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CONTRIBUTIONS TO THE MODEL PREDICTIVE CONTROL OF THE UOP FLUID CATALYTIC CRACKING UNIT

Maria MIHEŢ[a,b*] and Vasile Mircea CRISTEA[b]

[a] National Institute for Research and Development for Isotopic and Molecular Technologies, INCDTIM, 65-103 Donath Street, Cluj-Napoca 400293, Roumania
[b] Babeş-Bolyai University, Faculty of Chemistry and Chemical Engineering, 11 Arany Janos Street, Cluj-Napoca 400028, Roumania

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Fluid catalytic cracking (FCC) is one of the most important processes in oil refining due to its economic impact. Therefore, development of optimal FCC control strategies is of great interest, as it could lead to an enhancement of economic profitability. Model Predictive Control (MPC) of a FCC unit based on a reliable mechanistic model is presented in this work. This analytical model uses a five lumped species kinetic model by Ancheyta-Juarez *et al.*, [1] which was integrated in a previously developed mechanistic model (mass, energy, and momentum balance equations coupled with constitutive equations) based on construction and operating data originating from an industrial FCC plant. The mechanistic model based dynamic simulator revealed a very good fit between simulated and the industrial behavior of the unit, as well as a good description of the fluid catalytic cracking unit (FCCU) behavior at different scenarios of upsets in manipulated variables and disturbances. Several control structures were proposed and comparatively investigated by simulation, in the presence of two typical disturbances: main fractionator pressure drop and change in the properties of the feed (change of the cracking kinetic constants).

INTRODUCTION

The importance of fluid catalytic cracking (FCC) in oil refining is given by the fact that during this process less valuable heavy hydrocarbon molecules are converted into lighter molecules, with significantly greater economic value. Crude oil fractions are cracked to lighter fractions in a riser reactor which contains a fluidized bed of catalyst. Both hydrodynamic and chemical aspects make this process a very complex one,[2] leading to difficulties in process modeling as well in control.[3] From the chemical point of view, the complexity of the FCC process is a consequence of its complicated kinetics, as the feedstock consists of thousands of components. Thus, lumping of components according to the boiling range was the key concept for the development of several kinetic models with different number of lumps: 3 lumped species kinetic model,[4] 4 lumped species,[5] 5 lumped species,[1] or even 10 lumped species kinetic model.[6] Under these circumstances, the estimation of intrinsic kinetic constants which are necessary for the development of an analytical model of the process is extremely difficult. On the other hand, process control is a challenging task because FCC is a multivariable, nonlinear, and highly interactive process which is subjected to many operational, safety and environmental constraints.[7, 8]

The methodology of model predictive control (MPC) proved to be by now a good approach for FCC process control, due to its multivariable structure, optimal character and robustness to modeling and measurement errors.[8, 9] The main characteristic of MPC approach is that it calculates the control action based on the prediction of future dynamics of the system, instead on past or present information. Thus, early control action can be taken accounting for future behavior.

A typical UOP (Universal Oil Products) type FCCU with its control diagram is presented in

* Corresponding author: maria.mihet@itim-cj.ro

Fig. 1. The preheated feed is injected at the bottom of the riser and lifted up together with the hot regenerated catalyst flow. The hot regenerated catalyst flow entering the riser provides the required heat for the endothermic cracking reactions that take place in the ascendant flux. The obtained products are separated from the spent catalyst, and further processed in the main fractionator, while the spent catalyst is regenerated in a separate fluid-bed regenerator.

The main objective of this paper is to present several MPC control structures that were investigated and tested by dynamic simulation of the UOP Fluid Catalytic Cracking Unit, based on a previously developed analytical model that uses a five lumped species kinetic model by Ancheyta-Juarez *et al.*[1, 10]

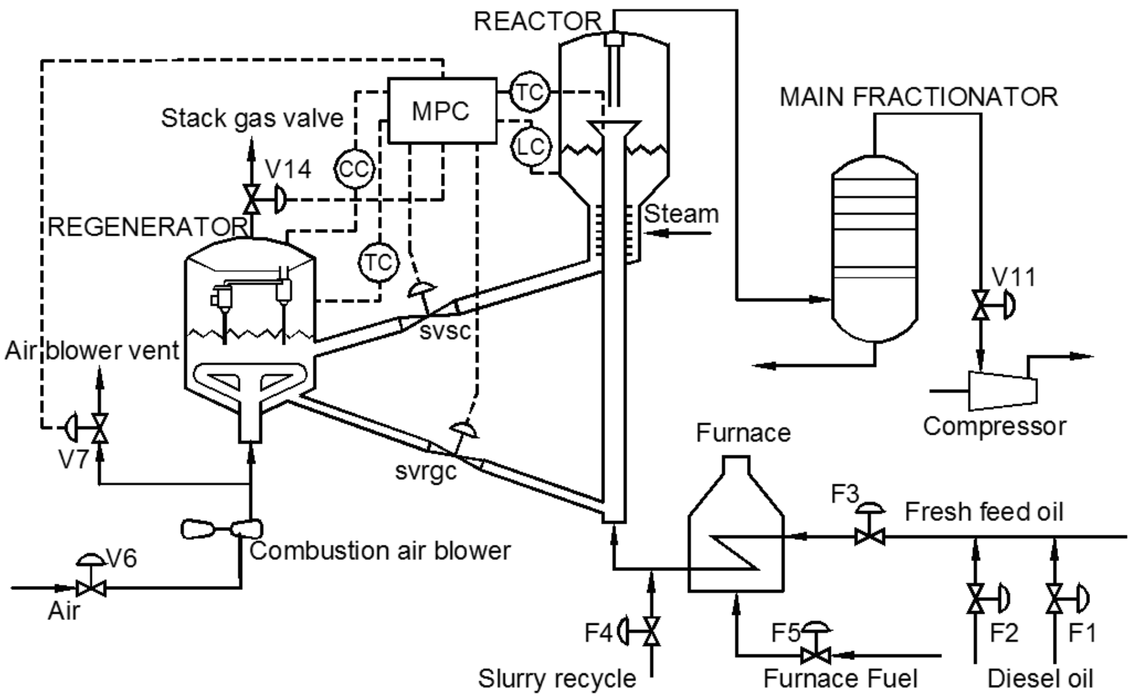


Fig. 1 – Schematic representation of a FCCU with control diagram.

RESULTS AND DISCUSSION

Dynamic simulations

A series of FCCU dynamic simulations were performed in order to study the response of the process to different upsets in manipulated variables and disturbances. The dynamic simulator is based on a previously developed analytical model,[3] which accounts for reference constructive characteristics and operation data from an industrial unit, and which includes the main reactor-regenerator subsystems. The reactor consists of two parts: the riser, where the cracking reactions take place, and the stripper, where traces of hydrocarbons deposited on the catalyst surface are removed using steam. The analytical model of the reactor is based on the following assumptions: (1) ideal plug flow and short transient times for the riser, and (2) the stripper is of continuous stirred tank reactor type (CSTR). However, a five lumped species kinetic model by Ancheyta-Juarez *et al.*[1] was considered instead of the 3-lumped species kinetic model by Weekman.[4] This modified analytical model is presented elsewhere.[10] The implemented 5-lumped species kinetic model considers gas oil, gasoline, liquefied petroleum gas (LPG), dry gas and coke as pseudo components of the feed. The most important advantage of this kinetic model is that it predicts the formation of coke (the source that supplies the required heat for the feedstock heating and vaporization and for the endothermic reactions), LPG and that it uses less kinetic constants than other 5-lumped species kinetic models.[11-13]

Dynamic simulations of the FCCU behavior were performed and investigated considering different scenarios of upsets in manipulated variables and disturbances, such as reactor-main fractionator pressure drop (Δp_{frac}), change of feedstock properties (coking rate constant, k_c), change of air blower flow or a change in the flow of regenerated (*svrgc*) or spent catalyst (*svsc*). The dynamic response at a disturbance given by a change of feedstock properties (coking rate constant) is presented in Fig. 2. A step coking rate disturbance of 1% was applied at time t = 500 s from the beginning of the simulation. Simulation results presented in Fig. 2 reveal the two time periods of change for the process variables.

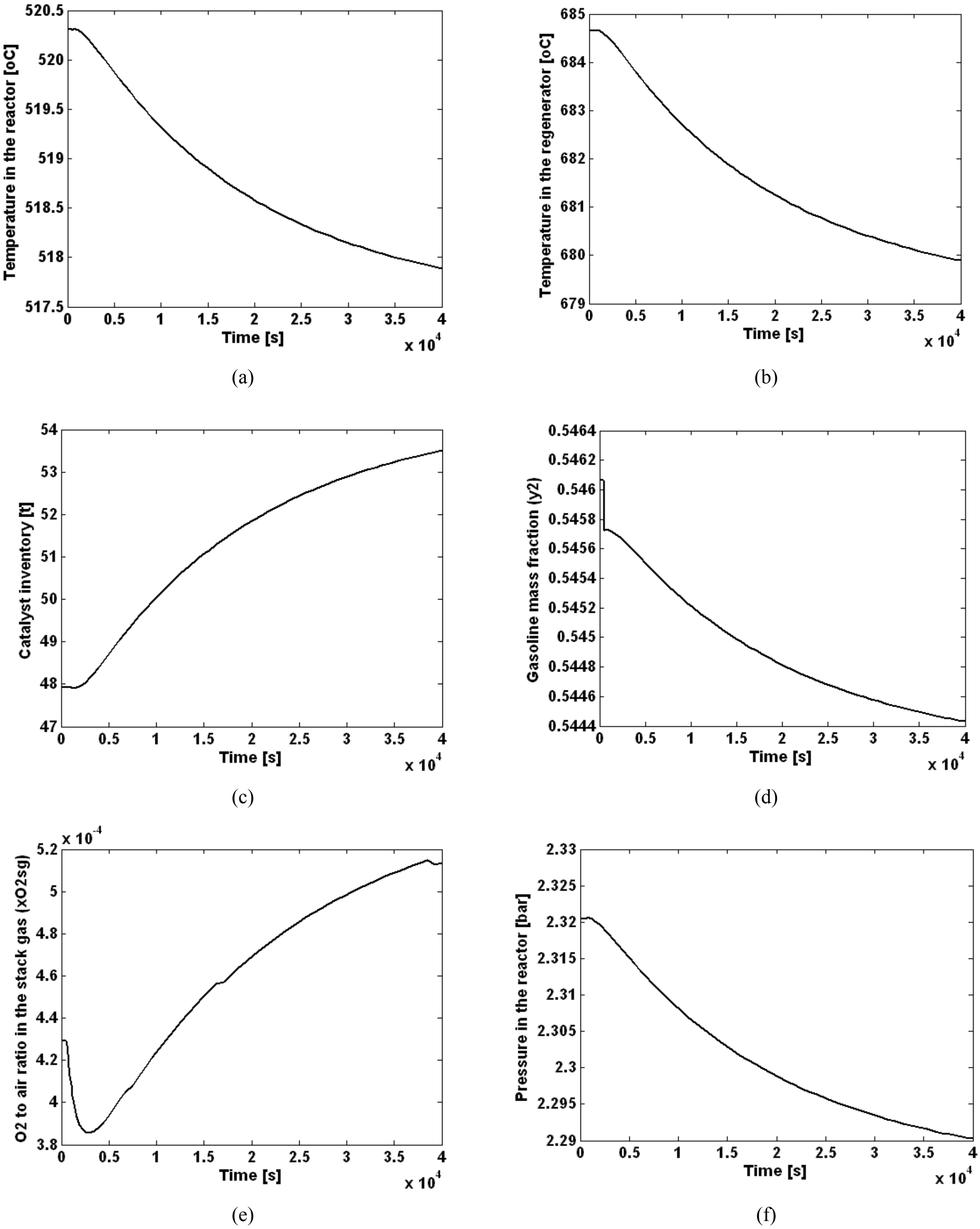


Fig. 2 – Simulation of FCCU dynamic behavior in the presence of a 1% coking rate disturbance. (a) Temperature in the reactor, T_r [°C], (b) Temperature in the regenerator, T_{reg} [°C], (c) Catalyst inventory in the reactor, W_r [t], (d) Gasoline mass fraction, y_2, (e) O_2 concentration in the stack gas, xO_2sg, (f) Pressure in the reactor, P_r [bar].

In the first period, an increase of coking rate constant leads to an increase of coke on the spent catalyst which further enters the regenerator. This hot spent catalyst induces a small temperature rise in the regenerator, and then in the reactor, leading to the intensification of cracking reactions in the reactor, and thus to an increase of pressure. The catalyst inventory in the reactor decreases in this period as spent catalyst flow rate becomes greater than the flow rate of regenerated catalyst.

In the second period, the decrease of temperature in the regenerator induces a decrease of reactor temperature, followed by a decrease of pressure in the reactor due to less intensified cracking reactions. This reduction of pressure in the reactor leads to the decrease of spent catalyst flow rate compared to the regenerated catalyst flow rate. As a consequence, during this second period, the catalyst inventory in the reactor increases. The duration of the first period is significantly shorter than that of the second period. These dynamic simulations prove the multivariable, nonlinear and highly interactive behavior of the FCC process.

Model Predictive Control of the FCCU

This section presents the investigated model predictive control structures which were used for the control of the FCCU.

MPC, also known as moving horizon control, is by now a widely used control strategy for both linear and nonlinear systems due to its flexibility and robustness. MPC is basically formulated as an open loop optimal control with finite horizons, which is subject to the system dynamics and to input or state constraints. The designed model predictive controller makes a prediction of the future behavior over an output prediction time horizon, *p*, based on the present time acquired information, and on the steady state model of the considered process.[8,14] This prediction is then compared to the desired trajectory, and by minimization of an objective performance function (usually the square error between setpoint and prediction) an open loop manipulated variable sequence (over an input horizon, *m*) is determined. Only the first step of the manipulated variable is implemented until the next time step, when a new set of measured values becomes available. Prediction and optimization are repeated again in order to obtain a new manipulated variable sequence, with input and output time horizons shifted one step into the future.

Design of a model predictive controller requires a plant model which defines the mathematical relationship between the inputs and outputs of the plant in order to predict the next process move and hence the appropriate control actions. For the purpose of this work, the modified analytical model presented above was linearized in order to obtain a linear, time invariant (LTI) model, and then transformed to a state space model, as required by the MPC algorithm.[14]

Several control structures were investigated and tested by dynamic simulation (Table 1). Both controlled and manipulated variables have been chosen based on the analysis of the industrial FCCU, in order to provide its efficient and safe operation. Consequently, as controlled variables, catalyst inventory, W_r, temperature in the reactor, T_r, and in the regenerator, T_{reg}, stack gas oxygen concentration, xO_2sg, and gasoline yield, y_2 were considered.[3] The manipulated variables were chosen from among those variables that are possible to be changed practically, such as: regenerated and spent catalyst flow rates (by the position of the corresponding slide valves, *svrgc*, and *svsc*), fuel flow of the preheating furnace, F_5, stack gas flow rate from the regenerator, V_{14}, or the air vent flow rate, V_7[3].

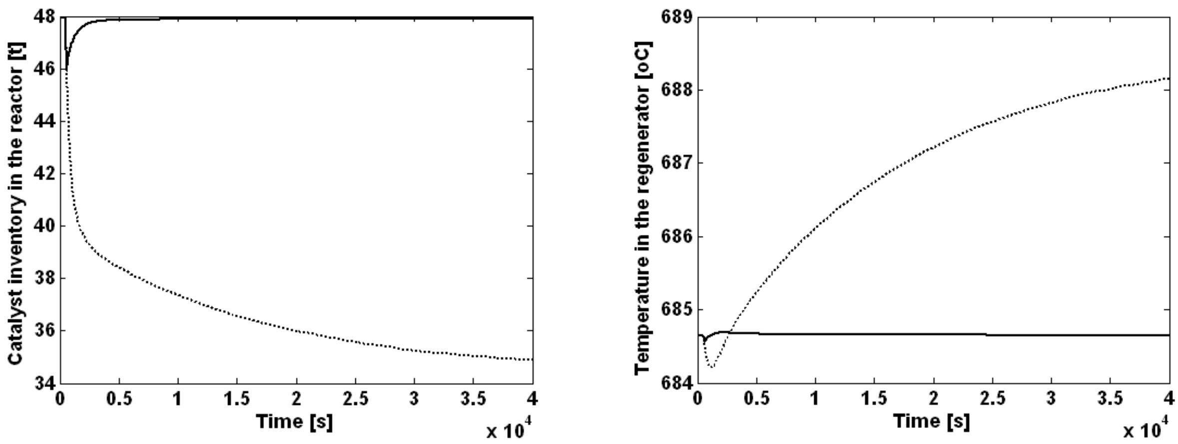
The investigated and tested control structures have a different number of controlled/manipulated variables: 2 x 2, and 4 x 4. The specific MPC controllers were designed in Matlab®, using its specialized MPC Toolbox. MPC simulations were performed in Simulink® with an s-function as FCC model of the plant. Each control structure was tested by dynamic simulation in the presence of the two main disturbances: reactor-main fractionator pressure drop (Δp_{frac}), and change of feedstock properties (coking rate constant, k_c). The MPC simulation results for two control schemes (S1: 2 x 2, and S3: 4 x 4) in the presence of the two considered disturbances are compared to the dynamic simulation results in the absence of the plant control. The results can be summarized as followings.

The first control structure, S1: 2 x 2 succeeds to counteract the disturbance effects, presenting small overshoot and short settling time, in case of both controlled variables W_r and T_{reg} (see Fig. 3 and Fig. 4). This behavior demonstrates a good capacity of the designed MPC controller to follow the established set points.

Table 1

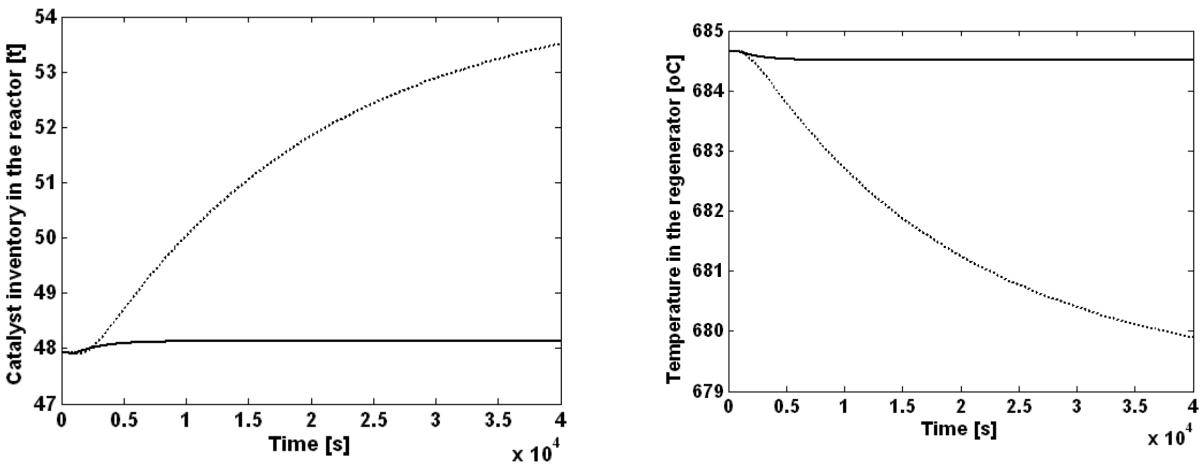
Investigated control structures

Control structure (name/dimension)	Controlled variables	Manipulated variables
S1: 2 x 2	W_r and T_{reg}	*svrgc* and *svsc*
S2: 2 x 2	T_{reg} and y_2	*svrgc* and *svsc*
S3: 4 x 4	W_r, T_{reg}, xO_{2sg} and T_r	*svrgc*, *svsc*, V_{14} and V_7
S4: 4 x 4	W_r, T_{reg}, y_2 and T_r	*svrgc*, *svsc*, V_{14} and F_5
S5: 4 x 4	W_r, T_{reg}, y_2 and T_r	*svrgc*, *svsc*, V_{14} and V_7



(a) (b)

Fig. 3 – MPC simulation results (solid line) in the presence of the pressure drop (Δp_{frac}) disturbance (10% step increase at t = 500s), for S1: 2 x 2 control scheme, compared to the disturbed process without control (dashed line). (a) Catalyst inventory in the reactor, W_r [t], (b) Temperature in the regenerator, T_{reg} [°C].



(a) (b)

Fig. 4 – MPC simulation results (solid line) in the presence of the coking rate disturbance (1% step increase at t = 500 s), for S1: 2 x 2 control scheme, compared to the disturbed process without control (dashed line). (a) Catalyst inventory in the reactor, W_r [t], (b) Temperature in the regenerator, T_{reg} [°C].

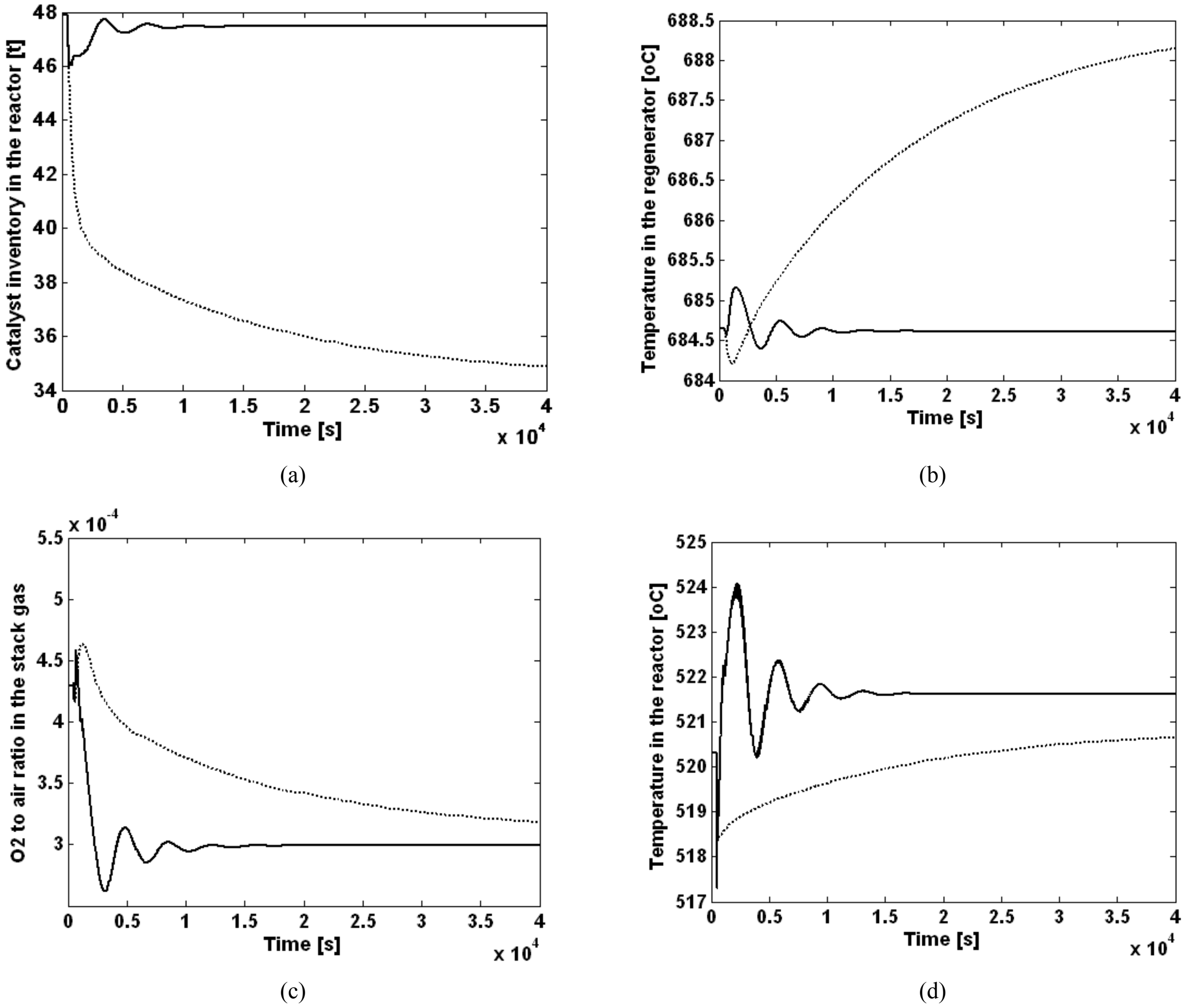


Fig. 5 – MPC simulation results (solid line) in the presence of the pressure drop (Δp_{frac}) disturbance (10 % step increase at t = 500 s), for S3: 4 x 4 control scheme, compared to the disturbed process without control (dashed line). (a) Catalyst inventory in the reactor, W_r [t], (b) Temperature in the regenerator, T_{reg} [°C], (c) O_2 concentration in the stack gas, xO_2sg, (d) Temperature in the reactor, T_r [°C].

Another control structure to be presented and discussed is S3: 4 x 4, which shows inferior control performance for the selected controlled variables: W_r, T_{reg}, xO_{2sg} and T_r, in the case of both disturbances (Fig. 5 and Fig. 6). Control of catalyst inventory in the reactor, W_r, and temperature in the regenerator, T_{reg}, showed good performance, with relative short settling times and small overshoot (Fig. 5 (a), (b), and Fig. 6 (a), (b)). However, control of oxygen concentration in the stack gas and of temperature in the reactor is less accurate due to increased overshoot, and larger offset (Fig. 5 (c), (d), and Fig. 6 (c), (d)).

The other investigated control structures presented similar or inferior performances. Thus, the second 2-dimension control structure, S2: 2 x 2, succeeds to counteract the effect of the 2 disturbances in an efficient manner, for both controlled variables (T_{reg} and y_2).

The other two control structures of 4 x 4 dimension (S4, and S5) show better control performances for W_r and T_{reg}, than for y_2 and T_r, at the disturbance represented by the reactor-main fractionator pressure drop, Δp_{frac}, as well for a change in feedstock properties (change of coking rate).

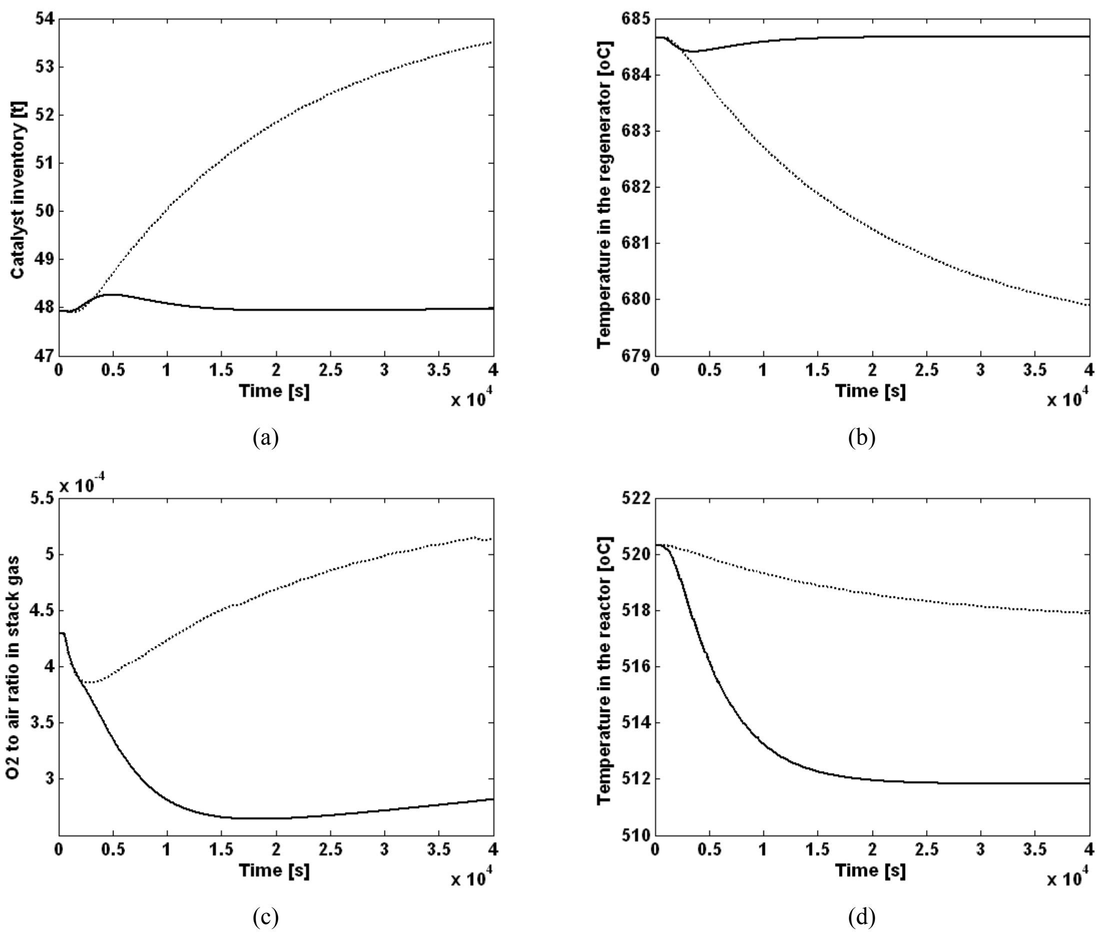


Fig. 6 – MPC simulation results (solid line) in the presence of the coking rate disturbance (1% step increase at t = 500s), for S3: 4 x 4 control scheme, compared to the disturbed process without control (dashed line). (a) Catalyst inventory in the reactor, W_r [t], (b) Temperature in the regenerator, T_{reg} [°C], (c) O_2 concentration in the stack gas, xO_2sg, (d) Temperature in the reactor, T_r [°C].

CONCLUSIONS

This paper presents dynamic simulation results for the FCCU based on an analytical model that uses a 5-lump kinetic model. The performed simulations of different scenarios of upsets in manipulated variables and disturbances highlight the multivariable, nonlinear and highly interactive character of the process. A series of model predictive control algorithms were proposed and investigated based on the previously developed first principle model, using Matlab® and its specialized toolboxes. MPC simulations compared to dynamic simulations for the two main disturbances that may occur in a FCCU showed good control performances in the case of the 2 x 2 dimension control structures (S1: 2 x 2, and S2: 2 x 2). The effect of the investigated disturbances was counteracted with short settling times and small overshoot. The 4 x 4 control structures showed an inferior control performance, as not all of the 4 controlled variables were accurately controlled. For each structure, control of W_r and T_{reg} was efficient, with short settling times and small overshoot. However, control of xO_{2sg}, y_2 and T_r was less efficient, due to longer settling times, large overshoot, or stationary errors. This study revealed advantages and limitations of different FCC control configurations from the perspective of a feasible control structure design.

Nomenclature

FCCU – Fluid catalytic cracking unit
UOP – Universal Oil Products
MPC – Model Predictive Control
svrgc – regenerated catalyst slide valve position [0 - 1]
svsc – spent catalyst slide valve position [0 - 1]
F_5 – fuel flow of the preheating furnace (m^3/s)
V_{14} – stack gas flow rate from the regenerator (valve position)
V_7 – air vent flow rate (valve position)

W_r – catalyst inventory in the reactor [t]
T_{reg} – temperature in the regenerator [°C]
xO_2sg – O_2 concentration in the stack gas
y_2 – gasoline yield [mass fraction]
T_r – temperature in the reactor [°C]

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