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Adaptive State Feedback Predictive Control and Expert Control for a Delayed Coking Furnace*

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Abstract An adaptive state feedback predictive control (SFPC) scheme and an expert control scheme are presented and applied to the temperature control of a 1200 $kt \cdot a^{-1}$ delayed coking furnace, which is the key equipment for the delayed coking process. Adaptive SFPC is used to improve the performance of temperature control in normal operation. A simplified nonlinear model on the basis of first principles of the furnace is developed to obtain a state space model by linearization. Taking advantage of the nonlinear model, an online model adapting method is presented to accommodate the dynamic change of process characteristics because of tube coking and load changes . To compensate the large inverse response of outlet temperature resulting from the sudden increase of injected steam of a particular velocity to tubes, a monitoring method and an expert control scheme based on heat balance calculation are proposed. Industrial implementation shows the effectiveness and feasibility of the proposed control strategy.
Keywords delayed coking furnace, adaptive control, state feedback, predictive control, expert control

1 INTRODUCTION

Heavy oil upgrading is an important method for refineries to enhance economic benefits. Since its first commercial application in 1930 [1], delayed coking has been a widely used process for bottom of the barrel upgrading, mainly because of its ability to handle different kinds of heavy oil, even the heaviest and contaminated residues [2–4]. The productive capacity of delayed coking in China holds the second position in the world and has increased rapidly in recent years [4].

Coking furnace is the key equipment of the delayed coking unit [5, 6]. The outlet temperature of the furnace has a direct influence on the yield of the final products. Also, it is one of the major operation parameters that affect the coking rate in the tubes. Coke deposited in the tubes raises the tube metal temperature (TMT) and increases pressure drop and, thus, shortens the run length of the coking unit. Applying advanced control strategy to get a tight control of the outlet temperature helps to maintain the product yield and decreases the coking rate caused by a temporarily high temperature.

The most commonly used regular control is the proportional-integral derivative (PID) cascade control, but it is difficult to maintain the temperature at a set point because the coking furnace suffers from more disturbances compared with other furnaces in a refinery plant. A model-free adaptive controller is implemented for a coking furnace [7]. In Ref. [8], fuzzy logic technique is used for a vacuum furnace. A feedforward variable structural PID controller is developed for the temperature control of polymerase chain reaction [9]. Originated from applications in petroleum refineries and power plants in the late 1970s, model predictive control (MPC) has been widely implemented in a variety of fields [10]. In Ref. [11], modular multivariable dynamic matrix control is applied to a refinery furnace. A predictive functional controller with a similar proportional integral optimal regulator structure is implemented to a coking heater [12]. The State Feedback Predictive Control (SFPC) has both the advantages of MPC and state feedback control, and it outperforms input-output model based MPC in dealing with unmeasured disturbances [13, 14].

Because of the inevitable coking in tubes and change of load, the dynamic characteristics of a furnace vary with time and with nonlinear property during one running period. For nonlinear systems, multi-model control is an effective method [15–18]. However, in practice the difficulty is how to obtain the linear models at different operating points. A rigorous model for the whole coking furnace [19] is too complex to implement or impossible in the current state for control purpose. In Refs. [20, 21], lumped-parameter models are used in the predictive function control of a heat exchanger and a chemical batch reactor, and favorable results are achieved.

For this furnace, steam is injected in its tubes to increase the velocity of the feed stream in the furnace to decrease the coking rate. The steam injected at a velocity sometimes increases suddenly and this will cause a large inverse response in the outlet temperature (this will be discussed in Section 3.3.1). This problem has not been addressed by others.

In this article, a simplified control model based on the first principles for a coking furnace is developed, and SFPC controllers are designed to stabilize the outlet temperatures. An online model adapting method is presented to accommodate the dynamics of process changes. An expert control algorithm is proposed to compensate the large inverse response resulted by a sudden increase in steam velocity. Online application results demonstrate that the proposed control strategy is effective and feasible.

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2 PROCESS DESCRIPTION AND ORIGINAL CONTROL SYSTEM ANALYSIS

2.1 Process description

As shown in Fig. 1, the feed is divided into four branches before entering the furnace. After rapidly passing through the furnace, it is heated to reach the thermal cracking temperature of approximately 500°C, then enters the downstream coking drum, where the main cracking and condensation reactions occur. Every feed pass has one steam injection point that lies in the convection section and two water injection points that lie in the radiation section.

The furnace is a double radiation furnace that has two symmetric chambers. In each of the chambers, there are two pass of feed tubes, and the two pass tubes are furthermore separated by a bridge wall, thus forming four separated fireboxes.

2.2 Original control system analysis

The disturbances that influence the outlet temperatures of the furnace mainly include: (1) variations of the flow rate, temperature, and components of the fresh feed and the recycle oil; (2) variations of pressure and heating value of the fuel gas; (3) variations of the flow rate and temperature of the water or steam that is injected in the tubes; (4) the drum switch.

Suffering from many disturbances that enter the process at different points, the original outlet temperature and chamber temperature PID cascade control cannot work satisfactorily under some situations. Fig. 2 is the response of a PID controller when there is

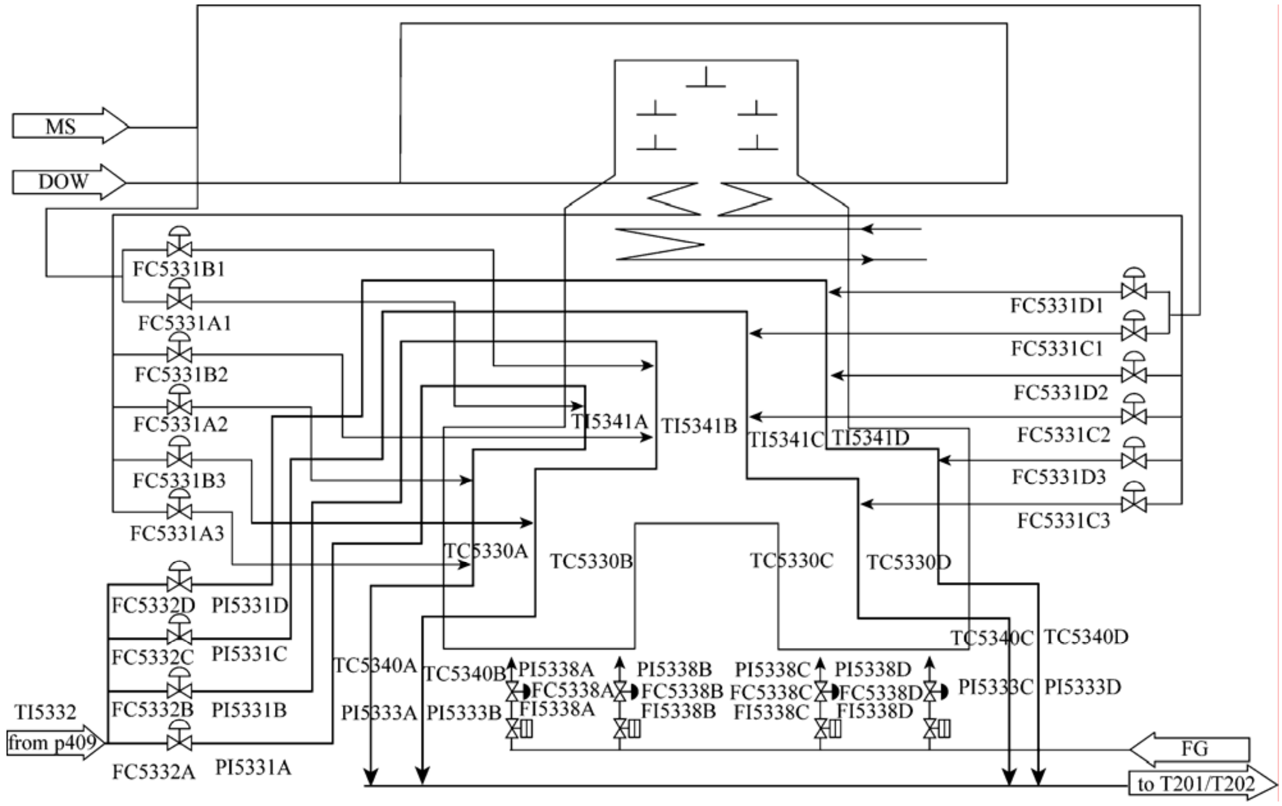

Figure 1 A furnace of the delayed coking unit

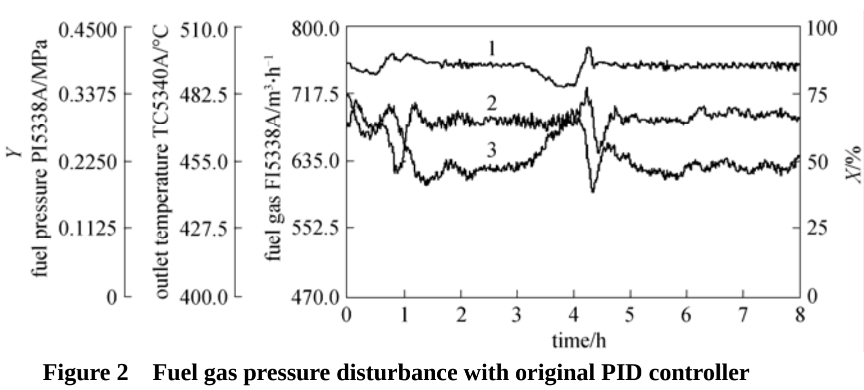

Figure 2 Fuel gas pressure disturbance with original PID controller

(X: scaled variable, $X = \dfrac{Y - Y_{\min}}{Y_{\max} - Y_{\min}} \times 100\%$, Y: original variable)

1—fuel pressure PI5338; 2—outlet temperature TC5340A; 3—fuel gas FI5338A a fuel pressure disturbance.

A large inverse response for the outlet temperature can be caused by a sudden increase of the steam of a particular velocity. Fig. 3 and Fig. 4 are the responses when the process is under manual and PID control. In Fig. 3, though the flow rate of the fuel gas was reduced quickly by manual operation from about 600 $m^3 \cdot h^{-1}$ to 350 $m^3 \cdot h^{-1}$, the outlet temperature still reached 515°C. In Fig. 4, the outlet temperature has reached 540°C under PID control. Such a high temperature greatly speeds up the coking rate in the tubes. Actually, the coking layer of the tubes in this pass is thicker than the others indicated by the fire side TMT after this fault. The large variations severely affect the safe operation of the furnace.

To sum up, the coking furnace is a multi-input single output (MISO) system, which has many disturbances and constraints, and sometimes a dangerous situation might occur when a disturbance is observed because of the large steam increase. The conventional PID cascade control can hardly get satisfactory results.

3 ADVANCED CONTROL SCHEME DESIGN

3.1 Control objectives and strategies

The goal of advanced control for the coking furnace is to stabilize the outlet temperature, to improve its disturbances rejection capability, and to prolong the run length of the furnace between two consecutive decoking times. The principle diagram of the advanced furnace control system is shown in Fig. 5, in which there are two controllers, one is adaptive SFPC and the other is steam increase expert control. In most of the time, the SFPC controller is applied in order to keep the outlet temperature at its set point under different disturbances. Because the flow rate of steam has significant influence on the outlet temperature, the expert controller handling steam increase events is

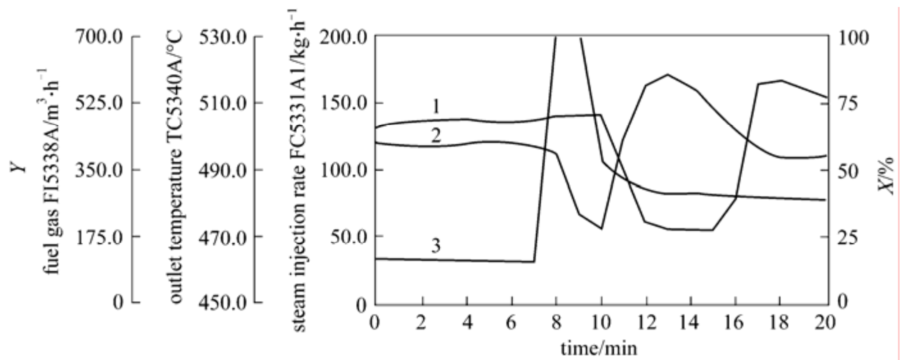

Figure 3 Influence of a sudden increase in steam injection on the outlet temperature
1—fuel gas FI5338A; 2—outlet temperature TC5340A; 3—steam injection rate FC5331A1

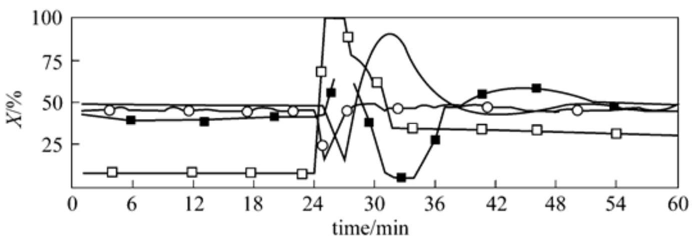

Figure 4 The response of outlet temperature to a sudden increase in steam injection under PID control
—— outlet temperature TC5340C; ○ feed FC5332C; ■ fuel gas FI5338C; □ steam injection rate FC5331C1

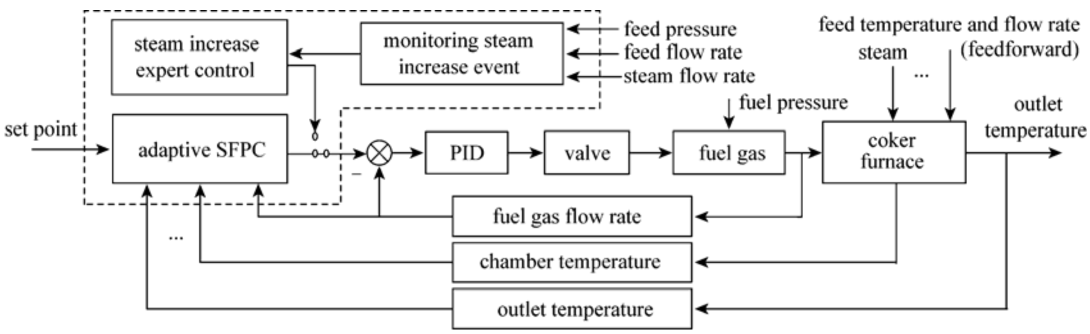

Figure 5 The principle diagram of furnace advanced control

applied separately and can be switched to and from the adaptive SFPC controller automatically.

3.2 Model predictive control for the furnace outlet temperature

3.2.1 *Control strategy*

As described in Section 2.1, the four passes of feed are heated separately in four fireboxes, so there is little heat couple among them. Four single-loop SFPC controllers are utilized to substitute for the original PID cascade controllers to improve control performance.

The temperature-pressure compensation has been made for the fuel gas. A PID controller maintains the compensated fuel gas flow rate at set point is added in the distributed control system (DCS). This is important to handle the disturbance caused by the variation of fuel gas pressure.

The chamber temperature is taken as a state variable. Disturbances that affect the chamber temperature first, such as the change of fuel gas heating value, will be suppressed by state feedback before they influence the outlet temperature.

A method for anti-windup is included in the controllers. When the valve position of the fuel gas PID controller reaches its high limit set by engineers, the SFPC controller stops increasing the set point of the fuel flow PID controller, and *vice versa*.

The feed temperature and flow rate are incorporated in the SFPC controller as disturbance variables to implement feedforward control.

The high limits of the chamber temperature and TMT are considered. When either of the limits is active, the flow rate of the fuel gas can only be decreased.

3.2.2 *Structure of the control system*

The controller structure for pass A is illustrated in Table 1.

Table1 Structure of the controller for pass A

Variables	Descriptions	Points①
controlled variable	pass A outlet temperature	TC5340A. PV
manipulated variable	set point of fuel gas flow controller A	FC5338A. SV
state variables	pass A outlet temperature	TC5340A. PV
	chamber temperature A	TC5330A. PV
	fuel gas flow rate A	FC5338A. PV
disturbance variables	pass A feed flow rate	FC5332A. SV
	feed inlet temperature	TI5332. PV

① PV＝process value; SV＝set value.

3.2.3 *A state space model for the coking furnace*

To a large extent, the type and accuracy of the model determine the performance of an MPC controller. The most widely used model form in the state of the art commercial MPC products is the input-output model. By careful design and implementation of test signals, a parametric or nonparametric model is obtained on the basis of the test data. However, the input-output model has its disadvantages. First, it is hard to carry out step response test for the process with high safety requirements, such as coking furnaces. Second, from the view point of control structure, MPC on the basis of input-output models uses only output feedback, and its disturbance rejection capability is not satisfactory [13, 22]. When MPC uses a state space model, it can have both state and output feedback and helps to realize a more flexible and comprehensive control scheme. Third, for a nonlinear process, the model cannot be expected to accurately predict process behavior beyond the range of the test data. And the nonlinear state space model needed by the model adapting method cannot be obtained by the plant test method. The best alternative is to build a model based on first principles.

Assuming that: (1) because an air preheater is installed, the heat loss of fuel gas changes little, and it is not included in the chamber energy balance equation, only the effective heat transferred to the feed is considered; (2) the heat transferred to the feed is proportional to the temperature difference between the chamber temperature and the outlet temperature; (3) thermophysical properties of the fluids are constant. The coking furnace can be described by the following energy and material balances:

$$\rho_1 V_1 c_{p1}\frac{\mathrm{d}T_\mathrm{o}}{\mathrm{d}t} = F_\mathrm{i} c_{p1}\left(T_\mathrm{i}-T_\mathrm{o}\right)+UA\left(T_\mathrm{l}-T_\mathrm{o}\right) \tag{1}$$

$$\rho_2 V_2 c_{p2}\frac{\mathrm{d}T_\mathrm{l}}{\mathrm{d}t} = -UA\left(T_\mathrm{l}-T_\mathrm{o}\right)+K_3F_3 \tag{2}$$

$$T_3\frac{\mathrm{d}F_3}{\mathrm{d}t} = -F_3+F_{3\mathrm{s}} \tag{3}$$

where T_o, T_l, and F_3 are the outlet temperature, chamber temperature, and temperature-pressure compensated flow rate of fuel gas, respectively; F_i and T_i are, respectively, the flow rate and inlet temperature of the feed; U and A stand, respectively, for the average heat transfer coefficient and the area for heat transfer; K_3 denotes heat transferred to the feed when a unit of fuel gas is burned; ρ_1 and c_{p1} are, respectively, the density and specific heat of the feed in the tubes; ρ_2 and $c_{\mathrm{p}2}$ are, respectively, the density and specific heat of the air in the chamber, and $F_{3\mathrm{s}}$ is the set point of F_3. The parameters are obtained by using the equipment dimensions and fitting to the measurement data. The heat capacity $\rho_1V_1c_{p1}$ and $\rho_2V_2c_{p2}$ are adjusted by respective correcting factors to fitting the real-time trend.

The temperature-pressure compensation for fuel gas flow is:

$$F_\mathrm{comp} = F\sqrt{\frac{T_\mathrm{ref}}{T}\frac{\left(p+p_0\right)}{\left(p_\mathrm{ref}+p_0\right)}} \tag{4}$$

where F_comp and F are the compensated and measured flow rates, respectively; T_ref and T are, respectively, the design and measured temperatures; p_ref and p are, respectively, the design and measured pressures, and p_0 is a factor to convert the gauge pressure to the absolute pressure.

By linearizing Eqs. (1)−(3) at the operating point $\boldsymbol{O}=\left(T_\mathrm{o}^*,T_\mathrm{l}^*,T_\mathrm{i}^*,F_\mathrm{i}^*,F_3^*\right)$, the following state space model of the coking furnace is obtained:

$$\begin{cases} \dot{x} = \boldsymbol{A}\boldsymbol{x} + \boldsymbol{B}\boldsymbol{u} + \boldsymbol{F}\boldsymbol{v} \\ y = \boldsymbol{C}\boldsymbol{x} \end{cases} \tag{5}$$

where $\boldsymbol{x} = [\Delta T_o \quad \Delta T_l \quad \Delta F_3]^T$, $\boldsymbol{v} = [\Delta F_i \quad \Delta T_i]^T$, $u = \Delta F_{3s}$, $y = \Delta T_o$, $\Delta F_i = F_i - F_i^*$, $\Delta T_o = T_o - T_o^*$, $\Delta T_l = T_l - T_l^*$, $\Delta T_i = T_i - T_i^*$, $\Delta F_3 = F_3 - F_3^*$, $\Delta F_{3s} = F_{3s} - F_{3s}^*$.

3.2.4 *Online model adapting method*

According to Eqs. (1)–(3), at a steady state we have

$$F_i c_{p1}(T_o - T_i) = K_3 F_{3s} \tag{6}$$

It means that when the process reaches a steady state, the heat absorbed by the feed is equal to the effective fuel gas combustion heat. For the outlet temperature, the following equation holds:

$$T_o = \frac{K_3}{c_{p1} F_i} F_{3s} + T_i \tag{7}$$

As can be seen from Eq. (7), when K_3 and c_{p1} are held constant, the gain of the model is changed corresponding to the feed flow rate. If F_i changes a lot, the parameters of the state space model should be adjusted to prevent deterioration of the controller's performance. In another case, while the thickness of the coking layer in the tubes increases, the thermal resistance is becoming larger and thermal efficiency is lowered, then U and K_3 are gradually decreased. In order to maintain the outlet temperature at its set point, the chamber temperature should be raised and the fuel gas consumption is also increased. If the chamber temperature or TMT reaches its upper limit, the feed flow rate should be decreased to avoid constraint violation. In this case the process behavior is also changed. So the parameters in the model should be re-evaluated when the furnace has operated over a relative long time.

To sum up, when the operating point is changed, it is necessary to adjust the model online. The online model adapting method is shown in Fig. 6.

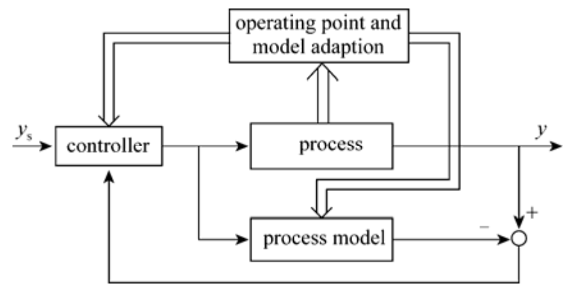

Figure 6 The basic method for on-line model adapting

The principle for the online model adapting method is to keep the structure of the model based on first principles unchanged, and then to calculate the new parameters for the model, and update controller parameters if necessary when the operating point is moving obviously. In every control cycle, a set of process variables is evaluated in order to decide whether or not the system is at a steady state. The evaluation criterion is:

$$\frac{1}{N}\sum_{i=1}^{N_y}\sum_{j=1}^{N}\left|\frac{y_{ij} - \overline{y}_i}{\overline{y}_i}\right| < \varepsilon \tag{8}$$

where N_y is the number of the variables that is selected to describe the system's status, N is the length of the history data, $\overline{y}_i$ is the average value for the i-th variable, and ε is a predefined value.

When the system is at a steady state, a new operating point $\boldsymbol{O}_{\text{new}}$ is determined. What should be decided is whether or not the operating point has shifted. This is done by using Eq. (9)

$$\|\boldsymbol{O}_{\text{new}} - \boldsymbol{O}\|_2^2 > d \tag{9}$$

where $\|\cdot\|_2$ is the 2-norm, and d is a predefined distance.

If there is a large change between the new operating point and that of the model is built, the state space model is updated on the basis of the new operating point. This helps to reduce the frequency of model updating, and the robustness of model predictive control can be fully utilized. The procedure for the algorithm is shown in Fig. 7.

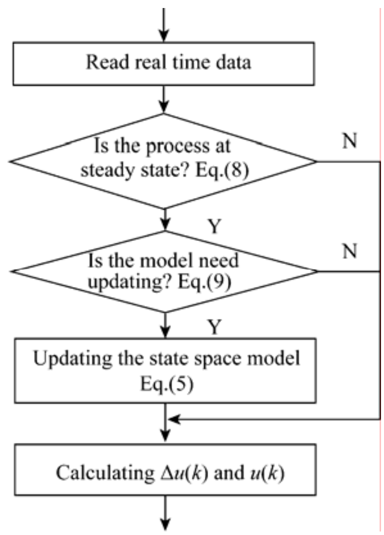

Figure 7 Procedure for on-line model updating

3.3 Expert control for the sudden increase of steam of a particular velocity

3.3.1 *Monitoring of the suddenly increasing of steam of a given velocity*

Because of the original design problem of the steam injection system, abnormal events of the steam injection system that the steam of a particular velocity is condensed to water in the pipe sometimes occur. When it happens, steam of a particular velocity will decrease. Then, the operator will adjust the flow rate of the steam of a particular velocity to its normal value. However, because of the dead band and high pressure difference for the steam flow control valve, the flow rate does not change as expected. It has a sudden increase when the valve position changed enough to overcome the dead band, as shown in Fig. 3 and Fig. 4.

The condensed water is vaporized when it mixes with the high temperature feed, and the steam flow rate is much higher than its normal value, which causes the pressure in the tube to increase. Then feed flow rate will be lowered temporarily.

To monitor the events of sudden increases of steam of a particular velocity, three conditions are determined as the criteria by fundamental analysis:

(1) The increase rate of the steam flow exceeds a low threshold that is set according to history data;

(2) The increased rate of the pressure downstream the feed flow control valve exceeds a predefined value;

(3) The decreased rate of feed flow is greater than a predefined value.

When the three conditions are satisfied at the same time, it is believed that the steam injection is suddenly increased and measures should be taken immediately to avoid the outlet temperature violate its upper limit.

3.3.2 *Measures taken after the sudden increase of steam at a particular velocity*

The vaporized condensed water and injected steam absorb much more heat than in normal operation, and the residence time of feed is decreased at the same time, which could cause the outlet temperature to decrease dramatically.

However, both decreasing the feed flow and increasing the heat transfer coefficient make the temperature increase subsequently. The increase of overall heat transfer coefficient from the tube skin to the feed flow can be obtained from Eqs. (10)–(12)

$$\frac{1}{K}=\frac{1}{\alpha_{\mathrm{i}}}+\frac{d_{\mathrm{w}}}{\lambda_{\mathrm{w}}}+\frac{d_{\mathrm{ck}}}{\lambda_{\mathrm{ck}}} \tag{10}$$

$$\alpha_{\mathrm{i}} \propto V_{\mathrm{f}}^{0.8} \tag{11}$$

$$V_{\mathrm{f}}=\frac{V_l+V_{\mathrm{g}}+V_{\mathrm{st}}}{A_{\mathrm{tube}}} \tag{12}$$

where K is the overall heat transfer coefficient; α_{i} is the inside heat transfer coefficient; d_{w} and d_{ck} are, respectively, the tube wall and coking layer thickness; λ_{w} and λ_{ck} denote, respectively, the thermal conductivity of the tube metal and coke, V_{f} is the velocity of feed in the tube; V_l and V_{g} stand for the liquid and vapor phase volume feed flow rate respectively; V_{st} is the steam flow rate, and A_{tube} is the cross sectional area of the tube.

Actually, the dynamics of outlet temperature to steam of a particular velocity exhibit inverse response, and the outlet temperature will settle at a value larger than the previous steady state value if the fuel maintains unchanged. However, the PID controller does not know this change. The fuel gas is increased rapidly in the initial temperature decrease phase as shown in Fig. 4, which is an error correction action. When the temperature trend inverts, due to superposition of inherent dynamics and effect of rapid increase of fuel, the temperature will increase sharply. It is then usually too late for the PID controller to take action to restore the temperature to its normal value. The operator has to switch the controller to manual mode.

This is because the whole transient process of outlet temperature is only approximately 10 min, in which the outlet temperature decreases about 3 min. It is not practical to compensate the temperature decrease. Measures taken should focus on preventing the outlet temperature from exceeding its upper limit. An expert control scheme is adopted here. The control rules are presented below.

(1) Keep the fuel gas flow for 1 min.

In this phase, a new set value of the fuel gas on the basis of heat balance is calculated.

(2) Decrease the fuel gas flow rate to the set value with a predefined rate, for instance, 100 $\mathrm{m}^3\cdot\mathrm{min}^{-1}$. The new set point is determined by Eq. (17).

(3) Keep the fuel gas flow rate unchanged until the outlet temperature reaches its highest value and begins to decrease.

(4) Increase the fuel gas flow rate to the original value with the same predefined rate.

(5) Keep the fuel gas flow rate for a period, and then switch to the SFPC controller.

The key point for the expert system is to decide the value to which the fuel gas flow should decrease after the sudden increase of steam of a particular velocity.

According to the energy equilibrium Eq. (6), if the outlet temperature remains unchanged after the sudden increase of steam of a particular velocity, the fuel gas should be

$$F_{\mathrm{3s,new}}=\frac{F_{\mathrm{i,new}}}{F_{\mathrm{i}}}\cdot\frac{K_3}{K_{\mathrm{3,new}}}F_{\mathrm{3s}} \tag{13}$$

The subscript "new" in Eq. (13) donates the values after the sudden increase of steam of a particular velocity. Both decrease of the feed and increase of the heat transfer coefficient require the decrease of fuel gas flow rate. The feed flow rate can be measured, but the effective heat $K_{\mathrm{3,new}}$ is not obtained readily. An approximate method is proposed here. When the system is at steady state, there is a relationship between the heat transfer rates:

$$Q=K_{\mathrm{r}}\left(T_{\mathrm{l}}^4-T_{\mathrm{m}}^4\right)=K_3F_{\mathrm{3s}} \tag{14}$$

where Q is the heat transfer rate; T_{m} is the temperature of the outer TMT, and K_{r} is the radiation heat transfer coefficient, which is assumed to be constant. In Eq. (14), the convective heat is neglected.

When the fuel remains unchanged after a sudden increase of steam of a particular velocity, the heat transfer rate is

$$Q_{\mathrm{new}}=K_{\mathrm{r}}\left(T_{\mathrm{l,new}}^4-T_{\mathrm{m,new}}^4\right)=K_{\mathrm{3,new}}F_{\mathrm{3s}} \tag{15}$$

From Eqs. (14) and (15), we have

$$\frac{K_3}{K_{\mathrm{3,new}}}=\frac{T_{\mathrm{l}}^4-T_{\mathrm{m}}^4}{T_{\mathrm{l,new}}^4-T_{\mathrm{m,new}}^4} \tag{16}$$

Equation (13) holds if the system is at a steady state, but the response of the outlet temperature for the fuel gas decrease does not reach its steady state in the handling process, and the increase of steam of a particular velocity decreases the severity of reaction in the tubes, with the reduce in absorbed reaction heat. So it is necessary to correct the result of Eq. (13) by α.

$$F_{\mathrm{3s,new}}=\alpha\frac{F_{\mathrm{i,new}}}{F_{\mathrm{i}}}\cdot\frac{K_3}{K_{\mathrm{3,new}}}F_{\mathrm{3s}} \tag{17}$$

where α is a correcting factor and $\alpha<1$.

4 CONTROL PERFORMANCE

It has been more than one year since the whole

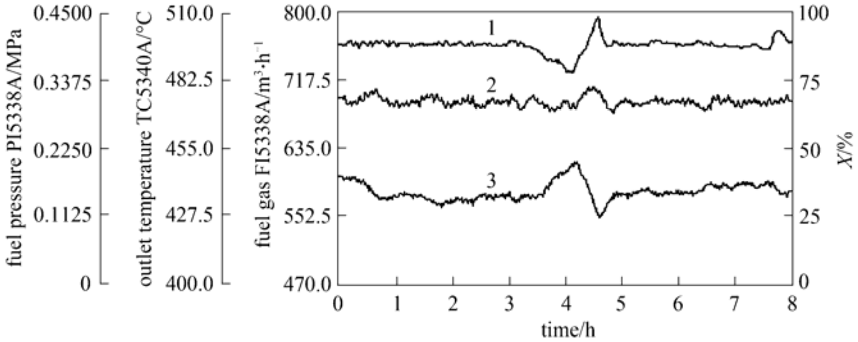

Figure 8 Fuel gas pressure disturbance rejection with adaptive SFPC
1—fuel pressure PI5338A; 2—outlet temperature TC5340A; 3—fuel gas FI5338A

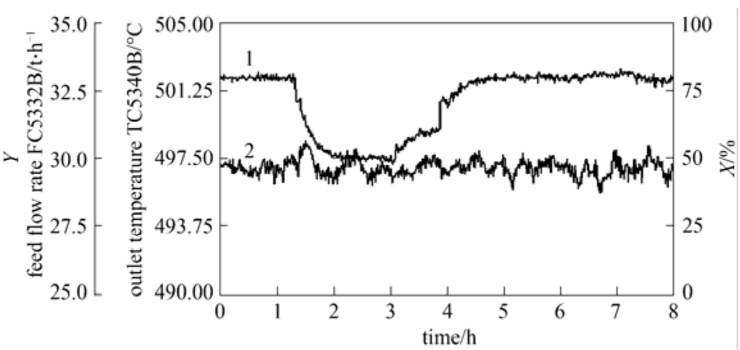

Figure 9 Load disturbance rejection with adaptive SFPC
1—feed flow rate FC5332B; 2—outlet temperature TC5340B

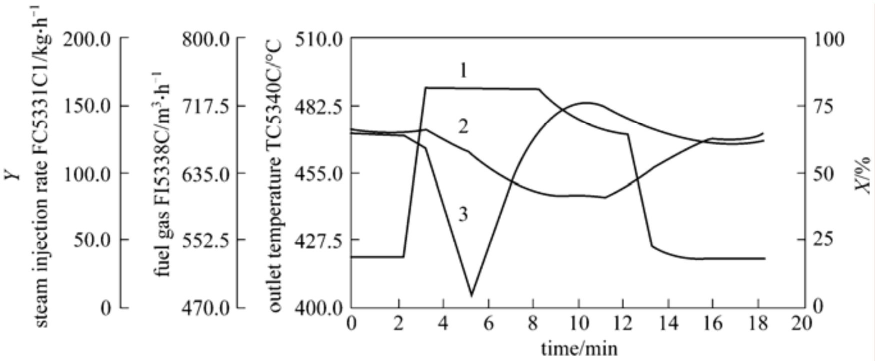

Figure 10 The performance of expert control
1—steam injection rate FC5331C1; 2—fuel gas FI5338C; 3—outlet temperature TC5340C

control strategy was implemented in 2006. The application has a service factor of approximately 95%, and the control results proved its efficiency and feasibility.

4.1 Adaptive SFPC in normal operation

Adaptive SFPC is used in this application. The control interval is chosen as 10 s, the predictive horizon is 42, and the control decay coefficient is 0.8. The control interval is chosen relatively short to monitoring the steam change.

Figure 8 illustrates the response of furnace control system to a fuel gas pressure disturbance. Compared to Fig. 2, the magnitudes of the disturbances in both figures are approximately the same, but the variation of the outlet temperature is reduced from 10°C to 2°C.

Figure 9 illustrates the response to a load disturbance. A 10% load decrease has little influence on the outlet temperature because of feedforward control and model adaptation.

4.2 Expert control for the increase of steam

The steam increasing with expert control is shown in Fig. 10. The highest temperature is reduced to approximately 505°C. The regulating effect is better than that attained by the skilled operator on condition of finding and treating in time as shown in Fig. 3. It should be pointed out that this procedure has greatly released the burden of the operators and unified the operation.

4.3 Control performance during commissioning

An 8 d commissioning test was conducted by the plant. The outlet temperature of pass A is shown in Fig. 11. The standard deviations of outlet temperatures before and after advanced process control (APC) are shown in Table 2.

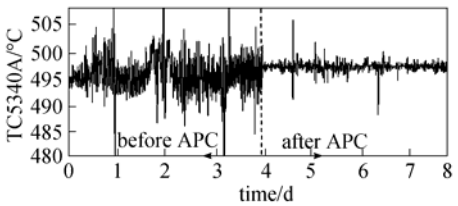

Figure 11 Outlet temperature of pass A before and after APC

Table 2 Standard deviations of outlet temperatures before and after APC

	Before APC/°C	After APC/°C	Decrease rate/%
pass A	3.209	1.066	67
pass B	2.749	1.357	50
pass C	1.258	1.079	14
pass D	1.896	1.453	23

5 CONCLUSIONS

In this article, an adaptive SFPC with an expert control scheme is proposed and applied to control the outlet temperature of a delayed coking furnace. Adaptive SFPC and online model adaptation on the basis of a simplified fundamental model significantly improves the control performance. The standard deviations of the outlet temperatures of the four passes are reduced by an average of 38%, of which the best one is 67%. The expert control prevents the outlet temperatures from violating their high limits and, thus, helps the safe operation of the furnace and decreases the coking rate. Compared to the previously implemented PID control scheme, the run length of the furnace between two consecutive decoking times is prolonged from 5 months to 11 months and the downstream process performance is also improved. Therefore, the proposed adaptive SFPC is effective for the complex process and the expert control method is effective for the special control problem.

NOMENCLATURE

A	area for heat transfer, m^2
A_{tube}	cross sectional area of the tube
c_{p1}, c_{p2}	specific heat of the feed and air, $kJ\cdot kg^{-1}\cdot K^{-1}$
d	a predefined value to judge if the operating point has shifted
d_w, d_{ck}	tube wall and coking layer thickness, m
F_i	feed flow rate, $kg\cdot s^{-1}$
F_3	temperature-pressure compensated flow rate of fuel gas, $m^3\cdot s^{-1}$
F_{3s}	set point for F_3, $m^3\cdot s^{-1}$
K	overall heat transfer coefficient between tube metal skin and feed, $W\cdot m^{-2}\cdot K^{-1}$
K_3	heat transferred to the feed per unit volume of fuel gas, $kJ\cdot m^{-3}$
N	length of the history data
N_y	number of variables to describe the system's operating point
$\boldsymbol{O}$	operating point
p	measured pressure for fuel gas flow rate compensation, kPa
p_0	factor to convert gauge pressure to an absolute value, kPa
T	measured temperature for fuel gas flow rate compensation, K
T_i, T_o	inlet and outlet temperature of feed, K
T_1	chamber temperature, K
T_3	time constant for a first order system
U	average heat transfer coefficient between flue gas and feed, $W\cdot m^{-2}\cdot K^{-1}$
$\boldsymbol{u}$	system input
V_f	velocity of feed in the tubes, $m\cdot s^{-1}$
V_l, V_g	liquid and vapor phase volume feed flow rate, $m^3\cdot s^{-1}$
V_{st}	steam flow rate, $m^3\cdot s^{-1}$
V_1, V_2	volumes of tubes and chamber, m^3
X	scaled variable
$\boldsymbol{x}$	state vector
$\boldsymbol{y}$	system output
α_i	inside convective heat transfer coefficient, $W\cdot m^{-2}\cdot K^{-1}$
ε	a predefined value to judge if the system reaches steady state
λ_w, λ_{ck}	thermal conductivity of the tube metal and coke, $W\cdot m^{-1}\cdot K^{-1}$
ρ_1	average density of the feed and steam stream in the tubes, $kg\cdot m^{-3}$
ρ_2	density of the air in the chamber, $kg\cdot m^{-3}$

Subscripts

i	index subscripts of variables
new	a new state or value
ref	reference value for fuel gas flow rate compensation

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