

Advanced process control re-engineering

Update on a refinery's model predictive control application that resulted in maximised FCCU feed while the unit was pushed against key constraints such as regenerator excess oxygen and main fractionator reflux temperature

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Industrial model predictive control (MPC) is well established in the process industries, with commercial MPC technology widely available and advanced process control project methodologies reaching maturity. MPC is dominated by applications using linear (or linearised) multivariable models, which map the dynamic and steady state behaviour between process manipulated variables and their related dependent or controlled variables.

In addition to the process dynamics behaviour, these models include in their structure the configuration and tuning of the base level PID loops, the models being obtained from identification of test data gathered by stepping the manipulated variables (flow, temperature or pressure controller setpoints) to obtain the dynamic response of the controlled variables (analysed product qualities, temperatures, flows, levels etc).

Throughout the life of most process plants there are physical changes made in terms of feedstocks processed and product grades produced. Often, these changes lead to structural changes in the process units (such as piping changes, distillation column packing/tray changes, catalyst changes etc) and regulatory control system changes (such as PID loops being opened or closed, new cascades and changes in analysers). In the past, changes in loop modes (closing a loop in "auto" which was previously in "manual") or significant changes in PID tuning would require a plant test to obtain new response data and subsequent re-identification of the data to obtain the new model before the MPC scheme could be updated to correctly represent the revamped unit.

This is an expensive option and many previously successful MPC applications have been taken offline following base level control system changes in the distributed control system (DCS), waiting until there is sufficient budget and/or time available to re-test the unit.

A new approach is available, where the MPC model can be directly manipu-

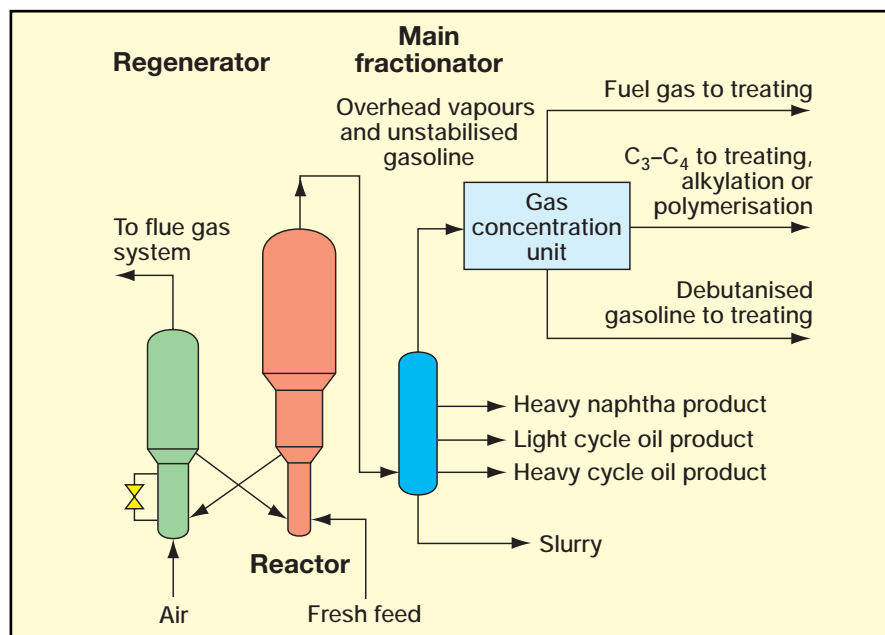


Figure 1 Fluidised catalytic cracking unit (FCCU)

lated to account for changes in PID loop modes and tuning, without the need for any further plant step-testing or identification. The Universal Process Identification (UPID) registered software from Cutler Johnston Corp is a multivariable model identification package used to identify finite impulse response (FIR) models from plant step-test data, with the ability to update the FIR model to compensate for PID tuning and loop configuration changes.

The FIR model structure includes an FIR model of the process response and the PID control loop response; these are convoluted to give the combined response. The ability to re-tune or re-configure the regulatory control system without re-testing improves the on-stream factor in advanced control applications and reduces application maintenance costs.

The first commercial project using the new UPID technology to revamp an old FCCU MPC scheme was carried out at BP Gelsenkirchen GmbH Horst Refinery in Germany. Gelsenkirchen is one of

four refinery locations of Ruhr Oel, a joint undertaking of BP Refining & Petrochemicals and Petroleos de Venezuela SA (PdVSA). BP Gelsenkirchen, a BP Refining & Petrochemicals subsidiary, manages the plants in Scholven and Horst. BP Refining & Petrochemicals operates the mineral oil and petrochemicals business in the Deutsche BP group, covering everything from fuels and heating oil to the wide range of products covered by petrochemicals.

The Scholven and Horst plants in Gelsenkirchen are operated as an alliance. With approximately 12 million tonnes crude oil capacity, this alliance is one of the largest and most efficient refinery and petrochemical complexes in Germany.

FCCU re-engineering

The fluidised catalytic cracking unit (FCCU) is designed to convert low value fuel oils into high value gasoline blend stocks by cracking the oil at high temperatures in the presence of a catalyst. The dynamics of the process are complex

and it must be operated to produce the maximum amount of gasoline subject to mechanical and economic constraints. The incremental benefits from increasing production in the unit are very large. The BP Gelsenkirchen FCCU processes approximately 4250 tonnes of feed per day.

The FCCU consists of a reactor and a regenerator as illustrated in Figure 1 (previous page). The catalyst is fluidised and moves between the regenerator and reactor by means of a pressure differential. The pre-heated oil and hot catalyst from the regenerator enter the reactor via a riser. The cracking reaction takes place in the riser and the oil vapours and the catalyst pass into the reactor where the catalyst is separated by means of cyclones. The catalyst returns to the regenerator for the removal of the coke formed on the catalyst during the reaction. The temperature of the reactor is controlled by manipulating the regenerated catalyst slide valve.

The "full burn" regenerator is a fluidised bed combustor, which burns off the coke. The air for combustion is supplied by a steam-turbine driven air compressor. The vapours from the reactor are passed to a fractionator where they are condensed into products: slurry oil from the bottom, heavy and light cycle oil (LCO) from the mid section and naphtha from the top section, the remaining gas stream passes overhead and is condensed. Liquid is pumped from the overhead accumulator to the gas concentration unit where further hydrocarbon products are separated.

A DMCplus model predictive controller (the present commercial implementation of dynamic matrix control) was originally installed in 1997. The scope of the application was across the reactor-regenerator, main fractionator, gas concentration (debutaniser and naphtha splitter), gas plant depropaniser plus associated inferential property estimators for quality control. The application was configured as one large multivariable controller covering the reactor-regenerator, main fractionator, naphtha splitter and debutaniser, with a separate controller for the depropaniser section.

The large application was required because of the tight interactions of the individual process units and downstream hydraulic limits.

The main benefit from this application came from the ability to continually maximise feed against the available regenerator excess oxygen – this was the critical constraint because the regenerator air blower was limiting throughput. In achieving this benefit the original multivariable controller design had opened the reactor outlet temperature

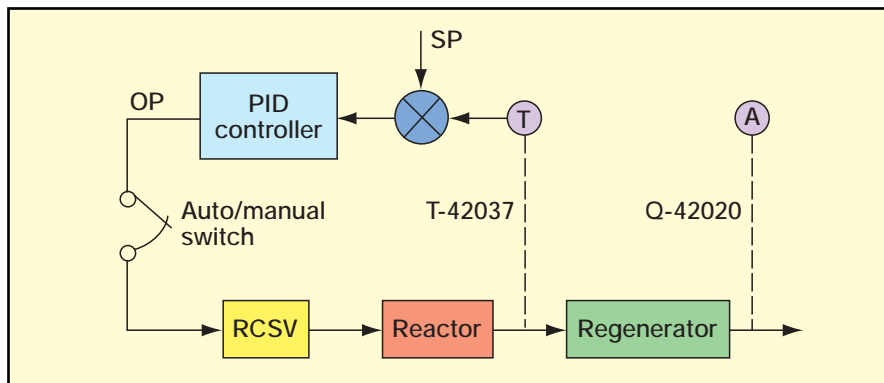


Figure 2 Reactor outlet temperature control schematic

PID controller. The reactor temperature controller behaviour was unstable because of mechanical disturbances in the process and this instability was affecting the stability of the excess oxygen; breaking the loop resolved this issue.

In 2002 a major turnaround on the FCCU resulted in significant process changes. The reactor technology was updated from a vented riser to a new vortex separation system. The catalyst type was changed, a new feed riser nozzle design installed and the air blower capacity increased. The changes also included strategic operating mode modifications including removal of the naphtha splitter column and routing of all naphtha to a new ultra low sulphur gasoline unit.

The physical process changes made in the reactor-regenerator section of the unit during the 2002 turnaround mean that the excess oxygen limit, which had been so critical prior to the unit turnaround, is no longer so difficult to maintain. The reactor outlet temperature has now been closed in the DCS (the interactive variation in excess oxygen due to temperature loop dynamics is less critical now). The fast DCS temperature loop means that the reactor temperature is in a tighter band than before the modifications; this is beneficial for selective reaction control.

The project to revamp the FCCU DMCplus controllers was carried out by Applied Manufacturing Technologies during the first half of 2003.

Model changes

Following the process changes, the DMCplus application on the reactor-regenerator and main fractionator sections of the plant remained unused, since the controller models were no longer representative of the plant. The main change required was to ensure that the reactor temperature control configuration in the DMCplus application matched the configuration now in use on the plant. Additional simplifications could be made; the original FCC_MAIN application has been split into three

separate applications to allow operators to work with them completely independently. The new controllers are FCC_RXRG (reactor/regenerator section), FCC_FRAC (main fractionator section) and FCC_DB (debutaniser). A separate FCC_C3C4 depropaniser has also been updated, but is not discussed here.

Temperature reconfiguration

Figure 2 shows a simplified schematic of the reactor outlet temperature control loop, also showing its impact on related dependent variables (for simplification, many other multivariable impacts on these dependent variables are not illustrated in the figure). The output of the reactor temperature controller T-42037 sets the regenerated catalyst slide valve (RCSV), which in turn varies the catalyst circulation rate and the resulting reactor temperature and many other process variables (the excess oxygen Q-42020 illustrated in the schematic).

Originally, the T-42037 PID controller was open-loop, with the DMCplus application setting the loop output directly; in the new configuration a significant change is that the reactor outlet temperature (ROT) T-42037 loop is now closed. The original multivariable controller matrix model had T-42037.OP as manipulated variable, with the T-42037.PV as a controlled variable, which was allowed to vary within a range specified by the operator. This is a significant change in the operation of the FCCU; without using UPID this modification to the DMCplus application would require a complete re-test of the entire application at a cost of several hundred thousand euros.

While UPID can perform model identification from plant test data and then allow reconfiguration of the model, it is not a necessary condition to have done the identification within UPID. The software can import FIR models from other sources; the control matrix model from the original DMCplus controller was imported into UPID and manipulated to match the current process operation.

Once imported, any model can be

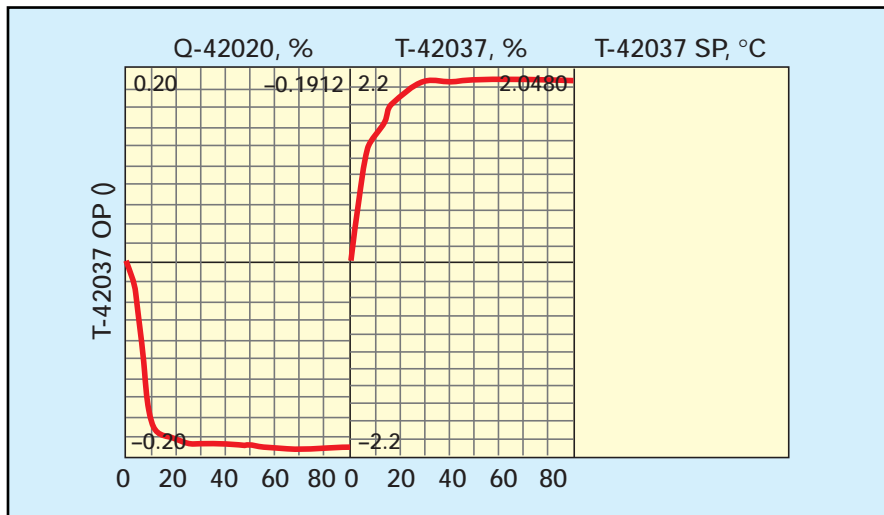


Figure 3 Dynamic matrix showing T-42037 variables with PID in open-loop (Manual)

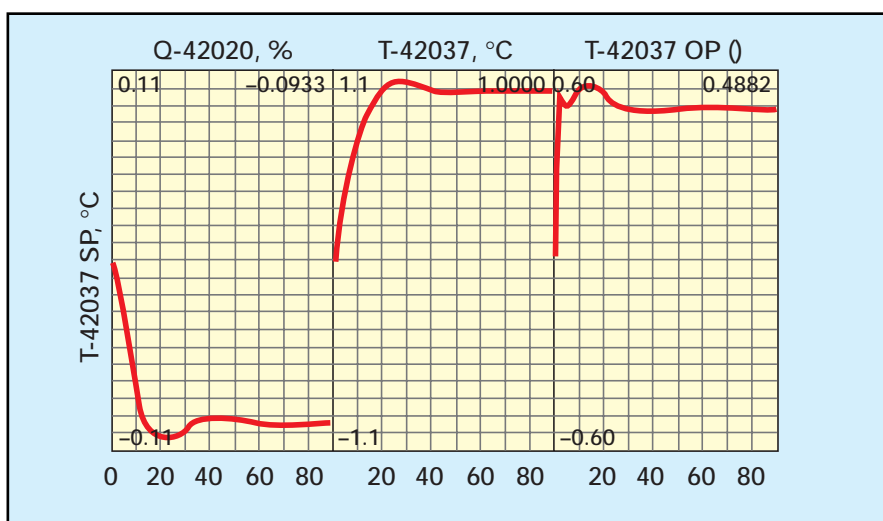


Figure 4 Dynamic matrix showing T-42037 variables with PID in closed-loop (UPID changes Manual to Auto)

adjusted using the UPID tools. Prior to changing the mode of the reactor temperature controller in the model, the whole matrix of model responses was reviewed and adjusted to ensure consistency, for example, between quality model responses and column temperatures responses in the main fractionator. This was an important step in ensuring the new “UPID converted” model would also have consistent gains and be representative of actual process behaviour.

The UPID model conversion process simply requires the addition of a T-42037.SP dependent column into the model matrix and inclusion of the PID loop tuning information. The dynamic matrix model section illustrated in Figure 3 shows the impact of a 1% change in the independent variable (reactor temperature controller output T-42037.OP) on the dependent variables reactor outlet temperature (T-42037) and excess oxygen (Q-42020).

(Each of the graphs represents the sequence of FIR coefficients in the

model relating the dynamic and steady-state relationship between the specific independent and dependent variables. In these dynamic matrix model plots the rows correspond to independent variables and the columns to dependent variables). A plant test PID data table in the UPID software requires configuring

for loops, which are involved in model changes, where the original mode for the PID control loops included in the MPC scheme is specified (the T-42037 loop mode would be manual). The specific PID algorithm used in the loop is specified in such a table; UPID includes generic control loop algorithms and several DCS specific control loop algorithms so that they can be applied to most commercially available DCS loops.

To perform the conversion the loop mode for T-42037 in the model is changed from Manual to Auto (where the new mode and tuning for PID loops is specified). This change triggers the convolution routines, which convert the open-loop models into closed-loop models. Note from Figure 4 that the independent variable following conversion is now T-42037.SP and that the corresponding response for the loop output behaviour (T-42037.OP) is now included as a dependent variable.

An additional dependent response for the temperature PV is calculated by UPID in the second column of Figure 4, which shows the closed-loop dynamics and a unity steady-state gain for a one degree change in SP. The figure also shows that the related dependent variable response for excess oxygen is automatically updated; compare the response in Figure 3 (column 1) with Figure 4 (column 1) to observe this feature. The automatic adjustment to the model is consistent on all related dependent variables; this has saved a significant amount of work in terms of re-step-testing, model identification and consistency checking.

Main fractionator controller

Other changes to the multivariable controller configuration include the removal of the open levels in the fractionator section of the FCC controller (Figure 5). The column bottom level L-42008 and overhead accumulator level L-42012 are now removed from MPC scope and the PID loops closed. Open

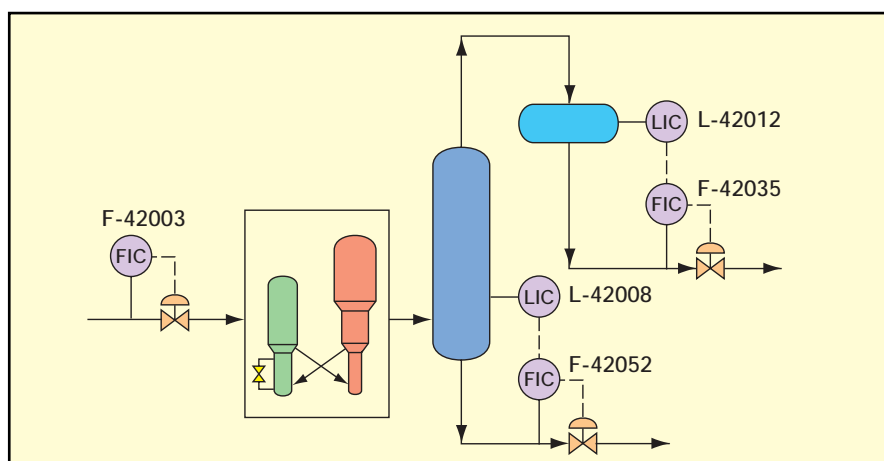


Figure 5 Main fractionator level controls

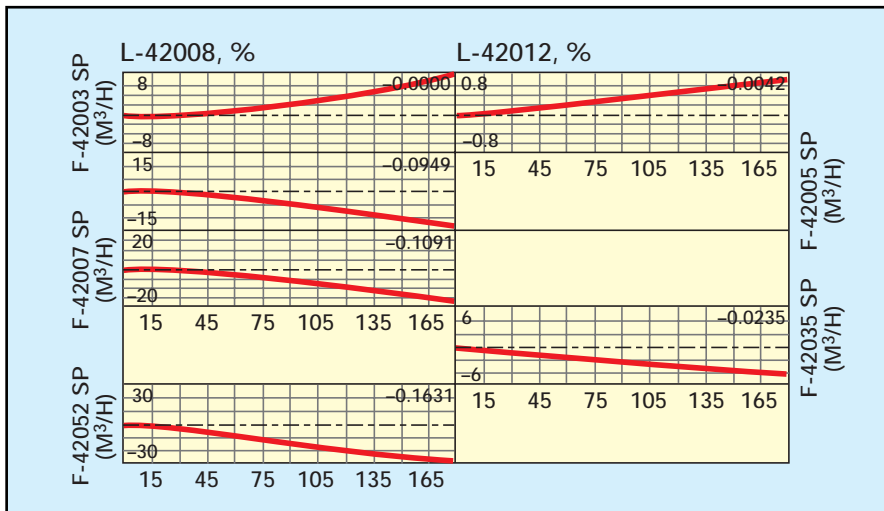


Figure 6 FIR models with fractionator accumulator level L-42012 and bottom level L-42008 in open-loop

levels controlled within MPC schemes often bring benefits, in terms of allowing the level to “drift” so that more important dynamic control objectives can be met. For example, the naphtha flow F-42035 was manipulated slowly by MPC to avoid flooding in the primary absorber, as well as for managing the main accumulator level.

Figure 6 illustrates the impact of the process flows (feed, heavy naphtha, LCO, light naphtha, HCO) on the overhead accumulator and column bottom levels when the DCS level controllers are in manual mode (open loop). The model shows that a 1m³/h increase in feed F-42003.SP (first row models as highlighted) causes both levels to increase their rate of change.

Following the FCCU revamp, there is at present no economic or operational benefit from having open levels. Therefore, the open levels were removed from the multivariable controller and returned to DCS control. UPID is used to close the level controller in the model to

match the new closed-loop DCS strategy. In Figure 7 the level responses change dramatically, representing the closed loop dynamics with the level controllers setting the flow controllers. The new model resulting from the closed level controllers shows L-42008 and L-42012 setpoints now appearing as independent variables.

The level SP to PV models (bottom two rows of FIR models in Figure 7) show a near unity response as expected, with oscillation reflecting the slow tuning of the loops. The dynamic response shown for the flows cascaded to the level controllers illustrate a classical behaviour for true material balance control applications – the flows must change quickly to move the level, and then return to their initial values.

(The SP to PV gain is not exactly unity because the model time to steady state should be much larger for exact modelling of these responses; these responses are in fact removed from the on-line scheme. Likewise the level SP MVs are

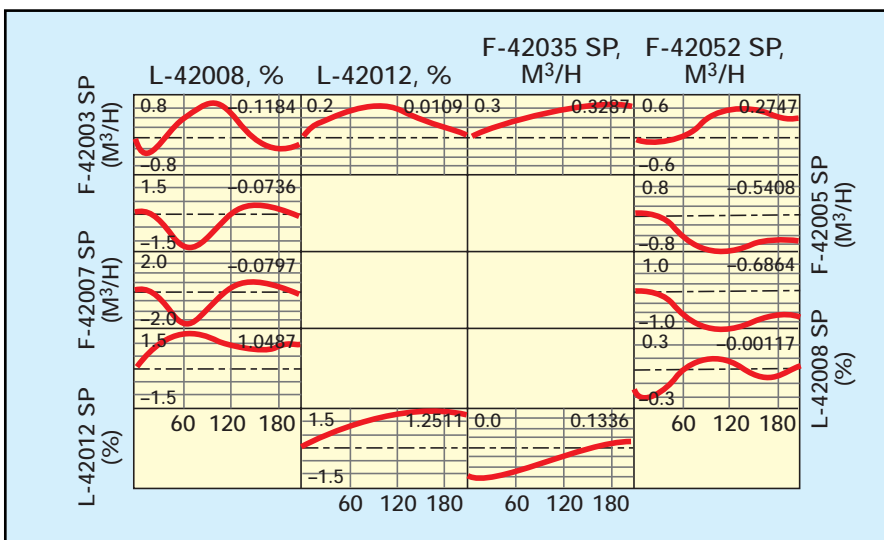


Figure 7 FIR models with closed-loop fractionator levels

removed from the final matrix model for the controller as there is no reason to manipulate them in the new DMCplus scheme.)

As a result of the model change the cascade flow slave setpoints F-42035 and F-42052 become dependent variables. The heavy cycle oil flow response F-42052 was retained in the model matrix to be used as a constraint in the reactor-regenerator section of the DMCplus controller. Naphtha flow to the primary absorber F-42035 was not added because the primary absorber is never constraining in the revamped FCCU. Returning these levels to DCS control was a key factor in breaking the original large controller into three smaller applications.

Naphtha stripper control

The FCCU DMCplus controller re-vamp project also took the opportunity to improve some of the operational issues, which needed to be addressed to improve operator acceptance of the multivariable controls. One issue was that the heavy naphtha stripper level control valve would periodically wind-up to 100% and operators had to intervene to cut back the heavy naphtha product flow – taking the flow out of DMCplus control. This was impacting on benefits, as the heavy naphtha should be maximised against its draw temperature (quality) to maximise profitability of the unit.

The fundamental problem that required addressing is that the control valve for the stripper level is very non-linear, but the DMCplus control technology uses linear FIR models to represent the control valve response. Control was improved by proper handling of this non-linearity, transforming the non-linear relationship into a linear relationship, which could be properly modelled and controlled within DMCplus.

Control valves may be characterised according to their inherent flow characteristics, which describe the flow rate through the valve as a function of the valve stem position with a constant pressure drop across the valve. These characteristics are described as decreasing, constant or increasing sensitivity. When control valves are combined with other equipment (pipes, orifice plates, bends etc), the installed flow-rate characteristics differ from the inherent characteristics of any single element in the system.

The effects of resistances resulting from piping, orifice plates, or other equipment in series with the control valve and the variation of available head with flow rate affect the flow vs. stem relationship.

Installed control valve characteristics can be approximated by linear or parabolic curves relating the fraction of

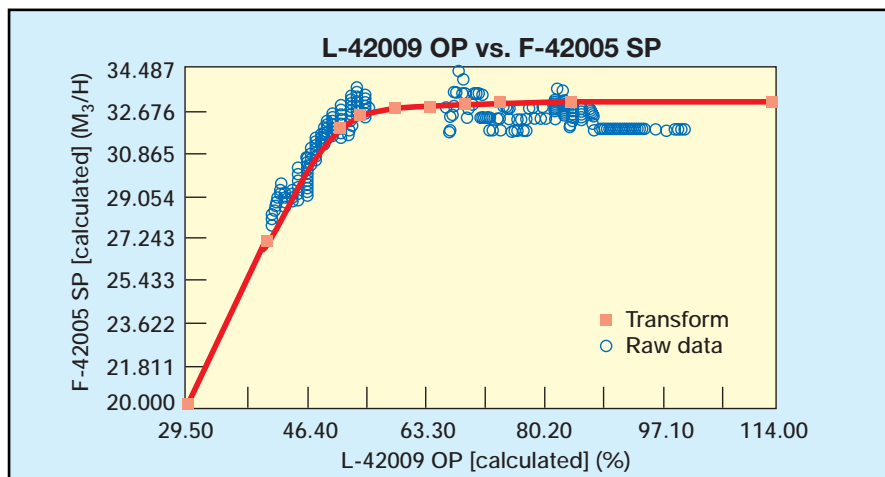


Figure 8 Piece-wise linear transformation

maximum valve stem position (L) to fraction of maximum rated flow (Q); these curves can be approximated by the following well known equations:

$$Q_{LIN} = \frac{L}{\sqrt{\alpha + (1 - \alpha)L^2}}$$

$$Q_{PAR} = \frac{L^2}{\sqrt{\alpha + (1 - \alpha)L^4}}$$

The α coefficient represents the valve head differential at maximum flow divided by the valve head differential at zero flow and is normally determined by regressing normalised measured data for the valve position and flow. A simpler approach is to use a piece-wise-linear (PWL) transform, since it is easier to prescribe exactly the desired transformation shape – the linear and parabolic approximations will often not fit all of the desired operating range.

UPID contains many standard transformations, including the linear and parabolic valve transforms and the PWL transform so that process data can be linearised prior to identifying FIR models. Noting that PWL is the most widely used approach, the UPID software contains a tool to automatically determine PWL coefficients.

The methodology is to plot the valve position against the measured flow through the valve; UPID contains a X-Y plotting tool for developing accurate PWL transformations. For example, the plotting tool can demonstrate how the heavy naphtha product flow F-42005 is plotted against the stripper level control valve L-42009.OP, with the non-linear behaviour being apparent such that for increasing valve opening beyond an output of 55%, there is no real flow increase (Figure 8).

The solid line through the valve vs flow data is fitted automatically by UPID; the user can choose the number of line segments and a smoothing penalty used in the derivation of the X-Y coordinates of the PWL function. The

smoothing penalty is added to the curve fitting algorithm and the resulting transformation is a trade-off between representing the data and obtaining a smooth function. In Figure 8 the automatically determined X-Y points were manually adjusted to extrapolate beyond the range of the data to ensure the transform would work even if new data was measured outside the flow/valve ranges used in the curve fit.

The resulting PWL transform can be implemented directly in DMCplus, or in the DCS loop, to give linearised valve behaviour. The stripper level controller output L-42009 has been transformed to better represent the severe non-linear behaviour when the valve opens beyond 55%. With the transform, illustrated in Figure 8, DMCplus knows that the flow F-42005 must be cut aggressively should the LIC output go beyond 50% (the upper limit set by the operator for this constraint). It is much easier to control the linearised valve position, and the DMCplus application is able to keep the stripper level properly on control by automatically adjusting the heavy naphtha flow by just the right amount to keep the valve on control when the stripper is limiting.

Conclusion

The UPID software is a remarkably useful and novel tool for the advanced control engineer. This article demonstrates its use in re-engineering control applications following base level control changes and its use in linearisation of process data so that commercial industrial model predictive control schemes can work well.

Having been re-engineered, the multivariable controller is once again maximising feed, pushing the unit against key constraints such as regenerator excess oxygen and main fractionator reflux temperature. The operators are now more comfortable using the DMC-

plus controllers, as they found the original large scope application difficult to understand. The cost of re-engineering the application using UPID was less than half of what it would have cost to re-step test and identify the new models.

The advanced FIR modelling features within UPID have not been discussed here. They include:

- Improved FIR smoothing
- Individual smoothing per curve
- Specifying dead-time per curve
- Model quality estimation
- Automatically optimising time to steady-state
- Imposing gain relationships.

Being able to generate open-loop responses for low level PID controls in the DCS gives the advanced control engineer a simple method for optimising all difficult-to-tune loops – a method which is fast and accurate. It is well known that good base level controls in the DCS are a critical success factor for MPC.

UPID can help an engineer to be successful by improving his effectiveness in setting up basic control schemes. The built-in graphical transformation utility gives a fast method for reviewing and, when necessary, automatically transforming all process relationships. Using this tool leads to more accurate controller models which in turn give improved on-line performance.

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