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Averaging level control of an industrial naphtha splitter reflux drum using a PID/MPC distributed control system function block

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Abstract: Averaging level control is used in process plants surge or buffer tanks to avoid propagation of inlet flow disturbances to downstream equipment while ensuring that pre-specified level constraints are not violated. The tank capacity is used to "filter out" inlet flow variation resulting in smoother process operation. The paper describes the design and application of the PID/MPC software function block PID_PLA offered as part of the Experion PKS distributed control system standard Control Builder library for averaging level control in an industrial refinery naphtha splitter reflux drum. During the design phase, the SISO model predictive control algorithm Profit®Loop is compared with traditional averaging level control approaches such as specifically tuned PI control or error-squared PI control, respectively.

Keywords: Averaging level control, Advanced Regulatory Control (ARC), Model Predictive Control (MPC), Distributed Control System (DCS).

1. INTRODUCTION

After flow control, liquid level control occurs most frequently in the process industries. In most cases, a cascade control structure is used as shown in Fig. 1, where the tank outlet flow is controlled by a secondary controller, and the level controller is the primary one. The inlet or feed flow acts as a disturbance. There are applications where the objective is to control the level tightly to a setpoint. Inlet flow disturbances are then propagated to the manipulated variable and disturb finally downstream process units. In other applications however, the purpose of level control is to dampen flow disturbances while respecting lower and upper level limits, i.e. avoiding tank overflow or running it empty. In this case, the tank acts as surge or buffer tank, and the process variability is shifted from the material flow to the level, see Fig. 1.

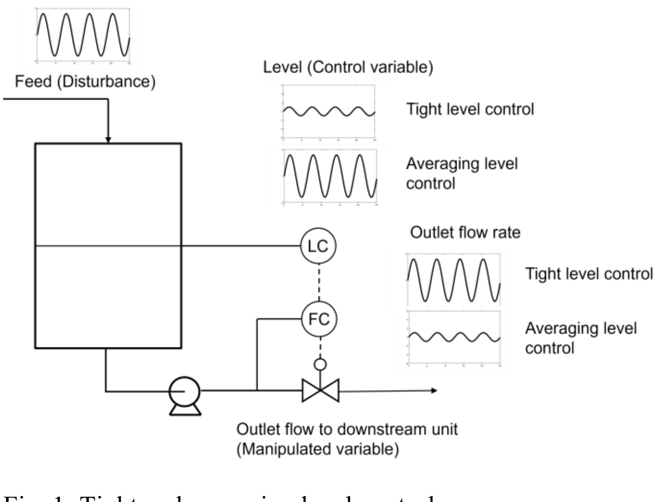

Fig. 1: Tight and averaging level control

The second case is called "averaging level control". Controller tuning for the two cases must be different, since for tight level control a high controller gain is required, whereas for averaging level control a low gain is needed. Practical experience shows, that particularly averaging level controllers in industry are often not tuned appropriately to meet their purpose.

Essentially three approaches for averaging level control can be distinguished: a) P and PI control with tuning rules developed specifically for this application, b) nonlinear control algorithms where the controller gain depends on the control error, and c) model predictive control.

All three concepts were widely studied in the literature. First tuning methods for P/PI level controllers were recommended six decades ago (Buckley, 1964; Shinskey, 1967). They later have been discussed and refined by (Cheung and Luyben, 1979). An extensive list of tuning rules for P/PI averaging level controllers for different process models and performance specifications can be found in (Wade, 2017). Tuning rules based on optimal control theory are presented in (Taylor and la Grange, 2002) and (Shin et al., 2008), based on optimal control and H_∞ theory in (Krämer and Völker, 2014). P/PI controllers that mimic the behaviour of a robust model predictive controller have been developed in (Rosander et al., 2012). An insightful review of P/PI controller application to liquid level control is given in (Luyben, 2020).

A nonlinear PI level controller was first elaborated in (Cheung and Luyben, 1980). PI(D) control algorithms with an error-dependent controller gain (such as the PI-e^2 controller) are available on different distributed control systems and PLCs, and have been applied to averaging level control in industry, too. Recently, (Reyes-Lua, Backi and Skogestad, 2018) presented a nonlinear control scheme composed of a low gain PI controller for setpoint control, two high gain P controllers to prevent level limit violations, and a mid selector. This structure can be implemented in DCS using standard software function blocks. (Gous et al., 2023) proposed an advanced regulatory control concept consisting of integral gap control and ramp horizon control which also can be implemented using standard DCS functions. The possibility to apply of MPC to averaging level control was first mentioned in (Campo and Morari, 1989).

Commercial multivariable MPC packages such as Aspen DMCplus® or Honeywell Profit®Controller are way too expensive and difficult to be applied for single-loop control. However, since several years smaller MPC function blocks running directly on DCS stations are available (e.g. Emerson DeltaV Predict, Siemens ModPreCon or Honeywell Profit®Loop) This offers the chance to apply MPC for averaging level control with a reasonable effort.

This paper describes the design and application of a PID/MPC DCS software function block for averaging level control in an industrial refinery naphtha splitter at Heide Refinery. During the design phase, the use of this block is compared with specifically tuned PI control or error-squared PI control, respectively.

The remainder of the paper is organized as follows: Section 3 describes the computer/DCS environment for the development of the application and for comparison with PI(D) control. Since details of the PID/MPC function block have only been published in the DCS documentation yet, a short description is put in front (Section 2). In Section 4, the industrial application is characterized, and the paper is concluded in Section 5.

2. THE PID_PLA FUNCTION BLOCK OF THE HONEYWELL EXPERION® PKS DCS

It is well-known in the advanced control community that the Honeywell Experion DCS offers a full-blown multivariable MPC controller (Profit®Controller). It normally runs on a dedicated PC connected with the DCS by an OPC interface. In another version, it runs embedded in the DCS control station (C300). It is, however, less popular that the DCS also provides an SISO MPC controller named Profit®Loop (Lu, 2004; Dittmar, 2009). This “little brother” of the Profit® Controller is part of the PID_PLA function block, combining the functionality of a PID controller with a predictive controller/optimizer. Although Profit®Loop can be used for setpoint control, it is particularly advantageous for range control applications with minimum output changes such as tank surge control. Range control requires the specification of upper and lower limits for the control variable.
In contrast to PID controller tuning, only one tuning constant has to be specified by the user: the so-called performance ratio PR which is the ratio between the desired closed-loop settling time t_s and the open-loop settling time of the process). PR=1 represents steady-state control and reaches the desired settling time at the speed of the process. While decreasing the PR value leads to faster and more aggressive control, increasing it slows the closed loop response down and increases robustness (tolerance with respect to plant model mismatch). In case of integrating processes, PR must be set to PR=1, and the desired closed-loop response time must be specified directly.
If range limits are not actually violated or predicted to be violated within the controller’s prediction horizon, a steady-state optimization function will usually be activated. The optimization objective can be (see Fig. 2):

- to drive the process to either the maximum or the minimum range limit,
- to strive to a user-defined optimization target value within the range limits, or
- to keep the optimal solution between extra “optimization limits” inside the control range.

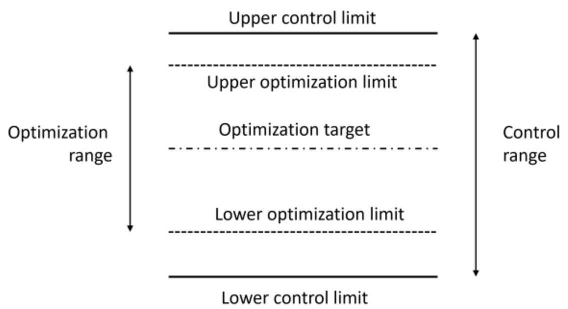

Fig. 2: Targets and limits in the PID_PLA function block

If no steady-state optimization is activated, the process will freely float between the range limits. For averaging level control, the optimization target value could, for example, be 50%, the control range between 20% and 80% of the level measurement span, and the optimization range between 30% and 70%. A unitless parameter called “optimizer speed” dictates how fast the optimizer drives the process to the optimal target. The optimizer speed can be specified independently from the performance ratio, i.e. the closed-loop settling time. Normally, the optimization speed will be selected to be slower than the “controller speed”. With a speed of OPTSPEED = 6.0, the optimizer will be as fast as the controller, with OPTSPEED = 3.0 the optimizer will be twice as slow. The settling time associated with changes induced through steady-state optimization is called “optimization settling time” and denoted by $t_{s,opt}$. $t_{s,opt}$ and the control loop settling time t_s are related by the equation $t_{s,opt} = \dfrac{6 \cdot t_s}{OPTSPEED}$. The default value of the optimizer speed is OPTSPEED = 2.0.

Like any other MPC controller, Profit®Loop needs a dynamic process model. For averaging level control, the “process” includes the tank and the flow control loop, i.e. the secondary control loop of the level-to-flow cascade. In order to develop this model, a separate system identification tool called Profit®Assistant is provided. Based on user information about the process, Profit®Assistant creates a PRBS test signal. For integrating processes (e.g. outlet flow rate to tank level), identification is executed in closed loop, i.e. the test signal is applied to the level controller setpoint. As a result, a transfer function model of the form

$$G(s) = K_P \frac{b_4 s^4 + ... + b_1 s + b_0}{a_4 s^4 + ... + a_1 s + a_0} e^{-sT_d} \quad (1)$$

will be generated, i.e. (maximum) fourth order plus time delay. This includes integrating process models such as

$$G(s) = \frac{K_I}{s(T_1 s + 1)} e^{-sT_d} . \quad (2)$$

To assess the model quality, Profit®Assistant provides a "model rank" and a graphical comparison between measured values and simulation results.

The Experion Control Builder is used for the configuration and parametrization of the control function. A continuous function chart (CFC) is developed as shown in Fig. 3. Here, the process value (PV) is read from an analog input and sent to a data acquisition block before it is processed by the PID_PLA block to generate the output (manipulated variable, MV). In a cascade structure , this MV is not sent to an output channel as in this Figure, but to the setpoint of a secondary PID controller block.

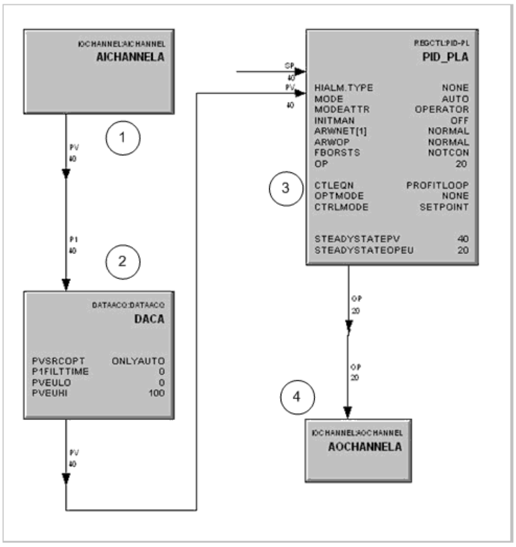

Fig. 3: CFC of a Profit®Loop control application

A mask related to the PID_PLA block can be opened which then allows to enter all necessary PID or Profit®Loop parameters, respectively. In Fig. 4, the "Advanced" tab is selected. Here, on the left-hand side, the process model can be entered and/or displayed, and optimization parameters can be specified on the right-hand side. Hitting the button on the right bottom of the mask starts the Profit®Assistant tool.

A PID_PLA faceplate and a detailed controller representation for the construction of the operator screens (Experion Graphics Builder) are provided as well.

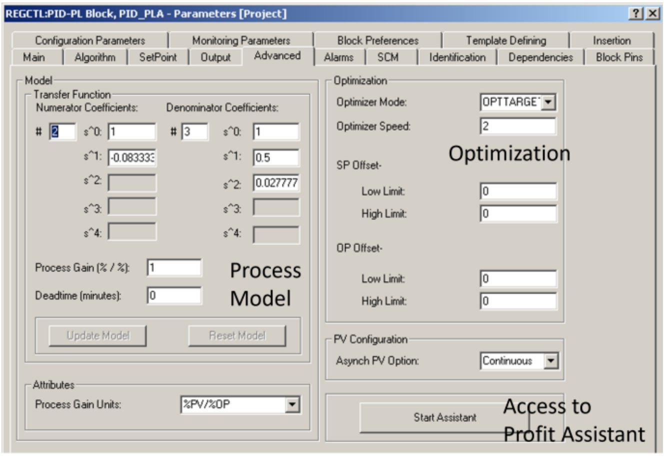

Fig. 4: Input mask for PID_PLA block (Control Builder)

3. ENVIRONMENT FOR THE DESIGN AND SIMULATION

For security reasons and to gain experience with the Profit®Loop and Profit®Assistant tools, the project did not start in the refinery unit right away. Instead, an existing DCS/PC system at the university was used to carry out preliminary simulations and tests. This university lab test environment is shown in Fig. 5.

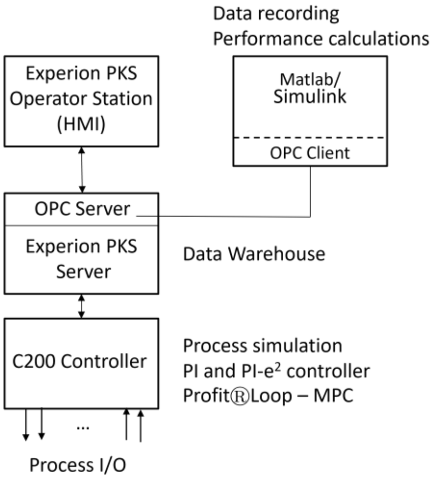

Fig. 5: Computer /DCS environment for experiments

It consists of a small industrial Experion DCS with a C200 controller, a Dell server holding the DCS database and an OPC server, and an operator station providing the DCS human-machine interface. MATLAB/Simulink is installed on another PC and connected via its OPC toolbox. It is used for data recording and calculation of controller performance. A process simulation was configured on the C200 controller using standard DCS software function blocks. For the flow rate, an FOPDT model with the transfer function

$$G_{P2}(s) = \frac{1}{3s + 1} e^{-s} \quad (3)$$

was used, the flow rate PI controller was tuned with $K_C = 1,5$ and $T_i = 3s$ (SIMC tuning), i.e.

$$G_{C2}(s) = 1,5\left(1+\frac{1}{3s}\right). \quad (4)$$

The level process was assumed to have integral behaviour with dead time (IPDT), its model was

$$G_{P1}(s) = -\frac{2}{s}e^{-2s}. \quad (5)$$

For averaging level control simulations, the following controllers were configured on the DCS C200:

a) PI controller tuned specifically for averaging level control according to (Wade, 2017):

$$K_C = \frac{0,74}{\sqrt{\left(1-T_d K_{P,I}\right)}} \cdot \frac{\Delta F_{in,max}}{\Delta L_{max}}, \quad T_i = \frac{4}{K_{P,I} K_C} \quad (6)$$

Here, $K_{P,I}$ and T_d denote the integral gain and time delay of the level, $\Delta F_{in,max}$ the expected maximum inlet flow rate step change and ΔL_{max} the maximally tolerated level control error.

b) PI-e²- or "error-squared" controller

$$u(t) = K_C\left|e(t)\right|\left(e(t) + \frac{\left|e(t)\right|}{T_i}\int_0^t e(\tau)d\tau\right) \quad (7)$$

(this controller can be configured within the PID part of the PID_PLA block). The Experion PKS version of the error-squared controller allows only the modification of the controller gain according to

$$K_C = K_{C,lin}\left(NLGAIN\frac{\left|e(t)\right|}{PV\ range}\right). \quad (8)$$

In the simulations, the NLGAIN parameter was selected to give $K_C = 0,9K_{C,crit}$ (90% of the ultimate gain) if the maximally tolerated level control error is achieved.

c) Profit®Loop MPC1 controller with a control range from 20% to 80% of the level measurement range, and an optimization target OPTTARGET = 50%. Since the level process is integrating, the performance ratio is set to PR=1 and the desired closed-loop settling time was set to 60 min. The speed of the optimization was on its default value of OPTSPEED=2.

d) Profit®Loop MPC2 controller with a control range from again 20% to 80% of the level measurement range, and an optimization range (i.e. not a target value) of 40% to 60% of the measurement range. The desired closed-loop settling time is again 60 min, and optimization speed OPTSPEED =2.

With all four controllers, the reaction to a step disturbance of 5% of the input flow rate measurement range was simulated. Since the process model was implemented on the DCS, sinusoidal inlet flow disturbances have not been investigated. To assess the control performance, the following performance criteria were calculated within MATLAB:

a) closed loop settling time t_s as the duration of time after the disturbance to enter and stay within 5% of the maximum deviation from its resting value,

b) the maximal absolute control error (either deviation from target or from range limits) $\left|e_{max}\right|$,

c) total variation of the manipulated variable (input usage)

$$TV = \sum_{i=1}^{\infty}\left|u_{i+1} - u_i\right| \quad (9)$$

d) maximum change of manipulated variable after step disturbance $\Delta u = \left|u_{extr} - u(0)\right|$.

The last two performance measures describe how input disturbances affect the flow to the downstream unit. Nominal case simulation results are presented in Tab. 1.

Table 1: Control performance for different control algorithms

Controller	$\left\|e_{max}\right\|$ [%]	t_s [min]	TV [%]	Δu [%]
PI	30,38	45,2	6,27	5,68
PI-e²	17,20	233	5,99	5,5
MPC1	28,24	51,1	5,3	4,8
MPC2	18,56	50,8	5,35	5,34

Compared with the PI controllers, with Profit®Loop a smaller output flow variation (expressed by TV and Δu) was achieved. In our example, with MPC1 (control to an optimization target) the reduction of TV and Δu amounts to more than 15%. A quantification of control robustness is difficult to achieve, since proprietary control algorithms have to be assessed. Therefore, the simulations were repeated with $\pm 20\%$ changes in the integral gain of the process. Selected results are presented in Table 2. The numbers describe the maximum absolute percentage deviation from the nominal case for the process gain range $-2,4 \le K_{P,I} \le -1,6$

Table 2: Robustness for 20% process gain changes

Controller	$\left\|e_{max}\right\|$	t_s	TV	Δu
PI	3,9%	2,2%	15,6%	2,4%
PI-e²	2,6%	NA	13,8%	2,3%
MPC1	4,0%	0,9%	22,4%	2,9%
MPC2	5,0%	8,3%	22,5%	1,5%

The sensitivity of the performance measures with respect to process gain changes is small for $\left|e_{max}\right|$, t_s and Δu. For the total variation TV, the sensitivity is still acceptable. Surprisingly, the model-based Profit®Loop controllers were almost as robust against model uncertainty as the PI controllers, although they internally use the "wrong" process model for prediction.

4. REFLUX DRUM LEVEL CONTROL OF AN INDUSTRIAL NAPHTHA SPLITTER

The Profit®Loop controller was then tested in collaboration with Heide refinery at a naphtha splitter condensate reflux drum. A simplified P&I diagram of the unit is shown in Fig. 6. In the column, a naphtha feed is thermally separated into heavier and a lighter boiling fractions. The overhead product is liquefied in a condenser, the condensate is collected in a reflux drum.

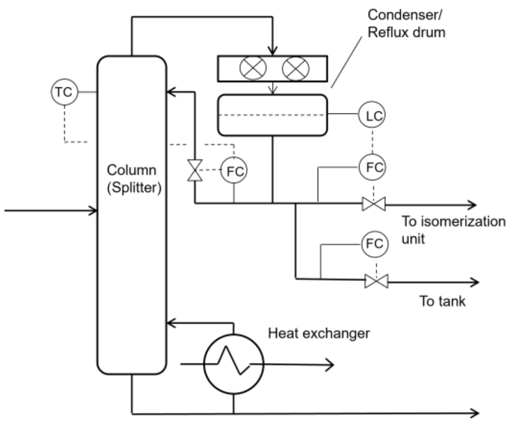

Fig. 6: Simplified P&I diagram of a refinery naphtha splitter

A part of the condensate is returned to the column, the reflux flow rate controller acts as secondary controller in a temperature-to-flow rate cascade structure. The reflux drum level is controlled by manipulating a part of the overhead product flow which is sent to a downstream isomerization unit. The level controller is cascaded to a secondary flow rate PI controller. A third part of the condensate is pumped with constant flow rate to a product tank. The objective of reflux drum level control is not to tightly control the level, but to dampen the flow rate to the downstream isomerization unit respecting level constraints.

First, a dynamic process model for the relation between overhead product flow rate an reflux drum level was developed. Therefore, a step test was executed putting the flow controller into manual mode, stepping the flow rate setpoint and recording the level PV. This experiment must be carried out with caution and close observation of the level limits. By manual evaluation of the recorded data, it was obvious that the process can be modelled by integrating behaviour plus a small dead-time. By visual inspection, the following IPDT model could be determined:

$$G(s) = -\frac{0,053}{s} e^{-3s} \tag{10}$$

This model consists of the flow control loop and the level dynamics, respectively. In addition, the flow rate to level dynamics was also modelled using the Profit®Assistant identification tool. Here, the identification was carried out in closed loop, i.e. the level controller setpoint was excited with a PRBS. Subsequent identification resulted in the IPDT model

$$G(s) = -\frac{0,058}{s} e^{-3s} \tag{11}$$

which is similar to the step test result. The second model was used for the Profit®Loop MPC.

In consultation with the refinery personnel, the lower an upper level limits for control were set to 30% and 70%, respectively. The closed-loop response time was selected to be $t_s = 60\,\text{min}$. The optimization target was specified with 50%, and the optimizer speed remained at its default value.

In the continuous function chart (Control Builder), the PIDA block was replaced by the PID_PLA block. In the operator screen (Display Builder), the PID controller faceplate for the level controller was replaced by the PID_PLA faceplate. This allows the operator to change the level limits and the optimization target if necessary. The most critical phase during the commissioning of the new control scheme is the download of the new CFC and display to the DCS on the fly, because level control is impossible then for a couple of seconds. This requires a close collaboration between engineers and the control room operator.

Figs. 7 and 8 show some of the results. Fig. 7 presents 1 min snapshot values for the reflux drum level over six months, before and after commissioning of the new averaging level control structure. The level standard deviation was increased from $\sigma_{LC} = 1,72$ to $\sigma_{LC} = 6,41$, i.e. by a factor of more than 3,5. Fig. 8 however, presents the trend of the flow rate to the isomerization unit. Here, 1 min snapshots for about 4 days before and after the Profit®Loop commissioning are shown. The linear upward trend on the right hand side of the chart is due to the overlapping of level and temperature control at the splitter. (The temperature controller manipulates the reflux flow rate which disturbs the drum level and, as a consequence, the overhead product flow rate.) If one subtracts the linear trend, the reduction of flow rate standard deviation can be calculated: it is $\sigma_{FC} = 0,885$ with PI control and $\sigma_{LC} = 0,553$ with Profit®Loop control. This is equivalent to a 38% reduction and has definitely improved the operation of the downstream isomerization unit.

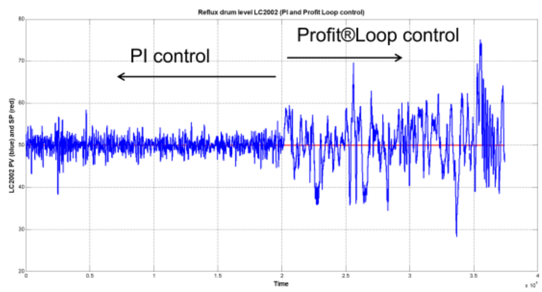

Fig. 7: Reflux drum level with PI and Profit®Loop control

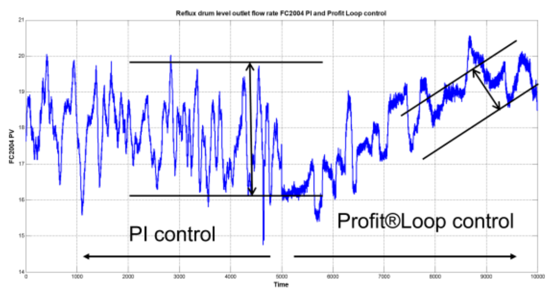

Fig. 8: Overhead product flow rate with PI and Profit®Loop control

Including PRBS tests, system identification, changes in CFC (Control Builder) and operator display (Graphics Builder), commissioning and test, the whole project could be finished within one refinery shift of eight hours.

5. CONCLUDING REMARKS

Most averaging level control applications in industry use the PI control algorithm. P control is rarely applied since operators want to see the process value running at the controller setpoint which is not the case for load disturbances. Model Predictive Control with its capability of range control and limiting the output rate of change is obviously a good choice for this type of application. But general-purpose MPC packages for industrial DCS are not only expensive, but also complex in design and commissioning, and not easy to tune. But with the availability of MPC algorithms embedded in DCS control stations, these arguments are no longer valid.

In this paper, we therefore presented the application of the PID/MPC function block PID_PLA (Profit®Loop) offered by the Honeywell Experion PKS system within an industrial refinery unit. As a result, an almost 40% reduction of the manipulated variable standard deviation could be achieved. This led to a smoother operation of the downstream isomerization unit. With the help of the embedded system identification tool, a dynamic process model could be developed without much effort. This was done by executing closed-loop tests with a PRBS test signal. Controller tuning was transparent and simple, since only one parameter had to be specified for the control and for the optimization part of the algorithm. All design, implementation and commissioning work could finished in a reasonable time frame.

Since it was the first application of this technology in the refinery, an Experion PKS system installed at the university control lab was used to become familiar with the Profit®Loop/Profit®Assistant tools. In addition, the results achieved with the PID_PLA block could be compared with PI and error-squared PI control.

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