holds BS and MS degrees in chemistry and a PhD in chemical engineering from Aristotle University. Dr. Komvokis was a researcher at CPERI and a research professor at the University of South Carolina.

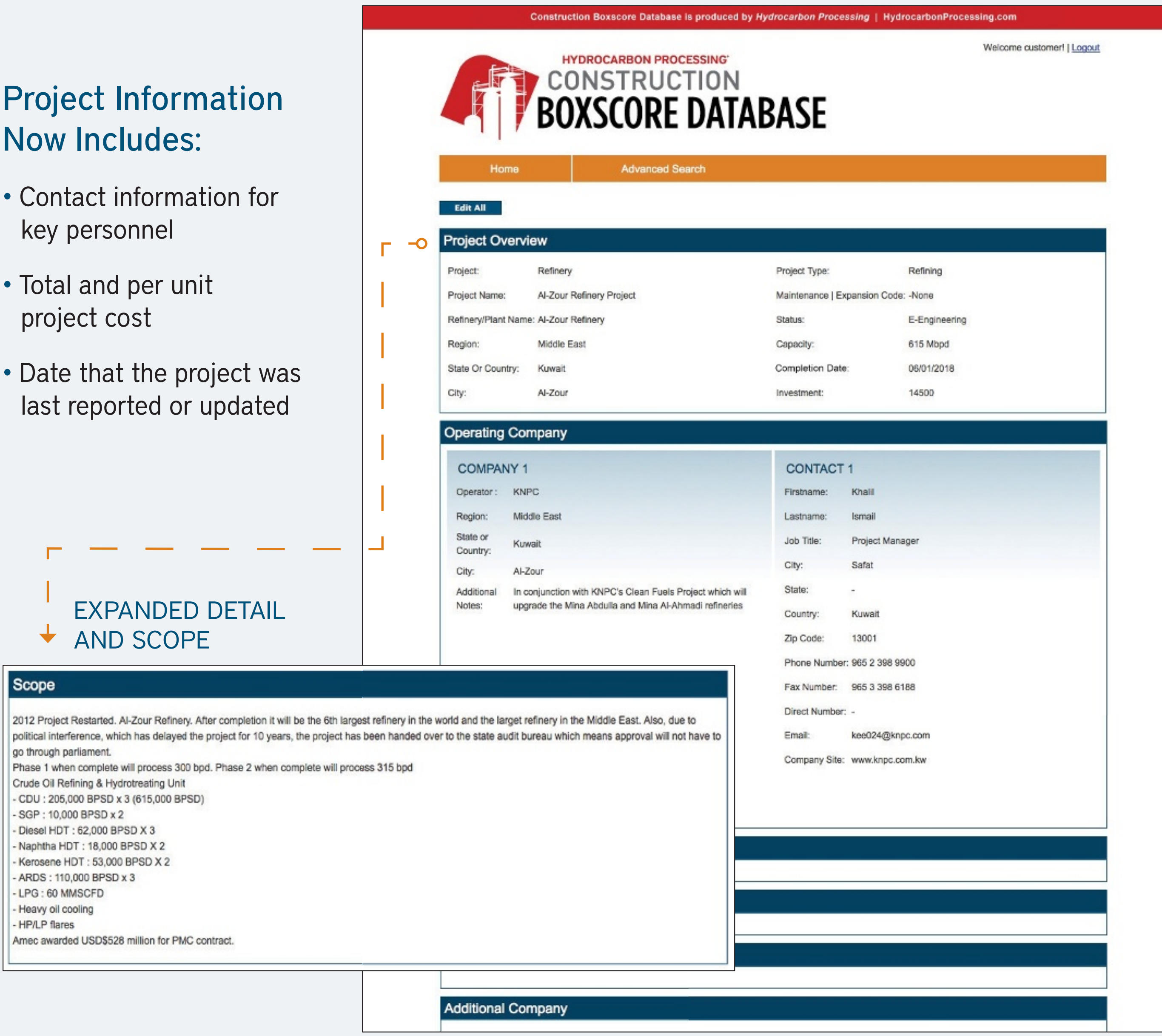
STEPHEN CHALLIS is a FCC technical advisor for refining catalysts with BASF. Mr. Challis has been a technical advisor for three years. Prior to this, he worked for ExxonMobil for 29 years, with 22 years spent providing technical support for FCCUs in EMEA.

- Project details for more than 3,500 active, global construction contracts including contact information for key personnel
- Advanced search that filters the listings by project type, scope, region, investment and more
- Daily updates for new and newly updated projects
- The weekly Boxscore Update e-newsletter with new listings and trends analysis

Project Information Now Includes:

- Contact information for key personnel
- Total and per unit project cost
- Date that the project was last reported or updated

EXPANDED DETAIL AND SCOPE



Sign up for a Boxscore Demo

ConstructionBoxscore.com/Demo

Subscribe Today

ConstructionBoxscore.com/Subscribe

Lee Nichols
Director of Data Division

Phone/Fax:
+1 (713) 525-4626
Lee.Nichols@GulfPub.com

K. J. BRYDEN and **E. T. HABIB JR.**, Grace Catalysts Technologies, Columbia, Maryland; and **O. A. TOPETE,** Grace Catalysts Technologies, Houston, Texas

Processing shale oils in FCC: Challenges and opportunities

As novel hydraulic fracturing technology with directional drilling continues to improve, shale oil will continue to be a game-changer for North American refiners. Although credited with many advantages, shale oil does not come without its challenges. Suppliers and processors alike are urgently working to adapt to the changing oil landscape. Just a few years ago, investments were focused on processing heavy crudes. Now, however, the industry is faced with lighter, sweeter crude streams from shale plays.

In varying degrees at each refinery, shale oil makes up only a percentage of the total feedstock. Present estimates put shale oil production at 5% of total US crude demand. The percentage could grow substantially as shale oil production increases and refiners invest in process modifications to handle this lighter feed.

The rapid introduction of shale oil to the refinery has come with interesting consequences for the fluid catalytic cracking unit (FCCU). To aid refiners in understanding the implications of shale oil, a detailed feed analysis and cracking studies of a representative Bakken shale oil and its fractions, compared to a typical Mid-Continent vacuum gasoil (VGO), are provided here. These results aid in understanding how best to optimize operations and maximize FCC value. A key element in this optimization is appropriate catalyst selection to overcome some of the challenges commonly associated with processing shale oil. In particular, the presence of iron and calcium in variable quantities can be addressed through the application of a catalyst with an optimized matrix and mesoporosity.

Properties of raw Bakken crude. Shale oil is highly variable. Density and other properties can show wide variation, even within the same field.[1–4] For this study, a sample of raw Bakken crude was obtained from a refinery. The crude was light and sweet with an API gravity of 42° and a sulfur content of 0.19

TABLE 1. Properties of straight-run shale oil feed used in this study, compared to published assay data

			Published assay data[5]		
		Bakken sample referenced in this work	**Bakken**	**WTI**	**LLS**
API gravity, degrees		41.9	> 41	40	35.8
Sulfur, wt%		0.19	< 0.2	0.33	0.36
Distillation yields		**wt%**	**vol%**	**vol%**	**vol%**
Light ends, C_1–C_4		1	3	1	2
Naphtha, C_5–330°F		32	30	30	17
Kerosine, 330°F–450°F		14	15	15	14
Diesel, 450°F–680°F		25	25	24	34
VGO, 680°F–1,000°F		23	22	23	25
Vacuum residue, 1,000+°F		5	5	7	8
Total		**100**	**100**	**100**	**100**
Conradson carbon residue, wt%		0.78	NA	NA	NA
Gasoline fraction properties	RON (G-con)	60.6	NA	NA	NA
	MON (G-con)	57.6	NA	NA	NA
LCO fraction properties, 430°F–650°F	Aniline point, °F	155.9	NA	NA	NA
	API gravity	37.6	NA	NA	NA
	Diesel index	58.6	NA	NA	NA

wt%. **TABLE 1** presents the properties of this sample, compared to published assays of Bakken, West Texas Intermediate (WTI) and Light Louisiana Sweet (LLS) crude oils.

The properties of the Bakken crude used in this study closely matched those in the published assay. Similar to other light crudes, raw Bakken crude has a low amount of FCC feed (< 28% at 680°F+). The straight-run (SR) Bakken sample was distilled into a 430°F– gasoline cut and a 430°F–650°F light cycle oil (LCO) cut, and the properties of these cuts were measured. The gasoline composition and properties were analyzed via proprietary octane calculation software based on detailed gas chromatography analysis.[6,7] The gasoline fraction from the straight-run Bakken sample was highly paraffinic and had low octane numbers [a research octane number (RON) of 61 and a motor octane number (MON) of 58]. The LCO fraction had an aniline point of 156°F and an API gravity of 37.6°, resulting in a diesel index of 59. While the Bakken crude sample was light and paraffinic, it also had a heavy end.

Processing challenges. Light sweet crudes are generally easy to process, although challenges arise when these crudes are the predominant feedstock in refineries designed for heavier crudes. Shale oils, like other light sweet crudes, have a much higher ratio of 650°F– to 650°F+ material compared to conventional crudes. Bakken shale oil has a nearly 2:1 ratio, while typical crudes, such as Arabian Light, have ratios near 1:1.

A refinery running high percentages of Bakken oil could become overloaded with light cuts, including reformer feed and isomerization feed, while at the same time growing short on feed for the FCCU and the coker. Refiners running predominantly shale oil could shut down the vacuum distillation and coker units, and send the entire atmospheric tower bottoms (ATB) portion to the FCCU. Many refiners would still be short on FCC feed, and some have considered bypassing a portion of the whole shale oil around the crude distillation unit to fill up the FCCU capacity.

Also, while the highly paraffinic Bakken ATB would crack to high conversion, the expected low delta coke would result in low regenerator temperatures and possible difficulty in circulating sufficient catalyst to maintain reactor temperature.

Shale oil cracking yields. To examine the impact of shale oil on FCC yields, cracking was performed with whole Bakken crude, a 430°F+ distillation of Bakken, a 650°F+ distillation of Bakken and a reference sample of a typical Mid-Continent VGO. Feed properties are presented in **TABLE 2.** Cracking was done over an FCC catalyst in a fixed-fluidized bed advanced cracking evaluation (ACE) test unit[8] at a constant reactor temperature of 980°F, using three catalyst-to-oil (C/O) ratios (4, 6 and 8) for each of the feeds. The catalyst used in the experiments was an FCC catalyst with an optimized matrix and mesoporosity, de-

TABLE 2. Properties of whole Bakken crude and distilled cuts of Bakken crude, compared to a Mid-Continent VGO

Property	Whole Bakken crude	430°F+ distillation of Bakken	650°F+ distillation of Bakken	Mid-Continent VGO
API gravity, degrees	41.9	28.6	23	24.7
Conradson carbon residue, wt%	0.78	1.34	2.27	0.32
K-factor	11.68	11.73	11.86	12.01
Sulfur, wt%	0.19	0.3	0.43	0.35
Total nitrogen, wt%	NA	0.07	0.12	0.14
Basic nitrogen, wt%	NA	0.023	0.038	0.046
Hydrogen, wt%	13.6	13.1	12.7	12.9
Percent boiling > 1,000°F	5.1	14.5	23.6	16.5
Molecular weight	NA	321	414	430
Molecular n-d-m analysis				
Ca, aromatic ring carbons, %	NA	16.9	22.1	17.6
Cn, naphthenic ring carbons, %	NA	21.7	17.3	20.3
Cp, paraffinic carbons, %	NA	61.4	60.6	62.1
D2887 simulated distillation, °F				
Initial boiling point	96	330	530	527
10%	157	470	658	691
30%	306	580	756	773
50%	478	699	844	848
70%	663	840	953	928
90%	909	1,057	1,135	1,045

TABLE 3. Deactivated catalyst properties

Total surface area, m^2/g	196
Zeolite surface area, m^2/g	110
Matrix surface area, m^2/g	86
Unit cell size, Å	24.3
Rare earth oxides, wt%	2.1
Alumina, wt%	52.1

TABLE 4. Interpolated yields at a C/O of 6

	Whole Bakken crude	430°F+ distillation of Bakken	650°F+ distillation of Bakken	Mid-Continent VGO
Conversion, wt%	83.5	71.7	74.3	74.4
H_2 yield, wt%	0.02	0.06	0.08	0.04
C_1 and C_2, wt%	0.9	1.2	1.5	1.3
Total C_3, wt%	4.5	5.1	5.2	5.1
C_3=, wt%	3.5	4.4	4.4	4.4
Total C_4, wt%	10.1	10.8	10.7	10.8
C_4=, wt%	4.3	5.7	5.9	6.1
LPG, wt%	14.6	16	15.9	15.9
Gasoline, C_5–430°F, wt%	65.4	52.1	52.9	54.1
RON, G-con	78	89.2	90.1	90.3
MON, G-con	70.9	78.9	79.6	79.5
LCO, 430°F–700°F, wt%	14.2	24.6	19.6	19.1
Bottoms, 700°F+, wt%	2.3	3.7	6	6.4
Coke, wt%	1.8	2.7	4.1	2.9

activated metals-free using a cyclic propylene steaming (CPS) protocol. Deactivated properties are given in **TABLE 3.**

Interpolated yields at a C/O of 6 are presented in **TABLE 4.** The whole Bakken crude resulted in low coke and a low-octane gasoline. While the whole Bakken crude yielded significant gasoline, much of the gasoline was from uncracked starting material in the feed. The yields of the 430°F+ and 650°F+ distillations of the Bakken crude were similar to those of the Mid-Continent VGO reference sample. The 650°F+ distillation of the whole Bakken crude had higher coke than the Mid-Continent VGO due to its heavier end and higher Conradson carbon number.

Results of processing SR shale oil. While FCC is typically used to reduce the molecular weight of the crude oil heavy fractions (such as VGO and ATB), in some cases refiners are charging whole shale oil as a fraction of the plant's FCC feed. As a model case to understand the cracking of whole crude oil in the FCC and the effect of process conditions on yields, the whole Bakken crude described in **TABLE 2** was processed in a circulating-riser FCC pilot plant at three riser outlet temperatures: 970°F, 935°F and 900°F.

As a reference case, the Mid-Continent VGO described in **TABLE 2** was cracked at a riser outlet temperature of 970°F.[9] The catalyst used in the experiments was a high-matrix FCC catalyst, deactivated metals-free using a CPS-type protocol. Deactivated properties are shown in **TABLE 3.**

FIG. 1 presents the yield structure of the starting feeds and the cracked products for a riser outlet temperature of 970°F. The Mid-Continent VGO is a typical VGO feed with a large portion of 650°F+ material and a small fraction of LCO-range material. When cracked, the LCO-range material cracks to liquefied petroleum gas (LPG) and gasoline, and the 650°F+ material cracks to the typical distribution of LPG, gasoline and LCO, resulting in a net increase in LCO.

The whole Bakken crude starts with large fractions of gasoline and LCO-range material and a low amount of 650°F+ material. The amount of gasoline produced after cracking is high since the LCO-range material cracks to predominantly gasoline, and much of the starting gasoline is unconverted. LCO yields are low since there is little starting 650°F+ material to crack to LCO.

For the three different reactor outlet temperatures, plots of C/O ratio, gasoline, LCO and coke yields vs. conversion are shown in **FIG. 2.** As expected, lowering the reactor temperature increases the amount of LCO produced. Cracking SR shale oil produces little coke and bottoms. At the same conversion level, lowering the reactor temperature results in slightly more gasoline yield (due to increased C/O), which is consistent with prior research.[10]

At a riser outlet temperature of 970°F, the whole Bakken feed produces more gasoline, less LCO and less coke than the reference Mid-Continent VGO. Compared to the VGO, which produced gasoline with a RON of 93 and a MON of 80, the SR Bakken oil produced a paraffinic low-quality gasoline (at all three reactor outlet temperatures) with a RON of less than 80 and a MON of less than 70.

Synthetic crude from the pilot plant runs was distilled to recover the 430°F–650°F LCO fraction. The aniline point and API gravity of the LCO were measured to calculate the diesel index, which is a measure of LCO quality. **FIG. 3** presents data for LCO yield and quality as a function of conversion. Increasing conversion lowers LCO quality as a result of increased cracking of the LCO-range paraffins to lighter hydrocarbons. As seen in prior research,[11] LCO quality follows LCO yield and did not appear to be influenced by reactor temperature at constant conversion. Diesel index values of the LCO produced by cracking

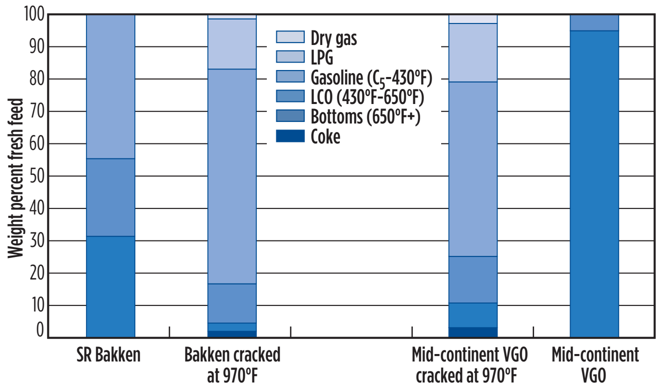


FIG. 1. Yield structure of starting feeds and cracked products for SR Bakken and Mid-Continent VGO.

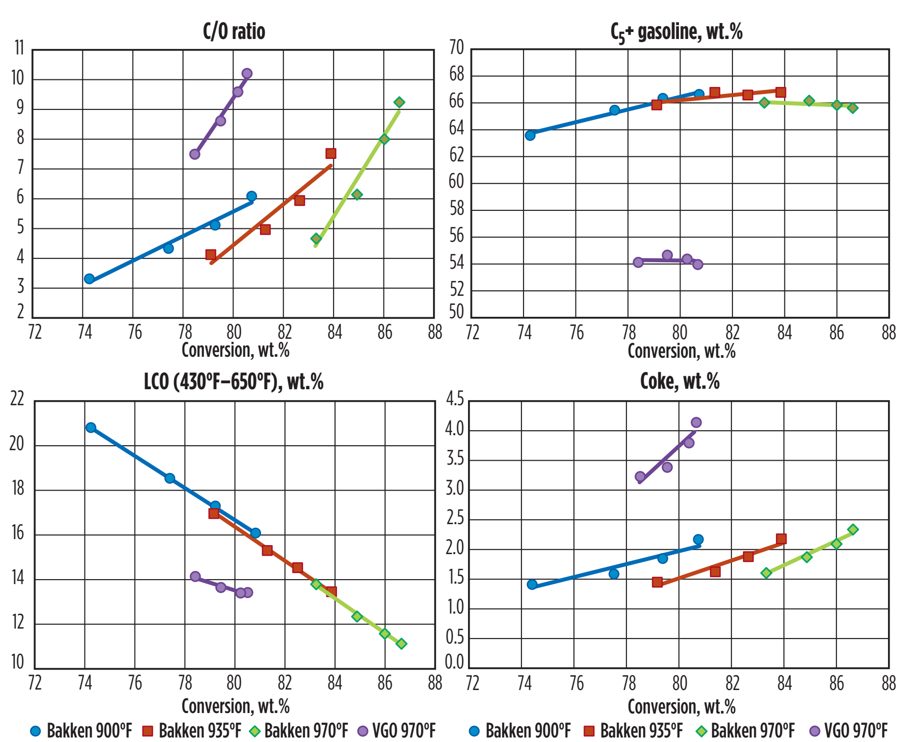


FIG. 2. Product yields as a function of riser outlet temperature and feed.

whole shale oil were significantly higher than those obtained when cracking the reference Mid-Continent VGO. At a conversion of 78 wt%, the whole Bakken sample gave an LCO with a diesel index of 40, compared to a diesel index of 10 obtained for the LCO produced from the Mid-Continent VGO.

This study of the effect of operating variables shows that whole shale oil responds to FCC operating conditions in a similar way to conventional oils. However, the product yield slate is substantially different in that good-quality (high-diesel-index) LCO is produced in the FCC along with large amounts of low-octane gasoline.

Metals in shale oil. While most shale oils are low in nickel (Ni) and vanadium (V), they have been found to be high in inorganic solids and in iron (Fe) and alkali metals.[2,12] **TABLE 5** presents metals analyses of whole Bakken crude, a Bakken 650°F+ distillation, and a sample of Mid-Continent VGO.

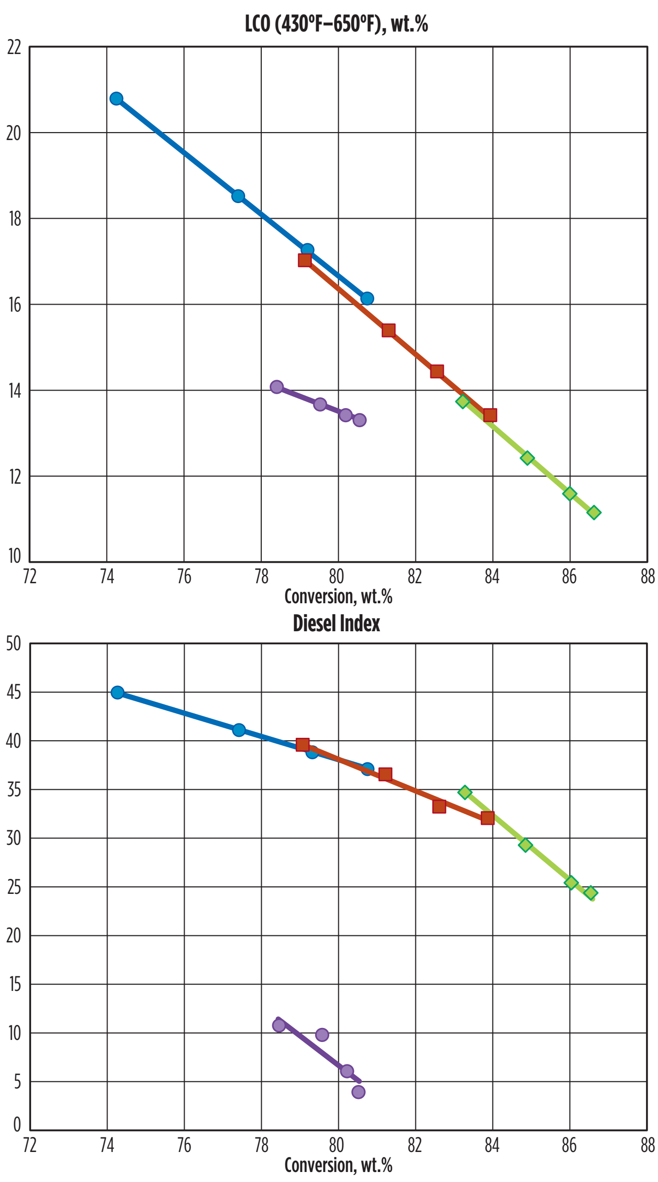


FIG. 3. Effect of conversion level and feed type on LCO yield and quality.

Also included in the table are other published metals analyses of shale oils. While metals levels in the samples vary, Fe and calcium (Ca) levels are generally high. Reports from the field indicate that Bakken crude is typically low in Ni and V, while crudes sourced from the Eagle Ford shale have higher Ni and V levels that can vary significantly based on their source.

To better understand the possible sources of metals in shale oil, the sample of whole Bakken crude was filtered through an 0.8-micron filter, and the solids were recovered. Scanning electron microscopy of the solids identified irregular micron- and submicron-sized particles. Energy dispersive spectroscopy maps of Fe, sulfur and Ca are shown in **FIG. 4.** The Fe in the sediments is associated with sulfur.

X-ray diffraction of the sediment identified the following crystalline phases: anhydrite (Ca_2SO_4), magnetite (Fe_3O_4) and pyrrhotite (substoichiometric FeS). Anhydrite and pyrrhotite have been mentioned in previous studies as being present in the Bakken formation.[13,14] Based on this analysis, it appears that much of the iron in the Bakken crude comes from very small particles of iron oxide and pyrrhotite.

Iron and calcium effects. Fe and Ca have negative effects on catalyst performance. While particulate tramp Fe from rusting refinery equipment does not have a significant detrimental effect on catalyst, finely dispersed Fe particles in feed (either as organic compounds or as colloidal inorganic particles) can deposit on the catalyst surface, reducing its effectiveness.[15,16]

The Fe deposits combine with silica (Si), Ca, sodium (Na) and other contaminants to form low melting temperature phases, which collapse the pore structure of the exterior surface, blocking feed molecules from entering the catalyst particle and reducing conversion.[17] Fe, in combination with Ca and/or Na, has a greater negative effect on catalyst performance than does Fe alone. The symptoms of Fe and Ca poisoning include a loss of bottoms cracking as feed particles are blocked from entering the catalyst particle, along with a drop in conversion.

Catalytic solutions for FCC processing of shale oil. The variability in shale oil properties requires a catalyst capable of process flexibility and metals tolerance. Several catalyst design

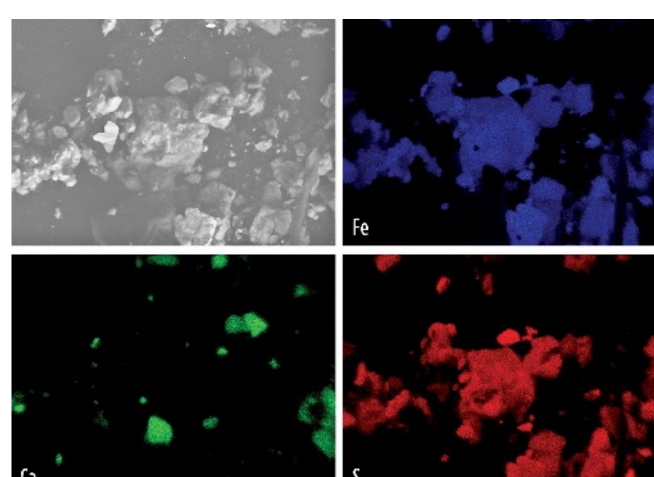


FIG. 4. Energy dispersive spectroscopy maps of sediment in Bakken crude.

TABLE 5. Metals analyses of several shale oils

Property	Samples referenced in this work: Mid-Continent VGO	Samples referenced in this work: Whole Bakken crude	Samples referenced in this work: 650°F+ distillation of Bakken crude	Published assay data[12]: Flashed Bakken crude	Published assay data[12]: 75% Eagle Ford stream, total	Published assay data[12]: 75% Eagle Ford stream, filtered	Published assay data[3]: Bakken crude	Published assay data[3]: Eagle Ford crude
Barium, ppm	< 0.01	0.2	0.1	NA	NA	NA	0.02	0.21
Calcium, ppm	< 0.1	0.5	1.2	0.6	15	1.4	0.54	9.8
Iron, ppm	< 0.1	7.5	7.8	4.1	16	3	0.7	2.3
Magnesium, ppm	< 0.04	0.2	0.2	< 0.2	1.6	< 0.12	0.05	0.34
Nickel, ppm	< 0.04	0.4	1.9	0.6	8	8	0.05	< 0.14
Potassium, ppm	< 0.04	0.4	0.3	< 0.2	1.2	< 0.3	0.1	0.5
Sodium, ppm	< 0.06	8.7	3.9	4.1	34	0.4	2.8	12
Vanadium, ppm	< 0.03	0.1	0.5	0.22	22	22	0.02	< 0.05

TABLE 6. Mesoporosity comparison of equilibrium catalysts

Catalyst	Mercury pore vol, cm³/g: Total	Micropores, 36 Å–100 Å	Mesopores, 100 Å–600 Å	Macropores, 600+ Å
Optimized matrix and mesoporosity catalyst	0.389	0.092	0.206	0.091
Optimized matrix and mesoporosity catalyst	0.412	0.107	0.232	0.071
Competing catalyst 1	0.386	0.116	0.102	0.168
Competing catalyst 2	0.413	0.092	0.089	0.232

factors are important. A catalyst that cracks deep into the bottom of the barrel increases FCC flexibility and maximizes total distillate yield. A moderate zeolite-to-matrix catalyst ensures that activity is not compromised while maintaining optimal bottoms cracking. The appropriate level of rare-earth exchange on zeolite is also a crucial aspect in maintaining optimal coke selectivity. A catalyst with optimized matrix and mesoporosity is a highly effective system for the effective processing of a variety of feedstocks.

Catalyst design can be optimized to resist the effects of contaminant Fe and Ca. High alumina catalysts, especially catalysts with alumina-based binders and matrices, are best suited to process Fe- and Ca-containing feeds because they are more resistant to the formation of low-melting-point phases that destroy the surface pore structure. Optimum distribution of mesoporosity also plays a role in maintaining performance because diffusion to active sites remains unhindered, despite high-contaminant metals.

The paraffin, aromatic and porphyrin molecules in the 700°F–1,000°F boiling-point fraction of FCC feed have dynamic molecular sizes between 10 Angstroms (Å) and 30 Å.[18] These molecules are too large to fit into zeolite pores (which are typically smaller than 7.5 Å) and must first be cracked by the matrix activity of the catalyst. For free diffusion of the 700°F–1,000°F boiling-range molecules to occur, the catalyst pore diameter needs to be 10 to 20 times the size of the molecule, or 100 Å–600 Å.[18]

Based on these considerations, a catalyst with an optimized alumina matrix and mesoporosity in the 100 Å–600 Å range was designed.[19] While two catalysts may have similar total pore volume, their mesoporosity can vary greatly. **TABLE 6** compares the mesoporosity of this optimized matrix and mesoporosity catalyst to a competing catalyst. **Note:** The mercury intrusion method measures pore sizes greater than 36 Å, so the values represent porosity associated with the catalyst matrix only. Micropores smaller than 100 Å are undesirable and lead to poor

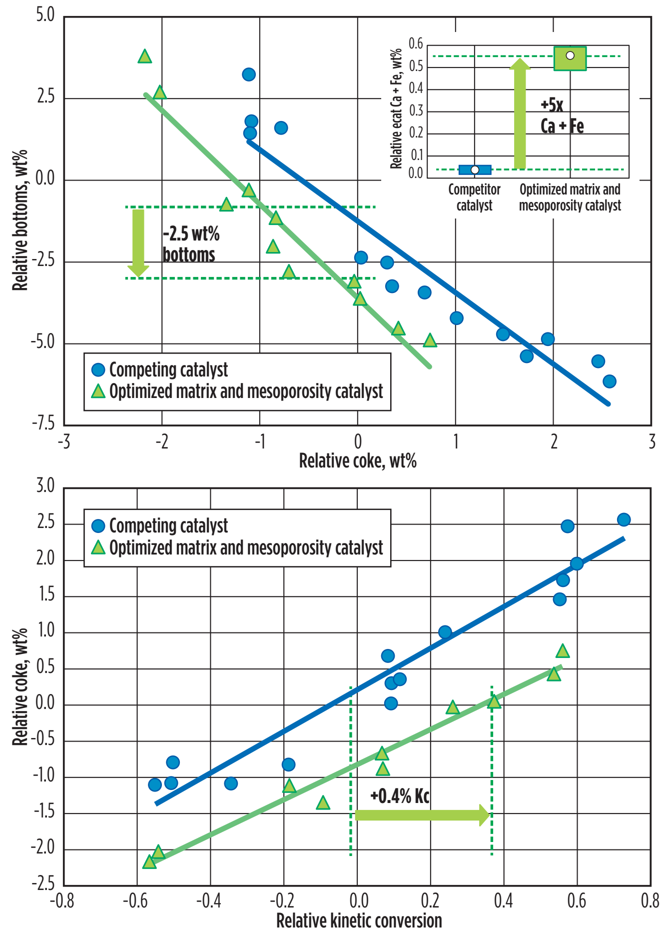


FIG. 5. Despite higher contaminant metals, the catalyst with optimized matrix and mesoporosity improved bottoms cracking and coke selectivity.

coke and gas selectivity as the result of poor diffusivity and over-cracking.[15] As seen in the table, the 100 Å–600 Å mesopore volume of the optimized matrix and mesoporosity catalyst was twice that of the competing sample.

The resistance of the optimized matrix and mesoporosity catalyst to Fe and Ca poisoning was demonstrated in a commercial application. A refinery was processing resid feedstock high in Fe and Ca. Over time, the unit exhibited the symptoms of Fe poisoning. As Fe nodules built up on the catalyst surface, equilibrium catalyst activity, unit conversion and bottoms cracking all began to suffer. The refiner switched from a competing catalyst to one with optimized matrix and mesoporosity.[a] Upon switching, activity, bottoms cracking and coke selectivity improved, even at higher contaminant metals levels (**FIG. 5**).

Takeaway. The shale oil boom has resulted in a renaissance in the North American refining industry. While shale oils are generally light, sweet and easy to crack, quality can vary greatly, and shale-derived feeds can contain sediments with high levels of iron and alkali metals. Catalyst formulations with optimized matrix and mesoporosity provide the best resistance to iron and calcium poisoning. Proper catalyst choice allows refiners to fully exploit the opportunities of shale oil while minimizing the detrimental impacts of processing. HP

ACKNOWLEDGMENTS

The authors thank their colleagues at Grace for their assistance with the testing for this article. Special thanks go to Larry Langan for his analysis of the Bakken sediment.

NOTE

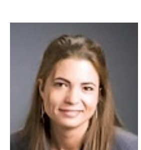
[a] Grace's MIDAS FCC catalyst contains a specialty alumina matrix and is designed to maximize the 100 Å–600 Å mesoporosity optimal for coke-selective bottoms conversion.

LITERATURE CITED

1 Marfone, P. A., "Refiners have a new learning curve with shale oil," *Hydrocarbon Processing*, March 2013.
2 Kremer, L., "Shale oil issues and solutions," AFPM Principles and Practices Session, Salt Lake City, Utah, October 2012.
3 Haynes, D., "Tight oil impact on desalter operations," Crude Oil Quality Association Meeting, New Orleans, Louisiana, November 2012.
4 Ohmes, R. and M. Routt, "Characterizing and tracking contaminants in opportunity crudes," 2013 AFPM Annual Meeting, San Antonio, Texas.
5 Hill, D., "North Dakota refining capacity study final technical report," DOE Award No. DE-FE0000516, January 5, 2011.
6 Haas, A., G. McElhiney, W. Ginzel and A. Buchsbaum, "Gasoline quality: The measurement of compositions and calculation of octanes," *Erdol Kohle-Erdgas-Petrochemie*, Vol. 43, pp. 21–26, 1990.
7 Cotterman, R. L. and K. W. Plumlee, "Effects of gasoline composition on octane number," ACS National Meeting, Miami Beach, Florida, 1989.
8 Kayser, J. C., "Versatile fluidized bed reactor," US Patent 6,069,012, Kayser Technology, 2000.
9 Bryden, K., G. Weatherbee and E. T. Habib, "Flexible pilot plant technology for evaluation of unconventional feedstocks and processes," American Fuel and Petrochemical Manufacturers (AFPM) Annual Meeting, San Antonio, Texas, 2013.
10 "FCC operation," *Grace Davison Guide to Fluid Catalytic Cracking*, 1993.
11 Ritter, R. E., "Light cycle oil from the FCC unit," National Petrochemical and Refiners Association (NPRA) Annual Meeting, San Antonio, Texas, 1988.
12 Savage, G., "Crude preheat management for challenged and unconventional crudes," Crude Oil Quality Association Meeting, San Antonio, Texas, March 2013.
13 Holubnyak, Y., et al., "Understanding the souring at Bakken oil reservoirs," SPE International Symposium on Oilfield Chemistry, The Woodlands, Texas, April 2011.
14 Cioppa, M. T., "Spatial variations in magnetic components of the Devonian Birdbear formation, Williston Basin," Geofluids 7 Conference, Rueil-Malmaison, France, June 2012.
15 Cheng, W.-C., E. T. Habib, K. Rajagopalan, T. G. Roberie, R. F. Wormsbecher and M. S. Ziebarth, "Fluid catalytic cracking," *Handbook of Heterogeneous Catalysis*, 2008.
16 Yaluris, G., "The effects of Fe poisoning on FCC catalysts: An update," *Catalagram* 91, W. R. Grace & Co., 2002.
17 Yaluris, G., W.-C. Cheng, L. T. Boock, M. Peters and L. J. Hunt, "The effects of Fe poisoning on FCC catalysts," NPRA Annual Meeting, New Orleans, Louisiana, 2001.
18 Zhao, X., W.-C. Cheng and J. A. Rudesill, "FCC bottoms cracking mechanisms and implications for catalyst design for resid applications," NPRA Annual Meeting, San Antonio, Texas, 2002.
19 Cher, Y.-Y., J. Koebel and R. Schiller, "Enhanced bottoms cracking and process flexibility with MIDAS FCC catalyst," *Catalagram* 112, W. R. Grace & Co., 2012.

KEN BRYDEN is the manager of the FCC Evaluations Research Group at Grace Catalysts Technologies. He holds a BS degree in chemical engineering from the University of California at Davis and a PhD in chemical engineering from the Massachusetts Institute of Technology. Dr. Bryden is a co-inventor on one issued patent and a co-author of nine peer-reviewed journal articles.

E. THOMAS HABIB, JR. is the director of customer research partnerships and the DCR licensing manager for Grace Catalysts Technologies. He is an industry-recognized expert with 40 years of FCC experience. Mr. Habib is the lead author of the "Fluid catalytic cracking" chapter in the second edition of the *Handbook of Heterogeneous Catalysts*, and he has authored numerous other publications, including the book *Fluid Catalytic Cracking with Zeolite Catalysts.*



OLIVIA TOPETE is the FCC marketing manager for Grace Catalysts Technologies North America. She holds a BS degree in chemical engineering from Purdue University and an Executive MBA degree from Rice University. She has 11 years of experience in the hydrocarbon processing industry, including roles in technical sales and service with Grace, technical service and engineering at Merichem and field operating services with UOP.


EAU DU
BITUMEN
FOR MEN

J. LACK and **B. HARRELL,** Baker Hughes,
Sugar Land, Texas

Reduce salt corrosion rates with stronger base amines

When processing discounted crude oils, one of the primary challenges refiners face is under-salt corrosion in crude unit distillation overhead systems. The combination of higher overhead chloride levels and the presence of amine contaminants in discounted crudes can greatly increase the risk of salt formation. For those overhead systems where amine salt corrosion cannot be mitigated by conventional methods, the application of nontraditional neutralizing amines can be beneficial. In particular, under-salt corrosion rates produced by stronger base amines are significantly less than those of weaker bases found as contaminants, such as ammonia or monoethanolamine. Corrosion rates can also be reduced when the weaker base contaminant salts are mixed with stronger base amine salts. Significant reductions in corrosion rates have been confirmed in both laboratory and field operating environments.

Controlling corrosion in the overhead condensing section of crude atmospheric distillation units is a constant challenge for refiners. The typical treatment strategy involves a delicate balancing act of neutralizing acids in condensed waters with ammonia or amines while avoiding the formation of corrosive salts via a vapor phase reaction with hydrogen chloride (HCl). Recent trends of declining crude quality, which result in less effective chloride removal and the frequent presence of amine contamination, have increased the risk of forming corrosive salts. Neutralizing amine selection that reduces the risk of salts caused by the neutralizer does not affect the risk of salt formation by contaminant amines or ammonia.

Contaminant reduction techniques, such as caustic addition for chlorides and desalter acidification for amines, can be successful at reducing risk of salt formation.[1,2] However, these approaches have limitations if significant reduction is required. Water wash applications have had mixed results as a means for reducing salt formation and are not always viable in some overhead configurations. With these options exhausted, a refiner is left with operational changes or enduring frequent bundle failures, both of which can be costly. There is a need for additional options, including chemical treatment at the overhead, which can reduce corrosion from salts.

Chemical suppliers have focused on filming inhibitors to reduce corrosion rates from acidic water and amine neutralizers to control pH. Various strategies have been used to optimize neutralizer performance and reduce the risk of salt formation. These have included the use of volatile amines, weaker base amines, amine blends and thermodynamic model guidance.[3–7] While these techniques have been successful at reducing salt formation by the neutralizer, they do not address the corrosion from salts formed by ammonia or contaminant amines. To reduce corrosion from salts formed by contaminants, first consider the chemical drivers that make salts corrosive.

Theory. Ammonia or amines can react with HCl to form a solid or liquid salt:

$$RNH_2 + HCl \rightleftarrows RNH_3Cl \quad (1)$$

These salts are highly hydroscopic and easily absorb water from the steam used for stripping the distillation column. In the presence of water, the salts ionize, forming ammonium or aminium ions with chloride ions:

$$RNH_3Cl \rightleftarrows RNH_3^+ + Cl^- \quad (2)$$

According to the Brønsted theory, which describes acid-base interactions, the reaction of an acid with a base produces a conjugate base and a conjugate acid.[8] The strength of the conjugate acid is inversely related to the strength of the base from which it is derived:

$$\underset{\text{(Acid)}}{A\text{–}H} + \underset{\text{(Base)}}{B} \rightleftarrows \underset{\text{(Conjugate base)}}{A^-} + \underset{\text{(Conjugate acid)}}{B\text{–}H^+} \quad (3)$$

The resulting ammonium or aminium ion is the conjugate acid of ammonia or the amine. This conjugate acid is "weak" by definition in that there is an equilibrium established, which depends on its acid constant:

$$RNH_3^+ \rightleftarrows H^+ + RNH_2 \quad (4)$$

where:

$$K_a = \frac{[H^+][RNH_2]}{[RNH_3^+]} \quad (5)$$

It is the production of the hydrogen ion that creates the corrosive environment associated with the salt. At the point where corrosion occurs, the local environment is essentially an aqueous solution of the ammonium or amine salt in contact with the metal surface. Although the morphology of attack is localized, the metal is oxidized by the hydrogen ion in a similar manner to that of general acid corrosion:

$$Fe^\circ + 2H^+ \rightarrow Fe^{2+} + H_2 \quad (6)$$

The rate of corrosion for this mechanism is proportional to the hydrogen ion concentration. Typically, hydrogen ion concentration is reported as a pH value where:

$$pH = -\log[H^+] \quad (7)$$

Therefore, corrosion should increase exponentially as pH decreases. In the same manner, the corrosivity of the salt can be related to its concentration and acid strength:

$$pH = pK_a + \log\left(\frac{[RNH_2]}{[RNH_3^+]}\right) \quad (8)$$

All else being equal, the corrosivity of a salt should theoretically be exponentially related to its acid strength. The strength of the conjugate acid is inversely proportional to the base strength of ammonia or the amine. The ionization constants are related by the ionization constant for water:

$$K_a \times K_b = K_w \quad (9)$$

or

$$pK_a + pK_b = pK_w \quad (10)$$

Because of this relationship, a stronger base will form a weaker (theoretically less corrosive) acid salt and a weaker base will form a stronger (theoretically more corrosive) acid salt.

Reducing salt corrosion rates. This presents an opportunity to reduce the corrosion rates by manipulating the chemistry of the salt. The two most common contaminant base species are ammonia and monoethanolamine (MEA). Each of these have a pK_b of 4.75 and 4.6, respectively, and conjugate acids with a pK_a of 9.25 and 9.4, respectively. Both are also commonly used in crude unit overheads as a neutralizing agent. Other contaminant amines and commonly used neutralizing agents have similar or much weaker base strengths. **TABLE 1** shows the pK_b and conjugate acid pK_a of these common contaminants and neutralizing agents.

Hydrochloride salts of ammonia and amines, whose conjugate acid pK_a is between 9 and 10, typically result in corrosion rates of 50 mpy to 200 mpy (1.3 mm/yr–5.1 mm/yr). In one case, the N-methylmorpholine (a weaker base) forming salts (pK_a 7.4), produced corrosion rates > 1,000 mpy (> 25.6 mm/yr).

Given the observed corrosion rates with weak bases (ammonia, MEA) and weaker bases (n-methylmorpholine), a study sought to determine whether salts of stronger base amines would result in lower corrosion rates compared to the salts formed by those commonly encountered in crude units. Additionally, the investigation considered whether corrosion rates of contaminant amine salts could be reduced in the presence of a stronger base amine salt.

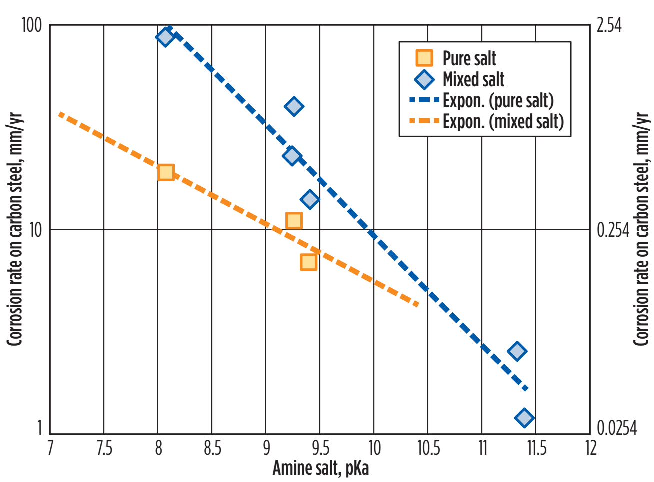


FIG. 1. Corrosion rates for 1M solution of amine hydrochloride salts at 71°C (160°F).

EXPERIMENTAL PROCEDURE

Corrosion tests were conducted to determine the relative corrosivity of amine hydrochloride salts in concentrated, aqueous solutions. In these tests, carbon steel specimens were immersed in various deaerated salt solutions at elevated temperatures in non-stirred pressure vessels.

Materials. Corrosion tests were carried out in non-stirred autoclaves with a capacity of approximately 300 ml. Each vessel was fitted with a glass liner that contained the amine salt solution and coupon. The heads of the autoclaves were fitted with two ports, an inlet tube and outlet, so the solutions could be deaerated by purging with nitrogen.

Amine hydrochloride salt solutions were made from common contaminant and neutralizing amines as well as from stronger base candidate amines. Salt solutions (saturated and 1M concentrations) were prepared using deionized water at room temperature. Additionally, saturated solutions were prepared in a dry nitrogen atmosphere by adding the solid salt to stirred deionized water until a precipitate persisted.

Procedure. A test solution of 100 ml was charged to a glass liner and placed in the autoclave. A head was placed on the autoclave and nitrogen was bubbled at a constant rate through the solution for 30 to 45 min. A coupon, mounted on an inverted polytetrafluoroethylene (PTFE) machine screw, was then quickly placed in the solution and the autoclave was sealed. The purpose of the PTFE screw was to suspend the coupon above the bottom of the glass liner. A final deaeration step was completed by sequentially pressurizing and depressurizing the autoclave with nitrogen. The autoclave was then placed in an oven that had been preheated to the test temperature.

After 25 h, the autoclaves were removed from the oven. Cooling was facilitated with a stream of compressed air. Once the vessels could be safely handled, the coupons were removed, rinsed with deionized water, isopropanol, n-heptane, and ultrasonically cleaned as described above. Scale was not observed on any of the coupons. Once dry, the final masses of the coupons were recorded and the corrosion rates were calculated.

OBSERVATIONS

The experimental results showed a very good exponential relationship of corrosion rates vs. pK_a for 1M solution of amine hydrochloride salts at 71°C (160°F). The blue diamonds labeled as "pure salt" in **FIG. 1** show the corrosion rates of single amine salts. The relatively dilute 1M solution was used for convenience and to confirm the theory by comparing the relative corrosion rates of the salts without factoring solubility. These measured corrosion rates are less than would be experienced in a process unit where concentrated salts near saturation would be expected.

To examine the corrosion rate of salt mixtures made from amines with different base strengths, a 1M solution with 80 mol% amine salt with a pK_a of 11.4 was mixed with 20 mol% of various salts of weaker bases. The resulting corrosion rates of these salt mixtures are represented in **FIG. 1** with the gold squares labeled as "mixed salt." The data, graphed at the pKa of the salt comprising the weaker base for comparison, show a significant reduction in corrosion rate relative to the rate of the isolated salts of the weaker bases.

To better examine the effects of amine salts under field conditions, corrosion rates were determined for saturated salt solutions at 71°C (160°F). As shown in **FIG. 2,** these results indicate a good exponential correlation, even with varying solubility among amine salts. The corrosion rates obtained with the 9 to 10 pK_a amine salts fell in the 50 mpy to 200 mpy (1.3 mm/yr–5.1 mm/yr) range typically experienced by these amine salts.

As with the 1M solution tests, a saturated salt mixture was created with 80 mol% of a 11.4 pK_a salt with 20 mol% of ammonia and MEA contaminant salts. The resulting corrosion rates, shown in **FIG. 3,** indicate significant reductions relative to corrosion rates of the isolated contaminant salts.

Given that the measured corrosion rates of the saturated salt solutions are within the range experienced in actual process units, the results of the saturated mixture test might be the best indicator of what can be expected from field applications. Though the corrosion rates of the mixture were not low enough to be considered "controlled," the significant reduction of corrosion suggested it to be a viable treatment option for systems where conventional methods of preventing salt formation fall short or are unavailable.

CASE HISTORY

A US refinery was successfully controlling salt formation in the crude atmospheric overhead exchanger with a water wash injected to the inlets of the exchangers, as shown in **FIG. 4.** Approximately five years into the bundle life, there was a reconfiguration of the unit upstream of the atmospheric tower. Though the reconfiguration of the unit had some effect on the process conditions of the atmospheric overhead, the wash rate remained well above the industry accepted minimum. However, after about a year under the new operations, tube leaks were discovered in the overhead exchangers. These tubes were plugged, pressure tested and the operation resumed. This cycle repeated every few months until the bundles had to be replaced. The new bundle had its first leak just five months into operation, representing an average corrosion rate of 328 mpy (8.3 mm/yr).

A corrosion risk monitor was used, which is a field program designed to enhance corrosion control programs via a combination of rigorous thermodynamic simulation and proprietary salt formation calculations, to determine what changed in the system to increase the corrosion risk beyond what the water wash system could handle. The results indicated that, under the new operation, salt formation potential had increased, which placed more burden on the existing water wash.

TABLE 2 summarizes the modeling results before and after unit reconfiguration. A combination of higher contaminant levels and lower stripping steam rates resulted in lower salt approach temperatures. A newly developed spray mass transfer model suggested that the level of contaminant removal required to prevent salt formation had been achieved prior to the reconfiguration. However, under the new operation, the required level of contaminant removal could not be achieved because the spray nozzles were too close to the exchanger inlets.

Consideration was given to relocating the wash injection further upstream to increase vapor-water contact time, but

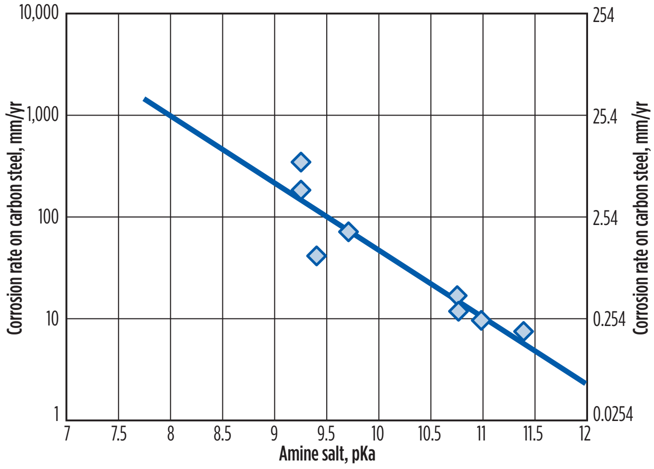


FIG. 2. Corrosion rates for a saturated solution of amine hydrochloride salts at 71°C (160°F).

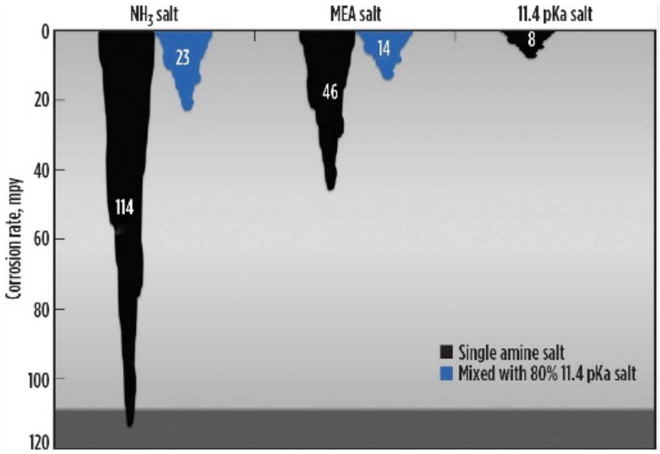


FIG. 3. Corrosion rate reduction of saturated salts.

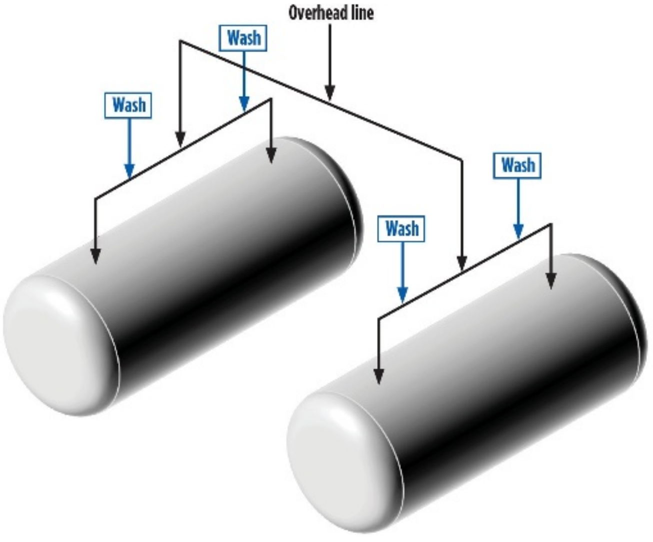


FIG. 4. Unit overhead wash injection locations.

higher velocities (85 to 105 ft/s, 26 to 32 m/s) presented an unacceptable risk. A project to address current limitations in the overhead was planned for the next unit shut-down, but was three years away from implementation. As an interim measure, the application of a patent pending, stronger base neutralizer was proposed in an attempt to slow corrosion rates until modifications could be made during the future turnaround.[9]

During the summer of 2011, the new stronger base neutralizer application was initiated to coincide with the installation of a new set of bundles. Although modeling calculations have indicated no change in the approach temperature for ammonia salts, these new bundles were in successful operation for 18 months without incident, more than tripling the life before treatment. Inspection of the bundles showed much of the tube thickness remained, suggesting an over three year life may have been achieved. The successful treatment program reduced corrosion >80% and allowed the refiner to avoid at least three charge rate reductions during that period, saving an estimated $1.5 million. FIG. 5 shows the bundle and process history to date.

TABLE 1. Base and conjugate acid strengths of common contaminants and neutralizing agents

Chemical name	Base strength pK_b	Conjugate acid strength pK_a
Ammonia	4.75	9.25
Monoethanolamine (MEA)	4.6	9.4
Diethanolamine (DEA)	5	9
Methyldiethanolamine (MDEA)	5.4	8.6
Methoxypropylamine (MOPA)	4.1	9.9
Dimethylethanolamine (DMEA)	4.75	9.25
Dimethylisopropanolamine (DMIPA)	4.6	9.4
Trimethylamine (TMA)	4.3	9.7
Morpholine	5.5	8.5
N-Methylmorpholine	6.6	7.4
N-Ethylmorpoline	5.9	8.1
Ethylenediamine (EDA)	4.1, 7.1	9.9, 6.1

TABLE 2. Significant operational changes at reconfiguration

	Before reconfiguration		After reconfiguration	
Parameter	Best case	Worst case	Best case	Worst case
Dew point temperature	90°C (195°F)	88°C (190°F)	85°C (185°F)	71°C (160°F)
Salt formation temperature	93°C (200°F)	104°C (220°F)	102°C (215°F)	121°C (250°F)
Salt temperature approach at dew point	–3°C (–5°F)	–17°C (–30°F)	–17°C (–30°F)	–50°C (–90°F)
Contaminant removal required to prevent salts	35%	90%	90%	99.95%
Effect on exchanger bundle life	Five years, no incident		Less than one year, tube leaks	

Results. This study has demonstrated that corrosion rates of amine salts in crude units can be reduced by altering the chemical composition of the salts formed. These results have been confirmed with acid-base theory, laboratory testing and a refinery trial application of the concept. Key takeaways:

- Acid-base theory predicts lower corrosivity for salts of stronger (high pK_a) bases
- Data from laboratory corrosion tests confirm theoretical predictions on relative corrosivities of salts
- Patent pending, stronger base amine has been applied to successfully extend bundle run length by preventing corrosion failures.

The results of these experimental studies and the field trial suggest a viable alternative to tolerating salt corrosion when other options are not available; however, it is our recommendation that all efforts to prevent salt formation be exhausted before relying on a stronger base amine treatment program. Preventing salt formation with guidance from a reliable ionic model remains the best course of action to achieve consistent, long-term corrosion control. HP

ACKNOWLEDGEMENTS

The authors wish to thank Ksenija Babic for contributions to the laboratory study and Aslam Khan for contributions to the refinery trial application. This article is a revised and updated version from an earlier presentation at CORROSION/2013, Paper C2013-0002037, NACE International, Orlando, Florida. Published with permission from NACE International, Houston, Texas.

LITERATURE CITED

1 Lack, J. E., "An In-Depth Look at Amine Behavior in Crude Units," National Association of Corrosion Engineers, Houston, Texas, March 2005.
2 Rechtien, R. G. and G. G. Duggan, "Identifying the Impacts of Amine Contamination on Crude Units," National Association of Corrosion Engineers, Houston, Texas, March 2006
3 S. Ferguson et al., US Patent 4 806 229, 1989.
4 S. E. Lehrer and J. G. Edmondson, US Patent 5 211 840, 1993.
5 V. K. Braden et al., US Patent 5 965 785, 1999.
6 J. E. Lack et al., US Patent 7 381 319, 2008.
7 Duggan, G. G. and R. G. Rechtien, "Application of Ionic Equilibria Process Simulation for Atmospheric Distillation Overhead Systems," National Association of Corrosion Engineers, Houston, Texas, March 1998.
8 Smith, M. B and J. March, March's Advanced Organic Chemistry: Reactions, Mechanisms and Structure, 6th edition, John Wiley & Sons, 2007.
9 J. E. Lack, US Patent Application 20120149615, filed 2011.

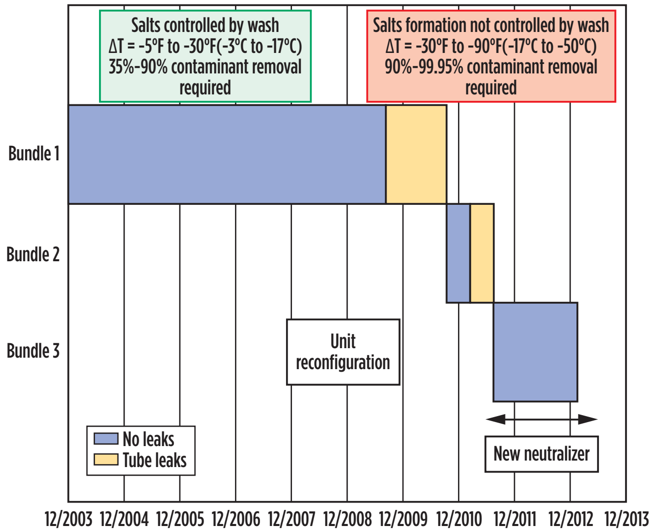


FIG. 5. Bundle and process history.

The new AMETEK Thermox WDG-V extractive combustion analyzer offers industry-leading safety support. First in its class to be third-party certified for SIL-2 implementation in safety-instrumented systems, the WDG-V provides a complete solution for combustion process control and safety.

Reliable detection of low-combustion oxygen and/or high CO in a fired heater or boiler is critical to burner management system effectiveness. The WDG-V analyzer monitors hot, wet flue gas to minimize excess oxygen, lower NOx emissions, and improve operating efficiency in power generation and petrochemical refining. It can also monitor methane levels to assure safe burner startup and shutdown.

The all-new WDG-V. Combustion management and safety capabilities so good, they make this guy jealous. Learn more at **www.ametekpi.com.**

AMETEK®
PROCESS INSTRUMENTS

A. W. SLOLEY, CH2M HILL, Bellingham, Washington

Mitigate fouling in crude unit overhead—Part 1

In a refinery crude distillation column, up to 40% of the total heat removal may be found in the tower overhead condenser. The quantity of duty available makes heat integration (HI) between the column overhead and unit feed an interesting project. Approximately 70% of the present crude units recover some portion of this duty. However, many of these units suffer from persistent and serious corrosion problems in the overhead system. In extreme cases, the viability of the entire refinery site may be compromised due to fouling and corrosion conditions.

Up to 50% of the refinery's entire corrosion expense may occur in the crude unit. The root cause of the fouling and corrosion issues can often be linked to the basic unit configuration choices. Once the unit configuration is set, anti-fouling chemical addition and unit operating conditions limit but never completely solve, the fouling and corrosion issues. An overview of major corrosion factors and a case study are presented.

CORROSION SYSTEMS—BACKGROUND

The major corrosion problem in crude unit overhead systems comes from chloride-induced corrosion. The "chloride-induced" corrosion includes two major mechanisms: 1) hydrochloric acid corrosion, and 2) under-deposit corrosion (UDC).

FIG. 1 illustrates the areas where the crude unit suffers from chloride-induced corrosion. Complexity results from the interaction of operating conditions, mitigation strategies and materials selection. Chlorides are not new problems; they are a continuation of old problems, but under more severe conditions.[1]

Chlorine source. Chlorides enter the crude unit as a component of dissolved salts in water mixed with the crude oil. The dissolved salts include sodium chloride (NaCl), magnesium chloride ($MgCl_2$) and calcium chloride ($CaCl_2$). In special cases, organic chlorides may be present. At one time, carbon tetra-chloride was used as a well-treating chemical. Although this is no longer done, crudes may become contaminated with synthetic organic chlorides.[2,3] To a lesser extent, chlorides can also enter the unit as entrained solids protected by an oil film.

Upstream of the desalter, the temperature is low enough that salt corrosion is not a major issue. The crude unit desalter uses electrostatic precipitation to extract most of the water and reduce the chloride content. However, the desalter does not remove all the chlorides, resulting in some chloride contamination in the atmospheric heater.

The native chlorides themselves are not the cause of corrosion. Instead, at 350°F to 400°F (175°C to 205°C), the $CaCl_2$

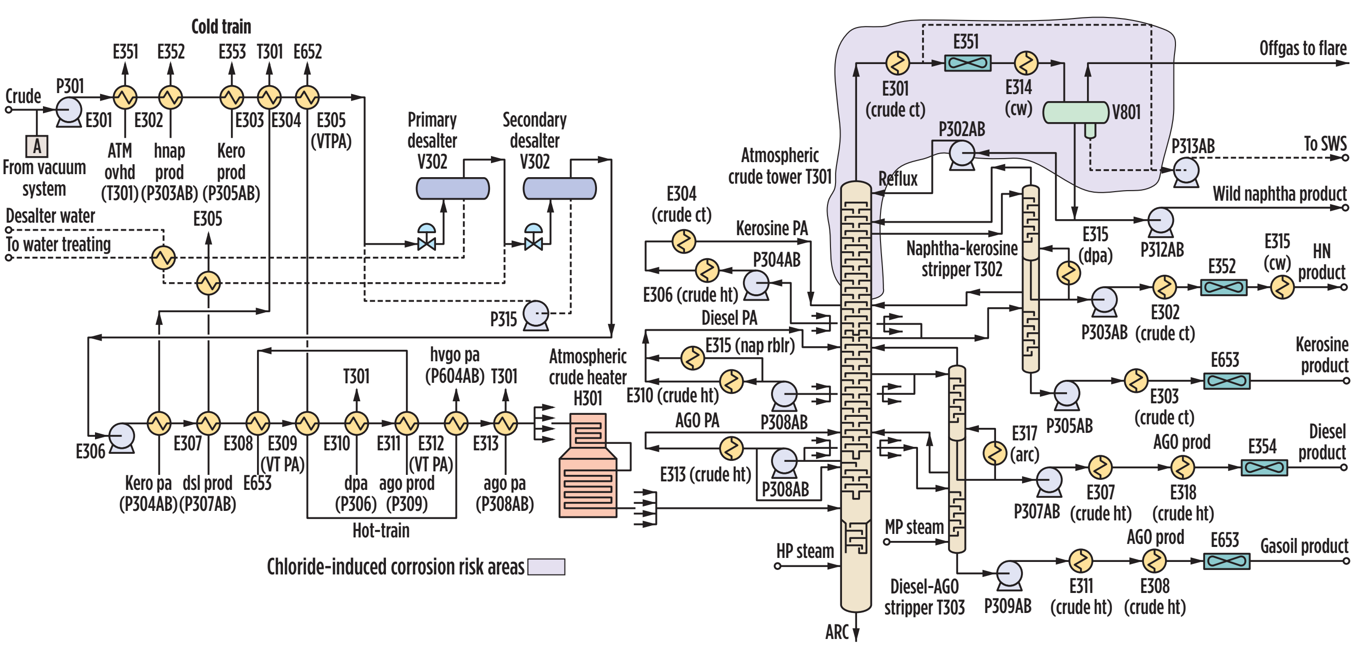


FIG. 1. Areas of chlorine-induced corrosion on a crude distillation unit.

and $MgCl_2$ thermally decompose (disassociate). The chlorine rapidly reacts with hydrocarbons, producing HCl. At normal crude atmospheric heater operating conditions, nearly all of the $MgCl_2$ and much of the $CalCl_2$ will hydrolyze. While NaCl may start to hydrolyze at temperatures as low as 450°F (232°C), temperatures above 800°F (425°C) are often required for significant cracking. NaCl decomposition is not a substantial chloride source for most crude atmospheric tower overhead systems.

Chlorine consequences. HCl, in the absence of water, does not significantly corrode carbon steel (CS). The overhead system of the crude tower condenses water. The water absorbs the HCl, creating hydrochloric acid. At this point, many outcomes are possible, depending on the location of where the first drop of water forms. The first drop of water will occur at the point where film temperatures reach the dewpoint. Bulk temperatures are often reviewed, but this will miss many problems, as surface temperatures are what count. The initial drop of water may have very high HCl concentrations, making it extremely corrosive.

The water also absorbs ammonia (NH_3). HCl combines with NH_4 and forms ammonium chloride (NH_4Cl). In situations where the water revaporizes, solid deposits of NH_4Cl form, thus creating the potential for UDC.

Temperature predictions for dewpoints are often based on partial pressures from steam table values. This method will underpredict the dewpoint temperature. Ionic interactions between multiple species—water, hydrogen sulfide (H_2S), NH_3, HCl, cyanide and others—increase the dewpoint temperature. In some reported cases, dewpoint temperatures exceeded 25°F (14°C) or more above steam table based predictions.

Processing configuration will also affect chloride corrosion:

- Preflash columns or drums often receive feed hot enough to contain HCl. Preflash configurations may also reduce the amount of naphtha in the atmospheric column overhead. This increases the partial pressure of water in the atmospheric column, increasing the water dewpoint temperature.
- Columns with heavy naphtha (HN) draws have more problems. The HN draw makes the column overhead composition lighter, reducing its temperature, making corrosion situations more likely.
- Common overhead systems may include either one or two overhead drums. However, some crude units have included up to five overhead drums. The cooling levels between each of the drums become significant in determining the initial point for corrosion. The purpose of multiple overhead drums is for heat recovery from the overhead vapor. Heat recovery implies corrosion locations where water condenses.
- Other situations may lead to localized water condensation. This includes cold reflux and cold pumparound (PA) returns, creating shock condensation inside the atmospheric columns. Shock condensation can also occur due to sub-cooled feed to the crude column when connecting the atmospheric column and preflash column.
- Higher stripping steam rates improve distillate recovery. They also increase steam partial pressure and make water condensation more likely.
- High NH_3 concentrations can lead to a NH_4Cl deposition directly from the vapor phase to a solid phase.[4, 5] This can occur inside the atmospheric column (most common), overhead lines or in overhead exchangers. The NH_4Cl then absorbs moisture, starting the process of UDC.

Common methods of alleviating chloride corrosion include additives, modification of unit design, selection of operating conditions, corrosion inhibitors and materials selection. Other strategies include using a water wash to ensure that the initial water condensation location is controlled, and the corrosive species are quickly diluted. Combining all of these factors creates tremendous complexity.[6]

Previous articles have discussed the options for materials selection and have focused on water wash and the results of water wash on unit HI.[7] The potential for reduced heat recovery poses the major question for water wash.

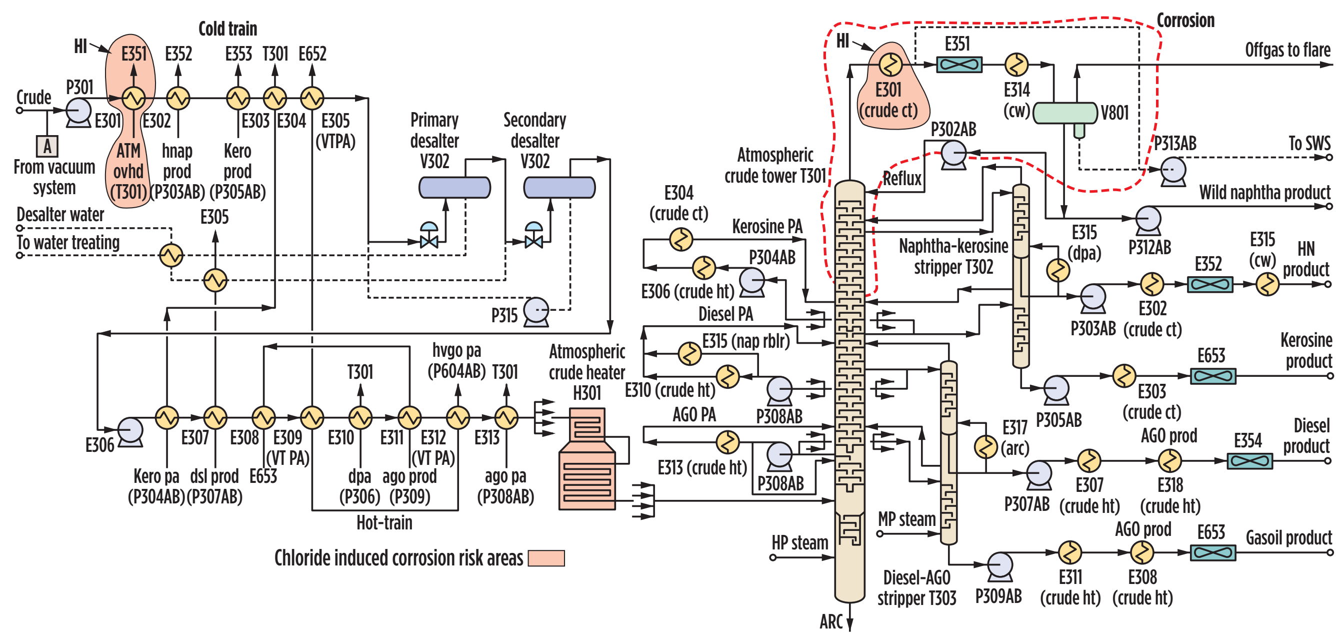


FIG. 2. Over head-to-crude heat integration and corrosion example.


SO$_x$
REGENERATIVE SO_2 SCRUBBING SYSTEMS
THIRD STAGE SEPARATORS
NO$_x$
LABSORB™
EDV® SCRUBBER
FCCU SCRUBBERS
SRU TAIL GAS SCRUBBERS
SEMIDRY AND WET SCRUBBING SYSTEMS
FEASIBILITY STUDIES
BELCO
a DuPont Company
SCR NO$_x$ CONTROL SYSTEMS
WET ELECTROSTATIC PRECIPITATORS
NO$_x$ CONTROL SYSTEMS
ACID MIST
PARTICULATES

UNIT CONFIGURATION

Crude overhead-to-feed HI is a relic of a bygone era. With low-density crude, making high yields of naphtha, without preflash drums or towers, and short-run durations (often 12–18 months) between shutdowns, these heat-recovery systems can work. In refineries processing difficult-to-desalt crudes, with preflash expansions, lower naphtha yields, and runs of four years to five years, the crude-overhead-to-preheat is a major and continual problem. Nevertheless, about 70% of operating crude units have HI between the overhead and the crude preheat. **FIG. 2** shows a typical older crude unit with the HI between the tower overhead with the crude feed highlighted. A well-heat integrated unit can recover all the necessary preheat without the overhead HI service. Crude overhead-to-feed preheat should only be used on new units in exceptional cases.

Configuration analysis criteria. To evaluate a crude unit, identifying these locations are important:

- Areas where water condensation starts
- Areas where water may revaporize
- Areas susceptible to solid salt deposition
- How water can move from one part of the unit to another.

Analysis of crude unit corrosion potential should examine all of these elements to determine the mitigation strategy required.

Where does water condensation start? The critical point in corrosion from chlorides is the initial point of water condensation, which is driven by surface temperatures rather than bulk temperatures. It is very difficult to measure surface temperatures, so bulk temperatures are typically measured. Common practice often adds temperature controls to the overhead of crude columns and preflash columns to prevent water condensation. Typical practice sets a tower overhead temperature 30°F to 50°F (17°C to 28°C) hotter than the calculated dewpoint.

Shock condensation occurs when cold reflux enters the tower and creates localized water condensation. This is a common mechanism for corrosion inside the tower. Operating at maximum reflux rates allows an increased reflux return temperature for the same heat removal. Within the constraints of tower and external equipment, reflux and PA rates should be maximized.

Where does water revaporize? Once formed, water flows with the bulk hydrocarbon stream. As the water flows down the column, the water enters a hotter zone. Eventually, the water will vaporize, increasing the concentration of corrosive solutes. Finally, the water will completely vaporize, leaving a solid deposit, which creates the potential for UDC.

Where can solid salts form directly from the vapor? Solid salts, especially of NH_4Cl, can directly condense from the vapor. The condensation can be analyzed similar to a solubility product with liquid solutes. If the product of the ammonia concentration, (NH_4^+), times the concentration of the chlorine, (Cl^-), exceeds a specific value, NH_4Cl deposits. This can also be expressed as an NH_4Cl partial pressure. The higher the NH_3 and chloride concentrations, the higher the temperature at which solid salts form.

Solid salts may form above the water dewpoint temperature. This creates solid salts even if the system remains dry. If the solid salt deposition temperature is below the water dewpoint, the NH_3 and chloride are absorbed by the water instead. Many additive systems have been proposed, and used, to lower the solid salt formation temperature below the water dewpoint temperature.

The solid salts can form in two major locations. The first is the crude column and may be as far down as the kerosine draw tray. The second is in high-temperature overhead systems. If the crude column takes kerosine (or heavier) overhead, then the salts can form in the overhead exchangers.

How does water move around? Extreme corrosion can result from overhead drum water returning to the crude tower. The water includes chlorides, other salts and treating chemicals. These create deposits in the tower that lead to UDC. Levels should always be carefully monitored to prevent water carryover to the crude column and preflash column.

CORROSION AND DEPOSIT CONSEQUENCES

The major consequences to the crude unit include direct corrosion and pressure drop due to fouling deposits. Direct corrosion from the aqueous phase reduces unit life and increases maintenance costs. It can also lead to unit shutdowns. UDC has the same direct consequences. When they occur, unplanned shutdowns incur the highest expense, and may cost $2 million or more per day for a large refinery. In extreme cases, units have had to shut down due to corrosion causing a leak. Leaks in the upper part of the crude unit are likely to contain some mix of H_2S, light gases and naphtha. H_2S would be a direct hazard. Naphtha and light gases are a significant fire hazard. With a flash point of –70°F (–57°C) or lower, naphtha will ignite easily.

Solid deposits have a second consequence, given their ability to cause fouling and block flow. Fouling in the atmospheric column may block draw nozzles, and it often plugs trays. Fouling in the overhead exchangers increases pressure drop in the system. Increased pressure drop raises the tower operating pressure. Many atmospheric crude units run at a feed temperature limit. Higher operating pressure at a fixed operating temperature drops liquid yield in the unit. Lower liquid yield forces more material downstream, creating constraints in downstream vacuum processing. Fouling in overhead exchangers also reduces heat removal capability of the overhead system. Reduced heat transfer may limit production rate or yield as well.

Exact consequences will vary from unit to unit. However, they are all bad. Maintenance costs increase, reliability drops, capacity is reduced, and yields decrease.

MITIGATION STRATEGIES

Such actions include corrosion inhibitors, operating condition limits, feed quality, materials upgrades and unit configuration choices. Unit configuration will be reviewed from two directions:

- What configurations reduce corrosion and fouling?
- What configurations, imposed by conventional refinery economics, force the plant to use configurations that increase the possibility of fouling and corrosion?

The basic configuration choices will focus on the crude tower overhead system and on the possible split between over-

head products on the crude unit. The overhead system configuration choices include three major parameters:

- Number and type of overhead drums
- Inclusion of water wash
- Integration of overhead duty into crude preheat.

The product split factors include:

- Full-range naphtha (FRN) product
- Split between light naphtha (LN) and HN products
- Heavy product overhead (naphtha plus kerosine).

In addition, some units have switched the sequence of exchangers to initially warm up the cold crude before it reaches the overhead-to-crude exchangers. This increases the minimum oil-film temperature in the overhead exchanger. Higher film temperatures reduce the risk of localized water condensation. All of the mitigation strategies that use a specific configuration (water wash, different preheat configurations) sacrifice recoverable heat from the crude tower overhead to reduce corrosion rates.

Introducing overhead systems.

Several possible options are available with different numbers of overhead drums used:

One-drum overhead system with reflux. As shown in FIG. 3, this atmospheric crude unit uses a one-drum overhead system. It is a similar concept to the crude unit, as shown in FIG. 1. Overhead vapors from the crude tower are condensed and collected in an overhead drum, and the reflux is sent back to the crude tower. It is a common configuration. The system illustrated uses an air-fin exchanger followed by a cooling water exchanger. Vapor loading in the overhead system is high. The overhead must generate all the product naphtha plus reflux.

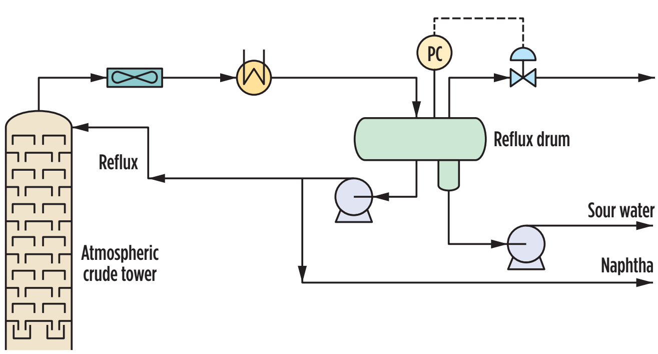


FIG. 3. One-drum overhead system with reflux.

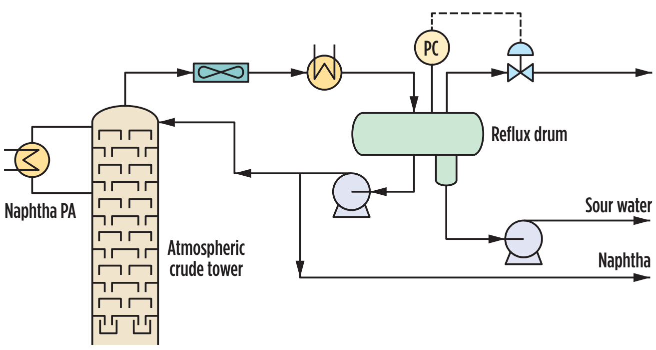


FIG. 4. One-drum overhead system without reflux.

One-drum overhead system without reflux. Pressure drop in the overhead system reduces unit liquid yield. One step to increase yield shifts some overhead heat removal from the overhead condensers into a PA-heat removal loop. FIG. 4 shows a naphtha PA added to the top of the crude column. This shifts heat removal from a condensing service to a subcooling service. The naphtha PA usually provides heat for feed heating. Corrosion and fouling problems in the naphtha PA exchangers are usually minor. However, the PA heat removal lowers the tower overhead temperature. Some reflux from the overhead drum may be added to the naphtha PA return to control naphtha product composition. With colder overhead temperatures, little heat is available in the overhead system for heat recovery. FIG. 4 shows the remaining overhead condenser duty going to air and cooling water.

Two-drum overhead system with dry reflux. The first condenser nominally operates at temperatures to support heat integration. The second condenser uses heat rejection. FIG. 5 shows a two-drum system. At average conditions in the first drum, it is too hot to allow water accumulation.

The intended benefit of this configuration is to allow heat recovery while minimizing water formation and corrosion problems. FIG. 5 shows a nominal heat recovery service as the first exchanger in the system. This system has a higher total pressure drop than the one-drum system without reflux, but it avoids investment in the naphtha PA pumps.

Two-drum overhead system with wet reflux. Flexibility to handle different crude compositions may make consistent dry-drum operation difficult. Having the capability to allow for water accumulation in the first drum allows the unit to operate across a wider range of operating modes. FIG. 6 illustrates a system with water condensation in both drums. Additionally, if correctly managed, the wet-drum system allows for higher energy recovery. The disadvantage of the wet-drum system is that corrosion and fouling benefits are less clear.

Three-drum (plus) overhead system. Adding more drums allows for more complicated heat removal systems. Conceptually, this improves heat recovery. FIG. 7 illustrates a three-drum system. While systems with more than two drums are unusual, some units have been built that have five drums in the overhead.

Units that send a mixed naphtha-kerosine stream to a common hydrotreater shift the benefits of selection criteria toward the multiple-drum configurations. The equilibrium conditions in the first drum of three-drum systems reliably lie above the water dewpoint. However, localized water condensation still

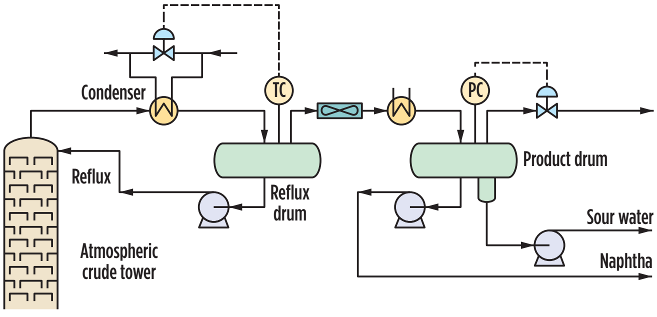


FIG. 5. Two-drum overhead system with dry reflux.

presents challenges. Multiple-drum systems also impose pressure drops in series. This increases unit operating pressure, decreasing liquid yields.

Product splits matter—no water wash. The previous illustrations stayed deliberately vague on product splits in each configuration. Specific product splits do matter. This set of examples uses a basis of no water wash to examine the possible consequences. Specifics in a given plant may vary greatly. Nevertheless, the analysis shows the important points to consider.

One-drum overhead system with reflux. The different operating modes for a one-drum system with reflux are illustrated in FIGS. 8 and 9. In FIG. 8, a FRN is produced as the overhead product. FIG. 9 shows the changes with a light-naphtha product overhead.

FRN product. A typical operation with FRN product will operate at approximately 300°F (149°C) overhead. This provides ample driving force for heat integration with crude feed. Normal practice has the overhead condenser as the first crude preheat exchange, as shown in FIG. 2. Crude feed temperature may vary from 65°F (18°C) to 100°F (38°C). Downstream of the preheat, overhead temperatures will drop to 200°F–240°F (93°C–116°C). Some combination of air and water cooling will drop the overhead drum temperature to 100°F (38°C). FIG. 8 shows kerosine as the upper side-draw from the crude column.

Under these conditions, the bulk of the water condensation may occur in the overhead exchangers, and surface condensation on the tubes is certain. Tube metal temperatures may drop as low as 135°F (57°C). Localized presence of hydrochloric acid in the overhead exchangers is nearly certain. Standard CS tubes will likely suffer from HCl attack. Fouling and UDC may occur in the exchangers as well. The exchangers are cold enough that, once condensation starts, the water is unlikely to completely revaporize. However, local flow regimes may create areas where deposits form.

The cold reflux returns to the crude tower at 100°F (38°C). Localized water condensation occurs on the top tray, making fouling and corrosion attack in the tower top highly likely. Corrosion is initiated by water formation on the top tray. Water vaporizes as it descends, causing deposits that can lead to UDC. The overhead drum rarely suffers from severe corrosion. Enough water is present to dilute the HCl. Revaporization does not occur, so solids deposits do not present major problems.

Product split between LN and HN. A typical operation with a light straight-run naphtha (LSRN) product will operate at approximately 200°F–240°F (149°C–116°C) overhead. Driving force remains sufficient for heat integration with crude feed as long as the crude exchanger is the first in the preheat sequence. As previously, crude feed temperature may vary from 65°F (18°C) to 100°F (38°C). After preheat, overhead temperatures will drop to as low as 140°F (60°C). Some combination of air and water cooling will drop the overhead drum temperature to 100°F (38°C). FIG. 9 shows HN as the upper side-draw from the crude column. Under these conditions, bulk condensation of water will occur in the overhead exchangers. Tube metal temperatures may drop as low as 100°F (38°C). Standard CS tubes are nearly certain to suffer from HCl attack.

Fouling and UDC, if present, will concentrate at the exchanger inlet area. The exchangers are cold enough that once condensation starts, water is unlikely to completely revaporize. However, local flow regimes may create areas where deposits form.

The cold reflux returns to the crude tower at 100°F (38°C). Localized water condensation occurs on the top tray. Fouling and corrosion attack in the tower top are highly likely. As the water descends and vaporizes, deposits are left and UDC occurs. The corrosion area may reach down to as low as the kerosine draw. The overhead drum rarely suffers from severe corrosion. Enough water is present to dilute the HCl. Revaporization does not occur, so solids deposits do not present major problems.

Plant economics and product specifications drive the necessity for splitting the naphtha into LSRN and HN. This split aids control of downstream isomerization and reformer unit operations. It also eases meeting gasoline benzene specifications in many plants. The economics behind the LSRN-HN split are usually so large, that, if this split is attractive, some method will be found to deal with the consequences.

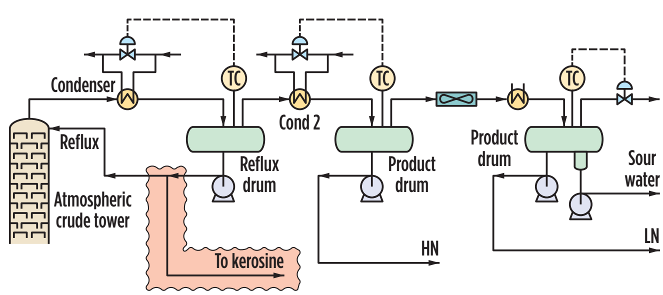


FIG. 7. Three-drum overhead system.

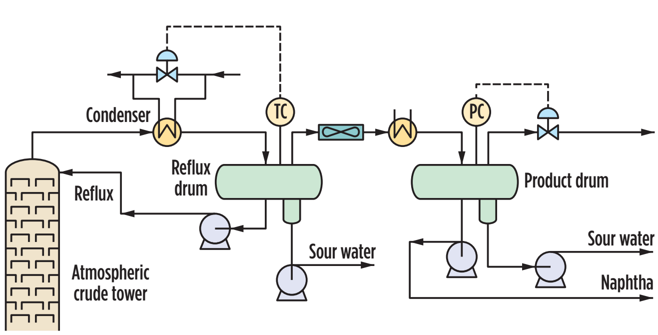


FIG. 6. Two-drum overhead system with wet reflux.

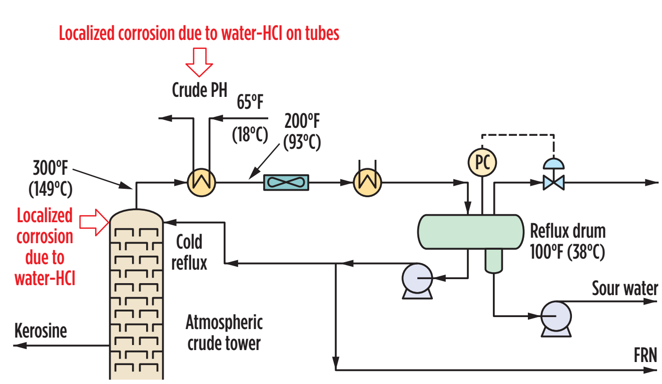


FIG. 8. One-drum overhead system for FRN overhead: typical operation.

One-drum overhead system without reflux. While slightly different, the performance of one-drum systems without reflux closely resembles two-drum systems with dry reflux.

Two-drum overhead system with dry reflux. The different operating modes for a two-drum system with dry reflux are represented in **FIGS. 10** and **11.** In **FIG. 10,** FRN is the overhead product. **FIG. 11** shows the changes with an LN-product overhead and an HN side-draw.

FRN product. A typical operation with an FRN product will operate at approximately 340°F (171°C) overhead. This provides ample driving force for heat integration with crude feed. In this case, the first overhead exchanger may not be the first exchanger in the crude preheat train. Crude feed temperature may vary from 65°F (18°C) to 150°F (66°C). **FIG. 10** shows operation with cold crude.

Downstream of the preheat, overhead temperatures will be 276°F (136°C). Some combination of air and water cooling will drop the overhead drum temperature to 100°F (38°C). **FIG. 10** shows kerosine as the upper side-draw from the crude column.

Under these conditions, bulk condensation is unlikely in the overhead exchangers. Surface condensation on the tubes is possible. The main determinant of localized water condensation is the crude inlet temperature. If the crude inlet temperature is low, 65°F (18°C), localized condensation is highly likely. If the crude inlet temperature is high, 150°F (66°C), tube metal temperatures may reach 220°F (104°C). In units with little steam and efficient desalting, water condensation may not occur with the hotter crude. However, operating upsets could lead to deposit formation and UDC. The first drum overhead temperature at about 276°F (136°C) remains high enough to add heat integration between the two drums. Consequences here closely resemble those in the one-drum system with the FRN product overhead.

Hot reflux returns to the crude tower at 276°F (136°C). Corrosion in the crude tower is unlikely. One complicating factor comes from the naphtha endpoint control. The hot reflux and FRN product systems inter-connect to allow for FRN endpoint control. Large amounts of FRN added to the reflux may cool the reflux to start localized water condensation in the crude column.

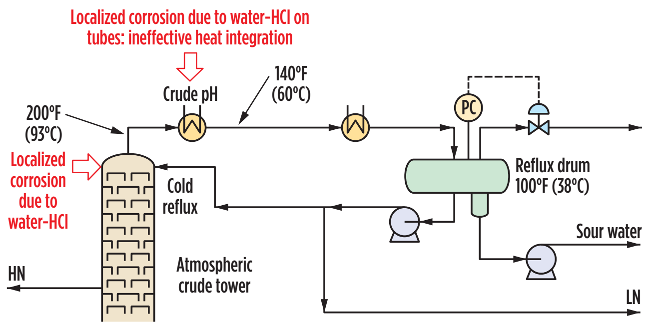


FIG. 9. One-drum overhead system for LN overhead: typical operation.

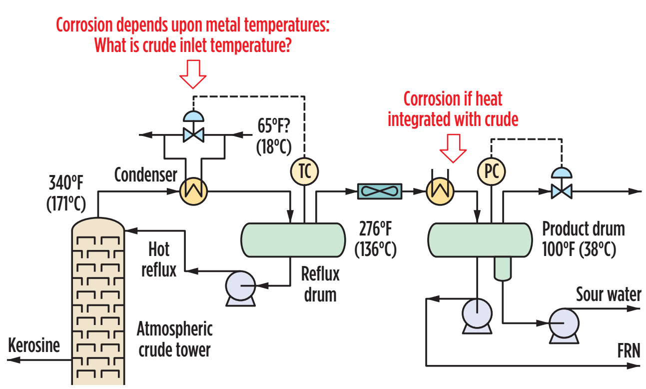


FIG. 10. Two-drum overhead system with dry reflux for FRN overhead: typical operation.

More split options for naphtha.

Splitting the LSRN and HN cools the tower overhead. The LSRN-HN split can be done by having an HN draw from the main column or by having segregated naphtha products from the two overhead drums. **FIG. 11** shows segregated products from the overhead drums. The tower overhead temperature drops to about 316°F (158°C) overhead. This still provides ample driving force for HI with crude feed.

Downstream of the preheat, overhead temperatures go lower, to about 200°F (93°C). This temperature is set by the need to get a specific composition for the LSRN-HN split. Some combination of air and water cooling will drop the overhead drum temperature to 100°F (38°C). **FIG. 11** shows kerosine as the upper side-draw from the crude column. Under these conditions, bulk condensation is highly likely in the overhead

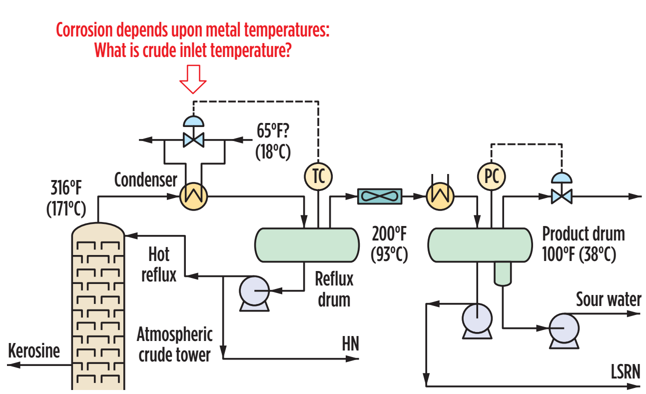


FIG. 11. Two-drum overhead system with dry reflux—two naphtha overhead: typical operation.

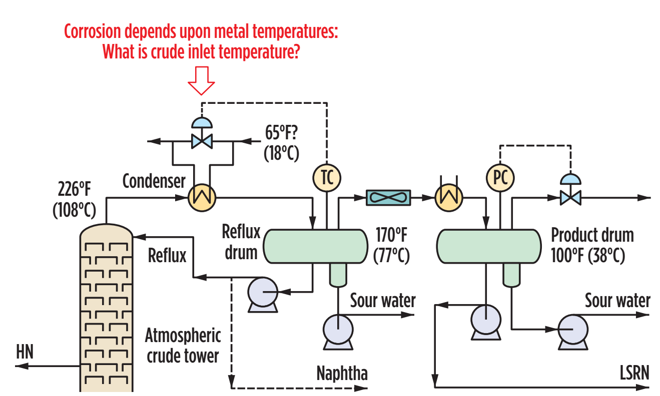


FIG. 12. Two-drum overhead system with wet reflux—two naphtha overhead: typical operation.

exchangers. To get the 200°F (93°C) drum temperature, cold crude is required. Localized water condensation is nearly guaranteed. Reflux returns to the crude tower at 200°F (93°C). Corrosion in the crude tower is likely to be minor. If it occurs at all, localized water condensation is slight and will be concentrated on the top tray.

Two-drum overhead with wet reflux and LSRN-HN split. Two-drum systems with wet reflux require a low first drum temperature. This configuration is unlikely unless the intent is to draw an HN product from the atmospheric column. **FIG. 12** shows this configuration. The upper draw from the main column is HN. A typical operation will operate at approximately 226°F (108°C) overhead. Driving force remains sufficient for HI with crude feed as long as the crude exchanger is the first in the preheat sequence. As previously, crude feed temperature may vary from 65°F (18°C) to 100°F (38°C). After preheat, overhead temperatures will drop to as low as 170°F (77°C). If the overhead temperature is much lower, then the system should really be considered a one-drum system with a booster drum. Some combination of air and water cooling will drop the overhead drum temperature to 100°F (38°C).

Under these conditions, bulk condensation of water will occur in the overhead exchangers. Tube metal temperatures may drop as low as 120°F (49°C). Standard CS tubes are nearly certain to suffer from HCl attack. Fouling and UDC, if present, will concentrate at the exchanger inlet area. The exchangers are cold enough that once condensation starts, water is unlikely to completely revaporize. However, local flow regimes may create areas where deposits form.

The reflux returns to the crude tower at 170°F (77°C). Modest localized water condensation occurs on the top tray, making fouling and corrosion attack in the tower top highly likely. The corrosion area may reach down to as low as the kerosine draw. The overhead drum rarely suffers from severe corrosion. Enough water is present to dilute the HCl. Revaporization does not occur, so solids deposits do not present major problems.

Next month. In Part 2, the author discusses the benefits of water-wash systems in crude units. HP

LITERATURE CITED

1. Kronenberger, D. L., "Corrosion problems associated with the desalting difficulties of Maya and other heavy crudes," Corrosion 1984, New Orleans, April 2–6, 1984.
2. NACE, "Effect of Nonextractable Chlorides on Refinery Corrosion and Fouling, "NACE, Houston, 2005.
3. Gutzeit, J., "Effect of organic chloride contamination of crude oil on refinery corrosion," NACE 00694, Corrosion 2000. Orlando, 2000.
4. Carlton, R. H., "Refinery condenser and exchanger corrosion by ammonium chloride," *Materials Protection,* January 1963, pp. 15–20.
5. Wu, Y.-M "Calculations estimate process stream depositions," *Oil & Gas Journal*, Jan. 3, 1994, pp. 38–41.
6. NACE, "Crude Distillation Unit–Distillation Tower Overhead System Corrosion," Houston, Texas, 2009.
7. Sloley, A. W., "Heavy oils processing requirements–Crude processing," Rio Oil & Gas 2012, Rio de Janeiro, Brazil, Sept. 17–20, 2012.

J. McMULLEN, Invensys, Boston, Massachusetts

Use modeling to evaluate processing solutions for heavy oils

In the oil segment of the global energy industry, feedstocks are moving toward the heavier end. The extraction, transport and processing of lighter crudes, using a stable of proven technologies, is no longer possible. These light oil supplies have become exhausted or more difficult to obtain, causing strain on the crude oil supply chain. Explosive economic growth in India, China and other emerging countries requires large volumes of conventional crude oils as well as increasing supplies of heavy oils.

Heavy crude oils have unique physical properties that often present costly refining challenges. Efficient processing solutions require a thorough understanding of the oils' characteristics and operating behaviors. Companies involved in heavy or extra-heavy oil refining must evaluate these characteristics and choose the best processing options for these oils.

Heavy oil trends. With increasing global demand and elevated prices for crude oil, heavier oils are still economical to produce using advanced technologies, despite the challenges associated with their production and processing. Their discounted market prices, relative to West Texas Intermediate (WTI), Brent and other benchmark light oils, can provide a financial incentive for refineries capable of processing these heavy oils. Conventional heavy oil reserves are plentiful in the Orinoco region of Venezuela, while the largest concentration of extra-heavy unconventional oil is found in the rich oil sands of western Canada.

Global heavy oil reserves have been estimated at more than twice those of conventional light oils, at six trillion barrels. This massive volume makes heavy oil an important energy resource, especially as conventional oil sources continue to diminish. US refineries import the majority of their heavy oil feedstocks from Canada. Amid global political volatility, Canada's heavy oil resources provide a politically and logistically secure supply of energy, despite the difficulty involved in their extraction, transport and refining.

There is a global need for alternatives to light, sweet, conventional oils. The industry has adopted methods to recover and manage increasingly heavier oils that require additional processing steps. These heavy and extra-heavy oils have complex characteristics that must be identified to be efficiently processed. The knowledge of how to process these oils is important for designing and operating the process correctly, for reducing downtime, for increasing efficiency, for optimizing profitability, for promoting higher environmental standards and for effectively eliminating safety concerns.

Heavy oil characteristics. Heavy oil is generally defined as a viscous crude oil with an API gravity between 10° and 22°, and a viscosity of less than 10,000 centipoise. Extra-heavy oils are unconventional oils with an API gravity below 10°; they are bitumen-like substances with extremely low flowrates at the reservoir. In many instances, they have been considered "bottom of the barrel," as compared to conventional petroleum sources with lower viscosity and higher API gravity. Their heaviness is generally attributed to high-molecular-weight compounds, including asphaltenes, which contribute greatly to feed viscosity and coking tendency. They contain relatively small amounts of paraffinic components and are more naphthenic and aromatic in nature. While heavy and extra-heavy oils are present in many global regions, the largest concentration of supply is found in North and South America.

In addition to high viscosity, high pour point and low API gravity, heavy and extra-heavy oils are characterized by higher levels of sulfur, nitrogen and heavy metals, including mercury (Hg). They have low hydrogen-to-carbon ratios and high carbon residue. These feedstocks often contain elevated amounts of particulates and water and are generally high in acids, particularly naphthenic acids.

Issues posed by heavy oils. Heavy and extra-heavy crude oils present a number of challenges, beginning with their extraction and continuing through refinery transportation and processing (FIG. 1). As oil sands recovery has progressed, the surface extraction of oil sands with strip-mining technology has been superseded by several thermal methods, with the most predominant being steam-assisted gravity drainage (SAGD). SAGD and similar technologies have allowed heavy oil to be recovered at lower depths as surface deposits have been depleted.

Newer methods have reduced the environmental impact of extraction while enabling access to a higher percentage of known subsurface reserves. SAGD technologies rely heavily on natural gas to raise steam. As a result, this extraction process generates a higher concentration of carbon dioxide, raising environmental concerns. However, low gas prices have helped make SAGD an attractive extraction process.

Due to the viscosity of heavy crude oils, especially bitumen-type material from Canada, these oils are not shipped via pipeline unless they are first blended with a diluent. This blend is often referred to as dilbit, a naphtha/bitumen blend, and it is needed to facilitate flow. Bitumen properties and ambient temperatures at origin and along the pipeline route determine the percentage of naphtha that is utilized. Natural gas condensate can also be used as a diluent.

Heavy oil can be diluted for pipeline transport with a synthetic crude oil (SCO), which is produced by partially upgrading the heavy oil through prerefining distillation processes to lower the viscosity. This synthetic oil/bitumen blend, known as synbit, adds to the cost of the heavy oil. As the quality and attributes of dilbit and synbit diluents can vary considerably from one upgrading facility or supplier to another, their removal and recycle present additional factors that must be understood to effectively refine heavy oils.

Supply trends impact, and will undoubtedly continue to affect, refining capabilities. Most existing refineries were not originally designed to be feedstock flexible or to accommodate heavy feeds.

Refining challenges. Supply trends impact, and will undoubtedly continue to affect, refining capabilities. Traditionally, refineries were designed to accommodate light, sweet crude oils—usually a specific blend that may no longer be available or that is presently cost prohibitive. Most existing refineries were not originally designed to be feedstock flexible or to accommodate heavy feeds.

Heavy oils require additional processing steps to remove impurities and to provide a full spectrum of products from a variety of feedstocks. For this purpose, the addition of hydrogen (H_2) in the hydrocracking process is necessary. The crude oil combines with the H_2 at high temperature and pressure, in the presence of a catalyst, to saturate aromatic molecules, separating out the lighter streams. Subsequent reactor stages further separate the hydrocarbon components, increasing the yield of low-boiling-point, high-value fluids and middle distillates, while leaving the heavier residue to be converted into coke in a separate operation.

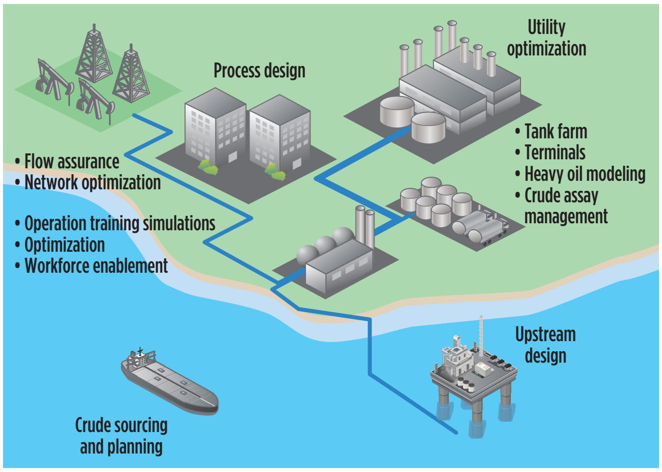


FIG. 1. Heavy oils pose a number of problems for upstream and downstream operations.

Heavy oils bring additional levels of heavy metals, such as Hg, vanadium, magnesium, nickel and iron. It is imperative to assess and remove certain metals, particularly Hg. Even if Hg levels are low, the presence of this metal in large volumes of liquid hydrocarbons represents significant exposure to processing equipment. Non-removal of Hg can cause metal embrittlement and failure due to corrosion, and it can pose a health hazard to refinery workers. Furthermore, Hg levels in C_3–C_6 product streams from the crude distillation column and in water effluent can poison catalysts.

Other materials commonly found in heavy oils can cause a myriad of operational problems if not eliminated. Calcium (Ca), present as calcium naphthenate, can cause fouling at the desalter, catalyst poisoning and scaling issues that require increased maintenance for heat exchanger tubes and other equipment internals. Ca deposits often require additional processes for elimination, such as demulsification.

Many of the new heavy oil sources have increased sulfur content and high levels of other impurities. Due to the compositional variability of heavy oils from different fields, processing technologies cannot be uniform, but instead must be structured to accommodate the specific characteristics of each oil.

High densities and viscosities require higher temperatures in refining units. In desalters, higher temperatures can compromise the heat energy balances designed for light oils. Additionally, the higher levels of solids found in heavy oils can lead to sludge, which can pose storage problems and have undesirable impacts on wastewater treatment and other offsite operations.

Delayed coker residue from the vacuum-reduced heavy oil presents another potential bottleneck in the refinery. As a batch process, this residue accumulates in the coker drums. If it is not removed in a timely manner, the refinery could back up and be forced to shut down.

Once a refinery has shifted its operational capabilities to produce high-value products from heavy oil, the facility must continue on this path, as significant modifications would be needed to convert the refinery feedstock slate back to light oil. A number of US Gulf Coast refineries have made the switch to heavy oil, and they rely on the delivery of Canadian crude oils. As a result of this scenario, pipelines have been reversed and extended to serve these facilities.

Modeling for heavy oil refining. As the availability of crude oil supplies changes and the characteristics of these feedstocks are more varied, existing refineries are forced to become more flexible in their operational capabilities and to adapt to the diversity of input materials.

Operations planning and optimization are critical in an existing facility, where margins fluctuate depending on processing costs, product yields and the adaptability of each processing unit. As an example, fractionation requires exact measurements of pressure, temperature and volume to maximize yields from feedstocks with varying compositions.

All refineries struggle to process heavy oils economically. Modeling methods that accurately depict the characteristics of these heavy feedstocks allow refiners to adapt their pro-

cesses accordingly, whether by altering existing operating processes or by designing new equipment.

The ability to predict the performance behavior of heavy oils at each stage of the refining process is important. It is difficult to use traditional modeling software, as these programs were developed to predict the behavior of light, sweet oils. There is no single solution for heavier oil processing to suit all refineries. Each refinery has its own set of parameters governed by its existing configuration, the composition of the oils being processed, the operating conditions within the plant and the desired range of product output.

For example, asphaltenes are prevalent in heavy oils. The use of processes to precipitate them to lower viscosities can assist in refining. When diluted with solvents, including condensates, the dilbits from oil sands are easier to transport via pipeline. Therefore, to optimize processing, refinery operators must be able to predict the amount of asphaltene precipitated as a function of the amount and nature of the solvent and the ranges of temperature and pressure.

Addressing issues jointly. Utilizing field-proven, proprietary modeling software, one company enlisted a consortium of global industry leaders (i.e., major oil companies, Canadian heavy oil producers, engineering firms and national oil companies) to provide a forum for defining and solving major issues in heavy oil processing.

TABLE 1. Predicted kinematic viscosities using API procedure 11A4.2

	Crude kinematic viscosity, cSt		
Temperature	**Measured**	**API 11A4.2**	**Heavy oil**
60°F	49.75	111.6	70.65
80°F	27.95	58.77	39.69
100°F	17.52	34.26	24.38
	Vacuum resid kenimatic viscosity, cSt		
Temperature	**Measured**	**API 11A4.2**	**Heavy oil**
210°F	6,787	454	4,784
250°F	1,261	147.5	858.5
300°F	274.6	50.83	176.8

TABLE 2. Predicted kinematic viscosities using API procedure 11A4.2 and heavy oil correlation

Viscosity method	Measured values	API procedure 11A4.2	Heavy oil prediction
Duty, 10^6 Btu/hr	3.654	8.976	5.432
Log mean temperature difference, °F	165	150	159
Heat-transfer coefficient, Btu/hr/ft²/°F	4.657	12.62	7.144
Shell side			
T_{out}, °F	265	245	258
ΔP, psi	58	8	27
Tube side			
T_{out}, °F	110	119	113
ΔP, psi	9	13	11

The company is using the latest modeling technologies to predict how heavy oils will perform at various refinery processing stages. This software has identified several areas in need of improvement, and new and expanded models have been developed to increase prediction and plant optimization accuracy through advanced process simulation.

These efforts have been directed at multiple procedures: Preparing crude oil feed, investigating viscosity and thermal conductivity, maintaining H_2 balance and solubility as part of the upgrading process, removing Hg and other contaminants, and performing molecular-based characterization.

Knowledge of viscosity is critical to understanding heavy oil properties and potential product yields. The newest method of liquid viscosity prediction advances the accepted Twu correlation, improving the predictive accuracy of viscosities in the 100 cSt–100,000 cSt range and closely estimating temperature dependence, especially for low temperatures. Utilizing data from more than 125 heavy oil assays provided by the consortium, the accuracy of viscosity prediction in simulations is greatly improved.

In addressing liquid-phase thermal conductivity in heavy oils, there is limited industry data. However, in recognizing the comparable characteristics of solvent refined coal II, for which some data is available, accurate correlations have been established. Recent testing has determined a strong performance using the Sato-Riedel method, requiring only an estimate of critical temperature to produce the best level of accuracy.

Hydroprocessing is required to remove impurities in heavy oils. Hydrogenation increases the yield and converts low-value feedstocks into higher-value end products. With the addition of H_2 in the process, refinery optimization relies on H_2 management and its solubility in hydrocarbons. Laboratory measurements for H_2, hydrogen sulfide and ammonia vapor-liquid equilibrium with defined hydrocarbons have been used to fit equation-of-state binary interaction parameters to improve the accuracy of predicting the elements' solubility in hydrocarbon mixtures.

Hg is a contaminant in heavy oils that poses multiple challenges during refining. It can poison catalysts, contaminate wastewater, destroy process equipment and impair processes, and be a human health hazard. An accurate understanding of Hg's solubility in hydrocarbons is necessary for its mitigation.

Similarly, naphthenic acid is a prevalent and increasingly corrosive element in heavier oils. The most problematic acids are those with a molecular weight having a boiling point of 430°F–750°F. The acid concentration, density and viscosity of the oil must be assessed to predict its corrosion potential.

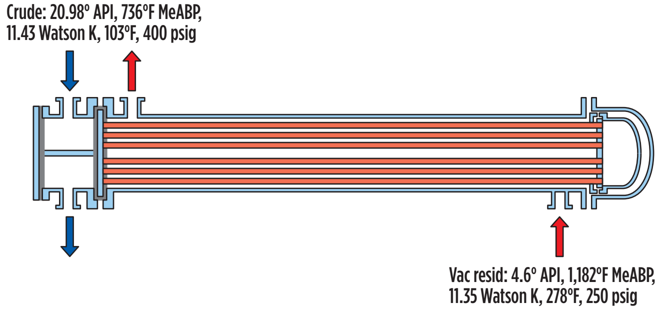
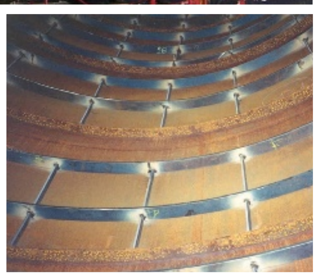


FIG. 2. A correlation developed for attributes of heavy oil and kinematic viscosity.


OVERSIZE LOAD




Our Safety Services are **a Breed Apart**

We're not one of the pack — We don't sell safety equipment. Safety isn't our side business either. Instead, we provide the safety manpower and methods to help you drive your safety culture and program from the top down and the bottom up during turnarounds, shutdowns and projects when your safety team is stretched thin.

Trained for better field performance — We augment your staff with specialists. They are specifically and rigorously trained to perform their assigned tasks. (Why pay craft/helper wages for holewatch/ firewatch?) We also provide permit specialists, planners, safety advisors, auditors/inspectors, coordinator services, HSE consulting and more. Select one, some or all of our services.

Good behavior gets rewarded — Our clients tell us how our services consistently save them time and money. One client estimated that our tracking process eliminated 900 hours of craft standby time alone. Another said our permitting processes impacted his critical path work saving around $800,000.

Call now to learn more about our reliable breed of safety.

CERTIFIED SAFETY

Bridging the Gap Between Safety and Productivity

Phone: 800-994-2339

E-mail: ssmith@certifiedsafety.net
www.certifiedsafety.net

League City, Texas • Benicia, California • Whiting, Indiana

Select 86 at www.HydrocarbonProcessing.com/RS

Kinematic viscosity predictability. The procedure for characterization using data from consortium members focuses on extrapolating the molecular weights of critical heavy oil components with normal boiling points exceeding 1,000 K. The importance of this procedure cannot be understated when designing process components, such as heat exchangers.

Using a newer correlation developed specifically for the attributes of heavy oil and kinematic viscosity, results can be compared to measured data and to data derived from the older API procedure 11A4.2. The new correlation exposes a significant difference in accuracy in heat exchanger sizing.

TABLE 1 shows predicted kinematic viscosities in centistokes at various temperatures using API procedure 11A4.2, compared to the heavy oil correlation shown in **FIG. 2.** In **TABLE 2,** the comparison shows that the duty would be almost 250% oversized. The newer heavy oil method predicts duty much closer to the measured data, especially as temperatures rise.

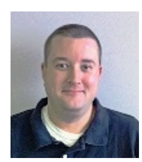
Takeaway. The nature of refining is continually changing with increasing global demand and feedstock supplies that are becoming heavier and more difficult to process. In addition, increasing government oversight and environmental constraints on emissions heavily impact refining operations.

To maximize profitability, refiners are faced with decisions to alter their operations to suit changing feedstock slates. Operational challenges are extremely difficult, if not impossible, to address without knowledge of the specific characteristics of different crude oils.

Heavy oils have become economical feedstocks and are adding to global hydrocarbon supplies. However, they require additional additives to combat high viscosity and high mass density. At the refinery, they need extra treatment phases that include adding H_2 during processing to eliminate or mitigate impurities and to alter their molecular structures.

The compositions of heavy oils vary, adding complexity to their processing. While models have been used for years to understand and optimize the processing of light oils, increased demand for heavy oil has led to the development of models to address the unique attributes of these oils.

To optimize processing and reduce design and operating costs, refiners should use accurate simulation models that are specifically tailored for heavy crude oils. Without this accuracy, operating and capital costs will escalate, and performance will diminish, at a time when refiners are seeking advantages to increase profitability. HP



JOSEPH MCMULLEN is the product marketing manager for the SimSci brand within Invensys. He earned a BS degree in chemical engineering in 2000 and an MBA degree in 2004, both from Villanova University in Pennsylvania. Since starting with Invensys in 2001 as a senior technical support specialist, Mr. McMullen has held multiple roles in product management and transitioned to product marketing in 2011. He regularly discusses simulation topics via his blog, http://simulationots.blogspot.com/

W. K. SHIFLETT, D. TORCHIA and **C. W. OLSEN,** Advanced Refining Technologies, Columbia, Maryland; and **D. BROSSARD,** Chevron Lummus Global, San Ramon, California

Catalyst systems offer flexibility for hydroprocessing applications

The dynamics of global gasoil (GO) and ultra-low-sulfur diesel (ULSD) demand are driven by regulatory demands for transport diesel and evolving needs for vastly improved marine-bunker quality in specific emission control areas (ECAs), as well as in economic growth. It is expected that this will drive regional variability in ULSD/0.1% sulfur GO differential over time and consequently provide incentives for refiners to maximize flexibility in both hydrocracker distillate yields and ULSD unit performance. Adding in an expectation of gasoline supply becoming long in the Middle East over the next several years, the refiner is further motivated to evaluate and select premier catalyst systems for the production of both diesel products from their hydrocracking and diesel treating units.

The growth in refined products is strongly driven by the demand for clean diesel and new regulations on GO between now and 2020. Notably, cumulative annualized global growth in diesel/GO demand is predicted to be about 2%, outpacing gasoline at 1.3% with the estimated gasoline/diesel-GO ratio dropping from 0.85 in 2012 to 0.81 in 2020.[1] In addition to steady growth in Asia and continued recovery elsewhere, new motor vehicle mileage standards, 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 and requiring 0.1% sulfur limits in ECAs in North America and Northern Europe will underlie a demand shift to diesel/GO products with an expected concurrent boost in diesel price primarily due to quality requirements.

Meanwhile, refining capacity additions will outstrip global demand in the 2015 to 2020 period, thus continuing pressure on refining margins. In short, it will be a period of opportunity for flexible refiners with hydrocracking capabilities, especially those coupled with a robust ULSD hydrotreater that can marry their catalyst system needs and operational responses to changing economic scenarios.

Molecular management and hydroprocessing units. 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 a hydroprocessing unit, largely due to the combination of olefins and the refractory nature of the LCO. It has the highest demand for hydrogen to produce a clean diesel or even 0.1% sulfur marine GO, and offers heat release management challenges when processed at higher fractions in a hydroprocessing unit feed. Provided there is adequate hydrogen supply, LCO is sometimes best processed in the hydrocracker along with other feeds such as automotive gas oils (AGOs) or vacuum GO (VGOs). Adding to the equation is that all LCOs are not created equal. Depending upon the refinery configuration, 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 GO can also be processed either in the hydrocracker or ULSD unit subject to individual unit capacities and infrastructure limitations such as hydrogen availability, pressure, cut point and impurities. Heavy coker GO (boiling well above the diesel range of > 975°F) will prove to be problematic for catalyst life cycles if processed in the ULSD unit and can present significant challenges to processing in significant quantities even in modern, robustly designed hydrocrackers. A hydrocracker originally designed or revamped for VGO service is a more suitable outlet.

This offers more potential to maximize diesel yields, especially in recycle flow configurations and higher pressures. With adequate hydrogen partial pressure and the appropriately tailored catalyst system to remove contaminants and provide sulfur conversion, light coker GO are readily processed to ULSD in the diesel hydrotreater.

Straight-run (SR) GO 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 the intent to pump hydrogen into the feedstock to convert heavier, higher-boiling materials into more valuable products while capitalizing upon aromatics saturation to increase volume swell, as well as product value parameters. Until recently, ULSD has been a secondary priority and generally not even a consideration in the original design of units operating today. With a robust ULSD in the refinery, this prioritization need not be overridden, but can be augmented by the utilization of the latest catalyst systems for hydrocracking that have been designed for maximum hydrodesulfurization (HDS) activity, as well as for the fundamental hydrodenitrogenation (HDN), hydrocracking and saturation needs.

Catalyst system flexibility. While several hydrocracker configurations are in usage, two dominate the landscape, especially when addressing clean fuels production: single-stage, once-through (SSOT) configurations and two-stage, recycle (TSREC) configurations.[2] A perspective over a decade of application is provided in **FIG. 1.**

Hydrocracking conversion spans a range from approximately 40% to 100%. The SSOT configuration is both simple and versatile, and it represents the simplest configuration when unconverted oil (UCO) has high value as either a lube plant feed or an FCC feed. This configuration dominates the low conversion market (> 70%). The SSOT process configuration is shown in **FIG. 2.**

Catalyst system optimization 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 HDS, while FCC feed will favor HDN, poly-nuclear aromatics removal and HDS. Balanced with these needs are the light product drivers: ULSD or the less-demanding 0.1% sulfur marine fuel. If ULSD production is a target and cannot be produced within the SSOT unit constraints, it is

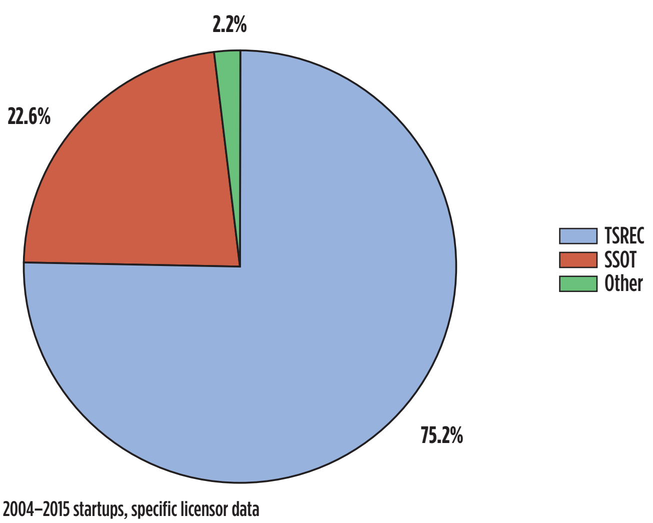


FIG. 1. Hydrocracker licenses by type on capacity basis.

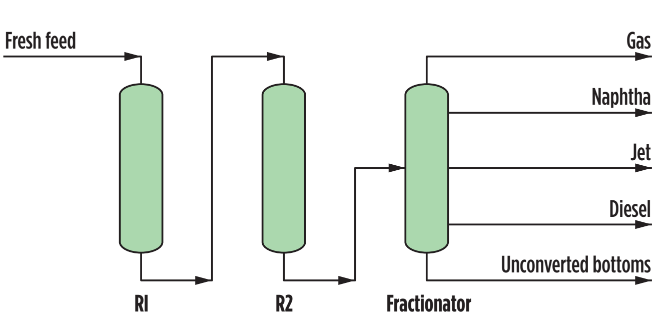


FIG. 2. SSOT process configuration.

critical to factor in and model the effect of this pre-processed 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 optimization are controlled by many constraints that must be accommodated in a single stage:

- Hydrodemetalization (HDM) needs driven by heavy VGO (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 requirements for the hydrocracking function in the lower catalyst system
- Hydroconversion targets to remove of heavy components
- HDS needs for products such as ULSD, marine GO and UCO
- Aromatics removal (lube or FCC applications)
- Isomerization (for lube needs).

SSOT catalyst system optimization is further challenged when the application involves a unit converted from a former service such as FCC pretreatment or, in less common cases, diesel 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 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 to meet HDN and HDS needs.

Lacking the flexibility to recycle and adjust the recycle cut point (RCP), product selectivity in the SSOT is controlled by the catalyst system choice. This choice relies mainly on hydrocracking catalyst component selection, along with setting the operating temperature regime and span.

Hydrocracking catalysts 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. **FIG. 3** shows the progression of such performance for the hydrocracking catalysts provided by one such supplier.

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 multi-catalyst combinations might instead point to a system of "A" and "C" catalysts and can actually achieve a better result than pure "B" alone.

TSREC configurations offers 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. The configuration is shown in **FIG. 4.**

TSREC units offer the refiner the ability to operate the two stages differently to simultaneously meet separate goals for each stage. This configuration also offers the flexibility to balance the stages to optimize 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 optimization of an additional catalytic component. In addition, feedstocks can be shifted within 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 optimization in the first stage is often influenced by feed quality and contaminants, with nitrogen, sulfur, metals, silicon and arsenic being the typical suspects. This is especially challenging for older, existing units but can be equally daunting even for new units.

Following feed contaminant removal, 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 stages. Depending upon unit objectives, the first stage cracking catalyst can be chosen from any of the "A," "B" or "C" groups.

Second-stage catalyst selection will largely be driven by the 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 high levels of conversion to the desired product, typically diesel or total distillate.

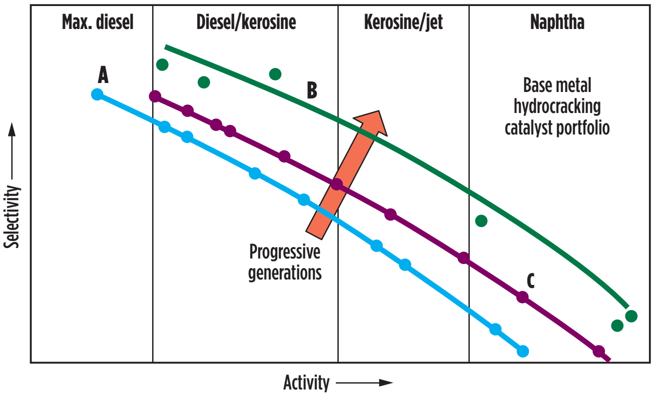


FIG. 3. Selectivity and activity improvements for hydrocracking catalysts by a hydroprocessing catalyst supplier.

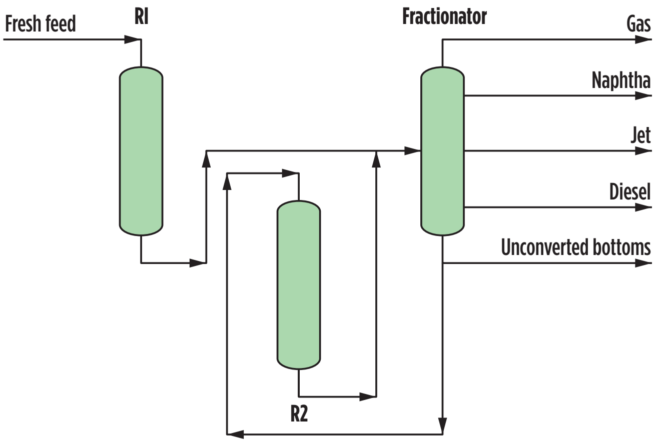


FIG. 4. TSREC process configuration.

Catalyst system optimization. ULSD units must also address molecular management. However, a properly designed catalyst system and optimized processing scheme can help the refiner maximize his profit goals. To help refiners deal with the severe demands of ULSD, a stage catalyst system in 2001. This system uses catalyst technology that is staged in the proper proportions to provide the best performance while also meeting individual refiner requirements. Catalyst staging is designed to take advantage of different reaction mechanisms for sulfur removal; a high activity cobalt and molybdenum (CoMo) catalyst efficiently removes the unhindered, easy sulfur via the direct abstraction route, and a high activity nickel and molybdenum (NiMo) catalyst then attacks the remaining sterically hindered, hard sulfur.

An important aspect for a staged catalyst system is designing the optimum proportions of the CoMo and NiMo catalysts that will deliver the best performance. This is dependent upon a number of factors, including the unit objectives, feed and operating constraints.[3,4]

A key advantage for this system is the efficient use of hydrogen. **FIG. 5** illustrates how the system can be tailored to provide the best balance of high HDS activity while minimizing H_2 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.

The figure also shows the relative H_2 consumption, and, as the percentage of the NiMo component increases, the H_2 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 all the CoMo system, and well below that for the NiMo catalyst. This is a direct result of the different kinetics for sulfur and aromatics removal, and it is a critical consideration when customizing the staged catalyst system.

For units that have a hydrogen constraint, the key to designing the proper catalyst system is increasing the hydrogenation selectivity to provide the highest HDS activity while minimizing hydrogen consumption. **FIG. 6** shows a rapid decrease

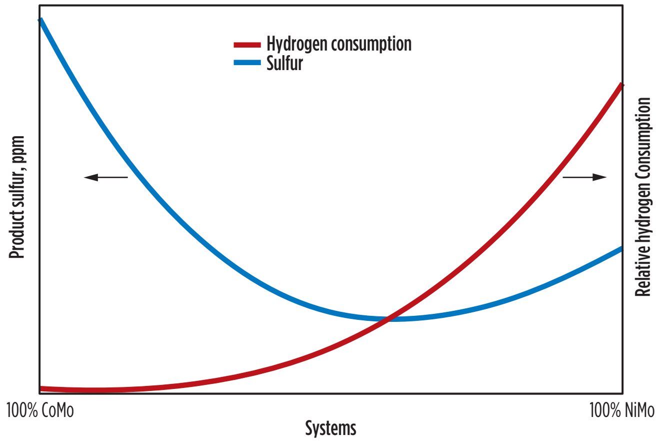


FIG. 5. Balancing high HDS activity while minimizing H_2 consumption.

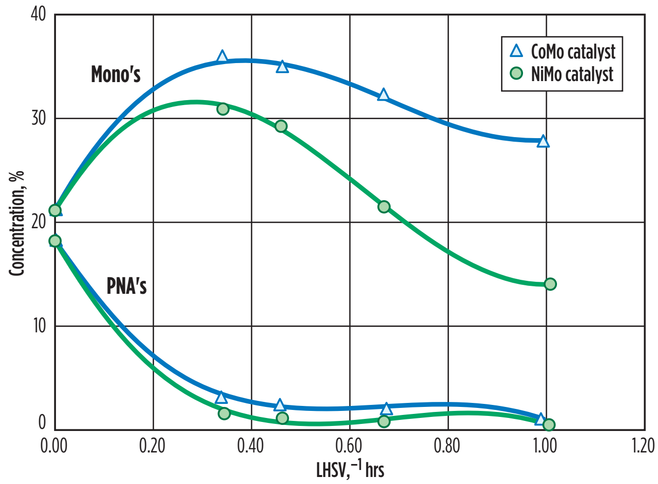


FIG. 6. Effects of increasing residence time.

in polyaromatics concentration and a corresponding increase in mono-ringed aromatics for both catalysts as the residence time is increased. 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 on the chart, the NiMo catalyst has about 15 absolute numbers with a lower mono-ringed aromatics concentration than the CoMo catalyst, and that corresponds to about 300 standard cubic ft per barrel in higher hydrogen consumption for the NiMo catalyst.

The data demonstrates that the system's hydrogenation activity can be tuned by adjusting the relative volumes in the CoMo and NiMo beds within the reactor. Of course, not all units have an H_2 constraint, and, in those cases, the incremental increase in aromatics saturation and the correspondingly higher hydrogen consumption obtained by the NiMo catalyst offers benefits such as cetane improvement and the ability to process more cracked stocks.

Experience 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 provide significant cetane uplift and volume swell.

In applications where there is sufficient H_2 availability and partial pressure, a NiMo catalyst is likely the most active system for HDS. However, it will consume significantly more hydrogen due to its efficiency at catalyzing hydrogenation reactions. If the incremental hydrogen consumption cannot be tolerated, a system can be designed that will deliver high HDS activity and minimize hydrogen consumption. In cases where the hydrogen pressure is lower, the staged catalyst system is often more active than either component alone. HP

LITERATURE CITED

1 IHS CERA, *Refining and Product Markets Annual Strategic Workbook*. 2013.
2 Torchia, D, Arora, A. and L. Vo, "Clean, green, hydrocracking machine," *Hydrocarbon Engineering*, June 2012.
3 Olsen, C. and L. D. Krenzke, "Custom Catalyst Systems for Meeting ULSD Regulations," 2005 NPRA annual meeting, Paper AM-05-17.
4 Olsen, C. and G. D'Angelo, "No Need to Trade ULSD Catalyst Performance for Hydrogen Limits," 2006 NPRA annual meeting, Paper AM-60-06.

R. G. VAZQUEZ, D. SEEGER and M. THUNDYIL,
GTC Technology US LLC, Houston, Texas

Improved oxidation process increases sulfur recovery with Claus unit

Since its development in 1936, the "modified Claus" technology has been the preferred process used by refineries and gas processing plants for the recovery of sulfur from acid-gas streams. Modern sulfur plants include a burner and a thermal reactor tied to a waste-heat exchanger as the first and second process units followed by a catalytic reaction section consisting of two or more catalytic converters. If sulfur recovery higher than 98% is required to meet the low-sulfur emission standards set by legislation, a tail-gas cleanup process must be included in the design of the sulfur recovery unit (SRU). The modified Claus process, however, has a number of operating challenges that are associated with the reactions and occur in the burner and thermal reactor. Some of these problems are: sustainable flame stability, operational reliability, minimum acid-gas concentration, strict control of the acid gas to air ratio, soot formation, catalyst fouling, and the formation of undesired byproducts such as carbonyl sulfide (COS) and carbon disulfide (CS_2).

New solution. A new process uses partial oxidation catalysis to improve the operation and reliability of the Claus unit.[a] The process lowers the capital investment in grassroots applications. The burner and thermal reactor are replaced with a catalytic combustor that uses a durable catalyst in a short-contact reactor to achieve near-equilibrium hydrogen sulfide (H_2S) conversion, high-sulfur selectivity and elimination of classic catalyst deactivation mechanisms due to sulfur poisoning and coke deposition on the catalyst.

Process description.

A simplified process flow diagram for the partial oxidation catalysis process is shown in **FIG. 1.** The figure shows a catalytic combustor in place of the conventional Claus burner and thermal reactor.[b] The acid gas and air streams are preheated to approximately 220°C before they are mixed and sent to the specially designed chamber to thoroughly mix the two gases before they enter the catalyst reactor bed.

As the gas enters the reactor, it comes into contact with the front face of the catalyst, passes through the catalyst bed in less than 0.1 seconds, and immediately enters the waste-heat exchanger or waste-heat boiler (WHB). In the catalytic reactor, over 80% of the H_2S present in the acid gas is converted to sulfur and sulfur dioxide (SO_2); most of the hydrocarbons are converted to hydrogen (H_2), carbon monoxide (CO) and water (H_2O); and the ammonia (NH_3) is converted to nitrogen (N_2) and H_2, or N_2 and H_2O, depending on the reactor temperature.

The gas exiting the WHB is sent directly into the first Claus converter, thus eliminating a sulfur condenser and a re-heater and achieving over 90% sulfur conversion after the first Claus converter. This technology can be applied to a conventional modified Claus configuration, and may be of particular interest in grassroots designs.

The gas path after the first Claus converter is the same for the partial oxidation catalytic process as that of a conventional Claus SRU.[a] The equipment following the WHB consists of conventional Claus catalytic reactors, sulfur condensers and re-heaters, followed by a tail-gas unit (TGU). However, since the COS and CS_2 formation is dramatically reduced due to the short contact reactor, the catalyst requirements for COS and CS_2 hydrolysis are, likewise, reduced.

The new process is designed in accordance with the principles of the classic Claus reaction.[a] In the catalytic combustor section, under sub-stoichiometric conditions, over 80% of the H_2S in the feed stream is converted to sulfur and to SO_2 and H_2O. If the sour gas being treated contains NH_3, and if the reaction temperature in the catalytic combustor section reaches 1,400°C, then the NH_3 will be destroyed through thermal dissociation to form N_2 and H_2. If the temperature does not reach 1,400°C, the NH_3 will be oxidized to form N_2 and H_2O at temperatures of about 1,200°C. Light hydrocarbons will be oxidized to form mainly CO and H_2 and some CO_2 and H_2O. Due to the short-contact reactor, only minimum amounts of COS and CS_2 will be formed in the reactor. Benzene, toluene and xylene are completely destroyed at temperatures above

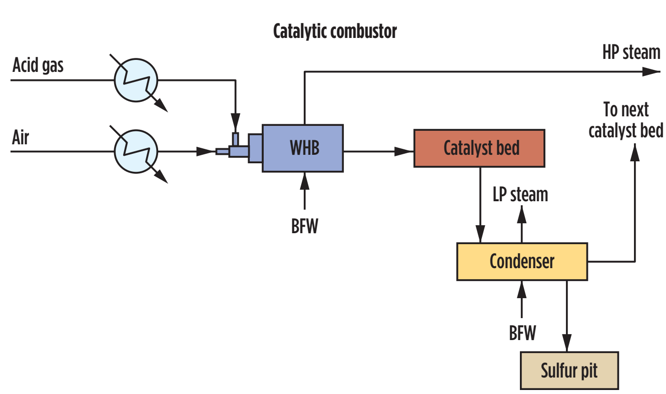


FIG. 1. Simplified flow diagram of the partial oxidation process integrated with the first Claus converter and first sulfur condenser.

900°C. Some of the reactions that occur in the catalytic combustor section are:

$$H_2S + 1.5O_2 \rightarrow SO_2 + H_2O$$
$$2H_2S + SO_2 \rightarrow 1.5S_2 + 2H_2O$$
$$2NH_3 + \text{Heat} \rightarrow N_2 + 3H_2 \text{ (thermal dissociation) or}$$
$$NH_3 + 0.75\ O_2 \rightarrow 0.5\ N_2 + 1.5\ H_2O \text{ (oxidation)}$$
$$H_2S \rightarrow H_2 + 0.5\ S_2$$
$$C_nH_{2n+2} + ((2n+1)/2)\ O_2 \rightarrow (n+1)\ H_2O + n\ CO$$
$$C_nH_{2n} + n\ O_2 \rightarrow n\ H_2O + n\ CO$$

In **FIG. 2**, a catalytic combustor used in the new process is compared to the conventional Claus burner and thermal reactor units.[a] Both the conventional burner and the thermal reactor and catalytic combustor were designed using the same acid-gas composition and operating conditions. **TABLE 1** summarizes the design basis.

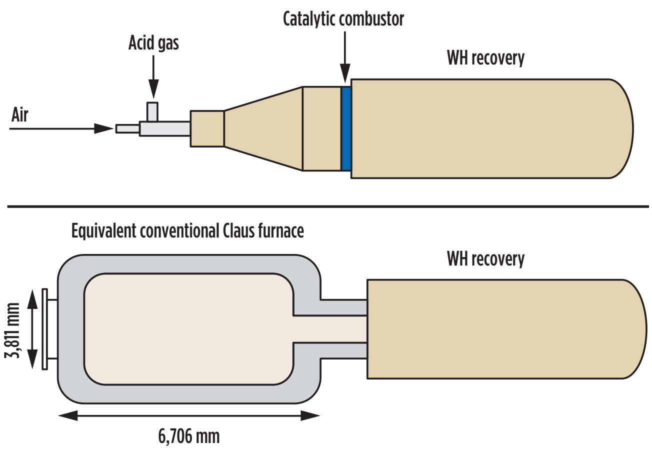


FIG. 2. A partial catalytic combustor (top) compared to a conventional Claus thermal reactor designed for the same application, showing the relative size difference. **Note:** The Claus furnace schematic does not illustrate the burner, which further accentuates the size difference.

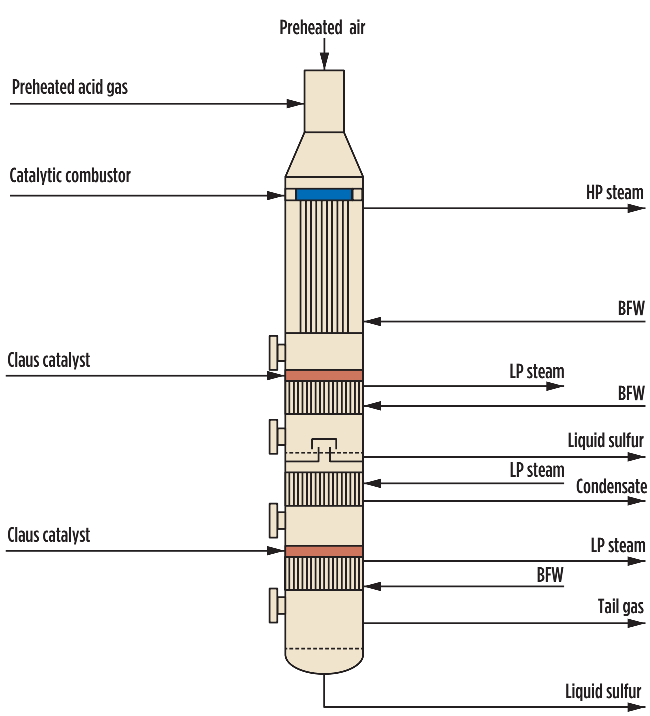


FIG. 3. Ultimate vertical design of new partial oxidation process with two Claus converters in series.

As shown in **FIG. 2,** the two units are drawn to scale, which shows the relatively larger size requirement for the burner and thermal reactor over the catalytic combustor. This size difference saves significantly on the footprint and amount of refractory material required. The refractory volume needed in the catalytic combustor is 1/40th that of the refractory volume required for the thermal reactor.

In some cases, the new partial oxidation process can be offered in a single vertical arrangement that also reduces interconnecting piping, sulfur rundown piping and total plant footprint; all further reducing capital cost.

In a vertical arrangement, the new catalytic combustor, integrated WHB and the catalytic converters and condensers are all

TABLE 1. Basis for the design of a catalytic combustor and a conventional Claus burner with a thermal reactor

Stream	Acid gas from amine unit	Sour-water stripper gas	Combustion air
Mass flow, kg/h	4,050	730	10,900
Flowrate, Nm^3/h	2,671	675	8,500
Composition, kg-moles/h			
C_1	0.48	0	0
C_2	0.72	0	0
C_3	1.19	0	0
C_4	0.60	0	0
CO_2	1.19	0	0
H_2O	2.86	6.00	5.4
H_2S	112.21	12.00	0
Hexane	0	0.04	0
Phenol	0	0.02	0
N_2	0	0	295.6
O_2	0	0	78.5
NH_3	0	12	0

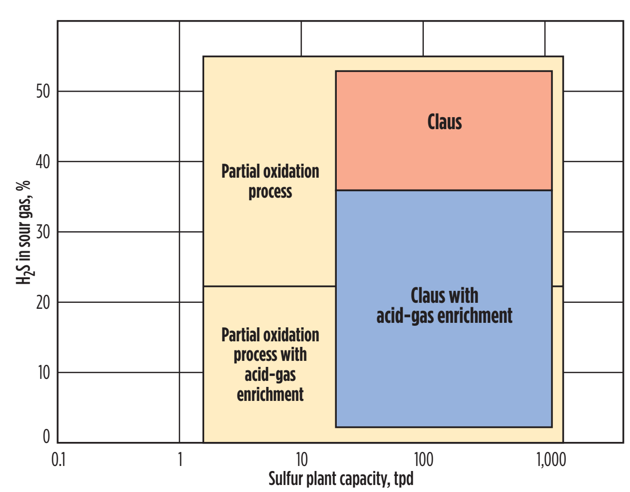


FIG. 4. Technology envelope for partial oxidation process and Claus.

combined in a single vertical tower.[a] **FIG. 3** is a simplified diagram of the vertical arrangement with two Claus converter stages.[a]

Sulfur-reduction optimization. The new partial oxidation process can provide several advantages and benefits over the modified Claus process, including: improved reaction efficiency, improved turn-down flexibility, reduced COS and CS_2 formation, lower soot formation, near-seamless switching between acid gas/fuel gas, and no oxygen breakthrough during startup or shutdown activities.[a] **TABLES 2** and **3** summarize the many process and economic advantages with a short-contact catalytic combustor vs. a conventional Claus burner and thermal reactor.

Applications. The key advantages of the new technology are that it dramatically reduces the capital cost of a "modified Claus" plant, and that it can handle lean sour-gas streams with ease. These advantages, either singly or together, expand the applicable envelope of "modified Claus" technology.

Example. The partial oxidation process has an application in a small refinery or a refinery processing sweet crudes with a total sulfur capacity of 2 tpd–20 tpd. In this situation, unlike iron redox technology, the new process can destroy NH_3, and thereby help refiners meet both sulfur oxide (SO_x) and nitrogen oxide (NO_x) targets with ease. Similarly, associated sour gas streams, with a total sulfur content of 2 tpd–20 tpd, can now be processed, allowing oil and gas producers to produce oil and sell pipeline-quality gas while meeting sulfur-emission targets. **FIG. 4** illustrates the new sulfur-removal technology envelope.

Opportunities. Modified Claus plants have a number of operating problems associated with the reactions that occur in the burner and thermal reactor units. A new technology can offer an alternative to eliminate most of the issues associated with flame stability, byproduct formation, refractory maintenance, etc. The new process, with a catalytic combustor at the front end of a modified Claus process, results in significant process and economic benefits and advantages including improved sulfur yield, reductions in COS and CS_2 formation, reduced impact of hydrocarbons on air demand, improved startup and shutdown operations, and reductions in the size of the Claus converter catalyst-bed sections.[a]

In addition, the new catalytic combustor allows upgrades of aging Claus burners and thermal reactors, and permits a 20%–30% reduction in capital costs associated with Claus plants, while reducing the TGU loading due to its higher conversion efficiencies. HP

TABLE 2. Operating conditions for new partial oxidation process and conventional modified Claus units

Item	Conventional modified Claus	New partial oxidation process[a]
Minimum acid gas concentration, %H_2S	50	25
Furnace temperature, °C	Base	150 to 200 higher
Reaction furnace residence time, sec	1–2	0.1
Air demand issue: Fuel gas to acid gas	Yes	No
Startup and shutdown	Time consuming	Fast
NH_3 destruction	Yes–2 chambers	Yes
COS, CS_2 formation	Yes	Traces only
First condenser/re-heater	Necessary	Can be eliminated
Claus converter beds	Base	50%–70% smaller
Heavy hydrocarbon destruction	Challenging	Yes (makes H_2/CO)
Plot space	1	0.50–0.70
Capital cost	1	0.70–0.80
Operating cost	1	0.95
Can conventional Claus units be upgraded?	–	Yes

NOTES

[a] GTC Technology now offers its unique GT-SPOC technology. GT-SPOC (sulfur partial oxidation catalysis) process, the burner and thermal reactor are replaced with a catalytic combustor.

[b] A catalytic combustor, GT-CataFlame, is in place of the conventional Claus burner and thermal reactor.

TABLE 3. Processing benefits with new partial oxidation unit[a]

Process advantages
Reactants are pre-mixed prior to passing through the catalytic combustor, creating a uniform flow through the cross-section of the reactor; thus eliminating the problems of post-combustion mixing for contaminant destruction
Lean H_2S gases (< 30 wt%) can be directly fed to the catalytic combustor
Close-coupling of the catalytic combustor and WHB improves the total sulfur yield and reduces air requirements due to rapid reaction quenching
No flames, fire-eyes nor burner management systems. A small retractable burner or access to a hydrogen-containing fuel is all that is needed to trigger the catalyst at startup
Fuel-gas oxidation for warm-up takes place at 25% air stoichiometry, eliminating oxygen breakthrough to downstream Claus converter beds during startup and shutdown
Air demand is identical for natural gas and H_2S; consequently, co-firing with natural gas is easier to control, and does not produce soot
Low-molecular-weight hydrocarbons are converted to reduction TGU friendly H_2 and CO by catalytic partial oxidation
Soot formation is virtually eliminated. Air/fuel ratios for fuel-gas oxidation closely resemble acid gas/air ratios making, switching nearly seamless
COS and CS_2 formation is almost completely eliminated
Sulfur product contains only 25%–33% of the dissolved H_2S/H_2S_x of normal Claus sulfur
Economic advantages
Significant reduction in refractory lining due to the smaller size of the catalytic combustor.
Small catalytic combustor volume and reduction of byproducts allow for design changes; all reducing the overall unit footprint
Design changes eliminate interconnecting piping and sulfur rundown piping and equipment
Less refractory and catalyst mass to heat up or cool down; this in turn reduces the time required for shutdown or startup
A new plant can be constructed in a vertical orientation where the unit is self-draining[a]
The first sulfur condenser and re-heater downstream of the WHB in conventional Claus designs are also eliminated, so that the gas stream exiting the WHB flows directly to the first catalytic converter

V. V. GALKIN, Gazprom Neft, Moscow, Russia;
V. MAKHIYANOV, Grubkin Russian State University of Gas and Oil, Moscow, Russia; and **M. I. LEVINBUK,** Topchiev Institute of Petrochemical Synthesis, Moscow, Russia

Update: Russia's capacity and modernization program

Since 2008, Russian regulations regarding the crude oil and refining industry have changed substantially. At present, Russia has 22 large refineries, each with an individual throughput capacity exceeding 4 million tpy (MMtpy); 8 medium-sized refineries with capacities ranging from 1 MMtpy–4 MMtpy; and over 200 small refineries. The government now requires that all Russian refineries be modernized or shut down. Such actions will involve technical and economic optimization parameters, especially for determining how to upgrade the medium- and small-sized refineries.

The new regulations will produce changes on refinery performance, capacity and processing profile (maximum output of gasoline, middle distillates or petrochemical feedstock). The maximum output of middle distillates is the most efficient refining model. Under new economic conditions, after the export duty for heavy petroleum products equal to 100% of the crude export duty will be introduced in 2015, refineries with the capacity under 1 MMtpy will be operating at a loss. Medium-sized refineries (capacity of less 1 MMtpy–4 MMtpy) updates cannot be carried out in several steps as profitability for these refineries will cover modernization costs. The presented approach will allow upgrading of small- and medium-sized Russian refineries to meet new government specifications in a cost-effective manner.

Changes in the Russian law. On Feb. 27, 2008, Resolution of the Government of the Russian Federation No. 118 enacted the Technical Regulations, "On Requirements for Motor and Aviation Gasoline, Diesel and Marine Fuel, Jet Fuel, and Fuel Oil" (the regulations). It resulted in a number of changes in the domestic oil industry. New fuel quality specifications and timelines for the conversion to cleaner transportation fuels caused Russian oil companies to reconsider refinery development programs and begin upgrades to existing refineries.

New export duties for heavy petroleum products (66% of the crude export duty in effect from October 2011 and 100% beginning in 2015) was another step taken by the Russian government to encourage refinery modernization.[1] Changes were also made to the excise rates for engine fuels: rates for low-grade fuels were increased, while rates for higher-grade fuels were decreased.[2] In addition, certain restrictions were imposed on connection of refineries in the process of design, construction or modification to main petroleum product pipelines and crude oil pipelines. Going forward in 2015, oil conversion ratio at refineries will be no less than 70%; otherwise, connection to main pipelines will not be possible.[3]

According to the Russian Ministry of Energy, these changes should promote investments by vertically integrated oil companies in developing secondary refining processes. The Ministry expects the investments in the refining industry modernization to reach 1 trillion rubles—\$34 billion (B)—from 2011 to 2015. At the same time, changes in the law will result in economic unviability of unsophisticated, export-oriented refineries and mini-refineries. Insufficient capital-output ratio of the mini-refineries will be forced to shut down over the medium term.

At present, Russia has 30 refineries in operation—22 large refineries (capacity over 4 MMtpy) and 8 medium refineries (capacity of 1 MMtpy–4 MMtpy) according to the Russian Ministry of Energy's register.[5] In addition, over 200 small (mini) refineries (capacity of < 1 MMtpy) are still in operation.[6] All refineries are in the process of modernizing their operations. There are several questions arising over the fuel quality and upgrading mandates:

- To what extent will the changes in the law affect site profitability, especially for small- and medium-sized refineries?
- What is the minimum capacity to ensure profitability and return on investment (ROI), while meeting tighter fuel quality requirements?
- What is the minimum investment necessary to upgrade small- and medium-sized refineries?

This article will investigate the standard refining models and approaches to address the Russian refining industry issues.

Basic data and limitations. Three main oil refining models were used as the base for the analysis:

Case 1. Gasoline based on catalytic cracking

Case 2. Diesel based on vacuum gasoil (VGO) hydrocracking

Case 3. Petrochemical for maximum production of petrochemical feedstock, based on fluid catalytic cracking (FCC) with maximum yield of C_3–C_4 olefins.

Estimates also comprise the review of sub-options to enhance the models' efficiency. Estimates for the three cases were applied for refining capacities from 1 MMtpy to 10 MMtpy, in increments of 1 MMton. This approach clearly shows the trends in economic efficiency for small, medium and large refineries. TABLE 1 summarizes the main petroleum product yields with the selected models.

The process model has a minimum set of processes to minimize project investments and produce Euro 5 transportation fuels. These models used several processes refining with the same feedstock. Processes, such as FCC and VGO hydrocracking are not covered in the models; small- and medium-sized refineries lack the investment assets for new FCC and hydrocracking units. The selected models are "reference" models; they may be used in the future to develop the optimum refining models for large refineries.

The models are analyzed in two stages. The first stage described here does not cover the variability of vacuum residue (VR) processing. All of the models include visbreakers and bitumen plants. The ratio of the capacities is based on the average data for Russian Federation refineries. The process models for the reference refineries use standard refining processes commonly installed at the existing refineries or are included in the long-term modification plans through 2020.[7]

Motor gasoline includes only AI-92 and AI-95 grades. In all options, methyl-tertiary butyl ether (MTBE) is purchased from third parties when needed to meet Euro 5 requirements. The maximum kerosine output is considered. However, for simplicity, the model does not include winter diesel fuel, which can be produced by mixing kerosine with summer diesel, if necessary. Crude quality is based on Urals oil, with sulfur and light petroleum product content of 1.4% and 47%, respectively. The export duties are per the law of the Russian Federation now in force. The crude price is set at $90/bbl.

The investment costs for plant construction are calculated using available data for construction of similar plants, as well as data received from licensing and consulting companies.[a] Investments are recalculated for different plant capacity as:[8]

$$I = I_a \times (C / C_a)^{0.6}$$

where I is the calculated investments for a plant with capacity, C; I_a and C_a are known investments and capacity of a plant taken as analogue, respectively.

Construction costs for auxiliary and power facilities, offsite facilities (OSF) and construction management expenses are taken as 70% of the total cost of main processing plants. This cost level corresponds to the construction of a grassroots refinery. If the refinery has a well-developed infrastructure, this ratio may be decreased. However, given the process models of small- and medium-sized Russian refineries, the modifications will be comparable to building a refinery. So, the 70% ratio may be used at this level.

Operating costs are calculated on the basis of actual plant performance at OJSC Gazprom Neft refineries and reference data.[9] To minimize operating and investment costs, assume refineries are constructed as complexes rather than individual processing units. The main assessment criteria are:

- **Technical parameters include:**
  - o Oil conversion ratio
  - o Yield of light petroleum products.
- **Economical parameters include:**
  - o Breakeven point—the capacity in which the refining margin (sales revenue less variable costs) will be higher than the level of semi-fixed costs. (The capacity selected by the refinery model must yield a profit.) **Note:** If the upgrading investments are high, then depreciation will also account for a larger share of the fixed costs.
- **Net present value** (NPV) is the ROI for the modification project. If the NPV on a 15-year planning horizon is more than zero, then the investments are paid back.[b]

All process models are made and optimized in MS Excel spreadsheets.

REFINING PROCESS MODEL ANALYSIS

The processing analysis investigated three options: gasoline only, diesel only and petrochemical integration:

Gasoline option. The process flow diagram for the gasoline option with minimum investments is shown in **FIG. 1.** This model allows for the oil conversion ratio at a level of 76% and the 67% yield of light petroleum products. The benzene and total aromatics content in the gasoline pool is a major problem. These

TABLE 1. Yields of main petroleum products for different refining models

	Gasoline		Diesel		Petrochemical	
Product	**Basic option**	**Sub-option 1A**	**Basic option**	**Sub-option 2A**	**Basic option**	**Sub-option 3A**
Liquefied petroleum gas (LPG)	4.9	3.3	2.4	2.4	9	9.9
Pyrolysis feedstock	–	–	4.7	7.9	2.3	–
Motor gasoline:						
AI-95	28.7	26.7	–	–	10.1	10.3
AI-92	3	2.1	–	–	–	–
Aromatic hydrocarbons	–	–	–	7.4	3.8	4
Reformate	–	–	10.5	–	–	–
MTBE, TAME and alkylate	–	4.1	–	–	6.3	7
Kerosine	8.8	8.8	27.4	27.4	12.1	12.1
Diesel fuel	27.7	27.7	29.8	29.8	27.5	27.5
Fuel oil	18.2	18.2	18	18	17.6	17.6
Bitumens	3.5	3.5	3.5	3.5	3.5	3.5
Sulfur	0.8	0.8	0.8	0.8	0.8	0.8
Fuel and losses	4.4	4.8	2.9	2.8	7	7.3
TOTAL	**100**	**100**	**100**	**100**	**100**	**100**

parameters can be achieved blending only MTBE or alkylate. Both MTBE and alkylate must be purchased or produced additionally from the refinery's own butylene. (In the gasoline option, purchased MTBE values were applied, as it is easier to get MTBE from the market.)

A sub-option was developed, which includes sulfuric acid alkylation, MTBE and tertiary-amyl methyl ether (TAME) units (option 1A). With these process units, the benzene and aromatics content in the gasoline meets Euro 5 emission requirements, without purchasing MTBE or alkylate from third parties. However, this approach will increase investments by 11%. Key performance indicators for the gasoline option and the additional sub-option are shown in **TABLE 2** and **FIG. 2.** Investments in refinery modification (gasoline option) are estimated at $ 4.3 B for a refinery with 10 MMtpy of capacity.

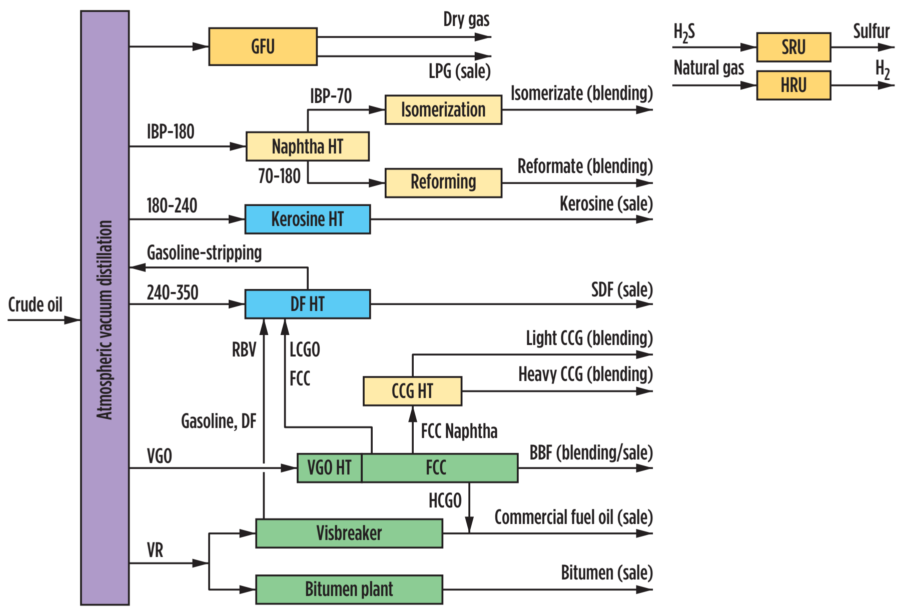


FIG. 1. Process flow diagram—gasoline option.

TABLE 2. Key performance indicators, gasoline option

Indicator	Basic option	Sub-option 1A
Conversion ratio, %	76	76
Yield of light petroleum products, %	67	71
Breakeven point, thousand tpy, at values:		
2012	710	710
2015	990	980

The breakeven point for the gasoline option is 0.71 MMtpy at 2012 values and 0.99 MMtpy at 2015 values. After the new duties for heavy petroleum products are introduced, the ROI can be achieved only if the refinery capacity exceeds 10.2 MMtpy. However, oil refining options with 6.8 MMtpy and more are paid back at present duty levels.

Diesel option. The process flow diagram for the diesel option with minimum investments is shown in **FIG. 3.** The diesel option allows for the oil conversion ratio at a level of 79% and 72% yield of light petroleum products. These parameters are higher than those for the gasoline option due to less unconverted residue from the VGO hydrocracker compared with FCC. The model is based on the VGO hydrocracker with maximum output of middle distillates and 95% conversion. The diesel option selected for the analysis (provided that gasoline is produced by the hydrocracker) does not allow for Euro 5 gasoline due to aromatic hydrocarbons, benzene content and saturated vapor pressure from components produced by the refinery.

Since a large amount of components with low aromatic hydrocarbons (alkylate, TAME, MTBE, etc.) must be purchased to bring the gasoline pool into Euro 5 compliance, it was proposed to rule out motor gasoline production under this option. In this case, the IBP-62°C fraction is sold as pyrolysis feedstock while the 62°C–140°C fraction is directed to the reformer for processing, and the reformate is sold as feedstock for aromatics production. This approach improves the refining efficiency by reducing straight-run (SR) gasoline production, reducing the hydrocracker capacity and lowering operating costs for hydrogen production. As reformate is not

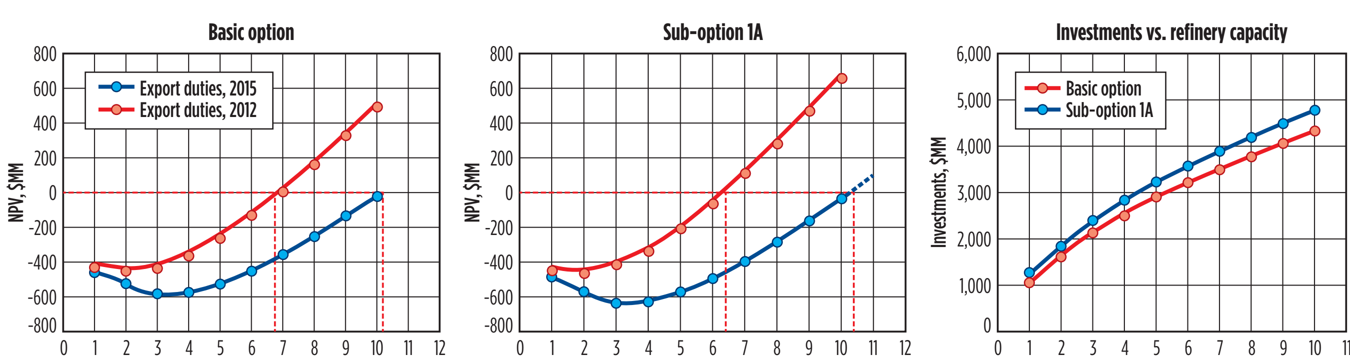


FIG. 2. NPV and investments—gasoline option.

Cameron's Centrifugal Compression.

Complete Solutions for Process Gas Applications.

Improved reliability and efficiency. Incredible peace of mind. From petrochemical and refining to fuel gas boosting and refrigeration, Cameron offers a wide range of centrifugal gas compressors that ensure peace of mind by providing solutions that keep you running smoothly, efficiently and profitably. In a variety of applications, our MSG® Series offers outstanding design flexibility for your exact needs. With installations in more than 90 countries around the world, a worldwide distribution and service network, and multiple packaging options, Cameron is sure to exceed your expectations with unequaled service and support. Add in CAMSERV™, the world's largest aftermarket network, and you can trust Cameron as your total solutions provider.

Learn more by calling 1 877 805 7911 or visiting www.c-a-m.com/cs.

Select 67 at www.HydrocarbonProcessing.com/RS

AD00304CC

RAISING PERFORMANCE. TOGETHER™

CAMERON

TABLE 3. Key performance indicators, diesel option

Indicator	Basic option	Sub-option 2A
Conversion ratio, %	79	79
Yield of light petroleum products, %	72	72
Breakeven point, thousand tpy, at values:		
2012	630	610
2015	870	810

TABLE 4. Key performance indicators, petrochemical option

Indicator	Basic option	Sub-option 3A
Conversion ratio, %	75	74
Yield of light petroleum products, %	65	64
Breakeven point, thousand tpy, at values:		
2012	680	630
2015	890	810

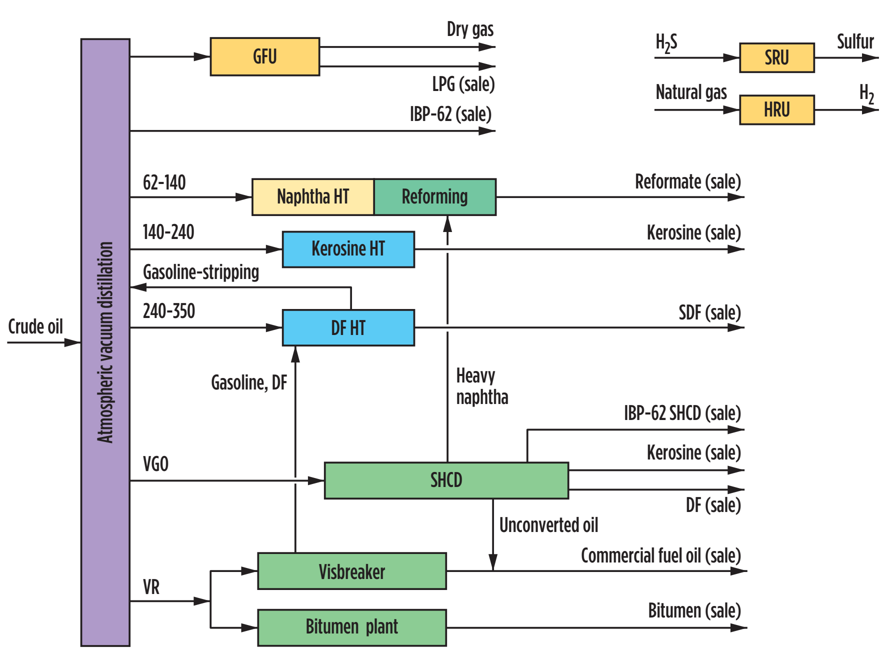


FIG. 3. Process flow diagram—diesel option.

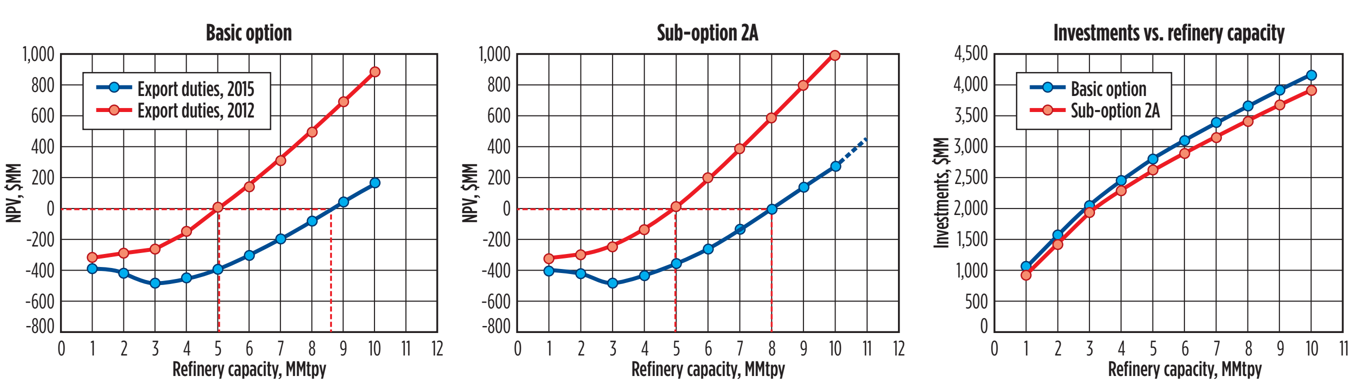


FIG. 4. NPV and investments—diesel option.

commonly used as a market product, an additional option was developed to produce aromatic hydrocarbons—benzene, toluene and xylene (BTX) as sub-option 2A. In this case, investments in refinery modification increased 7% (from the basic diesel option). Key performance indicators for the diesel option and sub-option 2A are shown in **TABLE 3** and **FIG. 4.** Investments in refinery modification (diesel option) are estimated at $3.9 B for a refinery with 10 MMtpy of capacity.

The breakeven point for the diesel option is 0.63 MMtpy at 2012 values and 0.87 MMtpy at 2015 values. After the new duties for heavy petroleum products are introduced, the ROI can be achieved only if the refinery capacity exceeds 8.8 MMtpy. However, oil refining options with 5 MMtpy and more are paid back at present duty levels.

Petrochemical option. The process flow diagram for the petrochemical option with minimum investments is shown in **FIG. 5.** This model is based on catalytic cracking with deep conversion and high yield of light olefins C_3–C_4, as well as on the reforming of the 62°C–140°C fraction to produce aromatic hydrocarbons. This model allows for the oil conversion ratio at a level of 75% and the yield of light petroleum products of 65%. The major problems for this refining model include low density of the gasoline pool (due to recovery of aromatic hydrocarbons) and low-octane gasoline/raffinate from aromatics extraction. The problem with density can be solved by adding toluene to the gasoline blend; the problem with raffinate was rectified by selling it as a petroleum solvent.

As the market for petroleum solvent is limited, a sub-option was also developed to direct raffinate to the catalytic cracker (sub-option 3A). Process licensors confirm that it is possible to direct raffinate for processing to deep FCC (DFCC) conversion.[10,11] Raffinate processing at DFCC will decrease the yield of light hydrocarbons and oil conversion ratio, but, also, it will remove the low-octane components from the gasoline pool and thus increases AI-95 gasoline production and the refinery's profit. Key performance indicators for the petrochemical option and the additional sub-option are shown in **TABLE 4** and **FIG. 6.** Investments in refinery modification (petrochemical option) are estimated at $5 B for a refinery with 10 MMtpy of capacity.

The breakeven point for the gasoline option is 0.68 MMtpy at 2012 values and 0.89 MMtpy at 2015 values, for sub-option 3A–0.63 MMtpy and 0.81 MMtpy, respectively. After the new duties for heavy petroleum products are introduced, the ROI can be achieved if the refinery capacity exceeds 9.3 MMtpy (8.2 MMtpy for sub-option 3A). However, oil refining options with 6 MMtpy and more (5.3 MMtpy and more for sub-option 3A) are paid back at present duty levels.

FUTURE OPTIONS FOR RUSSIAN REFINERS

The analysis shows that the model with maximum output of middle distillates is the most efficient model under the present Russian law. This option has minimum investments ($3.9 B –$4.1 B for a refinery with 10 MMtpy of capacity), a minimum breakeven point at a level of 600,000 tpy–650,000 tpy, and a minimum oil refining level at which refinery modification will pay back after the 100% duty for heavy petroleum products is introduced (8.2 MMtpy–8.8 MMtpy), excluding investments in heavy residue processing.

With regard to ROI, the petrochemical option with maximum yield of light olefins C_3–C_4 is comparable to the diesel option. But the investments are the largest among the options under discussion. Refinery modification under the gasoline option requires investments at $4.3 B. However, ROI can be achieved only for large refineries with the capacity exceeding 10 MMtpy.

The estimates show that the era of mini-refineries, with capacity less than 500,000 tpy is ending. The mini-refineries will shut down, as they are unprofitable and cannot undergo cost-effective modifications to meet legal requirements for product quality without substantially increasing throughput capacity. It is impossible to modify all of the small refineries and increase processing capacity to the minimum efficient capacity. Such an action would effectively double the refining throughput in Russia against the existing level. Such capacity is over the present domestic market demand and available crude oil supplies. Only those mini-refineries that can be upgrades are able to:

- Increase their capacity to 1 MMtpy, achieve 70% oil conversion ratio and find a market for produced petroleum products
- Find ways to purchase much cheaper equipment, given that present prices are peaking due to a greater market demand for modifications for other Russian refineries
- Consistently (on a phased basis) introduce new processes (processing units) within a reasonable time before the new customs duties for fuel go into effect.

As the total refining throughput of mini-refineries is approximately 11 MMtpy, large oil producers will be able to cover the market gap with petroleum products from present refineries on a competitive basis.

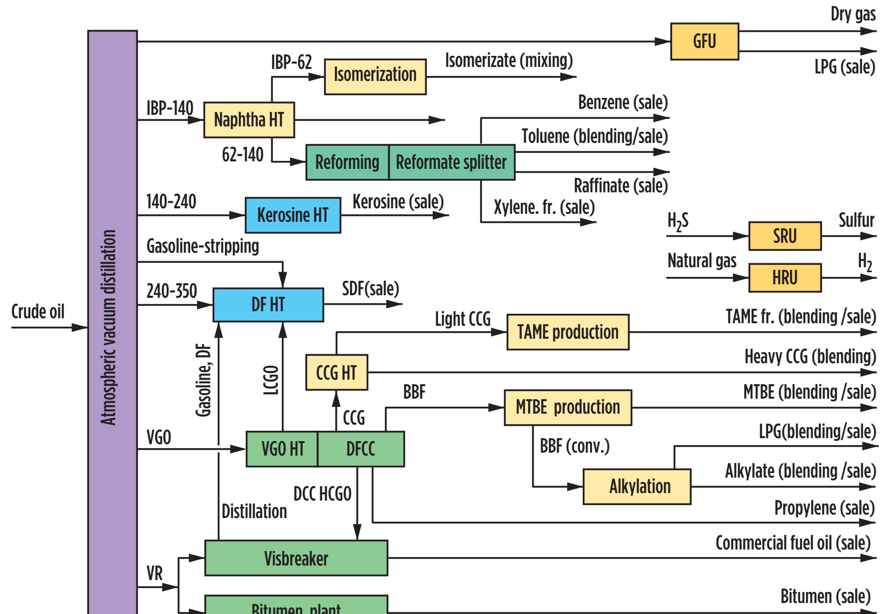


FIG. 5. Process flow diagram—petrochemical option.

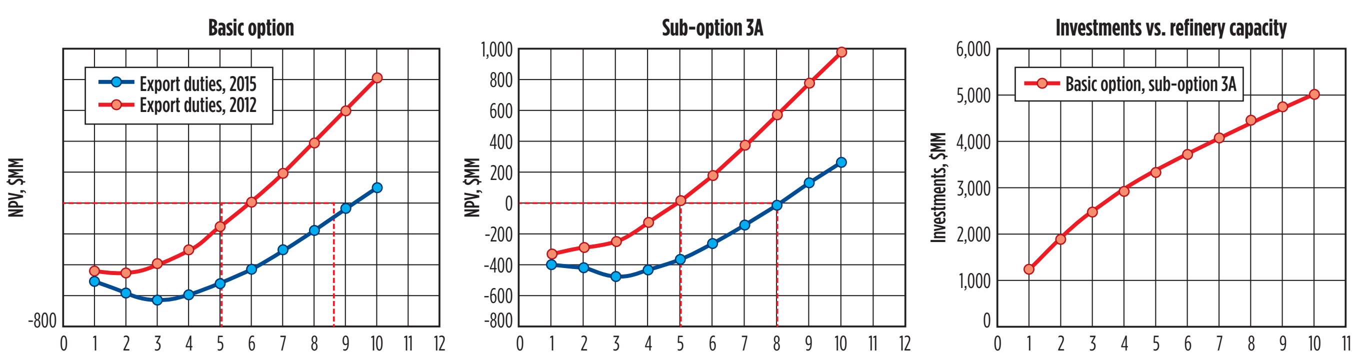


FIG. 6. NPV and investments—petrochemical option.

Medium-capacity refineries (1 MMtpy –4 MMtpy) are a more complicated issue. Their share in the total refining throughput in Russia is about 10%–15%. If the medium-sized refineries shut down, this may strongly affect the Russian market for petroleum products, especially in the regions where such refineries are located. According to the estimates, after the 100% duty for heavy petroleum products is introduced in 2015, only refineries with capacity exceeding 8 MMtpy will be able to get ROI for their modification if they are done in a conventional way. Construction of heavy-oil residue processing complexes may have a beneficial effect on ROI in modifying medium-sized refineries.

Conversely, it will mean that step-by-step modifications will be impossible because medium-sized refineries will not reach a sufficient profit level to start operation unless they construct deep-conversion facilities. The existing offsite facilities of a refinery will reduce the investment required for modification. Nevertheless, for most medium-sized refineries with a low level of secondary processes, such investments will be close to grassroots-refinery construction costs. It appears that the Russian government should reconsider the issues related to modification and further operation of medium-sized refineries. HP

NOMENCLATURE

HT	Hydrotreater
CCG	Catalytically cracked gasoline
DF	Diesel fuel
DFCC	Deep fluid catalytic cracking
FCC	Fluid catalytic cracking
HCGO	Heavy catalytic gasoil
HPU	Hydrogen production unit
HRU	Hydrogen recovery unit
LCGO	Light catalytic gasoil
SDF	Summer diesel fuel
SRU	Sulfur recovery unit
LCGO	Light catalytic gasoil
VR	Vacuum residue
VGO	Vacuum gasoil

NOTES

[a] Information attained from KBC and CLG.

[b] All process models are made and optimized in MS Excel spreadsheets. Validity of the models is verified using Honeywell's RPMS.

LITERATURE CITED

1 Resolution of the Government of the Russian Federation No. 716, "On Amending Resolution of the Government of the Russian Federation," Aug. 26, 2011; No. 1155 "On Calculating Rates of Export Customs Duties for Certain Categories of Goods Produced from Oil," Dec. 27, 2010.

2 Clause 1, Article 193 of the Tax Code of the Russian Federation as approved by Federal Law No. 338-FZ, "On Amending Part Two of the Tax Code of the Russian Federation and Certain Legislative Acts of the Russian Federation," Nov. 28, 2011.

3 Clause 6, sub-paragraph "zh" of the Rules for Connection of Refineries to Main Oil Pipelines and/or Petroleum Product Pipelines and Refinery Metering in the Russian Federation as amended by Resolutions of the Government of the Russian Federation, No. 864, Oct. 28, 2012, and No. 741, July 07, 2012.

4 Press Service of the Ministry of Energy of Russia, "The Russian Ministry of Energy held a meeting on implementation of quadruple refinery modernization agreements by oil companies in 2012,"/http://minenergo.gov.ru/press/min_news/12587.html, Aug. 9, 2012.

5 Register of Refineries Currently in the Process of Design or Construction or Those Put into Operation in the Russian Federation / Ministry of Energy of the Russian Federation / http://minenergo.gov.ru/activity/oilgas/reestr_npz/.

6 Rostekhnadzor Inspections Led to Shutdown of 80 Mini-Refineries / RBC daily, Nov. 25, 2010,/ http://www.rbcdaily.ru/tek/562949979211459.

7 OJSC VNIPIneft: Report at the 12th Russia and CIS Refining Technology Conference RRTC-2012, Sept. 20-21,2012, Moscow.

Complete literature cited and author biographies available online at Hydrocarbon Processing.com.


Overpressure risk never goes away.
That's why we offer total pressure relief solutions.
Farris Total Pressure Relief Solutions Include:
Pressure Relief Valves
PSM Engineering/Audit Service & Software
Asset Management Services & Repairs
The Total Solution.
It's simple. Safety relief valves and services to support your facility's entire lifecycle. Pressure relief systems designed right from the start. Pressure relief valves that help your plant operate safely. Global valve maintenance and repair support. And PSM software and audit services to ensure your plant operates the way it was designed to during an overpressure event.
Total Pressure Relief Solutions.
CURTISS WRIGHT Flow Control Company
Farris Engineering
We invite you to see all that we can do for you. http://farris.cwfc.com or +1 440-838-7690
Select 56 at www.HydrocarbonProcessing.com/RS

T. FRANK, Lubrizol, Wickliffe, Ohio; and **W. BRIDGES,** Coastal Chemical Co., Abbeville, Louisiana

Novel use of sulfiding agent prevents refinery shutdown

When an insufficient supply of acid gas threatened to shut down operations at a US refinery, quick thinking, along with the novel use of a sulfiding agent and a strong supply chain made all the difference.

A sulfiding agent is a chemical additive with a high percentage of sulfur that is typically dosed by the refinery or a third-party service company to help activate hydroprocessing catalysts. Traditionally, a refinery's need for a sulfiding agent is short term and event based. Hydroprocessing catalysts are typically delivered to refineries in an inert state. They then have to be sulfided to be activated, a process requiring a sulfiding agent. Under usual conditions, the sulfiding chemistry does its job, activating the catalyst and operations at the refinery startup as planned. Recently, however, this refinery found a new and unique application for its sulfiding agent: as an emergency sulfur supplement for its acid gas stream.

When the unexpected happens. The refinery was nearing completion of a partial plant maintenance turnaround. The crude, coker and several other units were down for final turnaround activities, while multiple units, including the fluid catalytic cracking (FCC), reformer and petrochemical units, remained operational on inventoried or purchased feedstocks.

Unexpected mechanical difficulties with the FCC forced it to shut down, along with the crude and coker units, leaving only a few small hydrotreaters running. With little acid gas being produced, there was not enough to satisfy the minimum requirements for the last operating train of the sulfur recovery unit (SRU). Without an SRU available to process sour gas, the refinery faced being out of environmental compliance, which would trigger a plantwide shutdown. Such a shutdown would result in serious economic and safety issues, and add three to four days of lost production while the refinery was restarted.

The right solution. Personnel at the refinery searched for answers to this dilemma. One possibility: quickly securing an outside source of sulfur to supplement the acid gas production in order to keep the SRU online. The refinery estimated that it needed an additional seven tons of sulfur per day, for at least a week, until the FCC unit could be returned to service. But where to get that much sulfur, that quickly? In this case, the answer was to put a sulfiding agent to work in a new way.

Late on a Thursday afternoon, the refinery turned to a strategic partner supplier with an unusual request. Could they provide enough of the sulfiding agent to meet the refinery's demand? The supplier answered in the affirmative. The refinery then determined that it needed four trucks of material as soon as possible to keep the refinery running. The supplier sourced the sulfiding agent and began loading the first two trucks by midday Friday. Tandem teams of drivers headed out from the supply terminal in Houston, Texas, and both trucks arrived at the Midwestern refinery by Saturday. The manufacturer also shipped a specially designed injection pump from the manufacturer's Ohio plant. This skid was specifically designed to inject the sulfiding agent into the refiner's process.

Success. By the time the sulfiding agent arrived at the refinery on Saturday afternoon, the refiner was in the midst of executing its modified operation plan (**FIG. 1**). The FCC-naphtha hydrotreater had been put back into service at reduced rates and was processing a stream of inventoried FCC naphtha containing approximately 0.5% sulfur. The sulfiding agent was immediately injected into the suction of the unit's charge pumps to supplement the unit's acid gas production. It worked. In the following days, two additional trucks of the sulfiding agent were shipped from Houston, and the refinery was able to maintain its existing operations until the FCC was returned to service nearly a week later.

FIG. 1. A sulfiding agent being offloaded at the refinery.

Bottom-line benefits. It is difficult to approximate the economic, environmental and safety benefits of keeping ancillary refinery process equipment in service. With proper operation of the SRU, the refinery was able to continue to treat refinery flare gases, avoid a complete refinery shutdown and continue production of refined products.

It can be estimated that avoiding a complete refinery shutdown allowed the refinery to process an additional 100,000 barrels of naphtha at the reformer and petrochemical plant. Not processing those naphtha barrels at the time would have meant backing out the processing of future crude barrels. Assuming that naphtha is approximately 20% of crude, the processed naphtha provided 500,000 barrels of future crude processing capacity. At an assumed incremental crude margin of $2/bbl, keeping the reformer feed pretreater and reformer online contributed an estimated $1 MM to the refinery's future operating income.

These figures do not take into account the additional economic benefits of keeping ancillary refinery process equipment in service, which can be substantial. As these figures show, it is clear that keeping the refinery running was the right thing to do.

The keys to the success of this effort were:

1. A willingness to try something different
2. Good communication with suppliers
3. A solid supply chain.

The result was a highly efficient, reliable and integrated system that allows for quick responses when issues arise. While it's not unheard of to expand the use of sulfiding agents into other areas (most notably as an anti-coking agent in refinery and petrochemical furnace applications), using the sulfiding agent as a supplemental source of sulfur is truly a novel idea, and a good one. Under the right conditions, it can save time and money, while avoiding potential environmental issues.

The ultimate goal of every refinery is to operate safely and profitably. However, when process units have to be unexpectedly shut down, both such objectives are jeopardized. By using a creative approach, and working closely with its suppliers, refinery personnel were able to find a way to maintain safe operating limits and realize additional profits. HP

TONY FRANK is a technology manager for fuel products at Lubrizol. In his role, he is responsible for the chemical technology behind finished fuel additives and process chemicals for the refining industry. He holds an MS degree in chemical engineering from the University of Toledo in Ohio. Prior to joining Lubrizol, Mr. Frank garnered refining experience working for Sunoco and Marathon in the areas of process engineer, plant economics and operations management.

WILL BRIDGES is regional sales manager for Coastal Chemical Co.. For the past 11 years, he has worked in various management roles, including product manager, projects manager and territory manager in southern Louisiana. Mr. Bridges holds a BS degree in chemical engineering from Louisiana State University.

A. K. SHUKLA, A. K. DUTTA ROY and P. S. CHAUHAN,
Oil and Natural Gas Corp., Uran plant, Maharashtra, India

Evaluate surfactants as antifoam substitutes in amine systems

Oil and Natural Gas Corp. (ONGC) operates an oil and natural gas processing complex at Uran in Maharashtra, India. Crude oil and associated gas produced from ONGC's western offshore fields are transported to onshore facilities at the Uran plant through subsea pipelines for processing. The crude oil is further processed, and the stabilized crude is sent to refineries via pipeline and oil tanker.

The associated gas is combined with the crude stabilization offgas and the condensate fractionating unit offgas. It is passed through gas sweetening units GSU12 and GSU13, and then through LPG recovery plants LPG1 and LPG2 for the recovery of LPG and natural gas liquids (NGL). After the recovery of liquefied petroleum gas (LPG), the gas is passed through an ethane/propane recovery unit to recover C_2/C_3 product from the gas. The lean gas is then routed to consumers via pipeline.

The two gas sweetening units were originally designed with diisopropanol amine (DIPA) and sulfolane solvent systems. However, due to a severe foaming problem in 2005, methyldiethanol amine (MDEA) and piperazine were added to the solvent as an immediate remedial measure. Since that time, the addition of DIPA in the system has been discontinued, and only MDEA and piperazine are added, infusing the solvent system with more amine. However, the system still contains DIPA, sulfolane and the degradation product oxazolidone.

The gas handling capacity of the plant is greatly impacted by frequent foaming problems. Foaming in the amine solution increases the differential pressure (dP) of the absorber column, and the solvent level also decreases, causing frequent process upsets. Foaming also leads to the loss of amine solution through carryover. To reduce foaming, silicon-based antifoam is used at the dosing rate of 3 liters (l)/shot (7 ppm–8 ppm) on an "as-required" basis, with the normal frequency being once every 2–3 days during foaming upsets.

The availability of antifoam has always been a problem, as this chemical is not produced locally. The requirement is very small (approximately 200 l/yr), so the procurement of such a small quantity from a foreign source is a difficult task. With the goals of developing alternate sources to reduce dependency on one source and improving availability of antifoam material, a study was undertaken to evaluate the performance of antifoam samples collected from different local sources on the amine solution used in GSU12 and GSU13.

Most of the silicone- and EO/PO-based surfactants have low surface tension, low interfacial tension, low interfacial shear viscosity and low intermolecular interaction, which are required for successful demulsification. Since these are also the properties required in a superior antifoam, the performance of in-house samples of ethylene oxide/propylene oxide (EO/PO)-based demulsifiers and silicon-based demulsifiers was evaluated along with commercially available antifoam samples. All of the samples were first evaluated for their antifoaming properties in the process and in the plant's product control laboratory, and then in actual field conditions. Results are shown in **TABLES 1–4.**

TABLE 1. Performance of silicon-based demulsifiers DS-A and DS-B

Train	Foam height without addition of surfactant, ml	Time required to break the foam	Foam height, ml, with DS-A, doses at 2 ppm	Time required to break the foam	Foam height, ml, with DS-A, doses at 5 ppm	Foam height, ml, with DS-B, doses at 2 ppm	Time required to break the foam	Foam height, ml, with DS-B, doses at 5 ppm
GSU12	65	50 sec	Slight	Just after stopping the gas flow	No foam	Slight	Just after stopping the gas flow	No foam
GSU13	70	60 sec	Slight		No foam	Slight		No foam

TABLE 2. Performance of EO/PO-based demulsifier D-EO

Train	Foam height without addition of surfactant, ml	Time required to break the foam	Foam height, ml, with D-EO doses at 2 ppm	Foam height, ml, with D-EO doses at 5 ppm	Foam height, ml, with D-EO doses at 10 ppm	Foam height, ml, with D-EO doses at 20 ppm	Time required to break the foam	Foam height, ml, with D-EO doses at 25 ppm
GSU12	68	52 sec	50	32	15	Slight	Just after stopping the gas flow	No foam
GSU13	74	66 sec	61	37	16	Slight		No foam

TABLE 3. Performance of antifoam A-11

Train	Foam height without addition of surfactant, ml	Time required to break the foam	Foam height, ml, with A-11 doses at 2 ppm	Foam height, ml, with A-11 doses at 5 ppm	Foam height, ml, with A-11 doses at 10 ppm	Time required to break the foam	Foam height, ml, with A-11 doses at 15 ppm
GSU12	62	52 sec	45	28	Slight	Just after stopping the gas flow	No foam
GSU13	78	65 sec	47	30	Slight		No foam

TABLE 4. Performance of antifoam A-12

Train	Foam height without addition of surfactant, ml	Time required to break the foam	Foam height, ml, with A-12 doses at 2 ppm	Foam height, ml, with A-12 doses at 5 ppm	Foam height, ml, with A-12 doses at 10 ppm	Time required to break the foam	Foam height, ml, with A-12 doses at 15 ppm
GSU12	66	54 sec	40	22	Slight	Just after stopping the gas flow	No foam
GSU13	75	66 sec	49	35	Slight		No foam

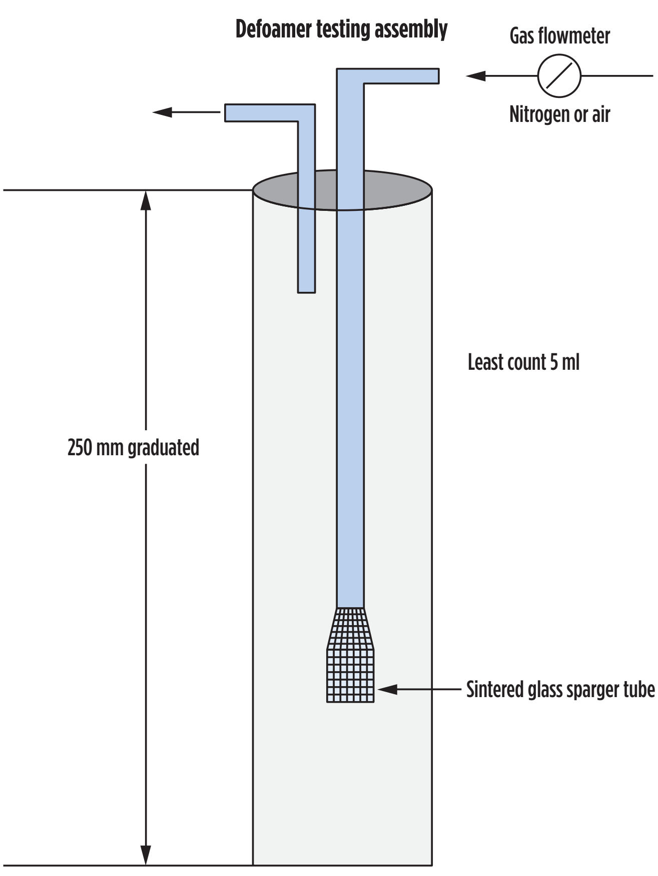


FIG. 1. Simplified defoamer testing assembly.

Foaming in amine solutions. The amine system is prone to foaming. Foaming problems in amine solutions are mainly due to contamination by materials that reduce surface tension and increase viscosity. Foaming occurs as a result of physical inclusion of gas bubbles enveloped by the amine solution. Clean amine solutions have a low tendency of foaming.

Foaming in amine solutions is not a problem unless the foam is stable.[1] Foam stability increases due to the presence of any material that increases the viscosity of the amine solution or reduces liquid drainage from the foam bubble. Similarly, the foaming tendency of the amine solution increases due to the presence of solids, acids and dissolved hydrocarbons in the amine solution. Three types of compounds are common in amine systems that increase the foaming tendency:[1]

1. Hydrocarbons dissolved in the amine solution can increase the viscosity of the solution, which results in the slow drainage of liquid from the surface film. Hydrocarbon solubility depends on the type of amine solution being used. Hydrocarbon solubility in MDEA is higher than in monoethanol amine (MEA) or diethanol amine (DEA). Due to the higher affinity of MDEA for hydrocarbons, the foaming tendency and formation of stable foam are greatly increased in MDEA solvent if higher hydrocarbons are present in the system.
2. The presence of acids, particularly organic acids that may be generated by amine degradation, increases foam stability. Organic acid increases the solubility of hydrocarbons in the amine solution and also increases the hydrocarbon/amine solution emulsion, which slows bubble drainage. Organic acids also raise the viscosity of the solution and promote the formation of gel-like, bubbled surface film.
3. Ingress of solids (particularly those less than 5 microns), iron-based corrosion products, and carbon particles from activated charcoal filters greatly increase the foaming tendency of the amine solution. These solids, having a tendency to concentrate on the liquid surface, increase the surface viscosity and slow the drainage of liquid from the bubble.

Antifoams and defoamers. Foams and emulsions have much in common; both are examples of immiscible fluids that are intimately dispersed. Surface elasticity and surface viscosity are of importance in the dispersal systems for both fluids. In a defoaming process, a gas phase is separated from a liquid phase, which requires interface destabilization and coalescence. In a demulsification process, a liquid phase needs to be separated from another liquid phase through interface destabilization, droplet flocculation and coalescence.[2]

Antifoams are chemicals designed to reduce and hinder the formation of foam. Defoamers are chemicals that break down existing foam. While there is a distinct difference between these two actions, the chemical interactions are similar, and many antifoam additives are added to act as both an antifoam agent and a defoamer.[2]

Generally, antifoam additives are insoluble in the foaming medium and have surface active properties. Any chemical that is to be used as an antifoam agent must interact with the foam-stabilizing surfactant; it will behave as antifoam if it is more surface active than the surfactant. It can then enter the foam bubble wall and destroy the film elasticity, thereby dropping the surface viscosity and facilitating gas diffusion.[2]

There are thousands of different antifoams available on the market. Generally, anything that has destabilizing effects on foam can be used as an antifoam.[3] Antifoams can be broadly classified into the following categories:

- Oil-based antifoams have an oil carrier; these may be further divided into two categories: Non-polar oils, such as mineral oil and silicones; and polar oils, such as fatty alcohols, fatty acids and alkyl amines
- Powder antifoams are oil-based defoamers on a particulate carrier, such as silica
- Silicon-based antifoams have a silicone compound as the active component
- Alkyl polyacrylates are suitable for non-aqueous systems where air release is more important than the breakdown of surface foam
- EO/PO glycol copolymer-based antifoams.

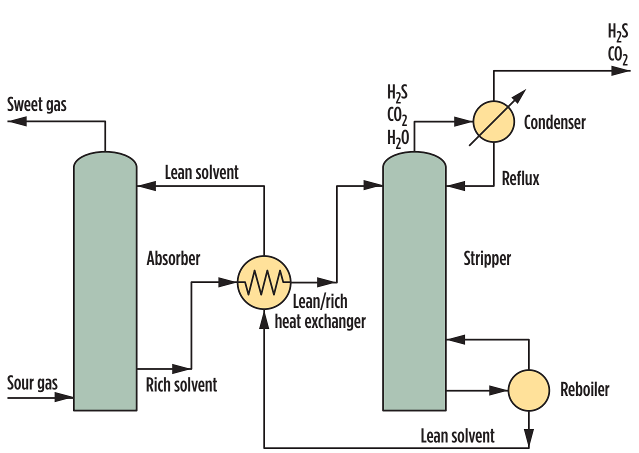


FIG. 2. Simplified amine gas sweetening process flow diagram.

TABLE 5. Scale of foaming tendency and foam stability

Foaming tendency	
Foam height, ml	**Foaming tendency**
< 10	None
10–60	Slight
60–125	Moderate
> 125	Severe
Foam stability	
Foam break time, sec	**Foaming tendency**
< 5	None
5–20	Slight
20–30	Moderate
> 30	Severe

Antifoams for amine systems. For a material to be an effective antifoam for an amine system, it must have extremely low solubility in the amine solution.[1] Three types of antifoams are available for amine systems: polyglycol, silicon and high-molecular-weight alcohols.

The selection of the right antifoam to mitigate foaming problems at a particular plant is often a trial-and-error process. The performance of antifoam samples received from different sources on media samples collected from GSU12 and GSU13 at different time intervals are recorded in **TABLES 1–4** and **FIGS. 1–4.**

Determination of foaming tendency and stability. The method used for the determination of foaming tendency and the stability of an amine solution is described below,[4] and a simplified sketch of a defoamer testing assembly is shown in **FIG. 1:**

1. Pour 50 ml of amine solution into a graduated cylinder
2. Establish gas flow at 0.2 l/min; insert sparger tube and record foam height after 5 min.
3. Stop gas flow; using a stopwatch, record the time required for foam to break to a liquid surface
4. Repeat steps 1–3 after adding antifoam product at the desired concentration.

Interpretation of results. The scale of foaming tendency and foam stability is shown in **TABLE 5.**[4] The experimental results show that, although the foaming tendency of the untreated amine solution is moderate, the foam stability is severe. However, after adding just 2 ppm of silicone surfactant DS-A or DS-B, there was only a slight foaming that disappeared immediately after stopping the gas flow.

When 5 ppm of either DS-A or DS-B surfactant was added to the amine solution, no foaming was observed in either GSU12 or GSU13. The experimental results of other samples (i.e., EO/PO-based surfactant D-EO and antifoams A-11 and A-12) showed similar antifoaming tendencies, but only at very high doses of 10 ppm and 20 ppm.

Field trial. Based on the above laboratory observations, silicone-based surfactants DS-A and DS-B were tried in actual field conditions. During the system upsets due to foaming, 2 l of surfactant DS-A (based on the laboratory-evaluated dose

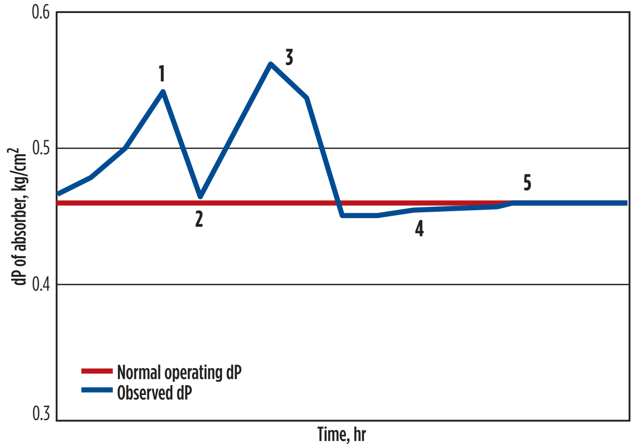


FIG. 3. Foaming upsets and response of DS-A surfactant.


TO BUILD A PROJECT THIS SIZE TAKES:
ENGINEERING & MODULAR CONSTRUCTION EXPERTISE
A SKILLED WORKFORCE
WORLD-CLASS FACILITIES
UNWAVERING COMMITMENT TO SAFETY & QUALITY
IT TAKES
AVONDALE
www.hii-avondale.com

of 4 ppm–5 ppm) was added to the amine solution in one shot to evaluate its performance in actual field conditions. The trial was conducted over a period of two months using surfactants DS-A and DS-B, one at a time, to evaluate the performance of both products.

Foam broke immediately after dosing the silicon surfactant into the recirculating amine solution. The dP of the system also increased for a short time, although it subsequently dropped below the normal operating level, possibly due to the breaking of the foam and the sudden release of the foam-trapped solution. After 10–12 hr of silicon surfactant dosing, the dP of the system stabilized to the normal operating level, and the plant was able to operate at a higher feedrate (**FIGS. 3–4**).

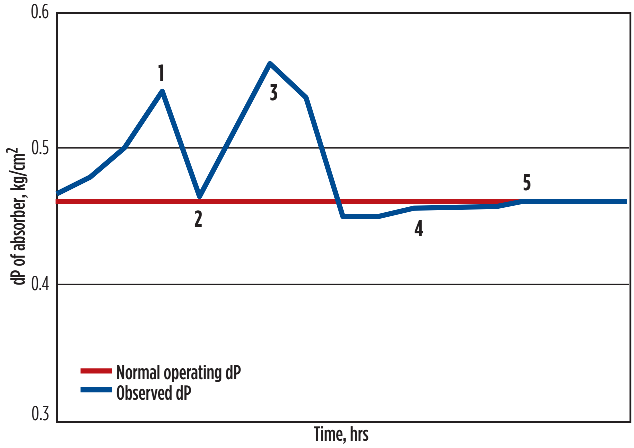


FIG. 4. Foaming upsets and response of DS-B surfactant.

The volume of amine solution in the tank also increased, giving more operational freedom to the operator as the plant ran at low volume. During the foaming upsets, after a dosing of surfactant DS-A or DS-B, the dP of the system was sustained at a normal operating level for a considerable time (7–10 days) without any foaming in the system, and the frequency of antifoam shots was reduced from once every 2–3 days to once every 7–10 days. The quantity of antifoam, which was required at 3 l/shot with previously used antifoams, was reduced to 2 l/shot, making the amine solution less contaminated.

FIGS. 3–4 show the foaming upsets and responses of surfactants DS-A and DS-B. In **FIG. 3,** the absorber dP was increased due to foaming, the feed was decreased to bring the dP down, surfactant was added to the system, and feed was increased slowly. As a result, the dP was decreased and maintained at a normal operating level. In **FIG. 4,** the absorber dP was increased due to foaming, and the feed was decreased to bring the dP down. Surfactant was added in the system, and the feed was increased slowly. As a result, the dP was decreased and maintained at a normal operating level.

Takeaway. During the lab study and field trial, it was observed that, at the dosing rate of 4 ppm–5 ppm (i.e., 2 l/shot in actual field conditions), both silicone-based surfactants (demulsifiers) DS-A and DS-B have served as excellent antifoam agents for the amine solution at the Uran plant.

The antifoam requirement for the Uran plant is very small (200 l/yr) and will decrease further by using either of the above silicone-based surfactants. Since both surfactants are being used in large quantities in offshore installations as a low-temperature emulsion breaker, their easy availability is ensured. The application of DS-A and DS-B surfactants as antifoam agents offers many performance advantages as compared to the conventional antifoam used at the Uran facility. HP

ACKNOWLEDGMENTS

The authors are grateful to U. S. Deshmukh, plant manager at the Uran plant, for his guidance and encouragement, without which this work would not have been possible. The authors also thank R. Nagarajan, G. P. Bahuguna, Dr. M. M. Siddiqui and Dr. A. K. Saxena for their valuable suggestions. Additionally, R. N. Tiwari, A. T. Jadhav and Mr. Jaymohan provided support during the field trial, while R. K. Bhatt and Mr. Jagmohan provided support during the laboratory studies.

LITERATURE CITED

1 GE Water & Process Technologies, technical bulletin, 2006.
2 Rome, C. and T. Hueston, "Silicon in the Oil and Gas Industry," 2002.
3 Von Phul, S. A. and L. Stern, "Antifoam: What is it? How does it work? Why do they say to limit its use?" 2004.
4 "Process Licensor's Operation Manual, Book 1," Oil and Natural Gas Corp. Uran refinery, *Plant Operation Manuals,* 1990.

DR. A. K. SHUKLA joined Oil and Natural Gas Corp. (ONGC) in 1989 as a graduate trainee and now serves as chief chemist. He has varied experience in quality assessment and process quality control of crude oil, natural gas and value-added products. Presently, Dr. Shukla is working at ONGC's process facility and looking after in-process, quality control and assessment of crude oil, gas and various value-added products. He has been instrumental in developing silicon surfactants as substitutes for antifoam agents for the gas sweetening units and the implementation of a third-generation gas sweetening solvent system.

Saving You Money Every Step of the Way

PowerDyne®

Pure Performance Thermodynamic Disc Traps

TLV. CORPORATION
13901 South Lakes Drive, Charlotte, NC 28273-6790
Phone: 704-597-9070   Fax: 704-583-1610
E-mail: TLV@TLVengineering.com
For Technical Service 1-800-TLV-TRAP
http://www.tlv.com

Member of FCI Fluid Controls Institute, Inc.

Manufacturer
TLV® CO., LTD.
Kakogawa, Japan
is approved by LRQA Ltd. to ISO 9001/14001

ISO 9001/ISO 14001

H. Z. HA, Fluor Canada Ltd., Calgary, Alberta, Canada;
and **D. TIWARY,** Canadian Natural Resources Ltd.,
Fort McMurray, Alberta, Canada

Consider accurate feed characterization with improved simulation methods

Assay definition is crucial for accurate process simulations. The performance of process simulators depends on the feed definition to the processing unit. Modern analytical laboratories can offer complete assay measurements on the bulk properties, distillation behavior, composition (elemental, molecules of light ends and hydrocarbon types of heavy fractions), heavy metals, salt content, etc.

A well-defined laboratory assay can establish the sound basis for simulation. However, simulation performance not only depends on the quality of data, but also on the way these data are implemented. Inputting the assay data into the simulator in a different way will generate different simulation results, which must be addressed and validated.

Models hold promise. Simulation software use pseudo-components defined from the true boiling point (TBP) distillation curve to calculate the thermo-physical behavior of petroleum when processed.[a] The TBP distillation usually refers to ASTM D2892 distillation (15 plates operating at a reflux ratio of 5:1) and D5236 vacuum pot-still method. These physical distillation methods are costly and time-consuming. The final boiling points (FBPs) of these TBP methods are also limited (for D2892 maximum of 400°C and D5236 to about 565°C). However, the simulated distillation data are usually provided by the refiner. Simulated distillation via gas chromatograph (GC) can provide a fast and cost-effective way to define detailed distillation curves for petroleum over a wide range of boiling points. Unlike the physical distillation methods (ASTM D86, D1160, D2892 and D5236), in which the boiling ranges are limited by the cracking of petroleum samples when heated over 360°C, the high-temperature simulated distillation (HTSD) can determine the boiling range distribution for petroleum from 36°C to 720°C (ASTM D7169).

These HTSD methods are crucial for characterizing oil sands and bitumen as more than 50% bitumen fractions and are in the vacuum residue (VR) range (> 525°C). In this study, the simulated distillations for ASTM D2887 and D7169 are used to determine the boiling point distribution of the diluent and bitumen samples. The study addresses the effects of data entry methods on model performance.

BACKGROUND

In using simulation software, the only option for simulated distillation curve input is D2887. ASTM D2887 provides a quantitative percent mass yield to the corresponding boiling point distribution of petroleum fractions with an IBP of 55°C and FBP of 538°C. The IBP refers to the boiling point for 0.5 wt% of the sample, while the FBP refers to the boiling point of 99.5 wt% of the sample accumulatively distilled. To obtain the TBP distillation for the simulation, the API conversion methods (API 1994 direct and indirect methods) are applied to convert the D2887 into a TBP distillation. The API 1994 conversions were developed based on the statistical studies of TBP and on simulated distillation data obtained from conventional crudes and refined products.

Standards. Conventional crudes are classified as light, and the D2887 test can cover the boiling range. Based on the extensive lab data collected from conventional oils, API Procedure 3A3.1 (*API Technical Data Book,* 1994) proposed that the TBP at 50 vol% distilled equals to the simulated distillation temperature at 50 wt% distilled. The TBPs below and above 50% boiling are modified with weighting factors correlated from lab data. The API 1994 conversions depict the relationship between TBP and simulated distillation of conventional crudes.

However, with the advances in modern GC technology, the simulated distillation method can not only extend the test boiling range to heavier petroleum materials, but it can also match the TBP curve within an acceptable deviation. By using a reference oil TBP calibration curve, the HTSD can generate a distillation curve matching the TBP distillation on the wt% off basis over the full TBP range measured by D2892 and D5236 within ± 2% difference.[1] This suggested that the HTSD can be used as the TBP distillation based on the wt%. A simulation study recommended replacing the TBP test with the HTSD results to improve the process simulation performance for the vacuum distillation unit.[2] For comparison, both the direct simulated distillation input (DSDI) method (the input simulated distillation data as the TBP on a mass basis) and the simulated distillation data with API 1994 conversions (referred as API 1994) are applied to illustrate the difference in simulation performance from different assay input methods.

PROCESS SIMULATION WITH THE PLANT DATA

A typical bitumen assay was developed based on the ASTM D7169 HTSD method. The assay was averaged from

50 bitumen samples from different mining sites or depths to represent the actual bitumen feed to the upgrader. The diluent is also defined from simulated distillation D2887 averaged from the daily sampling of recovered diluent from the diluent recovery unit (DRU) over a three-month-best operation period. Two different simulation cases were developed by commercially available simulation software.[a]

In one case, the D7169 HTSD data was directly input into the assay definition as the TBP distillation on a mass basis.[a] Diluent was also defined in the simulation software from D2887 data by the DSDI method.[a] In the other case, the API 1994 direct conversion method was applied to transfer the simulated assay into TBP distillations based on the same simulated distillation data of bitumen and diluent mentioned earlier. Since the simulators do not have any options for other simulated distillation tests besides ASTM D2887, the D7169 was input into the D2887 entry, as the essence of both tests are the same.[a] All the other simulation setups and specifications were identical for these two cases to outline the baseline for comparison.

Bitumen models. Diluted bitumen (dilbit) feed was prepared by blending diluent and bitumen in a 50/50 volumetric basis.[a] Peng-Robinson thermodynamic package was selected as the simulation basis to model a DRU in an oil sands bitumen upgrader. The feed TBPs defined by the DSDI method are compared with the TBPs defined by the API 1994 conversions, as shown in **FIG. 1.** The bitumen defined by DSDI is generally lighter than the bitumen defined from API 1994 conversions, as the DSDI derived TBP curve is (**FIG. 1**) in the TBP curve from API 1994 conversions over the entire distillation range. These differences are inherited into the TBP curves of the corresponding diluted bitumen, as shown in **FIG. 1,** in 50 vol%–100 vol% off the distillation range. However, the TBP profiles of diluent are quite similar regardless of the assay input methods due to the narrow boiling range of the diluent. The TBP of diluent is depicted in **FIG. 1,** and the boiling range of 36°C–200°C (referring to 0 vol%–50 vol% off distilled) shows there are minor overlaps in the TBP between diluent and bitumen samples.

The DRU plant operation data, averaged from continuous 30 best operating days, were used to evaluate the simulation performances with different assay input methods.[a] During the 30 best operating days, the DRU was running at, or close to, design capacity without any major operational issues. The simulations were tuned in the same way to match the averaged product specification based on lab results.

The plant data are compared with the simulation results of the two different assay input methods in **FIGS. 2** and **3.** The dilbit temperature profile as it passes through the preheating train is shown in **FIG. 2.** A general agreement of temperature profile was found between plant data and the results from the two simulations. A temperature drop was predicted by both simulations when the dilbit passes through the preheating exchangers.

The actual plant data indicated that this interesting temperature drop did occur when dilbit was being heated by gasoil (GO) product from a coker unit in the preheating exchanger. The dilbit temperature drop was caused by latent heat used when a significant amount of light hydrocarbons and water vaporized near 135°C in the preheating train. The general agreement between plant data and the two simulations indicate that the thermophysical (e.g., enthalpy) properties of pseudocomponents defined by two different assay entry methods are

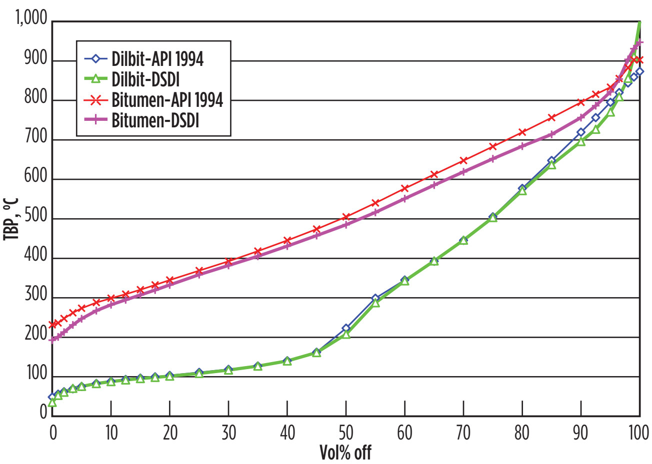


FIG. 1. Comparison of feed definitions during engineering and plant operation.

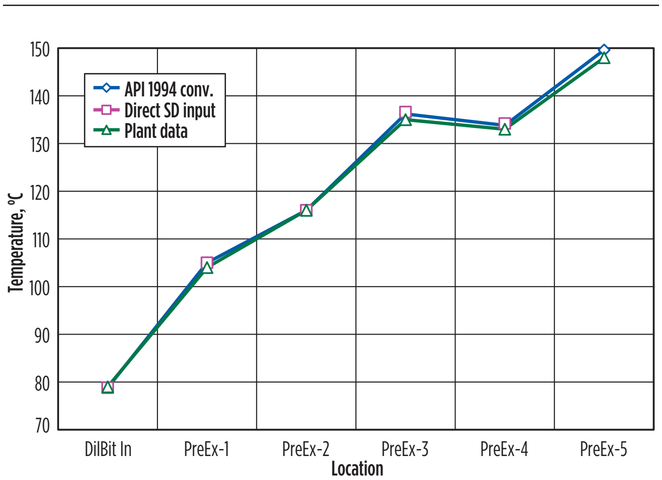


FIG. 2. The temperature profile of dilbit through the preheating train.

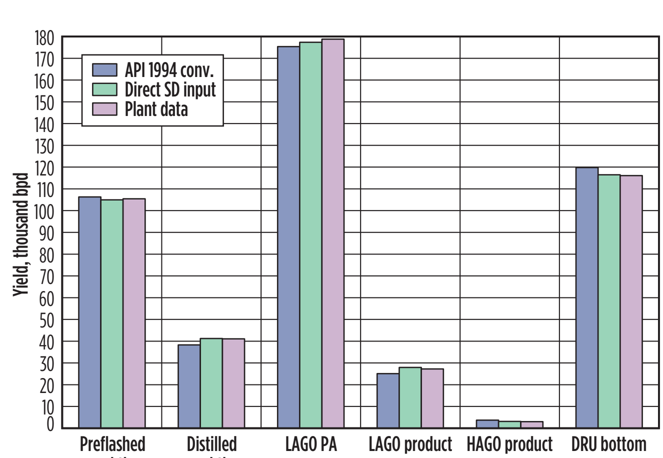


FIG. 3. The product yield distribution of a DRU.

very close. The difference caused by assay entry methods in thermal behavior is minor.

The simulated distillation to TBP conversion can distort the true distillation behavior of feed characterization simulations given the fact that the state-of-the-art gas chromatography techniques can virtually resemble the TBP distillations on a mass basis.

FIG. 3 illustrates the product yields and pumparound (PA) rate of the DRU unit as compared with the two simulations using different assay entry methods. As shown in **FIG. 3,** the method for inputting simulated distillation assay data into simulators has a substantial influence on the product distribution. The difference in product yields due to assay entry methods range from 1,000 bpd (pre-flashed naphtha) to about 4,000 bpd (DRU bottom). The total performance of DSDI defined simulation is superior to that of simulation using API 1994 conversions. The simulation performance with DSDI not only predicted good matches on individual product yields, but also predicted a closer light atmospheric GO (LAGO) PA rate to plant data as compared to the predictions from the simulation using API 1994 conversions. The averaged relative deviation (ARD) of DSDI defined simulation, defined as (predicted yield – plant yield)/plant yield, is less than ±2%. The ARD of the simulation using API 1994 conversions is about ±8.3%.

The economic impact on the project resulting from these differences in yield can be significant. The estimated profit difference on product loss alone could be over $100 million per year, regardless of any engineering cost for troubleshooting later. The predicted difference would also affect the unit equipment design and operating performance.

For example, the simulation with API 1994 conversions typically under-predicted the light product yields from the DRU distillation column due to the higher TBP profile obtained from the conversions. Consequently, the column overhead condenser and the corresponding product pumps would be undersized, which could become a bottleneck. Applying the DSDI method in assay definition in simulators would improve the simulation performance, define a more accurate basis for unit design and provide a better benchmark for plant operation and debottlenecking.

FINAL EVALUATION

A simulation study was conducted to investigate the effect of different assay input methods (based on the same simulated distillation data) on the simulation performance. With the DSDI method, the characterized bitumen TBP contains more light fractions as compared to the TBP defined with API 1994 conversions. As expected, the simulation defined by DSDI predicted higher naphtha and LAGO yields as compared to the predicted yields from simulation with API 1994 conversions. The predicted yields with the DSDI method match the plant data, with ARD < ±2%. The ARD of simulation with API 1994 conversions was about ±8.3%. The good agreement between the performance of the simulation with the DSDI method and the actual plant data suggests that the performance of the HTSD method (ASTM D7169) for characterizing the heavy oils such as bitumen is satisfactory. As a result, direct input of HTSD data as TBP data on a wt% basis in the assay definition of simulators is recommended. The conversion of simulated distillation data to obtain TBP curve is not recommended.

There were incidences of simulations in which the simulated distillation data were erroneously input as the TBP on a vol% basis. Inputting the simulated distillation data as TBP on a vol% basis is fundamentally wrong since the simulated distillation detects the hydrocarbon eluted on a mass basis, and it targets to resemble the TBP curve on a mass basis. If simulated distillation data is erroneously input as the TBP on a vol% basis, then the resulting vol%-based TBP will be identical to the actual TBP profile on a wt% basis. **Result:** The simulation will under-predict light-product yields and over-predict heavy-product yields due to the differences in the densities of light and heavy hydrocarbons. Care should be exercised when inputting an assay data into a simulator, and the DSDI method should be used. HP

NOTES

[a] Study results were done using HYSYS or Pro-II.

LITERATURE CITED

[1] Villalanti, D. C., J. C. Raia and J. B. Maynard, "High-temperature simulated distillation application in petroleum characterization," *Encyclopedia of Analytical Chemistry,* R. A. Meyers, Ed., pp. 6726-6741, John Wiley & Sons Ltd., Chichester, 2000.
[2] Golden, S. W., D. C. Villalanti and G. R. Martin, "Feed characterization and deep-cut vacuum columns: Simulation and design," AIChE 1994 Spring National Meeting, Atlanta, Georgia, April 18–20, 1994.
[3] Villalanti, D., J. Maynard, J. Raia and A. Arias, "Yield correlations between crude assay distillation and high temperature simulated distillation," The AIChE 1997 Spring National Meeting, Houston, March 9–13, 1997.

HARRY Z. HA is a senior process engineer/specialist with Fluor Canada Ltd. He has 25 years of experience in R&D of petrochemical industry. He joined Fluor in 2005 and has been working on several oil sands and heavy-oil upgrading projects. In addition to the working on process design, Dr. Ha focuses on data and methods development to support process modeling and simulations. Before joining Fluor, he worked as a physical scientist at the National Centre for Upgrading Technology of Canada for four years and was involved in the development of upgrading technology for heavy oil such as coking, visbreaking and hydrocracking. Dr. Ha holds an MS degree in environmental engineering from Hong Kong University of Science and Technology and a PhD in chemical engineering from the University of Alberta.

DHARMENDRA TIWARY is a lead process engineer with the Canadian Natural Resources Ltd. Prior to joining CNRL, he was a senior staff consultant with Aspen Technology Inc. (Calgary, AB) and a process engineer in a petroleum refinery of Indian Oil Corporation Ltd., at Panipat, India. He is a registered professional engineer with the Association of Professional Engineers and Geoscientists of Alberta. He received a bachelor's degree in chemical engineering from Indian Institute of Technology at Kharagpur, India and an MS degree in chemical engineering from the University of Calgary.

R. C. PATIL, M. YADAV, A. GUPTA and **A. DAS,**
Reliance Technology Group, Reliance Industries Limited,
Jamnagar, Gujarat, India

Optimize slop-wax bed height in the vacuum unit

The vacuum distillation (VD) unit in the petroleum refinery separates the heavier fraction of crude oil from the atmospheric distillation unit (ADU) into light vacuum gasoil (LVGO) and heavy VGO (HVGO) under a vacuum to avoid thermal cracking of the heavier hydrocarbons. The two-phase mixture of the reduced crude oil (RCO) liquid and vapor from the ADU enters the VD column flash zone and flashes further under low pressure.

While the lighter materials, mainly VGO as vapor, move up the column, the heavier liquid materials, vacuum residue (VR), move to the column bottom. However, some VR liquid droplets are entrained within the vapor and must be washed off to maintain HVGO product quality. A slop wax or wash bed is used to separate out the entrained liquids from the vapor.

Background. The vapors rising from the flash zone are contacted with the hot internal reflux of HVGO from the VDU inside the slop-wax bed, as shown in **FIG. 1.** Normally, structured packing/grid is used for contacting the wash oil with the vapor carrying the entrained droplets due to the low liquid retention property and, consequently, low residence time of the liquid. The slop wax is drawn from the bottom of the slop-wax bed, and it is considered as HVGO from the slop-wax bed in liquid form, i.e., overflash and entrained VR from the flash zone.

Operation trends. At present, a growing trend among refiners is to maximize VGO recovery from the crude unit. However, the 95% cutpoint of VGO exceeds the conventional operational temperature limits of 565°C to 580°C. Increasing the VGO cutpoint comes with the potential to increase coke formation inside the slop-wax or washing bed unless proper engineering design and operational measures are applied.

Engineering design plays an important role in determining the coking rate inside the slop-wax bed as defined by the liquid retention time on the packing, vapor distribution within the packing, entrainment from the flash zone, etc. From the operations point of view, controlling the HVGO internal reflux at optimal values is most critical. With an increase in the internal reflux, the HVGO product becomes a purer product but at the expense of lower HVGO yields. Conversely, providing very low internal reflux will establish zones, within the slop-wax bed that receive very low liquid flowrates (low wetting rates), and thus, these zones are more prone to coking due to the high residence time of the liquid.

Other issues. Coking also causes an increase in the pressure drop, which affects separation in the vacuum column. Eventually, high pressure drop conditions can shut down the unit, thus leading to significant production losses. The wetting rate is the key reliability parameter for sustainable operations of the vacuum column for longer service runs. Wetting rates must be monitored and maintained above the minimum value continuously during the operation.

It is therefore desirable to maintain an internal reflux at a minimum value based on HVGO product specification and the minimum specified wetting rate of structured packing. In other words, there exists an optimum bed height where the maximum VGO yield can be achieved with an acceptable VGO quality and minimum wetting rate to avoid coking.[1]

A systematic study was conducted to investigate the impact of slop-wax bed heights on VGO yield and reliability. Commercially available process simulators can be effectively applied to study these conditions from the process point of view, and engineering knowledge can be used in operations to increase the reliability of the unit. A case study investigated a vacuum distillation column operation to determine how to increase VGO yield while maintaining the product quality without impairing the reliability of the column.

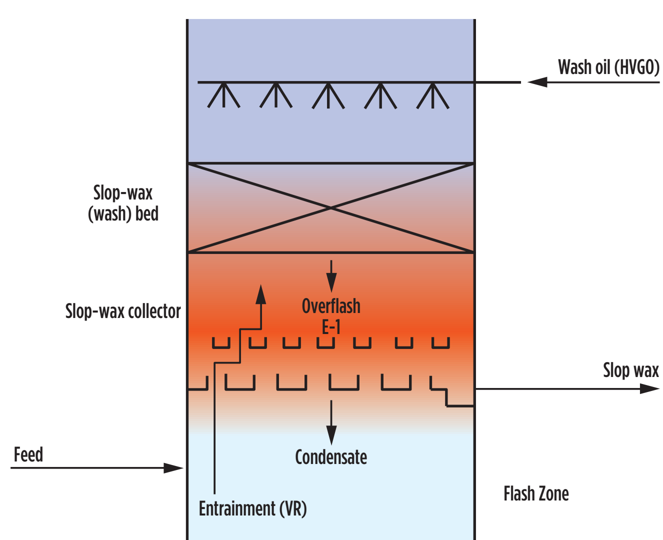


FIG. 1. Wash-zone section of vacuum column.

Simulation methodology. A commercial process simulator was used in this case study. The RCO from the furnace is sent downstream via a transfer line, exhibits the non-equilibrium flashing and generates superheated hydrocarbon vapors before entering the vacuum column flash-zone section. The degree of superheating depends on the transfer line design and furnace operation. To model this non-equilibrium flashing inside the transfer line, the multi-stage flashing at different pressures was incorporated into the simulation. The quality of HVGO was specified in terms of its 95% point, and the HVGO production rate was varied to attain the desired 95% point of HVGO. The required values of internal reflux of HVGO and slop-wax rates were calculated by the simulator. The entrainment and LVGO product rates were kept constant.

A systematic design calculation, for revamp of slop-wax bed in a vacuum distillation unit brings out some interesting observations, which on the hindsight appear very logical. There is a tradeoff between VGO yield and quality along with maintaining the reliability of unit.

In the actual operation, the wetting rate within the packing is monitored by estimating the overflash quantity. This quantity can be determined by a metal balance across the wash bed. The total slop-wax amount (liquid collected at the collector tray below the wash bed) is equal to overflash plus entrainment. By knowing the metals content—nickel (Ni) and vanadium (V)—or Conradson carbon residue (CCR) content of the slop wax, VR and HVGO, it is possible to calculate the amount of overflash in the slop-wax bed. The entrainment is assumed to be VR, and overflash is assumed to be the HVGO. In simulation, the wetting rate is estimated from the quantity of HVGO (i.e., overflash) sent to the slop-wax draw tray.

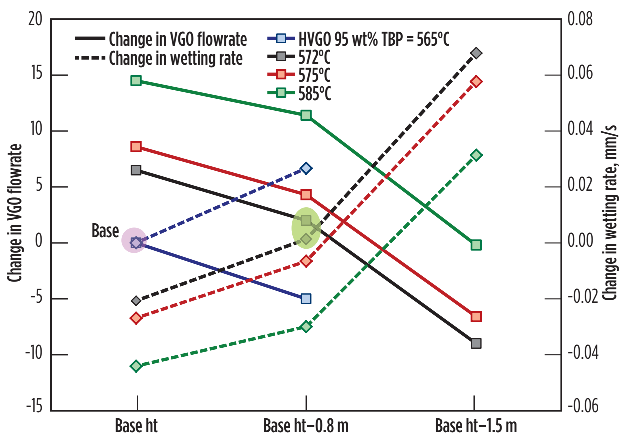


FIG. 2. Impact of slop-wax bed height on recovery.

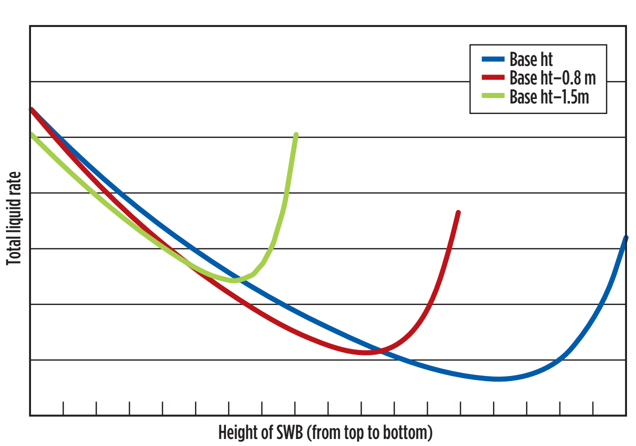


FIG. 3. Liquid traffic along the slop-wax bed height.

Case studies. The study was repeated for three bed heights, as shown in **FIG. 2.** The slop-wax bed height was varied by reducing the stages. At a constant 95% TBP wt point of the HVGO, as the height of the slop-wax bed is decreased, the HVGO yield also decreases and the wetting rate increases. For a given slop-wax bed height, as the HVGO draw rate increases, its 95% point increases while the required HVGO internal reflux, and, thus, wetting rate decreases. The Base Case point is circled, as shown in **FIG. 2.**

From **FIG. 2,** with the existing slop-wax bed, as the VGO yield increased by 8.6 tph by decreasing internal reflux (IR) and specifying VGO 95% TBP wt point as 575°C, the wetting rate decreased by 0.027 mm/s. As compared to the Base Case, there are concerns for potential rapid coking of the bed, and, thus, deteriorating operating conditions. Increasing the VGO 95% TBP wt point to 585°C by increasing the VGO yield to 14.5 tph worsens operating conditions due to the reduced wetting rate by 0.044 mm/s from the Base Case.

When the bed height was reduced by 0.8 m from the Base Case, the VGO yield decreased by 5 tph with the same (565°C) VGO 95% TBP wt point. However, the bed wetting rate improved considerably by 0.027 mm/s from the Base Case, thereby improving unit reliability. Drawing off more VGO product by 4.3 tph and 11 tph, respectively, will result in a 95% TBP wt point of 575°C and 585°C and the wetting rate is reduced by 0.007 mm/s and 0.030 mm/s with respect to the Base Case.

Reducing bed height further shifts the attention from the wetting rate constraint to the limits on VGO quality. By reducing the bed height another 1.5 m, even at the same VGO product draw as that of the Base Case, the 95% TBP wt point increases to 585°C, which may not be acceptable due to higher Ni and V content. However, it can be understood that it is very comfortable with regard to bed wetting (increase of 0.031 mm/s from the Base Case).

To maintain specified VGO 95% TBP wt. point, as the bed height is reduced, the HVGO IR rate must be increased, which decreases the VGO yield but increases the wetting rate of the packing. This defines the zero line, as shown in **FIG. 2,** which determines the maximum VGO yield achievable for a fixed slop-wax bed height with the minimum wetting rate. In general, every refinery applies separate guidelines for the minimum wetting rate based on operational experience. Above this line, the wetting rate would be sufficient to minimize coking within the packing, and, thus, the VGO yield can be improved by varying the bed height according to the VGO quality as defined by the 95% TBP wt point and metals contents. Below the line, either the wetting rate would be inadequate or there will be a loss in VGO yield.

The bed height can be altered to attain a maximum yield of VGO from the vacuum column while maintaining the same wetting rate as the Base Case and the HVGO 95% TBP wt point within reasonable limits with respect to the metals content in the final VGO product. Reducing the distance between VGO yield and wetting rate curves is necessary to attain the required results. In this case, the additional 2 tph of HVGO flow can be attained with the HVGO 95% TBP wt at 572°C while sustaining the same wetting rate as that of the Base Case when the bed height is reduced by 0.8 m, as shown in **FIG. 2.**

In this operating case, it is also necessary to address coking concerns. With the shorter slop-wax fixed bed, less coking occurs as compared to using a taller bed even at the same wetting rates. Less coking occurs for several reasons and includes:

- With the shorter bed, liquid channeling is less severe as compared to the taller bed. Channeling becomes a significant problem as the bed height increases. Redistributors are used to mitigate channeling.
- Due to poorer separation, the slop-wax quality is lighter with shorter beds. Again, less coking is expected.
- Shorter bed will cause lower pressure drop.
- In many refineries, the wash bed will coke up in the middle as the liquid rates are reduced in the center of the bed.

Coking problems. One of the explanations for increased coking is that the VR entrainment also helps in wetting the packing. The entrained VR becomes zero after a certain height within the bed as it is washed with the internal reflux of the HVGO. The HVGO liquid flowrate decreases from the top to the bottom of bed due to separation, as well as re-vaporization of liquid occurs. Therefore, a minimum liquid rate is needed between the top and bottom of the bed.

For example, if exponential profiles of HVGO liquid and VR flows are assumed, then the total liquid traffic in the slop-wax bed varies at different heights, and it can be estimated. **FIG. 3** illustrates the estimated liquid traffics within the bed. As shown from this figure, as the bed height is reduced, then the minimum liquid flowrate shifts to higher values. The different VR and HVGO profiles along the bed height may establish different liquid traffic profiles However, the lower bed height will have a flatter liquid flow profile, and the chances of localized coking within the middle of the bed are reduced.

The reliability of the wash-zone section depends largely on the flash zone entry device design, liquid distribution and packing used in the wash-zone section.[2, 3] One would like to have packing that would enhance the performance of the slop-wax bed. In general, lowering the packing height with high-separation and de-entrainment efficiency is desired to maintain high-quality HVGO and to minimize the wetting rate within the packing, and thus avoid coking.

Roundup. The simulation clearly demonstrates that there is a tradeoff between the VGO yield and quality along with ensuring reliability of the unit. There exists an optimum slop-wax bed height where a higher VGO yield can be achieved without violating the constraint on quality or minimum wetting rate, which is very important for reliability. Besides the simulation, other factors favor using the shorter bed and include:

- With the shorter bed, liquid channeling is less severe as compared to using a taller bed.
- Due to poorer separation, the slop wax is lighter than in the case for the shorter bed. **Result:** Less coking occurs.
- The axial liquid flow profiles in shorter beds are flatter; therefore, the profiles are better as compared to taller beds.
- Lower pressure drop occurs in the shorter bed.

Along with the bed height, the packing type, internal reflux distributor and flash-zone entry devices are important factors contributing to the sustainable operation of a vacuum distillation unit. HP

LITERATURE CITED

1 Golden, S. W., N. P. Lieberman and E. T. Lieberman, "Vacuum Column Troubleshooting," Ed. J. Mcketta, *Encyclopedia of Chemical Processing and Design,* Vol. 60, pp. 218–230.

2 Siegel, L. and W. Meier, "Distillation columns with structured packings in next decades," *Chemical Engineering Research and Design,* Vol. 81, No. 1, January 2003, pp. 39–47.

3 Pilling, M. and M. Bannwart, "Use of gravity flow distributor increases capacity in refinery vacuum tower wash sections," AIChE Petrochem Expo, Distillation Symposium, 2001.

RAHUL C. PATIL is the senior manager, working in the area of modeling and simulation at the refining R&D division, Reliance Industries Ltd., Jamnagar, Gujarat, India. He holds an MS degree in chemical engineering from IIT Kanpur and a BS degree in chemical engineering from ICT Mumbai. His experience includes debottlenecking, application of computational fluid dynamics, process optimization and troubleshooting in various refinery processes units comprising crude and vacuum distillation unit and delayed Coker.

MANOJ YADAV is the general manager, working in the area of fluidized catalytic cracking (FCC) process engineering at the refining R&D division, Reliance Industries Ltd., Jamnagar, Gujarat, India. He holds an MS degree in chemical engineering from IIT Delhi and a BS degree in chemical engineering from Punjab University, Chandigarh. His experience includes process optimization and troubleshooting in various refinery processes, process development in the area of separation and FCC.

DR. AJAY GUPTA is the assistant vice president and leads the modeling and simulation group of refining R&D at Reliance Industries Ltd., Jamnagar, Gujarat, India. He has BS and MS degrees and a PhD in chemical engineering from IIT, Delhi, India. His experience includes implementation of advanced process control strategies in FCC and crude distillation units in several Refineries, development of reactor models for various chemical processes including effect of scale, and application of computational fluid dynamics for solving problems in petroleum refining and petrochemical industry.

DR. ASIT K. DAS is vice president and head of the refining R&D at Reliance Industries Ltd., Jamnagar, Gujarat, India. He holds a BS degree in chemical engineering from Jadhavpur University, West Bengal, India; an MS degree in chemical engineering from IIT, Kanpur, India and a PhD in chemical engineering from Ghent University, Belgium. He has vast experience in developing petroleum refining processes including modeling and simulation, pilot-plant studies and scale up.

WGLNetwork.com

Facebook/WGLNetwork

Join Women's Global Leadership Network on LinkedIn

Registration Now Open

2013 WOMEN'S GLOBALLEADERSHIP CONFERENCE IN ENERGY & TECHNOLOGY HOUSTON™

OCTOBER 29–30, 2013 | HYATT REGENCY HOUSTON | HOUSTON, TEXAS

You're Invited: Help Celebrate Our Tenth WGLC

Gulf Publishing Company's tenth Women's Global Leadership Conference in Energy and Technology (WGLC) is taking place this October! It's an exciting time to be a woman in the oil and gas industry as new discoveries and technological innovations create global opportunities. We'd like to invite you to attend the 2013 WGLC conference **29–30 October, 2013** and take part in this special event.

As one of the largest women's events in the industry, and the only one that focuses on discussing key environmental, economic and professional development issues in oil and gas, this is the perfect forum to network with your peers and exchange valuable ideas and experiences. Targeted specifically to matters of female leadership, WGLC strives to provide meaningful discussion on all aspects of responsible stewardship ranging from energy security and geopolitics to personal career development. We hope to see you there.

2013 WGLC Delegate Fees

Registration Type	Regular Admission
Single Attendee	$880
Team of Two	$1,595
Small Group of 5	$3,300
Leadership Pack of 10	$6,325

For more information about the event or participating, please contact Melissa Smith at Melissa.Smith@GulfPub.com or +1 (713) 520-4475.

For group delegate passes, please contact Gwen Hood at Gwen.Hood@GulfPub.com or +1 (713) 520-4402.

Special Supplement to
HYDROCARBON PROCESSING®

ENGINEERING AND CONSTRUCTION

2013

CORPORATE PROFILES

COVER PHOTO

Photo courtesy of ConocoPhillips.

STREAMLINE PLANT INFORMATION FOR BROWNFIELD ASSETS

R. STANDISH and E. BOTTERILL, Intergraph Corp., Siriusdreef, The Netherlands

Brownfield assets suffer from handling large volumes of unstructured information. Often, such projects do not have up-to-date/as-maintained 3D models, data warehouses or integrated operation systems, along with high-quality structured information. The "as-exists" information is difficult to manage. With limited engineering, administrative and IT personnel on a plant site, organizing and keeping track of this information is a significant problem, particularly when the plant is subject to a continual cycle of updates, revamps, turnarounds and maintenance changes.

Unstructured information comes in such a wide variety of forms and file formats. **Result:** No one technology can solve all of the issues associated with the identification, classification, organization, storage and visualization of the plant-site information. Recent breakthroughs in information management (IM) and high-definition surveying (HDS) may change management of brownfield information, thus leading to higher productivity, safety and operational excellence. By pragmatically combining several of these technologies (IM and HDS), engineers can successful navigate, organize and structure disjointed asset data.

Object relationship and metadata. Many organizations have implemented document management systems (DMSs) that are at a corporate level, or now, in the "public cloud." These have been successful where there is structure in the content, and processes are applied to ensure that the information remains under control. Such projects rely heavily on metadata hierarchy to characterize the content via an index. However, there are drawbacks:

- It is more natural for users to navigate by relation as they understand the real world of physical objects and how they relate to each other. Then, the user can apply the discovered information that otherwise would not normally be available.
- The relationships between objects provide the context or meaning. Filtering out a particular relationship type is a powerful way to provide role-based views of the information.
- Hierarchical views, such as the traditional cabinet > folder > file paradigm of DMS, lead the user to suspect that documents will be in the lowest levels of the hierarchy. In an object/relational model, they can reside anywhere. This is a very useful aid to change impact analysis—if this is changed, what else will be impacted?

Find what you need by what you know. An example of a DMS is shown in FIG. 1. The user has set up folders that include supplier and vendor data. Under the vendor data folder, there are various vendor names, and under these files, the user has created folders representing purchase orders, against which the equipment was acquired or purchased.

This may be optimal for the person who originally set up the hierarchy. However, how would other users be able to traverse this and locate the documents they need? Would a safety engineer know to look for an accident report in a folder named to represent a purchase order?

Optical character recognition. For a number of years, optical character recognition (OCR) has been available. Primarily associated with the image/fax/document scanning industry, OCR was designed to recognize text from raster (pixel-based) images and turn it into computer-sensible/readable text. Of course, OCR has become more sophisticated. Not only has its recognition accuracy increased to 95%; but many vendors provide utilities to search for patterns and structure inside the raster image, such as zonal OCR, e.g., "search in the bottom right-hand corner of a drawing and extract title-block information."

Smart converters. With schematic drawings, P&IDs in particular, locating text within the drawing that matches patterns is important. As shown in FIG. 2, text may not be a single contiguous readable string to recognize the tag pattern. It often lies within and around symbology that represents objects to be identified and captured. It has to be linked together and examined to determine the true tag number pattern, which means there needs to be a rule base that is configurable to match these patterns.

Full-text retrieval. Besides all of the patterns used by OCR and smart converters in identifying unstructured information, there are also many useful words and phrases within the documents to be used by a search. Most people are familiar with this Google-like searching. For example, you could find all the documents with the text phrase "operating procedure" listed. However, as with the other techniques described in this article,

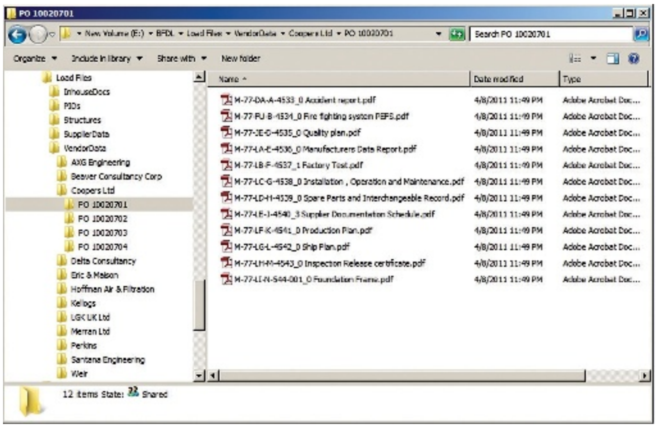


FIG. 1. Screen shot of a document hierarchy.[1]

this on its own is not the complete search answer. When you do such a search and it returns hundreds of pages of results, does anyone look at every page? This is usually not the case. Most people will refine the search more accurately, adding more or different search terms. The end result is full-text retrieval (FTR), in conjunction with the other classification technologies as outlined earlier.

Information aliasing. Expecting all parties to correctly follow your naming and numbering system for the life of the asset is unrealistic. For example, you may have required PA-4401, but instead, you found such variations as PA 4401 or PA/4401. It becomes more complex when there are primary and standby equipment with prefixes or suffixes added to the tag name, such as PA-4401-A or PA-4401-A/B/C. The same applies to drawings. Clearly, with these situations, the ability to support all naming and numbering patterns and provide aliases, such that each individual can locate the same information by their coding system, would be a great advantage.

Thumbnailing. Human vision can spot patterns, shapes and definition at an astonishing rate. An engineer is often able to identify a drawing from a small-scale pictorial representation quicker than trying to determine the correct drawing from a list of metadata. Modern printing and visualization software can provide accurate and efficiently rendered representations of the original drawing/document/image for quick display, as shown in FIG. 3.

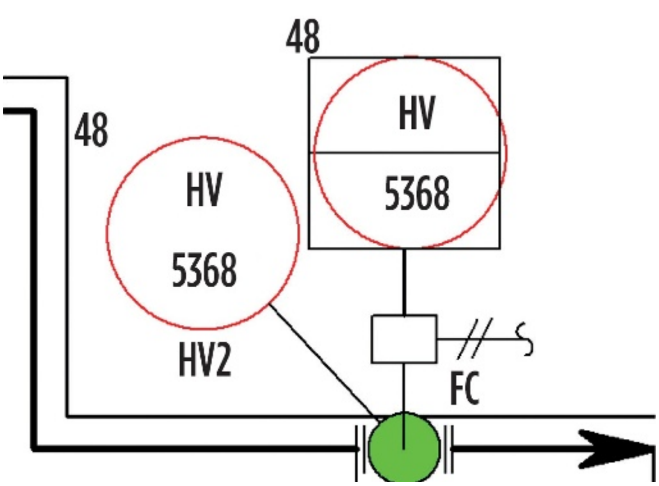


FIG. 2. Smart conversion of a flow controller.

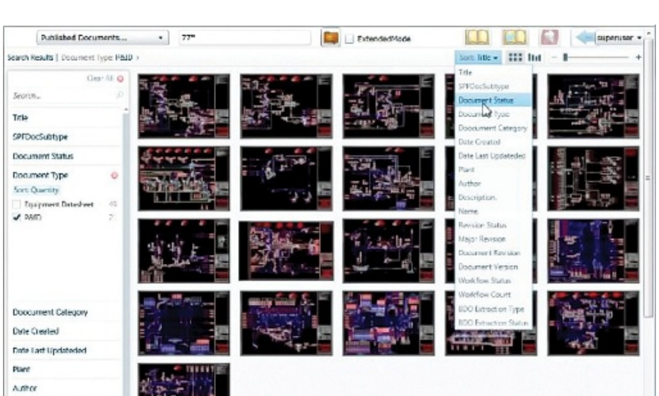


FIG. 3. An example of thumbnailing P&IDs.

High-definition surveying. When it is not cost-effective to use 3D CAD to model an existing plant (reverse engineering) for information navigation and review purposes, modern HDS systems can be used to rapidly integrate 3D point clouds and high-resolution 360° images into a single interactive display. As shown in FIG. 4, the user can quickly orient/rotate within the plant from this realistic rendering, and can traverse from one scan location to another. This gives the impression of walking through the plant from point to point. With this full 3D point cloud available, the remote user can study the actual plant and take measurements that are useful for plant reengineering studies, without having the cost and hazard of being at the plant.

When linked via "hotspots" to the unstructured information resource, this can provide a photo-realistic comparison between the "as-built" condition of the plant and the "as-exists" condition of the design basis. This is useful in design studies, safety and emergency planning, and other situations where cross-referenced information would be invaluable.

Crawlers. The Internet would not be the Internet without "crawlers" continuously reading, indexing and re-reading every updated page on the web. There is no way that you could keep up with the task and volume of information if you were to manually update and index your websites. The same applies to your unstructured information stores. Of course, you could manually do a "check-in" for each document, or, every once in a while, bulk-load a new batch of documents. It is more productive to borrow a practice from the Internet and continuously crawl the content, searching for new or updated unstructured information.

Workflow control. Your business runs on processes and procedures. Workflow control systems have the ability to enforce and automate the process of delivery, execution and notification. Not only can you do without the need to look for new things to do as they are delivered to you, but all of the capabilities that

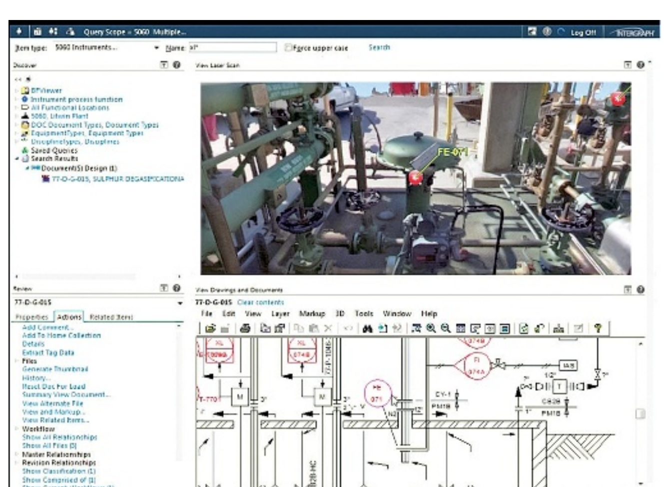


FIG. 4. Screen shot of high-definition survey to reengineer 3D model of existing plant.

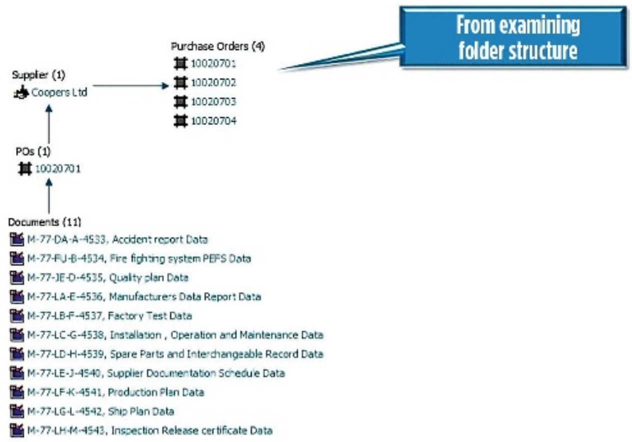


FIG. 5. A reading of the folder structure to create objects and the relationships between them.

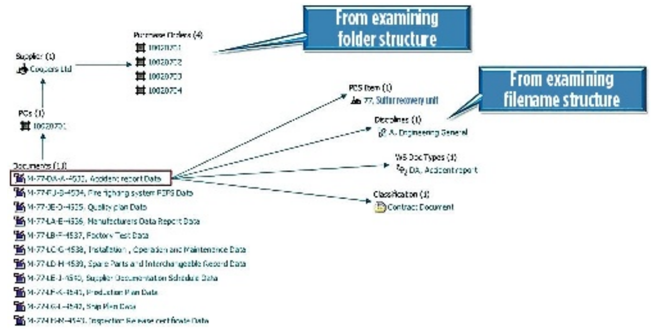


FIG. 6. A crawler reading filename structure to create additional objects and relationships.

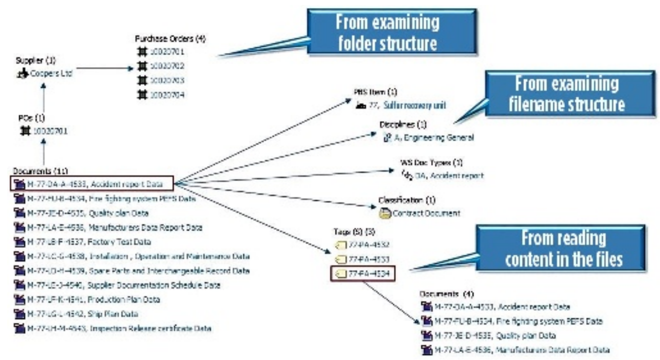


FIG. 7. Objects and relationships created from data-matching.

have been outlined earlier can be orchestrated to rapidly gather new unstructured information and to process it accordingly.

Web portals. Because of their ubiquitous application, web portals are undervalued. They have the ability to create "windows" into a number of different information systems/sources from a single Web page, and to have each of these windows updated and be triggered by actions from other windows on the same page. Users can also read and review content from these applications without necessarily having the originating application on their computer systems. In addition, users can browse the information remotely over the Internet. These are all beneficial traits in the context of unstructured information.

Utilizing organized 'unstructured information.' As shown in FIG. 5, the crawler could read the folder structure to create objects and relationships. A user could now search for a document and then navigate to the related purchase order, and, from there, navigate to the supplier to see how many purchase orders were in favor of the supplier, as well as the documents related to those purchase order objects, and so on.

In FIG. 6, the crawler has also read the filename and has decoded the constituent parts against a predefined pattern. Such codification is typical for document/drawing naming and numbering systems. You now can see that "77" represents the sulfur recovery unit as a plant structure item, A represents the discipline, DA represents the document type, and so on.

In FIG. 7, the document has been read to extract data-matching patterns—in this case, tag numbers. As indicated earlier, the appropriate reader used is determined by the file type, such as OCR for image. Here you can see tag numbers, e.g., "77-PA-4352" were extracted from the document "M-77-DA-A-4533, Accident Report Data," and relationships were established between the document and the tags.

This document-to-tag linkage is recognized as most important by owner operators and engineering, purchasing construction companies alike. Often, this is requested in the form of a document register at handover time. In FIG. 7, the tag 77-PA-4534 can be found in four documents. This reverse relationship is important for managing the impact of change. In addition, if the data regarding this one tag is changed, then it will potentially impact four documents. Such information is tremendously valuable once the entire document set is captured.

Path forward. The dirty secret about unstructured information is that this issue has existed for a long time even with the capability of computers to capture and maintain information/documents. Arguably, computers have exacerbated a problem that has been around a lot longer in the paper world. The answer may not be more rigors in your existing DMS, or an extension of your enterprise resource planning or computerized maintenance management system. Many companies have such systems already, and the issue still exists.

The real solution is to recognize that legacy information, which is often uncontrolled and unstructured, will always exist. Your ability to provide rapid and cost-effective decision support from previously disparate and unstructured information will be constrained by how well you deal with it. HP

NOTES

[1] Screen shot of a document hierarchy using Microsoft Windows Explorer.

RICK STANDISH is vice president of global business development at Intergraph PP&M. He is responsible for the strategic product direction of Intergraph's Information Management Portfolio including document management and control, data warehousing (SmartPlant Foundation), brownfield data acquisition (SmartPlant Fusion) and design tool integration (SmartPlant Enterprise). Mr. Standish holds a BEng degree in electronic and manufacturing engineering from Loughborough University.

EWAN BOTTERILL, Intergraph's vice president of business development global accounts, has worked for Intergraph for more than 10 years in various positions, including vice president for Greater China business development. Before joining Intergraph, he worked as an application engineer and as a sales manager for several design and technology companies.

AIR LIQUIDE GLOBAL E&C SOLUTIONS

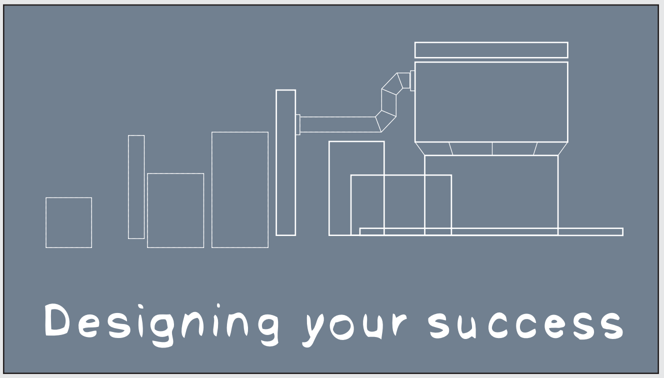
Air Liquide Global E&C Solutions is a technology partner of choice for the design, engineering and construction of leading-edge processing facilities and related infrastructures worldwide for the production of industrial gases, clean conversion projects, gas cleaning facilities, selected refinery projects as well as facilities for the production of oleochemicals and polymers.

We enable our customers to optimize the use of the planet's natural resources in order to provide clean and sustainable energy thanks to our people and their capability to innovate constantly. Through cutting-edge innovation applied to our proprietary technology portfolio bundled under the Lurgi, Cryogenics and Zimmer brands, we contribute to the transformation of the energy industry and help to preserve and protect the atmosphere of our planet.

Looking back on decades of operational expertise within the world leader in gases for industry, health and the environment, we develop creative, safe, reliable and competitive solutions for our customers, proposing a sustainable worldwide offer of best-in-class plants in a dynamically changing market place.

Offering a full spectrum of solutions and drawing on our expertise in engineering, procurement and construction through to manufacturing, technology development and Feed studies, Air Liquide Global E&C Solutions will meet the challenges of the most complex projects, maintaining control over reliability and delivering on performance guarantees while meeting the highest safety standards.

As part of our global portfolio, based on syngas, hydrogen production and clean conversion technologies for fuels or chemicals, the Lurgi technology portfolio offers innovative solutions that allow the operation of environmentally compatible plants with clean and energy efficient production processes. Our technological leadership is based on proprietary and exclusively licensed technologies which aim at converting all carbon energy resources (oil, coal, natural gas, biomass, etc.) into clean products.



Your success will be built on our global capabilities organized in a network of 15 Engineering centers, three manufacturing workshops as well as sales and business offices and representatives spread around virtually all continents with a strong local footprint to serve our clients.

With over 1600 patents to our credit, we foster innovation to continuously enhance our technological, engineering and project execution skill-set. Our technology focus and long term view of people development allow us the consequent nurturing and development of competencies of our 4,000 experts.

Global capabilities and local solutions, creativity and agility together with a long term vision are key to our capacity to make your EPC projects all over the world a success.

Air Liquide Global E&C Solutions. Designing your success.

Cryogenics
Lurgi
Zimmer

AIR LIQUIDE
GLOBAL E&C SOLUTIONS

THE TECHNOLOGY FOR THE WORLD'S HYDROCARBON PROCESSING INDUSTRY

In this demanding global industry, AVEVA technology is an essential enabler of the design, construction and operation of ever more complex and challenging projects.

BIGGER—FASTER—MORE EFFICIENTLY

For the design and construction of processing facilities, AVEVA offers advanced, powerful and fully integrated solutions that are mandated the world over by Owner Operators and used by their EPCs.

1. **The key to Lean Construction.** With the release of the ground-breaking new AVEVA Everything3D™ (AVEVA E3D™), AVEVA at last provides the key to unlock Lean methodologies in capital engineering projects. With the most powerful 3D design capabilities available and integrating as-built 3D laser scans with as-designed models in the same environment, AVEVA E3D closes the feedback loop between design, fabrication and construction.

2. **The key to integrated project execution.** AVEVA technology enables engineering and design data from all project disciplines to be created, controlled and shared in a single common project model. No other solution enables such extensive, flexible and controlled collaboration across the entire project team. And no other solution extends this capability reliably across a globally distributed network, to efficiently tackle the most complex engineering projects. Only with AVEVA technology can engineers and designers from opposite sides of the globe collaborate on a single project in real time. And only AVEVA technology enables integration of 3D plant design with project and workface planning to achieve effective 4D construction.

3. **The key to design integration.** A committed champion of information sharing and the removal of barriers to data exchange, AVEVA continues to integrate the entire design process. Its new AVEVA Bocad™ applications integrate the world's most powerful structural steel design technology with 3D plant design, increasing efficiency and capability and eliminating costly design errors.

4. **The key to efficient upgrades.** Upgrading in-service oil & gas facilities is a hazardous process in which every day's lost production counts. AVEVA's advanced laser scanning applications reduce site working and create accurate, finely detailed 3D representations of the as-operating plant, enabling the design of 'fit first time' upgrades that minimise asset downtime.

SAFER—LONGER—MORE PRODUCTIVE

Getting your assets running earlier and keeping them running safely, at optimum output, for longer, demands the management and use of a huge quantity of information. AVEVA's advanced Information Management solutions support many of the industry's most efficient producers and processors.

5. **The key to rapid ROI.** Achieving early start-up and rapid ramp-up to nameplate capacity requires close collaboration between Owner Operator and EPC from the outset. AVEVA's powerful Information Management technologies enable the efficient management, control and use of asset information, in a common environment, from FEED throughout design, construction, commissioning and operations.

6. **The key to safe, compliant operation.** Whether for the control of Work Orders, efficient shift handover, or the monitoring of plant KPIs and regulatory compliance, AVEVA provides the most powerful Asset Management solutions available. AVEVA's unique Digital Plant concept harnesses all asset information. It supports the most efficient operational processes, plant performance optimisation, troubleshooting and incident management.

An industry partner. AVEVA's deep understanding of its customers' needs, and the reliability of its software, has made it a trusted partner to the industry's leading operators and contractors. AVEVA works closely with its customers to maximise the value they gain from its products and to ensure that its technology continues to meet real-world needs. AVEVA technology inherently protects its users' information assets while offering ever more powerful ways to exploit their value.

Global solutions, local support. AVEVA's world-leading solutions are delivered and supported through a global network of offices, staffed by knowledgeable professionals. Technical expertise combined with industry knowledge enables rapid implementation of solutions which support best-practice working methods. Flexible configuration and expansion, backed up by expert local support, ensures that an AVEVA deployment will provide competitive advantage as a customer's business needs evolve.

AVEVA™

CONTINUAL PROGRESSION

CONTACT INFORMATION

10350 Richmond Avenue, Suite 400
Houston, TX 77042, USA
Tel: +1 713 977 1225
Fax: +1 713 977 1231
sales.us@aveva.com
www.aveva.com


Four
uest Energy
3001
Venting is Normal

NITROGEN SERVICES: ACCELERATED COOL DOWN (ACD)

WHY NITROGEN?

Nitrogen (N_2) is a colorless, odorless, and tasteless gas that makes up 78.09% (by volume) of the air we breathe. It has several properties that make it an ideal tool for use in the energy industry. Nitrogen is inert, nonreactive, nontoxic, and noncorrosive. It is also non-flammable: a nitrogen atmosphere will not support combustion.

Nitrogen is produced in large volumes in both gas and liquid form by cryogenic distillation; smaller volumes may be produced as a gas by pressure swing adsorption (PSA) or diffusion separation processes (permeation through specially designed membrane). Cryogenic processes can produce very pure nitrogen. When produced in cryogenic facilities, nitrogen is shipped in liquid form and transported in specialized insulated tank trucks suitable for handling the extreme low temperatures of nitrogen in its liquid state.

Refineries, petrochemical plants, and marine tankers use nitrogen to purge equipment, tanks, and pipelines of dangerous vapors and gases (for example, after completing a pipeline transfer operation or ending a production run) and to maintain an inert and protective atmosphere in tanks storing flammable liquids.

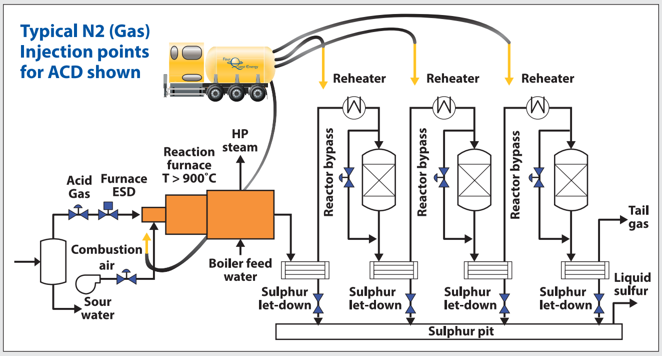
NITROGEN AS A COOLING AGENT

Another important application for nitrogen is to use it to cool and inert reactors filled with catalyst during the shutdown process for maintenance work. The inert environment created from nitrogen prevents any combustible reactions from occurring. Flowing cold nitrogen through a unit can also substantially reduce the amount of time it would take to cool the unit than if it were allowed to cool naturally.

Depending on the system design, nitrogen can be used for accelerated cool downs in either a gaseous or liquid phase ranging from –196°C to +400°C. During a once-through (gas-phased) cool down, nitrogen gas is pumped through a process system. As the gas moves though the reactor, it exchanges heat with any matter it comes into contact with, resulting in an accelerated cool down. When using nitrogen in its liquid phase, cryogenic nitrogen fluid is pumped into a recirculating process of the unit gas stream with a specially designed nozzle. The nitrogen is vaporized by the warm gas stream and forms mixed gas at a lower temperature. This cool gas mixture is used in the same manner as the gaseous cool down to accelerate the cooling of the reactor system. The liquid phase cool down, however, uses significantly less nitrogen due to the added cooling benefits of phase changing liquid to gas.

The nitrogen is transported to the site as a liquid via a nitrogen pump truck, equipped with a storage tank, diesel fired vaporizer, and cryogenic triplex pump. Additional nitrogen to complete the cool down is brought to site in specialized transport trucks. The liquid nitrogen can be pumped from the truck in its liquid form, or it may be converted to gaseous nitrogen, depending on the application. It then travels through injection piping into the process unit (hydrotreater, sulphur recovery unit, etc.) to be cooled.

The heat loss during a cool down can be determined using the following principle:



$$Q = m \times c\,(\Delta T)$$

Where:

Q	=	heat loss
m	=	mass of object to be cooled
c	=	specific heat capacity
ΔT	=	change in temperature (final-initial)

This calculation will define the overall heat transfer requirement from which the nitrogen consumption can be determined. Specific elements of the process unit, i.e., catalyst and metallurgy, each contribute to the total nitrogen requirement based on their individual masses and material properties. It is therefore important to compile as much information as possible regarding these parameters during the engineering assessment.

FOURQUEST NITROGEN SERVICES FOR ACD

Virtually any reaction vessel operating at extremely high temperatures can benefit from the thermally engineered accelerated cool down service provided by FourQuest Energy. The high-temperature vessels used in the energy industry for a wide range of processes and products can often take days or even weeks to cool down to manageable working temperatures. Proper engineering design optimizes the accelerated cool-down portion of the plant shutdown process, allowing our clients to better plan and execute shutdown activities beyond the time savings provided by the ACD. Put simply, shutdown time is significantly reduced so that maintenance work can start sooner in a safer environment and with lower overall shutdown expenses.

Four Quest Energy

CONTACT INFORMATION

4820 Railroad Street, Deer Park, Texas 77536
Office: +1 281-476-9249
info@fourquest.com
www.fourquest.com

HEURTEY PETROCHEM

Creating engineering solutions since 1953

HEURTEY PETROCHEM, PROVIDING ENGINEERING AND FABRICATION SOLUTIONS SINCE 1953

Heurtey Petrochem is an international oil and gas Group providing Engineering, Procurement and Construction solutions to the downstream and upstream hydrocarbon industries.

In 2013 the Group is celebrating in 60 years of projects serving the petroleum industry, which have together made our company a leader with an extensive international presence.

Listed in the Paris stock exchange market Alternext of Euronext since 2006, Heurtey Petrochem is part of the IFP Energies Nouvelles Group and recently acquired 60% of the capital of Prosernat, also based in France, extending its expertise to the gas sector.

The Group develops its business through an extensive network of subsidiaries and local representatives strategically spread around the globe: France, USA, Brazil, Russia, India, Romania, South Africa, China, Korea, Malaysia, Saudi Arabia and UAE.

Benefiting from key markets dynamics and a strong growth of its activity, most of Heurtey Petrochem's foreign offices were launched these past five years and now successfully provide a complete follow-up of turnkey projects.

A FULL RANGE OF FURNACES FOR THE REFINING, PETROCHEMICAL, HYDROGEN AND GAS INDUSTRIES

Downstream. Refining is the Group's main sector of activity, reflecting its experience of 60 years in this segment and its ability to provide a comprehensive range of refining furnaces for all hydrocarbon treatment and conversion units.

Heurtey Petrochem has also built up extensive expertise in the petrochemicals area, has developed its own design tools and collaborates with the best licensors as part of major projects. The Group offers a wide range of furnaces for ethylene and aromatics production units.

The Hydrogen sector is also a major field of activity, Heurtey Petrochem has, for several years, been working in cooperation with some of the largest licensors for steam-reforming technologies. The Group is also active in the area of waste-heat recovery units, with integration of an additional stage in the steam-generation process, notably for CO Boilers and for convection sections downstream from Steam Methane Reforming furnaces.

Upstream. In this market segment, Prosernat provides technologies and facilities for field treatment of natural gas in the following areas: Phase separation, Gas sweetening, Gas dehydration, Sulfur plants, NGL recovery, Modular units, Gas compression and CO2 Capture.

CONTROLLING ALL PHASES OF THE PROJECT EXECUTION PROCESS

Engineering and project management. Heurtey Petrochem has the full range of process engineering, mechanical design, detailed engineering, procurement management, fabrication and construction resources required for conducting projects.

Basic project engineering is performed mainly at the Group head office in France, or in the United States. The detailed engineering, procurement and construction phases of project implementation are optimized thanks to the expertise available in the relevant subsidiaries.

During the fall of 2012, Heurtey Petrochem successfully completed its first complete hydrogen production unit for the Pancevo refinery in Serbia (77000 Nm^3/h, Haldor Topsøe technology).

Fabrication and modularization. Heurtey Petrochem runs its own manufacturing workshop in Romania, strategically located near all transportation means. With a monthly production capacity of 300 tons of casing and steelwork, the Heurtey Petrochem Manufacturing workshop enables the Group to control each step in the furnace-fabrication process in terms of quality, costs and deadlines. Its extensive network of partners also enable the Group to offer a wide range of fabrication solutions worldwide.

Heurtey Petrochem provides a complete range of services, from detailed engineering to turnkey supply: feasibility studies, process and mechanical studies, detailed engineering, revamping and modernization, site construction, commissioning and start-up, training, spare parts sourcing etc.

HEURTEY'S 2013 HALF YEAR RESULTS, BACKLOG AT ITS HIGHEST LEVEL

Heurtey Petrochem announced on August 8th, 2013 a consolidated turnover of € 204 million over the first semester 2013 (growing by 25%) and a record level firm backlog of € 452 million (€ 231 million of order intake), growing by 6% compared to December 31st, 2012. With its good performance over the semester, the Group is confident in its ability to achieve over the year a turnover on the high hand of the €360–€ 390 million range.

HEURTEY PETROCHEM

CONTACT INFORMATION

8 cours Louis Lumière, 94306 Vincennes Cedex - France
Phone : +33 (0) 1 41 93 80 00 — Fax : +33 (0) 1 41 93 80 01
Email : heurtey@heurtey.com
Website : www.heurtey.com

WE BUY AND SELL COMPLETE PROCESS PLANTS, PROCESS LINES AND EQUIPMENT

International Process Plants (IPP) is a self-funded global buyer and seller of surplus manufacturing facilities, process plants, industrial real estate, and individual equipment. IPP's business model serves clients in two ways:

1. IPP provides companies the opportunity to acquire existing assets at competitive prices and in a fraction of the lead time of building or buying new
2. for companies looking to divest assets that have become surplus to their needs, IPP serves as an outlet to convert those assets into funds quickly in a fiscally and environmentally conscious manner.

IPP, started over 35 years ago by 2 brothers who still lead the company, has grown to company-owned operations in 17 countries serving a client base of 160,000 companies in the worldwide oil & gas, power generation, petrochemical, chemical, pharmaceutical and food processing industries. One of the largest companies in this business, IPP's inventory includes over 100 complete plants, more than 30,000 individual pieces of process equipment and multiple industrial real estate development sites.

BUYING EXISTING PLANTS: A STRATEGIC WEAPON

Many of IPP's clients compete in mature industries where today's challenging economic climate demands creative solutions for business success. Given that IPP's plants cost a fraction of the capital needed to build new ones (savings typically reach 40–50%), buying an existing plant is the only economically viable option to satisfy plant needs for many of IPP's clients. IPP's clients also save valuable uptime as most plants can be relocated and ready for operation within 18 months or less compared to new plant construction that can take 3–4 years.

IPP's stock of complete plants includes several properties of interest to the hydrocarbon processing industry:

- IPP's Gasification Plant: 3, 000 metric tons / day
- LNG Peak Shaving Plant:
- 11 Refinery Facilities: 15,000–275,000 barrels / day
- 5 Hydrogen Plants: 1,580–25,000 NM^3 / hr (99.9%+ pure)
- Methanol Plant: 400 metric tons / day

Gasification allows oil refineries to convert residue waste into valuable commodities. A study by the National Energy Technology Laboratory (NETL) estimated that typical US refineries using gasification to produce these commodities could save up to $55,000 / day. The syngas produced by gasification could also be sold. Refineries buying an existing gasification plant can save up to 50% of the capital required to build new (US $500–$800MM). Coupled with operating cost savings and potential revenue from syngas sales, refineries can fund a significant number of incremental projects.

IPP's LNG plant can be used to store surplus natural gas for demand spikes, to liquefy associated gas in oil fields while reducing emissions and increasing pumping capacity, in industrial sites with no natural gas pipeline to delivery LNG and for vehicle fueling stations.

WHY BUY ASSETS FROM IPP?

Buying existing plants or equipment assets from IPP maximizes the impact your fixed capital budget can have by saving you 30–50% vs. buying new plus you'll get to market or back online faster. Plant clients tell us they are up and running in ½ the time it would have taken for a new plant. Equipment clients love reducing downtime when equipment fails because IPP delivers within days.

Complete Process Plants, Process Lines, Industrial Real Estate and Major Pieces of Process Equipment

SELLING ASSETS TO IPP

Unlike some companies in this business, IPP utilizes its own assets to purchase the entire plant site outright. With no 3rd parties involved, selling to IPP maximizes your plant's value and generates your cash infusion faster, funds you can immediately use. IPP also purchases idle and surplus processing lines or individual equipment pieces, turning those non-productive assets into instant funds as well.

Selling your plant to IPP avoids the carrying costs of an idled plant. IPP's clients have saved as much as $700,000 a year in utilities, insurance, taxes, etc. From a human capital perspective, IPP's clients don't have to supervise plant or equipment shut downs, provide security or sell assets piecemeal. IPP can help manage any needed dismantling or environmental remediation, eliminating further costs.

IPP is highly experienced in extracting multiple revenue streams from our plant acquisitions, often achieving 100% recycling of the plant's contents, a process that benefits the seller too. For example, one plant site included timber that IPP sold to a client in the paper business, value that was reflected in the plant's purchase price.

IPP
INTERNATIONAL PROCESS PLANTS
Your Partner for Complete Plant Solutions

CONTACT INFORMATION

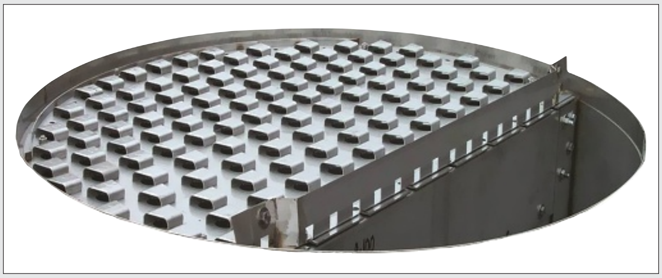
Address: 17A Marlen Drive, Hamilton, NJ 08691
Phone: +1-609-586-8004
Fax: +1-609-586-0002
sales@ippe.com
http://www.ippe.com

CELEBRATING A CENTURY OF INNOVATION IN MASS TRANSFER EQUIPMENT DESIGN

For many years, equipment designers searched for the ultimate mass transfer device—one that could be used for any application. Over time, they reached the conclusion that there is no single device for all applications; rather there is always a device that has distinct benefits and characteristics that make it the most suitable to meet the requirements for a particular application.

Koch-Glitsch put decades of practical knowledge, innovative thinking, and tremendous expertise into research and testing that led to the development of a comprehensive product portfolio. Along the way, Koch-Glitsch products set new standards and became the preferred technology in the industry.

INTALOX™ ULTRA™ random packing



PROVALVE™ tray

RANDOM PACKING

Random packing is frequently used as an inexpensive but very effective means to increase a tower's capacity and/or efficiency. The reduced height to diameter ratio of CASCADE MINI-RINGS™ random packing provided easy vapor passage resulting in lower pressure drop, additional vapor/liquid handling capability, and increased fouling resistance.

The improvements in capacity and efficiency in IMTP™ random packing made it the most widely used in the industry. Its low pressure drop reduced energy consumption in revamps, and the reduced foam generation made it a good choice for foaming systems.

As a fourth generation packing, the unique shape of INTALOX™ ULTRA™ random packing provides the highest effective surface utilization of any random packing available. It provides higher capacity and efficiency and lower pressure drop compared to previous generations of random packing.

STRUCTURED PACKING

Made of corrugated sheet metal, FLEXIPAC® structured packing was introduced to the market by Koch-Glitsch in the 1970s.

Conventional corrugated sheet metal structured packing accumulates liquid at the base of one element before it flows to the top of the next lower element. FLEXIPAC® HC™ structured packing was the first packing with edge modifications to the corrugation at the top and bottom of each packing layer that allows liquid to drain to the next lower element without excessive accumulation.

TRAYS

The FLEXITRAY™ valve tray has been the preferred tray technology in the industry for over 60 years. With proven performance in all liquid-vapor contacting applications, FLEXITRAY™ valve trays have been installed in tens of thousands of columns worldwide.

The SUPERFRAC™ tray has the highest combined capacity and efficiency of all single-pass cross-flow trays tested at FRI. It has been industry validated for more than 20 years in new tower design and revamping of existing trays. In new columns, SUPERFRAC™ trays can reduce vessel diameters, heights, or both. In existing columns, they can increase capacities, reduce utility costs, and improve separation.

With ULTRA-FRAC™ high performance trays, existing columns can be retrofitted, resulting in significant capacity increases from high to low liquid loaded systems.

The PROVALVE™ tray's unique valve design with no moving parts provides reliable unit operations in severe fouling applications such as coker fractionators, sour water strippers, and beer stripping columns.

SEVERE SERVICE GRID

With no horizontal surfaces and open structure, GLITSCH GRID™ and FLEXIGRID™ severe service grid packings do not allow solids to settle, which reduces the formation of coke on the packing surface.

PROFLUX™ severe service grid is a rugged assembly of rods welded to sturdy corrugated sheets. This construction combines the efficiency and de-entrainment performance of a conventional structured packing with the ruggedness and fouling resistance of a conventional grid packing that outperforms both. Greater capacity compared to conventional grid or structured packing allows increased flow rates in existing towers.

Understanding market demands and envisioning future requirements constantly drives Koch-Glitsch's search for improved column internals and services that will help operators address challanges with energy efficiency, product purity, operation reliability, and the environmental impact of processing operations.

CONTACT INFORMATION

4111 East 37th Street North
Wichita, KS 67220
Phone: 316-828-5110
Fax: 316-828-7985
www.koch-glitsch.com

THYSSENKRUPP UHDE—ENGINEERING WITH IDEAS

With more than 2,000 plants to its credit, ThyssenKrupp Uhde is one of the world's leading engineering companies in the design and construction of chemical, refining and other industrial plants. We have subsidiaries and associates in all four corners of the globe. This worldwide network with over 5,900 employees is active in a number of different fields: fertilisers, electrolysis, gas technologies, oil, coal and residue gasification, refining technologies, organic intermediates, polymers and synthetic fibres as well as coke plant and high-pressure technologies. We offer our customers not only cost-effective high-tech solutions in industrial plant construction and the entire range of services associated with an EPC contractor but also comprehensive service packages for the entire life cycle of their plants.

HIGH-TECH ENGINEERING. WORLDWIDE. CUSTOMISED.

In the international plant construction industry, it is essential for companies to provide first-class processes which guarantee industrial plant operators an optimum in quality and operational economy.

Not only does ThyssenKrupp Uhde promise to meet these demands, we also orientate our strategy towards expanding our presence on international markets and to adopting innovative new processes to boost our existing technology portfolio.

In addition, ThyssenKrupp Uhde has also invested in yet more on-site competence in the form of subsidiaries and international partnerships which enable us to keep up with the ever-increasing number of customers on the world market. ThyssenKrupp Uhde's special brand of corporate performance and efficiency is based on the fact that the company can offer complete production lines in many technological fields and has consequently accumulated a wealth of experience in dealing with process-related tasks.

Furthermore, our various central divisions enable us to provide additional services which include plant location selection, the arrangement of financing schemes, negotiations with authorities, plant management, maintenance, safety analyses, safety technology, training of operating personnel and project management.

Our affiliates and associates abroad work to the same quality standards as the head office in Dortmund. This ensures that any services connected with the name Uhde are synonymous with excellent quality irrespective of location.

In 2011, ThyssenKrupp Uhde successfully commissioned the 3,500 tonne-per-day urea plant built for Yara at their site in Sluiskil (Netherlands).

TECHNOLOGIES

We are constantly improving and expanding our process portfolio with the aim of providing our customers with even more cost-effective and, at the same time, safe, environment-friendly plants. Proprietary developments, cooperation with external partners and the acquisition of technologies all play a key role in achieving this aim.

We thus guarantee at all times up-to-date and technically optimum solutions for our customers, combined with comprehensive, state-of-the-art services for the entire industrial plant construction business.

ThyssenKrupp

ThyssenKrupp Uhde

CONTACT INFORMATION

Friedrich-Uhde-Strasse 15, 44141 Dortmund, Germany
Tel.: +49 231 547 0
Fax: +49 231 547 3032
info.uhde@thyssenkrupp.com
www.thyssenkrupp-uhde.de

H. POORKAR, M. AL-ZAHRANI and
H. AL-SHBRAIN, TASNEE, Jubail, Saudi Arabia;
and **K. SEFIDROU,** Rockwell Automation,
Houston, Texas

Advanced process control improves operation of a polypropylene unit

Saudi Polyolefins Co. (TASNEE Petrochemicals) operates two world-scale polypropylene (PP) units in Jubail, Saudi Arabia. The units include polymerization and extrusion sections. The polymerization section for each line has two reactors operating in parallel mode for homopolymer production. Hydrogen is added continuously in a fixed ratio to the propylene feed to control the PP's molecular weight. The molecular weight is expressed as the melt flowrate (MFR). Each reactor train has its own degassing section, intermediate product storage and extruder. The total capacity of the plant is 720,000 tpy.

In early 2009, TASNEE decided to deploy an advanced process control (APC) solution for its recently debottlenecked PP plant in Jubail. Selected APC vendors were supplied with an extensive set of plant operating data, allowing them to perform a preliminary benefits analysis. The project scope included a reactor composition controller and virtual online analyzer for MFR and xylene solubility (XS). Quality control information and automated transition records for the four reactors were also included.

Benefit study. To estimate the potential benefits related to implementing an APC solution, a detailed benefits study was carried out in 2009 for the two PP lines. The benefits study methodology was agreed to by both TASNEE and the APC vendor. The benefits estimates were based on actual process data and the APC vendor's experience with similar projects. The benefits were seen in three key areas: Measurable improvements in process stability; the ability to accurately predict key process variables (such as MFR and XS), which can be difficult to measure online; and improvements in process operation consistency.

Another key benefit discovered is that APC will push the process closer to its constraints without violating them, thereby increasing the plant production rate. This benefits study focused on three primary benefit areas; including increased production, reduced off-spec product and lower operating costs.

Results of the initial benefits study showed that less than six months return on investment (ROI) was achievable. In other words, production increased by over 2%, while the process capability index (C_{pk}) soared by over 50%.

APC technology selection. The TASNEE project team invested significant time and effort for vendor and technology selection. Factors such as technology capability, polymer experience, commitment to success, and cost were considered in the selection process. The vendor selection process took approximately six months. Each vendor was given the opportunity to present its benefits analysis and its technical solution. The TASNEE team paid special attention to the APC solution that provided remedies for chronic operational issues.

Project implementation. The project kickoff meeting was held shortly after the contract was finalized and initial preparations were completed. The initial preparation work involved things like installing the APC server, interfacing it with the distributed control system (DCS) and then making the subsequent DCS configuration changes.

TASNEE assigned a project team made up of operations experts, process engineers, process control gurus, DCS experts and IT support. The project team was overseen by a dedicated project manager who also acted as the technical lead. The project team was involved in all aspects of the implementation. Within the first three months, the majority of the APC solution was in place and commissioned. Improvements in process stability and product quality were noted shortly after commissioning began.

To improve reactor stability, the reactor pressure controller was commissioned, along with the polymer production rate controller. For the next four months, the project team focused on fine tuning the controller and installing and testing of the automated transitions. The implementation was completed seven months after the kickoff meeting. The performance test and audit took an additional three months. The project exceeded all performance criteria outlined in the project objectives.

Initial challenges. This polymer APC project had to overcome some difficult issues related to the process while remaining compatible with overall objectives. One issue was that the process makes multiple products with varying catalyst activities (due to fluctuating hydrogen concentration) in the reaction loop. This introduced a high degree of non-linearity in the gains and the dynamics.

Another issue was that some of the controlled variables were open loop unstable. This required that the controller respond in a quick and sufficient manner so as not to introduce further instabilities. The team had to be aware that the process was sensitive to impurities in the feedstock and to those generated internally. The impurities tended to affect the catalyst activity and thus the response of the process. Further, the APC solution had to be programmed in a way to recognize that the steady-state location of process variables for a product will vary from run to run.

Another red flag to consider was that some of the final product properties that are important to the end user are not characterized by the parameters measured by the lab or controlled in the process. Clear lines of responsibility also had to be established, so that appropriate technology transfer to the operations and technical groups for maintenance and support could be undertaken. Fortunately, the analyzer performance issues and lab result variability were flagged early in the project and the project team was able to offer solid technical solutions.

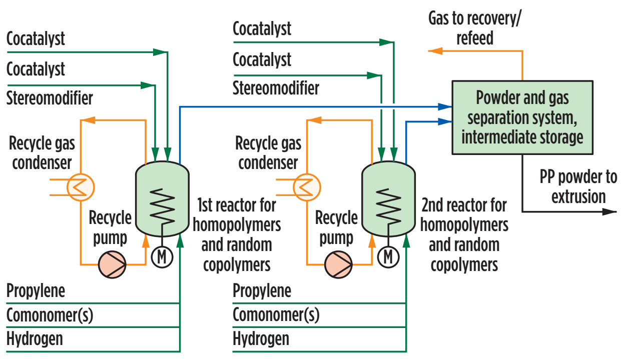


FIG. 1. Typical process flow diagram with APC implementation. Source: CB&I

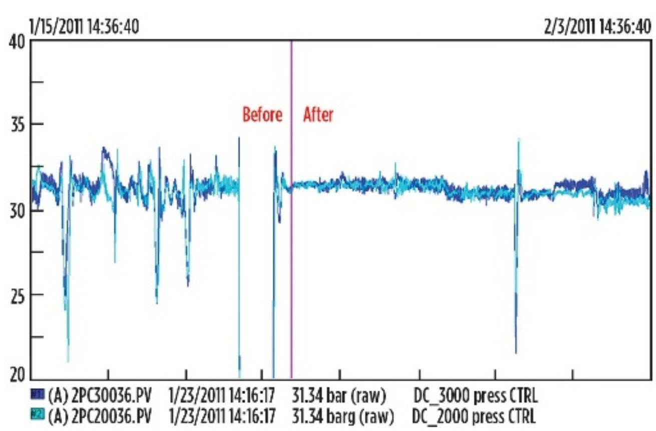


FIG. 2. Reactor's pressure control before and after APC implementation.

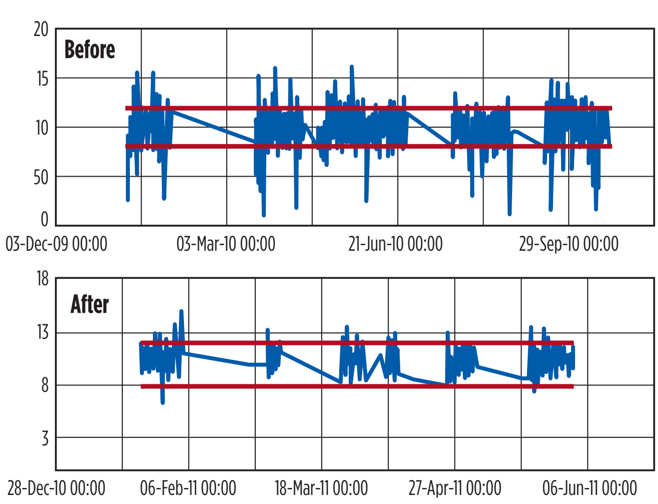


FIG. 3. Quality control performance before and after APC implementation.

Results achieved. The main project objectives were to increase production rate, improve process stability (reduce off-spec) and improve product quality or C_{pk}. APC solutions for polymers achieve these objectives by implementing a nonlinear multivariable control solution (**FIG. 1**). An important aspect of reactor stability is the ability to control critical process parameters such as reactor pressure and reactor temperature. **FIG. 2** illustrates the pressure control improvement for one of the polymer lines.

TABLE 1 and **TABLE 2** show the improvements in process stability relative to the baseline during the three month performance period. **TABLE 3** illustrates the project's performance guarantee objectives and **TABLE 4** provides the actual results for the three month performance period.

In addition to reactor stability, improvement in the product quality was an important objective. The product quality im-

TABLE 1. Improved process stability for Train 1

	Standard deviation		
Train 1	**Baseline**	**After APC**	**Improvement**
Reactor 1 pressure	0.74	0.31	58%
Reactor 2 pressure	0.68	0.27	61%
Reactor 1 temperature	0.93	0.47	50%
Reactor 2 temperature	0.93	0.68	27%

TABLE 2. Improved process stability for Train 2

	Standard deviation		
Train 2	**Baseline**	**After APC**	**Improvement**
Reactor 3 pressure	0.63	0.25	60%
Reactor 4 pressure	0.71	0.2	72%
Reactor 3 temperature	0.52	0.34	35%
Reactor 4 temperature	0.66	0.34	48%

TABLE 3. Performance guarantee

Performance guarantee terms	**Train 1**		**Train 2**	
	RX 1	**RX 2**	**RX 3**	**RX 4**
Production increase	2%	2.50%	2.50%	2.50%
Quality improvement for MFR-C_{pk}	50%	50%	50%	50%

TABLE 4. Performance achievement

	Train 1		**Train 2**	
Results achieved	**RX 1**	**RX 2**	**RX 3**	**RX 4**
Production increase	3.50%	2.80%	2.80%	2.85%
Quality improvement for MFR-C_{pk}	> 50%	> 50%	> 50%	> 50%
Quality improvement for XS-C_{pk}	> 50%	> 50%	> 50%	> 50%

provement was measured by the C_{pk} number. This is a statistical measure of the plant's capability to operate between product limits and near the designated product target. For this undertaking, C_{pk} for MFR was a project objective. In the end, APC implementation resulted in all products exceeding the C_{pk} objective.

Improvements in other product properties, including XS, were also observed. **FIG. 3** shows the quality improvement for one of the product grades.

The APC solution not only improved process stability, increased production and offered better quality control, it also improved product transitions by automating the changes. The product transitions are now faster and more consistent. Transition times were reduced from 50% to 80% depending on the direction of the transitions and the magnitude of the change. This translates to making less off-spec material due to product transitions. A typical MFR transition is illustrated in **FIG. 4.**

From then to now. The project was completed seven months after the formal kickoff meeting. The total benefits achieved shows that the ROI was less than three months. A few important factors that contributed to the project's success included the initial work on the project base line, scope definition and preparation. The essential elements are:

1. Close collaboration between the TASNEE team and the APC vendor team
2. Ongoing training
3. Technology transfer to the TASNEE team
4. Effective project management.

The application performance was monitored during the evaluation period (three months after commissioning) and the post-audit report shows that all performance criteria were exceeded.

The installed APC solution has been online for over 24 months, with an average utilization of over 93%. The nonlinear model predictive control solution is maintained by TASNEE personnel so that the expected results will continue to be delivered. Before implementation of this solution, TASNEE was focused on improving its lab quality control performance. Implementing this APC solution has allowed TASNEE to reduce the number of lab samples required. This has reduced the lab workload, resulting in more consistent results and improved product quality control. HP

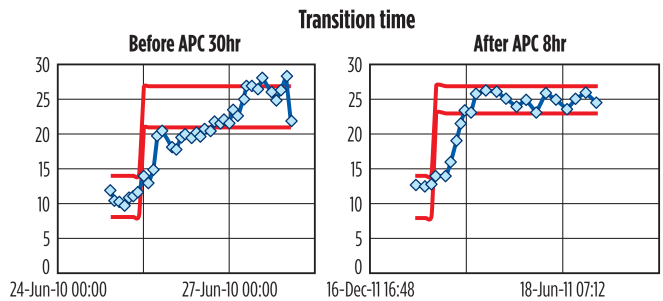


FIG. 4. Product transitions before and after APC implementation.

HYDROCARBON PROCESSING®

Bret Ronk, Publisher
Phone: +1 (713) 529-4301
Fax: +1 (713) 520-4433
E-mail: Bret.Ronk@GulfPub.com
www.HydrocarbonProcessing.com

SALES OFFICES—NORTH AMERICA

IL, LA, MO, OK, TX
Josh Mayer
Phone: +1 (972) 816-6745, Fax: +1 (972) 767-4442
E-mail: Josh.Mayer@GulfPub.com

AK, AL, AR, AZ, CA, CO, FL, GA, HI, IA, ID, IN, KS, KY, MI, MN, MS, MT, ND, NE, NM, NV, OR, SD, TN, TX, UT, WA, WI, WY, WESTERN CANADA
Diana Smith
Phone/Fax: +1 (713) 520-4449
Mobile: +1 (713) 670-6138
E-mail: Diana.Smith@GulfPub.com

CT, DC, DE, MA, MD, ME, NC, NH, NJ, NY, OH, PA, RI, SC, VA, VT, WV, EASTERN CANADA
Merrie Lynch
Phone: +1 (617) 357-8190, Fax: +1 (617) 357-8194
Mobile: +1 (617) 594-4943
E-mail: Merrie.Lynch@GulfPub.com

DATA PRODUCTS

Lee Nichols
Phone: +1 (713) 525-4626, Fax: +1 (713) 520-4433
E-mail: Lee.Nichols@GulfPub.com

SALES OFFICES—EUROPE

FRANCE, GREECE, NORTH AFRICA, MIDDLE EAST, SPAIN, PORTUGAL, SOUTHERN BELGIUM, LUXEMBOURG, SWITZERLAND, GERMANY, AUSTRIA, TURKEY
Catherine Watkins
Tél.: +33 (0)1 30 47 92 51
Fax: +33 (0)1 30 47 92 40
E-mail: Watkins@GulfPub.com

ITALY, EASTERN EUROPE
Fabio Potestá
Mediapoint & Communications SRL
Phone: +39 (010) 570-4948
Fax: +39 (010) 553-0088
E-mail: Fabio.Potesta@GulfPub.com

RUSSIA/FSU
Lilia Fedotova
Anik International & Co. Ltd.
Phone: +7 (495) 628-10-333
E-mail: Lilia.Fedotova@GulfPub.com

UNITED KINGDOM/SCANDINAVIA, NORTHERN BELGIUM, THE NETHERLANDS
Michael Brown
Phone: +44 161 440 0854
Mobile: +44 79866 34646
E-mail: Michael.Brown@GulfPub.com

SALES OFFICES—OTHER AREAS

AUSTRALIA—Perth
Brian Arnold
Phone: +61 (8) 9332-9839
Fax: +61 (8) 9313-6442
E-mail: Australia@GulfPub.com

CHINA—Hong Kong
Iris Yuen
Phone: +86 13802701367, (China)
Phone: +852 69185500, (Hong Kong)
E-mail: Iris.Yuen@GulfPub.com

BRAZIL—São Paulo
Alfred Bilyk
Phone/Fax: 11 23 37 42 40
Mobile: 11 85 86 52 59
E-mail: Brazil@GulfPub.com

INDIA
Manav Kanwar
Phone: +91-22-2837 7070/71/72
Fax: +91-22-2822 2803
Mobile: +91-98673 67374
E-mail: India@GulfPub.com

INDONESIA, MALAYSIA, SINGAPORE, THAILAND
Peggy Thay
Publicitas Singapore Pte Ltd
Phone: +65 6836-2272
Fax: +65 6634-5231
E-mail: Singapore@GulfPub.com

JAPAN—Tokyo
Yoshinori Ikeda
Pacific Business Inc.
Phone: +81 (3) 3661-6138
Fax: +81 (3) 3661-6139
E-mail: Japan@GulfPub.com

PAKISTAN—Karachi
S. E. Ahmed
Intermedia Communications
Phone: +92 (21) 663-4795
Fax: +92 (21) 663-4795

CLASSIFIED SALES

Gerry Mayer
Phone: +1 (972) 816-3534, Fax: +1 (972) 767-4442
E-mail: Gerry.Mayer@GulfPub.com

REPRINTS

Rhonda Brown, Foster Printing Service
Phone: +1 (866) 879-9144 ext. 194
E-mail: RhondaB@FosterPrinting.com

The first number after the company name is the page on which an advertisement appears. The second number, appearing in parentheses, after the company name, is the Reader Service Number. There are two ways readers can obtain product and service information:

1. Go to www.HydrocarbonProcessing.com/RS. Follow the instructions on the screen, and your request will be forwarded for immediate action.
2. Go online to the advertiser's Website listed below.

Company / Website	Page	RS#
AB Sandvik Materials Technology www.info.hotims.com/45683-61	116	(61)
Advanced Refining Technologies www.info.hotims.com/45683-107	32	(107)
aeSolutions www.info.hotims.com/45683-72	38–39	(72)
Aerzen USA Corporation www.info.hotims.com/45683-157	30	(157)
Air Liquide www.info.hotims.com/45683-101	E-140	(101)
Altra Couplings www.info.hotims.com/45683-152	14	(152)
AMACS www.info.hotims.com/45683-155	21	(155)
American Petroleum Institute www.info.hotims.com/45683-168	110	(168)
Ametek Process Instruments www.info.hotims.com/45683-73	72	(73)
Aveva AB www.info.hotims.com/45683-59	E-142	(59)
Avondale www.info.hotims.com/45683-77	121	(77)
AW Chesterton Company www.info.hotims.com/45683-160	44	(160)
Axens www.info.hotims.com/45683-51	164	(51)
Bartec GmbH www.info.hotims.com/45683-83	18	(83)
BASF Corporation www.info.hotims.com/45683-70	50	(70)
BiLFINGER Water Technologies www.info.hotims.com/45683-55	47	(55)
Bluebeam Software www.info.hotims.com/45683-58	26	(58)
Burckhardt Compression AG www.info.hotims.com/45683-79	7	(79)
Cameron www.info.hotims.com/45683-67	107	(67)
Carver Pump Company www.info.hotims.com/45683-158	34	(158)
CB&I www.info.hotims.com/45683-109	90	(109)
CertifiedSafety www.info.hotims.com/45683-86	88	(86)
Curtiss Wright Flow Control Company, DeltaValve www.info.hotims.com/45683-65	2	(65)
Curtiss Wright Flow Control Company, Farris Engineering www.info.hotims.com/45683-56	111	(56)
Dresser-Rand www.info.hotims.com/45683-96	35	(96)
DuPont BELCO Technologies Corp www.info.hotims.com/45683-97	75	(97)
DuPont Clean Technologies www.info.hotims.com/45683-151	6	(151)
Emerson Process Management (DeltaV)	93	
Emerson Process Management (Fisher)	15	
Exxon Lubricants & Specialities	45	
Flexitallic LP www.info.hotims.com/45683-93	5	(93)
Foster Wheeler www.info.hotims.com/45683-52	77	(52)
FourQuest Energy www.info.hotims.com/45683-79	E-144	(79)
GE Water & Process Technologies www.info.hotims.com/45683-104	124	(104)
Gulf Publishing Company		
Circulation	136	
Events—EMGC	160	
Events—IRPC	157	
Events—WGLC	134	
GPC Software	122	
Marketplace	158–159	
Webcasts	46, 114	
Haldor Topsøe A/S www.info.hotims.com/45683-103	48	(103)
Heurtey Petrochem www.info.hotims.com/45683-64	55, E-146	(64)
Hytorc www.info.hotims.com/45683-54	23, 25	(54)
Inpro/Seal Company www.info.hotims.com/45683-74	129	(74)
International Exposition Co. www.info.hotims.com/45683-53	130	(53)
International Process Plants (IPP) www.info.hotims.com/45683-94	E-148	(94)
Invensys www.info.hotims.com/45683-63	112	(63)
ITW Polymer Technologies/Chockfast www.info.hotims.com/45683-162	51	(162)
JobElephant.com www.info.hotims.com/45683-156	22	(156)
John Zink Company www.info.hotims.com/45683-80	31	(80)
Jonell, Inc. www.info.hotims.com/45683-60	36	(60)
Koch–Glitsch www.info.hotims.com/45683-111	E-150	(111)
Linde Engineering NA www.info.hotims.com/45683-85	102	(85)
Linde Process Plants www.info.hotims.com/45683-82	163	(82)
Lurgi GmbH www.info.hotims.com/45683-101	E-140	(101)
Man Diesel & Turbo www.info.hotims.com/45683-102	71	(102)
Merichem Company www.info.hotims.com/45683-84	97	(84)
Metso Automation www.info.hotims.com/45683-98	8–9	(98)
Milliken Workwear www.info.hotims.com/45683-76	17	(76)
OMI Industries www.info.hotims.com/45683-91	65	(91)
Paharpur Cooling Towers, Ltd. www.info.hotims.com/45683-100	42	(100)
Pagell www.info.hotims.com/45683-166	81	(166)
PCC Energy Group www.info.hotims.com/45683-90	95	(90)
Pentair www.info.hotims.com/45683-105	105	(105)
Prosernat www.info.hotims.com/45683-163	57	(163)
Quest Integrity Group LLC www.info.hotims.com/45683-164	57	(164)
Sabin Metals Corporation www.info.hotims.com/45683-68	29	(68)
Saint-Gobain NorPro www.info.hotims.com/45683-106	12	(106)
Samson GmbH www.info.hotims.com/45683-169	96	(169)
Scott Safety www.info.hotims.com/45683-71	66	(71)
Servomex Ltd. www.info.hotims.com/45683-167	89	(167)
SO.CA.P SRL www.info.hotims.com/45683-161	51	(161)
Spraying Systems Co www.info.hotims.com/45683-66	40	(66)
Sulzer Chemtech, USA Inc. www.info.hotims.com/45683-88	55	(88)
Summit Industrial Products, Inc. www.info.hotims.com/45683-154	20	(154)
Team Industrial Services www.info.hotims.com/45683-99	98	(99)
ThyssenKrupp Uhde GmbH www.info.hotims.com/45683-81	E-152	(81)
Tiger Tower Services www.info.hotims.com/45683-75	87	(75)
TLV America Corporation www.info.hotims.com/45683-57	123	(57)
Total Safety www.info.hotims.com/45683-108	85	(108)
Trachte USA www.info.hotims.com/45683-165	64	(165)
UOP, a Honeywell Company	115	
Velan www.info.hotims.com/45683-153	16	(153)
Wood Group Mustang www.info.hotims.com/45683-89	119	(89)
Zeeco www.info.hotims.com/45683-110	82	(110)
ZymeFlow Decon Technology www.info.hotims.com/45683-92	127	(92)

LORAINE A. HUCHLER, CONTRIBUTING EDITOR
Huchler@martechsystems.com

The case for change

When should you change water-treatment suppliers? After a failure? When you are dissatisfied with the quality or price of your incumbent supplier? When you receive an attractively-priced proposal? When your trusted service representative retires? When your competitor buys your refinery?

Answer: Any of these events can be a compelling reason to make change. The real question is: Is your organization prepared to make a change? Changing suppliers might be the absolute right decision. However, if your organization is not ready to make the change, you have created risk—risk of damaging your equipment, lost production, and, perhaps, unsafe conditions that cause injury or death.

Preparing your organization for change. It seems so obvious: If you want to make change, you must understand what things need to change. Too often, organizations violate all of the best management practices and change suppliers based on a flawed assumption, such as: change will solve present problems. Implementing changes without understanding the cause can create additional problems.

The most logical first step is a situation analysis. The classic method, SWOT—Strengths, Weaknesses, Opportunities and Threats—is an excellent framework. The source of data for this analysis is an audit of the plant's utility water systems that include inspection of the systems and interviews with key stakeholders about the operation of their systems and performance of the incumbent water-treatment supplier. This audit should address these key aspects: system reliability, system operability, data acquisition and management, documentation and procedures, organizational, water and energy efficiency, and costs.

TABLE 1. Excerpt of a SWOT Analysis—US Petrochemical Plant

Strengths	Weaknesses
System reliability and operability—Robust pretreatment infrastructure due to recent upgrades and capital investment	**System reliability—**Many water-related failures in the cooling water circuits caused over $1 million of lost opportunity and capital costs. **Organizational—**Operators lack knowledge about the operating, monitoring and maintenance practices for utility water.
Opportunities	**Threats**
Organizational—Improve operator competency	**System reliability—**Failures in the cooling water circuits will continue to compromise operability, safety and profitability until personnel increase their knowledge and ownership of the utility water systems.

The audit is also an opportunity to understand the value (or lack of value) of the incumbent supplier services. One caveat—it is important to understand the supplier's responsibilities. It is human nature for plant staff to wish that their supplier would take responsibility for as many tasks as possible. When plant personnel abdicate responsibility for managing their utility water systems instead of delegating responsibility, it is a path to certain disaster.

Audit objectives. The primary objective of the audit is to identify risk, especially if the plant has experienced a failure, lost production and/or damage. The audit should drive corrective action to change the circumstances leading to the failure. Often, the audit team will identify opportunities for improvement—possibly reducing the chemical treatment costs. In all cases, the results of an audit will lead to greater understanding and ownership of the utility water systems by plant personnel—a proven path to reduce risk.

Finally, plant personnel should understand the inherent costs and risks of changing water-treatment suppliers. Costs include time, for the audit team members and for the procurement and contract specialists, along with capital costs—new chemical feed control systems and integration of data acquisition and control systems into the plant's distributed control system.

Audit outcomes. An example SWOT analysis from an audit of a large petrochemical facility is summarized in **TABLE 1.** The audit team created a set of prioritized tasks with estimates of costs and return on investment, identification of the person who "owns" the task, and a negotiated completion date. Once plant personnel have implemented changes to correct the deficiencies and improve the utility water system, the organization can begin the process to change chemical suppliers.

Think before changing. Making decisions based on unsubstantiated assumptions is a dangerous practice—including the decision to change water-treatment suppliers. Instead, use this apparent problem as an opportunity to benchmark your operating costs and competency. Strengthen your staff's ownership of the utility water systems. If change is the best solution, create a long-term, partnership relationship with the new supplier. HP

LORAINE A. HUCHLER is president of MarTech Systems, Inc., a consulting firm that provides technical advisory services to manage risk and optimize energy- and water-related systems including steam, cooling and wastewater in refineries and petrochemical plants. She holds a BS degree in chemical engineering, along with professional engineering licenses in New Jersey and Maryland, and is a certified management consultant.