

Implementing coker advanced process control

This project resulted in about a half-year payout

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How would you convince someone to install advanced process control (APC) on a coker unit? Such discussions involve not only dollar figures but also one's own conviction.

The arguments in favor of APC are known:

- Use of a complex relationship between controlled variables (CVs) and manipulated variables (MVs) in qualitative and quantitative ways
- Prediction capability of two to three hours
- Optimization potential
- Reduced standard deviation permits operating closer to the limits.

To go beyond the numbers we asked one of the coker operators to come up with reasons for installing APC:

“APC maximizes throughput in compliance with all given product qualities and security limits. Unit security as well as product yields would improve. APC monitors all the process variables of the unit [Authors' note: That depends on the controller setup of course] and reacts on changes instantly inside prespecified limits. APC saves energy, for example with optimized heater control. Exceeding CO emission limits would be practically eliminated. APC does what no operator can do: comprehend all changes simultaneously and react. APC works according to chemical engineering principles [Authors' note: Again, that depends on the controller setup]. This is a chance for the operator to get to know the unit better, which should help him or her in certain situations (when APC would not be available). . .”

This project was executed using the best available technology:

- Multivariable predictive controller (MVPC)
- Application performance monitor
- Platform for coding the inferential models
- Inferential control models based on first-principles modeling technology
- Experienced APC company to implement the APC application.

An earlier article¹ and an editorial² on this application set the scene, giving a progress report in the early stages of the project and highlighting some of the issues. At the time of writing the article

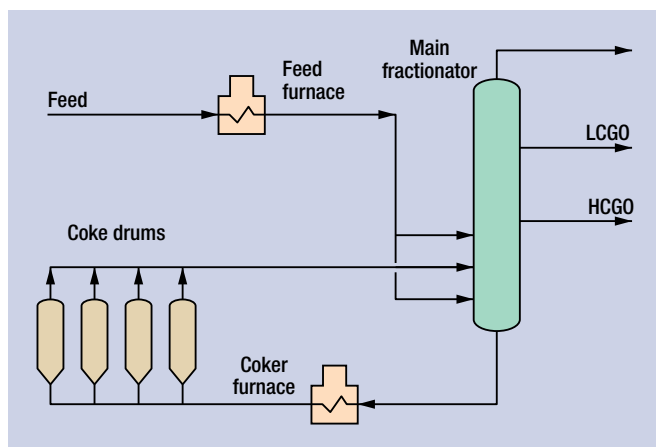


FIG. 1 Delayed coking is one of the most difficult refinery units to operate and control.

we were still struggling with commissioning problems, whereas at the time of the editorial commissioning was just concluded. In this article we present a simplified summary of our post project audit, omitting details that are not in the general interest but keeping the spirit and many numbers of the original document. Ruhr Oel has used a whole year's worth of data for this audit, certainly a very extensive statistical sample.

A 2001 benefit study has identified large benefits associated with feed maximization, yield maximization to quality constraints and keeping the unit within equipment constraints. In general the application performance has exceeded our expectations and we calculated the actual payout time on this project at roughly half a year.

Introduction. Delayed coking is one of the most difficult refinery units to operate and control. Fig. 1 shows a very simplified diagram of delayed coking. The unit takes vacuum resid (fresh feed), heats it and injects it into the main fractionator bottom, where it is mixed with an internal reflux recycle of heavy cracked material. The total fresh and recycled feed is then heated

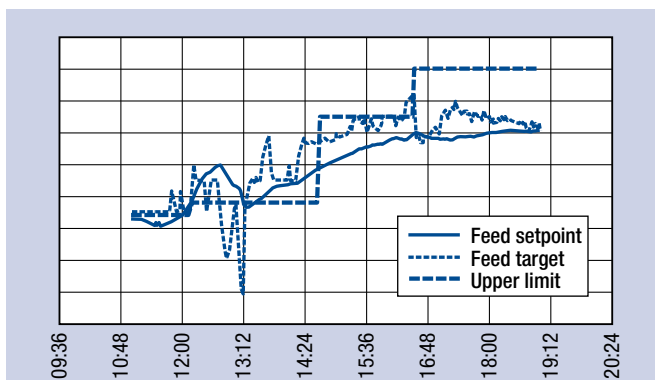


FIG. 2 The increase in feed flow made APC very profitable even if no other benefit is considered.

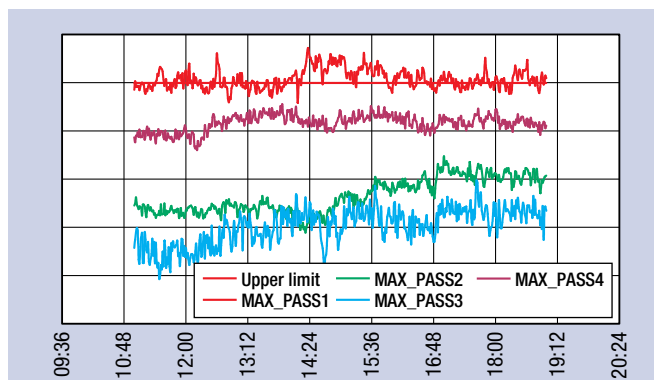


FIG. 3 Coker furnace pass skin temperatures.

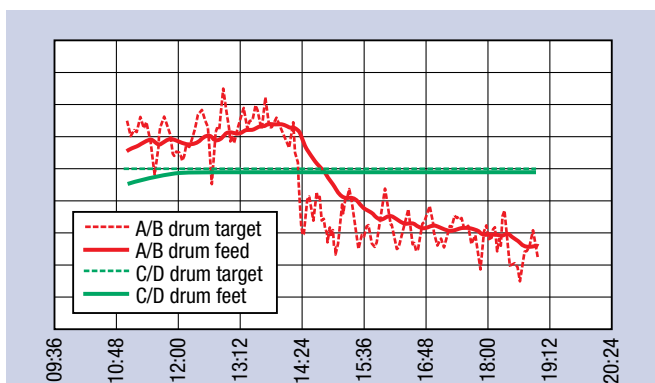


FIG. 4 The feed targets for drums A/B are not achieved because of the dual constraints.

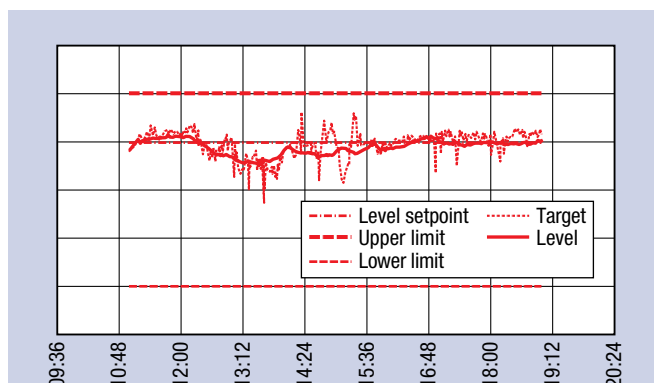


FIG. 5 The fractionator level is stable while the unit is constrained.

in the coker furnace to a high cracking temperature. Hot partially cracked feed flows from the coker furnace into coke drums, where the reaction continues. Cracked distillate vapor ascends in the coke drum and flows into the fractionator for separation.

Coke remains in the drums and is periodically removed. That is the main reason for the coker being a difficult unit to operate. Twice daily filled coke drums are switched off for coke removal and empty drums are connected. The drum that was just filled then goes through a cycle of steaming out, cooling, opening, coke removal, closing, steaming and pressure testing, heating and finally reconnecting to the furnace and fractionator. Heating cold drums creates significant disturbances because the heating is done by sending hot cracked vapor through the cold drum, depriving the fractionator of both heat and material. The main disturbance, however, comes later upon connecting the warm empty drum. The drum temperature needed for cracking is around 500°C, but the new empty drum cannot practically be heated to such temperatures. Drums are typically switched in at 400°C, which quenches the reaction almost completely, causing a major disturbance that lasts about an hour until the newly connected drum reaches normal operating temperature.

And there are other characteristics that make coker operation difficult: unforgiving high temperatures, danger of coking trays in the lower sections of the fractionator and instrument failures due to coking or high temperatures. The fractionator bottom serves as a surge drum for the coker furnace, but it is not a large surge capacity and the handles for controlling this level—changing fresh feed, coker furnace feed or recycle affect

other unit constraints. Operators must practically cut feed to cope with simultaneous level and unit constraints, and that is a heavy economic penalty.

If that is not enough, the matrix of CV responses to MV is quite dense. Usually in refinery units each CV is associated with one or sometimes two main MVs, and process response matrices are fairly sparse, but in a coker each of the main MVs affects many CVs, presenting a challenge for the operators. Operators must manipulate many handles to reach a single objective. To do it right they should have the dynamic and steady-state responses of the unit “at their fingertips,” but very few do.

For the same reasons, implementing APC on a delayed coker is also difficult. In addition to the drum switch challenge, good control of a dense response matrix requires high accuracy of the unit dynamic response models. In absence of such accuracy constraint balancing cannot be achieved without constraint violation.

Nevertheless, we aimed high and implemented the following objectives into the control strategy:

- Maximize fresh feed to constraints. A frequently asked question is “Has the throughput changed?” because it often appears as hardly “increasing.” Detailed analyses reveal that as long as the product qualities are within specification and increased throughput can be accepted, APC looks to increase fresh feed.
- Balance the constraints when there are degrees of freedom, and increase the feed further. For example, use equal pass flows when the furnace is not constrained, but when it is constrained permit some increase of unconstrained passes over constrained ones.
- Maximize the product draws to quality (or other) con-

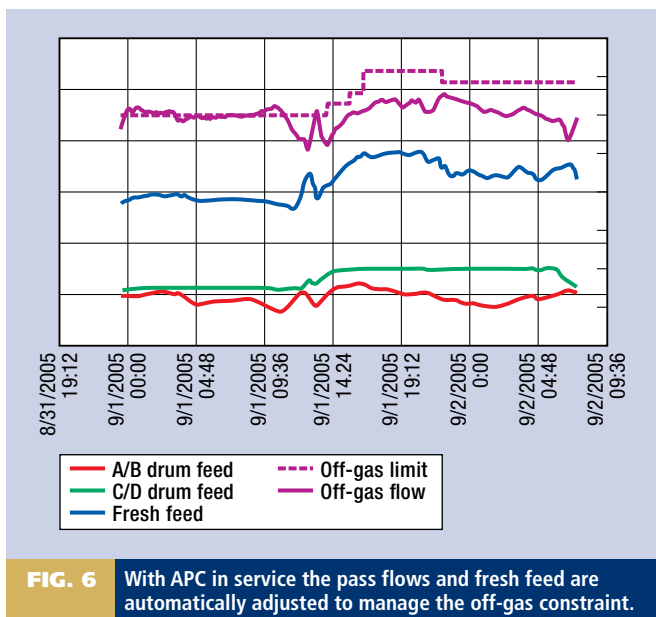


FIG. 6 With APC in service the pass flows and fresh feed are automatically adjusted to manage the off-gas constraint.

straints. The two side products: HCGO and LCGO are under control, but the HCGO product quality target was questioned because of a downstream hydrotreater problem. Very heavy HCGO increases the hydrotreater reactor pressure drop and shortens its run time.

- Work smoothly during drum switches without sacrificing approach to constraints
- Control fresh feed mixture recipe.

Fresh feed discussion. The coker operation is so varied that it is difficult to judge to what degree the feed is being truly maximized. We have tightened two important coker constraints and hence expected to see a decline in throughput. Tightening the HCGO quality constraint causes increased recycle, which penalizes capacity. Skin temperature constraints have not been officially tightened except before APC they were not rigorously enforced, and violations of up to 50°C were not uncommon, but after commissioning the APC this situation was corrected and skin temperature constraints were not to exceed their target of 660°C.

After several months of monitoring the application's performance it was a pleasant surprise to find out that while the maximum observed throughput came down somewhat, due to the more constrained environment, the average throughput still came up significantly. Before APC the average fresh feed flow was 165.2, with variability as high as 31 t/h, whereas after commissioning APC the average feed has gone up to 173.7, with variability of 14.5 t/h. That is a 5% increase, rendering APC very profitable even if no other benefit is considered.

Fig. 2 shows a September 1, 2005, 8-hr trend of:

- Operator maximum target for fresh feed flow
- Controller prediction of feed flow when the unit reaches steady state
- Actual coker feed flow.

As the period began, feed was clamped by operator limit, probably unnecessarily. At about 12:00 the operator permitted higher throughput and APC started to take up the slack. At about 13:00 a drum switch disturbance began and feed was temporarily reduced to cope with the shortage of heat. Following recovery from the disturbance the operator permitted further throughput

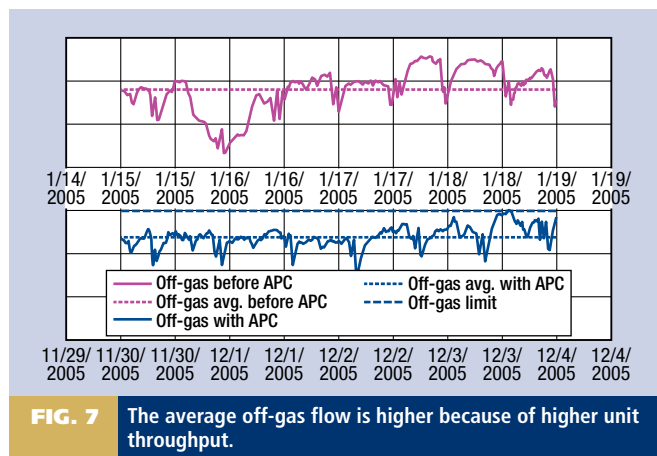


FIG. 7 The average off-gas flow is higher because of higher unit throughput.

increase and the feed continued to climb, reaching the point of 20% increase, settling this time below the operator limit due to other unit constraints.

To illustrate the multitude of constraints on the unit and their effect on throughput, Fig. 3 shows a September 2, 2005, 8-hr trend of coker furnace pass skin temperature. At 19:00 pass 1 was constrained on high skin temperature, HCGO draw temperature and LCGO 90% point. Pass 2 was constrained on HCGO draw temperature, LCGO 90% point and the upper MV pass flow limit. Passes 3 and 4 were constrained by HCGO draw temperature, LCGO 90% point and the feed target for drum pair C/D. Fig. 4 shows the feed flows and corresponding feed targets for each pair of drums at the same time, September 2, 2005. It is clear that the feed target for drums A/B (furnace passes 1 and 2) is not achieved because of the dual constraints of skin temperature on pass 1 and the maximum pass flow MV constraint on pass 2. This, along with the C/D feed target and the product quality constraints are restricting throughput of fresh feed to the unit.

Fractionator bottom level control. The fractionator bottom level is a surge inventory for the coker furnace and must be managed between limits of 45% to 65%. A setpoint of 60% serves to drive toward the higher side of the range. We have mentioned three handles for the level control: fresh feed, coker furnace pass flows and HCGO pump-down. There is actually another handle: the fractionator fresh feed has upper and lower feed injections; the upper injection absorbs significant amounts of heavy HCGO and it increases bottom level as well as recycle ratio. When pass flows are constrained, either by heater constraints or operator target, then the fresh feed would eventually be cut to ensure that the level remains within limit. Likewise, upon an increase in the coker pass flows, fresh feed is also increased proportionally.

Fig. 5 is an 8-hr plot, again on September 2, 2005, of the fractionator level, showing good stability, while the unit is completely constrained.

Off-gas. Control of off-gas (to the amine treatment column) is important since the downstream unit often has problems. In the past operators simply cut the fresh feed and pass flows manually. With APC in service the pass flows and fresh feed are automatically adjusted to manage the off-gas constraint. Fig. 6 shows the off-gas and pass flows. Initially on August 31 the maximum off-gas target was held low, and APC suppressed the feed flow. On September 1 the limit was relaxed, permitting a feed increase.

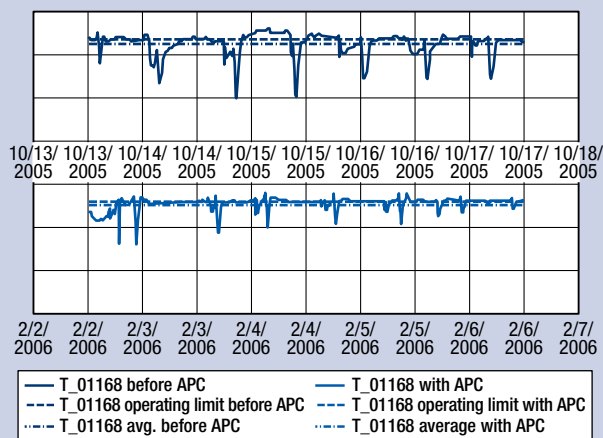


FIG. 8 HCGO is controlled better with the MVPC in closed loop.

TABLE 1. Product qualities' variability (standard deviations) before and after APC

	Naphtha 90% dist.	LCGO 90% dist.	LCGO 90% dist.	HCGO 90% dist.
Before APC	7.7°C	15.1°C	20.7°C	12.8°C
After APC	4.5°C	7.2°C	9.7°C	5.7°C

TABLE 2. Relative product yields before and after APC

	Naphtha	LCGO	HCGO
Before APC (base)	100%	100%	100%
After APC	117%	107%	101%

TABLE 3. Standard deviation of product yields before and after APC

	Naphtha	LCGO	HCGO
Before APC	± 4.6%	± 11.8%	± 6.4%
After APC	± 2.5%	± 4.8%	± 3.4%

And beyond the simple amine absorption constraint, coker off-gas shortages during drum switches cause site-wide refinery heating gas disturbances. If smoother control of coker off-gas were possible, all refinery furnaces would also run more evenly. Fig. 7 illustrates that this objective was partially accomplished. The figure compares off-gas flow before and after APC. In addition to the reduced variability it turns out that the average off-gas value is higher when APC is on because of higher unit throughput.

Product quality controls. One of the differences between a delayed coker versus other refinery units is that the coker is never at steady state. There are two drum switches a day and each switch is associated with two major disturbances: start of drum warm-up and switch. In addition to those four daily drum events, cracking conditions drift gradually as drum levels increase. When a unit is at steady state, a 04:00 morning sample typically yields lab results at about 09:00, at which point the operator would correct unit conditions. But when a unit is not at steady state the meaning of a sample taken at 04:00 is completely lost at 09:00. Indeed Ruhr Oel is aware of these difficulties and has specified very infrequent

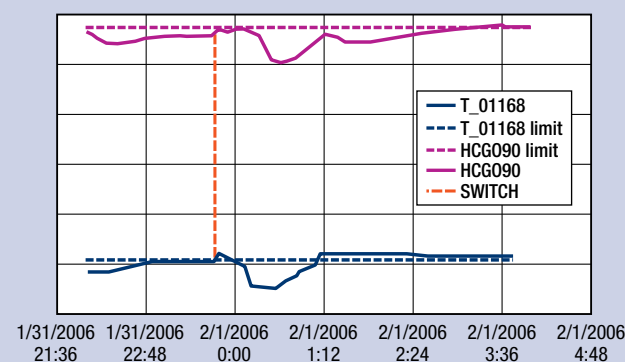


FIG. 9 Although the MVPC treats the draw temperature with higher priority, HCGO cutpoint is also under control.

lab tests on the unit. Hence, coming up with reasonable inferential models was an important project task.

The most important quality to control on the coker turned out to be HCGO endpoint. HCGO is taken into a hydrotreater where the heavy HCGO components shorten the hydrotreater run length. At the time that this problem became evident, no inferential models existed and operators started paying attention to HCGO draw temperature, T_01168. This draw temperature, while theoretically not a reasonable indicator of endpoint, became nevertheless a yardstick, and a draw temperature constraint target was negotiated between the coker and hydrotreater management teams. With the implementation of inferred properties it became easier to adjust operation to achieve a desired set of product properties, but because of the inter-unit agreement HCGO draw temperature continued to play a major role, and it is the most watched MVPC's CV.

Fig. 8 shows a comparison of how well the HCGO draw temperature is controlled with and without the MVPC in closed loop.

We have of course realized that the wrong yardstick is being used, and in fact had taken steps to prevent the APC optimizer from playing games, cutting HCGO much deeper while shifting LCGO into HCGO and keeping the draw temperature under control. We have used the APC opportunity to provide better indication, but the current practice cannot be changed overnight. With time we would gradually move to controlling real product qualities instead of draw temperature. Fig. 9 is an 8-hr trend of HCGO draw temperature and HCGO 90% inference through drum warm-up and switch events. It can be seen that although the MVPC treats the draw temperature with higher priority, HCGO cutpoint is also under control.

Generally the heavier the product, the more difficult it is to determine the boiling end point. And the repeatability and accuracy of a boiling end point is much lower than the 95 vol% or even 90 vol% point. Hence, the 90 vol% points are inferred for the products SBi (heavy naphtha), LCGO and HCGO. For LCGO the flash point is calculated too, because part of the LCGO can go to storage to avoid hydraulic bottleneck in downstream units.

Validation of the inferential models was not easy because of the very low frequency of sample taking, the uncertainty of sample time and the poor repeatability of lab tests on heavy, thermally unstable streams such as LCGO and HCGO. Taking those uncertainties into account we have come up with a way of defining lower and upper plausible range for the samples. Fig. 10 shows the ASTM 90% point lab data for naphtha, LCGO and

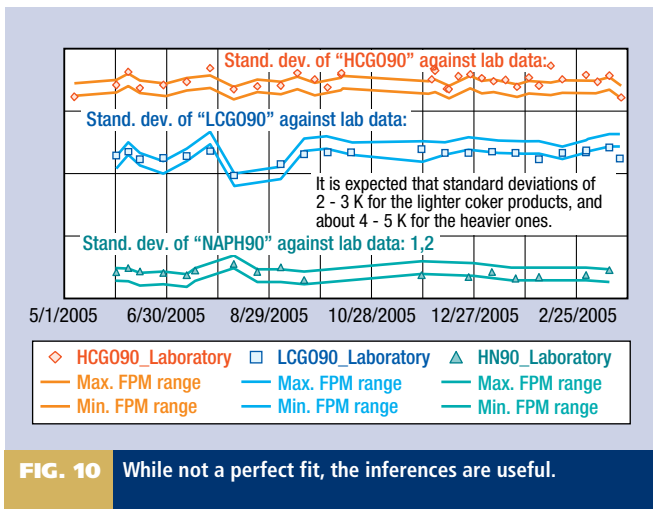


FIG. 10 While not a perfect fit, the inferences are useful.

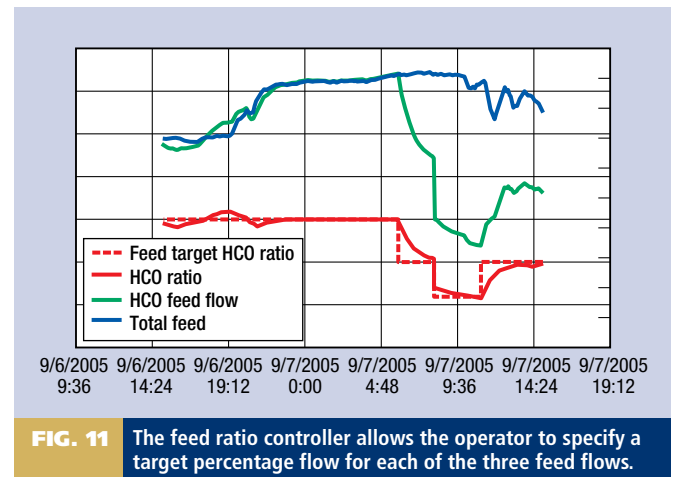


FIG. 11 The feed ratio controller allows the operator to specify a target percentage flow for each of the three feed flows.

HCGO, and how well it fits into the plausible range. While not a perfect fit, the inferences are certainly useful keeping product cutpoints at target during the ever-shifting unit conditions.

Table 1 shows how the APC has reduced the variability in the unit. The numbers are based on lab results from before and after commissioning APC.

With this better control it would be of interest to compare the distillate yields. We expected the distillate yields to come down because the quality constraints have been tightened, and that was the case for the maximum yield, but on average the yield did go up as shown in Table 2. The increased distillate yield at the expense of coke is quite a large benefit to the refinery.

Again, APC is able to reduce standard deviation of flows which allows a closer approach to the limits. Table 3 shows that halving of the standard deviation is possible.

Recipe control (feed ratios). The quality of green coke produced in the drums is determined by the feed mixture. There are two constraints on managing the feed system:

- Ensuring the feed mixture remains constant
- Ensuring a feed tank is emptied at a given rate.

The feed ratio controller allows the operator to specify a target percentage flow for each of the three feed flows. Then as APC adjusts the fresh feed out of the feed drum it would also adjust the feed mixture into the drum to maintain the ratio at target. Fig. 11 is a 24-hr trend showing the behavior of APC upon operator adjustment of feed ratio (red) for one of the three feed flows (green), while the fresh feed (blue) is also changing to meet the various unit constraints.

Other considerations. In a coker application, and in our opinion in general, the more restrictions that exist and the more control handles that can be used (i.e., the more complex the task) the better APC can help because precise control of such a complex unit is beyond some operators' ability. Of course such a

statement can be made only if the application is set up correctly, not a trivial task in itself. And to increase APC effectiveness, operators should always question whether the current limits on controlled and manipulated variables are appropriate. Insofar as we can make money by setting correct limits, incorrect limits would be counterproductive and lose money. Regular monitoring of the application is safeguarding continued benefits, with strong input from lead operators supporting the on-site APC engineers ensuring that limits are regularly reviewed and the performance checked against actual requirements. **HP**

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Volker Haseloff has 20 years of experience, starting out in process operation, then becoming a project engineer, and since 1998 he has concentrated on advanced control projects. At BP Gelsenkirchen he covers advanced control applications on different units especially the coker unit. Mr. Haseloff is a ChE graduate of the Technical University of Dresden.

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