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Control of a fluid catalytic cracking unit based on proportional-integral reduced order observers

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Abstract

The fluid catalytic cracking (FCC) unit is one of the most complex interactive processes in the refining industry, and is difficult to operate and control. This work deals with the regulative control law design for stabilization of the reactor and regenerator temperatures, considering that the kinetics terms are poorly known. The proposed control law makes use of on-line estimates of the input/output modeling errors, obtained from a proportional-integral reduced order observer. The structure of the proposed controller is similar to a proportional-double integral compensator (PI^2) such that a new parameterization of the controllers gains is given in terms of the closed-loop and estimation time constants. The performance of the control scheme proposed here is analyzed via numerical simulations. © 1999 Elsevier Science S.A. All rights reserved.

Keywords: FCC units; Modeling error; PI^2 compensators; Control gains tuning

1. Introduction

The FCC units are one of the most important process units in oil refineries for the largest yield of products reached and the importance of them. A small benefit in this process is economically attractive; besides, the environmental regulations related with gas emissions like CO, CO_2, NO_x and SO_x, are to date, very important. Therefore, advanced process control is needed. However, FCC units are process physically complex and prove several difficult to operate and control. The regenerator and the reactor sections of FCC units are highly interactive with difficult dynamics such knowledge is poor for the complexity of the chemical kinetics mainly.

The FCC process is an excellent opportunity to realize substantial benefits for both process industry and control theory. From the control theory point of view, the study of the regulation problem for uncertain systems is an important issue in robust control theory. The main task is to design of a control law in the presence of significant system uncertainties like modeling errors, load disturbances, variations of system parameters, etc.

A great development in nonlinear control theory in the last years has been the characterization of linearizable systems. That is, systems that can be linearized by means of a change of coordinates and state feedback. Linearization of nonlinear systems is related with the cancellation of input/output nonlinearities under the assumption of the perfect knowledge of these terms. A feedback control scheme designed with this approach guarantees closed-loop stability and output tracking [1].

In general, the problem of obtaining an exact knowledge of the nonlinearities present in a system is not an easy task. Given that the nonlinear terms are not fully known, the linearizing controller can provide a poor performance or even induce instabilities [2]. For the case of chemical reactors stabilization, several approaches are based on Taylor linearization of the reactor dynamic with the assumption that the uncertainties belongs to certain conic sector [3,4]. These approaches have several weaknesses. In the local linear approximation, the main properties of the chemical reactors are not exploited. On the other hand, many uncertainties and disturbances can not be included in a conic sector. This situation can lead to conservative control law designs and consequently poor closed-loop performance.

Recently, using filtering techniques [5] and calorimetric balances [6,7], new techniques to get on-line estimates of the uncertainty terms in chemical reactors, for both modeling and control purposes have been developed. The advantage of

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these approaches is their easy computational implementation. Besides, their structure has a strong physical meaning, without the restrictions of the methods above mentioned. However, the calorimetric balance techniques become unstable when the measurements are noisy.

On the other hand, the main drawback of the classical feedback controllers (i.e. P, PI and PID controllers) is that they cannot provide sufficient stability conditions when modeling errors and disturbances are present in the process. Besides, the optimum tuning of the control gains is difficult to determinate [8].

The FCC control problem has been addressed with several approaches, beginning with the classical work of Kurihara [9] where the design of a control system for non-linear multivariable processes was developed from optimal feedback control theory. Other of this approaches commonly used is related with model predictive control, where the control computation is based on a general performance index and parameterization of the control variables in a non-linear model, which includes the relevant constraints [10–12].

In the practice the FCC units are frequently regulated by means of standard PID controller and manual control actions based on the knowledge of the operators of the refinery. Some authors, have studied the open-loop and the closed-loop behavior of the FCC units under the above approaches [13,14] where a optimal controllability regions were found as a function of the kind of control inputs and control objectives.

In this work, we propose a control design approach for temperature regulation in FCC units. In the spirit of calorimetric balances, we construct a second order observer to estimate the uncertainties (chemical kinetics terms) and use such estimates within an input/output linearization framework. The resulting controller is a linear one and only temperature measurements are required for its implementation. We show that this control law is equivalent to a proportional plus double integral compensator, which is known to be efficient in handling slow varying load disturbances [8].

This paper is organized as follows: Section 2 is related with the mathematical model of the FCC unit, which is employed in this work. Section 3 shows the design of the ideal control laws for the regulation of the temperatures in the regenerator and the outlet of the riser. In Section 4, the on-line uncertainty estimation using a reduced order filter and the coupling with the linearizing control for the design of the non-ideal control laws are presented. The numerical simulations carry-out to show the closed-loop performance of the FCC unit are presented in Section 5. Finally the conclusions of the paper are presented in the Section 6.

2. Mathematical model of the reactor–regenerator FCC units

A FCC unit consists mainly of a reactor–regenerator section, a fractionator and gas processing facilities. Feed to the reactor is preheated in a fired-heater, and it mixes with regenerated catalyst along the reactor riser, during the cracking reactions the catalyst is deactivated by coke deposited. The catalyst used is sent back to the regeneration section where the coke is burned off with air, after the regenerator reactor the catalyst is recirculated to the reactor. The regenerator can operate in either complete or partial combustion mode. In the last mode, coke remains on the regenerated catalyst and there is unconverted CO and CO_2 in the regenerator flue gas. In complete combustion mode, most of the carbon on regenerated catalyst is burned off and there is O_2 in excess in the regenerator [15].

The Lee and Groves model is employed in this work. This model made use of three lumps scheme [16] to describes the kinetics in the riser, considering plug flow behavior. The regenerator model was developed by Errazu et al. [17], where the coke deposition on the catalyst is taken from Kurihara [9], assuming perfect mixing (CSTR) behavior of the dense phase. In this work is considered the partial combustion mode, which has become more important on the industrial point of view [11].

2.1. Riser reactor model

Considering the residence time in the riser is of the order of ten seconds, the assumption of steady state ideal plug flow reactor model is used. The mass balance is applied to describe the gasoil and gasoline according with Weekman and Nace [16], producing the following equations:

1. Mass balance for gasoil (y_f)

$$W_{f,ri}\frac{dy_f}{dz} = F_f\frac{dy_g}{dz} - W_{f,ri}K_1(T_r)y_f^2[\mathrm{COR}]\Phi/\rho_f, \tag{1}$$

where K_1 is the kinetic constant, expressed by Arrhenius model, COR is the Catalyst–Oil ratio, $W_{f,ri}$ is the catalyst holdup in the riser and ρ_r is the density of the gasoil-feed.The deactivation of the catalyst by coke deposition Φ is given by

$$\Phi = \Phi_0\exp(-\alpha t_c[\mathrm{COR}]z) \tag{2}$$

as a function of the residence time of the riser t_c, the Catalyst–Oil ratio and the axial coordinate z.

2. Mass balance for gasoline (y_g)

$$W_{f,ri}\frac{dy_g}{dz} = F_f\frac{dy_g}{dz} - W_{f,ri}\alpha_2K_1(T_r)y_f^2 - K_3(T_r)y_g[\mathrm{COR}]\Phi/\rho_f. \tag{3}$$

The amount of coke produced is calculated employing the Kurihara's correlation [9]

$$C_{cat} = k_c\frac{t_c^{1/2}}{C_{rc}^{N/2}}\exp\left(\frac{-E_{cf}}{RT_{ro}}\right), \tag{4}$$

which is a function of the residence time t_c, the outlet temperature if the riser and the coke in the catalyst regenerated C_{rc}.

3. Energy balanceAdiabatic operation with no change of phase is assumed. In this way the energy required to perform the cracking reactions is taken from the gas–solid mixture. With this assumption, the following equation is obtained

$$W_{\mathrm{ri}}\frac{\mathrm{d}T_{\mathrm{ro}}}{\mathrm{d}z}=(F_{\mathrm{f}}+F_{\mathrm{s}})\frac{\mathrm{d}y_{\mathrm{f}}}{dz}+W_{ri}\Delta H_{\mathrm{f}}K_1(T_{\mathrm{r}})y_{\mathrm{f}}^2[\mathrm{COR}]\Phi/\rho_{\mathrm{ri}}Cp_{\mathrm{ri}}, \tag{5}$$

where ΔH_{f} is the reaction heat and F_{f} and F_{s} are the flows of the gasoil and catalyst respectively.

2.2. Regenerator reactor model

Considering that the residence time of the catalyst in the regenerator is frequently of the order of ten to twenty minutes and that the temperature and the amount of coke on catalyst are uniform throughout of the reactor dense bed, the assumption of CSTR behavior for control purposes is adequate [17].

1. Mass balance for coke on regenerated catalyst (C_{rc})

$$W_{\mathrm{rg}}\frac{\mathrm{d}C_{\mathrm{rc}}}{\mathrm{d}t}=F_{\mathrm{s}}(C_{\mathrm{st}}-C_{\mathrm{rc}})-R_{\mathrm{cb}}. \tag{6}$$

2. Energy balance

$$W_{\mathrm{rg}}Cp_{\mathrm{s}}\frac{\mathrm{d}T_{\mathrm{rg}}}{\mathrm{d}t}=T_{\mathrm{st}}F_{\mathrm{s}}Cp_{\mathrm{s}}+T_{\mathrm{a}}F_{\mathrm{a}}Cp_{\mathrm{a}}-T_{\mathrm{rg}}(F_{\mathrm{s}}Cp_{\mathrm{s}}+F_{\mathrm{a}}Cp_{\mathrm{a}})-\Delta H_{\mathrm{cb}}R_{\mathrm{cb}}/M_{\mathrm{c}}, \tag{7}$$

where R_{cb} represent the kinetic expression for the coke combustion, which is given by the following equation

$$R_{\mathrm{cb}}=k_{\mathrm{cb}}\exp(-E_{\mathrm{cb}}/RT_{\mathrm{rg}})O_{\mathrm{d}}C_{\mathrm{rc}}W_{\mathrm{rg}}. \tag{8}$$

3. Mass balance of oxygen in the regenerator dense bed (O_{d})

$$W_{\mathrm{a}}\frac{\mathrm{d}O_{\mathrm{d}}}{\mathrm{d}t}=F_{\mathrm{a}}(O_{\mathrm{in}}-O_{\mathrm{d}})/M_{\mathrm{a}}-((1-\sigma)n+2+4\sigma)R_{\mathrm{cb}}/(4(1+\sigma)M_{\mathrm{c}}), \tag{9}$$

where W_{a} is the oxygen holdup in the regenerator, M_{a} is the molecular weight of the oxygen, M_{c} is the molecular weight of the coke and σ is the CO_2/CO ratio.

2.3. Separator model

Hydrocarbon stripping of the catalyst and gases is performed in a cyclone, which is modeled as a mixing tank. The effect of this separator is to produce a lag time between the riser outlet and the catalyst return to the regenerator. The governing equations for the coke (C_{st}) and the temperature (T_{st}) dynamics are given respectively by

$$W_{\mathrm{st}}\frac{\mathrm{d}C_{\mathrm{st}}}{\mathrm{d}t}=F_{\mathrm{s}}(C_{\mathrm{sc}}-C_{\mathrm{st}}), \tag{10}$$

$$W_{\mathrm{st}}Cp_{\mathrm{s}}\frac{\mathrm{d}T_{\mathrm{st}}}{\mathrm{d}t}=F_{\mathrm{s}}Cp_{\mathrm{s}}(T_{\mathrm{ro}}-T_{\mathrm{st}}). \tag{11}$$

The FCC model involves a large number of other variables and parameters that are not directly relevant to the subject matter of the paper. A complete description of the model is presented in Hovd and Skogestad [18] and Balchen et al. [10].

3. Ideal control law design

For partial combustion mode a common choice of regulated variables is the temperature at the riser outlet T_{ro} and the temperature of the regenerator dense bed T_{rg}. Because the product distribution is determined by the reaction conditions inside the riser; there is then a good incentive to control both T_{ro} and T_{rg}. The control of T_{rg} is critical to avoid irreversible deactivation of the catalyst in the coke burning stage and cyclone damage. Another common choice of manipulated variables is the regenerated catalyst flow from the regenerator and the airflow on the regenerator input. If the pairing $T_{\mathrm{ro}}-F_{\mathrm{s}}$, $T_{\mathrm{rg}}-F_{\mathrm{a}}$ is selected to design a decentralized control strategy, the riser–regenerator control structure is obtained [18].

In the present section it is showed how to achieve temperature regulation of the temperatures T_{rg} and T_{ro} using input–output linearizing state feedback [1]. It is assumed that all the parameters and reaction rates are known. In addition, it is assumed that all states are available for on-line measurements. Of course these are not realistic assumption, however they will be used as intermediate assumptions towards the final control laws to be designed in the next section.

3.1. Regulation of T_{rg}

The energy balance equation for the regenerator can be expressed like

$$\frac{\mathrm{d}T_{\mathrm{rg}}}{\mathrm{d}t}=L_{\mathrm{rg}}+D_{\mathrm{rg}}F_{\mathrm{a}}+\eta_{\mathrm{rg1}}, \tag{12}$$

Where L_{rg} includes the linear terms of the Eq. (7), D_{rg} is the coefficient of the control input F_{a} which is assumed to be known and η_{rg1} is the heat generated by the burning of coke in the regenerator dense bed, if linearizing input/output feedback control law [1] is used to regulate the regenerator temperature, the following expression is obtained for the control input

$$F_{\mathrm{a}}=\frac{1}{D_{\mathrm{rg}}}(-L_{\mathrm{rg}}-\eta_{\mathrm{rg1}}-g_{\mathrm{rg}}(T_{\mathrm{rg}}-T_{\mathrm{rg}}^*)), \tag{13}$$

where $g_{\mathrm{rg}}>0$ is the control gain, T_{gr}^* is the desired regenerator temperature (set point). Note that $D_{\mathrm{rg}}\neq 0$ for all the operation conditions. The above feedback linearizes the dynamics of the regenerator temperature dense bed, such that the closed-loop dynamics behaves as an asymptotically stable linear system.

$$\frac{\mathrm{d}T_{\mathrm{rg}}}{\mathrm{d}t}=-g_{\mathrm{rg}}(T_{\mathrm{rg}}-T_{\mathrm{rg}}^*). \tag{14}$$

Notice that $g_{\rm rg}^{-1}$ can be seen as the nominal closed-loop time constant, which can be chosen of the order of the catalyst residence time into the regenerator [18].

3.2. Regulation of T_{ro}

A distributed parameter model (under the plug flow reactor assumption) governs the dynamic behavior of the riser reactor. The Eq. (5) is spatially discretized in order to derive a linearizing feedback controller. A first order discretization of the spatial derivative of the Eq. (5) at the reactor outlet conditions (z=1) provides an approximation to the dynamics of the temperature at the riser reactor outlet

$$\frac{{\rm d}T_{\rm ro}}{{\rm d}t} = \frac{(F_{\rm s} - F_{\rm f})\Xi}{W_{\rm ri}} + \eta_{\rm ro1}, \tag{15}$$

where $\Xi=((T_{\rm ro}-T_{\delta})/\delta)$ and the kinetics related term is giving by

$$\eta_{\rm ro1} = \Delta H_{\rm f} K_1(T_{\rm ro})[{\rm COR}]\Phi y_{\rm f}^2/\rho_{\rm ri} Cp_{\rm ri}.$$

The above term is assumed to be unknown and correspond to the input/output modeling error, the parameter δ is a small distance from the riser reactor outlet and T_δ is the riser temperature at z=1−δ. If $T_{\rm ro}^*$ is the desired temperature at the riser outlet and considering that one wish that the regulation error $e_{\rm ro} = T_{\rm ro} - T_{\rm ro}^*$ has a closed-loop exponentially stable dynamical behavior, the control input $F_{\rm s}$ is calculated as follows:

$$F_{\rm s} = -F_{\rm f} + W_{\rm ri}\frac{(-\eta_{\rm ro1} - g_{\rm ri}(T_{\rm ro} - T_{\rm ro}^*))}{\Xi}. \tag{16}$$

Under the state feedback Eq. (16), the temperature $T_{\rm ro}$ converges asymptotically to the desired temperature value $T_{\rm ro}^*$ with closed-loop dynamics given by

$$\frac{{\rm d}T_{\rm ro}}{{\rm d}t} = -g_{\rm ri}(T_{\rm ro} - T_{\rm ro}^*). \tag{17}$$

As in the case of regulation of $T_{\rm rg}$, $g_{\rm ri}^{-1}$ can be seen as the nominal closed-loop time constant. Since the dynamics of the regenerator are slower than the dynamics of the riser, the overall dynamics of the FCC units are slower than the dynamics of the riser. Hence, to avoid exciting unmodeled dynamics it is suggested to choose $g_{\rm ri}$ of the order of $g_{\rm rg}$ [18].

4. Temperature regulation using uncertainty estimation

In the above section a perfect knowledge of reaction rates (the modeling errors) for the control design was assumed. In practice this is not a realistic assumption, particularly for the FCC process where the conversion of gasoil to lighter compounds and the burning of coke for catalyst regeneration are realized spell through a complex network of reactions [19], which are poorly known. However, for temperature regulation, the exact knowledge of the reaction rates functionalities is not necessary. Instead, a control strategy must to have access to the instantaneous input flow and output flow enthalpies and to the instantaneous heat production rates due to reaction activities into the reactor. In most cases, it is suitable to assume that the convective transport of energy is well known, so that in order to have a good performance in temperature regulation, a problem of estimating heat production rates must be confronted.

4.1. Uncertainty estimation with proportional-integral filtering

The methodology proposed to estimate heat generation rates by a proportional-integral filtering, is an extension of the work presented by Aguilar et al. [20], where a control scheme based on a proportional observer and an open-loop estimation methodology based on a proportional-integral Luenberger observer is developed. In this work, it is considered that the uncertainty terms can be expressed as new states (whose dynamics are unknown) to construct a new system with dimension n+1. However, the methodologies mentioned above made use of a full-order observer. In the process under study this is not necessary, because some of the state variables (the regenerator and riser outlet temperatures) are directly the system outputs, such that a reduced-order observer to estimate the unknown terms related with the heat generations in the riser and the regenerator are proposed.

4.1.1. Regulation of T_{rg}

The energy balance equation for the regenerator reactor can be expressed as

$$\frac{{\rm d}T_{\rm rg}}{{\rm d}t} = L_{\rm rg} + D_{\rm rg}F_{\rm a} + \eta_{\rm rg1}, \tag{18}$$

where $\eta_{{\rm rg1}_1}$ is the uncertainty term (modeling error) related with the heat generation rate and its dynamic behavior is given by

$$\frac{{\rm d}\eta_{\rm rg1}}{{\rm d}t} = \Omega(T_{\rm rg}, O_{\rm d}, C_{\rm rc}), \tag{19}$$

where $\Omega(T_{\rm rg},O_{\rm d},C_{\rm rc})$ is a unknown function depending on the states of the system. By taking the regenerator reactor temperature as the measured output, the subsystem given by the Eqs. (18) and (19) can be written in an observability form [21], so that it satisfies the uniform observability condition. Hence, $\eta_{\rm rg1}$ can be reconstructed by means of a state observer. Such observer could be constructed as a copy of the subsystem (18)–(19) corrected by an observation error. However, this typical observer structure can not be realized because the term $\Omega(T_{\rm rg},O_{\rm d},C_{\rm rc})$ is unknown. Then in order to estimate the modeling error $\eta_{\rm rg1}$, a proportional-integral reduced-order filter is proposed. By considering that the variable state given for the Eq. (18) is known.

we propose the following filter

$$\frac{d\hat{\eta}_{rg1}}{dt} = \hat{\eta}_{rg2} + 2\tau_{rg}^{-1}(\eta_{rg1,m} - \hat{\eta}_{rg1}),$$
$$\frac{d\hat{\eta}_{rg2}}{dt} = \tau_{rg}^{-2}(\eta_{rg1,m} - \hat{\eta}_{rg1}), \tag{20}$$

where τ_{rg} is the estimation characteristic time and $\eta_{rg1,m}$ is the "measured" output, which can be solved from the Eq. (18), to produce the following equations

$$\frac{d\hat{\eta}_{rg1}}{dt} = \hat{\eta}_{rg2} + 2\tau_{rg}^{-1}\left(\frac{dT_{rg}}{dt} - L_{rg} - D_{rg}F_a - \hat{\eta}_{rg1}\right),$$
$$\frac{d\hat{\eta}_{rg2}}{dt} = \tau_{rg}^{-2}\left(\frac{dT_{rg}}{dt} - L_{rg} - D_{rg}F_a - \hat{\eta}_{rg1}\right). \tag{21}$$

If the temperature measurements are noisy, the evaluation of the derivative of dT_{rg}/dt in the above equation is difficult or even impossible. Considering that the regulation error is given by $e_{rg} = (T_{rg} - T^*_{rg})$, to avoid the evaluation of the derivative of dT_{rg}/dt, the following variables are introduced

$$\theta_{rg,1} = \hat{\eta}_{rg1} - 2\tau_{rg}^{-1}e_{rg}, \quad \theta_{rg,2} = \hat{\eta}_{rg2} - \tau_{rg}^{-2}e_{rg} \tag{22}$$

which lead to the following expressions to obtain the estimated modeling error

$$\frac{d\dot{\theta}_{rg,1}}{dt} = \theta_{rg,2} + 2\tau_{rg}^{-1}e_{rg} + 2\tau_{rg}^{-1}(g_{rg}e_{rg}),$$
$$\frac{d\dot{\theta}_{rg,2}}{dt} = \tau_{rg}^{-2}(g_{rg}e_{rg}), \quad \hat{\eta}_{rg1} = \theta_{rg1} + 2\tau_{rg}^{-1}e_{rg}. \tag{23}$$

Note that all the terms are assumed to be known.

From the Eqs. (22) and (23), it is not hard to see that the estimated modeling error can be expressed as

$$\hat{\eta}_{rg1} = 2\tau_{rg}^{-1}e - 2\tau_{rg}^{-1}g_{rg}\int_0^t e_{rg}(\tau_1)\,d\tau_1 - \tau_{rg}^{-2}g_{rg}\int_0^t\left[\int_0^{\tau_1} e_{rg}(\tau_2)\,d\tau_2\right]d\tau_1. \tag{24}$$

The idea behind the observer construction is the following. The modeling error η_{rg1} is of feedback nature because its dependence on the system states and control inputs. It is expected that the dynamics of η_{rg1} be composed of low and high frequency modes. If one is able to estimate the η_{rg1} dynamics in a wide range of frequencies, and use this information in a feedback loop, then better closed-loop performance can be obtained.

A second-order (nominal) model for the modeling error, would be $(d^2\eta_{rg1}/dt^2 = \Gamma(t)$ where $\Gamma(t)$ is a unknown function. In this way, the observer Eq. (20) provides an estimate of $\eta_{rg1}(t)$ and its first time derivative η_{gr2}, which can be seen as a predictor of high frequency uncertainty dynamics.

With the estimate of η_{rg1}, the practical control law is expressed by the following equation

$$F_a = \frac{1}{D_{rg}}(-L_{rg} - \hat{\eta}_{rg1} - g_{rg}e_{rg}). \tag{25}$$

Now, introducing the definition for the uncertain term given by the Eq. (24), the control law takes the following structure

$$F_a = \frac{1}{D_{rg}}\left[(g_{rg} + 2\tau_{rg}^{-1})e_{rg} + 2g_{rg}\tau_{rg}^{-1}\int_0^t e_{rg}(\tau_1)\,d\tau_1 + g_{rg}\tau_{rg}^{-2}\int_0^t\left[\int_0^{\tau_1} e_{rg}(\tau_2)\,d\tau_2\right]d\tau_1 - L_{rg}\right]. \tag{26}$$

This control structure corresponds with a PI^2 controller [8], where the control gains are parameterized in terms of the closed-loop characteristic time (the inverse of g_{rg}) and the estimation characteristic time τ_{rg}, which are parameters with a strong physical meaning. On the other hand, with this parameterization, the identification step for the tuning of the control gains is avoided. In fact, concerning the notation in Eq. (26), the corresponding gains of the PI^2 controller are given by

$$K_{p\text{-}rg} = \frac{1}{D_{rg}}(g_{rg} + 2\tau_{rg}^{-1}),$$
$$\tau_{I\text{-}rg} = \frac{1}{D_{rg}}(2g_{rg}\tau_{rg}^{-1} + 2\tau_{rg}^{-2}),$$
$$\tau_{II\text{-}rg} = \frac{1}{D_{rg}}g_{rg}\tau_{rg}^{-2}. \tag{27}$$

where $K_{p\text{-}rg}$ is the proportional gain, $\tau_{I\text{-}rg}$ is the integral gain and $\tau_{II\text{-}rg}$ is the double-integral gain.

4.1.2. Regulation of T_{ro}

The methodology used above to the control of T_{rg} can be extended to the case of regulation of T_{ro}. The structure of the uncertainty estimator for the unknown heat consumption rate is given by the following equations

$$\frac{d\hat{\eta}_{ro1}}{dt} = \hat{\eta}_{ro2} + \tau_{ro}^{-1}(\eta_{ro1,m} - \hat{\eta}_{ro1}),$$
$$\frac{d\hat{\eta}_{ro2}}{dt} = \tau_{ro}^{-2}(\eta_{ro1,m} - \hat{\eta}_{ro1}), \tag{28}$$

where the "measured" output η_{ro1}, can be obtained from the Eq. (15) and introduced in the Eq. (28). Again, if the measured temperature at the riser output T_{ro} is noisy, the evaluation of the temperature derivative is difficult, in order to avoid this problem the following variables are defined

$$\theta_{ro,1} = \hat{\eta}_{ro1} - 2\tau_{ro}^{-1}e_{ro}, \quad \theta_{ro,2} = \hat{\eta}_{ro2} - \tau_{ro}^{-2}e_{ro}. \tag{29}$$

These variables, produce an input/output modeling error observer with a similar structure to the Eq. (23).

The corresponding practical control law can be expressed by

$$F_s = -F_f - \frac{W_{ri}}{\Xi}\left[(\tau_{ro}^{-1} + g_{ri})e_{ro} + (\tau_{ro}^{-2} + \tau_{ro}^{-1}g_{ri}) \int_0^t e_{ro}(\tau_1)\,d\tau_1 + \tau_{ro}^{-2}g_{ri}\int_0^t\left(\int_0^{\tau_1} e_{ro}(\tau_2)\,d\tau_2\right)\right]d\tau_1. \quad (30)$$

Where the corresponding gains for the riser are given by the following equations

$$K_{p\text{-}ri} = (\tau_{ro}^{-1} + g_{ri})W_{ri}/\Xi,$$
$$\tau_{I\text{-}ri}(\tau_{ro}^{-2} + \tau_{ro}^{-1}g_{ri})W_{ri}/\Xi,$$
$$\tau_{II\text{-}ri} = \tau_{ro}^{-2}g_{ri}W_{ri}/\Xi.$$

4.1.3. Comments

1. The idea of using a proportional-integral filter to estimate uncertainties is to reduce the effects of the persistent disturbances acting in the process [8]. In this way the proportional-integral filters generates estimates of the unknown reaction rates for both reactors in the FCC unit, which are subsequently used in the corresponding feedback controllers to yield a closed-loop behavior that is close to the ideal one.
2. For the system given by the Eqs. (20) and (28), the convergence analysis can be made using Laplace domain. Analyzing the structure of the above equations, it can be observed that they are linear ones, and can be represented by means of the following equation in the Laplace domain, where, in general, for both regenerator and riser, N is the uncertainty term (input/output modeling error) in the Laplace domain and τ is the estimation characteristic time

$$s\hat{N} = 2\tau^{-1}(N - \hat{N}) + \frac{\tau^{-2}}{s}(N - \hat{N}). \quad (31)$$

After some algebraic manipulations, the above equation can be represented by

$$\frac{\hat{N}}{N} = \frac{2\tau^{-1}s - \tau^{-2}}{s^2 + 2\tau^{-1}s + \tau^{-2}}. \quad (32)$$

which is a stable and minimum-phase transfer function. In the limit, when $s\to 0$ ($t\to\infty$) we have that $(N/\hat{N}) = 1$, this proves that the estimated uncertainties converge to the real value of the uncertain term. Hence, the practical control laws, Eqs. (26) and (30) recover the structure of the ideal control laws Eqs. (13) and (16). It should be stressed that this is only a robustness result and is far from realistic since small values of the estimation time constant render the closed-loop system highly sensitive to practically unavailable measurement noise and unmodeled (high-frequency) dynamics.
3. The practical control laws Eqs. (26) and (30) are second-order compensators. Roughly speaking, these control laws are output feedback compensators composed by a feedback functions and an uncertainty estimators. Their implementation requires available measurements (temperatures, some physical properties and flow rates) and estimated physical parameters (steady-state gains).

5. Numerical simulations

Numerical simulations were performed to show how the controllers Eqs. (26)–(30) work. In the simulation experiments to be discussed next, the following sequence of step disturbances acting on the regenerator–riser unit are assumed: A 5 K step increase in air temperature T_a occurs at 30 min, a 5 K decrease in the gasoil feed temperature T_f occurs at 180 min, a 2.5% increase in the combustion coke rate K_c occurs at 270 min and finally a 4 kg/min decrease in the gasoil feed flow F_f occurs at 390 min. This class of step disturbances is expected to be present in an industrial scale process [11]. Besides, several changes in the set points occurs for both reactors; the original set point for the regenerator temperature changes from 965 to 975 K at 85 min and from 975 to 960 K at 250 min, both in a step way. For the riser outlet temperature the set point changes from 765 to 770 K at 165 min and from 770 to 760 K at 333 min.

In the section it is showed via numerical simulations that the proposed control scheme yields closed-loop stability despite a certain class of unmodeled dynamics. This situations arises when the dynamics of the actuators are not considered in the control design stage. In general, the dynamics of actuators (valves) are faster than process dynamics. Assume that the control input for the regenerator is subjected to (non-modeled) actuator dynamics. That is,

$$\tau_r\frac{dF_{ar}}{dt} + F_{ar} = F_a, \quad (33)$$

where F_{ar} and F_a are respectively the real and the computed control input for the regenerator and τ_r is the time constant of the actuator. Notice that, if τ_r=0 (actuators with very fast dynamics) then F_{ar}=F_a and the closed-loop stability is assured. For the sake of numerical simulation the value of τ_r=1.5 min with the same control parameters as in the regulation case are taken.

Numerical simulation were made to illustrate the behavior of a standard PI controllers in order to compare its performance with the PI^2 controllers. The PI controllers were tuned by means of a classical identification methodology based on a step perturbation in both regenerator and riser reactors employing the Ziegler–Nichols method. The Figs. 1 and 3 show the closed-loop behavior of the temperatures in both reactors where standard PI controllers were used. It can be seen that the performance is very poor

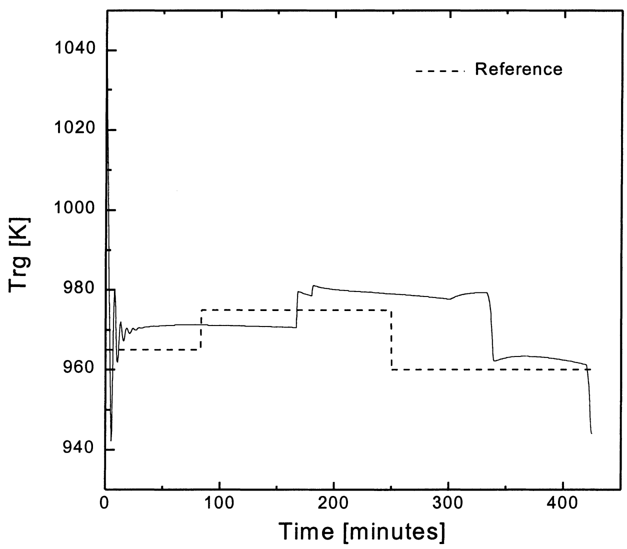

Fig. 1. Closed-loop behavior of T_{rg} under PI controller.

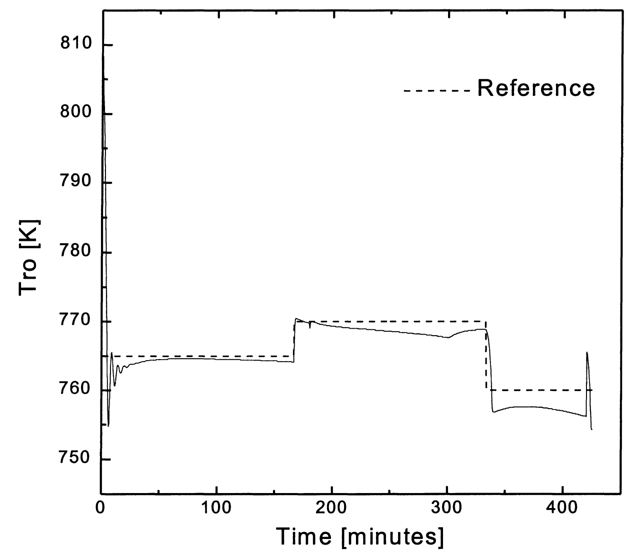

Fig. 3. Closed-loop behavior of T_{ri} under PI controller.

at short times, the temperatures present a large oscillatory behavior with large frequency and amplitude. Moreover, both temperatures show offset from the set point, which is due to the saturation of the catalyst flow rate. Notice that the control input for the regenerator show large changes which induce large perturbations in the catalyst temperature at the riser inlet. Since FCC units are recycling processes, these perturbations are feedback into the regenerator, thus leading to an overall instability and persistent control input saturation. These behaviors are undesirable from the operation point of view, because the compressor that provide the air flow is operating under demanding conditions. Regarding the riser, a highly oscillatory behavior of the control input would induce unfeasible flow rates of regenerated catalyst Figs. 2 and 4.

Figs. 5 and 7 show the simulation results when the temperatures T_{rg} and T_{ro} are regulated, by the actions of the feedback controllers Eqs. (26) and (30). During the first minutes of operation, the FCC unit presents a dynamical behavior caused by the startup of the process, and the transient generated by the initialization of the estimation procedure. When the disturbances enters the system, the regenerator–riser temperatures are moved off the set point. The disturbances above mentioned are detected by the estimation algorithm and this information is used by the control laws to increase or diminish the air or catalyst flows respectably to offset this changes and keep the temperatures in the desired values. For instance, the 5 K step increase in the air temperature T_a (occurring at 30 min) is detected by the filter Eq. (23) as an excess of energy entering the

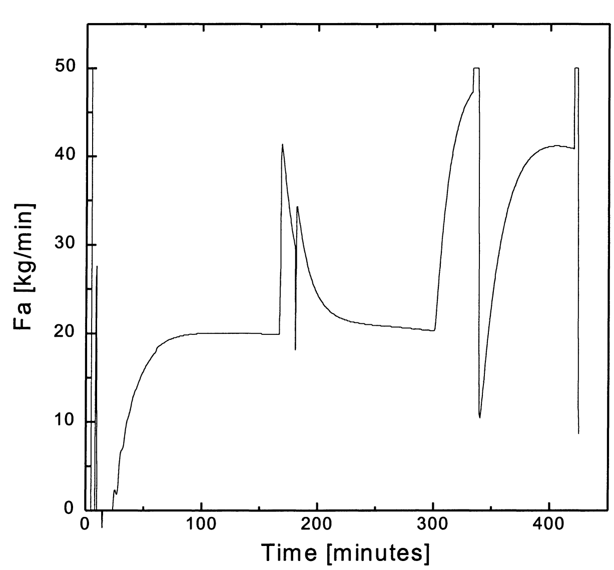

Fig. 2. Dynamic behavior of the regenerator input F_a under PI controller.

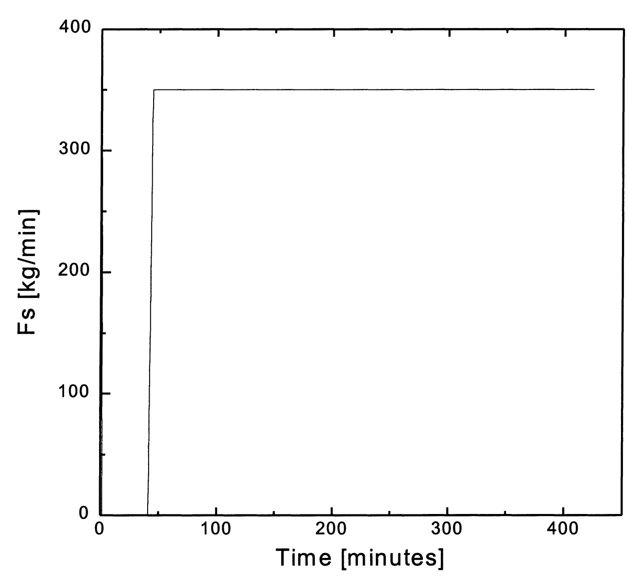

Fig. 4. Dynamic behavior of the riser input F_s under PI controller.

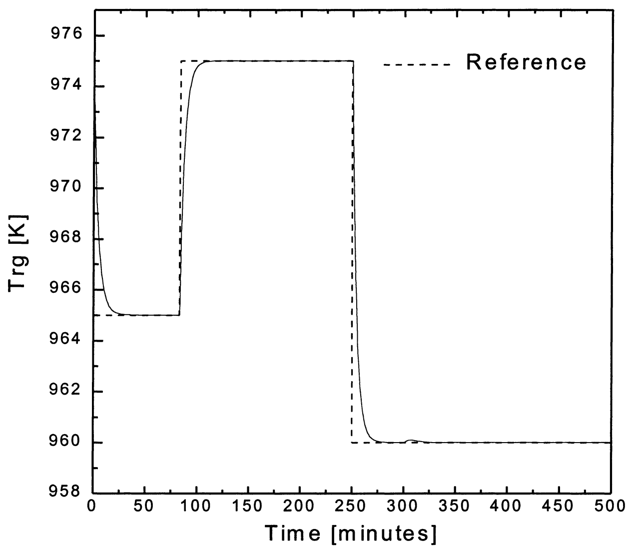

Fig. 5. Closed-loop behavior of T_{rg} under PI^2 controller.

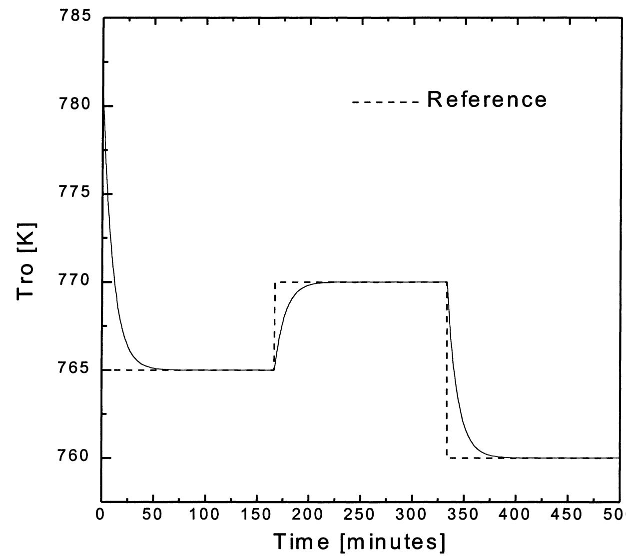

Fig. 7. Closed-loop behavior of T_{ri} under PI^2 controller

regenerator. To counteract the effect of this perturbation, the controller Eq. (26) augments the flow rate of air, so more heat is extracted from the regenerator via output convective flow. Now, when a 2.5% increase in combustion rate of coke K_c occurs at 270 min, more coke is produced, which increases the heat production in the regenerator and the gasoline production decreases due to an additional catalyst deactivation. To spell the effect of this perturbation, the controllers Eqs. (26)–(30) increase the flow rate of regenerated catalyst and air. The effects of this disturbances in the control variables F_s and F_a are showed in Figs. 6 and 8. When a step change in the set point occurs in both reactors the main effect consists in an excess or reduction of energy in the process. These changes are detected by the estimation algorithm and the information generated is used for the controllers, which counteracts the adverse effects by increase or reducing the flow rate of air and regenerated catalyst in order to keep the temperatures in the desired values.

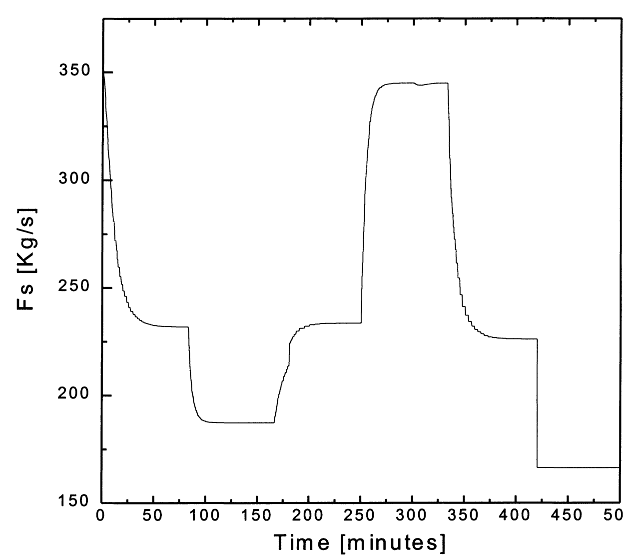

Fig. 8. Dynamic behavior of the riser input F_s under PI^2 controller

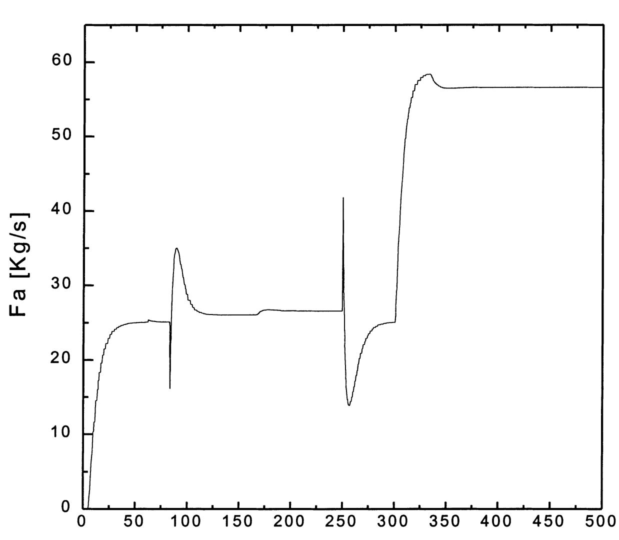

Fig. 6. Dynamic behavior of the regenerator input F_a under PI^2 controller.

6. Conclusions

In this work a model-based control strategy for the temperature regulation in a regenerator–reactor FCC unit was proposed. The unreliable modeling terms related with the kinetics of the process are considered unknown, and they are estimated by means of a proportional-integral reduced order filter. This estimation procedure allows us to synthesize adaptive input–output linearizing controllers, which become equivalent to a PI^2 controller, which is robust against input/output modeling errors and set point changes.

The resulting controllers are similar in form to standard controllers and can be tuned easily by choosing the closed-loop and the estimation time constants.

7. Nomenclature

F_{ar}	delayed control action
F_a	flow rate of air to the regenerator (25.0 kg/s)
F_s	flow rate of regenerated catalyst (294.0 kg/s)
g_{rg}	regenerator control gain (1/s)
g_{ro}	riser control gain (1/s)
T_{ro}	temperature at riser outlet K (controlled variable)
T_{rg}	temperature in the separator K (controlled variable)
T^*_{ro}	riser set point temperature K
N	input/output modeling errors in Laplace domain

Greek letters

τ_c	characteristic time in the riser (1.5 min)
τ_{rg}	parameter of the observer in the regenerator
τ_{ro}	parameter of the observer in the riser
τ_r	actuator time constant

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