Model-based control of crude product qualities

Unique advanced controls improved operation, particularly during crude switches

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Crude unit advanced control at the Statoil refinery in Kalundborg, Denmark, features a model-based method for controlling product qualities. For completeness, a conceptual summary of all controls involved are included, namely:

- Crude feed control
- Crude furnace
- Crude fractionator heat recovery
- Crude fractionator product qualities
- Crude light ends.

Model-based quality control uses simplified heat-balance equations for estimating the crude TBP curve. Once the TBP curve is known, yields can be set such that all products meet their quality targets. The technique is very useful during crude switches because as new crude enters the fractionator, it immediately affects the heat balance. Consequently, timing of increasing or decreasing sidestream yields is correct. This technique was first reported in 1985.[1] It was installed during 1983 in a refinery that switched crudes every three to four days. This model is an improved version of that early application. A crude unit simulation program was used to test and develop more accurate crude unit correlations.

Crude unit. The crude unit (Fig. 1) has a preheat section (not shown in the figure), a furnace section, a crude column and a light ends section. In the furnace section are three furnaces. Two of the furnaces process one mix of crudes. The third can process a different crude. It is equipped with a flash drum which separates the bottoms of the different crude.

The crude column takes a full-cut naphtha as its overhead product and three sidecuts: kerosine, light gas oil (LGO) and heavy gas oil (HGO). The side products as well as the bottom product are stripped by steam.

There are three pumparound circuits: top, middle and bottom pumparounds (TPA, MPA and BPA). It is also possible to pump naphtha product reflux to the top tray, but the reflux is minimized. For simplicity, it is excluded from the discussion (although it is included in the control application). The crude light ends section has two columns: a stabilizer and a naphtha splitter.

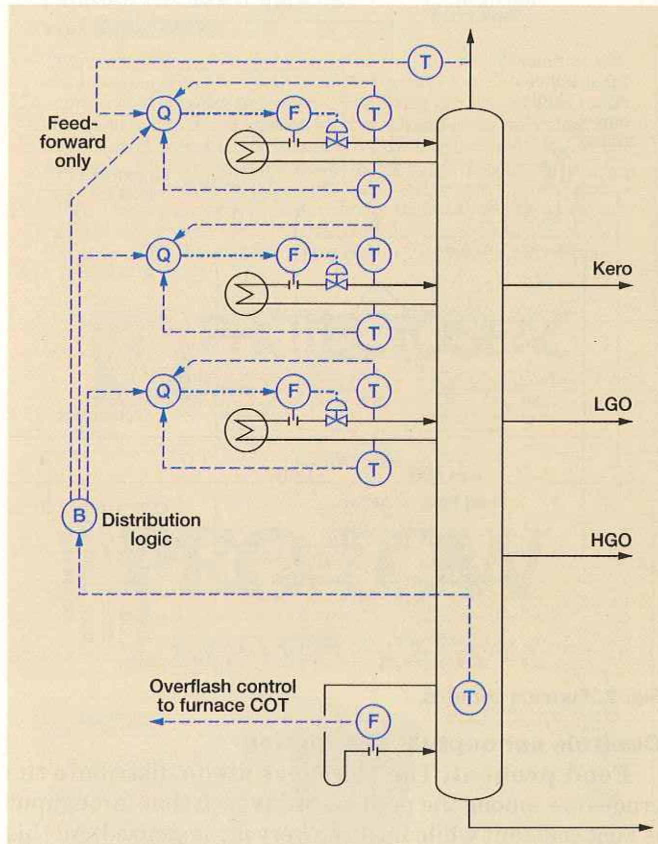


Fig. 1. Crude unit overview.

Control scheme organization. The advanced control applications are organized into modules, one or more for each section of the plant. A module is divided into schemes, where each scheme is a logical entity designed to bring about one operational objective. And finally, each scheme is built from basic loops such as feedforward, PID or design correlations.

Care has been taken to make the modules maintainable and reliable by:

- Standardizing control functions and implementation uniformity
- Schemes designed in such a way that each can work independently

Continued

• Logic to check validity of almost every input into the multivariable control schemes.

An example that illustrates scheme independence is a furnace module with two schemes: one for controlling coil outlet temperature (COT) and the other for controlling draft pressure. The draft-pressure scheme keeps the furnace operating efficiently, not too far from the stack oxygen constraint. Best results, of course, are achieved when both schemes work together. However, if for example the stack damper became stuck and the draft scheme is temporarily decommissioned, the COT scheme would continue to work without jeopardizing furnace safety.

The COT scheme includes its own constraint monitor which normally allows free fuel manipulation. But if a constraint is reached, the fuel is clamped and the COT logic overruled until the constraint is relieved.

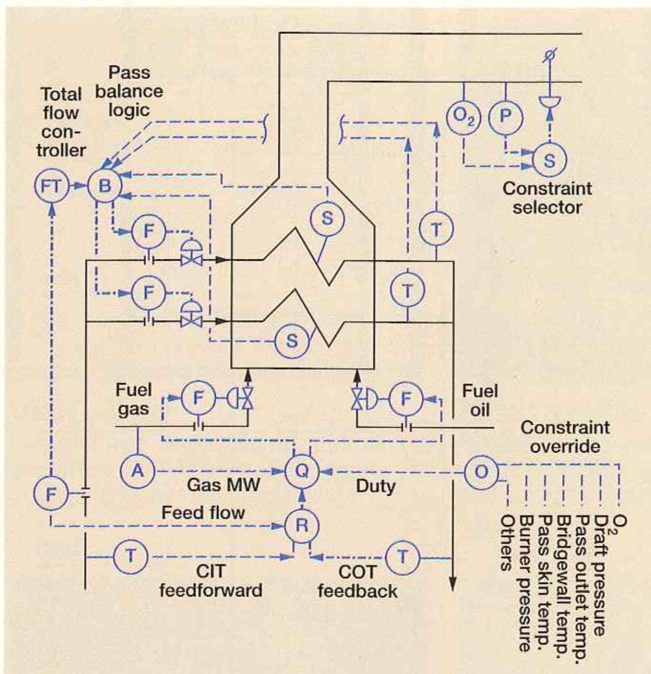


Fig. 2. Furnace controls.

Controls conceptual description

Feed preheat. The objectives are to distribute the crude flow among the preheat trains such that throughput is kept constant while heat recovery is maximized. All this is achieved without violating the many constraints on the preheat system such as:

• Desalter temperatures
• Hydraulic limitations
• Furnace loads
• Integrated heat requirements (i.e., light ends reboilers)
• Premature crude flashing before reaching the furnace. This became a serious problem with light crude operation
• Column internal reflux (satisfying the column cooling load balance requirements described later).

The control is structured as a rule-based optimizer whose optimizing logic can be overridden by constraint controllers.

Furnace. Furnace control addresses three operating objectives:

• Maintain the coil outlet temperature at its desired value and minimize deviation from that value.
• Operate at low excess air and only slightly negative draft pressure
• When necessary, overrule the first two objectives to keep the furnace always within safe operating limits.

Fuel calorific value calculation (Fig. 2). This scheme calculates fuel calorific value from density. Fuel gas density is measured by an on-line analyzer while fuel oil density is a manual input. The reasons for these calculations:

• Sometimes the gas molecular weight changes quickly, such as when LPG is injected into the fuel gas or the visbreaker changes its operating mode. Feeding forward the fuel calorific value eliminates disturbance to COT which would otherwise ensue.

• Abrupt changes in fuel oil do take place, for example, when an oil burner is taken out of service. Upon such a change, fuel gas is automatically increased to make up the heat shortage, and a precise knowledge of calorific value helps again to completely eliminate COT disturbance.

• As a side benefit, reliability is improved for monitoring the crude furnaces and unit performance.

COT control. The COT scheme manipulates fuel gas and/or fuel oil to maintain the COT at its target. It uses feedforward of coil inlet temperature, feed flow and fuel calorific value. Temperature control is so tight now that almost no disturbance can cause a deviation of more than 1°C. This includes severe disturbances such as taking a heat exchanger out of service, igniting a new burner or even losing one of the feed pumps. Usually the deviation is much lower at about 0.2°C (which is the normal noise level).

In addition to the COT control logic, this scheme also has a constraint-monitoring and override feature. When any furnace constraints are violated, the COT logic is interrupted and the fuel flow lowered to keep the furnace operating safely until the constraint is relieved.

The following constraints are monitored:
• Tube skin temperature
• Bridge-wall temperature
• Draft pressure (this constraint only becomes active when the draft control scheme described previously is not working)
• Oxygen reading in the stack
• Fuel gas burner pressure
• Fuel oil burner pressure.

Pass balancing. The pass-balancing scheme distributes feed flow among passes to keep pass-outlet temperatures reasonably close together while observing certain constraints on the skin temperatures and pass flows. Economically this scheme is not very important, as the furnace coils have not coked up. But it is very convenient for the operator, keeping the temperatures within an acceptable range even when burner distribution is uneven.

Draft pressure control. For a natural-draft furnace, this scheme is constructed as a constraint monitor, reading two constraints: draft pressure and stack oxygen, and manipulating the stack damper to keep both within safe limits.

Pumparound control. This addresses several objectives:

• Stabilize pumparounds such that the column is not disturbed upon changes in individual heat exchanger heat duties
• Respond quickly to heat balance disturbances on the column. Such disturbances occur when there are crude

changes or when there are furnace disturbances

- Adjust internal reflux in the column such that a reasonable economic balance is achieved between heat recovery and product separation and that tray loading does not exceed the flooding limit
- Adjust the furnace COT to always maintain sufficient overflash.

For the following, refer to Fig. 3. Note, however, that the figure is simplified such that inputs for each calculation block are only partially shown.

Heat-duty controllers. There are three such schemes, one for each pumparound. Using temperature and flow measurements, the schemes calculate heat duties and manipulate pumparound flows and/or return temperatures to keep that duty on target. Fig. 3 conceptually illustrates the schemes, although in reality the cooling circuits have more than one exchanger. Moreover, each scheme has additional logic to satisfy constraints. Example: reboilers must receive sufficient heat and the return temperatures must not drop below a certain value.

To respond to heat-balance disturbances, the schemes monitor tower temperatures and feedforward these to the heat-duty targets, such as when the tower heats the cooling loading increases, and vice versa. In addition, decouplers are implemented such that if one of the pumparound controllers cannot output its cooling load disturbances are redistributed to the other pumparounds. Thus, tight heat balance control is maintained even when one of the pumparounds is at its full capacity.

Tower top temperature control. Standard practice on crude units is to control the tower-top temperature by manipulating either reflux or top pumparound. This controller keeps the naphtha cut point at target while closing

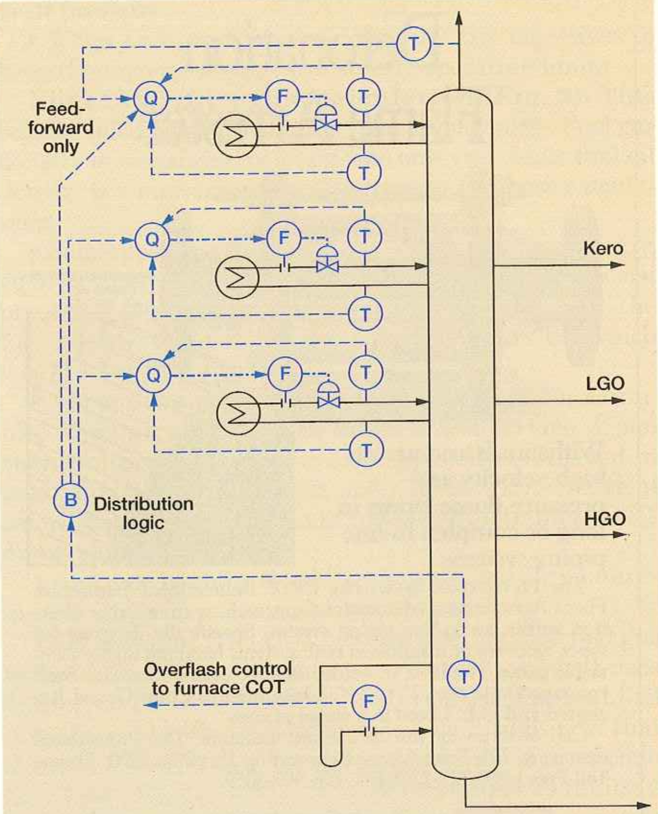


Fig. 3. Pumparound controls.

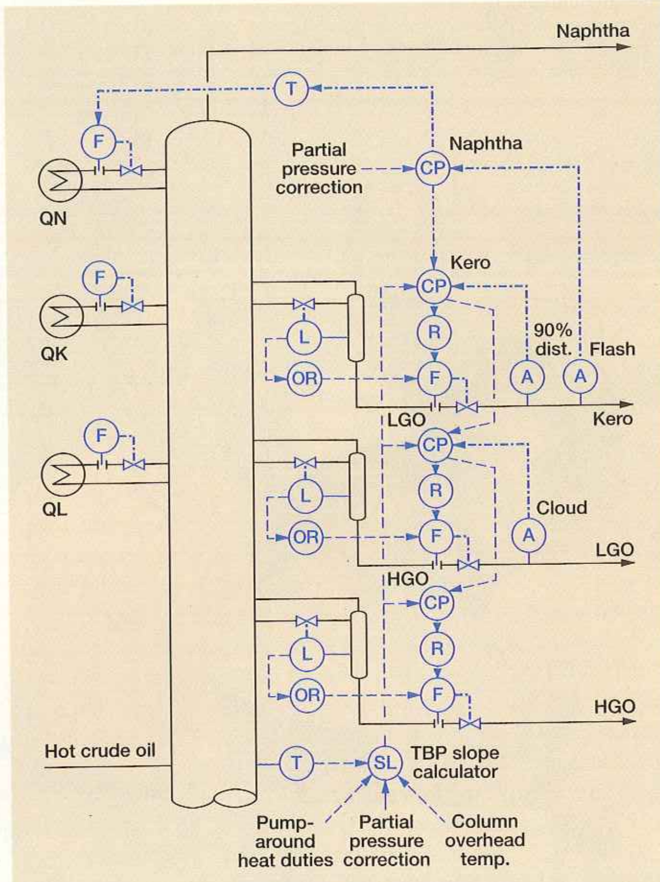


Fig. 4. Crude unit quality controls.

heat balance on the fractionator. The temperature controller is cascaded to the top pumparound heat-duty controller. It, in turn, can manipulate several variables (including top reflux) to control heat duty.

The important benefit of this scheme is that before implementation the temperature controller was cascaded to only one variable, a bypass valve on an airfin cooler. The valve would often saturate, and control was sluggish even within the controllability range. After implementation of the scheme, temperature variations around its target have reduced from above 4°C to 0.2°C.

Overflash control. Two alternative schemes control overflash percentage. One changes heat balance (furnace heat duty). The other manipulates material balance (HGO product flow). The heat-balance method description follows.

This scheme adjusts the furnace COT controller setpoint to maintain overflash at its desired target. The target is a matter of optimizing separation and dividing furnace load between the crude unit and vacuum unit. It was determined by off-line studies.

Overflash control schemes have the reputation of being difficult to implement. This is because when the furnace starts delivering hotter vapor it creates a heat disturbance in the column. The heat disturbance gradually moves up the column, and the top temperature controller finally takes action to increase the internal reflux. The increased reflux now works its way down the column and eventually shows up as overflash.

This process may take hours, often exhibits inverse response, and disturbs liquid quality on all trays.

The feature that helps shorten response time is a feedforward function that monitors tower heating or cooling and resets the pumparounds to stabilize the tower as explained earlier. With this logic, pumparounds respond immediately to an increase in flash-zone temperature. So loop delay drops from hours to minutes. The feedforward coefficient was tuned with the aid of a simulation program.

Cooling load balance optimization. This is a simplified optimization scheme that splits cooling load among the pumparound circuits such that there is a reasonable economic balance between fractionation quality and heat recovery. The scheme was developed with the use of a simulation program. Several scenarios with each crude were studied. From those studies evolved a simplified set of rules for optimally splitting cooling load. The scheme also computes proximity to flooding and weeping in column sections.

A crude flashing problem in the last heat exchanger before the furnace, and the need to wait for commissioning some feed preheat control schemes until this problem was solved was discussed earlier. To date, simple distribution logic to perform pumparound-to-pumparound decoupling and feeding forward of column temperatures has been implemented. The logic maintains a desired balance of the heat duties when no constraint is violated.

Stripping steam. Simple stripping steam-to-product ratio controllers have been implemented on the column bottoms and all side strippers. The operator determines ratio targets based on guidelines.

Product quality. The most important part of any crude unit control is overhead and sidestream product quality control.

The method uses an engineering model of computing the crude TBP. The calculation requires process measurements of column temperatures, pressure and pumparound heat duties. The TBP calculation program performs a heat balance around the column, partial pressure corrections and EFV-to-TBP conversion to calculate the crude TBP slope.

Another commonly used method for TBP-curve reconstruction is the "30% method." It assumes that the side draw temperature is equal to the 30% point on the sidestream TBP curve, and it reconstructs the TBP curve from measurements of sidestream yields and temperatures.

The heat-balance model has some advantages over the 30% model. These are:

- The 30% method uses sidestream flows in the TBP estimator. There is a fundamental flaw in using measured sidestream yields to determine sidestream yield targets. It is equivalent to treating one's own control actions as disturbances, and it introduces unintended feedback loops which may create instabilities.
- The heat-balance model applies standard engineering procedures whereas the 30% method is more of a "rule" than a method.
- The heat-balance model relies on vapor temperatures which respond rapidly to crude quality changes; the 30% method relies on liquid draw temperatures.
- The heat-balance model adapts to the crude being run and requires no tuning.

Based on the model, each sidestream yield is manipulated to maintain a constant cutpoint. The method does not require any operator or engineers' inputs, it simply monitors the heat balance. When a new, for example more volatile, crude comes into the column, cooling load necessarily increases to cope with more vapor, and the TBP scheme immediately notices the change and computes a new TBP. Moreover, TBP computation timing is just right. There is no need to lag or lead it as the increased cooling load is in phase with the increase in vapor flow. (If it is not in phase, then either the tower pressure or temperatures would go out of control.)

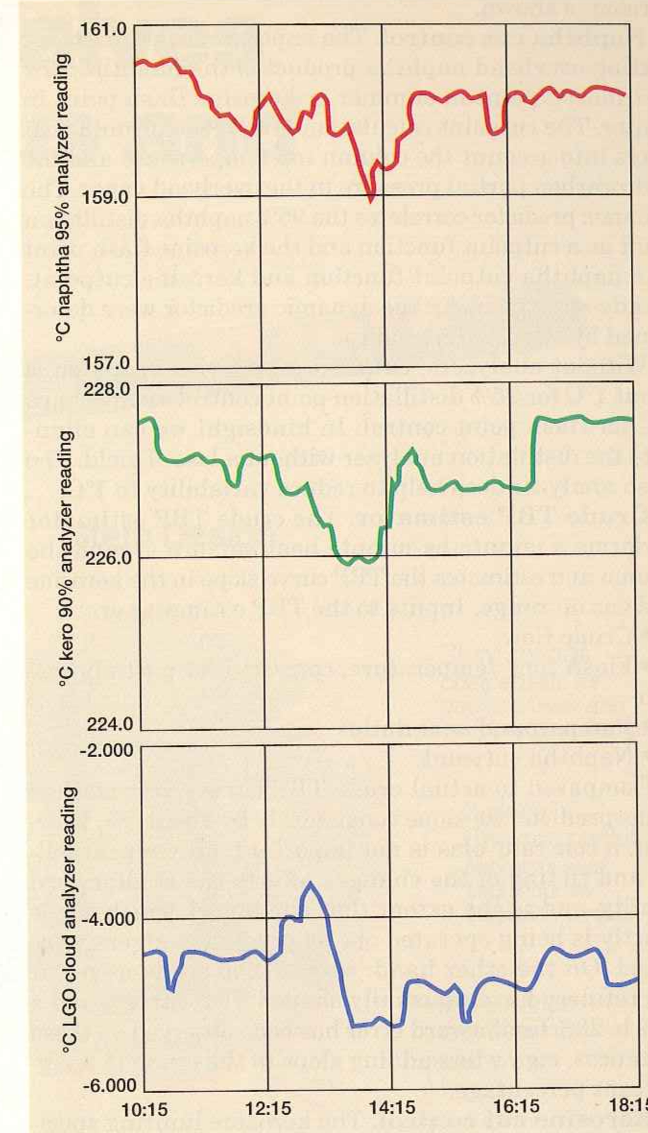


Fig. 5. Analyzer plots through a crude switch.

Another feature is use of dynamic predictors in conjunction with analyzer readings and other slow signals to compensate for deadtime. These are multivariable internal model-type predictors that use a dynamic equation form of the response. While most commercial packages use a table (or a dynamic matrix) form, it is sometimes advantageous to use an equation form. In this case, the advantage is that the model parameters can be changed "on the fly" whenever a new TBP curve is computed.

In addition, the dynamic predictor is tuned for robustness. It can withstand modelling errors of up to 40% in any parameter without going unstable.

For descriptions of specific schemes that follow refer to Fig. 4, though again for ease of reading only a simplified version is shown.

Naphtha cut control. The important constraints for cutting overhead naphtha product is the naphtha 95% distillation point in summer or kerosine flash point in winter. The cutpoint calculation involves a formula that takes into account the column top temperature and the hydrocarbon partial pressure in the overhead vapor. The dynamic predictor correlates the 95% naphtha distillation point as a cutpoint function and the kerosine flash point as a naphtha cutpoint function and kerosine cutpoint. Steady-state gains for the dynamic predictor were determined by simulation studies.

Without analyzers, cutpoint calculation precision is about 1°C for 95% distillation point control and perhaps 3°C for flash point control. In hindsight we can eliminate the distillation analyzer without a loss of yield. The flash analyzer does help to reduce variability to 1°C.

Crude TBP estimator. The crude TBP estimator performs a minute-by-minute heat balance around the column and estimates the TBP curve slope in the kerosine and gas oil range. Inputs to the TBP estimator are:

- Crude flow
- Flash zone temperature, corrected for partial pressure
- Pumparound heat duties
- Naphtha cutpoint.

Compared to actual crude TBP curves, our method underpredicts the slope consistently by about 5%. However, a constant bias is not important, only repeatability and timing of the changes affects the feedforward quality, and to the extent that one knows which crude exactly is being operated on, no prediction errors were found. On the other hand, some of the crude mixes at the refinery have unusually shaped TBP curves, and a 10% to 20% feedforward error has been observed on these instances, e.g., when adding slops to the crude in a significant percentage.

Kerosine cut control. The kerosine limiting specification is 90% distillation point. This quality is obtained in the first instance by holding the kerosine cutpoint constant. Knowing the crude TBP slope and the naphtha cutpoint, kerosine yield can be manipulated to precisely maintain the cutpoint at target. Should the naphtha cutpoint be changed, kerosine yield is automatically adjusted to hold the kerosine cutpoint constant.

Secondly, we use the kerosine 90% point analyzer in conjunction with a dynamic predictor to make a cutpoint trim adjustment. Without an on-line analyzer we can predict the 90% distillation point with repeatability about 3°C. With analyzer control the precision is narrowed to 1°C.

LGO cut control. The limiting specification on light gas oil is cloud point. The refinery cloud-control scheme models the LGO cloud as a linear function of LGO cutpoint and kero cutpoint. The cloud-predictor then resets the LGO cutpoint every time the kerosine cutpoint changes. Cloud-model coefficients change somewhat with crude type and cloud target. Current coefficients were obtained by a plant test, and there are plans to further improve on them by simulation studies of the variety of crudes at the refinery. This is not easy, however, as not all of the crudes in use have assays.

Given uncertainty in the cloud correlation, the cloud analyzer has proven useful in a predictive way to reset a bias in the cloud correlation. With correlation feedforward and analyzer feedback we were also able to obtain 1°C cloud variability around its target.

HGO cut control. HGO is an internal stream. Its yield is driven not by quality but rather by an economic balance between crude and vacuum units. Sometimes it is drawn at a fixed cutpoint and other times it is maximized subject to the overflash constraint.

Light ends. The control objectives are:

- Minimize pentane content in the LPG. This pentane is mostly iC_5, which is a valuable gasoline blending component.
- Control butane content in the light naphtha to about 2% in winter and 0.5% in summer.
- Split the naphtha to keep paraffinic hexanes in the light naphtha away from the reformer feed.

Naphtha stabilizer. Naphtha stabilizer control relies on one analyzer, $\%iC_4$ in naphtha, while top purity is inferred from reflux ratio. The analyzer-based controller resets the reboiler, and the inferential controller handles the reflux. A decoupler is configured to eliminate interactions between the two loops and permit a change in one without disturbing the other.

Naphtha splitter. Naphtha splitter control takes feedback from two analyzers: LVN 95% and HVN 5%, and in addition, to speed up response there is an inference of the split from a tray temperature.

The column controls are configured such that there is an analyzer to tray temperature to reflux cascade to control the cut. The LVN 95% analyzer reading is used for that purpose via a standard internal model predictive algorithm as described earlier.

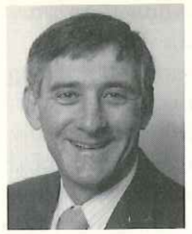
To control the gap between the two products, i.e., HVN 5% minus LVN 95%, there is an analyzer-based controller that manipulates the reflux.

In this case selection of the two control loops is such that there is little interaction between them and a decoupler is not necessary.

Implementation. Implementation was accomplished in stages, but first some interesting problems had to be overcome. Most were because, depending on which fuel was fixed, furnace controls were not always stable. Flash-zone temperature oscillations caused cloud measurement to also oscillate no matter how good the control was. Once that happened it was difficult to establish problem cause and effect. The implementation schedule called for furnace advanced controls to be implemented last. But this could not be done because some new instruments had to be installed before the furnace advanced control could be commissioned. However, once the problem was understood, the existing furnace controls tuning was improved and implementation continued smoothly.

The refinery also had problems with crude flashing. The problem affected the crude flow and temperature out of the heater. This in turn affected the flash-zone temperature.

We "moved away" from flashing by using less pre-heat and higher crude flow pressure. To this end we installed an advanced control application to maximize the pressure in the crude line. However, there is a limit to what can be done by control to solve the basic design issue. We are now considering installing a pre-flash drum and/or crude pump modifications. *Continued*



The author

Y. Zak Friedman is a consultant in the field of process control and on-line optimization. He has pursued research in the area of using process models for real-time closed loop control and optimization of refinery units, and in standardization of control technology. His experience spans 25 years in the refining industry, working for such employers as Exxon Research and Engineering and KBC Advanced Technology. Dr. Friedman holds a BS degree from the Israel Institute of Technology and a PhD degree from Purdue University. The work presented in this paper was accomplished when Dr. Friedman was working for KBC Process Automation.

Performance. The crude fractionator control has substantially improved product quality control performance, and hence the yield of the more valuable products. At steady-state operation the model consistently keeps the product qualities within 0.5°C of targets, i.e., the analyzer readings of naphtha 95%, kerosine 90% and LGO cloud deviate from their setpoints by about 0.5°C.

This compares to off-target operation on the order of 5°C in the naphtha 95%, 10°C in kero 90% and up to 4°C in LGO cloud point.

An even more dramatic result was observed during crude switches on the unit. Before implementation crude switches were so unpredictable that the operator resorted to a conservative quality target, putting up with losses of about 4% yield for four to eight hours before he could line out the unit. Now, with the product cut control scheme working, deviations of qualities from their setpoints are one to two degrees and line out occurs in about an hour.

Fig. 5 shows a plot of the three analyzers: naphtha 95%, kero 90% and LGO cloud before, during and after a typical crude switch. This particular switch was from a very volatile North Sea crude to an intermediate Middle Eastern crude. The bottoms-reduced crude yield went up from 25% to 50%, and the white product yields went down from a combined 75% to 50%. The pumparound flows were cut in half and their heat duties were cut by about 30%. The overflash was changed from 3% to 6% on crude. The crude switch started at 11:30 at the tanks but it took about 45 minutes to reach the unit. Line out was reached there in about one hour. The plot shows the analyzer readings remaining off target for about two hours, but that is only related to the measurement deadtime. All yields were established by 14:15, and the predictive controllers anticipated the analyzers' recovery from deviation and took no further action on the yields.

It can be seen that for the crude switch the maximum deviations of naphtha 95% and kero 90% from targets were 1°C, and the LGO cloud deviation was 1.5°C. The deviations lasted about one hour.

Financially, outcome of the advanced control is:

- An increase in yield of white products; about 1.5% on crude
- An increase of throughput; about 3% on crude
- A decrease in fuel required per cubic meter crude; about 5%
- A reformer yield gain of about 1% on feed due to more precise naphtha splitter operation.

LITERATURE CITED

[1] Friedman, Y. Z., "Control of Crude Fractionator Product Qualities During Feedstock Changes by use of a Simplified Heat Balance," Paper presented at the 1985 American Control Conference.