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THE DEVELOPMENT OF A MPC CONTROLLER FOR A HEAT INTEGRATED FLUID CATALYTIC CRACKING PLANT

MIHAELA-HILDA MORAR[a] AND PAUL ŞERBAN AGACHI[a]

ABSTRACT. The use of the advanced control techniques is necessary because, now, the PID control is no more competitive in the 20% of the industrial applications when special dynamics are involved. By implementing an advanced control system one can push the process unit to a more profitable region without affecting the operation constraints. In previous works an industrial FCC plant from a Romanian refinery was studied from the point of view of heat integration and steady state performance of the new heat exchanger network (HEN) design. In this study the improvement of the same FCC plant was done by implementing an advanced control scheme capable to maintain stabilized the heat transfer through the plant. Model Predictive Control (MPC) is one of the most used advanced control techniques in process control. A MPC controller was developed for an industrial fluid catalytic cracking (FCC) plant using Aspen HYSYS. The developed MPC controller results enable to establish that the strategy of the advance control imposed is a very efficient one in case of a FCC heat integrated plant.

Keywords: fluid catalytic cracking, heat integration, model predictive control, dynamic state.

INTRODUCTION

Due to its complexity the interest of solving problems related to the FCC process is worldwide spread. There has been a continuous effort to improve the efficiency and yield of the FCC unit during the time. There are many articles that present the problem of the FCC process modeling, simulation and control the most significant being [1] - [12] and just a few articles in which the study is related to the problem of energy integration [15].

The FCC process description is very difficult due to several reasons: the complexity of the chemical reactions mechanism, complex hydrodynamics, strong interaction between the operation of the main reactor and of the regenerator and due to the operation constrains imposed by the new HEN.

The aim of our research was to develop a MPC control scheme capable to control the FCC plant with the new HEN design obtained in a previous

[a] *Universitatea Babeş-Bolyai, Facultatea de Chimie şi Inginerie Chimică, Str. Kogălniceanu, Nr. 1, RO-400084 Cluj-Napoca,Romania, mmorar@chem.ubbcluj.ro*

work [13], [14]. The implementation of a MPC control scheme can provide a high stability of the plant knowing that the heat integration induces more instability in the process.

For reaching the purpose of this research, a FCC plant dynamic simulator was build in Aspen HYSYS using real industrial data related to material fluxes, temperatures, pressures, equipments size and geometry, etc. The data have been provided by a Romanian refinery.

MODEL PREDICTIVE CONTROL

The Model Predictive Control (MPC) is one of the most used advanced control technique in process control. In 1980s it was developed to meet the specialized control needs of power plants and petroleum refineries (ex. [15], [16]). Nowadays, can also be found in a wide variety of application areas like: chemicals, food processing, automotive, aerospace, metallurgy, pulp and paper, etc.

As it can be seen in Figure 1, the main idea of MPC is to choose the control action by repeatedly solving an optimal control problem. Therefore, MPC is based on iterative, finite horizon optimization of a plant model. At time "t", the current plant state is analyzed and a "cost minimizing" control strategy is computed, using a numerical minimization algorithm (Euler-Lagrange numerical method), for a very short time horizon in the future.

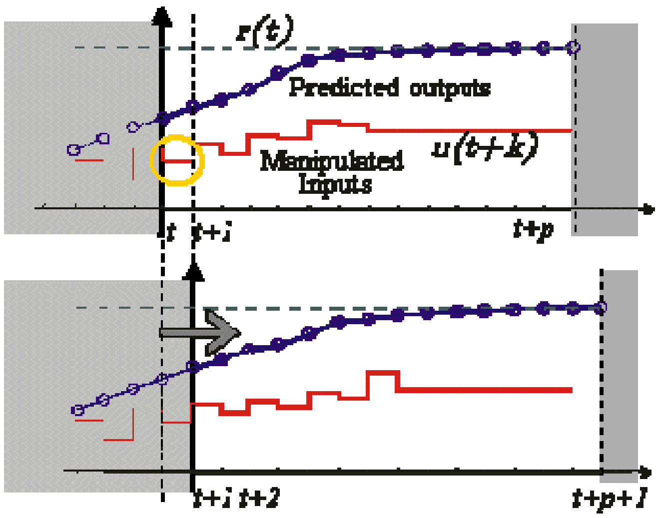

Figure 1. The MPC finite horizon

The MPC is also called receding horizon control because the prediction horizon keeps changing forward.

In Figure 2 the basic structure of MPC is presented. The model predictive control algorithm uses the models and current plant measurements in every moment in time to calculate the future moves for the safety.

Consequently, the process model plays an important role in the controller. The chosen model must be able to catch the process dynamics in order to predict the future outputs with high precision. Also, the model must be simple to be implemented and understood.

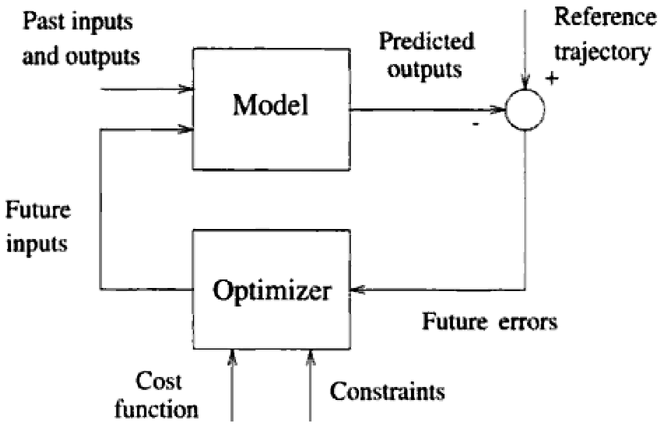

Figure 2. The basic MPC structure [16]

There are many types of models used in different formulations such as truncated impulse response model, step response model, state space model, transfer function model, etc. A typical MPC contains the behavior of multiple Single Input Single Output (SISO) controllers and de-couplers and uses a process model like first order model or step response model.

Comparing with the PID controller, the MPC controller main difference is represented by the necessity to build the process model that has to be implemented in the MPC controller structure.

THE DEVELOPMENT OF THE MPC CONTROLLER

The dynamic simulations performed revealed that only 5 controllers of PID type are necessary in order to obtain a proper stability of the FCC column. The stability of the column is reflected in the quality of heat transfer through the HEN. The development of the MPC controller for controlling the FCC column is based on the data provided by those 5 controllers that were able to settle the FCC column behavior at normal functioning conditions. The PID control scheme of the FCC column is presented in Figure 3.

MPC controller from the Aspen HYSYS object palette allows implementing up to a maximum 12 pairs of controlled variables (CV) and manipulated variables (MV). At the moment, the only MPC Control Algorithm that is implemented in the Aspen HYSYS software is the unconstrained MPC. This algorithm does not consider constraints on either controlled or manipulated variables.

In this case, 5 inputs (process /controlled variable – PV/CV) and 5 outputs (output target objects/manipulated variable – OP/MV) MPC is developed.

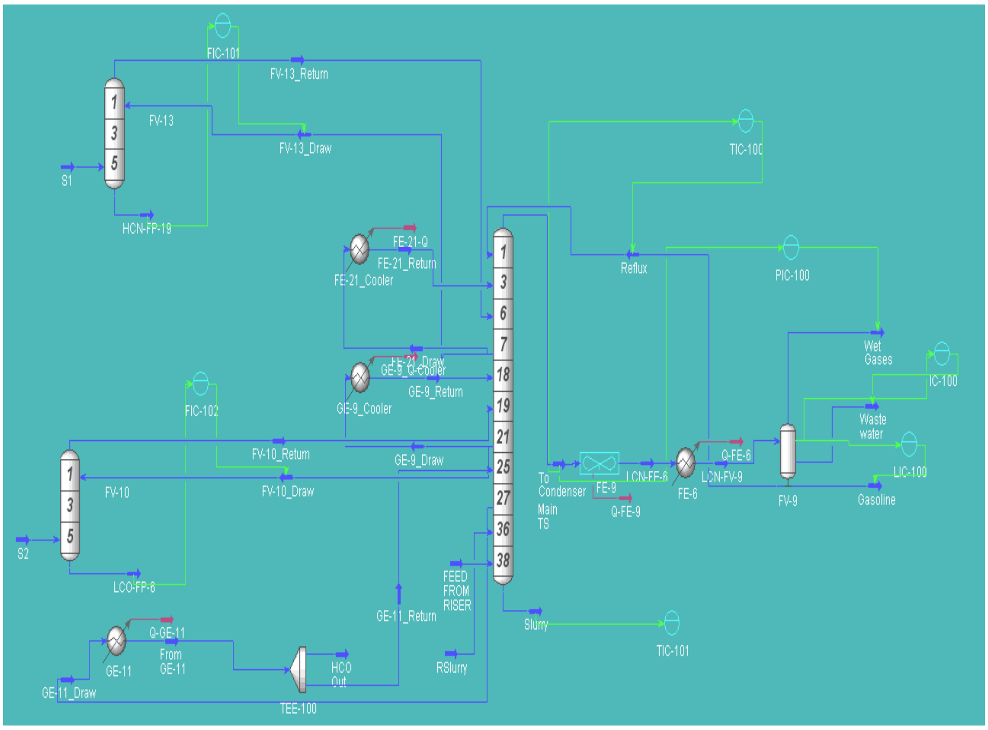

Figure 3. The PID control scheme of the FCC column

The controlled and the manipulated variables are summarized in Table 1.

Table 1. MPC control scheme selected variables

Controlled Variable		Manipulated Variable	
CV1	To Condenser stream temperature	**MV1**	Reflux flow rate
CV2	Condenser Liquid % Level	**MV2**	Gasoline flow rate
CV3	FV-13 bottom HCN flow rate	**MV3**	FV-13 HCN feed flow rate
CV4	FV-10 bottom LCO flow rate	**MV4**	FV-10 LCO feed flow rate
CV5	Slurry temperature	**MV5**	Column recycle slurry flow rate

"To Condenser" stream represents the top column output stream that passes two heat exchangers and enters in the condenser. The FV-10 and FV-13 (see Figure 3) are the side strippers of the FCC column. The stream which represents the bottom product of FV-13 (HCN stripper) is named HCN-FP-19. The stream which represents the bottom product of FV-10 (LCO stripper) is named LCO-FP-6. The Slurry stream is the bottom product of the FCC column.

Having all these specified the way of the MPC strategy implementation is presented as follows.

The Aspen HYSYS has implemented two options to specify the MPC controller model: the Step response model and the First order model. The model used in this case, for the MPC is the First Order model.

The first order model implementation is possible if the process gain (Kp), the process time constant (Tp) and the delay are known. With these process parameters it is possible to obtain the step response matrix necessary to build the internal MPC model.

The identification of those parameters could be realized by developing step response tests for each manipulated variable using the implemented PID control scheme from the dynamic state model. Consequently, a +10% step was used for each manipulated variable and the effect on the controlled variable was registered.

The step response tests provided the process gains, the time constants and the delays necessary for building the MPC controller. These parameters are presented in Table 2.

Table 2. First order model process parameters

MV / CV	MV1	MV2	MV3	MV4	MV5
CV1	Kp = -2.35 T = 2.56 τ = 0	Kp = 0.567 T = 15.66 τ = 0	Kp = -1.5182 T = 45.83 τ = 6.66	Kp = -1.5526 T = 77.083 τ = 0	Kp = -0.9611 T = 16.183 τ = 0
CV2	Kp = -0.15685 T = 3.033 τ = 0	Kp = -0.312 T = 8.33 τ = 0	Kp = -0.068 T = 58.75 τ = 0	Kp = -0.0775 T = 58.583 τ = 0	Kp = -0.0738 T = 20.016 τ = 0
CV3	Kp = -0.2043 T = 29.03 τ = 0	Kp = 0.1147 T = 32.16 τ = 0	Kp = 1.41 T = 0.01 τ = 0	Kp = -0.061 T = 113.5 τ = 7.33	Kp = -0.00456 T = 42.316 τ = 0
CV4	Kp = -0.66 T = 34.783 τ = 0	Kp = 0.39 T = 24.08 τ = 0	Kp = 0.437 T = 21.75 τ = 1.583	Kp = 7.436 T = 0.01 τ = 0	Kp = -0.1452 T = 26.85 τ = 0
CV5	Kp = -0.3741 T = 14.06 τ = 0	Kp = 0.32 T = 12.33 τ = 0	Kp = 0.28 T = 8.916 τ = 0	Kp = 1.156 T = 6.5 τ = 0	Kp = -1.151 T = 4.28 τ = 0

The process parameters were used to determine the 5x5 MPC step response matrix necessary for implementation of the MPC controller internal model. The dimensions of the generated step response matrix are 50x25.

After establishing the MPC controlled and manipulated variables and after determining the internal MPC model, the MPC controller was implemented in the FCC heat integrated plant dynamic model.

RESULTS AND DISCUSSIONS

The MPC control performances were analyzed in order to verify if the MPC controller was properly developed.

In order to perform the MPC performance analysis, several diagrams were developed in which the colors established for the variables are: red – the setpoint, green – the controlled variable, blue – the manipulated variable.

During the performance analysis it was observed that the MPC controller can not entirely stabilize the LCO-FP-6 flow rate, but the control action is good enough in order to have a proper operation of the HEN. The Figure 4 shows the behavior of the LCO-FP-6 flow over the time.

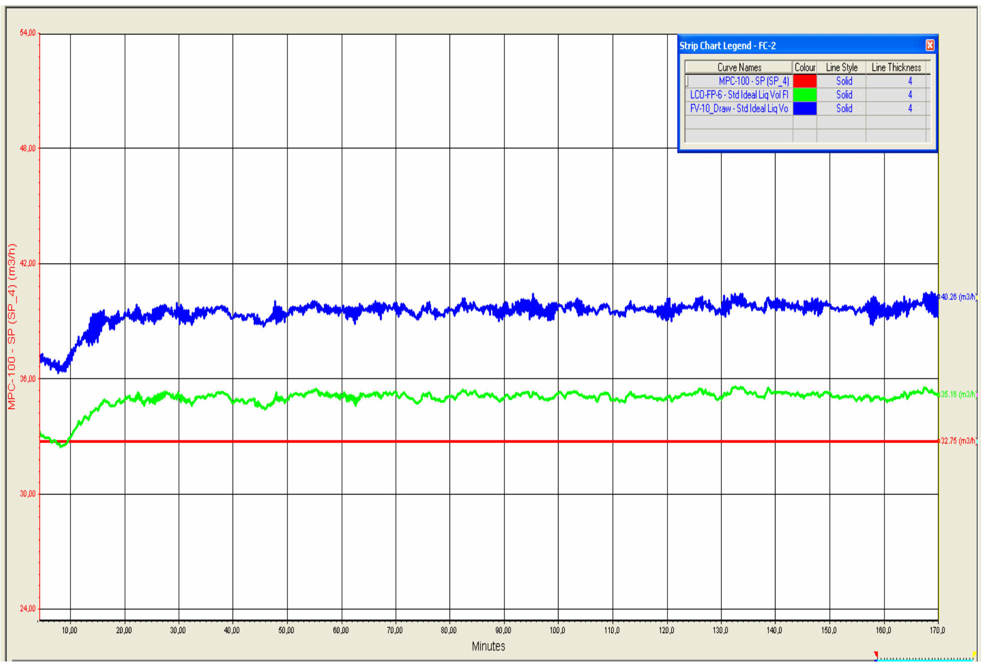

Figure 4. The performance of the LCO FP-6 flow control

As it can be seen in Figure 5, despite the LCO-FP-6 flow variation after stopping the noise amplification, the HCN-FP-19 flow control, which was the most affected by the noise, presents a good behavior and time stability. The HCN-FP-19 stream represents the bottom product stream of the FV-13 side-stripper which enters in the FE-24 heat exchanger.

The MPC control could reach the setpoint from industry for the Slurry temperature meaning that the controller model was successfully developed.

Further analysis regarding the HEN operation suggested that working at the set point value for the slurry temperature destabilizes the heat transfer of the HEN. This because working at the setpoint the quantity of the slurry entering in the HEN is not sufficient enough to sustain an efficient heat transfer as it was imposed in order to have a successfully heat integrated plant. This problem appears, as has been explained before, due to the geometrical data implemented in the FCC column model. Those data are not entirely the one that the real FCC column has. In order to solve the problem with the optimum heat transfer in the HEN, the system was operated near the value of the setpoint which assure the quantity of the slurry needed for an optimum operation of the heat exchangers. In these conditions the slurry temperature is approaching the setpoint and is stabilized at the value of 352^{0}C as can be seen in Figure 6.

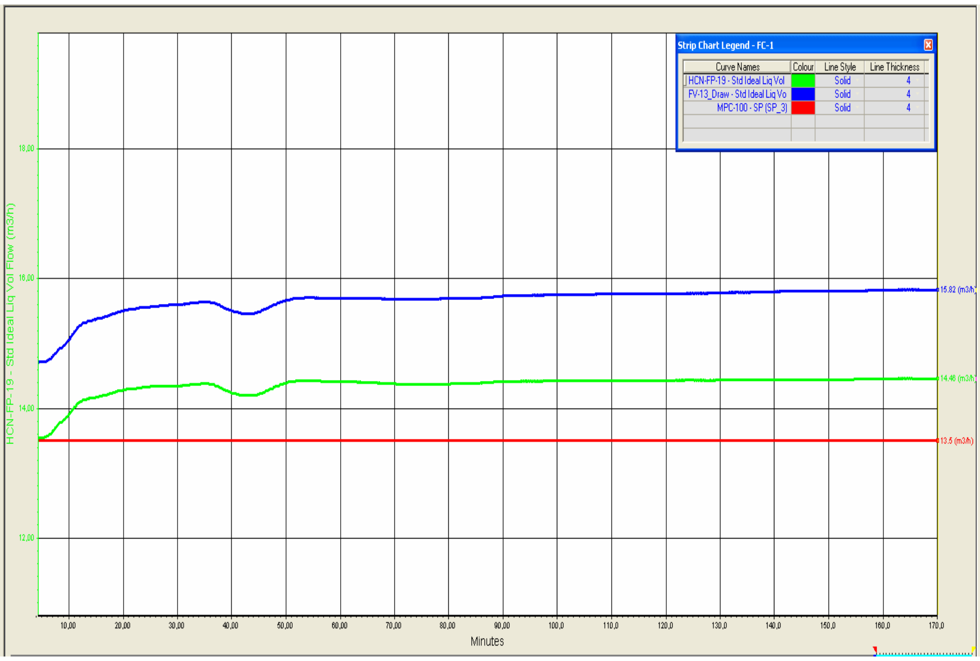

Figure 5. The performance of the HCN-FP-19 flow control

Beside the negative impact on the heat exchangers, the decreasing of the RSlurry flow rate presents another inconvenience. The role of the RSlurry stream in the column is to cool the effluent vapor stream from the riser that enters at the bottom of the column. If the cooling is not sufficient the column fractionation efficiency decreases considerably. The quality of the column products is compromised. Therefore, it is vital to have sufficient RSlurry quantity to handle the vapors proper cooling.

Consequently, the RSlurry flow control valve was set to the following values: 0% opening – 200.1 m^3/h and 100% opening – 346.8 m^3/h.

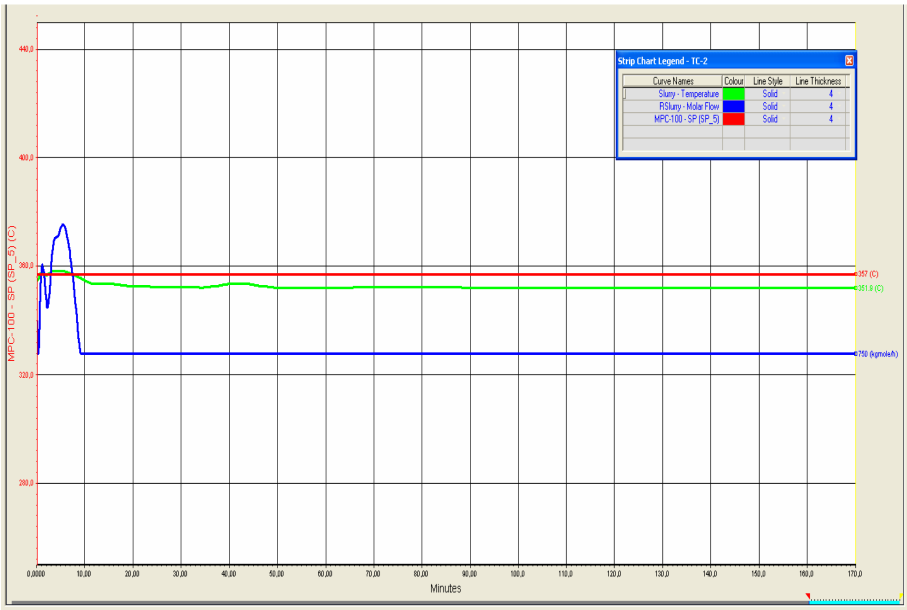

Figure 6. The performance of the Slurry temperature control

However, it is known from the real plant that the range of temperature for the slurry stream, in order to have a proper operation, is 345 – 360^{0}C. This means that the obtained value of the slurry temperature (352^{0}C) is an appropriate one. In these conditions the approach adopted is correct. The goal of maintaining an optimum heat transfer in the HEN has been realized. Another observation is that even if a lower temperature than the setpoint of the slurry stream is used this doesn’t affect the heat transfer in the HEN.

Comparing with the PID control, the MPC provides better results in controlling the temperature of the top output stream of the column (To Condenser). The PID temperature controller – TIC-100 achieves the set point after 30 minutes from starting the simulation. Meanwhile, as it can be seen in Figure 7, the MPC controller needs only 10 minutes from the simulation start to attain the setpoint.

Also, better control performance can be observed in controlling the liquid percent level of the FV-9 condenser. The MPC control has faster response than the PID level controller LIC-100 which follows the set point line after 40 minutes. The MPC controller needs only 25 minutes to bring the level at the setpoint. The MPC liquid level control performance is presented in Figure 8.

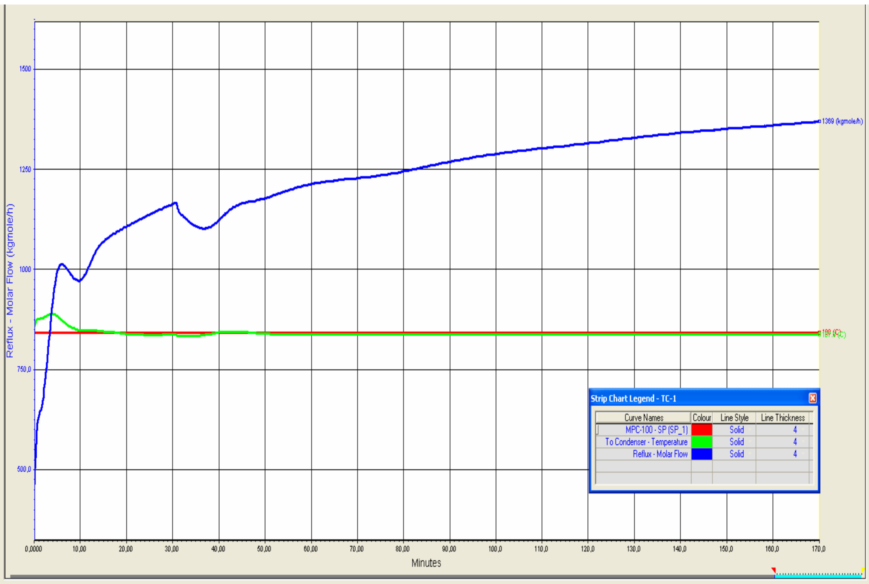

Figure 7. The performance of the temperature control of the column top product

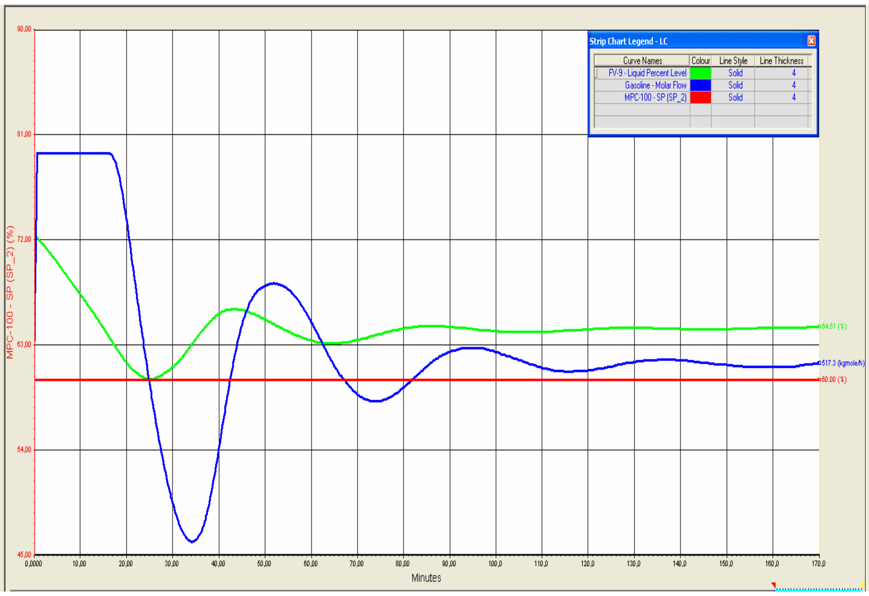

Figure 8. The performance of the condenser liquid percent level control

The liquid level from the condenser can be exactly maintained at 60% neither by PID controller nor by MPC controller because it is dependent on the column calculation of the output streams composition. The level reaches the steady state at around 65%.

The performance analysis of the MPC controller revealed that this controller is very efficient in case of the FCC heat integrated plant model proposed in this research.

CONCLUSIONS

In the last years, the model predictive control (MPC) has become a preferred control strategy for a large number of processes. This advanced control method was developed to meet the specialized control needs especially of the petroleum refineries.

The MPC can decrease the operating costs with approximately 2% - 6% [17] of the existing operating cost related to the real PID control scheme of a real plant because a MPC controller is capable to maintain the variation of the controlled variables much closer to the setpoint than PID controllers. Therefore, the plant is exploited at its maximum capacity.

Comparing with the PID controller, the MPC controller presents the necessity to build a process model. Many types of models such as truncated impulse response model, step response model, state space model, transfer function model, etc. can be used.

Based on the control demands of the heat integrated FCC plant and on the results obtained with PID control strategy, 5x5 MPC controller was developed using the First Order model.

In this case the estimation of model states and parameters is critical. The successful implementation of the MPC controller depends on them.

The identification of the process parameters that are needed for the first order model development (the process gain - Kp, the process time constant - Tp and the delay) was realized by developing step response tests for each manipulated variable.

After the parameter were calculated they were used to build the 5x5 MPC step response matrix necessary for determining the MPC controller internal model. In this way arise a new MPC controller capable to handle the control of the FCC column. The developed MPC controller results enable to establish that the strategy of the advanced control imposed is a very efficient one in case of a FCC heat integrated plant.

Regarding the MPC controller development, the novelty is using an easy method for the internal MPC model development. This method is fast and very useful in implementing an advanced control scheme at industrial scale. The step response tests in the real plant are insecure and costly especially in a continuous process. Any kind of these step control tests are able to compromise the quality of the products and to destabilize the industrial plant followed by undesirable incidents and costs.

LIST OF ABREVIATIONS

CV	controlled variable
FCC	fluid catalytic cracking
HCN	heavy cat naphta
HEN	heat exchanger network
LCO	light cycle oil
MPC	model predictive control
MV	manipulated variable
OP	output target object
PID	proportional-integral-derivative
PV	process variable
SISO	single input single output

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