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IMPLEMENTING ADVANCED PROCESS CONTROL FOR REFINERIES AND CHEMICAL PLANTS

Abstract

In today's globally competitive marketplace, chemical plants and refineries are looking at new ways to increase plant efficiency, production rates, safety and reliability. In the process control arena, base-level PID tuning optimization, APC (Advanced Process Control) and MPC (Model Predictive Control) remain attractive and under-utilized options. This paper describes various new and novel ideas harnessing the power from primary PID control improvements and APC/MPC implementation. The techniques and methodologies described can increase a plant's profit margin from 2 to 10 %. Spectacular increases in plant profits as high as 15 to 20 % (equivalent to 2 mil Euros/year) have been achieved and demonstrated in some cases. A few real examples from actual industrial plants have been provided in this paper.

***Key words**: closed-loop system identification, closed-loop controller tuning and optimization, advanced process control, model predictive control*

Importance of PID control quality monitoring and optimization

DCS (Distributed Control System) and PLC (Programmable Logic Controller) are now ubiquitous in all plants. Currently, the power of the DCS and PLC is efficiently harnessed for the safe and reliable operation of the plant, described in Schuppen et al. 2011, Bolton, 2009, Cauffriez et al. 2004 and Rullán, 1997. However, less than 15% of plants use modern software for PID control quality monitoring, PID tuning optimization, advanced process control (APC) or optimization. This is because of several reasons:

a) Absence of engineering knowledge and unavailability of practical and robust process control software tools for system identification, PID/APC parameter optimization and control quality monitoring.
b) New and inexperienced or changing personnel in the control room.
c) Expensive and complex vendor software.
d) Academic courses that are research oriented and not catering to the needs of the control room environment.
e) Fast changing software and hardware technologies.
f) Running plants conservatively because of fear of causing shutdowns and plant problems.

Plant management is often unaware of the poor condition of the process control quality in the plant because of absence of easy methods to trend and quantify the control quality metrics. A plant can have anywhere from few tens to several thousands of PID controllers and tens to hundreds of APC schemes. No matter how large or small is the plant, monitoring of all PIDs and APC schemes is a prerequisite for continuing to achieve the benefits from PID tuning optimization and APC implementation. Changes in process conditions, market/economic conditions, hardware and equipment changes result in a need for changes in the PID and APC controller tuning parameters. Often these changes are rarely made and optimal tuning parameter calculations are not precisely calculated due to lack of available tools and/or the required skill set.

The first step in improving the plant control quality is to install online control quality monitoring software. The second step is to analyze process data on poorly performing control loops and identify the system dynamic, described in paper of Ingimundarson et al. 2005. The third and final step is to calculate the new and improved PID or APC tuning parameters and then download them into the DCS or PLC. Following these three steps systematically can tremendously improve a plant's control quality and in turn, the plant's profit margins.

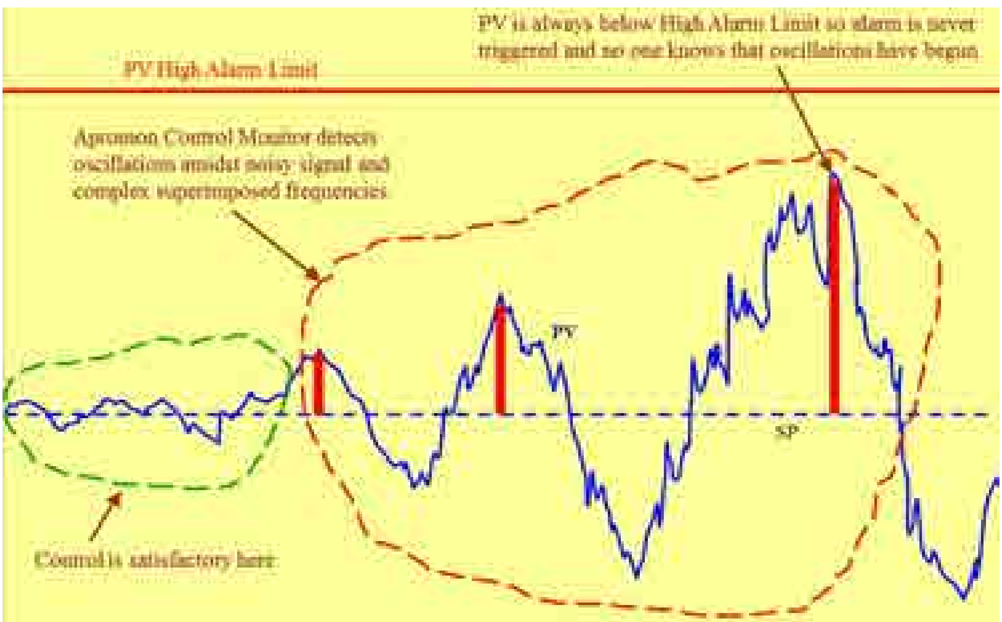

Figure 1: Online identification of Control Deterioration

If there is no way to generate a list of poorly controlling PID control loops and/or APC loops, then there is no way to start the improvement process. To assist plants and refineries on monitoring their control loops quickly and easily, PiControl developed Apromon software. Apromon connects easily to any DCS or PLC in just half day using standard OPC connectivity. It can monitor anywhere from 5 to over 5000 PIDs in a plant. When there is deterioration in a PID control loop, the deterioration typically manifests as oscillations, sluggish control and sustained deviation from setpoint (target). See Figure 1 illustrating a case when control was satisfactory at one time and then oscillations begin later. The highest and lowest PV values happen to lie within the alarm limits so that the operator is never alerted. Despite the noise and complex superimposed frequencies, upon detection of deterioration in any PID loop, software package Apromon sends automatic emails to the technicians and control engineers. All this happens automatically with no human intervention. Corrective action with manual tuning changes or automatic changes using online adaptive control helps to maintain tight control action. This methodology of providing proactive instead of reactive maintenance and support is critical for running the plants efficiently. Such regular monitoring and real-time alerting technology helps to ensure that manpower is properly focused on the right (poorly performing) PID controllers and APC schemes. The control quality criteria reported by Apromon are listed below. Apromon generates a daily report showing all criteria for all PID and APC tags.

Benefits of improved process control in chemical plants

The benefits of PID tuning improvements and APC/MPC implementation can be understood by examining Figure 2.

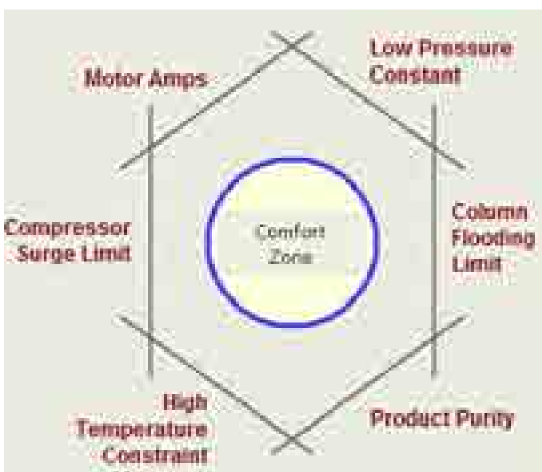

Figure 2: Moving the plant in direction of increased profit

Every process has various constraint limits that if not controlled tightly will result in process upsets, equipment failures, shutdowns or other undesirable situations. If a process is moved in the direction of some constraint to maximize production rates and thereby profit, one or more of the constraint limits may be violated. Any operator wants to avoid such violations and hence the plant is often run in a “comfort zone”, away from the direction of increased efficiency, higher production and max. profit.

If PID tuning control quality can be improved and a stable, effective APC or MPC system is installed, then the amplitude of deviation or oscillation can be reduced significantly. Reduction in amplitudes by a factor of three to ten times is possible and has been demonstrated in several plants. This allows the process to be run closer to the constraint limits or shutdown limits as shown in Figure 3. The ability to reduce the amplitude of deviation or oscillation allows raising the production average in a plant, described in Chen, 1989. This increase can be typically 2-10 % and can amount to several hundreds of thousands of Euros of increase in the plant’s profit margins.

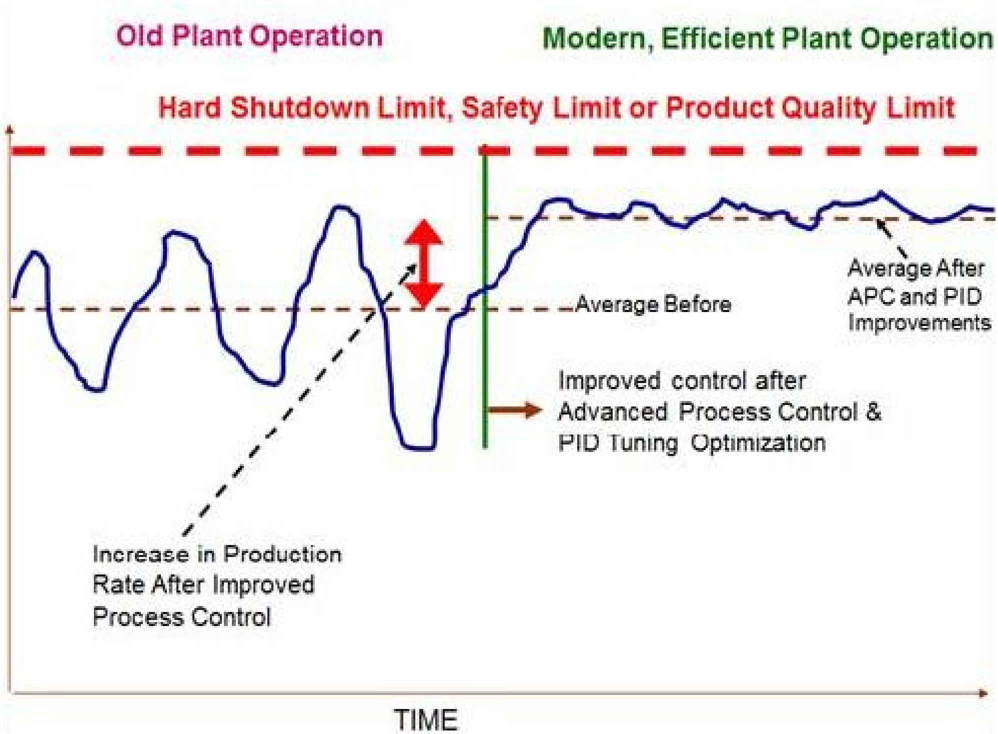

Figure 3: Monetary benefits because of PID control improvements, APC and MPC

Advantages of DCS/PLC-resident APC

Most plants are neither equipped with PID monitoring/tuning software nor any APC schemes. Some plants implement MPC/DMC (Dynamic Matrix Control) systems without paying detailed attention to the base-level PID and APC control quality. This is a common mistake and always results in failure to harness the full benefits of the MPC/DMC system. Though MPC/DMC technology is widely applied and successful in most truly multivariable interactive processes in refining and large scale olefins plants, powerful and robust DCS/PLC resident APC schemes can be quickly and easily designed and commissioned for many chemical processes. The main reason why most plants do not use PID tuning and APC applications at the DCS/PLC level is the absence of easily available practical, robust and user-friendly software for system identification and control optimization that is capable of working in the practical control room environment. Process control data from industrial plants are often complicated by noise, nonlinearities and complex disturbances that make system identification and tuning optimization complex and difficult. Furthermore, availability of quality data required for accurate system identification is often hindered by the inability to conduct conventional intrusive open-loop tests due to process sensitivity and non-linearity. The software tool needs to be capable of processing short-duration closed-loop data amidst significant measured and/or unmeasured disturbances. If system dynamics can be identified amidst the above-listed challenges, then it is possible to estimate the optimal PID tuning and APC system controller parameters. These parameters include cascade and slave PID tuning parameters, constraint override selector control scheme parameters, feedforward parameters and APC/MPC tuning parameters.

This paper describes novel methods for system identification, PID/APC tuning parameter optimization and APC implementation at the DCS/PLC level. It describes powerful, modern methods for multivariable, closed-loop system identification followed by PID/APC tuning parameter calculations. The novelty of this methodology is in the ability to identify multivariable closed-loop process model parameters amidst disturbances with ultra-short duration data which has been previously not possible. The technology and underlying algorithm is packaged inside PiControl's Pitops software product. With this approach using Pitops, closed-loop APC schemes can be easily designed and implemented for a variety of chemical, petrochemical and refining processes at lower cost, faster commissioning time with high on-stream factors. The uniqueness of Pitops is the ability to identify gains, dead times and settling times on multiple-input models simultaneously using complete closed-loop data. The APC schemes designed using Pitops improve control quality, which in turn increase the plant's profit margin.

This paper also describes new, novel and practical approaches to optimize the performance of base-level PID controllers which is different from other developed methods, described in Ljung, 1999. It describes new methods for using the DCS or PLC directly to implement robust APC strategies to improve process control to maximize the plant's profit margin.

It also shows a unique method to improve dynamic models in commercial MPC/ DMC systems. The above concepts are explained below using practical illustrations on the design and implementation of APC schemes directly inside a DCS or PLC. The first illustration describes production rate maximization of a chemical plant making lubrication oil. The second case describes product grade transition control and simultaneous production rate maximization of a polymer process. Both illustrations involve APC design and implementation directly inside a DCS/PLC without the use of external host computers or any commercial vendor software. The new approach succeeds even when the process data are superimposed and complicated by noise and unmeasured disturbances typically present in most plants. Such DCS/PLC-resident APC systems can produce as high as 10 % or more increase in the plant's profit.

Production rate maximizer APC inside the DCS or PLC

Figure 4 shows a process flow diagram of a lubrication oil manufacturing process.

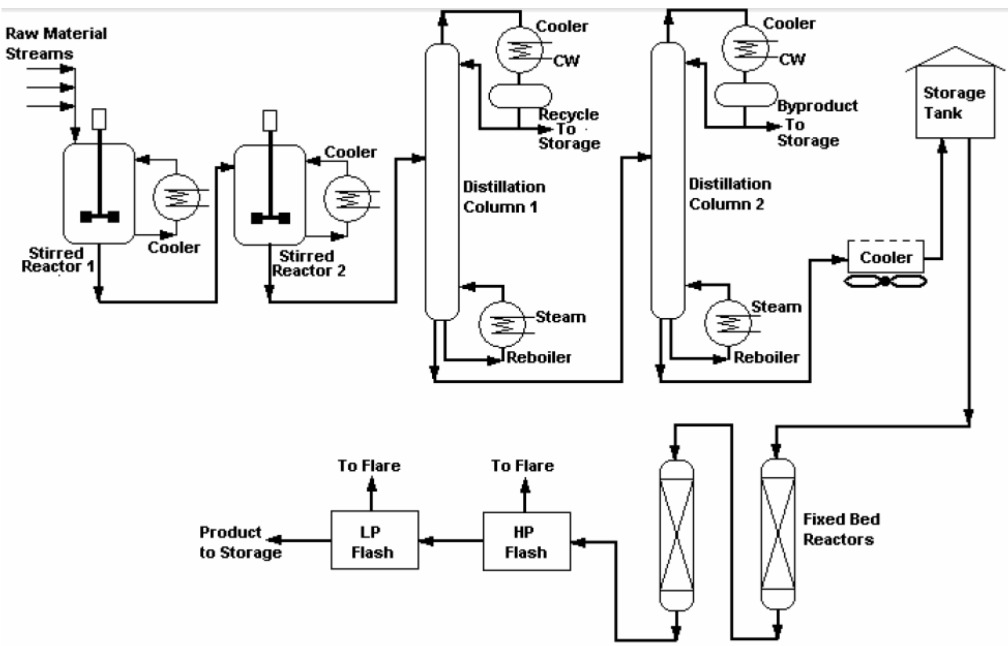

Figure 4: Process flow diagram for lubrication oil manufacturing process

Raw materials are mixed in the first liquid phase stirred reactor. Effluent from the first reactor is allowed more residence time in a second reactor. Both reactors are equipped with coolers to remove the exothermic heat of reaction.

The effluent from the second reactor is then separated in two distillation columns before being cooled and sent to storage as intermediate product. The intermediate product is fed to fixed bed reactors and the lights are flashed in two separator drums before cooling and final product storage. The plant is sold-out and any increase in production rate can significantly increase the plant’s profit margin. The plant is run in semi-continuous mode, meaning that the process conditions on the reactor, distillation and other major equipment are changed every few days make different product grades using different catalysts and raw materials. Absence of effective closed-loop APC or an inattentive operator can result in constraint violation on one or more of the CVs. This in turn can result in one or more of the following undesirable conditions:

a) Producing off-spec/lower-cost product;
b) Loss of expensive recycled material and/or lower product;
c) Damage to catalyst causing loss of catalyst life;
d) Causing unsafe operating conditions.

Table 1: Control matrix for lubrication oil manufacturing process

CVs / MVs	Desired Production Rate	Reactor#1 Level	Reactor#2 Level	Reactor#2 Temp.	Distil#1 Temp.	Distil#1 Online Anlysis	Distil#2 Temp.	Distil#2 Online Anlysis	Interm. Cooler Temp.	Storage Tank Level	Feed Limit to Fixed Bed Reactors	HP Flash Drum Pres.
Feed Flow to Reactor #1	X	X		X	X		X		X		X	
Feed Flow to Reactor #2			X	X								
Feed to #1 Distillation					X	X						
#1 Column Reflux					X	X						
#1 Column Steam					X	X						
Feed to #2 Distillation							X	X				
#2 Column Reflux							X	X				
#2 Column Steam							X	X				
Feed to Fixed Bed Reactors										X		
HP Flare Flow												X

The control matrix for this process is shown in Table 1. The boxes labeled "X" in Table 1 show a relationship between a given pair of variables. The variables in the left vertical column are the MVs (Manipulated Variables) and the variables in the upper horizontal row are the CVs (Controlled Variables). The MVs will be manipulated by the new DCS-resident APC strategy. The CVs are variables that the new APC strategy will strive to maintain at the desired limits specified by the production team.

The dynamic relationship between the pair of variables for every box checked with an "X" needs to be determined. Most chemical processes, including this process can be characterized by one of the common industrial process transfer function models (zero, first or second order transfer functions with dead-time). Pitops software was used to identify the transfer function process model parameters using existing historical data stored in the plant's data historian described in Pitops manual.

Most control systems today have the capability for archiving historical data for several years. The long-term data trending and archiving is adequate for the design of APC schemes, especially for APCs with a slow dynamics with settling times in the range of 20 minutes to over 4 hours. This settling time range covers most chemical processes. Figure 5 shows real plant data collected from a plant's data historian. The reboiler steam flow was pulsed and the process data was used to identify the process model parameters between the reboiler steam flow and the distillation temperature. This whole identification process was completed in less than ten minutes. The identified process model parameters are shown below:

Dead Time = 5.3 minutes,
Process Gain = 0.251 °C / (m^3/h),
1st order Time Constant = 45.2 minutes,
2nd order Time Constant = 67.1 minutes.

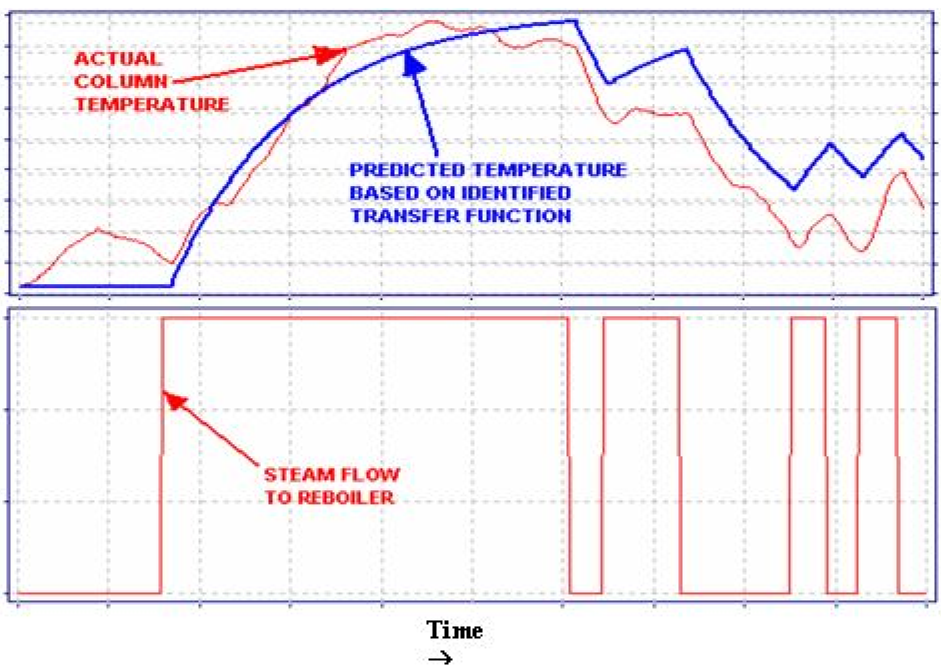

Figure 5: Open loop process model parameters identification

Many modern processes are highly heat and mass-balance integrated. Abrupt step-tests on the MV shown in Figure 5 may not be possible due to process sensitivity and inherent instabilities characteristic of heat and mass-balance integrated modern processes. Some refining processes, cryogenic distillation processes and exothermic polymer reactors are good illustrations where intrusive plant tests can severely disturb the whole plant and may not be permissible. Even in such processes that pose challenges for the identification of the process model, there are windows of data available from past history where a grade change, throughput change or some other change may have produced good quality closed-loop data with useful cause-and-effect information. Figure 6 shows such an example.

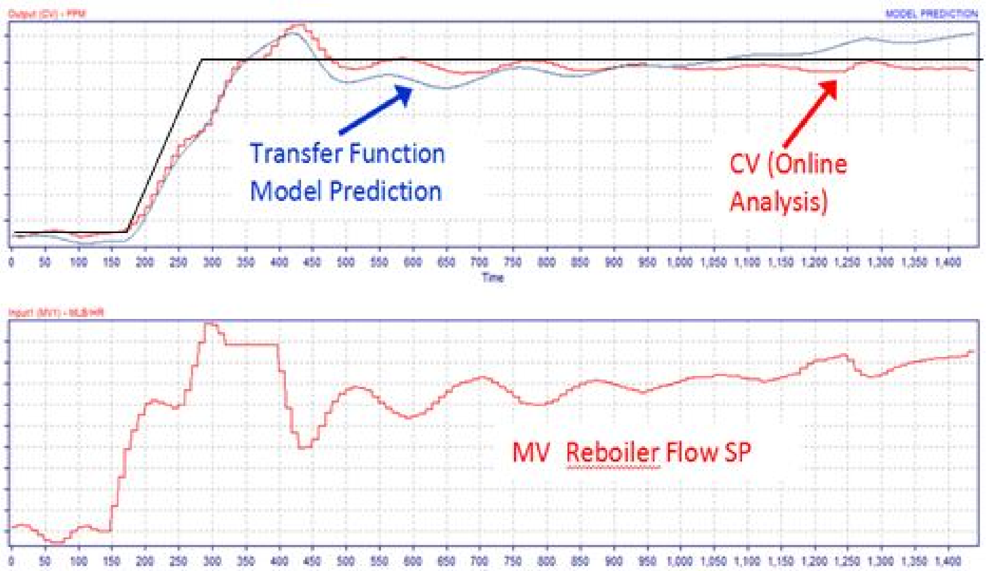

Figure 6: Closed-loop process model parameter identification

Notice that this illustration has neither open-loop pulses in manual mode on the controller output nor closed-loop pulses on the setpoint in auto mode. The new system identification algorithm using Pitops identifies the process model purely based on this closed-loop short duration data (without any step tests at all). The time needed for this calculation is just ten minutes. The process model parameters identified by Pitops are all calculated and displayed entirely in the easy-to-understand time domain, without the need to use or convert from the more complicated and abstract Laplace (S) or the Discrete (Z) domain.

The transfer function parameters identified by Pitops are shown below:

Dead Time = 19.7 minutes,
Process Gain = 196.9 ppm / (m^3/h),
1st order Time Constant = 80.5 minutes.

Figure 7: Simultaneous multi-input closed-loop process model parameters identification with short duration data and no pulse tests

Long-duration repetitive intrusive pulse tests on the process as required by currently known products and technologies are not required by the new Pitops algorithm. Figure 7 shows a complex multiple-input process model parameters identification illustration around a distillation column, described in paper of Guidorz et al. 2003. The three MVs (reboiler duty, reflux flow and the column feed flow) are changing simultaneously in complete closed loop mode.

The changes in the three MVs impact the CV (distillation column temperature). The new Pitops algorithm is able to identify all three process models simultaneously amidst the high frequency noise and unmeasured disturbances typically encountered in distillation processes. It also identifies the dead time automatically in addition to the other process model parameters. The identified process model parameters are shown in Table 2.

Table 2: Process model parameters for multivariable closed-loop case

	Dead Time (minutes)	**Process Gain**	**Time Constant** (minutes)
Reboiler Duty	7.7	13.9 kcal/h	26.2
Reflux Flow	5.4	-1.09 m^3/h	12.7
Column Feed Flow	6.9	-5.78 m^3/h	15.5

These above illustrations show how to use historical data from the past and identify process dynamics quickly and easily. This ability to use short duration multi-input closed loop data amidst unmeasured disturbances and even oscillations allows improving slave/cascade PIDs and also in the tuning of DCS/PLC-resident APC strategies.

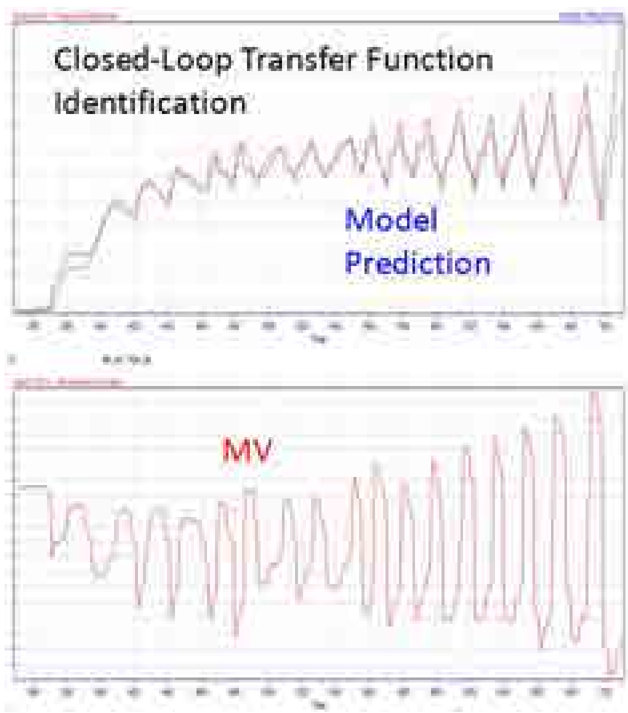

Figure 8: Closed-loop process model parameters identification amidst oscillations

Figure 8 illustrates closed-loop process model parameter identification amidst strong oscillations seen on another control loop. The identified process model parameters are shown below:

Dead Time = 1.3 minutes,
Process Gain = -121.7 psi/%,
1st order Time Constant = 17 minutes,
2nd order Time Constant = 27 minutes.

Note that the various illustrations above are based on pure closed-loop data without any step tests at all – neither on the setpoint in auto mode nor step tests on the controller output in manual mode. This clearly sets a new, unprecedented and unparalleled standard in the industry for successfully identifying multivariable closed-loop system dynamics without the need for any step tests. This ability was previously considered to be not possible and has now been successfully demonstrated using the above illustrations. The new fully closed-loop system identification technology does not need step tests in the plant. It can use historical data from the past stored in a plant's data historian and without the need for new, lengthy and difficult step tests in the control room. This is a major improvement in process control technology providing new, novel and powerful method for fast and accurate system identification followed by PID tuning optimization and APC implementation.

Initial Control Performance

Improved Control after Model Identification and Tuning Optimization

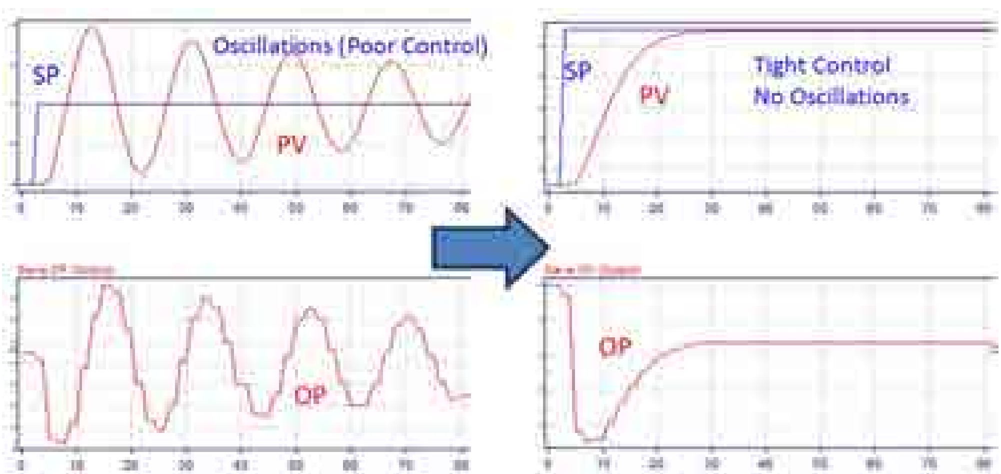

Figure 9: PID tuning optimization results

Knowing the process model parameters, a control simulation can be quickly setup using Pitops as shown in Figure 9. The left side of Figure 9 shows poor control with bad tuning parameters causing oscillations. The right side shows optimized PID tuning parameters showing non-oscillatory and crisp control. The optimal tuning parameters were calculated using Pitops based on the closed-loop process model identification using the oscillatory closed-loop data shown in Figure 8. Table 3 shows the initial PID tuning parameters causing the oscillatory control and the Pitops optimized tuning parameters producing non-oscillatory, tight and crisp control.

Table 3: PID parameters before and after controller tuning

	Initial PID Tuning	Optimal PID Tuning
Controller Gain (P)	1	0.269
Integral (I)	20 minutes	24.8 minutes
Derivative (D)	0 minutes	0 minutes

Once the process model relationships have been identified, APC schemes can be implemented on a DCS or PLC. Such an APC system can be comprised of various strategies including model-based control, inferential control, adaptive control, feedforward control, constraint over-ride control and cascade control.

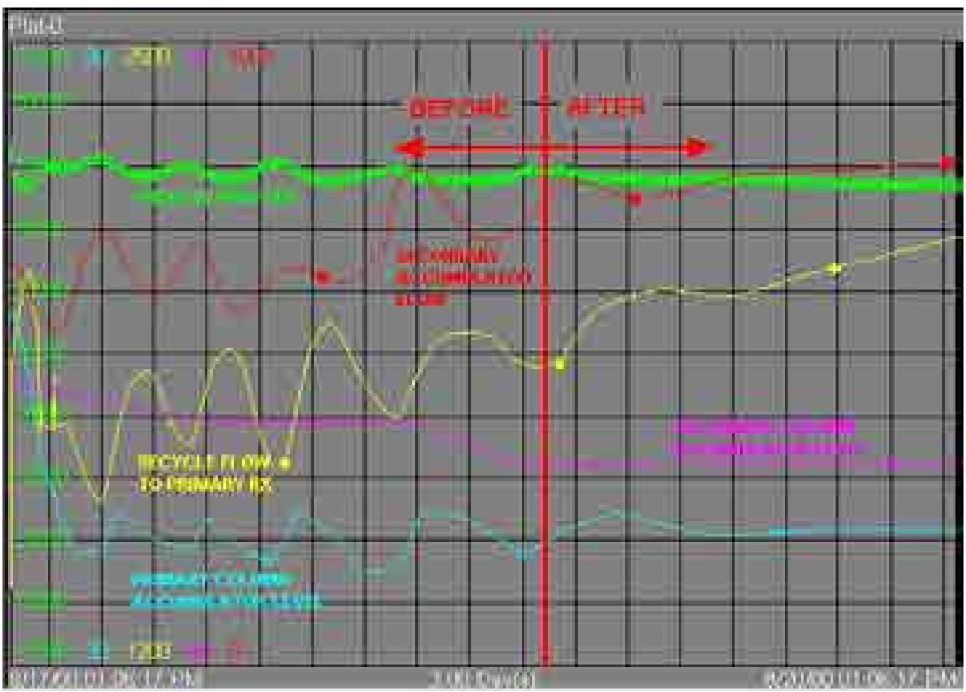

Figure 10: Improved lubrication oil plant operation after PID optimization and APC (real plant data)

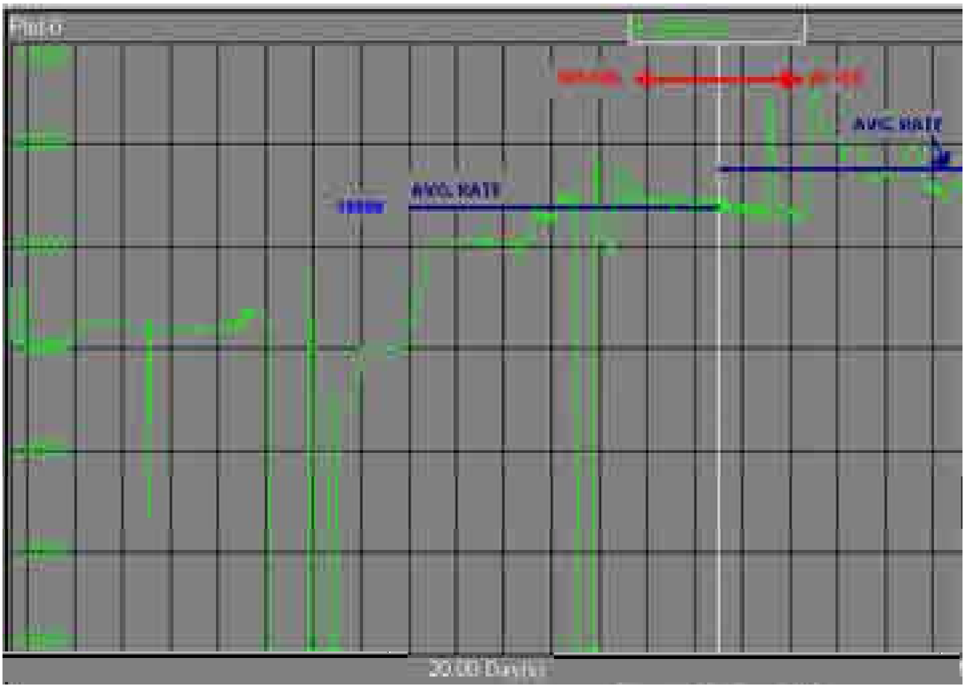

Figure 11: Increase in production rate because of process control (real plant data)

Figure 10 shows the plant's oscillatory behavior before PID tuning optimization and then the plant's improved and more stable behavior after PID tuning optimization. The increase in stability and removal of the oscillations enables the plant to push against operating constraints and thereby achieve higher production rates. Figure 11 shows the increase in the plant's production rate and also the improvement in the plant operation because of reducing upsets and problems caused by poor control. After implementing the process control optimization and APC, a noticeable increase in operating efficiency and plant throughput could be successfully demonstrated.

The power and value of this APC design is that all control parameters are calculated scientifically and precisely based on the accurate knowledge of the process dynamics. The key for APC success is the ability to identify process dynamics accurately using short duration closed-loop data. The lubrication oil APC system consisted of 12 MVs (manipulated variables), 28 CVs (controlled variables) and 11 FFs (feedforwards). The entire APC was installed on the existing DCS. It took just 2 weeks to install the DCS configuration, identify all models and to calculate the control parameters in the DCS. The duration for commissioning and tuning was one additional week. The operator feedback was excellent and the onstream factor was close to 100 %. The APC system produced a 5 % increase in production rate, equivalent to annual recurring benefits of 1,300,000 Euros.

Inferential model-based control on exothermic reactor

An application of model-based control is illustrated on a fluidized bed reactor control scheme shown in Figure 12.

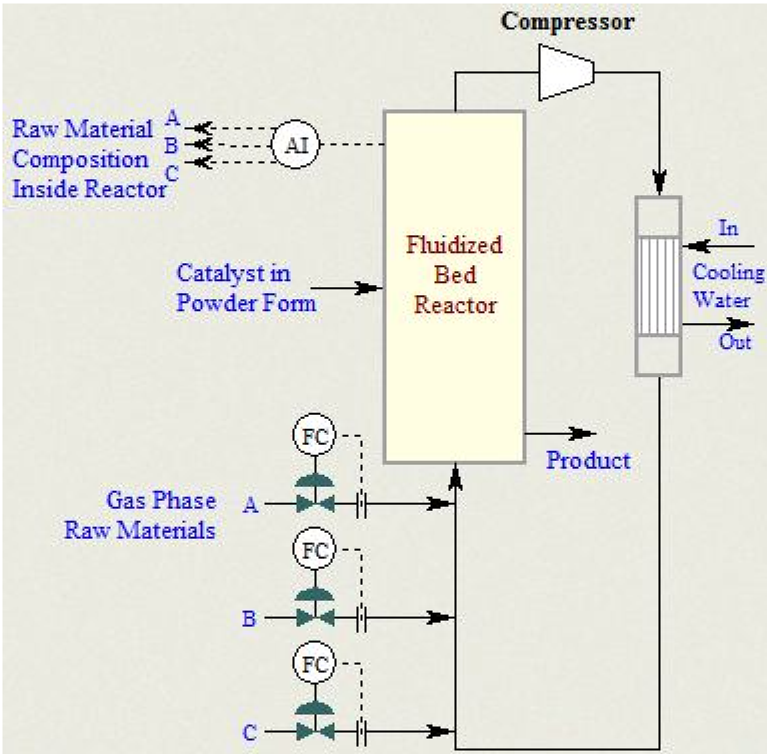

Figure 12: Fluidized bed exothermic reactor control

There are two opportunities for improving the control in above reactor. One is in the control of the reactor gas composition that impacts the product quality and the second on the temperature control of the exothermic reactor where a TC-TC cascade manipulates the cooling water flow into the external cooler on the recycle stream. The gas composition control is characterized by very slow dynamics with a settling time of five hours. The PV signal comes from an online gas chromatograph with a sample time of 10-20 minutes. With the slow dynamics and sample time, tight control with pure base level control is not possible.

Using time series process data, we can calculate the process dynamics and also develop a mass balance model of the process and then incorporate it into the inferential model-based control scheme, illustrated in Figure 13.

Reduction in the composition control deviation has been demonstrated to be reduced by ten-fold with significant improvement in product quality.

The second opportunity for improving control is on reactor temperature. The temperature control in exothermic reactors can be very complex because of the open-loop unstable process dynamics as shown in Figure 14 and 15 and described in Panda, 2009.

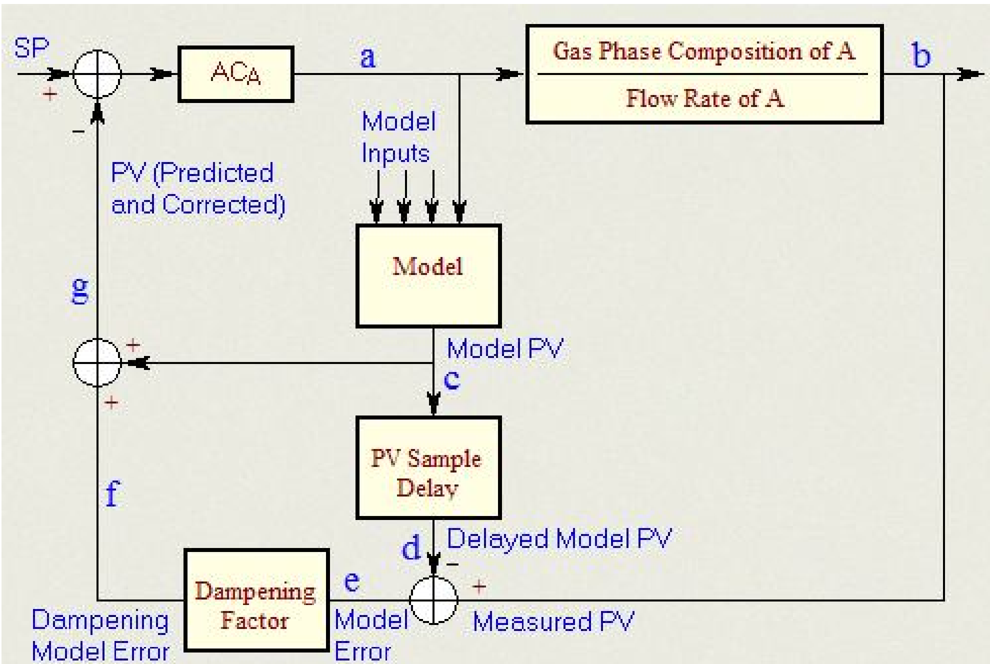

Figure 13: Inferential model-based control on fluidized bed exothermic reactor

The optimal tuning of this TC-TC cascade is difficult without the Pitops-based system identification, control optimization tools and the methodology described in this paper. Closed loop data analysis of the TC-TC data are shown in Table 4.

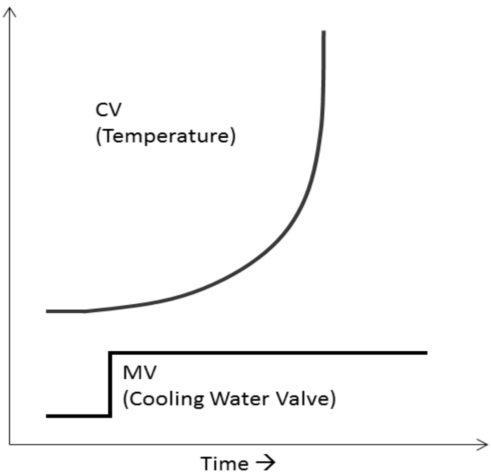

Figure 14: Open-loop unstable process model dynamics

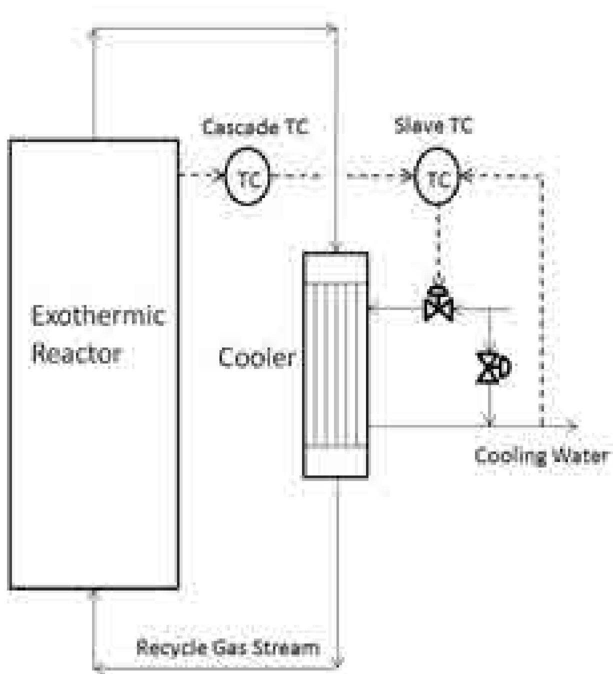

Figure 15: Exothermic reactor TC-TC cascade control with open-loop unstable dynamics

Table 4: TC-TC cascade process model and PID tuning parameters

	Slave Loop	Cascade Loop
Dead Time	2.3 minutes	3.2 minutes
Process Gain	1.01 °C/%	0.595 °C/%
1st Time Constant	38.4 minutes	95.3 minutes
2nd Time Constant	0 minutes	9.1 minutes
Controller Gain (P)	9.25	0.33
Integral (I)	8 minutes	15 minutes
Derivative (D)	0 minutes	0 minutes

Note that it is impossible to make open-loop step tests on such exothermic reactors since the process is extremely sensitive and unstable and even a few minutes of open-loop (manual) mode can cause the reactor temperature unacceptably outside safe operating range.

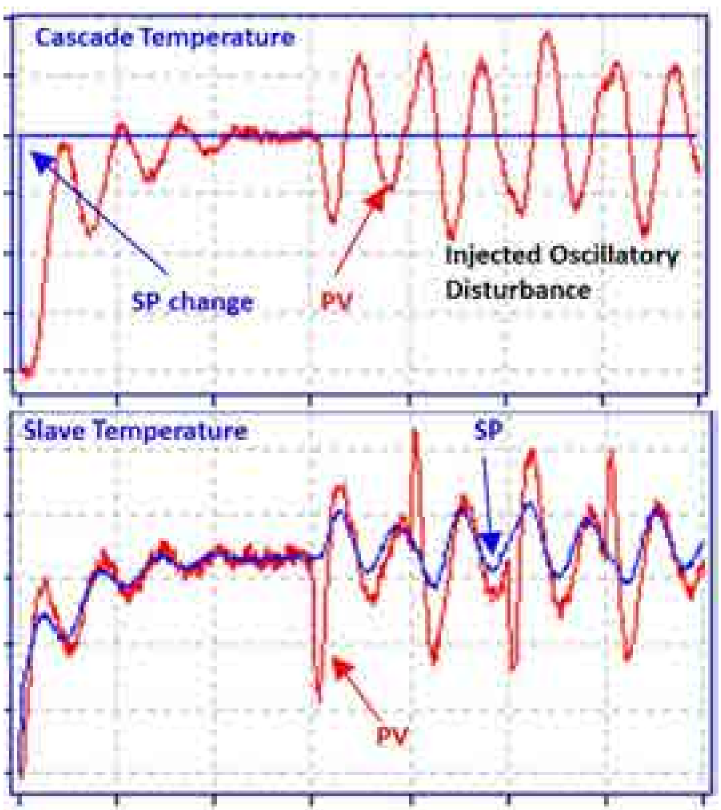

Figure 16: TC-TC cascade and slave dynamics on open-loop unstable process

In Figure 18, the PID tuning parameters of the cascade TC and the slave TC have been simulated and optimized for a simultaneous setpoint change, pulse and ramp disturbances and signal noise that are typically present in such control loops. The optimization of the TC-TC tuning parameters is not heuristic-based but is based on optimizing the PID parameters based on a custom simulation comprising of a typical SP change and typical disturbances, exactly as encountered in the real plant on the real DCS. The control parameters calculated based on the simulation and optimization will match exactly with the real controllers inside the DCS. This is one of the powerful practical functionalities available to the control room personnel, made possible by the use of Pitops.

The oscillations seen on the right side of the trend plot in Figure 16 are because of pulse and ramp disturbances that are injected into the control simulation. The inherent nature of the process model and presence of dead time make it impossible to completely eliminate these oscillations. However the controller tuning parameter optimization by Pitops minimizes the amplitude of these oscillations.

Improving MPC models using closed-loop identification

The ability to identify the process models with Pitops using short-duration closed-loop data has other novel and powerful benefits. A notable one is the ability to improve the process model gains identified by MPC systems. During the model identification stage of a MPC project, the MVs typically cannot be stepped more than about 2-3 % around their setpoint. This limit avoids process upsets and allows holding the effect of the steps for a longer time required by conventional technologies for identifying the steady state gains and the settling times. However, during the closed-loop control action of the MPC system, multiple MVs are moved and the magnitude of the MV changes is much larger than the 2-3 % change made during the model identification stage. This difference in the move sizes can result in significant differences (as large as 100-200 %) in the closed-loop gains of various critical MV-CV relationships causing the closed-loop control response of such a controller to be oscillatory and unacceptably poor.

An instance of such a problem is shown in Figure 17 below showing closed-loop DMC control action on an air separation unit. On a demand change, the DMC moved various MVs resulting in strong undesirable oscillations caused by inaccurate models. The data shown in Figure 17 is complete closed-loop data with an active DMC without any step changes. The oscillations and poor control were traced to erroneous models identified based on the 2-3 % (small) step tests during identification that were significantly smaller than the model response seen during the larger steps (30-40 %) made by the active DMC in closed-loop mode.

Using the novel closed-loop system identification methodology in Pitops, three second-order process models were simultaneously calculated using data from the trends in Figure 17. The model parameters calculated by Pitops are shown in Table 5. These process models were identified based on the pure closed-loop data from Figure 17.

The newly identified process models were used to reshape the models inside the DMC system. Subsequent closed-loop control action from the DMC system was vastly superior with crisp and tight control without any oscillations. This is a complex identification example comprising of completely closed-loop data superimposed with noise and unmeasured disturbances amidst an active DMC controller. The duration of the data window is ultra-short compared to what is needed for successful identification with known conventional software and methodologies.

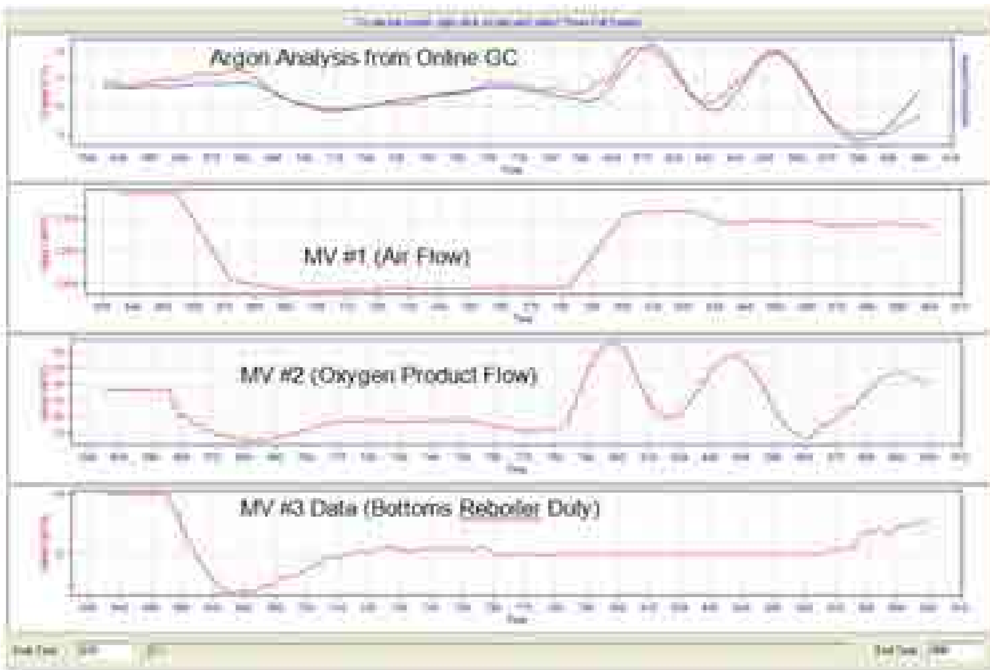

Figure 17: Poor and oscillatory closed-loop control from DMC on a gas plant

Table 5: Process models based on closed-loop identification

	1st Process model	2nd Process model	3rd Process model
Dead Time	4.1 minutes	7.4 minutes	13.3 minutes
Process Gain	-0.0273 %/SCFM	0.1439 %/SCFM	1.114 %/SCFM
1st Time Constant	59.4 minutes	57.1 minutes	357.2 minutes
2nd Time Constant	14.3 minutes	19.5 minutes	28.7 minutes

Multivariable control of processes with fast dynamics

Equipment like mechanical extruders, compressors, turbines, turbo-machinery, hydraulic devices, robots and others show fast process dynamics that settle in under 2-5 minutes. For such applications, DCS/PLC-resident APC strategies are extremely attractive. These can be run at fast scan rates. Their PID/APC tuning parameters can be mathematically optimized based on accurately identified dynamic models. The DCS/PLC-resident APC control action can be noticeably superior to a corresponding MPC/DMC system.

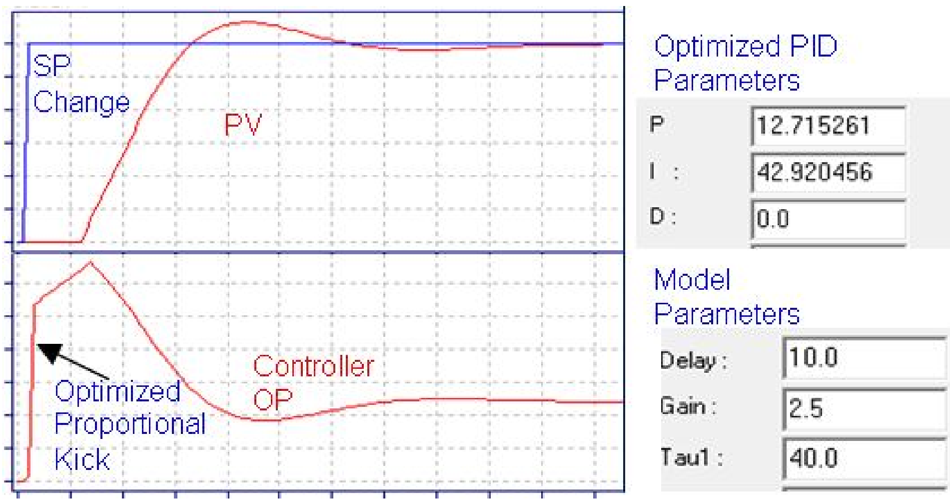

Figure 18: Precise optimized control action based on accurate system dynamics

DCS/PLC-resident APC schemes have several advantages over MPC/DMC. First, they can run with super-fast scan rates of 1 second or even faster. Second, feedforward control schemes, constraint overrides and cascades can all be precisely designed and optimized based on knowledge of process models. This provides excellent disturbance rejection especially since most disturbances on such processes can be fast. Third, their MV trajectory can be easily designed for crisp and non-oscillatory control action. Figure 18 shows the generation of optimized proportional kick illustrated for a setpoint change. The strong proportional kick is based on mathematically calculated PI tuning parameters which are based on the identified process model shown in Figure 18. The ability to identify process models and then generating the optimal control parameters mathematically is an important capability in the design of APC schemes. Without model identification and tuning tools, many PID controllers in industry are run with incorrectly small integral times, which either cause oscillations or force the proportional gain to be also incorrectly small. Both these incorrect adjustments produce poor control, large deviations from setpoint and sustained oscillations.

Due to unavailability or absence of robust and practical system identification tools like Pitops, many plants have chosen to hand-over turnkey projects to a consultant or vendor to implement a MPC/DMC system on such processes characterized by fast dynamics. Many such MPC/DMC systems have been ineffective or inferior to DCS/PLC-resident APC.

Summary and conclusions

Though the hardware side of DCS and PLC systems is mature, the use of advanced applications functionality in the DCS and PLC for providing APC and optimization is not yet fully explored in many chemical plants and refineries. Many new and inexperienced process control engineers and technicians enter the control rooms every year. The academic knowledge from current college courses is focused heavily on Laplace domain and the discrete (Z) domain but lacks the practical time-domain based exposure needed in the control room environment. A plant can have hundreds or even thousands of PIDs and no human group can monitor all PIDs and keep them controlling optimally all the time. Process changes, equipment changes, grade changes and other reasons can causes changes in system dynamics causing controllers to become oscillatory or sluggish. This is turn forces production rates to be reduced and control quality to be impacted. The absence of easy to use control monitoring tools, system identification tools and PID/APC tuning optimization tools results in poor control and missed opportunities in many chemical plants and refineries. The new process control software tools, technology and ideas explained in this paper are novel, modern, robust and simple. They can be used by new and inexperienced control engineers, even technicians without advanced college education and process control specialization. All over the world, an increasing number of skilled and experienced people are close to retirement age and many new persons are entering the control rooms. The need for practical approaches and powerful new software tools for training, for control quality monitoring of all PIDs and APC schemes, for quick and easy system identification using available data from the plant's historian are few of the novel ideas that can help tremendously improve the control quality and the plant's profit margin.

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PRIMJENA NAPREDNOG VOĐENJA PROCESA U RAFINERIJAMA I KEMIJSKIM POSTROJENJIMA

Sažetak

Na današnjem globalnom i konkurentnom tržištu, kemijska postrojenja i rafinerije traže nove načine za povećanje učinkovitosti rada produktivnosti, sigurnosti i pouzdanosti postrojenja. U području automatike, optimiranja regulacije, naprednog vođenja i vođenja poduprijetog modelom procesa i dalje postoji niz izazova. Ovim radom prikazano je nekoliko novih metoda povećanja učinkovitosti rada postrojenja koji se temelje na optimiranju rada standardne i napredne regulacije. Također je prikazano nekoliko primjera iz realnih industrijskih postrojenja. Opisane metode i tehnike rada mogu povećati dobit prostrojenja od 2 do 10 %. Moguće je ostvariti i povećanja od čak 15 do 20 % (ekvivalentno 2 milijuna Eura godišnje), što je potkrijepljeno primjerima.

Ključne riječi: *identificiranje sustava, ugađanje i optimiranje regulatora, napredno vođenje procesa, prediktivno vođenje*

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