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Universal Process Identification Revamps FCCU APC

1 Introduction

This paper describes the first commercial project using a new advanced control technology to revamp an old model predictive control (MPC) application on a fluid catalytic cracking unit (FCCU). The project was carried out at the Gelsenkirchen Horst Refinery in Germany. Gelsenkirchen is one of four refinery locations of Ruhr Oel, a joint undertaking of Veba Oil Refining & Petrochemicals (VORP, a BP company) and Petroleos de Venezuela, S.A. (PdVSA). Veba Oel Verarbeitungs-GmbH, a VORP subsidiary, manages the plants in Scholven and Horst.

MPC was originally installed on the FCCU in 1997, using DMCplus™ technology from AspenTech. The application was configured as one large multivariable controller covering reactor-regenerator, main fractionator, unsaturated gas plant (debutaniser and naphtha splitter), with a separate controller for the depropaniser section. The main benefit from this application came from the ability to continually maximise feed against the available regenerator excess oxygen. In achieving this benefit the controller design had opened the reactor outlet temperature controller. At the time of the first application implementation there were mechanical problems with the riser temperature slide valve that caused the temperature and circulation to be unsteady. With the instability in the reactor/regenerator circulation rates, and temperatures and resulting instability in the regenerator excess oxygen while the riser temperature control loop was in automatic the original application was designed with the slide valve in manual.

In 2002 a major turn-around on the FCCU unit made significant changes; the reactor technology was updated from a vented riser to new vortex separation system, the catalyst type was changed and a new feed riser nozzle design installed. The changes included strategic operating mode and process equipment modifications:

- Naphtha splitter column removed
- All naphtha now routed to the new Prime G+ ultra low sulphur gasoline (ULSG) unit
- Naphtha 95% distillation point no longer a constraint
- Increased the air blower capacity
- With the refurbishment of the fresh catalyst slide valve the riser temperature could now be operated in closed loop temperature control in DCS.

After the turnaround the reactor temperature control loop was much more stable which greatly improved the ability to control the regenerator excess oxygen.

Most commercial MPC technology, including DMCplus, uses 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 dynamic 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 (for example flow, temperature or pressure controller *setpoints*) to obtain the dynamic response of the controlled variables (for example, analysed product qualities, temperatures, flows, levels, etc.).

Throughout the life of most model based predictive control applications one should expect to encounter situations where the regulatory control structure should be changed for optimal performance of the control system. Going from winter to summer operation can result overloading of condenser capacity resulting in column pressure control systems going "wide open". If the overhead pressure control system is operated in automatic some capacity must be given up in order to allow the pressure control loop the ability to adjust condensing rate to control pressure. In this situation the optimum operation is to run with the pressure control saturated and the pressure held just at maximum. Another example of this same type of situation occurs when light olefins prices rise and become high relative to FCCU gasoline product. In this situation the FCCU severity is raised to generate maximum light olefins. The gas compressor loading increases and the compressor suction pressure loop can become saturated. The FCCU economic optimum occurs when severity is raised until the pressure is at target and the gas compressor speed is at maximum. On the other hand if the gas compressor suction pressure control loop is operated in automatic some capacity must be given up in order to keep the control loop from saturating. In each of the examples above the model based control system should be changed to reflect the required change in the regulatory control system. The Horst FCCU Riser temperature control loop change is just another example of this type of situation.

In the past, changes in loop modes (e.g. 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 off-line following base level control system changes in the DCS (distributed control system), waiting until there is sufficient budget and/or time available to re-test the unit.

A new advanced control technology is available, where the MPC model can be directly manipulated 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™] 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 unique 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; which are combined to give the overall system 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 project to revamp the FCCU DMCplus controllers using UPID was carried out by Applied Manufacturing Technologies (AMT) during early 2003.

2 FCCU Process and Control Description

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 Veba FCCU processes approximately 4,250 tonnes of feed per day.

The FCCU consists of a reactor and a regenerator as illustrated in Figure 1. The Horst unit is a 'full burn' regenerator with the air for combustion is supplied by a steam turbine driven air blower. The main fractionator separates the reactor products into bottoms slurry oil, heavy and light cycle oil, heavy naphtha, and a wet gas stream that is separated in a gas recovery section.

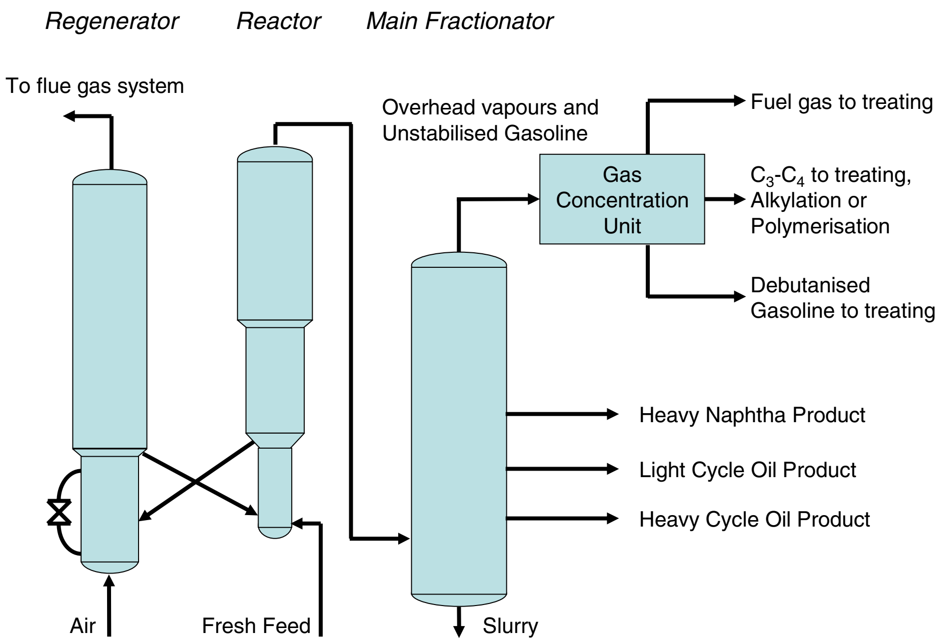


Figure 1 Fluidised Catalytic Cracking Unit (FCCU)

The DMCplus™ model predictive controller originally installed in 1997 had a scope which included:

- Reactor-regenerator
- Main fractionator
- Gas concentration (Debutaniser and Naphtha splitter)
- Gas plant depropaniser

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. Incorporating all plant sections into one large controller allowed the application to maximize feed and severity against downstream gas recovery section 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. To achieve this benefit the original multivariable controller design had opened the reactor outlet temperature PID controller in the DCS. The reactor temperature controller would not behave stably because of mechanical disturbances in the process and this instability was affecting the stability of the excess oxygen; breaking the loop resolved this issue.

The 2002 major turn-around on the FCCU made significant process changes. The physical process changes made in the reactor-regenerator section of the unit mean that the excess oxygen limit, which had been so critical prior to the unit turn-around, is no longer so difficult to maintain. With the riser temperature control loop operating much more stably than prior to the shutdown the reactor outlet temperature has now been closed in the DCS. The Horst FCCU operates very smoothly with temperature control closed in the DCS and now the original model needed to be changed to reflect this changed underlying regulatory control structure

The MPC revamp project set out to update the DMCplus model structure to match the new operating mode for the reactor temperature controls. Additionally the project identified key issues with the control models, and simplified the overall controller structure to improve operator acceptance. The following sections describe the steps taken, illustrating how the UPID technology was used.

3 Model changes using [UPID]

Following the process changes the DMCplus application on the reactor-regenerator and main fractionator sections of the MPC remained unused – because 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 – the reactor/regenerator section, FCC_FRAC – the main fractionator section and FCC_DB – the debutaniser. A separate FCC_C3C4 depropaniser has also been updated, but is not discussed in this paper.

3.1 Reactor 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.

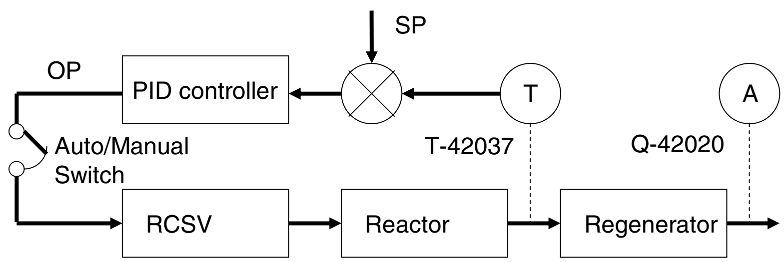


Figure 2 Reactor outlet temperature control schematic

Originally the T-42037 PID controller was open-loop, with the DMCplus application setting the loop output directly; in the new configuration 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 and without using UPID this modification to the DMCplus application would require a costly re-test of the application.

The control matrix model from the original controller was imported into UPID and manipulated to match the current process operation. First the existing dynamic model responses were reviewed 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 requires the addition of a T-42037.SP dependent variable column in the model, as shown in Figure 3. 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). (In these dynamic matrix model plots the rows correspond to independent variables and the columns to dependent variables.) The 'Plant Test PID Data' table in the UPID software, Figure 4, contains the regulatory control system specification for control loops involved in model changes; here we specify the original mode for the PID control loops included in the MPC scheme – (note that the T-42037 loop mode is Manual.) The specific PID algorithm used in the loop is specified in this table; UPID includes generic control loop algorithms and several DCS specific control loop algorithms so can be applied to most commercially available DCS loops.

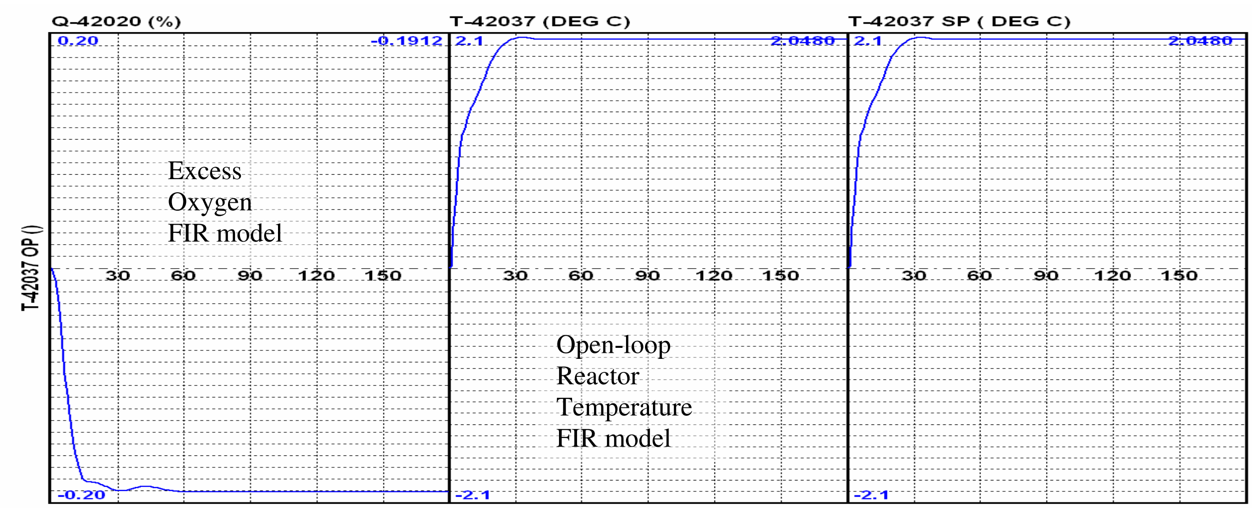
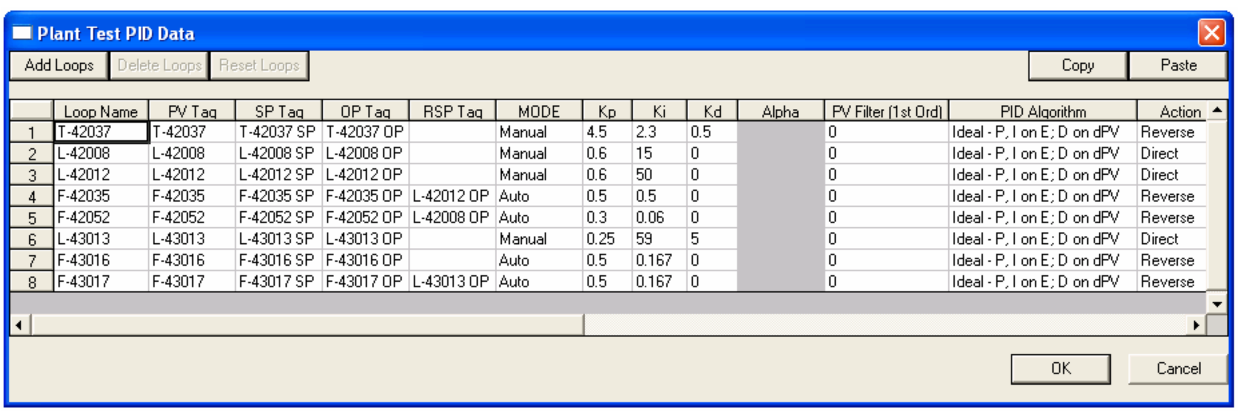


Figure 3 Dynamic matrix showing T-42037 variables with PID in open-loop (Manual)



Plant Test PID Data

Add Loops | Delete Loops | Reset Loops | Copy | Paste

	Loop Name	PV Tag	SP Tag	OP Tag	RSP Tag	MODE	Kp	Ki	Kd	Alpha	PV Filter (1st Ord)	PID Algorithm	Action
1	T-42037	T-42037	T-42037 SP	T-42037 OP		Manual	4.5	2.3	0.5		0	Ideal - P, I on E; D on dPV	Reverse
2	L-42008	L-42008	L-42008 SP	L-42008 OP		Manual	0.6	15	0		0	Ideal - P, I on E; D on dPV	Direct
3	L-42012	L-42012	L-42012 SP	L-42012 OP		Manual	0.6	50	0		0	Ideal - P, I on E; D on dPV	Direct
4	F-42035	F-42035	F-42035 SP	F-42035 OP	L-42012 OP	Auto	0.5	0.5	0		0	Ideal - P, I on E; D on dPV	Reverse
5	F-42052	F-42052	F-42052 SP	F-42052 OP	L-42008 OP	Auto	0.3	0.06	0		0	Ideal - P, I on E; D on dPV	Reverse
6	L-43013	L-43013	L-43013 SP	L-43013 OP		Manual	0.25	59	5		0	Ideal - P, I on E; D on dPV	Direct
7	F-43016	F-43016	F-43016 SP	F-43016 OP		Auto	0.5	0.167	0		0	Ideal - P, I on E; D on dPV	Reverse
8	F-43017	F-43017	F-43017 SP	F-43017 OP	L-43013 OP	Auto	0.5	0.167	0		0	Ideal - P, I on E; D on dPV	Reverse

OK | Cancel

Figure 4 PID loop configuration information in UPID

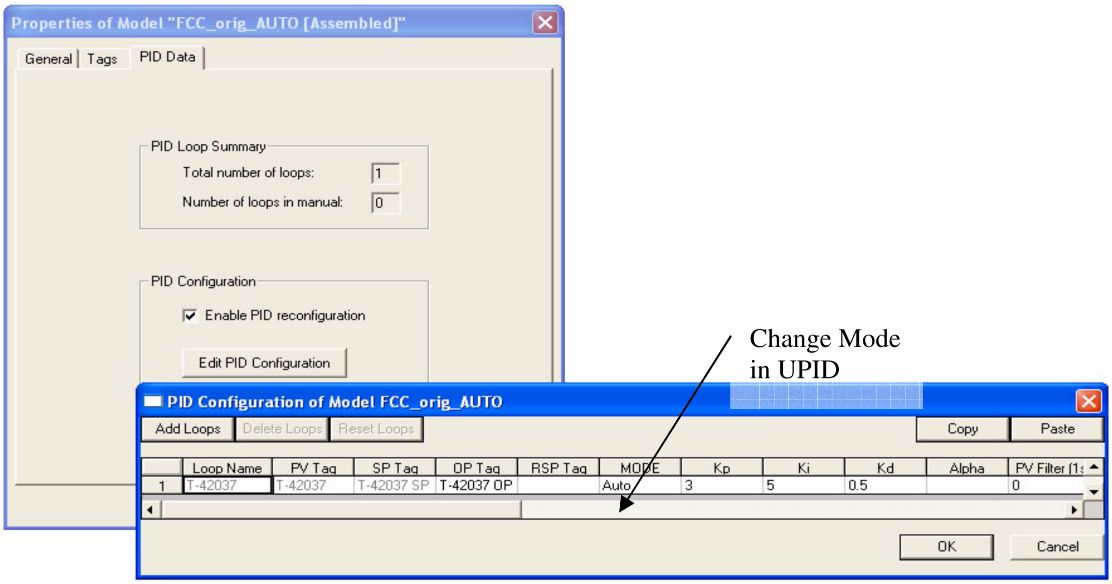


Figure 5 Model properties dialog and editing PID configuration

To perform the conversion the loop mode for T-42037 in the model is changed from Manual to Auto. To do this the desired loop configuration is set in the software as illustrated in Figure 5; specify the new control mode for T-42037 and its PID tuning constants. This change causes the software to reassemble the model into a new dynamic model where the underlying reactor temperature control loop is closed. Note from Figure 6 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, see Figure 6 (column 2) 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 6 (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.

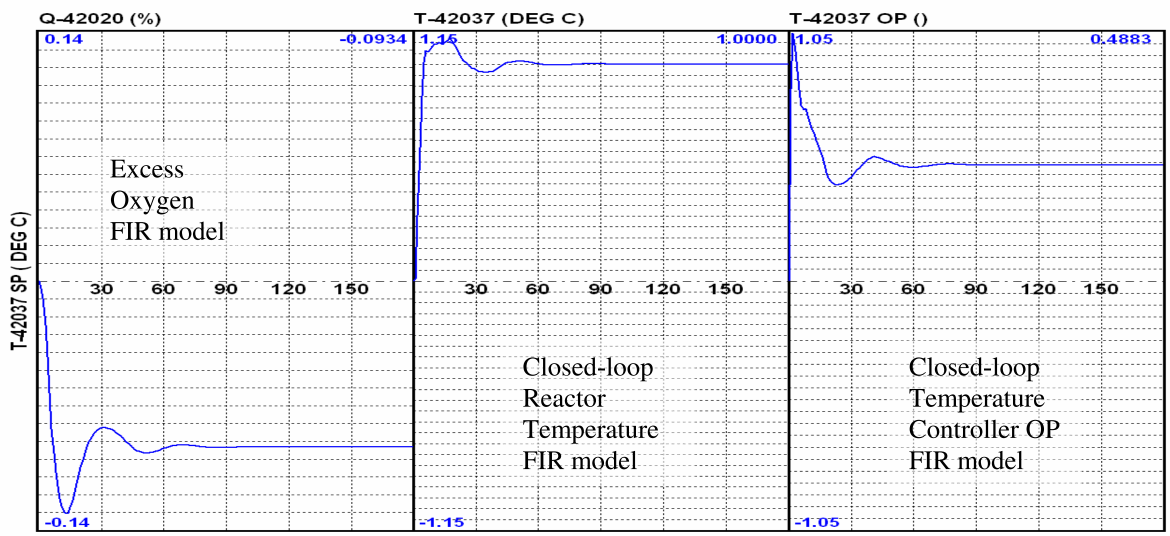


Figure 6 Dynamic matrix showing T-42037 variables with PID in closed-loop (UPID change Manual to Auto)

3.2 Reviewing the entire newly assembled model

As discussed earlier the riser temperature control loop (T-42037) was operated in manual mode in the original control system design. The purpose of this effort was to change this regulatory loop operating mode to automatic under the new control system design. Changing the base regulatory loop mode not only changes the response curve of the riser temperature controller but it also modifies many of the associate control loops. Any response curves representing independent variables that effect riser temperature while this loop is operated in manual would be expected to be different when the riser temperature control loop is operated in automatic. This is shown in Figure 7 below as those response curves in the column under T-42037 highlighted by shading. Any response curves representing response of the other dependent variables such as heavy naphtha draw temperature, main fractionator bottom temperature, LCO draw temperature (T-42050, T-42052 and T-42053) or fractionator overhead drum level (L-43013 (%)) to a change in the riser temperature control valve (T-42037.OP) would also be expected to change. The two cases previously discussed are directly effected response curves that one would expect to see changed as a result of switching the riser temperature independent variable from the valve (T-423037.OP) to the PID controller setpoint (T-42037.SP).

When a regulatory controller mode is changed there is a much broader effect on the model where in fact many more curves than those previously discussed are affected by this simple regulatory control system structural modification. Care must be taken to closely evaluate each of these response curves.

Take the case of the feed temperature (T-42010.SP); this variable affects the riser temperature (T-42037) as well as the main fractionator base temperature (T-42052). Since the reactor temperature control valve (i.e. the regenerated catalyst slide valve, T-42037.OP) also affects the main fractionator temperature when the riser temperature control loop is switched to automatic mode from manual one would expect a different response for the main fractionator temperature response to a feed temperature change. This happens as a consequence of the action of the reactor temperature control loop; feed temperature effects reactor temperature, the reactor temperature PID control loop adjusts the reactor temperature valve to bring the reactor temp back to set-point, the action of moving the regenerated catalyst slide valve cause a change in the spent catalyst temperature and the circulation rate resulting in a regenerator temperature change. In fact any independent variable that affects reactor temperature when that loop is in manual will cause the regenerated catalyst slide valve to move when this control loop is in automatic. Once the reactor temperature is switched to automatic all such independent variables that cause the reactor temperature controller to move the slide valve will now affect every dependent variable that responds to a change in reactor temperature slide valve movement (T-42037.OP). The response curves shaded in Figure 7 will all be affected by switching the reactor temperature controller from manual to automatic in this same way. Even the independent/ dependent variable pairs where there is no response but where this same type of relationship exists will have a resulting curve when the control mode is changed as described above.

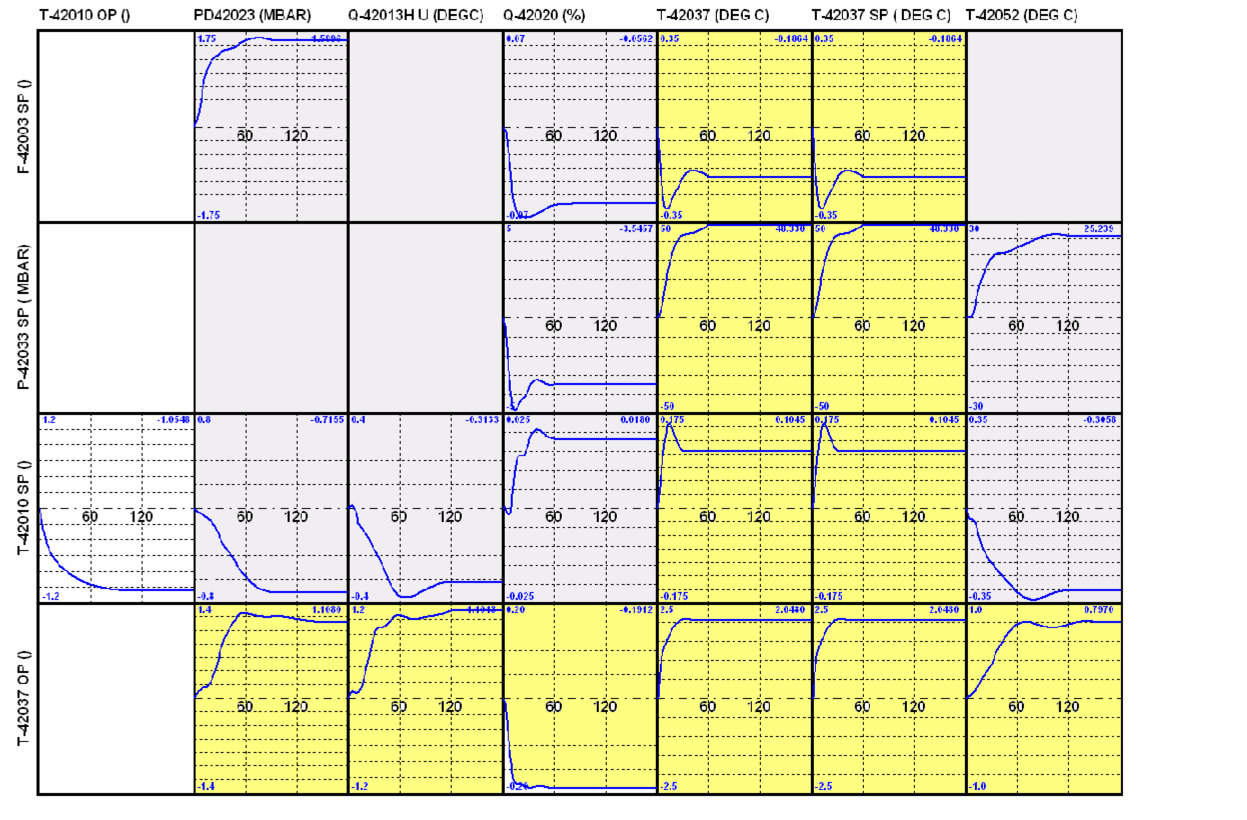


Figure 7 Section of Horst FCCU Dynamic Model in Manual

All possible response curves will be generated by the software package, regardless of how significant any particular response is in terms dynamic path and final steady state gain. The quality of the resulting new or modified response curves are a function of the quality of the original model. Since these models are empirical linear approximations there is always some amount of error present in the model, which is dealt with via real time feed back in the controller. The combination of small errors in the original model resulting from using these empirical approximations can generate models that are obviously incorrect. In addition, since the software accounts for every response possible as outlined in the example above, many small responses curves will be generated and added to the transformed or newly assembled model. Each new response curve that is created when changing the structure of the underlying regulatory control system must be reviewed for reasonableness and significance. If there are significant and important models that are obviously incorrect this means that the original model had significant errors in it. In this case the original dynamic response must be re-examined and the original error corrected. Utilizing this software in this way will uncover model errors that can then be corrected to improve the overall MVC system performance.

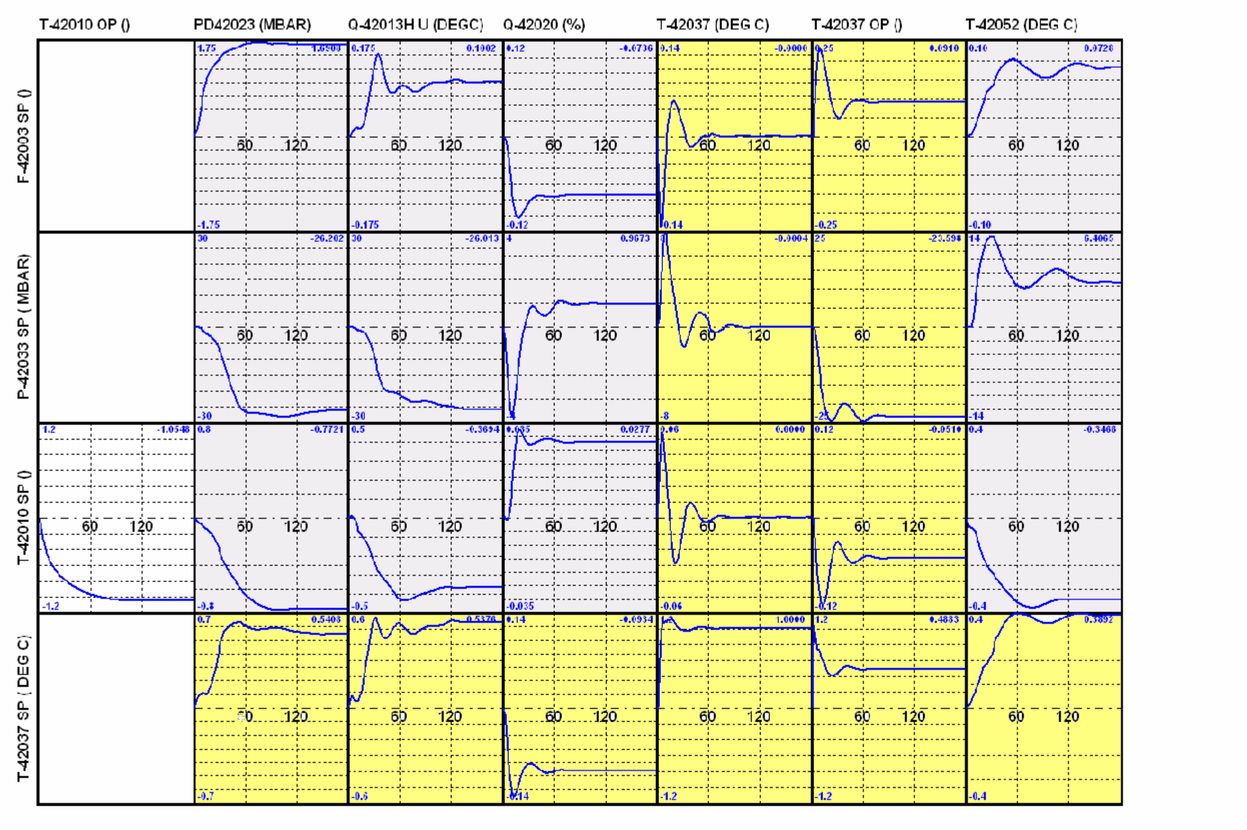


Figure 8 Section of Horst FCCU Model with reactor Temperature Control Loop (T-42037) in AUTO

In the case of the small response curves that are generated when transposing the model as a result of changing the regulatory control structure; a judgement must be made regarding whether or not to retain these small response curves. Each curve must be reviewed and carefully considered and those that are deemed to be insignificant should be removed from the model.

3.3 Main fractionator controller simplification

Other changes to the multivariable controller configuration include the removal of the open levels in the fractionator section of the FCC controller (see Figure 9). The column bottom level L-42008 and overhead accumulator level L-42012 are now removed from MPC scope and the PID loops closed. Open 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.

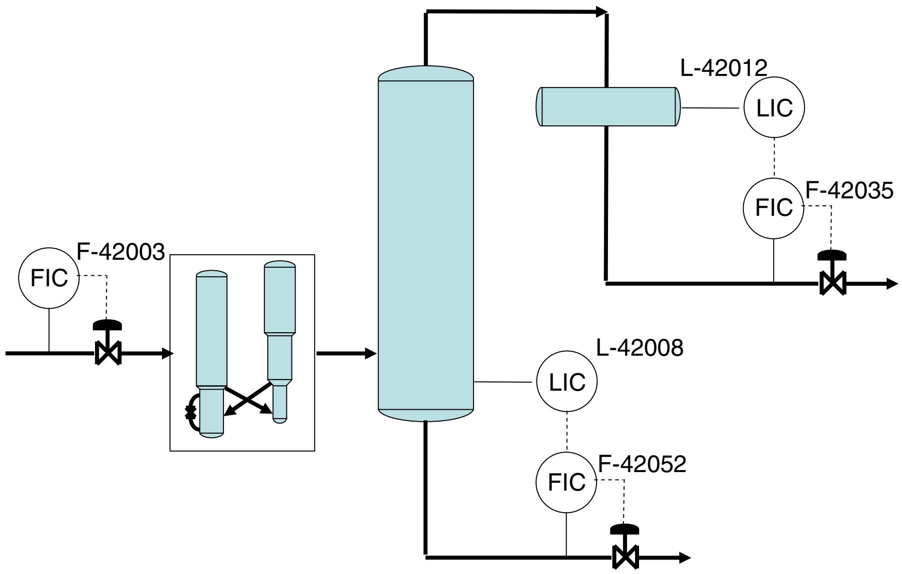


Figure 9 Main fractionator level controls

Figure 10 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 (i.e. open loop).

Following the re-vamp of the FCCU there is presently 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 11 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 11) 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 setpoint MVs are removed from the final matrix model for the controller as there is no reason to manipulate them in the new DMCplus scheme.) Other responses in the updated model show the disturbance impact of feed, heavy naphtha, LCO light naphtha and HCO – the curves oscillating as the slow level controller tuning reacts to the disturbances.

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 as a controlled 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.

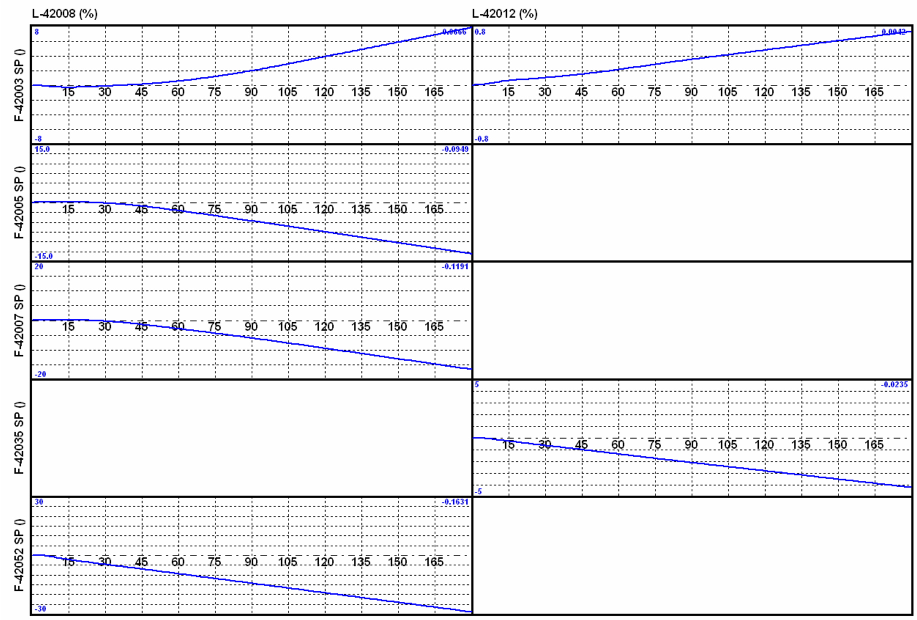


Figure 10 FIR models with fractionator accumulator level L-42012 and bottom level L-42008 in open-loop

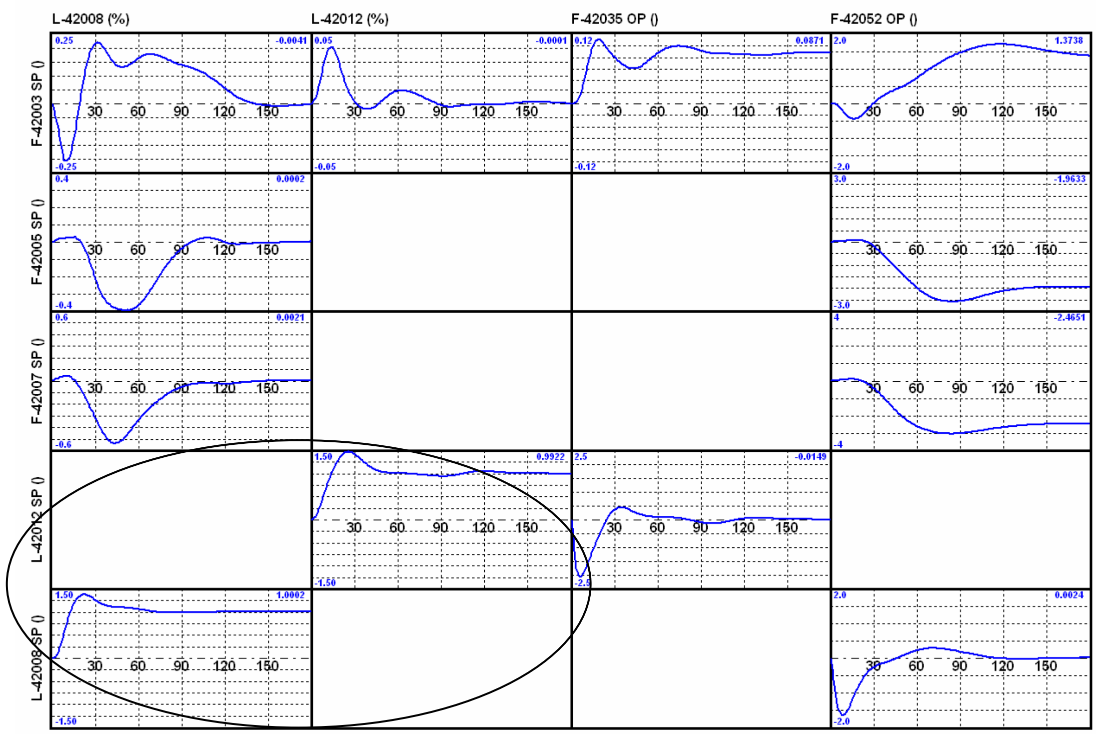


Figure 11 FIR models with closed-loop fractionator levels

4 Loop tuning

A novel feature of the UPID identification technology is that it can be used to assist in the tuning of difficult PID loops, either in single loop or multi-loop situations. The process involves either:

- developing a closed-loop model of the response with its existing PID tuning parameters, and then determining the open-loop response by opening the model in UPID, or
- developing an open-loop model directly.

In either case a short plant test can obtain the required data for model identification. There are many PID tuning calculation methods which can be used to derive initial tuning coefficients based on open-loop response, even guessing is a safe way to start in the UPID environment. UPID does not contain any 'loop tuning' algorithm. It contains most of the widely used PID algorithms, so these can be used in conjunction with any external method for determining the P, I and D coefficients of the controller. Changing any of the PID coefficients, or specific PID algorithm, results in an update of the convoluted FIR model; the FIR coefficients are updated and a new closed-loop response presented to the user.

Thus, using the UPID tool it is easy to manually adjust PID tuning and see the predicted impact on the loop response without running a dynamic simulation, and without having to use a trial and error approach on the actual plant. In our experience the PID settings determined using the UPID approach are normally acceptable first time – avoiding the need for trial and error tuning on the process. Figure 12 illustrates an investigation into shortening the rise time of the reactor temperature controller (the original tuning was left alone, as the predicted impact on the loop output was too aggressive).

This approach has delivered significant performance gains on other recent projects conducted by AMT, both in terms of loop performance and in project team effectiveness during pre-testing.

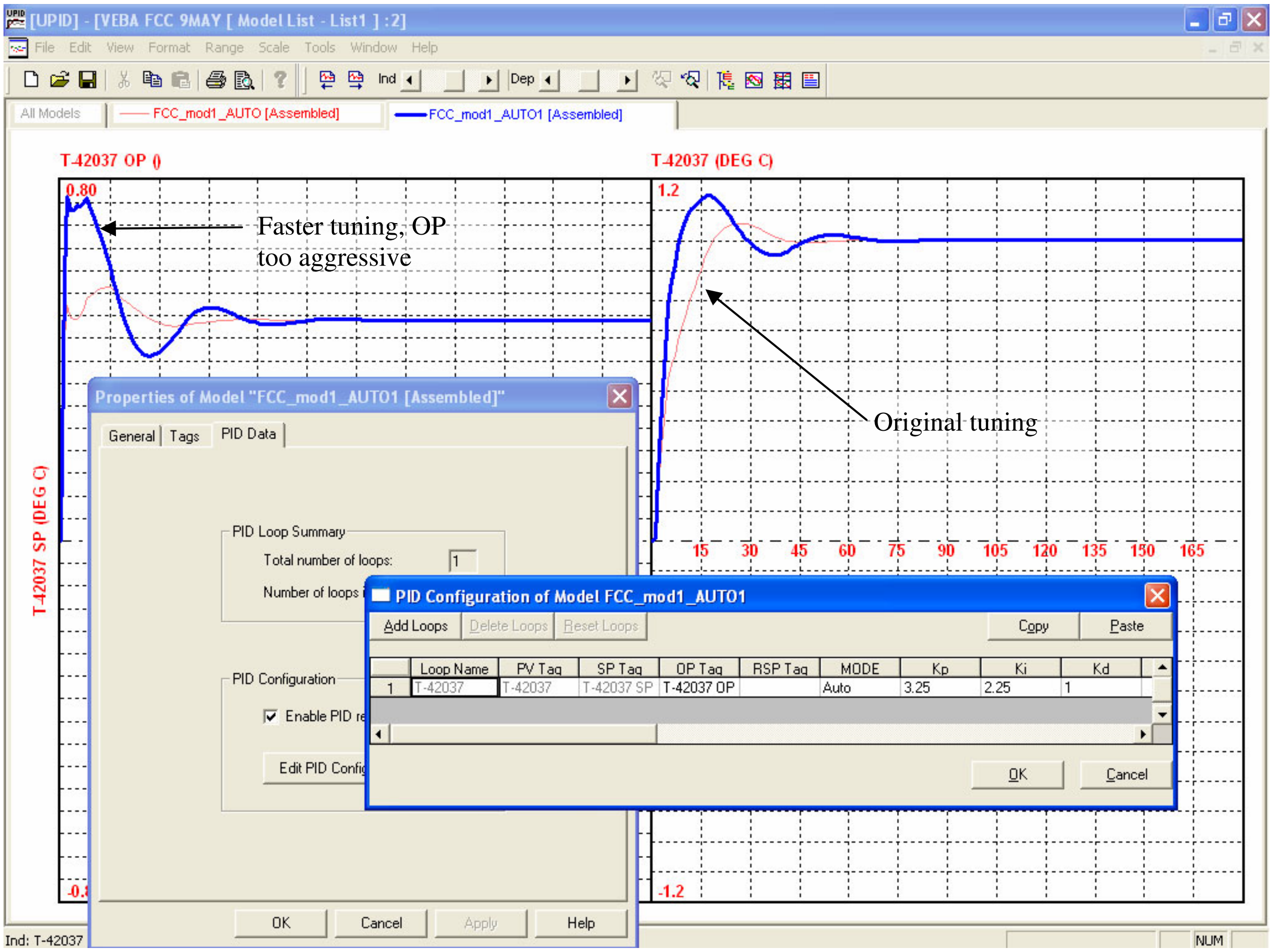


Figure 12 Using UPID to evaluate loop tuning coefficients

5 Concluding Remarks

The UPID software is a remarkably useful and novel tool for the advanced control engineer. This paper has illustrated its use in re-engineering control applications following base level control structure and tuning changes, its use in the linearization of process data, and its use in improving the performance of base level PID tuning, so that commercial industrial model predictive control schemes can work well even when the underlying control systems are changed.

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 DMCplus 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
- Specify dead-time per curve

- Model quality estimation
- Automatically optimise time to steady-state
- Impose 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 optimizing 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 in UPID 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.