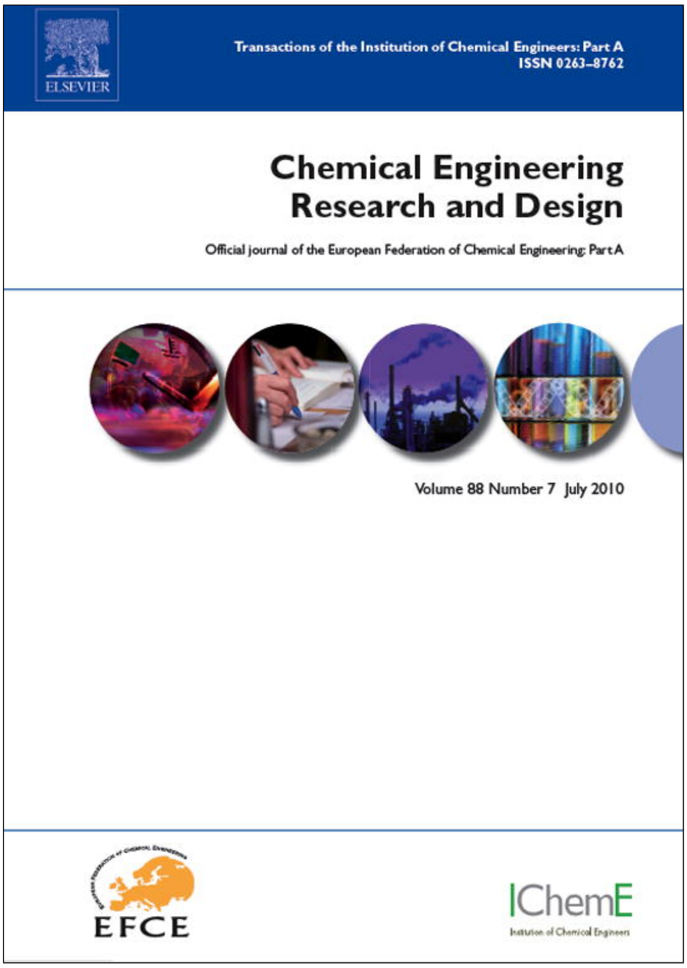
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Design of Controllers for a Fluidized Catalytic Cracking Process

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ABSTRACT

Fluidized-bed catalytic cracking (FCC) units, are one of the most complex and interactive processes in the refining industry, hence they are difficult to operate and to control; on the other hand they are one of the main producers of gasoline, which is an incentive to improve operation policies. Control of the FCC has been and continues to be a challenging and important problem. A control system is composed of several interacting control loops. The number of feasible alternative configurations of control loops is very large. Multiloop control systems are often used for industrial multivariable processes because of their simplicity. Relative gain array is used to minimize the interaction and to select the variable pairings. This work is concerned with the design of multivariable feedback control configurations for FCC units. Sufficient conditions to achieve regulation in terms of the steady-state gain matrix are provided, allowing to obtain a systematic procedure for analyzing multiloop control configurations of complex and interacting processes. Numerical simulations on a dynamical model based on a FCC unit operating in the partial-combustion mode are used to show the effectiveness of several control configurations under disturbances.

Keywords: Conventional control; Multiloop control; Fluid catalytic cracking; Internal model control; Relative gain array; Decoupling

1. Introduction

Catalytic cracking has been one of the key processes in petroleum refining in the last decades. The fluid catalytic cracking (FCC) process is an important process in refiners for upgrading heavy hydrocarbons to more valuable lighter products mainly gasoline; hence small enhancements in their operation are economically attractive. In order to achieve this objective, it is necessary to implement advanced process control techniques. In a refinery that produces a substantial amount of gasoline, FCC gasoline makes up about 40% of the overall refinery gasoline pool. The remaining gasoline is typically derived from straight run naphtha, coking, hydro cracking, and other molecular conversion units such as alkylation and reforming operations. Due to its large throughput and the high product feed upgrade, the overall economic benefits of a refinery could be considerably increased if proper control and optimization strategies are implemented on FCC units.

Analysis and control of FCC processes have been known as challenging problems due to the following process characteristics: (i) very complicated and little known hydro dynamics, (ii) complex kinetics of both cracking and coke burning reactions, (iii) strong interaction between the reactor and regenerator, and (iv) many operating constraints.

Control of the FCC has been and continues to be a challenging and important problem. As will be seen, its steady-state behavior is highly non-linear, leading to multiple steady states, input multiplicities, and all that this implies. In earlier years before the development of zeolite catalysts, the major control problem was one of stabilization, of just keeping the unit running. Later with zeolite catalysts the emphasis shifted to increasing production rates in the face of unit constraints and to handling heavier feeds. The requirements for reformulated gasoline have added the need to control product composition. This is a more complex problem since the number of process variables that one would like to control

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Nomenclature

Parameters

F	feed
G	gasoline
T_1	temperature at riser exit
T_a	temperature of air to the regenerator
T_{cy}	regenerator cyclone temperature
T_{st}	temperature in reactor
T_{rg}	regenerator bed temperature
F_s	catalyst recirculation rate
F_a	the regenerator air rate
T_{ro}	temperature at riser end
$K_{p11}, K_{p12}, K_{p21}, K_{p22}$	relative gains of the multiloop controllers
K_p	steady-state gain matrix
D_1, D_2	decoupling controllers for loop 1, loop 2 respectively.
$G_p(s)$	transfer function of the process
$G_c(s)$	transfer function of the controller
$G_m(s)$	transfer function of the model
$G_m^-(s)$	invertible part of the model
$G_m^+(s)$	non-invertible part of the model
$G_f(s)$	transfer function of the feedback element

Greek letters

λ	positive tuning parameter
τ	time delay
n	order of the filter

substantially exceeds the number of manipulated variables that are available for the task. One is faced with a juggling act of sorts in which one tries to maintain all the variables within acceptable bounds by controlling a subset of dominant variables.

1.1. *History of FCC modeling: a short literature survey*

Numerous papers have been published concerning different modeling approaches and control strategies for the FCC process, which deal with the strong interactions and many constraints from the operating, security and environmental point of view. The potential of yielding more market-oriented oil products, increasing production rate and stabilizing the operation become the major incentives to search for more accurate and practical models, high performance, and cost effective and flexible control strategies.

Several studies on the dynamic modeling of the whole FCC unit have been presented in recent papers. Fundamental modeling work on FCCUs has been reported by Lee and Groves (1985), Felipe (1992), and McFarlane et al. (1993). All the models except the one by McFarlane et al. assume that the FCCU is equipped with slide valves to control catalyst circulation rate. Einashaie and Elshishini (1993) extended their steady-state model to a simple dynamic model, and investigated the sensitivity and stability of a bed-cracking type FCC unit. Arbel et al. (1995) developed a model that can describe both the steady-state and dynamic behavior of an FCC unit being operated in either the partial or full combustion modes. Ansari and Tade (2000) proposed a multivariable control to the reactor-regenerator. Three of the most commonly used models in the literature which incorporate the features of earlier models are the one by Lee and Groves (1985), Avidan and Shinnar (1990), McFarlane et al. (1993) and Arbel et al. (1995).

Several works have been conducted in order to model the regeneration process. Han and Chung (2001) developed a dynamic model of a FCC in which the regenerator was divided in two regions. Ali et al. (1997) published a dynamic model of a FCC converter and later, it was modified by Malay et al. (1999).

Alvarez-Ramirez et al. (2004) proposed a linear cascade (master/slave) control configuration which leads to asymptotic regulation of the riser output composition (e.g. gasoline yield) about a feasible set point. Sufficient conditions to achieve regulation in terms of the steady-state gain matrix and response time-constants are provided, allowing to obtain a systematic procedure for analyzing multivariable control configurations of complex and interacting processes. Some tuning guidelines issues are discussed.

2. The FCC process

Several authors have made substantial efforts to model the behavior of FCC units. A detailed review of recent work on FCC modeling can be found in a paper by Arbel et al. (1995), Bristol (1969). Since FCC feedstocks consist of thousands of components, the estimation of intrinsic kinetic constants is very difficult; thus, the lumping of components according to the boiling range is generally accepted. Contributions to the modeling of FCC units vary from regenerator models over kinetic models for the reactions taking place in the reactor riser (Ansari and Tade, 2000; Arbel et al., 1995). The model used for this case study is one developed by Lee and Groves (1985) with slight modifications introduced by Hovd and Skogestad (1993). It is based on the three-lump reactor model, which comprises the main components in a FCC unit. The cracking is then described by the following three reactions:

$$\text{feed}\quad (F)^- \rightarrow 1 \quad \text{gasoline}\quad (G)$$

$$\text{gasoline}\quad (G)^- \rightarrow 2 \quad \text{coke} + \text{gas}\quad (L)$$

$$\text{feed}\quad (F)^- \rightarrow 3 \quad \text{coke} + \text{gas}\quad (L)$$

In general, FCC processes are highly reactive in the sense that almost every molecule in the feed undergoes some change, but overall conversion as used here is typically 30–40 wt.%. The gasoline yield increases with conversion up to a maximum and then decreases as the second reaction predominates, and gasoline cracks to lighter products and to coke. A FCC model with capability to describe the main dynamical aspects (e.g. interactions, convergence rates, etc.) for a feedback control study can be found in Hovd and Skogestad's paper. Control design results and numerical simulation tests described in subsequent sections will be based on such a nonlinear FCC model.

2.1. *FCC unit description*

A schematic overview of the FCC process is shown in Fig. 1. There are two main stages in the process: the cracking where pertinent reactions take place, and the regeneration where the catalyst is regenerated by burning off the coke deposited on the bed. Feed oil is contacted with hot catalyst at the bottom of the riser, causing the feed to vaporize. The cracking reactions occur while the oil vapor and catalyst flow up the riser. The residence time of the catalyst and hydrocarbon vapors in the

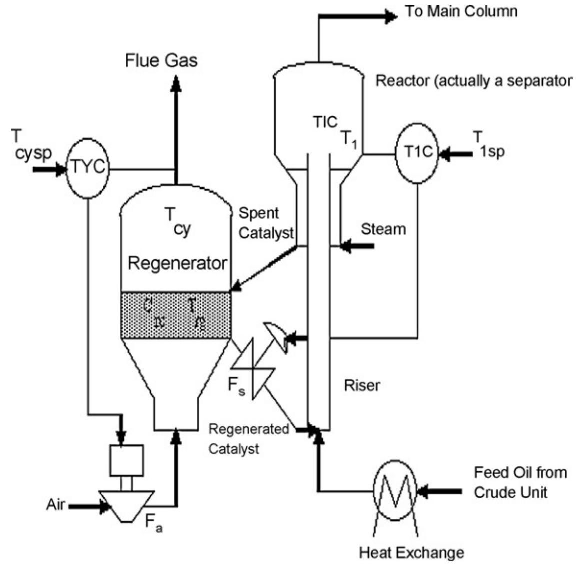

Fig. 1 – FCC unit.

riser is typically in the range 5–8 s. The riser top temperature is typically between 750 and 820 K and is usually controlled by regulating the flow of hot regenerated catalyst to the riser. As a by-product of the cracking reactions, coke is formed and deposited on the catalyst, thereby reducing catalyst activity. The catalyst and products are separated in the reactor. Steam is supplied to the stripper in order to remove volatile hydrocarbons from the catalyst. In the regenerator, which is operated in the fluidization regime, the coke is burnt off the catalyst surface by the air blown into the bed. This combustion reaction serves to reactivate the catalyst activity and to maintain the bed temperature (950–980 K for a gas oil cracker and 980–1080 K for a resid cracker) high enough to supply the heat required for the vaporization and cracking reactions of the feed in the reactor.

Depending on the coke producing tendency of the feed, the FCC process can be operated in two distinct modes: partial-combustion and the complete combustion modes. In the partial-combustion mode the conversion of coke to CO_2 is not complete, which means that relatively large amounts of both CO are formed (this CO-rich regenerator flue gas can be sent to a CO boiler for further combustion to produce high pressure steam). It is not always possible operate a FCC unit in the complete combustion mode, specially if the feed has a large coke production tendency and there exists also an economic incentive operating in the partial-combustion mode, as the heat recovered in the CO boiler is valuable.

2.2. *Constraints*

The optimal operating point for an FCC lies at one or several constraints. The control structure which allows operation closest to the constraints is preferable.

Common constraints include:

- Maximum regenerator cyclone temperature constraint.
- Maximum wet gas compressor capacity.
- Maximum air blower capacity.

The important measured variables are chosen to be the reactor temperature/riser outlet temperature (T_1), the regenerator gas temperature (T_{cy}) and the regenerator bed temperature (T_{rg}). The manipulated variables are the catalyst recirculation rate (F_s) and the regenerator air rate (F_a).

2.3. *Mathematical model of the FCC units*

The modeling of complex chemical systems for the simulation of process dynamics and control has been motivated by the important economic incentives for improvement of plant operation and plant design, as in the case of FCCU. Here one of the major economic incentives for process control results from moving the steady state operation to a better operating point. This demand gives the complexity of the FCCU, a profound knowledge of the process itself and their dominant dynamics, which could be incorporated into a model. Most studies concerning FCC units have been dealt with the process control based on simplified reactor models, which in principle incorporate major observed dynamics. On the other hand very little is known about existing proprietary models. Typically the regenerator which dominates the process dynamics has been modeled as a continuous stirred tank reactor. Other studies have incorporated the two phase nature of the dense bed in the regenerator measurements in industrial units have shown non-isothermal operation of the dense bed with temperature gradients up to 15 °C. This paper is concerned with the model given in Hovd and Skogestad (1993).

3. Multiloop control design

The research on the dynamic characteristics of FCC units reveals that FCC processes consist of a MIMO system with two inputs and two outputs. The manipulated variables are the catalyst circulation rate F_s and the air flow to the regenerator F_a. The control variables are the reactor temperature and regenerator cyclone temperature.

The simulink representation of FCC is shown in Fig. 2.

In the process industries, where there is a higher degree of uncertainty about process behavior, for control systems design purposes the input–output (transfer function) model approach is generally adequate. Furthermore, there is a correspondence between state-space models and their input–output counterparts. This section therefore considers only structures of input–output models of multivariable processes used in control systems design and analysis. The open-loop performance of a FCC unit is shown in Fig. 3. Dynamic simulation of the FCC process was performed according to the parameters and model given in Hovd and Skogestad (1993). Each simulation run started from the steady state corresponding to the base case operating conditions and the subsequent transient response was obtained as each simulation variable went through a series of step changes.

The transfer function of the dynamic of the FCC unit was determined from the reaction curve of the process obtained

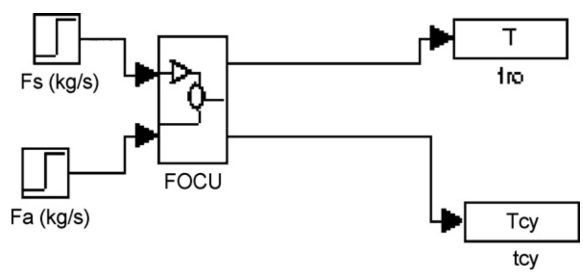

Fig. 2 – Simulink of FCC.

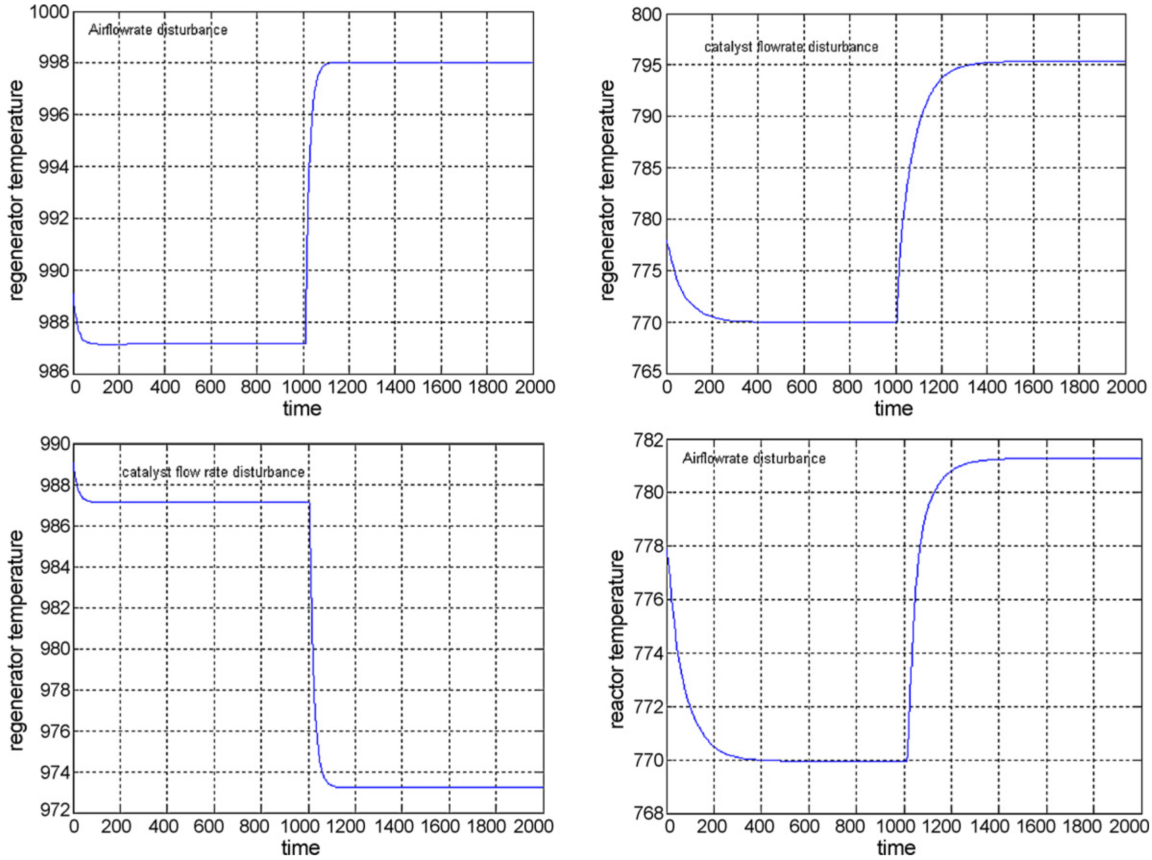

Fig. 3 – Response of FCC for control design.

by giving step disturbances to catalyst circulation rate F_s and the air flow to the regenerator F_a. From the step responses, we can see that the dynamics of both riser and regenerator units are dominated by a first order response. Then we have

$$\begin{bmatrix} T_{ro} \\ T_{cy} \end{bmatrix} = \begin{bmatrix} \dfrac{K_{p11}}{1+s\tau_1} & \dfrac{K_{p12}}{1+s\tau_2} \\ \dfrac{K_{p21}}{1+s\tau_3} & \dfrac{K_{p22}}{1+s\tau_4} \end{bmatrix} \begin{bmatrix} F_s \\ F_a \end{bmatrix}$$

The elements within the blocks in the transfer function matrix define the relationship between the respective input–output pairs. From the step responses the steady-state gain matrix K_p of the FCC multivariable system is then given as

$$K_p = \begin{bmatrix} 6.1000 & 3.3880 \\ -15.9000 & 8.9250 \end{bmatrix}$$

The response of multiloop controller is given in Fig. 4.

4. RGA

One of the most important factors, common to all process control applications, is the correct (best) pairing of the manipulated and controlled variables. A number of quantitative techniques are available to assist in the selection process. One of the earliest methods proposed was the relative gain array (RGA) (Bristol, 1969). The original technique is based upon the open loop steady-state gains of the process and is relatively simple to interpret.

The relative gain array matrix is then calculated using the procedure given by Skogestadt and Morari (2009). The RGA matrix is obtained as shown in below

$$\text{RGA} = \begin{bmatrix} 0.5026 & 0.4974 \\ 0.4974 & 0.5026 \end{bmatrix}$$

From this RGA matrix, we can conclude that our pairing of manipulated variables with control variables (i.e., F_s is used to control reactor temperature and F_a is used to control regenerator cyclone temperature) is best. But there is some interaction between the two loops. To remove interaction completely we go for decoupler design.

5. Decoupling control systems

Basic idea is to use additional controllers to compensate for process interactions and thus reduce control loop interact ions. Ideally, decoupling control allows set point changes to affect only the desired controlled variables. Typically, decoupling controllers are designed using a simple process model (e.g. steady-state model or transfer function model). Naturally, the main goal of decoupling is to make the design of diagonal proportional-integral (PI) (or similar) control of multiple-input multiple output systems possible by eliminating interactions; i.e., the closed-loop response of each loop is the same as it would be if the other loops were on manual control.

The dynamic elements of decoupler are designed as given below

$$D_1 = \frac{465.1s + 15.9}{120.6s + 8.925}$$

and

$$D_2 = \frac{-92.32s - 3.388}{275.85s + 6.1}$$

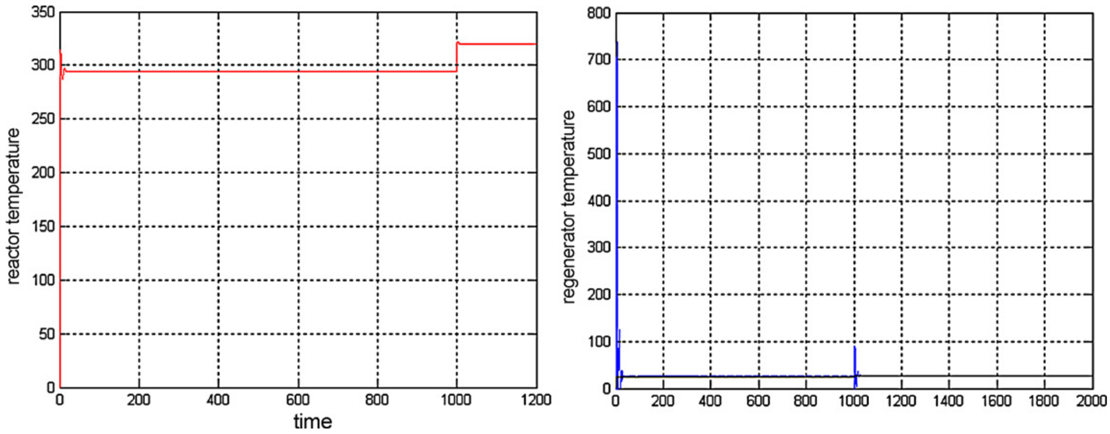

Fig. 4 – Multiloop response.

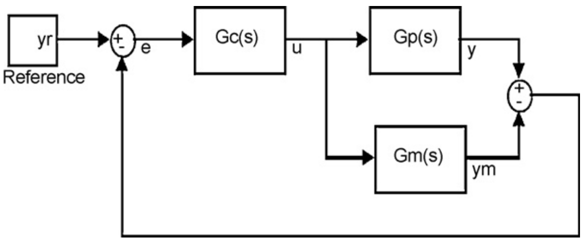

Fig. 5 – Internal model control.

6. Internal model control

A more comprehensive model-based design method, internal model control (IMC), was developed by Garcia and Morari (1982) and Rivera et al. (1986). The IMC method, like the DS method, is based on an assumed process model and leads to analytical expressions for the controller settings. These two design methods are closely related and produce identical controllers if the design parameters are specified in a consistent manner. The IMC method is based on the simplified block diagram shown in Fig. 5.

The IMC method is derived as follows. The process to be controlled (here is FCC) is denoted by $G_p(s)$ and a model of it $G_m(s)$. If the controller, $G_c(s)$, is set $G_c(s) = G_m(s)^{-1}$ and the model is correct then the open-loop response is equal to the set point. This is the ideal case and cannot work in practice, because there are always disturbances and modeling errors. Additionally, the model is not always invertible. Therefore a feedback of the error between the model and the process output is introduced.

The process model $G_m(s)$, is then split into an invertible $G_m^-(s)$ and a non-invertible $G_m^+(s)$ part:

$$G_m(s) = G_m^+(s)G_m^-(s) \tag{1}$$

The non-invertible part, contains all positive zeros and time delays, which upon inverting becomes unstable or non-realisable. The rest of the model consists of the invertible part which is incorporated into the controller. The non-invertible part is treated as un-modeled dynamics and is handled by the feedback.

The controller is then

$$G_c(s) = G_m^-(s)^{-1}G_f(s) \tag{2}$$

where

$$G_f(s) = \frac{1}{(\lambda s + 1)^n}$$

is a filter with appropriate n to make the controller proper. λ is a positive tuning parameter that can be manually set to achieve a suitable response.

The comparison graph of multiloop controller and internal model controller is shown in Fig. 6a and b.

The comparison graph of decoupling controller and internal model controller is shown in Fig. 7a and b.

From the above graph one can found that internal model control is better than multiloop and decoupler. The performance index calculated for the three controller designs are given in table. The performance analysis for loop 1 is shown in Table 1, and for loop 2 is given in Table 2.

The performance analysis suggests that internal model control is better than the other two controllers.

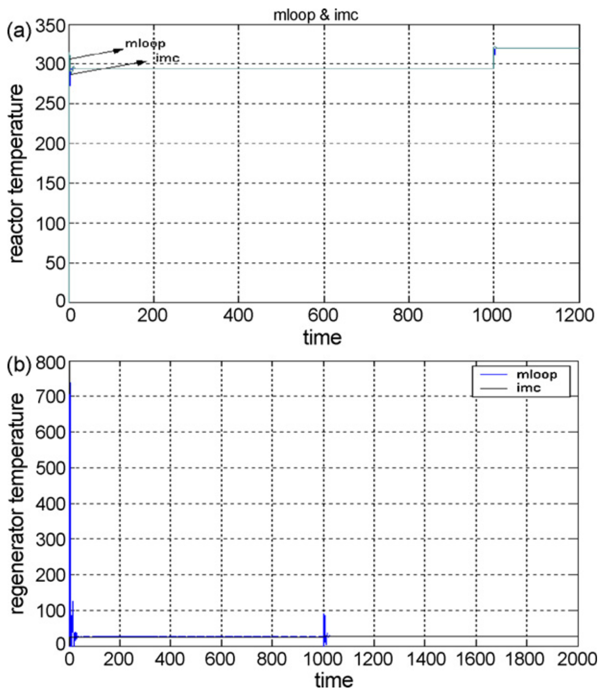

Fig. 6 – (a) Comparison of multiloop with IMC:loop 1 and (b) comparison of multiloop with IMC:loop 2.

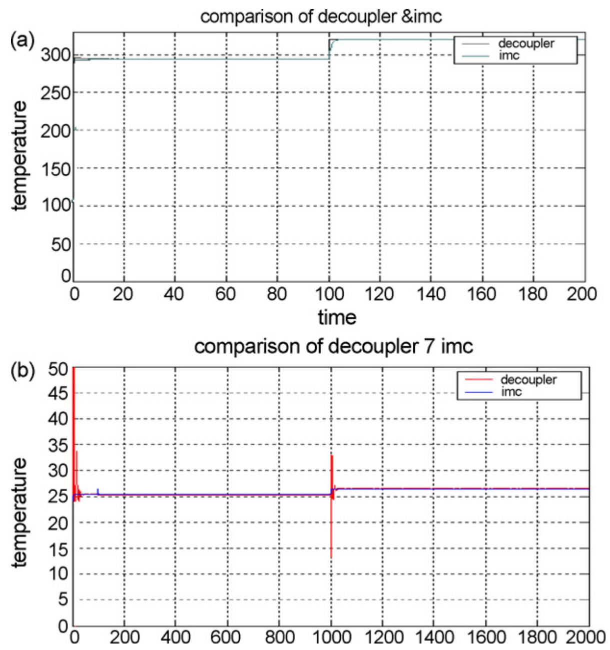

Fig. 7 – (a) Comparison of decoupler with IMC:loop 1 and (b) comparison of decoupler with IMC:loop 2.

Table 1 – Performance index for loop 1.

Index	Decoupler	Multiloop	IMC
ISE1	6926	10,630	4397
IAE1	90.3	153.3	45.37
Settling time	10 s	20 s	4 s
Peak overshoot	297	315	294

Table 2 – Performance index for loop 2.

Index	Decoupler	Multiloop	IMC
ISE2	2940	33,700	790
IAE2	253.3	5733	81.3
Settling time	20 s	45 s	10 s
Peak overshoot	35	715	25.35

7. Conclusion

In this paper, a systematic method is proposed for the design of multiloop control configurations for complex processes. The amount of interaction and selection of variable pairing is found by RGA analysis. Decoupler is designed to remove the interaction completely. Internal model control design is proposed for this FCC process. Numerical simulation on a FCC unit shows that the resulting control performance with the proposed design is very satisfactory. An advantage of the proposed control design is that easy tuning procedures can be designed. Furthermore, the compensator design procedure is relatively simple and can be implemented easily.

References

Ali, H., Rohani, S. and Corriou, J.P., 1997, Modeling and control of a riser type fluid catalytic cracking (FCC) unit. Transactions of the Institution of Chemical Engineers, Part A, 75: 401–411.

Alvarez-Ramirez, J., Valencia1, J. and Puebla, H., 2004, Multivariable control configurations for composition regulation in a fluid catalytic cracking unit. Chemical Engineering Journal, 99: 187–201.

Ansari, R.M. and Tade, M.O., 2000, Constrained nonlinear multivariable control of a fluid catalytic cracking process. Journal of Process Control, 10: 539–555.

Arbel, A., Huang, Z., Rinard, H.I., Shinnar, R. and Sapre, A.V., 1995, Dynamics and control of fluidized catalytic crackers. 1. Modelling of the current generation of FCCs. Industrial Engineering Chemistry Research, 34: 1228.

Avidan, A. and Shinnar, R., 1990, Development of catalytic cracking technology: a lesson in chemical reactor design. Industrial Engineering Chemistry Research, 29: 931–942.

Bristol, E.H., 1969, On a new measure of interactions for multivariable process control. IEEE Transaction on Automatic Control, AC-11: 133–134.

Einashaie, S.S.E.H. and Elshishini, S.S., 1993, Digital simulation of industrial fluid catalytic cracking units-IV. Dynamic behavior. Chemical Engineering Science, 48: 567–583.

Felipe, I.L., 1992, Dynamic modeling of an industrial fluid catalytic cracking unit, In *Presented at European Symposium on Computer Aided Process Engineering*—1 Escape 1, Elsinore, Denmark, , pp. S139–S148.

Garcia, C.E. and Morari, M., 1982, Internal model control. 1. A unifying review and some new results. Industrial, Engineering Chemistry Process Design and Development, 21(2)

Han, I.S. and Chung, C.B., 2001, Dynamic modeling and simulation of a fluidized catalytic cracking process. Part I: process modeling. Part II: property estimation and simulation. Chemical Engineering Science, 56: 1951–1990.

Hovd, M. and Skogestad, S., 1993, Procedure for regulatory control structure selection with application to the FCC process. American Institute of Chemical Engineers Journal, 39: 1938–1953.

Lee, E. and Groves, F.R., 1985, Mathematical models of the fluidized bed catalytic cracking plant. Transactions of the Society for Computer Simulation, 219.

Malay, P., Milne, B.J. and Rohani, S., 1999, The Modified Dynamic of a Riser Type Fluid Catalytic Unit. The Canadian Journal of Chemical Engineering, 77: 169–179.

McFarlane, R.C., Reineman, R.C., Bartee, J.F. and Georgakis, C., 1993, Dynamic simulator for a model IV fluid catalytic cracking unit. Computers & Chemical Engineering, 17(3): 275–300.

Rivera, D.E., Morarl', M. and Skogestad, S., 1986, Internal model control. 4. PID controller design. Industrial, Engineering Chemistry Process Design and Development, 25: 252–265.

Skogestadt, S. and Morari, M., 2009, Implications of large RGA elements on control performance. Chemical Engineering, 206–241.