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ANALYSIS OF MULTIVARIABLE CONTROLLER FOR FLUID CATALYTIC CRACKING UNIT IN A PETROLEUM REFINERY

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ABSTRACT

The projected paper has been scrutinized and revised the circulated experimental results related to reactor-regenerator structure of fluid catalytic cracking unit (FCCU). It correspondingly focuses the limitations of existing methods for controlling the heat in the reactor-regenerator system. Now the extant study, complex dynamic model of the reactor–regenerator system has been industrialized and successively castoff with the controller. FCC process is considered as a primary conversion unit in a incorporated refinery and epitome FCC operation can have a significant impact on the refinery profitability. Control of the FCC continues to be a stimulating and momentous problem due to interaction between the loops. So there is a tangible requisite for a control logic that effectively utilizes the available process measurements and archetypal information representative of the process. Internal Model Controller (IMC) discloses the set point tracking Stabilization can be succeeded through simulated implementation of a model predictive control (MPC).

Keywords

FCCU, Control, PID,IMC, MPC

INTRODUCTION

The fluid catalytic cracking is unique of the chief imperious processes in the oil refineries. Its design and operation are predominantly targeted for the making of gasoline with a higher octane rating and olefinic gases. The ratio of cracking and the culmination products are sturdily reliant on the temperature and presence of catalysts. The fluid catalytic cracking unit (FCCU) devours developed the taxing workbench of many advanced control methods. Today, in cooperation academia and industry are articulating great interest in the enhancement of new control algorithms for their proficient industrial FCC implementation. Scrutiny and control of FCC process have been originate perplexing problems owing to the following process characteristics, (i) very tortuous and tiny known hydrodynamics, (ii) composite kinetics of together cracking and coke burning responses, (iii) durable interaction between the reactor and regenerator, (iv) voluminous operational restraints. FCCU's steady state comportment is remarkably nonlinear, leading to manifold steady states, input multiplicities etc. This is auxiliary composite problem meanwhile the numeral of progression variables that lone would like to control expansively outstrips the number of manipulated variables that are obtainable for the task.

PROCESS MODELLING

A condensed process representation and instrumentation diagram is shown in the figure below.The archetypal FCC component encompasses of three instruments, a riser reactor, a catalyst stripper and a regenerator. Reactor-Regenerator section is meticulous as core of the FCC unit. For modeling FCCU, these mechanisms should be pondered separately then they are integrated to pretend the entire FCC unit. In this mode the crucial measured variables are ideal to be the reactor temperature/riser outlet temperature (y_1) and the regenerator gas temperature (y_2).The manipulated variables are flow rate of regenerated catalyst (m_1), flow rate of spent catalyst and flow rate of air to the regenerator(m_2).The demonstrating of composite chemical systems for the reiteration of course dynamics and control has been enthused by the economic enticements for enrichment of plant operation and plant strategy, as in the case of FCCU. The delinquent is to treasure official schemes that are**(1)** Active, **(**2) Sparingly reasonable, **(**3**)** Related to prevalent practice, and **(**4**)** Able to afford adequate operator interface when favored.A inert model is hand-me-down for the riser .In this toil, an operative pragmatic control scheme of Joseph A.Bromley and Thomas J.Ward (1981) is cast-off. In this arrangement, feed is gasoil which be talented to cleft into gasoline or light gas.

The equilibrium equation for Hold up of catalyst is,

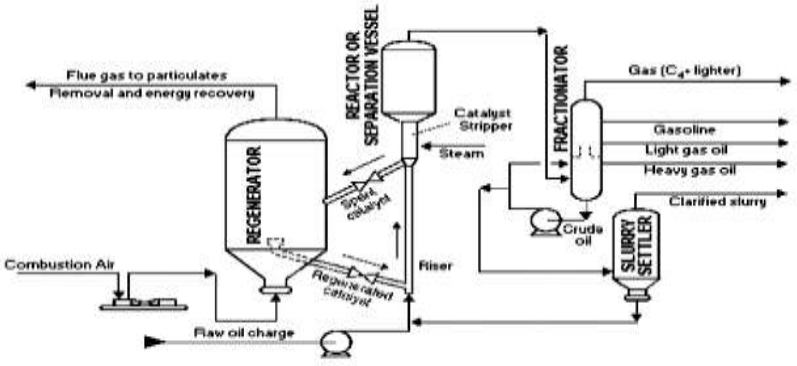

Figure 1:Schematic representation of FCC unit

$$\frac{dH_{RA}}{dt} = [R_{RC} - R_{SC}] \quad (1)$$

Note : $\frac{dH_{RG}}{dt} = -\frac{dH_{RA}}{dt}$

The balance equation for concentration of spent catalyst:

$$\frac{dC_{SC}}{dt} = \frac{1}{H_{RA}}[R_{RC}C_{RC} - R_{RC}C_{SC} + 100R_{CF}](2)$$

And also, $R_{CF} = R_{CC} + F_{CF}R_{TF} + 0$; (3)

The balancing equation for concentration of catalytic carbon:

$$\frac{dC_{CAT}}{dt} = \frac{1}{H_{RA}}[-R_{RC}C_{CAT} + 100R_{CC}](4)$$

The balancing equation for concentration of carbon on regenerated catalyst:

$$\frac{dC_{RC}}{dt} = \frac{1}{H_{RG}}[R_{SC}(C_{SC} - C_{RC}) - 100R_{CB}](5)$$

Where,

$$R_{CB} = \left(\frac{R_{AI}}{C_1}\right)(21 - O_{FG})/100(6)$$

Reactor model

The catalyst is steam wide-open in the reactor vessel to eliminate hydrocarbons. The reactor slide is separated in a cyclone to exterminate catalyst and the product vapors pass to a product fractionator. The poise equation of reactor temperature is given by,

$$\frac{dT_{RA}}{dt} = \frac{R_{RC}}{H_{RA}}(T_{RG} - T_{RA}) + \frac{1}{S_C H_{RA}}[-S_F D_{TF} R_{TF}(T_{RA} - T_{TF}) - \Delta H_{FV} D_{TF} R_{TF}] - \frac{\Delta H_{CR} R_{OC}}{S_C H_{RA}} \quad (7)$$

And also, $R_{OC} = D_{TF}R_{TF}C_{TF}(8)$

$$\frac{C_{TF}}{1 - C_{TF}} = \frac{K_{CR}P_{RA}H_{RA}}{R_{TF}} \ (9)$$

Regenerator model

The poise equation of Regenerator temperature is given by,

$$\frac{dT_{RG}}{dt} = \frac{R_{SC}}{H_{RG}}(T_{RA} - T_{RG}) + \frac{1}{S_C H_{RG}}[-S_A R_{AI}(T_{RG} - T_{AI}) + \Delta H_{RG} R_{CB}] \ (10)$$

MIMO SYSTEM

Contemplate a technique with two measured outputs and two operated inputs in figure 2.

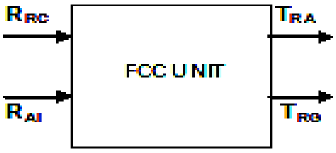

Figure 2. MIMO system

The input-output relationships are given by,

$y_1(s) = G_{11}(s)m_1(s) + G_{12}(s)m_2(s)$ (11)

$y_2(s) = G_{21}(s)m_1(s) + G_{22}(s)m_2(s)$ (12)

Where $G_{11}(s)$,$G_{12}(s)$,$G_{21}(s)$ and $G_{22}(s)$ are the four transfer functions relating the two inputs and the two outputs. These equations elect that the transformation in m_1 or m_2 , will distress both controlled outputs.Two prospective hitches arise from this process interaction (1) it may unsettle the closed loop system and (2) It inclines to sort controller tuning promote problematic.

CONTROL SCHEMES

PID Controller

In more than 95% of the control loops are of PID type has been actually PI control. PID control is an imperative ingredient of a distributed control system.

The basic features of the PID controller is described by,

$$u(t) = K\left\{e(t) + \frac{1}{T_i}\int_0^t e(\tau)d\tau + T_d\frac{de(t)}{dt}\right\} \quad (13)$$

PID control is ever-present. While modest in theory, design and implementation of PID controllers can be problematic and time intense in run-through.

- PID control implicates numerous responsibilities that include:
- Choosing an applicable PID algorithm (P, PI, or PID)
- Improvements in controller tuning
- Pretending the controller in contradiction of a plant model

Internal Model Controller

The IMC-PID tuning instructions disclose good set point tracking nevertheless sluggish disturbance rejection, which befits austere when a process has a small time-delay. The mock trainings of several process models show that the proposed design mode affords progressed disturbance rejection for lag-time dominant process, when the various controllers are all regulated to have the identical degree of sturdiness letting to measure of maximum sensitivity.

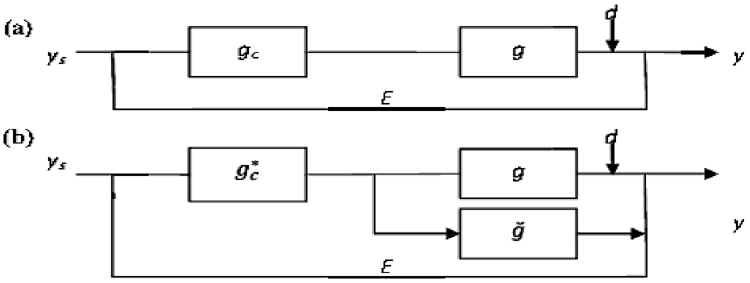

Figure 3.(a) :Conventional configuration and Figure 3.(b) :Internal Model Control configuration

The goal of control system task is to realize a wild and accurate set-point tracking. This implies that the effect of external disturbances should be concise as proficiently as possible and also being guaranteed of tactlessness to modeling error.

The PID parameter tuning edict established on the relationship of the IMC and the PID controller has been proposed by Rivera *et al*. [1986]. PID control erection is shown in Fig.3 (a), where g_c and g_c* are the PID controller and the controlled process, respectively. They are given by

Proportional gain(K_c) $K_c = \frac{T+0.5\theta}{K(\lambda+0.5\theta)}$ (14)

Integral Time(T_i) $T_i = T + 0.5\theta$ (15)

Derivative Time(T_d) $T_d = \frac{T\theta}{2T+\theta}$ (16)

Model Predictive Controller

Model Predictive Control (MPC) is a reformist method of process control that has been in practice in the process activities in chemical plants and oil refineries meanwhile the 1980s. MPC practices an explicit dynamic plant model to predict the influence of imminent reactions of the deployed variables on the output and the control signal acquired by minimizing the cost function. This predication proceeds into account, restraints on both the inputs and outputs of the process. MPC is based on iterative, determinate horizon optimization of a plant model. At time t, the state is sampled and a cost curtailing control tactic is work out for a relatively short time horizon in the future.

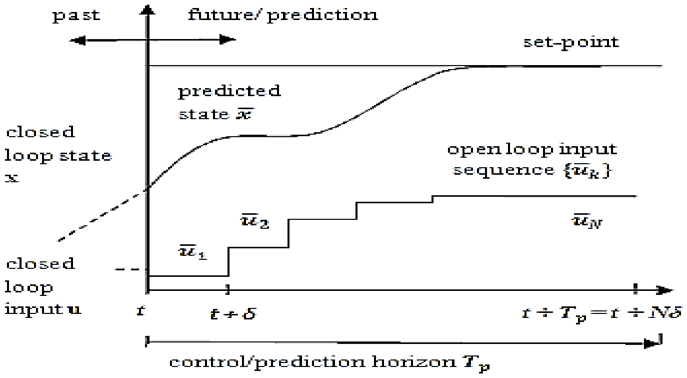

Figure 4: Principle of MPC

As articulated formerly, MPC is a multivariable control algorithm that habits an internal dynamic model of the process or system, past control passages and an optimization cost utility J over the retreating prediction horizon, to reckon the optimum control passages. The optimization cost function is as trails:

$$J = \sum_{i=1}^{N} \omega_{xi} \, (r_i - x_i)^2 + \sum_{i=1}^{N} \omega_{ui} \, \Delta u_i{}^2 \quad (17)$$

The prediction horizon (P) conveys the controller how many sample steps ahead should be recycled when curtailing the object function. The control horizon (M) conveys the controller how many control steps should be recycled when curtailing. These precincts can lead to an infeasible solution set for the controller. Some advantages of MPC embrace straightforward origination, based on well understood concepts and explicit treatment of limits. Its improvement time much tinier than for challenging radical control methods. Its ability to antedate imminent events and can take control action consequently is an imperative factor.

RESULTS AND DISCUSSION

FCCU process embraces of two inputs and two outputs.In practical combustion mode, the commonchoice of variables have been regulated are the riser outlet temperature (T_{RA}) and regenerator's temperature (T_{RG}).If the pairings T_{RA} -R_{RC} and T_{RG}-R_{AI} were selected to design a decentralized control strategy, a classical rise-regenerator control structure has been obtained.

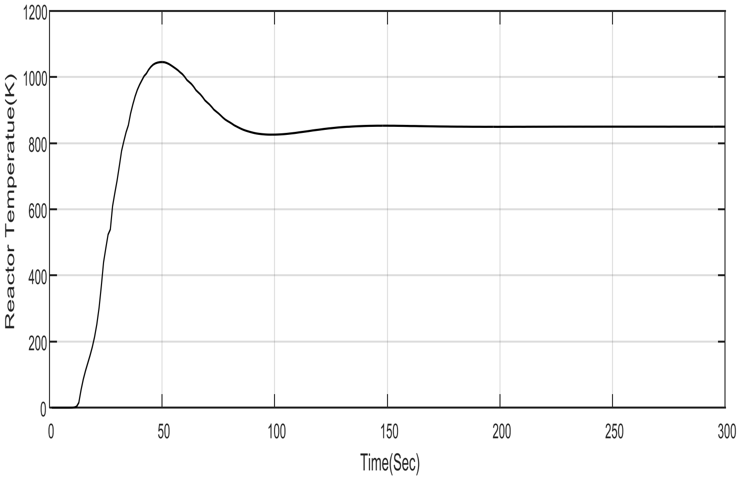

Figure 5: Response of Reactor Temperature using with PID Controller

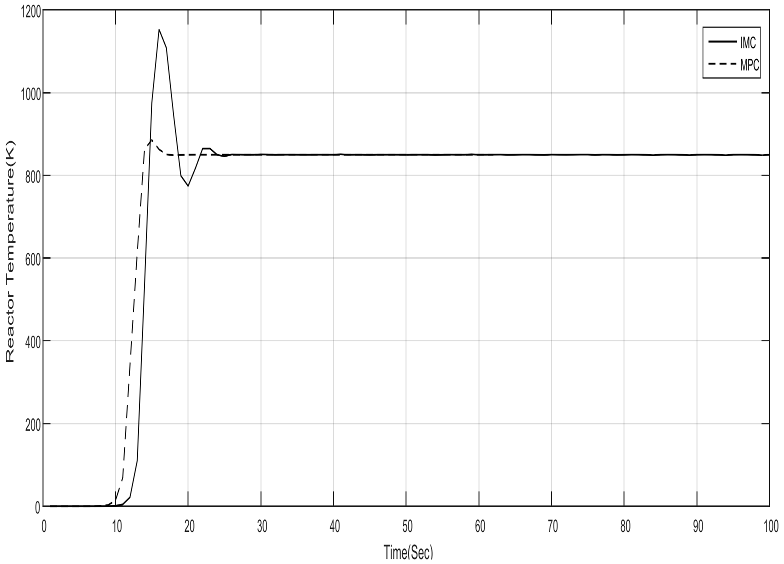

Figure 6:Response of Reactor Temperature using with IMC-PID and MPC Controller

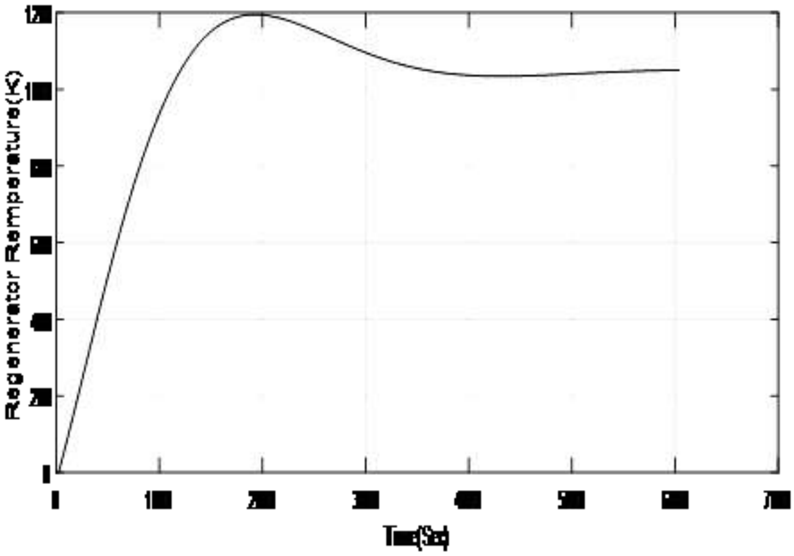

Figure 7: Response of Regenerator Temperature using with IMC-PID and MPC Controller

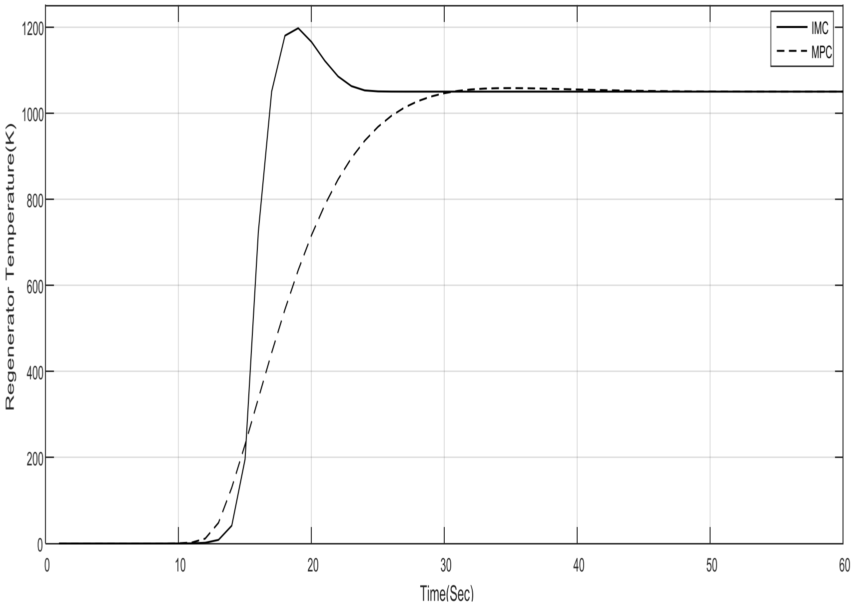

Figure 8: Response of Regenerator Temperature using with IMC-PID and MPC Controller

Table 1.Performance Analysis of Reactor temperature using various controllers

Performance evaluation	PID	IMC – PID	MPC
ERROR	0.1	0.02	0
ISE	4	0.12	0
IAE	40	6	0
ITAE	16000	1800	0
SETTLING TIME(SEC)	400	300	42
DELAY TIME(SEC)	350	12.5	12.2
RISE TIME(SEC)	30	2.5	2
PEAK TIME(SEC)	80	16	15
PEAK VALUE	967.5	1153	885.6

Table 2.Performance Analysis Regenerator temperature using various controllers

Performance evaluation	PID	IMC – PID	MPC
ERROR	0.1	0.3	0.1
ISE	4	9	0.32
IAE	40	9	3.2
ITAE	16000	270	102.4
SETTLING TIME(SEC)	400	27	32
DELAY TIME(SEC)	350	15.8	11.2
RISE TIME(SEC)	30	2.3	9
PEAK TIME(SEC)	80	19	20
PEAK VALUE	967.5	1197	1058

CONCLUSION

FCCU is engaged as the prime system model. This non-linear 2 input 2 output system, is controlled with multi loop IMC-PID and MPC controller. Presentation guides like settling point, overshoot, and ISE, IAE, ITAE errors of MPC controller is compared with IMC-PID controller. The end result portrays MPC is far better than the IMC-PID in all etiquettes as it affords smooth reference tracking with reduced peak overshoot and improved closed loop performances such as ISE, IAE, ITAE. This confirmations why MPC is copious proper for industrial applications PID and indorse better stabilization while using MPC. In sight, the effect of reactor temperature and regenerator temperature dissect through conversion of total feed (volume fraction) using MPC.

NOMENCLATURE

C_1 - Fitting constant for particular data

C_{CAT} - Concentration of catalytic carbon on catalyst, wt%

C_{RC} - Concentration of regenerated catalyst, wt%

C_{SC}- Concentration of spent catalyst, wt%

C_{TF} - Conversion of total feed, volume fraction

D_{TF} - Density of total feed, kg/m^3

F_{CF} - Factor for carbon formation of feed, (kg carbon/s) / (m^3/s)

H_{RA} - Hold up of catalyst in the reactor, Kg

H_{RG} - Hold up of catalyst in the regenerator, Kg

O_{FG} - Oxygen in flue gas, mol%

R_{AI} - Rate of regenerator air, kg/s

R_{CB} - Rate of coke burning, kg/s

R_{CC} - Rate of catalytic carbon formation in the reactor, kg/s

R_{CF} - Rate of carbon forming on catalyst, kg/s

R_{OC} - Rate of gas oil cracking, kg/s

R_{RC} - Rate of regenerated catalyst, Kg/s

R_{SC} - Rate of spent catalyst, kg/s

R_{TF} – Rate of total feed, m^3/s

S_A - Specific heat of air, J/kg-k

S_C - Specific heat of catalyst, J/kg-k

T_{AI} - Temperature of air, K

T_{RA} - Temperature of Reactor, K

T_{RG} - Temperature of Regenerator, K

T_{TF} - Temperature of feed, K

ΔH_{CR} - Heat of cracking, J/kg

ΔH_{FV} - Heat of feed vaporization, J/kg

ΔH_{RG} – Heat of regeneration (coke burning), J/kg

x_i - i-th controlled variable

r_i - i-th reference variable

u_i - i-th manipulated variable

w_{xi} - weighting coefficient for relative importance of x_i

w_{ui} - weighting coefficient of relative big changes in u_i

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Bibliography

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