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Implementation of PFC (Predictive Functional Control) in a petrochemical plant

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Abstract: PFC (Predictive Functional Control) is a simple but very effective SISO (Single-Input Single-Output) predictive controller which does not require any matrix inversion or numerical minimization of a cost function, therefore it is easy to realize in a PLCs (Programmable Logical Controllers) or a DCS (Decentralized Control Systems). The manipulated variable and/or its increment, as well other variables can be constrained easier than when using PI(D) (Proportional, Integrating, and Derivating) controller. Furthermore PFC handles the dead time process better than with PI(D) controller. Three industrial applications of PFC for two distillation columns and a reactor in a petro-chemical plant were implemented and the experiences are presented here.

Keywords: predictive functional control, distillation column, programmable logical controller

1. INTRODUCTION

PFC (Predictive Functional Control) is a simple predictive control algorithm which does not require any matrix inversion or numerical minimization of a cost function. PFC algorithms can be easily realized for all types of SISO (Single-Input, Single-Output) processes. The manipulated variable and/or its increment, as well other variables can be constrained easily. Furthermore PFC as a MBC (Model Based Controller) can compensate the dead time of the process. Also bumpless switching between a backup controller (PID) and PFC are available in practice for simplicity of installing PFC algorithm; see (Richalet and O'Donavan, 2009; Haber and Zabet, 2011).

Therefore, PFC has growing usage in industries and has been implemented successfully for many chemical processes; see e.g. (Richalet, *et al.*, 1978). In the course of the implementation of a new process control system three PFC were successfully applied for the control of two distillation columns and a reactor in a petro-chemical plant. In these cases PFC partly released a much more complicated predictive controller and partly a not sufficient working PI(D) control.

The paper is structured as follows. In Section 2 the PFC algorithm is shortly shown for a delayed first-order model, because a first approximation of a controlled process is usually satisfactory for the PFC controller design. In Section 3 bumpless switching between PI and PFC controllers is presented briefly. In Section 4, three different case studies of the distillation columns controls using PFC are presented. Finally in Section 5 some conclusions of the PFC implementations are summarized.

2. PREDICTIVE FUNCTIONAL CONTROL

The principle of PFC with a constant set-value is that the controlled variable y achieves the reference trajectory at the target point using one change in the manipulated variable u. The desired change in the controlled variable y during the prediction horizon n_p (from the actual time k) is calculated from the desired change of the reference trajectory and the predicted change of the model output y_m, see Fig.1.

The desired changes in the controlled variable y during the prediction horizon n_p can be defined as follows:

$$\Delta\hat{y}(k+n_p \mid k) = (1-\lambda_r^{n_p})[y_r(k)-y(k)] \tag{1}$$

where y_r is the reference signal, $\lambda_r = \exp(-3\Delta t/T_c)$ is the reduction ratio of the control error, Δt is the sampling time.

The reference trajectory is chosen an exponential function for simplicity. It provides the closed loop settling time $t_{95\%}=T_c$.

A PT1 (proportional, first-order) process model without dead time (chosen for simplicity) is described in discrete-time as:

$$y_m(k) = -a_m\, y_m(k-1) + K_m(1+a_m)u(k-1) \tag{2}$$

where y_m is the model output, u is the model input, K_m is the model's gain and $a_m = -\exp(-\Delta t/T_m)$ is the discrete-time model parameter and T_m is the time constant of the model.

The predicted change in y_m during n_p steps becomes:

$$\begin{aligned}\Delta\hat{y}_m(k+n_p \mid k) &= \hat{y}_m(k+n_p \mid k) - y_m(k) \\ &= [1-(-a_m)^{n_p}][K_m\, u(k) - y_m(k)]\end{aligned} \tag{3}$$

Comparing the predicted change of y_r in (1) and the predicted change of y_m in (3) results in the manipulated variable:

$$u(k) = k_0[y_r - y(k)] + k_1 y_m(k) \tag{4a}$$

where:

$$k_0 = \frac{1-\lambda_r^{n_p}}{K_m[1-(-a_m)^{n_p}]}\ , \qquad k_1 = \frac{1}{K_m} \tag{4b}$$

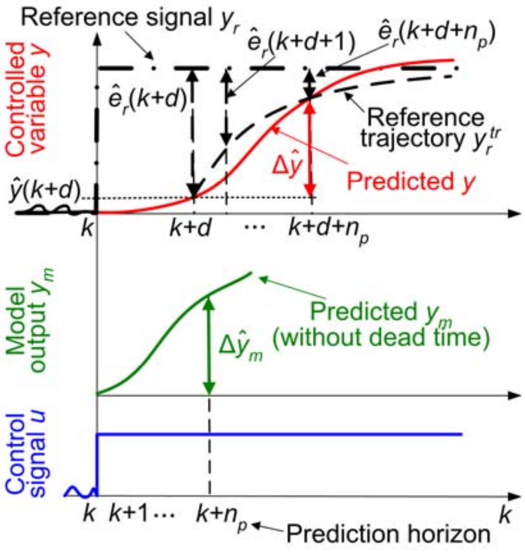


Fig.1. PFC principle

If the process has dead time $d = d_m$ then $y(k)$ in (4a) has to be replaced by $y(k) + [y_m(k) - y_m(k-d_m)]$.

3. BUMPLESS SWITCH BETWEEN PI(D) AND PFC

Bumpless switch means that the manipulated variable does not change abruptly when switching between different controller modes to avoid any discontinuity in the controlled variables, see (Aström and Murray, 2008; Ping, *et al.*, 1996).

In this paper switching between backup PI and PFC is presented; see (Richalet and O'Donavan, 2009; Haber and Zabet, 2011). A common strategy is to let run the passive controller in such a way that the calculated manipulated variable equals always to the manipulated variable of the active controller, see the scheme in Fig. 2. The following points are taking into account during switching:

- Manipulated variable of the backup PI and PFC controllers are computed always, i.e. independent whether they are used or not.
- The controlled variable $y_{active}(k)$ is set as set-point of the passive (not used) controller $y_{r,\,passive}(k) = y_{active}(k)$.
- The manipulated variable $u_{active}(k\text{-}1)$ is set as output signal of the passive controller $u_{active}(k\text{-}1) = u_{passive}(k\text{-}1)$.
- When PFC is installed for the first time, the steady-state value of the manipulated variable calculated based on the static gain of the model is set to all stored values of u_{PFC}.
- If the process is not in steady-state or the set-points of the control modes differ then the set-point change should be filtered. The filter is initialised with the existing controlled variable, and the desired set-point is set to the input of the filter. As the new set value is neared slowly there shall be no jump in the signals during the transients.

If a process is controlled by a PI controller $y_{r,PFC}(k) = y_{PI}(k)$, then the manipulated variable of the passive PFC equals the steady state value as:

$$u_{PFC}(k) = u_{ss}$$

During PFC mode $y_{r,PI}(k) = y_{PFC}(k)$, the manipulated variable u_{PI} is computed as follows:

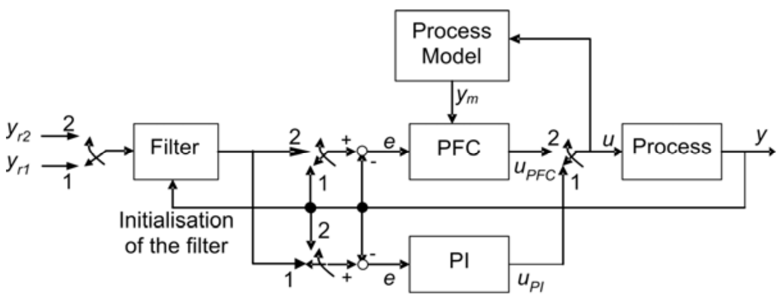
$$u_{PI}(k) = u_{PFC}(k-1)$$

4. CASES STUDY - PETROCHEMICAL PLANT

Three PFC controllers have been implemented in the Shell Rhineland Refinery in Wesseling near to Cologne.

4.1 Description of reactor temperature control with PFC

In a reactor usually the most important temperature is the highest temperature due to safety reason and asset protection. Thus this temperature is the one that should be controlled. Many reactions are exothermic and therefore hard to control if there is no ability for intercooling of the reactor (e.g. by quench or integrated heat exchanger). The only possibility to control such a reactor is by setting the inlet temperature, which is either pre-heated or pre-cooled.



Mode 1 : PI controller
Mode 2 : PFC controller

Fig. 2. Scheme of bumpless switching between PI and PFC controllers

Due to complex dynamics, nonlinearity, dead time (especial with liquid-phase) and feed changes which influence the exothermic behavior of the reaction the temperature control of the outlet temperature of a reactor is quite challenging for a control algorithm.

The reactor is a part of an aromatics complex of a petrochemical plant where different high purity aromatics products are produced. This reactor is settled in a recycle stream, where a side product is recovered to a feed component of the plant. The outlet temperature of a fixed bed reactor without intercooling should be controlled as the hottest measured temperature of the reactor. The temperature is controlled by a cascade structure. The inlet temperature of the reactor, which is the outlet temperature of a natural draft furnace, is controlled by the inner basic controller, see Fig. 3.

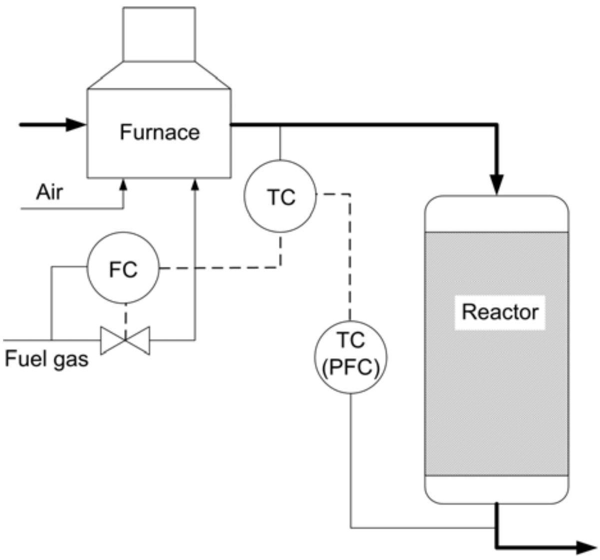


Fig. 3. Scheme of the reactor control

The dynamic behavior between the inlet and the outlet temperature of the reactor was identified by a step test and it was identified by a delayed first-order model using measured temperatures at each 1 min., see Table 1.

Table 1: Approximated delayed first-order reactor model

Gain [°C/°C]	Time constant [min]	Dead time [min]
1.037	8	7

The relation of dead-time and time-constant shows, that a conventional PID would have significant problems controlling this process, thus a PFC was chosen. Also a robust control algorithm is necessary, because the natural draft furnace produces stochastic disturbances due to the uncontrolled air intake of the furnace.

The outlet temperature of the reactor was measured two days and is plotted together with its set-value in Fig. 4a. The three peaks between the sampling points 1000 and 1800 descend form fluctuations in the heating gas. The manipulated variable is the set-point of the PID controller of the inlet temperature of the reactor. The inlet temperature of the reactor and its manipulated variable, the heating gas flow were measured two days and are plotted in Fig. 4b.

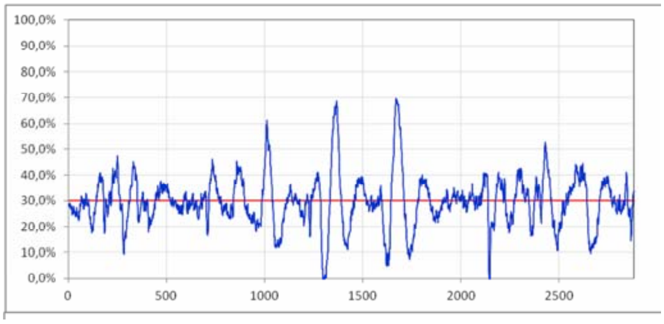


Fig. 4a. Outlet temperature of the reactor using PFC (red line: set-point, blue line: outlet temperature)

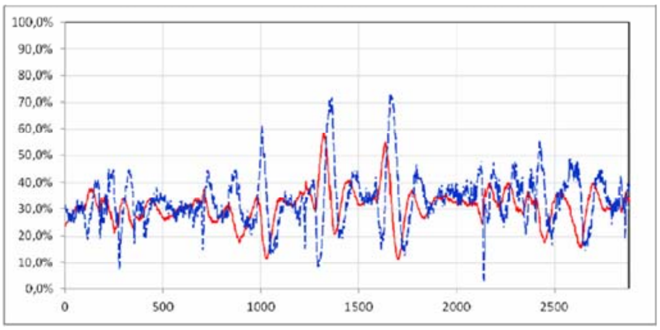


Fig. 4b. Inlet temperature of the reactor and heating gas flow using PID (red line: inlet temperature, blue dashed line: heating gas flow)

Fig. 4 shows a robust PFC control of the reactor outlet temperature and the furnace outlet temperature (reactor inlet temperature). The PFC tries to compensate the disturbances from the furnace and from the reaction as good as possible and it tracks the desired set-point of the hottest reactor temperature pretty reliable with minimum noise.

4.2. Topping distillation column control

A mayor part of a petrochemical plant is the distillation column in which intermediate products are separated by using the different boiling points of the products. The component with the lower boiling point is called the light end and leaves the column as the top-product. The component with the higher boiling point is called the heavy end and leaves the column as the bottom product. In the separation processes usually a solvent is used to support the separation and that has to be separated from the wanted products afterwards, usually by distillation.

In this case a column should be controlled which increase the purity of the produced heavy aromatic product after separation of the solvent. The product can contain small amount of lighter components which have to be toped from the product by distillation. Because of the relative small amount of top product and the high purity of the bottom (main) product this kind of columns are called "topping column". Typical for these kinds of columns is an extremely

high reflux to distillate ratio (based on the small amount of distillate off course).
The loss of higher value bottom product in the top stream is predicted by a quality estimator based on the temperature on tray 4 (from top) and the column-pressure. Due to the very low distillate flow, the losses of heavy product are absolutely not very high, even with high concentrations. Nevertheless, these losses should be minimized. The content of light product in the bottom product is hard limited by the product specification (<0.05%).

An APC structure was chosen, where the pressure compensated temperature at tray 4 is controlled by the reflux flow using a PFC algorithm to stabilize the column and minimize the temperature against the specification of light components in the bottom, see Fig. 5.

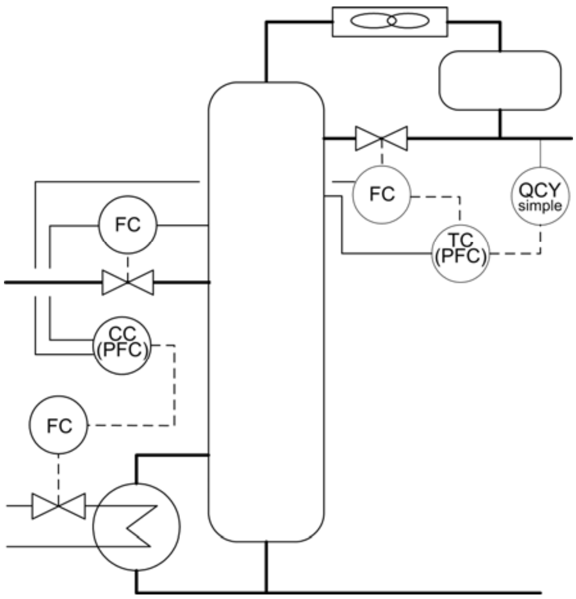


Fig. 5. Scheme of the topping distillation column control

The identified parameters of an approximated first-order model between the heating medium flow and the ratio reflux-to-feed using sampling time 0.5 min are shown in Table 2.

Table 2: Approximated delayed first-order model between the reflux flow and the temperature on tray 4

Gain [°C/(t/h)]	Time constant [min]	Dead time [min]
-2.38	7.8	3

The quality estimator for heavy components in the top of the column is based on the pressure compensated temperature. A simple calculator based on a linear model was chosen which calculates a temperature set-point based on the model and the set-point for the product losses, see the plot in Fig. 6a.

A PFC was installed at the bottom of the column which controls the reflux-to-feed (R/F) ratio by modifying the heating medium flow in order to minimize the energy consumption. If the reflux ratio is too high, the controller reduces the heating medium flow which results in a dropping temperature whereas the temperature controller reduces the reflux flow. It is clear that the feedback of the R/F controller takes place by the temperature controller in service - a simple logic ensures this behavior.
The identified parameters of an approximated first-order model between the heating medium flow and the ratio reflux-to-feed using sampling time 0.5 min are shown in Table 3.

Table 3: Approximated delayed first-order model parameters between the heating medium flow and the reflux-to-feed ratio

Gain [%/(t/h)]	Time constant [min.]	Dead time [min.]
6.7	11	3

A slow closed-loop settling time of 360 min was chosen to ensure minimum disturbance by this - just optimizing – controller, see the plot in Fig. 6b.

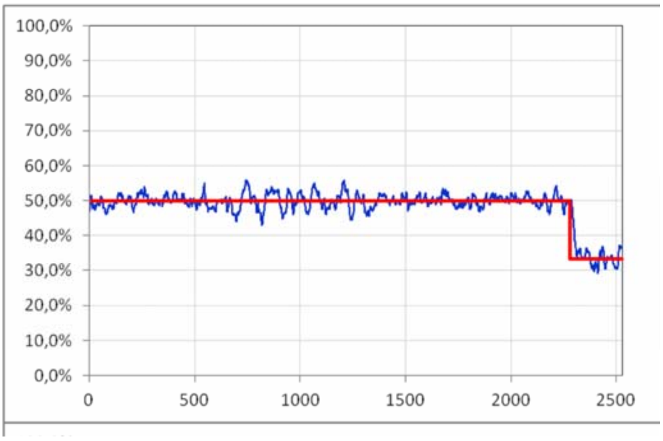


Fig. 6a. Control behavior of the predicted heavy component in distillate (red line: set-point, blue line: heavy components)

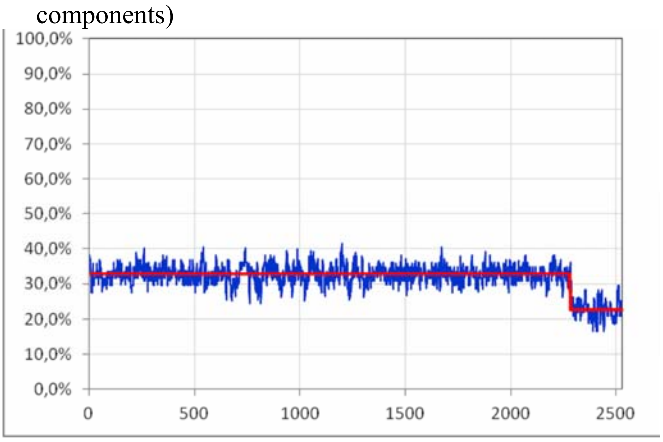


Fig. 6b. Pressure compensated temperature of the topping column using PFC (red line: set-point, blue line: pressure compensated temperature)

The control behavior of the top temperature controller shows a fast and reliable tracking of the desired set-point. Only some white noise is observable on the controlled variable in the plot. The heavy light end content controller is calculated from the same temperature.
Cascade control of the topping column is applied using PFC for the external closed-loop and using PID controller for the internal closed-loop. The reflux-to-feed ratio was controlled using PFC for a constant set-point and it is plotted in Fig. 7a.

The manipulated variable was the set-point of the heating medium flow which was controlled by a heating medium flow valve (not plotted) using a PID controller. The heating medium flow is plotted in Fig. 7b. The plot shows that the control of the reflux-to-feed ratio had a good performance.

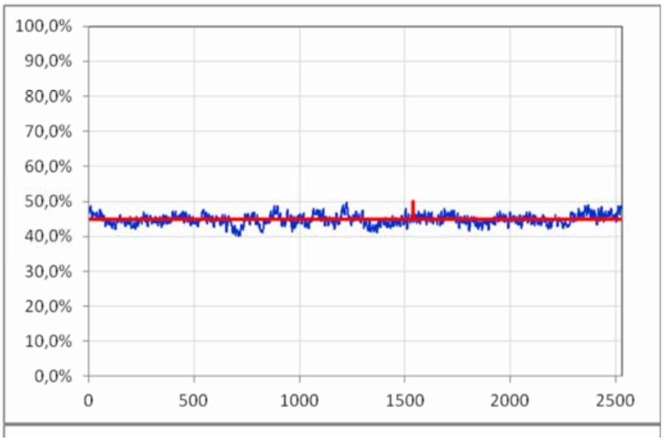


Fig. 7a. Reflux-to-feed ratio of the topping column using PFC (red line: set-point, blue line: reflux-to-feed ratio)

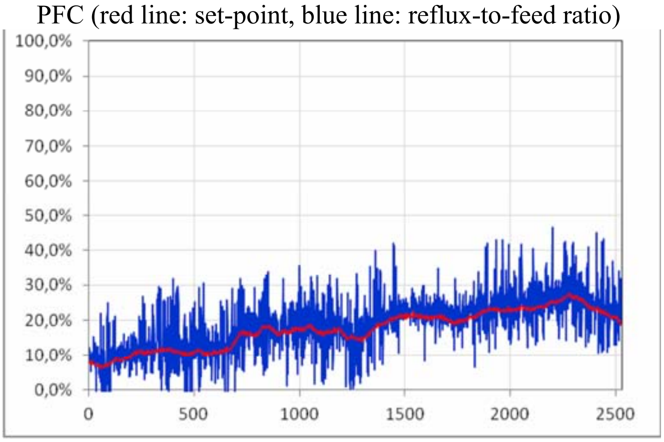


Fig. 7b. Heating medium flow of the topping column using PID (red line: set-point, blue line: heating medium flow)

4.3. Tailing distillation column control

In this case a column should be controlled which increases the purity of a product which is the bottom product of a leading distillation column. To fulfill the specification of this product some heavy components have to be removed from the major stream which is the light end in this case. Because the heavy end components have to be tailed from the product by distillation this kind of columns are called "tailing column". The column has 62 trays. Heating medium and reflux are only flow-controlled; no temperature controller is installed at the column in the basis system. The content of heavy components in the distillate is constrained by specification (<1.0%). Therefore an analyzer is installed at the distillate stream.

An APC structure was chosen, where the temperature at tray 53 (from top) is controlled by the heating medium flow using a PFC algorithm to stabilize the column. The scheme of the tailing distillation column control is shown in Fig. 8. The identified parameters of an approximating delayed first-order model between the heating medium flow and the temperature on tray 53 (from top) are shown in Table 4.

Table 4: Approximated delayed first-order model between the heating medium flow and the temperature on tray 53

Gain [°C/(t/h)]	Time constant [min]	Dead time [min]
1.78	16.3	0.5

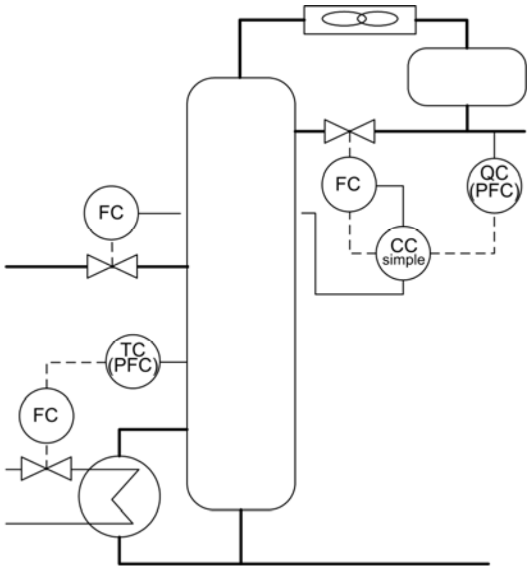


Fig. 8. Scheme of the tailing distillation column control

At the top of the column a simple ratio controller controls the reflux-to-feed ratio by modifying the reflux flow due to the direct relation by a simple algorithm

$SP_{flow}=SP_{R/F}{\cdot}y_{flow}/100$; $y_{R/F}= 100{\cdot}SP_{flow}/y_{flow}$

A three day long record of the controlled variable (temperature of tray 53) is plotted in Fig. 9a. Heating medium flow and the valve position are plotted in Fig. 9b.

Fig. 9 shows that control of the pressure compensated temperature at tray 4 had a fast and reliable tracking of the set-point, both during disturbances and for set-point tracking.

Furthermore a PFC was installed on top of the R/F controller which controls the heavy component measurement of the analyzer by modifying the set-point of the R/F controller. The identified parameters of an approximating first-order model between the reflux-to-flow ratio set-point and the heavy component are shown in Table 5.

Due to the optimizing characteristics of this loop and to prevent unnecessary disturbances to the column, this loop was tuned very slowly with a closed-loop settling time of 600 min (approximately 20 times the time constant).

Table 5: Approximating delayed first-order model between the reflux-to-flow ratio set-point and the heavy content

Gain [%/%] [heavy-content/(R/F)]	Time constant [min]	Dead time [min]
-0.00534	33	9

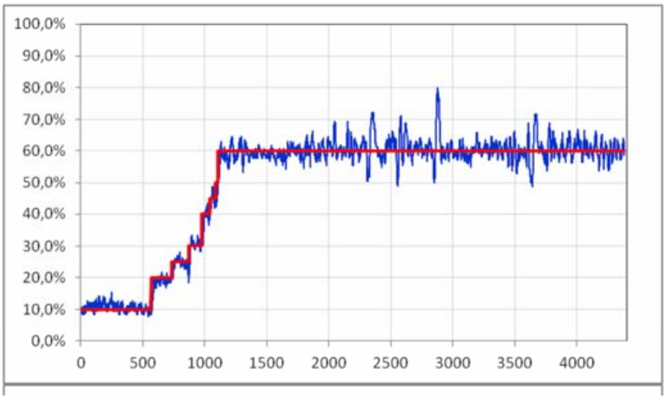


Fig. 9a. Pressure compensated temperature of the tailing column (tray 53) using PFC (red line: set-point, blue line: temperature at tray 53)

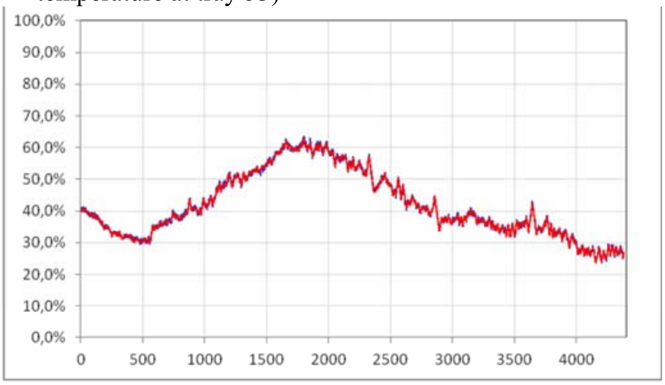


Fig. 9b. Heating medium flow of the tailing column using PID (red line: set-point, blue line: heating medium flow)

The temperature controller shows a fast and reliable tracking of the desired set-point, both during disturbances (see reflux flow plot) and for set-point tracking. The quality controller shows an adequate control of the quality, but it has to kept in mind, that the analyzer in this column is running at the lower end of the measurement range (with the depending disadvantages). The small spikes observable in the R/F plot descend form stepwise change of the feed set-point by the operator. They are compensated by the R/F controller without influencing the quality. At the end of the plot the R/F controller runs into saturation mode because the minimum allowed reflux for the column has been reached.

A three day long record of the controlled variable (pressure controlled temperature of tray 53) is plotted in Fig. 10a. Heating medium flow and the valve position are plotted in Fig. 10b.

5. CONCLUSION

PFC algorithm was installed successfully for three petro-chemical plants. The PFC was realized in a plant-wide Honeywell DCS (Decentralized Control System) (TDC2000/TDC3000). A generic implementation for the application module in this DCS was chosen. A bumpless switching algorithm presented by Haber and Zabet (2011) and an automatic initialization detection were realized.

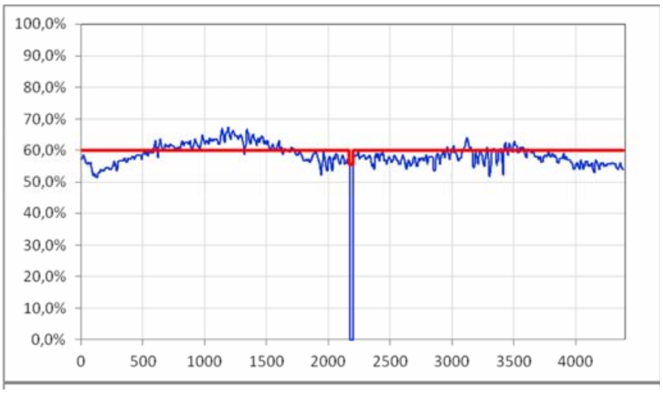


Fig. 10a. Control behavior of heavy component in distillate flow using PFC (red line: set-point, blue line: heavy component)

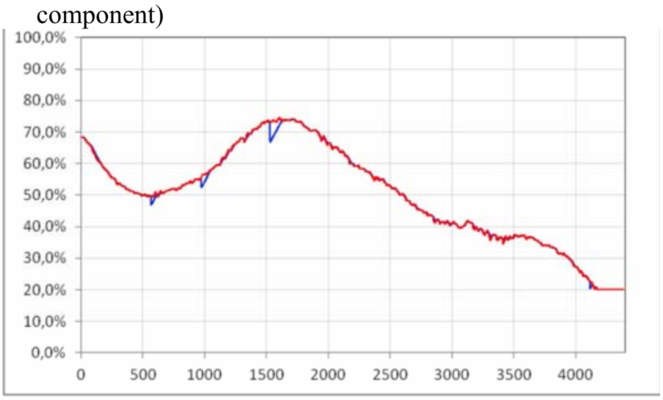


Fig. 10b. Reflux-to-feed ratio (red line: set-point , blue line: reflux-to-feed ratio)

The tuning of the PFC algorithm was easier than that of the PI(D) controller since PFC tuning is just based on a single parameter indicating the closed loop settling time. The control behavior was also much more robust and reliable than with PI(D) control. A numeric comparison of the control performance could not be done as the old control system was substituted by the new one. The PFC has proved itself to be a reliable and suitable replacement for the old partly more complex algorithms used at the site for many years before.

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