Application of Predictive Control techniques to the Synthesis Section of a Urea Plant

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Abstract
This document presents the results of a research project oriented to evaluate the benefits of applying MPC techniques to the synthesis section of a urea plant. In the first part of this work, the research efforts were focused on improving the performance of a multivariable controller called MUVAC, which controls the reactor level and the ratio between ammonia (NH_3) and carbon dioxide (CO_2) at the outlet of the urea reactor. In the second part, the research efforts were concentrated on designing and implementing an MPC strategy for controlling and maximizing the production.

Key words: Predictive control, Process Control, Chemical Industry, Multivariable systems, P controllers, PID controllers.

1 Introduction

Model predictive control (MPC) has been widely accepted industrially during the last decades, mainly because of its ability to handle constraints explicitly and the natural way in which it can be applied to multivariable processes [10]. For processes with strong interactions, MPC can offer a substantial performance improvement compared with the traditional single-input single-output control strategies. In this paper we present the results obtained when MPC techniques were used to control the synthesis section of a Urea Plant based on Stamicarbon technology.

In order to test the different control schemes, a complete urea model has been used. The model contains more than 2000 system parameters and 5000 differential and algebraic equations for calculating more than 5000 system variables [12][13]. Such a model was developed by *Protomation* and it has been implemented in a software package called *Process Studio*, which was created by the same company. It is important to observe that along the study we only have accessed to the simulation facilities from an input-output perspective since the equations of the rigorous model remained hidden. This situation resembles industrial reality where under normal conditions no model of the process is available. Hence multivariable identification techniques have been used to derive local linear models used by the different MPC controllers implemented along the study. The urea model implemented in *Process Studio* is quite accurate with respect to industrial plants and is widely trusted in industry; in fact, the urea simulator is used to train operators and to tune the control systems of some Stamicarbon urea plants during the commissioning phase. The model is based on a standard Stamicarbon 3200 T/d plant design with a pool condenser. All basic information for building an actual urea plant (e.g. PFDs', mass and energy balance, P&ID's, logic

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diagrams, equipment data sheets) has been used to develop such a model. Stamicarbon provided all the information necessary including the latest thermodynamic property and kinetic function libraries which are the core of the simulator. The project team who developed the simulator consisted of Stamicarbon experts on simulation and design of the urea process, experts on production, commissioning and start-up of urea plants, and Protomation experts on dynamic simulation and simulator development [12].

For the Implementation of the MPC strategies, an industrial MPC controller developed by *IPCOS* has been used. This MPC controller is called *INCA Suite* and it uses the OPC (OLE for Process Control) technology to communicate with the DCS (Distributed Control System) [11]. Since *Process Studio* and *INCA Suite* work with the OPC standard, the urea model that is running in *Process Studio* can be controlled by the *INCA Suite*.

This paper presents a general description of the urea process in section 2. This section includes also a detailed description of the synthesis section. Section 3 presents two MPC schemes that were designed in order to improve the behavior of a traditional controller called MUVAC, which controls the reactor level and the NH_3/CO_2 ratio at the outlet of the urea reactor in the synthesis section. Section 4 shows an MPC scheme for controlling and maximizing the production of the synthesis section and finally section 5 presents the main conclusions.

2 Description of the Urea Process

Urea (NH_2CONH_2) is produced at industrial scale by the reaction between ammonia (NH_3) and carbon dioxide (CO_2) at high pressure (13–30MPa) and high temperature (170 –200 ºC) [5]. The chemical reactions that are involved in the urea formation are the following:

$$2NH_3 + CO_2 \rightarrow NH_2COONH_4 + (\text{heat}) \quad (1)$$

$$NH_2COONH_4 \Leftrightarrow NH_2CONH_2 + H_2O - (\text{heat}) \quad (2)$$

In reaction (1), ammonia and carbon dioxide are converted into ammonium carbamate (NH_2COONH_4). This reaction is highly exothermic and fast, and goes essentially to completion under normal industrial processing conditions. In the reaction (2), ammonium carbamate dehydrates to produce urea and water. Unlike reaction (1), reaction (2) is slow and endothermic and therefore it needs long residence time to reach equilibrium (usually it does not reach thermodynamic equilibrium under processing conditions) [3][4]. Since more heat is produced in the first reaction than it is consumed in the second reaction, the overall reaction is exothermic.

Complete conversion of ammonia and carbon dioxide into urea and water is limited. An important differentiator between processes is the way that these non-converted materials are handled [3]. So, we can find different urea processes in industry. These processes are called once-through, partial recycle and total recycle. In the total recycle process, which is widely used, all the ammonia leaving the synthesis section is recycled to the urea reactor and the overall conversion of ammonia to urea reaches 99%. Stamicarbon and Snamprogetti processes are the most common examples of such kind of process [5].

The urea simulator developed by *Protomation*, models a "Stamicarbon Urea CO_2 Stripping Process" [2], which is a total recycle process. The following subsection presents an overview of this type of process and detailed description about the synthesis section.

2.1 Total Recycle CO_2 Stripping Urea Process

In a Total Recycle CO_2 Stripping Urea Process, the following sections can be distinguished (see figure 1):

- Synthesis
- Recirculation
- Evaporation and Finishing Technique
- Waste water treatment (Desorption and Hydrolysis)

In the CO_2 stripping process, ammonia and carbon dioxide are fed directly to the *synthesis section* where

the urea formation takes place according to reactions (1) and (2). In the *recirculation section* the unconverted ammonia and carbon dioxide are removed from the urea solution. These unconverted materials are condensed, forming a carbamate solution which subsequently is recycled to the *synthesis section*. The urea solution after the removal of the unreacted carbamate, ammonia and carbon dioxide is sent to the *evaporation section* where the water in the urea solution is removed. Afterwards, the urea melt is sent to the *granulation unit* (finishing technique) where the formation of the urea granules takes place. Finally, the process condensate that is coming from the *evaporation stage* is sent to the *waste water treatment* section, where the residual ammonia and carbon dioxide present in the process condensate are removed and sent back to the *recirculation stage*. The treated water from this stage can be used in other sections of the process. The exhaust gases (inert gases, NH_3, CO_2 and H_2O) from the different sections are washed in the absorbers in order to minimize the ammonia emissions.

2.2 Description of the Synthesis section

The block diagram of the Synthesis section is shown in figure 2. In order to achieve maximum urea conversion at synthesis conditions (pressure ≈ 140 bar and temperature ≈ 183 ºC), a molar ratio (NH_3/CO_2) of ammonia to carbon dioxide of 3 at the outlet of the urea reactor is required [3].

Ammonia and carbon dioxide (with a small amount of air to prevent corrosion) are fed at a high pressure into the synthesis section. Therefore, a multi-stage compressor and a high-pressure pump are used for feeding the CO_2 and the NH_3 respectively. The compressor is driven by a steam turbine and the speed of the turbine is controlled by the amount of HP-steam supplied. A H_2 removal reactor located between the compression stages is used to remove the remaining hydrogen in the fresh CO_2 stream (the CO_2 usually originates from an associated ammonia plant and therefore nearly always contains hydrogen).

The urea formation takes place in the pool condenser and in the urea reactor. Since part of the ammonia and carbon dioxide are not converted to urea and water, the reaction mixture coming from the urea reactor is subjected to a stripping process (it occurs in the CO_2 Stripper), where carbon dioxide is used to remove the unreacted ammonia and carbon dioxide. The off-gases (NH_3 and CO_2) from the stripping process are introduced into the pool condenser together with the carbamate solution from the High-Pressure scrubber and fresh ammonia. The heat released by the formation of carbamate is recovered to generate low-pressure steam in the pool condenser. The gases leaving the reactor are fed into the High-Pressure Scrubber. In the scrubber the gases are washed with the carbamate solution from the low-pressure recirculation stage [3]. Thus a major portion of gases is recovered and returned to the pool condenser [6]. The inert gases and some ammonia and carbon dioxide coming from the scrubber are washed in a low-pressure absorber, thereby minimizing ammonia emissions.

3 Control of the Reactor Level and the NH_3/CO_2 Ratio at the reactor outlet

As it was mentioned in section 2.2, the NH_3/CO_2 ratio at the reactor outlet must be 3 in order to achieve maximum urea conversion at the nominal operating conditions. This constitutes a very important control goal in the synthesis section. In addition, the NH_3/CO_2 ratio presents a considerable degree of coupling with the reactor level demanding the use of a multivariable control strategy to keep these variables simultaneously at a desired setpoint. Traditionally a multivariable controller called MUVAC has been implemented in order to control simultaneously the reactor level and the NH_3/CO_2 ratio.

Figure 3 shows the MUVAC control scheme of the synthesis section. The MUVAC controller consists of a pair of PID controllers (LIC201 and AIC202) with a static decoupling system. From figure 3 it is clear that the MUVAC has two manipulated variables, the setpoints of the NH_3 (FRC106) and CO_2 (FIC108) flow rate PID controllers, and two controlled variables, the reactor level and the NH_3/CO_2 ratio at the outlet of the reactor. In the MUVAC strategy the level changes are associated with the changes in the total feeding flow, and the NH_3/CO_2 ratio changes are associated with the changes in the proportion of the NH_3 and CO_2 feeding flows. The idea of this control strategy is to correct the deviations of the reactor level through LIC201 and the static decoupler without influencing the NH_3/CO_2 ratio, and to correct the NH_3/CO_2 ratio deviations through AIC202 and the static decoupler without affecting the reactor level [7].

However, in practice the MUVAC controller presents an undesirable oscillatory behavior and in spite of the use of the decoupler there is still a strong interaction between the reactor level and the NH_3/CO_2 ratio. The reason of such interaction is because the decoupler does not take into account the dynamics of the process. Finally, there is a large dead time (around 18 minutes) between the feeding flows and the measurement of the NH_3/CO_2 ratio. This is probably the main cause of the oscillatory behavior of the MUVAC controller, since its PIDs are not designed to handle such dead time.

To improve the performance of the control system, two control schemes based on the MPC technique were designed [1]:

- Scheme 1: an MPC without feed-forward action in the PID Loops.
- Scheme 2: an MPC with feed-forward action in the PID Loops.

In the following subsections a detailed description of each control scheme will be given and the results of applying such control strategies to the urea process will be presented.

3.1 MPC without feed-forward action in the PID Loops

Similar to the MUVAC case, the MPC controller should manipulate the ammonia and carbon dioxide feeding flows in order to control the NH_3/CO_2 ratio and the reactor level. Some preliminary tests were carried out in order to determine how the controlled variables (level and NH_3/CO_2 ratio) are affected by the manipulated variables (feeding flows). In such tests it was clear that the reactor level shows an integral behavior, once a step increment is made in one of the feeding flows (or in both), the level will rise until it reaches the top of the reactor. In addition, it was observed that changes in the setpoint of the NH_3 flow rate controller have more impact on the level than changes in the setpoint of the CO_2 flow rate controller.

Since *INCA Suite* (the industrial MPC controller) can deal only with stable processes, it was necessary to stabilize the reactor level by means of a proportional controller (LIC201). The manipulated variable of this proportional controller is the setpoint of the NH_3 flow rate controller. The NH_3 flow rate was chosen as manipulated variable because it has more influence on the level than the CO_2 flow rate. Additionally, this new topology of PID loops keeps the NH_3/CO_2 ratio stable.

The schematic diagram of the implemented control system is shown in figure 4. For the MPC controller, the manipulated variables are the setpoint of the controller used to stabilize the level (IRL – Internal Reactor Level) and the setpoint of the CO_2 flow rate controller (FCO2). The controlled variables are the reactor level and the NH_3/CO_2 ratio.

Using MPC to generate the setpoints of the PID controllers demands a good tuning of the PID controllers. This condition is important in order to have a good performance with a model of a reasonable low order. Therefore, the PID controllers that are manipulated directly (LIC201, FIC108) or indirectly (FRC106, SIC101) by the MPC controller were retuned using a software package called RaPID [8]. In RaPID, the system is identified using the data acquired from the process. Subsequently, the control objectives based on engineering specifications are defined and finally an optimization algorithm is executed in order to find the optimal PID parameters that fulfill all these requirements [8]. In general terms the PID controllers were adjusted to have a settling time as short as possible with a minimum overshoot.

In order to obtain the mathematical model (in this case a multivariable linear model) that allows the MPC to predict the future behavior of the process, multivariable identification techniques were used. In the identification experiment, step changes were used rather than PRBS (Pseudo Random Binary Signal) signals due to the slow dynamics of the process and the long duration of the simulations (the simulator was only 5 times faster than the real process). The sampling time used in the data acquisition process was 60 seconds. It was selected by dividing the faster time constant of the process by 20. The order of the model was selected by means of the singular values of the Hankel Matrix, the estimated order was 4. The multivariable model was found using the sub-spaces method N4SID [9]. To validate the model, it was simulated and the predicted outputs were compared with the process outputs. The mean squared error (mse) between the process outputs and the model outputs were calculated to measure the quality of the

model.

For this MPC controller, a control horizon of 50 samples was selected. The prediction horizon was chosen in the following way: Prediction Horizon = Control Horizon + Largest Settling Time = 300 samples. The tuning of the MPC Controller was done by means of a trial-and-error procedure.

INCA Suite is a generally applicable supervisory Model Predictive Controller [11]. One important feature of this MPC controller is that it provides a weighted (least-squares trade-off) and hierarchical (ranking) optimization. Controller objectives can be ranked according to their relative importance. In this way, the controller will start to fulfill the objectives in order of importance. In addition, we can have controller objectives (zone violations, setpoint errors and setpoint moves of manipulated variables) with the same rank. In this case, a least-squares trade-off among these objectives is performed. This is done under the condition that the higher ranked objectives are not violated.

The control objectives of this MPC controller were ranked in the following way: the highest priority of the controller is to keep the manipulated variables (MV's) within the operating range (this is a hard constraint). The second objective for the controller is to maintain the controlled variables (CV's) within their operating limits. The third goal for the controller is to keep NH_3/CO_2 ratio on its desired value and the fourth objective is to maintain the reactor level on its setpoint. The lowest priority goal is to bring the MV's to their ideal resting value (this is the value where the MV should go to, when the higher control objectives has been satisfied), if it is possible.

In order to test the performance of the control system the following tests were designed:

- Test 1: In this test the setpoint of the NH_3/CO_2 ratio is changed from 3 to 3.1, while the setpoint of the reactor level is kept on 50%.
- Test 2: The setpoint of the reactor level is changed from 50% to 30%, while the setpoint of NH_3/CO_2 ratio is kept on 3.
- Test 3: The position of the valve HIC201 is changed from 56 %(nominal position) to 53%. As this valve has a direct impact on the reactor level and on the residence time of the reactor, the change of this valve position represents a big perturbation for the system. The setpoints for the reactor level and the NH_3/CO_2 ratio are kept on 50% and 3 respectively.

In addition, the carbamate solution coming from the recirculation stage constitutes one of the main perturbations for the synthesis section. Therefore, this perturbation is present in all the tests mentioned before.

In the subsequent plots of the manipulated variables, the NH_3 flow rate is presented rather than IRL (setpoint of the LIC201). This is because at low level, the LIC201 controller manipulates the NH_3 flow rate to stabilize the reactor level. Figure 5 presents the results of test 1. Unlike the MUVAC controller, in the MPC controller the reactor level does not have oscillations. In addition, in the MPC case the NH_3/CO_2 ratio presents a smaller overshoot (peak of 0.01) than in the MUVAC case (maximum peak of 0.02). However, the rise time for the NH_3/CO_2 ratio with the MPC controller was longer. The control efforts of the MPC controller are "softer" compared with the control actions of the MUVAC controller.

The results of test 2 are shown in figure 6. For a setpoint change in the reactor level, the MPC outperforms completely the MUVAC controller. Practically there is not overshoot for the reactor level, and for the NH_3/CO_2 ratio. The maximum peak over the setpoint is 0.005 (0.038 for the MUVAC) and under the setpoint is 0.011 (0.031 for the MUVAC), which are smaller than the peaks observed when the MUVAC is used. Once more, the control actions in the MPC controller are smoother in comparison with the control efforts of the MUVAC.

Finally, figure 7 presents the response of the system when the position of the valve HIC201 was changed from 56% to 53%. In this test the MUVAC controller keeps the NH_3/CO_2 ratio very close to 3 (maximum peak of 0.025 over the setpoint) whereas the MPC controller has big peak of 0.1324 under the setpoint. For the reactor level, the MPC keeps it closer to 50% than the MUVAC controller. Since the most important controlled variable is the NH_3/CO_2 ratio, it is clear that the MUVAC was superior to this MPC since the MUVAC kept the NH_3/CO_2 ratio closer to its desired value.

In general terms, the MPC controller without feed-forward actions in the PID loops was much better than the MUVAC controller when the setpoints of the controlled variables were changed, but it was not as good as the MUVAC for rejecting perturbations.

3.2 MPC with feed-forward action in the PID Loops

In order to improve the disturbance rejection performance of the previous MPC controller, a new structure of PID control loops was considered. The key idea for this new scheme was to keep the proportion of the feeding flows constant (feed-forward action) when the reactor level is being changed. Only when a change in the setpoint of the NH_3/CO_2 ratio is required, this proportion will be changed temporarily until the NH_3/CO_2 ratio reaches its new steady state value. The schematic diagram of this new control system is presented in figure 8.

Once more, a proportional controller (LIC201) is used to stabilize the level through the ammonia flow rate controller (FRC106). The setpoint of the CO_2 flow rate controller (FIC108) is determined by the NH_3 flow rate multiplied by a constant *K* (feed-forward action) plus the output of the AIC202 proportional controller. The function of the AIC202 controller is to stabilize and to allow setpoint changes of the NH_3/CO_2 ratio at the reactor outlet. It is important to remark that if we remove the AIC202 controller and we change the constant *K* a little bit, the system will exhibit an integral behavior. In steady state conditions, the setpoint of the FIC108 is only determined by the FRC106 controller, since the output of the AIC202 proportional controller is approximately zero. The constant *K* was calculated by dividing the CO_2 flow rate over the NH_3 flow rate in steady state conditions when the output of the AIC202 controller is zero.

With this new PID configuration, the manipulated variables for the MPC controller would be the setpoint of the LIC201 controller (IRL –Internal Reactor Level) and the setpoint of the AIC202 controller (INC – Internal NH_3/CO_2). The controlled variables would be the reactor level and the NH_3/CO_2 ratio at the outlet of the reactor.

Following a similar procedure as the one applied to the previous MPC design; the new proportional controllers were tuned using RaPID, a new multivariable linear model (4th order) was found using the sub-space identification and the same validation criterion was used.

A control horizon of 50 samples was set and a prediction horizon of 200 samples was chosen applying the criterion used for the previous controller. The tuning of the MPC controller was performed by means of a trial-and-error process. The control objectives of this new MPC scheme were ranked as before.

In order to test this new MPC controller configuration, the systems was subject to the same tests performed to the previous MPC. Figure 9 shows the results of the test 1. It is clear that there is no overshoot for the NH_3/CO_2 ratio in the MPC case whereas a maximum peak of 0.02 is observed when the MUVAC is used. In addition, the NH_3/CO_2 ratio presents similar rise times in both controllers. The setpoint change of the NH_3/CO_2 ratio has a minor impact on the reactor level for the MPC case (maximum peak of 0.45% over the setpoint), but a large impact for the MUVAC case (maximum peak of 8.73 % over the setpoint). For the MPC case, the changes in the feeding flows are softer than the changes observed in the MUVAC case.

The results of the test 2 are presented in figure 10. Once more, the MPC controller outperforms completely the MUVAC controller. Practically there is no overshoot in the reactor level, and for the NH_3/CO_2 ratio the maximum peak over the setpoint is 0.0104 (0.038 for the MUVAC) and under the setpoint is 0.004 (0.031 for the MUVAC). So, it is clear that the impact of the level changes on the NH_3/CO_2 ratio is smaller in the MPC case than in the MUVAC case. As in the previous test, the changes in the feeding flows are smoother in the MPC case than in the MUVAC case.

Figure 11 shows the results of the test 3, where the position of the valve HIC201 is changed from 56% to 53 %. In this test the MPC controller keeps the NH_3/CO_2 ratio a little bit closer to 3 than the MUVAC controller. The impact of the disturbance on the reactor level is minor (max peak of 2.8%) in the MPC

case, but really significant (max peak of 12.36 %) in the MUVAC case. In order to test even further the disturbance rejection capability of the controllers, a more severe test was performed. In this test, the position of the HIC201 valve was changed from 56% to 48% (this represents a change three times larger than the change performed in test 3). The results of this test are presented in figure 12. From this figure it is clear that the MPC is far superior compared to the MUVAC controller. For the MPC controller the maximum peaks for the NH_3/CO_2 ratio and the reactor level are 0.027 and 8.21% respectively, whereas for the MUVAC controller these peaks are 0.092 and 40.37%. With the MUVAC controller the high-level alarm was triggered when the reactor level rose beyond 80%.

It is clear that the MPC with feed-forward action in the PID loops has outperformed the MUVAC for rejecting disturbances. There is no doubt that the incorporation of a feed-forward action in the PID loops has improved the performance of the previous MPC controller. In conclusion, unlike the previous MPC, the MPC controller with feed-forward action in the PID loops is better than the MUVAC controller in setpoint tracking and disturbance rejection.

4 Control of the production of the Synthesis Section through an MPC Controller

So far, the two MPC control schemes discussed in previous sections aimed to solve the problem of stabilization of the synthesis section. The application of MPC offers opportunities beyond the stabilization, including the possibility to exploit the optimization to drive the plant towards the maximization of the production or the minimization of the use of energy. The control scheme shown in figure 13 has been designed to maximize the production of the synthesis section of a urea plant [1]. The main goal of this control system is to maximize the throughput of the synthesis section by maximizing the feeding flows. This objective is achieved while keeping the NH_3/CO_2 ratio, the reactor level and the reactor pressure on their desired values and the reactor temperature within a defined operating range.

In this control scheme the production of urea can be set by selecting the setpoint of the NH_3 flow rate controller (FRC106). In this way, if the setpoint of the NH_3 flow rate controller is augmented, the amount of urea produced will be increased and vice versa. Since the FRC106 controller is used to set indirectly the throughput of the synthesis section, it will not longer be used to stabilize the reactor level. Instead, the valve HIC201 was selected as the manipulated variable, since it has a direct impact on the reactor level. This valve also influences the reactor residence time (conversion) and traditionally it has been operated manually in the industrial urea plants; with this control scheme such valve is manipulated directly by the LIC201 controller and indirectly by the MPC. The MPC controller can manipulate such valve in the best way because it has process knowledge embedded in the mathematical model used by the controller.

This control system also maintains the feed-forward action implemented in the MPC described in section 3.2 in order to maintain a good disturbance rejection for the NH_3/CO_2 ratio. Therefore the setpoint of the CO_2 flow rate controller (FIC108) is determined by the NH_3 flow rate multiplied by a constant K (feed-forward action) plus the output of the AIC202 proportional controller. Once more, the function of the AIC202 controller is to stabilize and to allow setpoint changes of the NH_3/CO_2 ratio at the reactor outlet, as it was explained in section 3.2.

For the MPC controller, the manipulated variables are the setpoint of the LIC201 controller (IRL), the setpoint of the AIC202 controller (INC), the setpoint of the PRC902 (LPS – Low pressure steam controller), the setpoint of the PIC904 controller (HPS – High pressure steam controller), the position of the HIC202 valve (HIC202) and the setpoint of the FRC106 controller (FNH3 – NH_3 flow rate controller). The controlled variables are the reactor level, the NH_3/CO_2 ratio at the outlet of the reactor, the reactor pressure (PR201) and the reactor temperature (TI201).

The low pressure steam and the high pressure steam were selected as manipulated variables because they have a direct impact on the carbamate reaction and the stripper efficiency respectively. After some experiments, a significant impact of the position of the valve HIC202 on the reactor pressure was observed. Therefore the valve HIC202 was also taken into account in the MPC strategy. Finally, the NH_3 flow rate was selected as manipulated variable because this flow sets indirectly the throughput of the synthesis section. The main goal of this controlled scheme is to maximize the amount of urea produced in the synthesis section. Observe that there are more MV's than CV's this feature provides an extra freedom

for the system to search the optimal operating point.

Once more, RaPID was used to tune the controllers LIC201 and AIC202. Subspace multivariable system identification was used to derive the linear model used by the MPC. The selected model exhibits dynamics of 9th order.

The operating ranges of the manipulated variables of the MPC controller were chosen very carefully to avoid problems with the downstream sections of the urea plant. For example, if the low pressure steam is reduced below 3 barg, then the conductivity of the waste water in the waste water treatment section is increased beyond the maximum admissible value (12 μs/cm). If the low pressure steam is increased above 3.55 barg the reactor pressure can reach values above 150 barg together with the CO_2 line that is carrying the CO_2 from the compressor to the stripper. This is a severe operating constraint. The pressure in the reactor and in the CO_2 line must not exceed 150 barg. If the high pressure steam is decreased below 17 barg, the conductivity of the waste water will be affected. The maximum pressure for the High pressure steam is limited by the process itself to a maximum of 19.6 barg. Beyond such value the PIC904 controller (it controls the high-pressure steam) saturates. Additionally, if we make fast and large decrements of the pressure in the high pressure steam, then the pressure of the rectifying column (this device is located in the recirculation section) that is connected to the stripper output will increase beyond the safe limits (6 barg is the maximum operating pressure). If the position of the valve HIC202 is larger than 16%, then the pressure of the LP absorber will be out of control because its pressure controller will be saturated (the pressure will be larger than 3 barg which is the nominal pressure). In addition, if the position of the HIC202 valve is below 9% then the pressure of the reactor will be larger than 150 barg. Finally, it is important to mention that the ammonia flow rate must be increased in a slowly to avoid large increments in the pressure of the rectifying column.

A control horizon of 50 samples was selected and a prediction horizon of 470 samples was chosen for the MPC controller. The tuning of the MPC controller was performed by trial-and-error. The ranking of the control objectives of this MPC controller is described in the following lines. The highest priority of the MPC controller is to keep the MV's within their operating limits. The next priority is to maintain the CV's within the operating range in the following order of importance: NH_3/CO_2 ratio, reactor level, reactor pressure and finally reactor temperature. The next control objective is to keep the NH_3/CO_2 ratio on its desired value. The next control task is to maintain the NH_3 flow rate on its ideal resting value. As the throughput of the synthesis section is set by the ammonia flow rate, we can maximize such throughput by making large the ideal value of the ammonia flow rate. The subsequent control priorities are to keep the level, pressure and temperature (only if it is possible!) of the reactor on their setpoints. Finally, the last goal of the controller is to bring the IRL, INC, LPS, HPS and HIC202 to their ideal resting values (again if it is possible!). In order to avoid large deviations in the setpoint of the high pressure steam and the low pressure steam, the priorities of the LPS and HPS variables were made higher than priorities of the IRL, INC and HIC202 variables.

In order to evaluate the behavior of the designed MPC controller the ideal resting value of the NH_3 flow rate was changed from 77.2 t/h to 80.5 t/h. The setpoints (nominal values at full load) of reactor level, the NH_3/CO_2 ratio, the reactor pressure and the reactor temperature are kept on 50%, 3, 144.6 barg and 181.96ºC respectively. With this test the production of synthesis section is increased in a controlled way by manipulating the ammonia feeding flow. As it was mentioned before, the carbamate solution coming from the recirculation stage constitutes one of the main disturbances for the synthesis section, and it is present in this test.

Figure 14 shows the increment of the flow of urea solution leaving the synthesis section (throughput) during the experiment. At initial steady state conditions, the throughput of the synthesis section was 241.82 t/h of urea solution. Once the NH_3 flow rate was increased from 77.2 t/h to 80.5 t/h the new throughput of the process in steady state conditions was 253.63 t/h of urea solution. We have obtained an increment of 11.81 t/h in the throughput of the synthesis section.

The controlled variables and the manipulated variables of the MPC controller along the test are presented in figures 15 and 16. From figure 15 it is remarkable how the controlled variables are kept close to their setpoints. So, the maximum peak for the NH_3/CO_2 ratio is only 0.0053, for the reactor level is only 1.79 %, for the reactor pressure is 0.88 barg and for the reactor temperature is 0.3°C. From figure 16 we can

observe how the ammonia flow rate is changed by the MPC controller slowly from 77.2 t/h to 80.5 t/h. In this way the controller does not produce large pressure changes in the rectifying column of the recirculation section. In addition from figure 16 we can notice how the new steady state values of the low pressure steam and the high pressure steam are not far away from their initial values.

The control scheme presented in this section has been capable not only of achieving stabilization but also it has been successful in increasing the production. Although it is not shown in this paper, this control scheme can also perform setpoint changes for the NH_3/CO_2 ratio, the reactor level and the reactor pressure. Independent setpoint changes in the reactor temperature are not possible since the reactor temperature is linked to the reactor pressure.

5 Final comments and Conclusion

In this paper the results of applying MPC control techniques for controlling the synthesis section of a urea process have been presented.

Traditionally the control of the NH_3/CO_2 ratio at the reactor outlet and the reactor level has been done by a multivariable controller called MUVAC. The MUVAC controller is characterized by an oscillatory behavior mainly due to its limitations to handle the large dead time present in the NH_3/CO_2 ratio measurement and the high degree of coupling between the reactor level and the NH_3/CO_2 ratio.

In order to correct such problem a first MPC controller without feed-forward action in the PID loops was designed. The results were very promising for setpoint tracking. But the performance in disturbance rejection was not better than the one exhibited by the MUVAC. Consequently, a second MPC was designed incorporating a feed-forward action in the PID loops. This second MPC outperformed the MUVAC controller in all aspects, since it was much better in performing setpoint changes and rejecting disturbances.

A very important conclusion is that the structure of the PID loops that are manipulated by the MPC controller plays a very important role in the overall performance of the MPC strategy. Therefore such structure of PID loops must be selected using process knowledge.

In order to control and maximize the production of the synthesis section of a urea plant a new MPC scheme was proposed. The main goal of this control system was to maximize the throughput of the synthesis section by maximizing the feeding flows. In addition, with this control scheme the NH_3/CO_2 ratio, the reactor level, the reactor pressure and the reactor temperature were kept under control. The results obtained with this new control scheme were fairly good from the control point of view and from the production point of view. The MPC increased the throughput of the synthesis section keeping at the same time the NH_3/CO_2 ratio and the reactor variables under control. The increment of the production is only limited by the maximum feeding flows.

Finally, from this work it is clear that we can obtain substantial benefits by applying MPC techniques on the synthesis section of a urea plant.

Acknowledgments

The authors would like to acknowledge the support of IPCOS ISMC office, Rob Faessen from Protomation and Klement K. Riegman from Stamicarbon. Additionally this research was supported by: ● Research Council KUL: GOA-AMBioRICS, BOF OT/03/12, several PhD/postdoc & fellow grants; ● Flemish Government: - FWO: PhD/postdoc grants, G.0407.02 (support vector machines), G.0080.01 (Collective Behavior and Optimization), G.0211.05 (Nonlinear Systems), research communities (ICCoS, ANMMM); - IWT: PhD Grants; - Tournesol 2005; - Belgian Federal Science Policy Office IUAP P5/22 (“Dynamical Systems and Control: Computation, Identification and Modelling”) . Dr. Joos Vandewalle is a full professor at the Katholieke Universiteit Leuven, Belgium.

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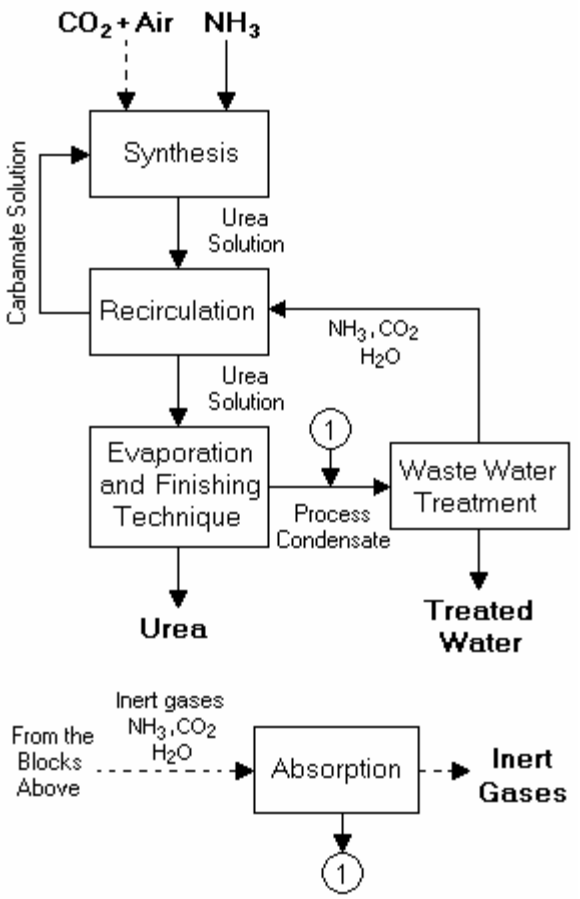

Fig. 1. Block diagram of a Total Recycle CO_2 Stripping Urea Process.

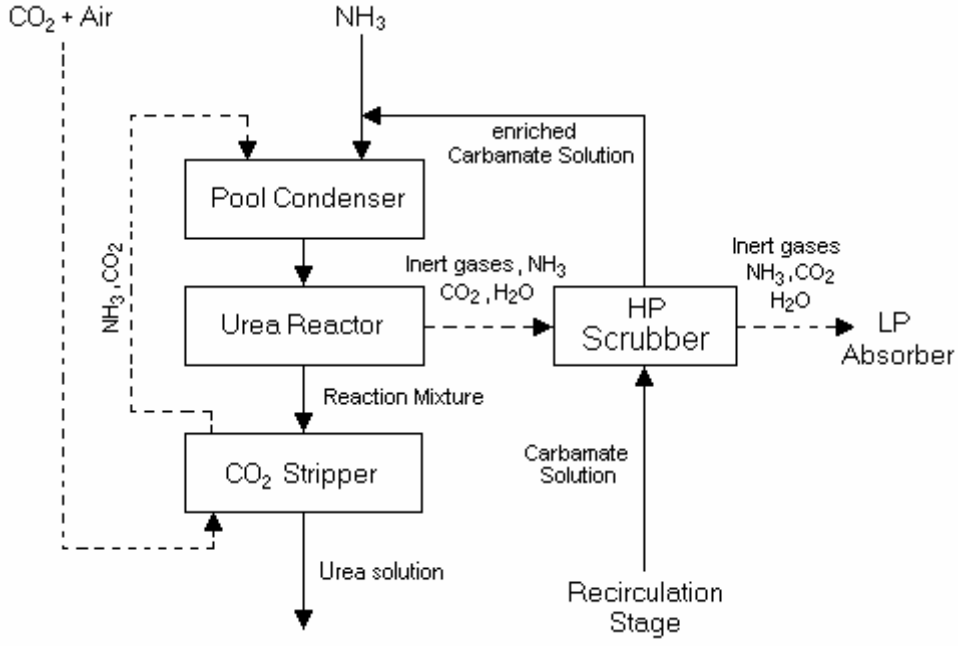

Fig. 2. Block diagram of the Synthesis Section

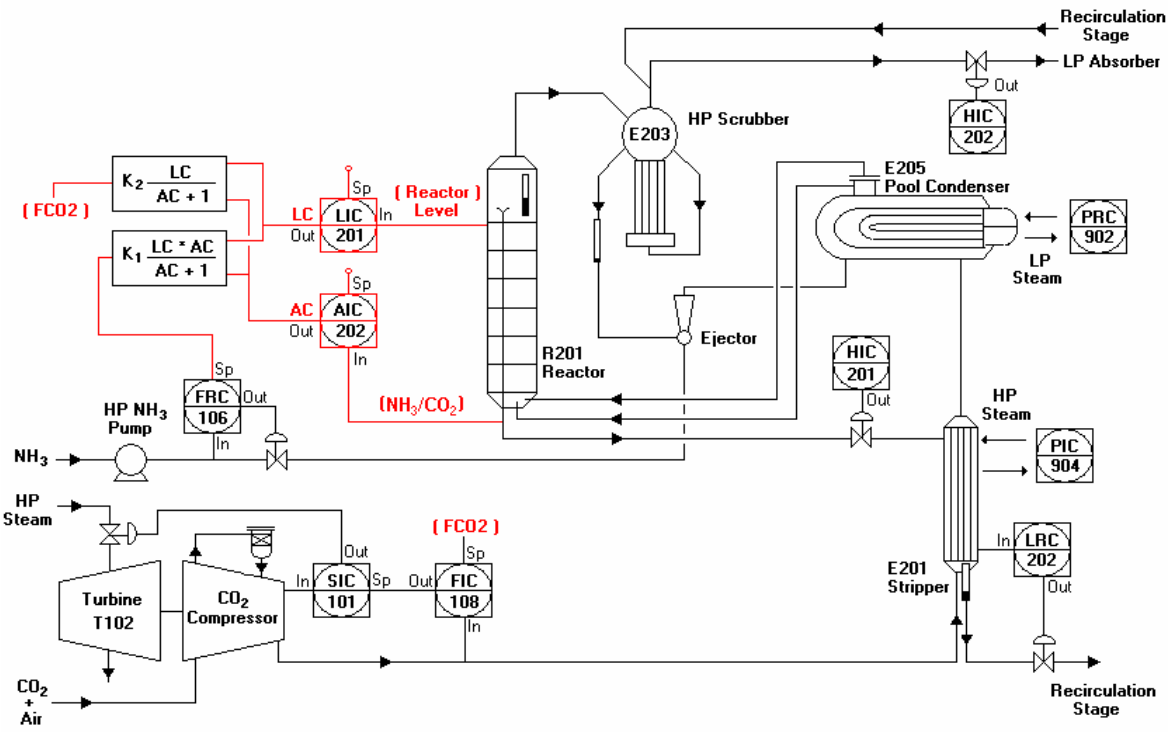

Fig. 3. Traditional control scheme of the synthesis section

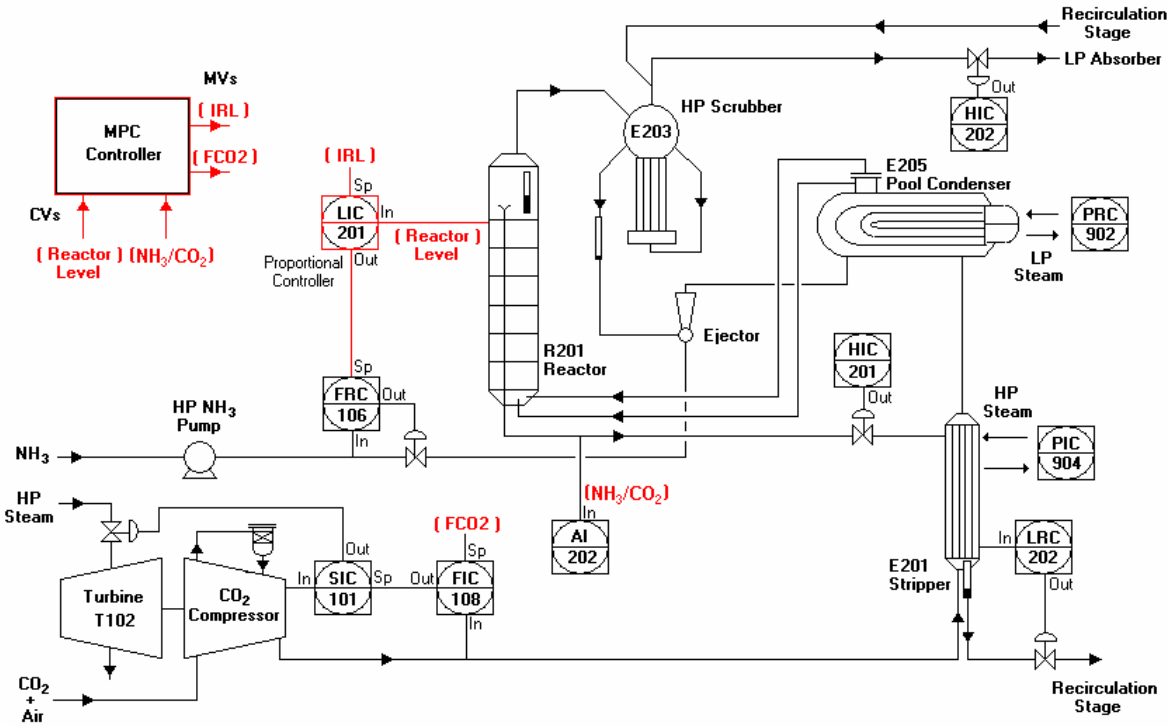

Fig. 4. Control Scheme of the MPC controller without feed-forward action in the PID Loops

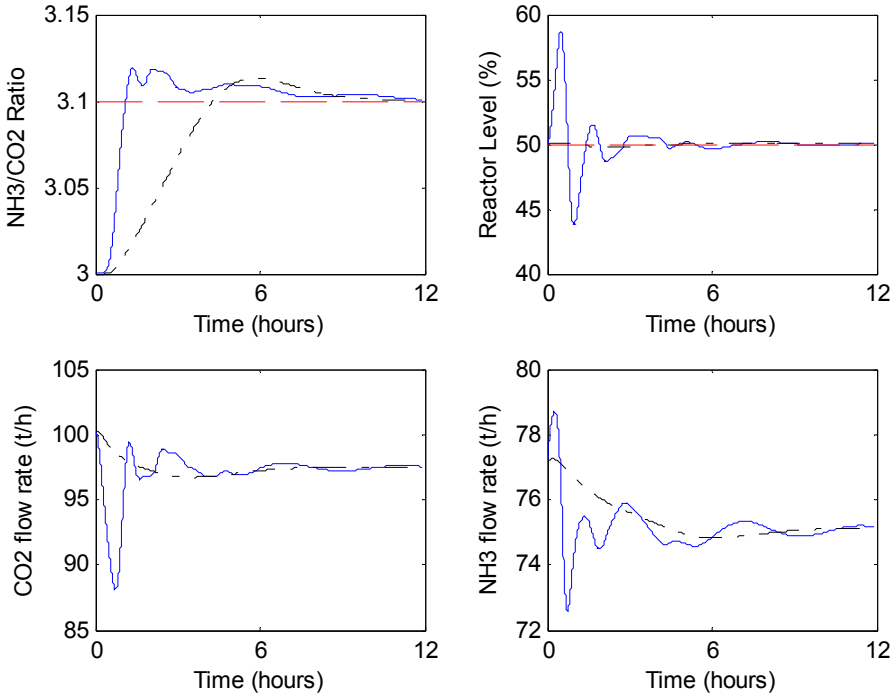

Fig. 5. Controlled (top plots) and manipulated variables (bottom plots) when the setpoint of the NH_3/CO_2 ratio is changed from 3 to 3.1 (Test 1). Blue solid line- MUVAC. Black dash-dotted line– MPC without feed-forward action in the PID loops.

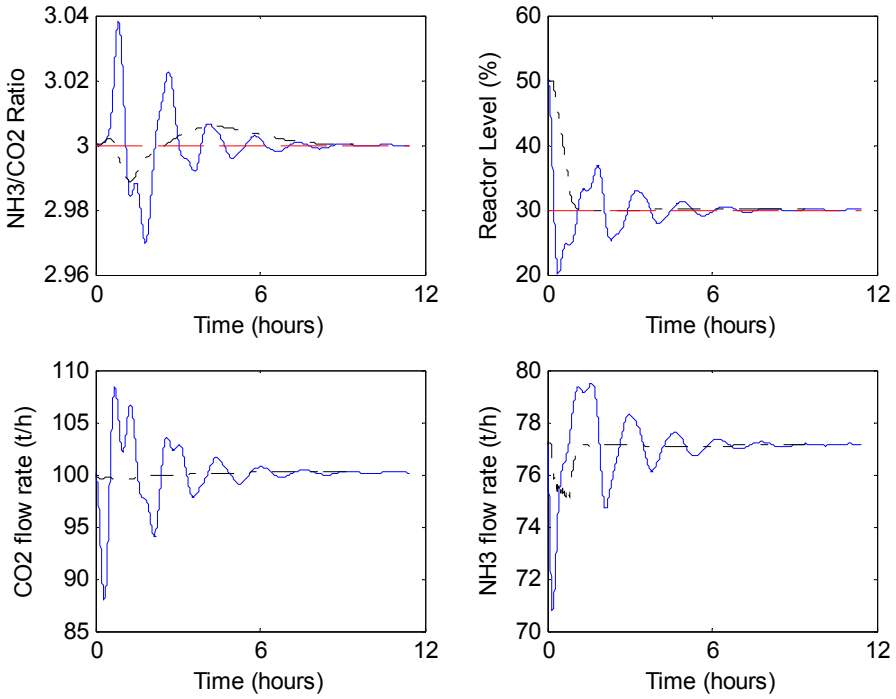

Fig. 6. Controlled (top plots) and manipulated variables (bottom plots) when the setpoint of the Reactor Level is changed from 50% to 30% (Test 2). Blue solid line- MUVAC. Black dash-dotted line– MPC without feed-forward action in the PID loops.

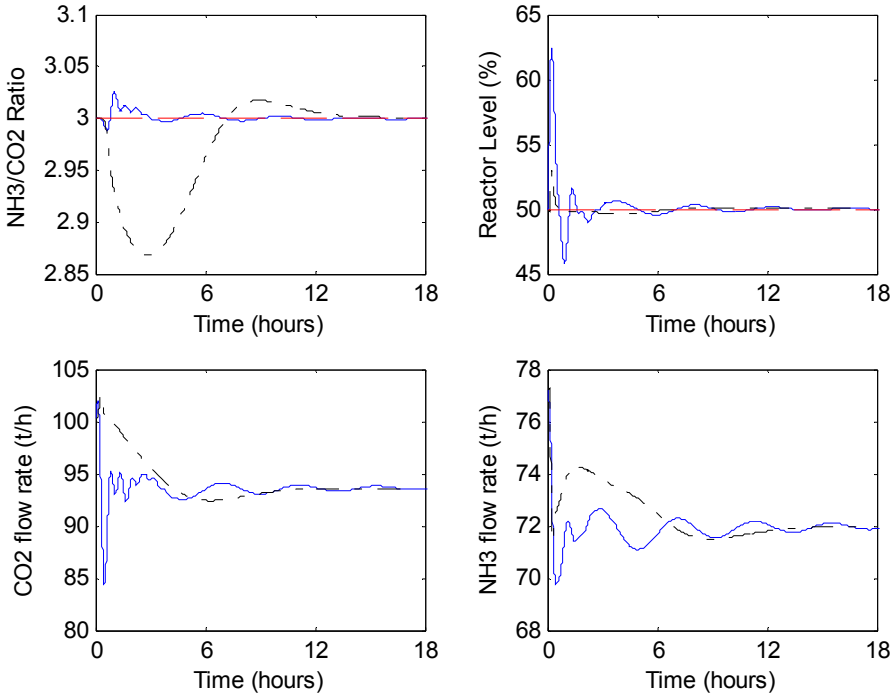

Fig. 7. Controlled (top plots) and manipulated variables (bottom plots) when the position of the valve HIC201 is changed from 56% to 53% (Test 3). Blue solid line- MUVAC. Black dash-dotted line– MPC without feed-forward action in the PID loops.

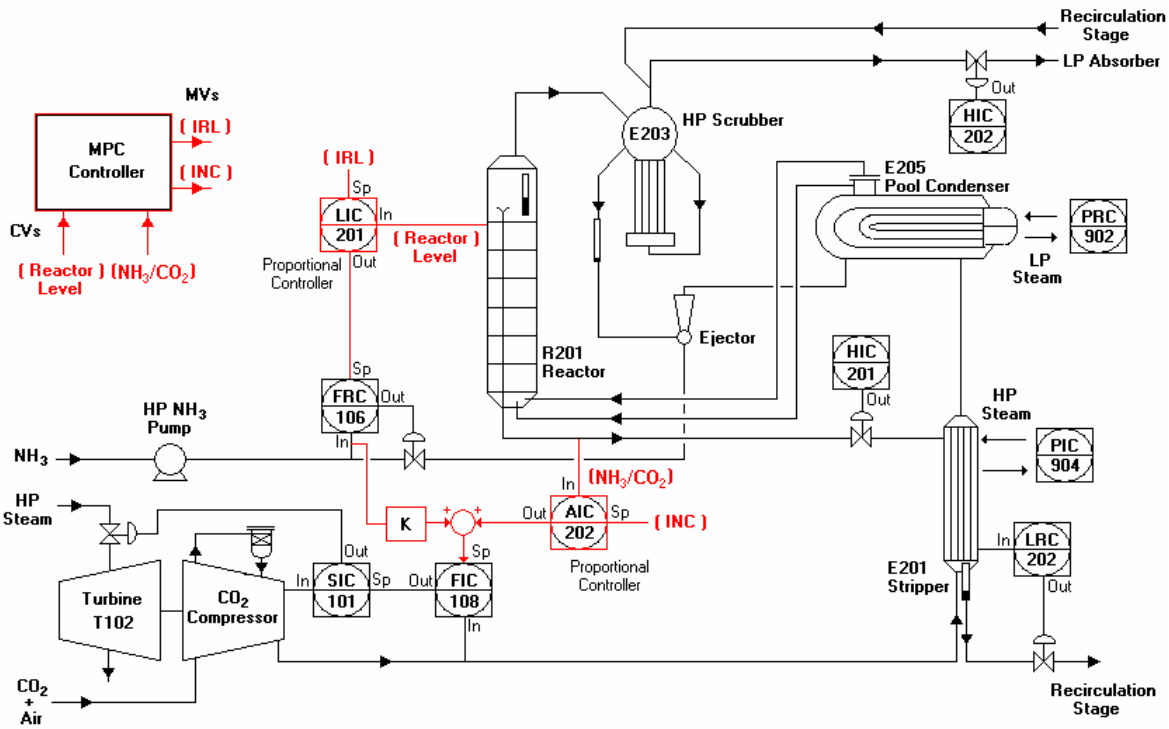

Fig. 8. Control Scheme of the MPC controller with feed-forward action in the PID Loops.

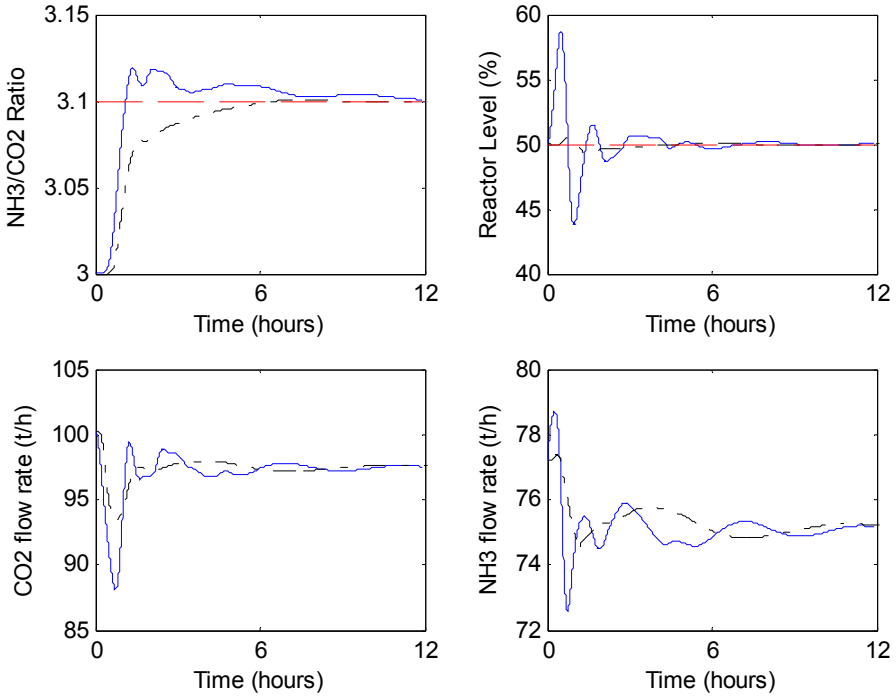

Fig. 9. Controlled (top plots) and manipulated variables (bottom plots) when the setpoint of the NH_3/CO_2 ratio is changed from 3 to 3.1 (Test 1). Blue solid line- MUVAC. Black dash-dotted line– MPC with feed-forward action in the PID loops.

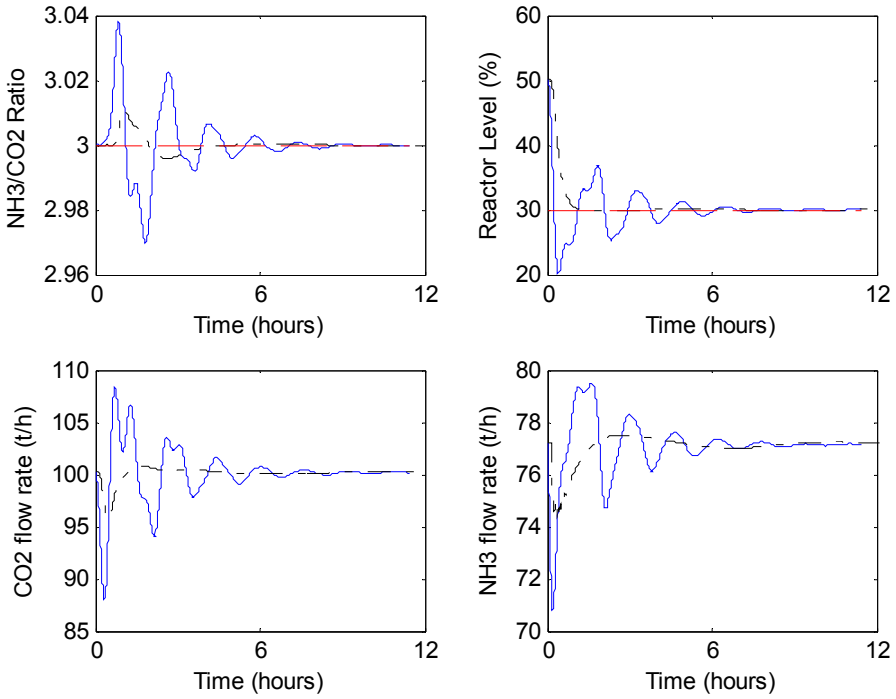

Fig. 10. Controlled (top plots) and manipulated variables (bottom plots) when the setpoint of the Reactor Level is changed from 50% to 30% (Test 2). Blue solid line- MUVAC. Black dash-dotted line– MPC with feed-forward action in the PID loops.

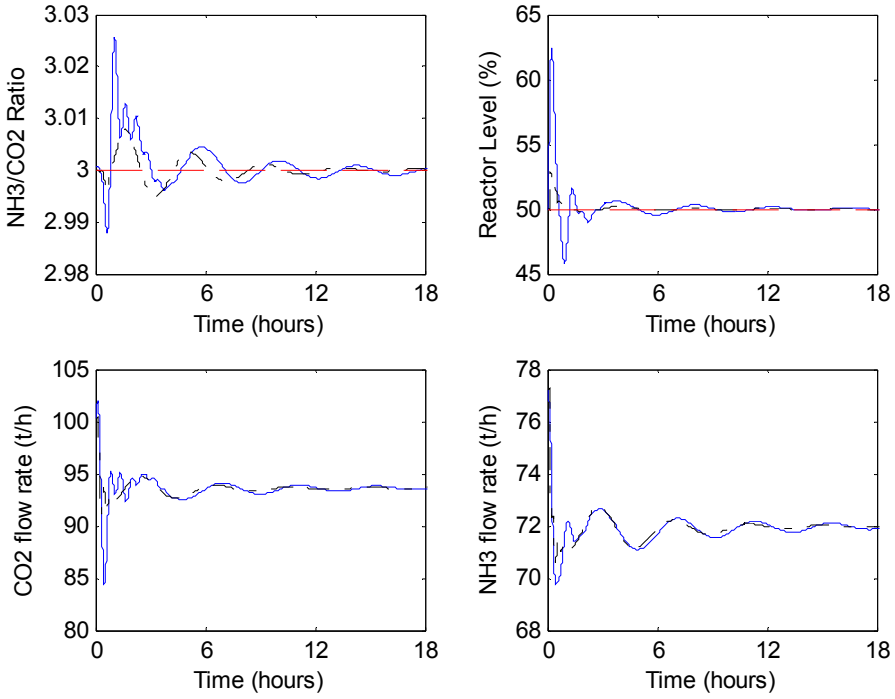

Fig. 11. Controlled (top plots) and manipulated variables (bottom plots) when the position of the valve HIC201 is changed from 56% to 53% (Test 3). Blue solid line- MUVAC. Black dash-dotted line– MPC with feed-forward action in the PID loops.

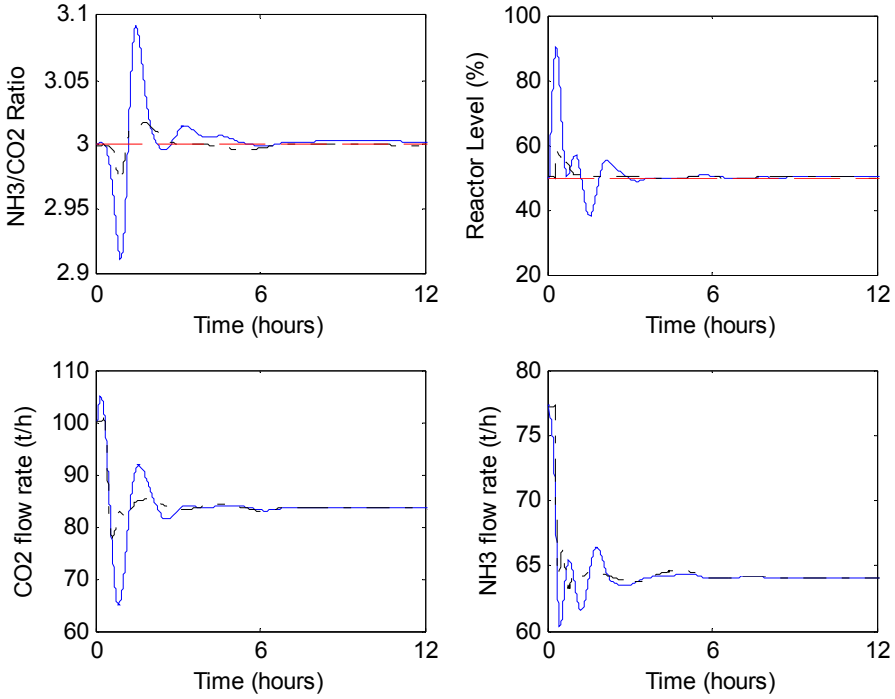

Fig. 12. Controlled (top plots) and manipulated variables (bottom plots) when the position of the valve HIC201 is changed from 56% to 48% (Test 4). Blue solid line- MUVAC. Black dash-dotted line– MPC with feed-forward action in the PID loops.

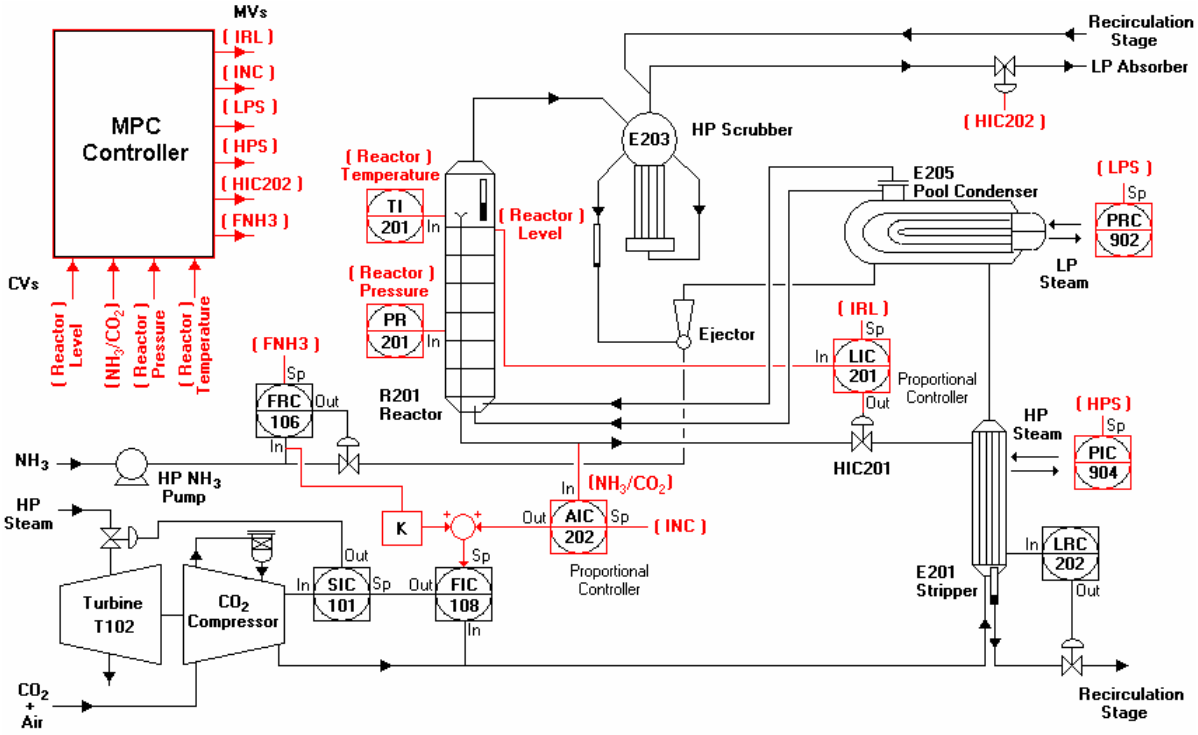

Fig. 13. Control diagram for controlling the production of the synthesis section.

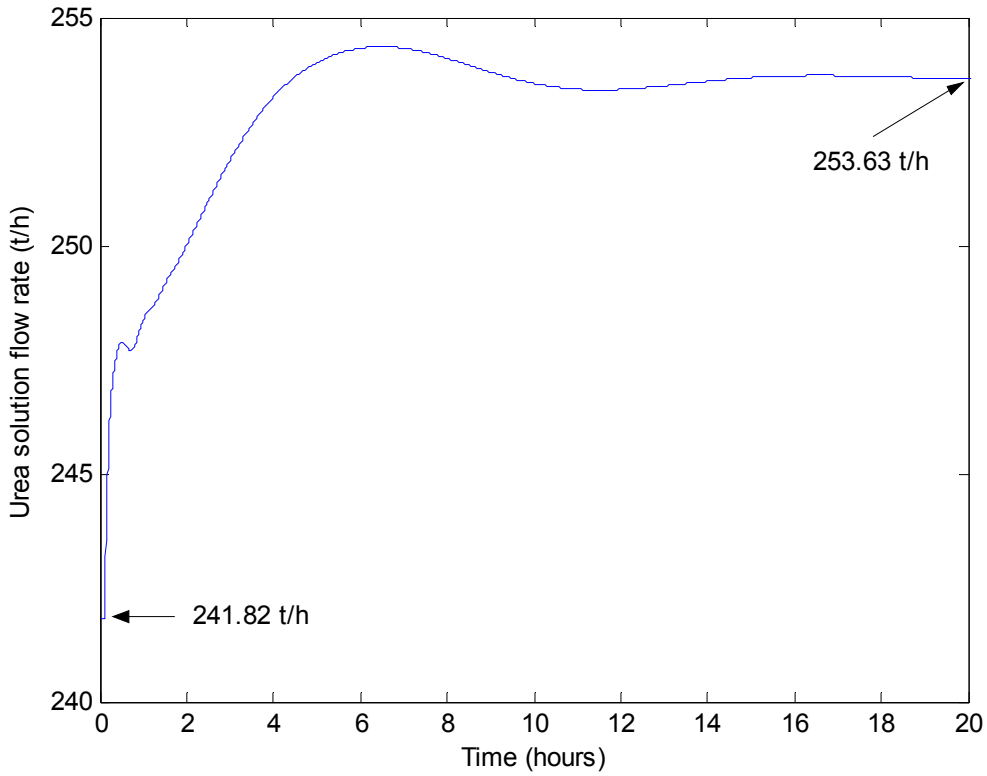

Fig. 14. Urea solution flow rate leaving the Synthesis section when the ideal resting value of the NH_3 flow rate is changed from 77.2 t/h to 80.5 t/h.

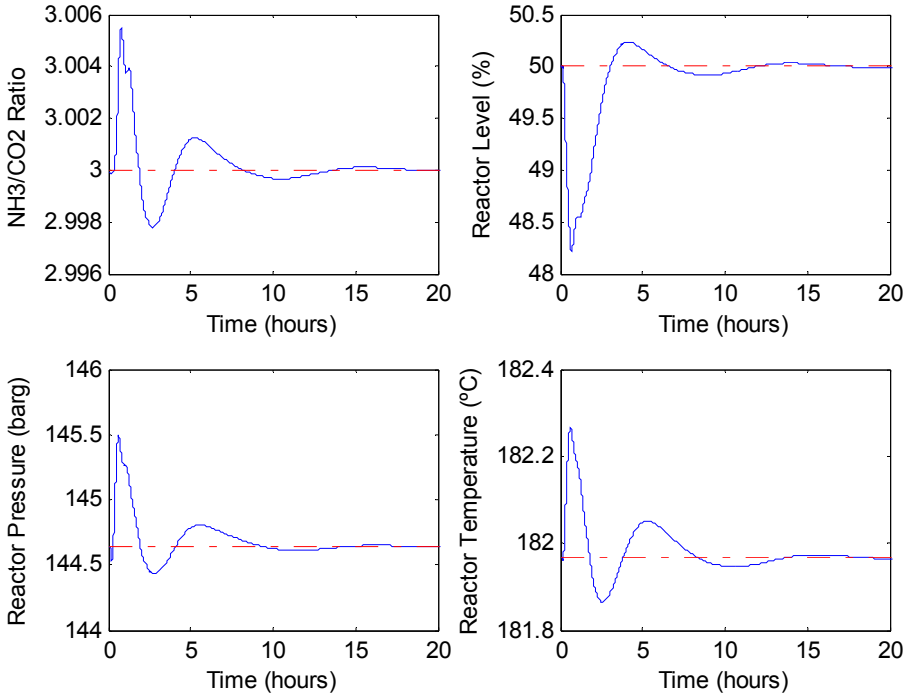

Fig. 15. Controlled variables when the ideal resting value of the NH_3 flow rate is changed from 77.2 t/h to 80.5 t/h .

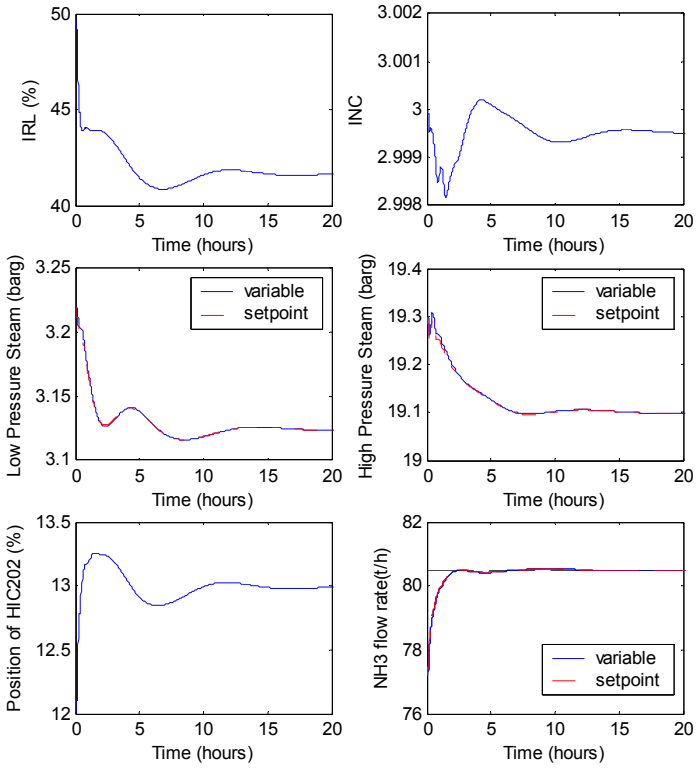

Fig. 16. Manipulated variables when the ideal resting value of the NH_3 flow rate is changed from 77.2 t/h to 80.5 t/h.