A MULTIVARIABLE CONTROL STRATEGY FOR AN INDUSTRIAL GAS-PHASE POLYETHYLENE REACTOR

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Abstract. In an industrial gas-phase polyethylene reactor, the safe operating range of temperature is rather narrow. Even within this temperature range, temperature excursions must be avoided because they can result in low catalyst productivity and significant changes in product properties. Using a first-principles model, including the recycle stream and the heat exchange system, a PID temperature controller with robust performance was designed via optimization in the frequency domain for different operating points. For the reactor total pressure, ethylene partial pressure, H2/C2 and C4/C2 molar ratios, PI controllers were designed. In the PID temperature controller, if the manipulate variable (cooling water valve opening) saturates, then the reactor operates without a feedback temperature controller, leading to oscillatory behavior and limit cycles. It has been demonstrated that the manipulate variable saturation and the nonlinear dynamic behavior are removed when auxiliary variables are used in a multivariable control strategy for the reactor temperature control. In this work, two control structures were compared to define the most suitable proposal for implementation in an industrial reactor and the impact of these control structures in the reactor production and in the polymer melt index are analyzed. The first control structure considers the control problem using the designed PID controller for the reactor temperature and includes a switching strategy with a PI controller for the auxiliary variables. The second control structure considers the control problem also using the designed PID controller for the reactor temperature, however including a MPC controller for the auxiliary variables.

Keywords: Multivariable Control Strategy, Polymerization Reactor and Stability.

1. Introduction

Stabilization of polyethylene reactors is a challenging problem and needs to be addressed through a good control. Dadebo et al. (1997) have demonstrated that without feedback control, industrial gas-phase polyethylene reactors are prone to unstable steady states, limit cycles and excursions toward unacceptable high temperature steady state. In their work, the ability of the controllers to stabilize desired setpoints of industrial interest is evaluated using a bifurcation approach. Seki et al. (2001) have studied the stabilization of gas-phase polyethylene reactors through an adequate tuning of PID controller, which is applicable regardless of the pole locations of the transfer function and have been proposed a retuning method of a PID controller, which can be performed at the onset of closed-loop instability.

In their paper, Ali et al. (2003) have investigated a multivariable control problem of an industrial gas-phase polyethylene reactor, where are considered the two time-scale of the process and the multirate sampling of the process variable control. Specially, the control of the reactor temperature and pressure in addition to the gas partial pressures is considered in their paper. Two control scheme and two algorithms are tested and compared.

All of aforementioned studies dealt only with the tight regulation of the bed temperature to ensure reactor stability in an operating condition where the saturation does not occur. In this work, a more comprehensive model is used to reproduce the process of an industrial gas-phase polyethylene reactor. In this model, the manipulate variable of bed temperature reactor controller is the control valve opening (the real situation in the industrial reactor), instead of the cooling water temperature and, in this way, the saturation of this valve can be accounted for. Besides, in the practical point of view, if the control valve saturates, then the reactor operates without a feedback temperature controller, leading to oscillatory behavior and limit cycles. Therefore, the manipulate variable saturation and the nonlinear dynamic behaviors should be removed from the system by using auxiliary variables. In this work, two multivariable control structures for the reactor temperature control are tested and compared to define the best proposal for implementation in industrial reactors.

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2. Process Model

A fluidized-bed zone and a disengagement zone compose the reactor, as shown in Figure 1. A heat exchanger is used to remove the reaction heat from the compressed recycle stream, and then the cooled gas is mixed with the feed stream to be re-injected in the base of the reactor. The solid catalyst (chromium based) is fed in a stream of nitrogen and then dragged to the fluidized-bed. The product is removed from the reactor by a discharge system operating in cycles determined by the production rate. In the disengagement zone the gas composition is analyzed by chromatography.

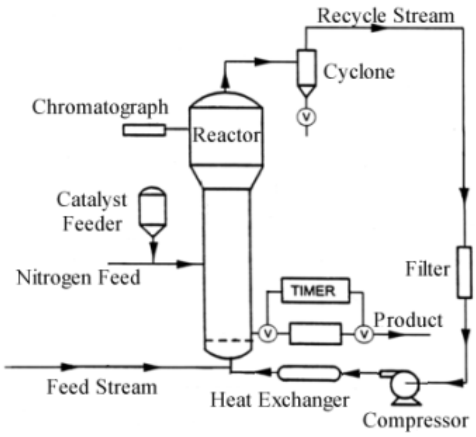


Fig. 1. UNIPOL® process.

The UNIPOL® process model is showed in Figure 2. McAuley et al. (1995) have demonstrated that the additions of the recycle stream and the external cooler models to the reactor model make possible the study of complex dynamics and location of bifurcation points, which could not be reproduced just by the reactor model. Besides, the addition of the heat exchange system, including all their equipments, makes possible an optimal design for the reactor temperature controller, using the cooling water valve opening as manipulated variable. The equations of the model used in this work are detailed in Salau et al. (2005).

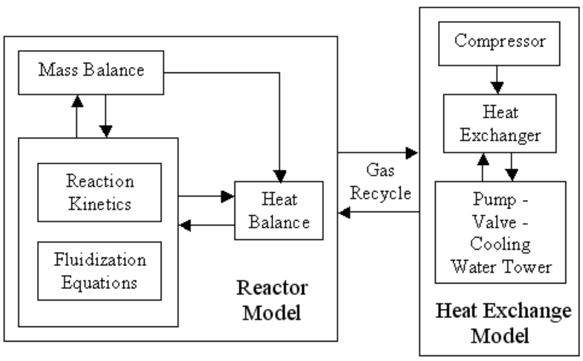


Fig. 2. UNIPOL® process model.

3. Description of the Challenging Control Problem

The studied industrial process presents a critical control loop. This loop does not show a suitable performance, decreasing the unit productive capacity. In Figure 3 is showed the equipments and the instrumentation of the heat exchange system in the industrial plant. The temperature controller was designed with a cascade loop structure. In the primary loop, the control variable is the reactor temperature and the manipulate variable is the recycle gas temperature setpoint. In the secondary loop, the control variable is the recycle gas temperature and the manipulate variable is the cooling water valve opening.

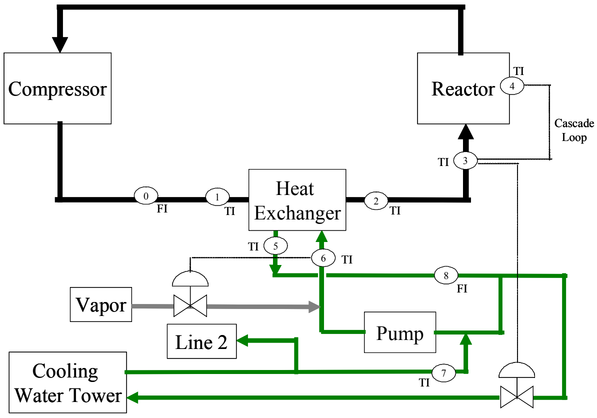


Fig. 3. The heat exchange system of UNIPOL® process.

In order to evaluate the cascade control benefits, two alternative control structures were studied: the use of a cascade loop where the cooling water temperature is the control variable of the secondary loop, instead of the recycle gas temperature, and the direct control of the reactor temperature through the cooling water valve opening, without cascade loop.

In Figure 4 is showed the graph with the step response to a change in valve opening of models G, GOC, and GAC, that represent, respectively, the reactor temperature (direct control without cascade loop), the recycle gas temperature (original cascade loop) and the cooling water temperature (alternative cascade loop). It can be seen the similarity of the dynamics in these variables. Therefore, the use of a cascade loop (original or alternative loop) in this industrial plant would not bring any benefit, because the variables dynamics are closer, carrying problems in the primary control loop adjust. Additionally, the use of cascade loop would bring a design additional cost of a second control loop, an additional delay in the system, and a competition between the two controllers (master and slave), due to high unit interaction, which presents an well pronounced oscillatory behavior. Thus, it was chosen a direct control loop to design the reactor temperature controller.

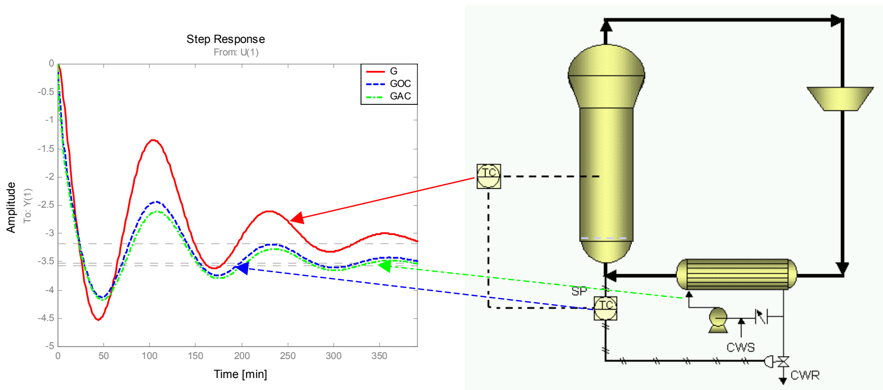


Fig. 4. Step response of models G, GOC and GAC to a change in the valve opening.

In Figure 5, it can be shown the nonlinear relationship between the reactor temperature and the cooling water valve opening. The gain in the reactor temperature is reduced as the valve moves in the fully open direction. Beside, after certain valve opening condition, heat exchange does not occur due to the system thermal limitation. This event can also be observed in Figure 6, where is shown the reduction in gain variation of cooling tower water flow as the valve moves in the fully open direction.

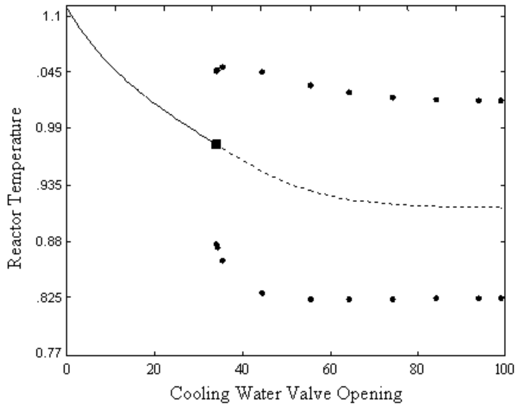


Fig. 5. The effect of cooling water valve opening on stability: — stable steady state; --- unstable steady state; ■ Hopf bifurcation; ● stable limit cycle.

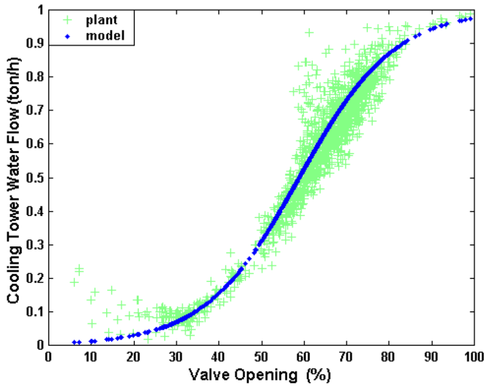


Fig. 6. Comparison between data plant and model for the relation between the valve opening and the water flow to the cooling tower.

Due to high non-linearity between the reactor temperature and cooling water valve opening, the relation between these variables were characterized by a multi-model system (many representative linear models of different operating region). The linear models were built at the points that belong to the unstable regions of valve opening, showed in Figure 5. A more detailed analysis of the dynamic behavior of this process has been provided by Salau et al. (2005).

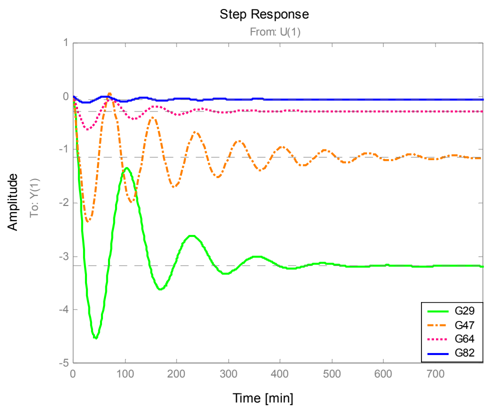


Figure 7. Different dynamics of valve opening step response.

In Figure 7, step perturbations in four valve opening points, belonging to the instability region showed in Figure 5, were made. It can be seen in Figure 7 the different dynamics of the step responses of these points and the multi-model characteristics of this system. Besides, it was observed that in high valve opening zones, the system becomes less sensitive (low gain).

It has been demonstrate that the polymerization reactor needs a suitable reactor temperature control. There are two possibilities to control the reactor temperature: through the reaction heat generation decreasing, where the catalyst rate is reduced to decrease the production rate, and through the heat removal from the recycle gas, where the cooling water rate and the inert saturated organic feed rate are increased to get a better heat exchange. Therefore, the costs of this variables is the determinant factor to choose the manipulate variables. Being the cooling water the cheapest option, its use is maximized but limited by the valve opening saturation. Thus, the second cheaper option is the increasing in inert saturated organic feed rate, which improves the heat exchange, but its use is limited by the reactant gas dew point. The last option is the reduction in the catalyst rate to decrease the heat generation, but it also decreases the production rate and generates off-specification products. Due to small sensibility of catalyst feeder, the reactor temperature decreases beyond the necessary to stabilize in the setpoint value.

4. Proposed Solution for the Regulatory Temperature Control Problem

The reactor system is easily stabilized with a proportional controller. However, the addition of integral and derivative action eliminates offset and leads to faster controller performance (Dadebo et al., 1997). The advantages of a PID controller in this case are the speed and thereby the sample time, the rejecting disturbances, the robustness and the trustworthiness and thereby the fail tolerance because this controller is implemented in DCDS (Distributed Control Digital System).

The PID temperature controller was designed using the SIOM-MMA (Sequential Iterative Optimization Method - Multi-Model Approach, Faccin and Trierweiler, 2004), based on a frequency domain optimization problem for a given desired performance. The method is able to design a PID controller for different operating points with robust performance, as shown in Figure 8. Thus, different unstable regions of valve opening, shown in Figure 5, were used in this method to the robust controller design. Satisfactory results were also obtained for rejecting disturbances. For the reactor total pressure, ethylene partial pressure, H2/C2 and C4/C2 molar ratios, PI controllers were designed to a better representation of industrial plant operation conditions.

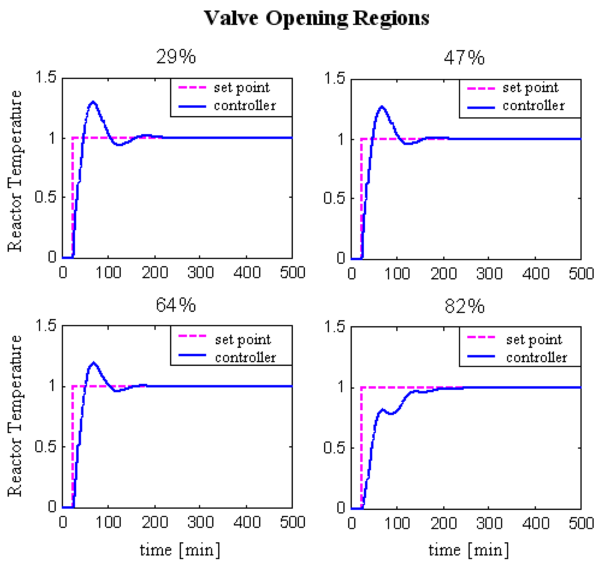
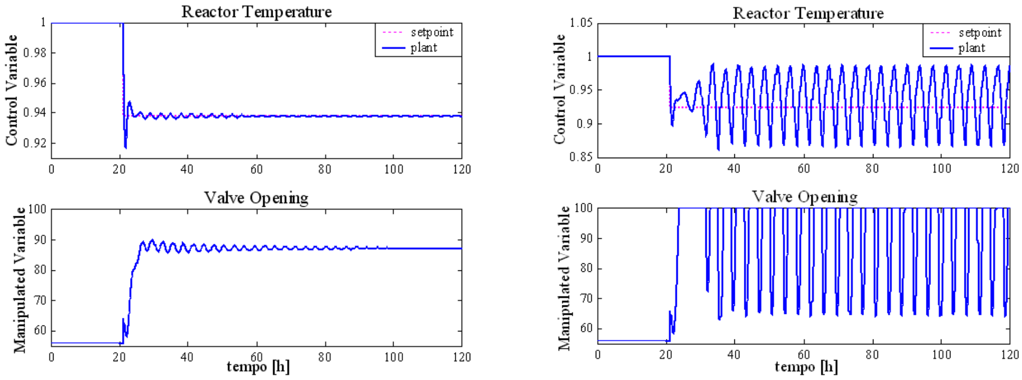


Fig. 8. Temperature response for a step setpoint change using the robust PID controller.

It is shown in Figure 8 that as the valve moves in the fully open direction, the controller performance degrades. Thus, some tests were carried out, using the nonlinear model, to demonstrate the controller performance in regions where the valve moves in the fully open direction. In Figure 9(a), it can be seen that the control valve achieves a valve opening of almost 90%, being able to stabilize the system, but presenting a poor performance.

A better controller performance could be achieved using a gain-scheduling strategy. However, according to Figures 5 and 6, the gain in the reactor temperature as valve moves after an opening of 80% in the fully open direction is practically ineffective. Besides, if the control valve saturates (a common situation in the industrial reactor), then the reactor operates without a feedback temperature controller, leading to oscillatory behavior and limit cycles, as shown in Figure 9 (b). Thus, the use of gain-scheduling strategy was ruled out.



(a) (b)

Fig. 9. Temperature response for a step setpoint change: (a) leading to control valve fully open direction and (b) leading to control valve saturation.

Therefore, manipulation of valve opening alone may not be sufficient to bring the temperature back to the desired level. The limit cycle can be removed from the system when auxiliary variables are used in the reactor temperature control. For instance, catalyst feed rate, inert saturated organic feed rate, and ethylene partial-pressure controller setpoint can reduce the rate of heat generation and stabilize the reactor temperature controller in the desired setpoint (Salau, 2004). However, all of these auxiliary variables reduce the production rate and other problems may arise with their use, some of them are shown in Table 1.

Tab. 1. Problems in using auxiliary variables to help control the reactor temperature

Auxiliary Variables	Problems
Catalyst feed rate	It is difficult to determinate the flow rate in the feeder due to its small sensibility. High impact in the reactor production.
Inert saturated organic feed rate	If the gas mixture dew point is achieved, the distributed plate of gas in the reactor will block. To prevent accumulation, the inert has to be removed from the reactor through the product or purge streams.
Ethylene partial-pressure controller setpoint	It causes a decrease in the recycle heat exchange capacity, by increasing the nitrogen/ethylene concentration ratio.

In the industrial unit, due to the high correlation between the reactor total pressure and the ethylene partial pressure, the last one is controlled in open loop through the nitrogen feed rate or the bleed stream, owing to the reactor total pressure be controlled through the ethylene flow rate manipulation. However, to use the ethylene partial-pressure controller setpoint in the multivariable control strategy would be necessary to design a controller strategy for the reactor using the total pressure and the ethylene partial pressure, simultaneously. Thus, only two of these variables: catalyst feed rate and inert saturated organic feed rate are used in this work

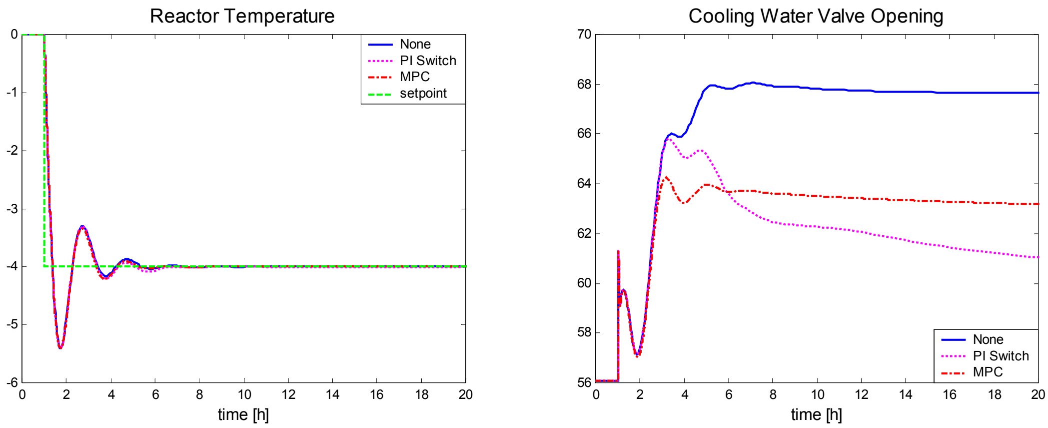
5. Proposed Solution for the Supervisory Control Problem

In the supervisory control problem, the reactor temperature PID controller, designed using the SIOM-MMA (Faccin and Trierweiler, 2004), was maintained and the valve opening, which is the manipulated variable of this controller, becomes the controlled variable in two different control structures within a multivariable strategy: a sequential PI switch controller and a model predictive controller (MPC). The idea of these control structures is, through the reactor temperature PID controller, opening the control valve until achieving a region where the gain between the valve opening and the cooling water becomes low. When a pre-determinate value of the valve opening is achieved, the inert saturated organic feed rate, with high specific heat capacity, is increased to improve the heat exchange between the recycle gas and the cooling water. Thereby, more reaction heat is removed from the reactor, decreasing the reactor temperature back to the setpoint. However, the maximum quantity of this inert saturated organic feed rate is limited by the gas mixture dew point. When this maximum quantity is achieved without stabilizes the reactor temperature in the setpoint, the catalyst feed rate is reduced to decrease the reactor temperature.

The first control structure considers the control problem using the designed PID controller for the reactor temperature and includes a switching strategy with a sequential PI controller for the auxiliary variables. The second control structure considers the control problem also using the designed PID controller for the reactor temperature, however, including a MPC controller for the auxiliary variables. The responses of both sequential PI switch controller and MPC are compared to define the most suitable proposal for implementation in an industrial reactor.

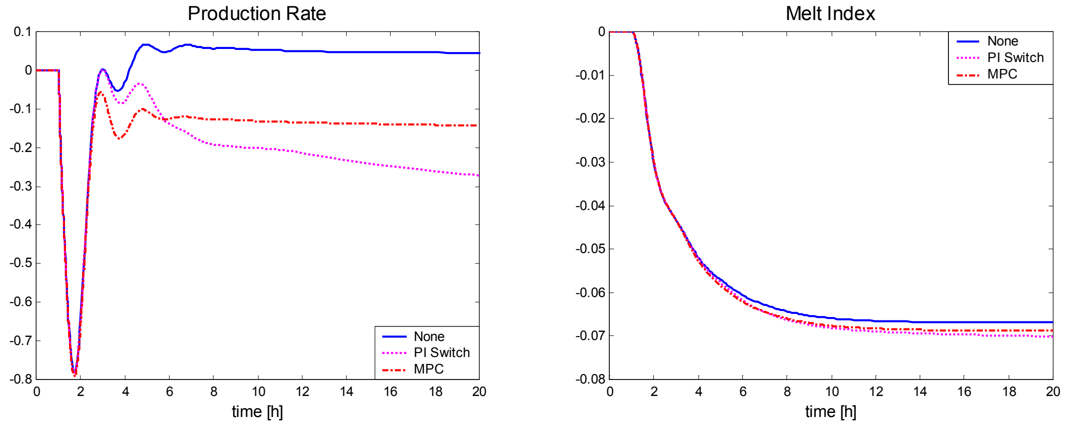
5.1. Results for a Nominal Operating Condition

It was chosen a nominal operating condition to compare the control problem using only the reactor temperature PID controller and the control problem using the two proposed multivariable control structures. In the multivariable control structures, the inert saturated organic feed rate is increased after the valve opening achieves 60%, and the catalyst feed rate is decreased after the inert saturated organic feed rate increases 3 dimensionless units.



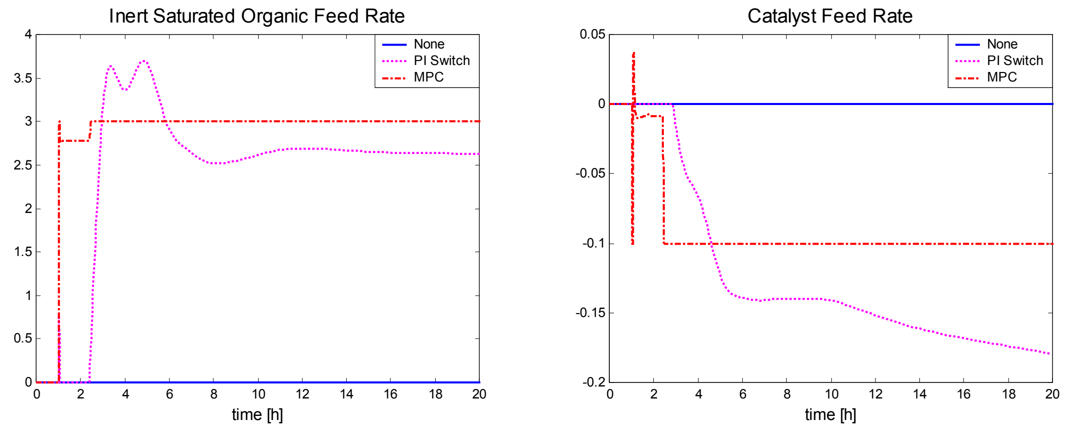
(a) (b)

Fig. 10. (a) Reactor temperature and (b) valve opening cooling water to a step change reactor temperature setpoint without and with the multivariable strategy (sequential PI switch and MPC).



(a) (b)

Fig. 11. (a) Production rate and (b) melt index to a step change in reactor temperature setpoint without and with the multivariable strategy (sequential PI switch and MPC).



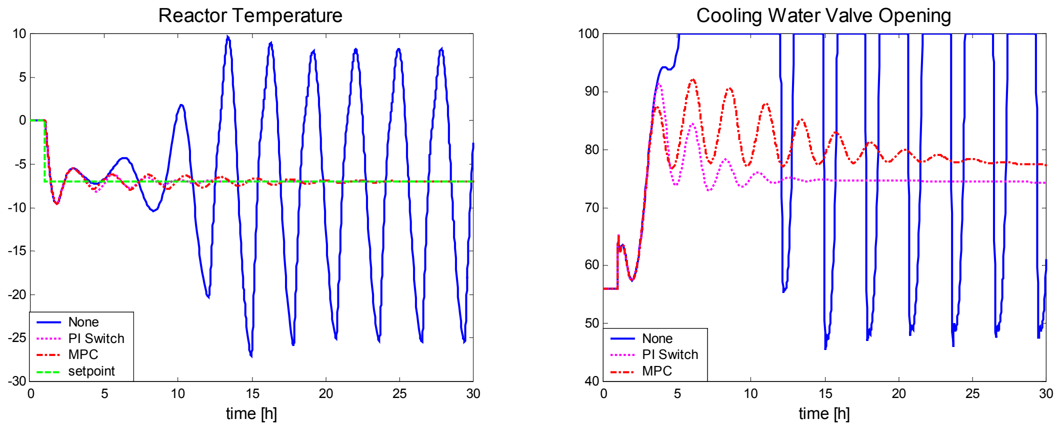
(a) (b)

Fig. 12. (a) Inert saturated organic feed rate and (b) catalyst feed rate manipulation without and with the multivariable strategy (sequential PI switch and MPC) to stabilize the system in the reactor temperature setpoint.

According to Figures 10-12, the performances of both proposed controllers to stabilize the reactor temperature in the desired setpoint are similar. Thus, both controllers are able to get quickly the new polymer melt index, avoiding off-specification product. However, the MPC controller presents a better result than the sequential PI switch controller for a nominal operating condition, because this controller is able to achieve the maximum quantity of the inert saturated organic feed rate before to manipulate the catalyst feed rate and, thereby, reducing production loss.

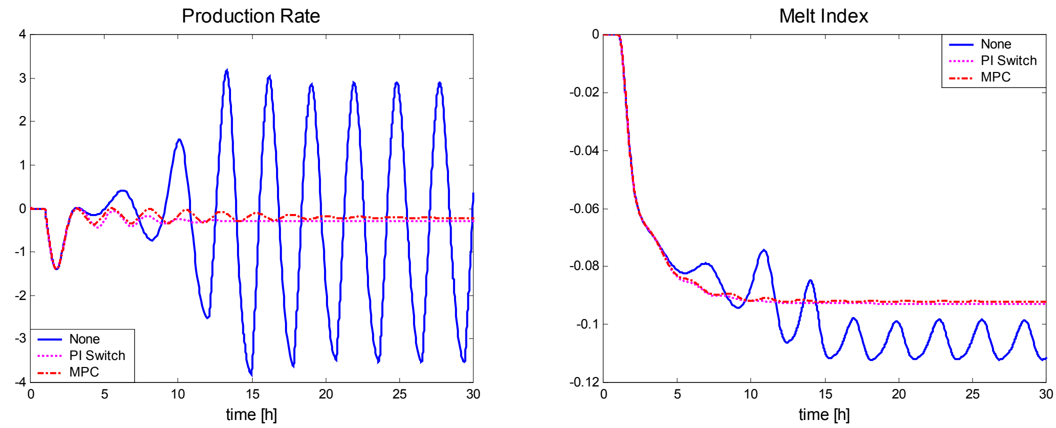
5.2. Results for a Saturation Condition

It was chosen a valve opening saturation condition to compare the control problem using only the reactor temperature PID controller and the control problem using the two proposed multivariable control structures. In the multivariable control structures, the inert saturated organic feed rate is increased after the valve opening achieves 75%, and the catalyst feed rate is decreased after the inert saturated organic feed rate increases 6 dimensionless units.



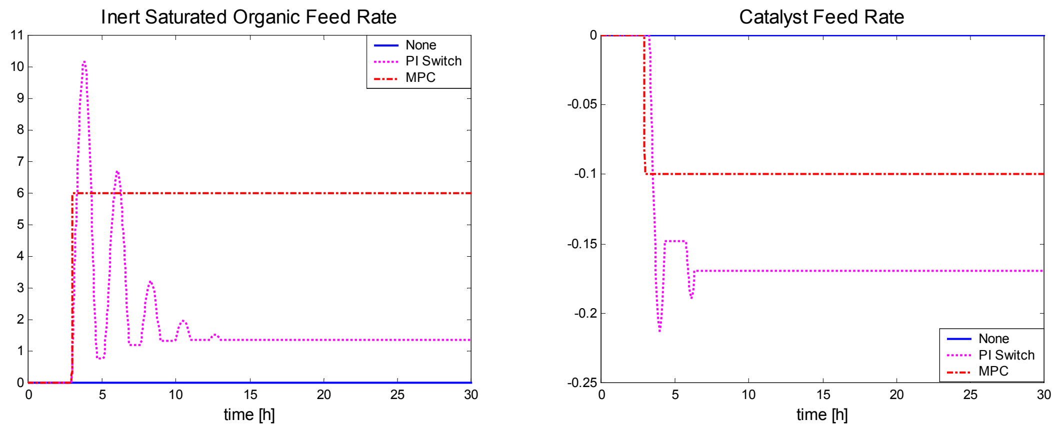
(a) (b)

Fig. 13. (a) Reactor temperature and (b) valve opening cooling water to a step change in reactor temperature setpoint without multivariable strategy, leading to the limit cycles, and with the multivariable strategies (sequential PI switch and MPC).



(a) (b)

Fig. 14. (a) Production rate and (b) melt index to a step change in reactor temperature setpoint without multivariable strategy, leading to the limit cycles, and with the multivariable strategies (sequential PI switch and MPC controller).



(a) (b)

Fig. 15. (a) Inert saturated organic feed rate and (b) catalyst feed rate manipulation without multivariable strategy and with the multivariable strategies (sequential PI switch and MPC) to stabilize the system in the reactor temperature setpoint.

In a valve opening saturation condition, the results for the proposed strategies are similar to the results obtained in a nominal operating condition. Without the multivariable strategies the oscillatory behavior leads to a limit cycle. In this condition, the MPC controller also is able to achieve the maximum quantity of inert saturated organic feed rate before manipulating the catalyst feed rate. Therefore, the MPC controller is more sensitive to the oscillations in the valve opening than the sequential PI switch controller, because the reactor temperature PID controller is operating in a region where the PID performance is poor, according to Figure 8.

6. Conclusion

It was shown that control of gas-phase polymerization reactors is a difficult task due to high non-linearity of the system and the strong interaction of the process variables. For this challenging control problem, a PID temperature controller with robust performance was designed via optimization in the frequency domain for different operating points. However, when the manipulation of valve opening alone is not sufficient to bring the temperature back to the desired level and this equipment saturates, limit cycles arises in the system. Thus, the use of auxiliary variables can reduce the rate of heat generation and stabilizes the reactor temperature controller in the desired setpoint. However, all of these auxiliary variables reduce the production rate and other problems may arise with their manipulation and, consequently, the use of these variables in a multivariable control strategy is limited. The results of this work suggest that the use of an appropriate strategy in a MPC controller for closing the reactor temperature control loop avoids the saturation of manipulate variable and, hence, the undesired non-linear dynamic behavior. Besides, this controller reduces the production loss and improves the product quality because of its ability to stabilize the reactor temperature in the setpoint with good performance. In order to obtain a better performance of the MPC controller in regions where the control valve moves in the fully open direction, a re-tuning in the reactor temperature PID controller must be done. In this case, the SIOM-MMA (Faccin and Trierweiler, 2004) approach, using different operating points of the region where the gain between the reactor temperature and the valve opening is low, seems to be a good choice.

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Acknowledgments

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