



ART Guidelines for Troubleshooting Hydrotreater Performance Problems

The volume of cracked feeds and heavy crudes processed through hydrotreaters has increased at many refining facilities. In addition, the crude market includes higher percentages of crudes with relatively high levels of sulfur, metals and thioiphenic compounds that require upgrading through new or revamped hydrotreaters. Concerning these challenges, the Advanced Refining Technologies (ART) Seminar recently held in Houston included a presentation by Woody Shiflett on systematically troubleshooting hydrotreater performance problems that can occur at the operating conditions required for these aforementioned feedstocks.

Shiflett and the team of hydro-processing experts at ART leverage resources from Chevron, Grace Davison and Japan Energy to provide hydrotreater operating best practices that benefit distillate hydrotreaters, fixed bed and ebullating bed hydroprocessing units, and other linked process assets. Some of the most important recommendations for troubleshooting hydrotreater performance problems evolving from this resource base are discussed in further detail with the following recommendations and case studies.

Feed Properties

To ascertain the effect of feed contaminants on unit performance, actual feed properties should be compared with expected properties, including feed contaminants, feed API (or density), feed boiling range (especially the tail end) and feed composition. Contaminants include nitrogen, sulfur, ConCarbon, asphaltenes and catalyst poisons (Ni, V, Fe, Na, Si, As, Hg, etc.).

Table 1. LCO properties

Feed Composition	Expected	Actual
LCO (vol%)	20	34
LCO Properties		
D2887 EP, °F (°C)	724 (384)	794 (423)
Sulfur, wt%	0.51	0.7
API Gravity	25.6	19.2
Nitrogen, ppm	393	790
Hard Sulfur	1227	4034

To demonstrate how feed properties can affect operating conditions, one case study on start-of-run (SOR) activity evaluation showed that actual weighted average bed temperature (WABT: determines the level of sulfur and nitrogen) was 20-25 °F (11 °C) higher than expected (about 638 °F) from pilot plant testing. In this case, loading and sulfiding went according to plan, while analytical and activity testing on samples from the lots met expectations. The higher than expected WABT can be accounted for by the difference in feed properties (expected vs actual). Catalyst were exposed to 40-60% LCO very early in the run leading to some premature coking (Table 1).

Key Operating Conditions

Actual operating conditions should be compared to expected conditions for feedrate/LHSV, make-up H₂ rate and purity, recycle rate and purity (H₂ partial pressure), H₂S concentration in treat gas and operating temperature profile. In a case study where the WABT for a low sulfur diesel operation is about 35 °F higher than ex-

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Advanced Refining Technologies

A Case Study - Over Conversion in ULSD

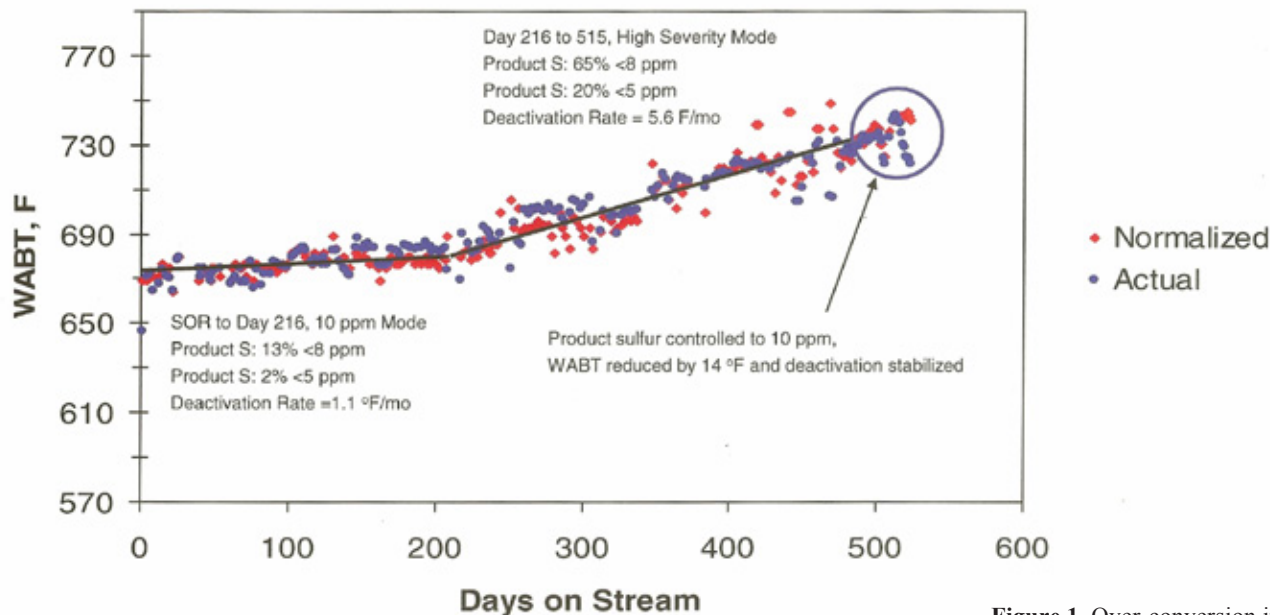


Figure 1. Over-conversion in ULSD

pected, the H₂S in the recycle gas was actually 8-10 vol% and H₂ purity was 60-65 vol%, whereas the proposal was based on only 2.0 vol% H₂S in the recycle gas. Reducing WABT in ULSD hydrotreaters operating in high severity mode will stabilize deactivation rates and reduce over-conversion, as shown in **Figure 1**.

Loading and Sulfiding Guidelines

Catalyst wetting with feed is important for best in-situ sulfiding performance, preferably at maximum liquid rate for good flow distribution. The catalyst should never be left in hydrogen at temperatures and pressures greater than 450 °F and 435 psig, respectively, as there is a potential to reduce the metals. To avoid possible metals reduction and/or coking, the exotherm should be controlled to less than 30 °F.

Half the sulfur is consumed during the low temperature sulfiding step. The oil used for sulfiding should be straight run (SR) with a final boiling point (FBP) less than 670°F, while the final temperature should be in the 600-650°F range. Use of cracked stocks should be avoided for at least three days after completion, and then

gradually introduced.

In a diesel unit case study where problems during sulfiding lead to the catalyst having to be dumped, regenerated and loaded back into the unit and sulfided properly, the WABT was about 40°F lower than expected at SOR. This could not be explained by feed and operating conditions. The catalyst lot analysis was favorable and loading densities were consistent with expectations. Problems during sulfiding included premature H₂S breakthrough observed at low temperature and only 1/3 of the sulfur was consumed. Decreased feedrate resulted in a large exotherm with 610°F inlet temperature and outlet reaching 662°F. The final high temperature hold was too short (< 30 minutes).

If the reactor temperature response is unusually low the problem may be due to recombination (for naphtha applications), hydrogen starvation, feed bypassing the catalyst and equilibrium limited reactions (e.g., polynuclear aromatic [PNA] hydrogenation). To determine the occurrence of hydrogen starvation, H₂ requirements should be calculated and compared to available H₂. Feed bypassing catalyst may be occurring due to a leaking feed/effluent heat exchanger or

feed going to a bypass line.

Mercaptan Recombination

Problems with mercaptan formation occur due to the recombination reaction between olefins and H₂S. The reaction is favored by:

- High temperature and low H₂ pressure, which increases the equilibrium concentration of olefins
- High H₂S concentrations
- Processing cracked feedstocks (e.g., VB/coker naphtha), which increase H₂ consumption leading to low partial pressure at the reactor outlet.

In a mercaptan recombination case study where SR naphtha was processed at 8 LHSV, 410 psig and 400 scfb H₂/oil ratio, decreasing reactor inlet temperature resulted in decreasing product sulfur, with inlet temperatures ranging from about 545 to 625°F, product sulfur ranged from about 0.2 ppm at 545°F at the lower end and about 1.2 ppm at the higher end at 630°F. The expected temperature at SOR was 602°F.

In this case, catalyst loading and sulfiding went fine and feed and op-

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erating conditions were actually less severe than design.

H₂ Availability

In a case study for determining H₂ availability when switching between 0.2 and 0.04 wt% sulfur (35% LCGO, 0.7 LHSV & 520 psig), SOR WABT was 10-15°F higher than expected. In this case, sock loading was involved instead of dense loading with feed-only sulfiding. Instead of avoiding the use of cracked stocks for at least three days after startup, LCGO was processed immediately upon startup. The temperature response was lower than expected with an apparent activation energy of 15 kcal/mole vs the 25 kcal/mole (or higher) typically expected under these conditions. Producing a product with less than 0.04% sulfur resulted in extremely high deactivation rates. H₂/oil ratio is way too low with an actual 500 scfb compared to an expected 1000 scfb. The “rule of thumb” is that the H₂/oil ratio should be three-to-four times greater than H₂ consumption. The H₂/oil ratio should be even higher for ULSD.

Liquid Maldistribution

Non-uniform liquid flow (maldistribution) leads to poor catalyst utilization, resulting in lower than expected activity and shorter cycle length. Potential causes include absence of a liquid distribution tray or else a distribution tray that’s operating poorly because it may have been installed incorrectly (not level) or it is in poor condition (e.g., not properly cleaned, leaking, etc). Also leading to poor catalyst utilization is poor catalyst loading, demonstrated by variable loading densities or non-uniform void space in reactor beds, along with these beds not being level. Other causes of maldistribution are due to objects left in catalyst beds during loading, uneven coking (“coke balls”), low flow rates (<0.5 psi/ft) and low liquid mass flux (minimum = 2000 lbs/ft²-hr [2.7 kg/m²-sec]).

Signs of non-uniform flow (maldistribution) include hot spots, uneven radial temperatures (>10°F or 5°C) and poor catalyst activity. Potential solutions include increasing gas rate, dense loading catalyst, change (reduce) catalyst size, improve internals design and/or installation, reduce upsets and im-

$$\frac{LHSV_1}{n-1} \left(\frac{1}{S_{Prod_1}^{n-1}} - \frac{1}{S_{Feed}^{n-1}} \right) = \frac{LHSV_2}{n-1} \left(\frac{1}{S_{Prod_2}^{n-1}} - \frac{1}{S_{Feed}^{n-1}} \right)$$

Figure 2. Feed rate response test

prove emergency responses and employ an activity grading catalyst system.

Increasing the feed rate reduces the effect of flow maldistribution or channeling. In feed rate response testing, flow distribution performance is compared at two feedrates (LHSVs) while holding temperature, H₂/oil ratio and feed quality constant. The rate constant for each feedrate is calculated as shown in Figure 2. Since the temperature is constant, the rate constants should be equal as per the equation in Figure 2.

Unit Case Studies

In a ULSK (ultra-low sulfur kerosene) case study, the unit started up with nearly 50°F lower activity than expected after catalyst loading and sulfiding had proceeded smoothly and a satisfactory catalyst lot analysis. Feed and conditions were actually easier than expected as shown in Table 2.

Sulfur speciation results on feed and outlet product showed the presence of “easy” sulfurs including mercaptans. The easy sulfurs in the product suggest a maldistribution or channeling problem. The mercaptans disappeared when temperature was increased. There was no recombination. In addition, product distribution of easy sulfurs was different from that of the feed and was not due to an exchanger leak.

The unit also showed unusually low pressure drop despite dense loading the catalyst. A subsequent flow rate test indicated a reaction order of 4.3 compared

to the expected range of 1.0 to 1.5. The 4.3 reaction order indicates that 25-30% of the catalyst is not being utilized due to maldistribution.

In a ULSD unit that started up consistent with expectations, it quickly experienced rapid activity loss that resulted in a shut down after only a few months compared to the expected one-year cycle length. Refinery personnel indicated that current and prior cycle feed and operations were not significantly different. Other important considerations included:

- The lot analysis for the catalyst were well within specifications and similar to other successful lots
- Loading and sulfiding proceeded as expected
- Operational issues could not be excluded (H₂ partial pressure potentially low, only sporadic feedstock endpoint data)
- Caustic contamination from upstream disulfide vessel was possible
- Sulfur speciation on feed and product indicated no exchanger leaks.

The spent catalyst showed significant contamination with the top of the bed containing a lot of powder along with the catalyst.

In another case study involving a coker naphtha unit, the spent catalyst in this unit also showed [Cont. page 4](#)

Table 2. ULSK unit case study expected and actual feed and operating conditions.

	Expected	Actual
API	44	46
Sulfur, ppm	2500	1100
LHSV, hr ⁻¹	1.75	1.47
H ₂ /Oil, scfb	870	930
Product Sulfur, ppm	10	11
SOR WABT, °F	575	623

significant contamination. In this case, the naphtha unit processing a SR/coker blend started up with the expected activity but then began to experience higher than expected deactivation part way through the cycle. Increases in temperature resulted in decreases in product sulfur, but the unit eventually reached its maximum inlet temperature.

Feed and operating conditions were typical with anti foam usage being similar to previous cycles. Loading and sulfiding both went fine. The unit's temperature response indicated that recombination was not occurring. Sulfur speciation on feed and product showed no easy sulfur species in the product, eliminating the possibility of an exchanger leak. Poisoning

was suspected, but analytical testing on the feed stream was inconclusive. Upon further examination, the spent catalyst showed significant contamination due to silica (Si) and arsenic (As). Although Si contamination was expected, As was not expected.

Troubleshooting Recommendations

Troubleshooting hydrotreater performance requires a systematic approach. It should first be verified if there really is a catalyst performance issue while simultaneously verifying that current feed and operating conditions have not changed. If there are changes, correlations should be used to ascertain if the changes explain the performance difference. In addi-

tion, the loading and sulfiding procedure should be reviewed to confirm if anything unusual occurred.

To determine the cause of performance issues, test runs should also be performed, including evaluation of the unit's temperature response and checking for potential maldistribution with feedrate response testing.

Collection of a sufficient number of (corresponding) feed and product samples is critical. The suppliers laboratory tools should be used to verify internal analyses and to have analyses com-



Editor's note: *Refinery Operations* extends its appreciation to Advanced Refining Technologies (ART) and Grace Davison for supplying this information on hydrotreater performance guidelines. For further elaboration, the reader should contact Woody Shiflett, Ph.D., Ch.E, Director of Global Marketing at ART (wosh@chevron.com; +1-510-242-1166).

PROCESS OPERATIONS

Increasing Sulfur Recovery Unit Capacity

Many refiners are finding that their existing sulfur recovery units (SRUs) do not have sufficient capacity to meet expected increases in upstream hydrotreating and hydrocracking capacity. In many cases, 99.9% sulfur recovery is expected from existing SRUs.

The primary units of the industry for sulfur recovery and processing continue to be conventional amine sweetening units followed by Claus units and tail gas cleanup units (TGCU). Since Claus units are capable of recoveries of about 94 to 97% for typical, three-bed units, TGCUs may need to be added to comply with stringent emissions regulations and achieve 99.9% sulfur recovery. Overall, current sulfur recovery efficiencies are typically in the 96-99% range, depending on the feed gas composition (CO₂,

mercaptans, ammonia, etc.).

Much of the focus for achieving 99.9% sulfur recovery pertains to partial oxidation of ammonia (NH₃) bearing gas streams, potentially containing minor but significant quantities of H₂S. Refining processes that result in H₂S byproduct formation from complex organic sulfur compounds also tend to convert nitrogen compounds to NH₃.

The NH₃ byproduct has normally been combusted in the Claus reaction furnace, where the amount of NH₃, which can be processed, is generally considered to be limited to 30-35% of the total Claus feed. To deal with this limitation, Worley Parsons has patented a process in which the ammonia acid gas combustion is carried out in a tail gas furnace.

Because of the complexities of the calculations in designing and analyzing

these three units (amine + Claus + TGCU), process simulation programs have become a necessity to optimize the units in any reasonable length of time. With TGCU units now involved in many cases, the overall simulation becomes a complex iterative process because of the TGCU's recycle stream. According to previously published information available from Bryan Research and Engineering, Inc., through the use of simulation technology, refiners may investigate and implement optimal operating conditions in these linked units.

A more detailed discussion pertaining to sulfur recovery developments and applications will be discussed in the March 2011 special report sponsored by *Refinery Operations*: "Innovations in Hydrocracking and Hydrotreating." ■



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Dealing with Chlorides Found in Refinery Streams

It is not uncommon for refinery feedstocks to contain chloride salts, including chlorides of alkali and alkaline earth metals in amounts ranging from 1 to 2000 ppm. It is known that hydrocarbon streams containing these chloride contaminants, at elevated temperatures and in the presence of water, will hydrolyze to form hydrochloric acid, causing problems to processing equipment in the absence of any chemical treatment programs.

With crude feedstocks first treated in a desalter, much of the salts and other water soluble contaminants are removed prior to introducing the hydrocarbon stream through the refinery process, including heat exchangers, furnaces, distillation columns and the associated processing equipment such as pumps, valves, piping, etc.

The process stream exiting the desalter is typically introduced into an atmospheric distillation column together with steam to make a rough fractionation into generally four cuts that includes an overhead stream containing C1 to C8 hydrocarbons. It is common practice to stream strip the crude in the atmospheric column. Thus, any hydrochloric acid formed upstream of the atmospheric column will be carried over in the light fraction (C1 to C8) to be condensed with water. Subsequent treatment of this condensed fraction will result in the hydrochloric acid

coming in contact with and causing corrosive damage to process equipment used to treat the condensed fraction.

The usual method for dealing with the overhead corrosion resulting from the hydrolysis reaction is to apply neutralizers and corrosion inhibitors, of which there are many available on the market. These inhibitors are costly and in many instances cause foaming and deposition problems, which can be more damaging than the corrosion problem.

According to information currently available from Champions Technology, Inc., a method has been developed for reducing hydrolysis in a hydrocarbon stream wherein a hydrocarbon stream containing a chloride compound which undergoes hydrolysis at elevated temperatures and in the presence of water to form hydrochloric acid is treated with an effective amount of a treating agent comprising at least one over-base complex of a metal salt and an organic acid complexing agent. Details on the application of this novel technology are beyond the scope of this discussion and should instead be referred to the subject matter experts at Champions Technology. According to information currently available, the treating agent is introduced into the hydrocarbon stream when the stream is at a temperature below which any substantial hydrolysis of the chloride containing compound occurs. ■

Preventing Furnace Tube Failures

Refinery processing involves heating hydrocarbon streams to as high as 1400°F (760°C), making furnace reliability an important aspect of refinery operations. More importantly, furnace tube metallurgy is exposed to temperatures close to 1750°F (950°C). Furnace tube degradation occurs primarily due to corrosion, metallurgical changes and creep.

Furnace tube failures occur primarily

due to creep. Creep is the time-dependent deformation occurring when tube metallurgy is subjected to stress (internal pressure) at elevated temperatures. For example, unscheduled coker outages due to furnace tube failures are costly within the unit as well as throughout the refinery. This is because the entire refinery can be impacted due to reduced crude processing rates or adjustments to crude **Cont. page 4**

blends if the coker must be shut down or operated at reduced capacity.

The coking process temperature is about 950°F (510°C) and 9Cr-1Mo tubes in a coker furnace typically are heated to a maximum of 1250°F (705°C) when coked. Therefore, being able to predict furnace tube life is

essential to maintaining refinery furnace reliability and refinery profitability. Fortunately, creep can be simulated and the simulation used to predict what will occur in the future—the remaining life of a furnace tube. There is a general trend, according to “show of hands” survey at the most recent NPRA Q&A that

periodic infrared (IR) scans are being used more often to help monitor tube metal

temperatures and check for hot tubes and hot spots. Periodic IR scans ensure that the tubes monitored are representative of furnace conditions. ■

Hydrogen Purification and Production Strategies

Increasingly strict environmental and product-quality regulations and the trend towards processing heavier and more sour crude oils has resulted in higher hydrocracking and hydrotreating capacities. The resulting increase in hydrogen consumption and limited or even decreased generation are creating tight hydrogen balances in many refineries throughout the world.

Efficient use of hydrogen is a necessity, with refineries facing eroded margins since the fourth quarter of 2008 due to constrained refinery operations combined with the onus for significant investments in hydrogen

generation and purification, particularly with regard to steam methane reforming (SMR) operations.

During the production of hydrogen in an SMR plant, CO₂ is also produced. In fact, the SMR process in centralized plants emits more than twice the CO₂ than hydrogen produced. To avoid CO₂ emissions into the atmosphere, CO₂ can be concentrated, captured, and sequestered.

Sequestration concepts and technologies are relatively new and there is no long-term test evidence to prove that these technologies will be successful. SMR is a mature technology, but the

problem with SMR is that it is operating at or near its theoretical limits. To deal with these challenges, hydrogen technology licensors such as Praxair have introduced improvements to various sections of SMR units and hydrogen purification units, such that a refining facility can plan on increasing H₂ production from their existing hydrogen production assets on a very cost-effective basis. Information relevant to these developments can be searched and downloaded from the “Archives” section of www.refineryoperations.com. ■

Operating Parameters for ULSD Units

According to a white paper by Sweers, et al, Grass-roots ULSD units tend to be designed for much higher operating pressure (1100+ psig).¹ The Sweers et al paper noted that in the past, typical operating condition requirements for hydrotreating diesel streams yielded satisfactory run lengths and basic bulk properties were sufficient to predict catalyst performance. Commercial run lengths are shorter for ULSD production because the operating conditions required for ULSD are more severe. It has been found in commercial units that some combinations of feed properties and operating conditions lead to unexpected accelerated deactivation rates. These rates cannot be predicted simply

by the traditional monitoring parameters and feed bulk properties used in the refining industry and are not typically due to catalyst poisons.

According to a paper by C. Buccisano, predicting the performance of key refinery units such as ULSD hydrotreaters requires understanding of the unit’s operations with other units upstream and downstream, storage capacities and re-run capabilities.² Buccisano noted the methodology and software technology employed by Jardine (a DNV company) to correctly capture the impact of various parameters on a ULSD unit’s performance.

Buccisano noted that quantification of expected performances is the key

step in the design optimization process, but the optimization can only be carried out if cost data and revenue data are forecasted for each one of the design options. The conclusions from a ULSD optimization study can therefore be used to increase return on investment as justify additional capital expenditure.

1. H. Marcel Sweers, et al, “Impact of Vaporization on Catalyst Deactivation: Part I – Low Pressure ULSD Production, Criterion Catalysts & Technologies Co.
2. Carolo Buccisano, “ULSD – LSG Upgrade Project: Maximize Utilization, Minimize Risk, Justify CAPEX,” DNV Consulting. ■

Improving De-isobutanizer Performance

A sulfuric acid alkylation plant had excess feed and reaction zone capacity, with the deisobutanizer (DIB) being the limiting factor. Specifically, the DIB reboiler duty was at its maximum and the reboiler steam-supply valves were wide open. A new alkylate Rvp specification of 5.0 psia also was constraining operation. An engineering review by Ascent Engineering suggested minor design modifications and maintenance

items which, when implemented, increased alkylate production capacity by more than 25%.

In another DIB optimization project, Petro Control's Generalized Distillation Shortcut (GDS) model was used on an alkylation unit DIB column. GDS employs first principles models by the way of a short cut simulation of a section of the column, hence the name of the model: General

distillation shortcut. According to information from Petro Control, GDS works by fitting a bottom (or top) column composition that would best agree with column temperature profile. The advantage of this approach, according to Petro Control is that it correctly accounts for the very nonlinear effects of temperature, pressure and vapor-liquid ratio. ■

UNIT AUTOMATION

Optimizing Hydrocracker Unit Operations to Reduce Costs and Improve Yields

Achieve better performance, energy efficiency, asset reliability, and safer operations through more precise control.
By Tim Olsen and Gary Hawkins, Emerson Process Management

With over 100 hydrocracking units operating around the globe and more under development, many opportunities exist for refiners to achieve higher levels of safe, reliable performance. No matter how efficient a unit may seem today, that process can be improved through advanced automation, resulting in reduced energy usage, greater production flexibility, increased quality and yields, and higher reliability with lower maintenance costs.

Because hydrocracking units are key contributors to a refinery's profitability, making improvements to optimize process operations makes real economic sense, especially in view of the high cost of designing and building these units. By employing advanced technologies to enhance performance, the refining company can maximize the return on its original investment.

Most hydrocracker units in North America were designed for gasoline production, but refiners are now looking to produce low and ultra-low sulfur diesel fuel as well. Achieving greater flexibility involves many factors, including more precise temperature control of the reactor beds as well as the upstream heaters and downstream fractionators. A variety of automation solutions is available from

Emerson Process Management to help refiners obtain greater operational flexibility from the day they are implemented.

For example, many existing units were designed for catalysts that are now outmoded. Today's catalysts are much more active and selective, but they cannot deliver high performance without tight temperature and pressure control. That's why the accurate and reliable measurements delivered by digital instrumentation are so important for controlling the reactor yield and other parts of the overall process.

Many factors contribute to better utilization of process units including good heater management, reduced control valve variability, extensive field instrument diagnostics, and vibration monitoring of a wide range of mechanical assets. In essentially every case, the precise control afforded by combining accurate measurements with modern automation solutions contributes significantly to better utilization.

The upgrades suggested for existing hydrocracking processes apply equally to greenfield units currently being designed or planned. Smart digital control should be specified wherever designers are seeking to build world-class performance into hydrocracking operations. The

commissioning and calibration of smart instruments are faster and more accurate with these features, resulting in benefits at project startup. After startup, the diagnostics generated by these same smart instruments support ongoing benefits through better performing operations and enhanced reliability. These advantages ensure a faster return on the substantial investment in hydrocracker units with continued performance for sustainable returns. ■



Editor's note: This discussion by Tim Olsen (tim.olsen@emerson.com) and Gary Hawkins (gary.hawkins@emerson.com) is based on a more detailed paper on "Optimizing Hydrocracking Operations to Reduce Costs and Improve Yields" to be published in *Refinery Operation's Innovations in Hydrocracking & Hydrotreating* special report (March 2011).

EDITORIALLY SPEAKING

Investment in Process Assets

While waiting for the next industry upcycle, the refining industry will rationalize process assets in 2011. Some analysts project an upturn in the global business cycle in 2011, when credit availability stabilizes. However, a growing percentage of non-oil based fuels are another reason that the refining industry could face margins pressures throughout 2011.

Although 2012 may seem like a long way off, demand could actually increase sooner than later. Today, there are more than 6 billion people who are more dependent than ever on hydrocarbon-based products in spite of all the green energy hype. China recently surpassed the United States in new car purchases. This is one of the reasons crude prices may increase steadily in 2011, this is even taking into consideration the expansion of the non-oil based fuels, such as liquefied natural gas (LNG), compressed natural gas (CNG), biofuels, etc.

However, due to multiple discoveries of large gas reserves in shale formations, liquefied fuels from these gas resources ranging from naphtha to Fischer Tropsch based diesel, could eventually account for a higher than expected percentage of the transportation fuels market.

Credit and capital are currently unavailable on the scale needed for increasing the level of processing complexity

required to transform oil into products for fuel and petrochemical markets. For example, the European refining industry generally does not have enough capacity to make enough diesel from the types of crudes available to them.

Investments will begin to accelerate only if refined product prices increase relative to any increases in feedstock costs. Of course, certain investments have already taken place and will continue to take place throughout the world; there are just too many people and too many businesses in need of oil products, even today.

In addition, the cost of building or revamping refining capacity has also decreased. What are needed are investors and a reduced capital cost structure. For example, the cost of steel fabricated processing equipment, such as heavy-walled, high pressure reactors (e.g., hydrocrackers for diesel/distillate production) has decreased. The cost of many technical services required by refiners, such as those for monitoring corrosion and fouling, has also been reduced.

One of the most important cost-saving measures in refinery operations is minimizing catalyst losses. This is also important from an emissions compliance perspective. While a detailed discussion of all the causes for increasing catalyst losses is beyond the scope of this article, some of the most notable causes include



Rene Gonzalez, Editor
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mechanical failures and changes with the circulating catalyst quality.

When embarking on a process control and optimization project, there are many stakeholders that will have to be convinced new computer control systems are the right direction for their facility. Any issues about the systems under consideration need to be resolved, as there are challenges when using new technology for the first time. Weighing the benefits and subsequent savings that automation can bring, as opposed to relying on old, tried and tested technology, which can carry the risk of obsolescence, is the first course of action. ■

CALENDAR OF EVENTS

JANUARY (2011)

18-21, *Refineries Asia 2011*, IBC Asia Pte Ltd, Singapore, info@ibcasia.com.sg, www.abc-asia.com

23-26, *Chemtech World & Industry Automation & Control*, Jasubhai Group, Mumbai, India, sales@jasubhai.com, www.jasubhai.com

FEBRUARY

13-16, *Hydrogen Conference & Expo*, National Hydrogen Association, Washington D.C., info@hydrogenconference.org, www.hydrogenconference.org

23-25, *ERS FCC & Hydrocracking*, Eurotek Refining Services Ltd, Windsor, London, enquiries@eurotek-refining.co.uk, www.eurotek-refining.co.uk

MARCH

8-11, *European Fuels Conference, 12th Annual Meeting*, World Refining Association, Paris, +44 (0) 20 7067 1800, www.wraconferences.com.

20-22, *NPRA Annual Meeting*, NPRA, San Antonio, Texas, +1 202 457 0480, www.npra.org

30-31, *14th Annual ARTC Meeting*, Singapore, Incisive Media & Global Technology Forum, +852 3411 4829, www.gtforum.com

APRIL

3-6, *The Middle East Downstream Week, 12th Annual Meeting*, World Refining Association, Paris, +44 (0) 20 7067 1800, www.wraconferences.com.

MAY

2-6, *Coking Safety Seminar*, Coking.com, Galveston, Texas, +1 360 966 7251, www.coking.com.

24-27, *NPRA Reliability & Maintenance Conference & Exhibition*, NPRA, Denver, Colorado, +1 202 457 0480, www.npra.org.

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