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FREE SET POINT TRACKING AND DISTURBANCE REJECTION OF BINARY DISTILLATION COLUMN USING MODEL PREDICTIVE CONTROL

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ABSTRACT

Off set free set point tracking is a challenging problem in distillation column control. This is usually due to the multivariable nature of the distillation column and hence the interaction between the loops. This study designed a model predictive controller using MATLAB and Simulink for a Lubyen Vinate Distillation Column. The results show that the model predictive controller performs an off set free set point tracking and disturbance rejection of about 70%. The system has a fast settling time of 5 minutes and 3 minutes for the 17th and 4th column respectively. This shows that the designed system has an efficient control as compared to the response of distillation column 60 minutes and 40 minutes for the 17th and 4th column respectively.

1. INTRODUCTION

A large number of industrial processes are found to be multiple-input and multiple-output (MIMO) in nature [1-5] . A good number of such MIMO processes may be categorized as two-input-two-output (TITO) process [6, 7] where two SISO (single-input-single-output) loops interact with each other[8]. Presence of interaction among the loops[9, 10] provides the intricacy in designing feedback controllers for such TITO processes. Adjustment of controller parameters for one loop influences the performance of other loop; in an extreme case it may destabilize the entire system [11]. Hence, in such cases PID controllers with conventional tuning usually fail to provide satisfactory performance. Moreover, the presence of dead time in process loop makes the tuning task more difficult[11]. Processes with large dead time exhibit undesired oscillations with prolonged settling time [12, 13]. So, controlling such TITO loops with time delays is an open problem[11].

Many researchers have focused on control methods for multiple-input and multiple-output (MIMO) systems. MIMO with *N* input/output processes are characterized by significant interactions between their inputs and outputs[2]. The control of MIMO processes is usually implemented using sets of single-input single-output (SISO) control loops. Interaction is a phenomenon that the loop gain in one loop depends on the loop gain in another loop. This interaction between controlled loops, leads to deterioration in the control performance of each loop [2]. The control of MIMO processes requires proper input–output pairing and development of decoupling compensators unit [4]. Decoupling control has emerged as one of the most popular techniques in the industrial process. The basic idea is to weaken, or even eliminate, the interactions between different input and output signals by decoupling methods.

2. COMPARATIVE ANALYSIS

To decide the control method to be used a comparative study was done on different control schemes.

2.1 Pole placement-based PI controller

The PID control families are most widely used control techniques in the industries due to their design simplicity and applicability to many processes[9, 11]. The proportional gain increases the system responses and improved the closed loop stability. However, it leaves a trace of an offset error and large proportional gain lead to instability of the system. The integral gain clears the steady state offset but larger integral time delayed the settling time of the process. In the presence of overshoots, the derivative gain is needed for effective control[14]. The addition of derivative gain however, amplifies noise in the systems. Therefore, appropriate selection of the control depends on the system dynamics. For any of the PID controls, the error value is calculated by taking the difference between the set-point and the measured controlled variable. The control tries to make the as close to zero as possible [15].

2.2. Cohen-Coon PID Tuning

In this type of control gains tuning, an open loop response is considered. As an empirical approach, similar procedures as outline above are followed. The Cohen – Coon formulations requires process with dead time [1]. In this work, PI gains are considered and based on the open loop response, the PI control gains are obtained.

2.3 PI plus Feedforward control

Feedback plus feedforward technique is one of the advance control schemes commonly used in the process industries. This scheme significantly improves the performance of a process in the presence of a measured disturbance. The feedback control takes care of the measured variable with respect to the set point while the feedforward eliminates the effect of external disturbance to the process. In this study, the effect from system 1 is considered as the disturbance to system 2. Thus, the feedforward control can be obtained by solving system 2 model at static equilibrium [16].

2.4 Internal Model Controller (IMC)

The main idea of IMC scheme is to obtain a good closed loop response from the open loop dynamic model [17]. The internal model law states that acceptable control can be achieved if and only if the closed loop control encapsulates some dynamics of the process. Thus, the controller depends on the accuracy of the derived model, because the controller would have the inverse dynamics of the plant in order to perfectly track the reference input [17]. The IMC design is in two phases. First, the process model (Gm) is factored into the invertible part (Gm-) and the non-invertible part (Gm+). The Gm+ contains the time delays and the right half plane zeros. The controller (Gc) is given in the equations below, where Gf is a low pass filter with the general form, N represents the order of the model in order to get a perfect poles zero cancellation, τc is the controller time constant which is the critical design parameter of IMC scheme. Depending upon the system, the filter time constant can have the value of the dead time of the system. After block reduction technique, the closed loop controller (Gcc) can be expressed as the final equation with negligible valve dynamics (Gv ≈ 0).

$$G_c(s) = \frac{G_f(s)}{G_m^-} \quad (1)$$

$$G_f(s) = \frac{1}{(\tau_c s + 1)^N} \quad (2)$$

$$G_{cc}(s) = \frac{G_c(s)}{1 - G_c(s) G_m(s)} \quad (3)$$

2.5 Model Predictive Controller (MPC)

Model predictive control refers to the class of control algorithms that compute a manipulated input profile by utilizing a process model to optimize an open loop performance objective subject to constraints over a future time horizon [13, 18]. Recently, the popularity of MPC has been increased for industrial applications and academic world. The reason is the ability of MPC designs to produce high performance control systems having capacity of operating without expert intervention for long durations[19]. The process model is the most important characteristic of MPC[19, 20]. The model is very vital for ability to implement MPC. Many alternative categories of MPC models exist, namely, linear or nonlinear, continuous or discrete-time, distributed parameter or lumped parameter, deterministic or stochastic, input output or state-space, frequency domain or time domain, first principles or black box[18]. Therefore, the step response model for MIMO system with two inputs and two outputs is developed by using principle of superposition. A model-based controller can be designed basing on the step response model for MIMO systems. Initially, error must be defined since the controllers behave according to the error. If constraints are imposed on controller and system's output, the minimization becomes more complex due to adding the constraints to objective function. Therefore, the solution cannot be solved explicitly.

Investigations into level control techniques for a TITO system using the PI controller, PI plus feed-forward and IMC scheme have been presented. Simulations of the dynamic model of a coupled tank have been performed to study the effectiveness of the controllers. The results of the proposed controllers showed a significant tracking performance using all the controllers. The performances of the controllers demonstrated that MPC scheme provides the best level tracking followed by IMC, PI plus feed-forward control as compared to the single PI controller

3 PLANT MODEL

The model in use is of a Lubyen and Vinate distillation column model relating temperatures on the 4th and 17th trays from the bottom of the column (T_4, T_{17}) to the reflux ratio R and steam flow rate to the reboiler S given as:

$$\begin{bmatrix} T_{17}(s) \\ T_4(s) \end{bmatrix} = \begin{bmatrix} \frac{-2.16e^{-s}}{8.25s+1} & \frac{1.26e^{-0.3s}}{7.05s+1} \\ \frac{-2.75e^{-1.8s}}{8.25s+1} & \frac{4.2e^{-0.35s}}{9.0s+1} \end{bmatrix} \begin{bmatrix} R(s) \\ S(s) \end{bmatrix} \quad (4)$$

Below is a Simulink representation of the above stated model.

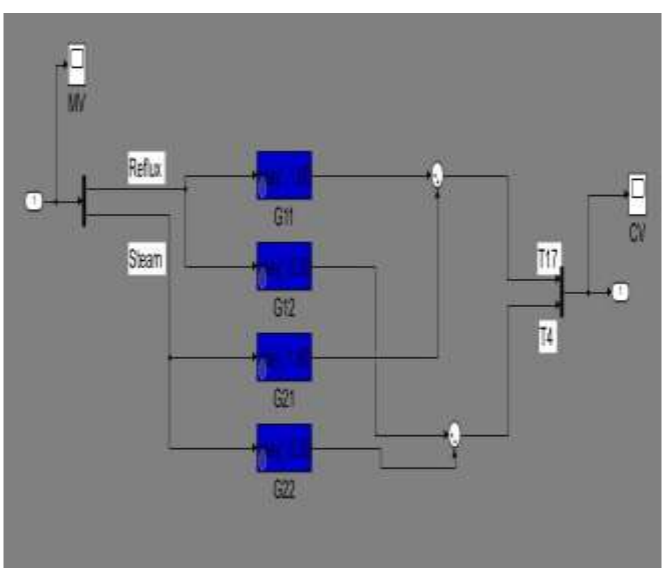

Figure 1: Distillation Column Model in Simulink.

4 PLANT MODEL

A linear A linear MPC was used on the plant. Below are the MPC design parameters considered for this work.

a. Controller sample time
b. Prediction Horizon
c. Control Horizon
d. Input and Output Weights

These values will be obtained from tuning in order to give us the desired control action. The aim is to achieve efficient trade-off between complexity and set-point/disturbance changes.

4.1 Controller Sample Time

Sampling too slow will have a negative impact on performance. Sampling fast will not necessarily provide better performance, though it may lead us to spend more than necessary on high-end instrumentation and computing resources**.** We have selected a sample time of 0.2s to offset both of these extremes.

4.2 Prediction Horizon

The choice of prediction horizon is important in model predictive control system design.

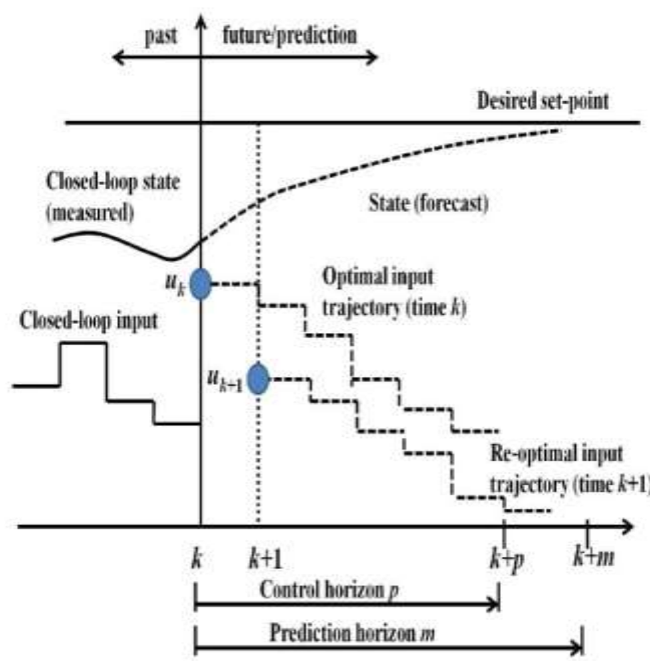

Figure 2: Model Predictive Control Representation.

It is the number of future control intervals the MPC controller must evaluate by prediction when optimizing its manipulated variables at control interval *k*. And we have chosen the value of 10 as ours.

4.3 Prediction Horizon

The control horizon, *m*, is the number of manipulated variables moves to be optimized at control interval *k*. The control horizon falls between 1 and the prediction horizon. The default is *m* = 2.

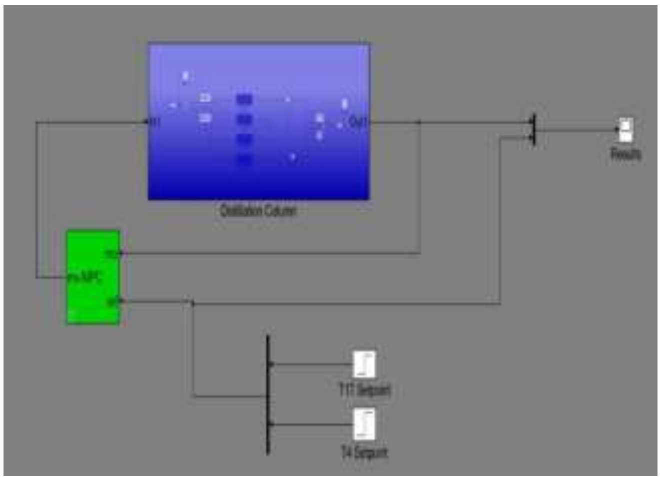

Figure 3: MPC design on Simulink.

Regardless of your choice for *m*, when the controller operates, the optimized MV move at the beginning of the horizon is used and any others are discarded. Here, we have used a value of 4.

5 SIMULATION AND RESULTS

The following values were obtained after tuning the controller in Simulink;

a. Controller sample time: 0.2s
b. Prediction Horizon: 10
c. Control Horizon: 4
d. Input and Output Weights: 0.396

The step response of the plant model before controller was applied to the system is shown in figure 4. The figure shows that each output has two contrasting responses. Without the controller the system has tendency to react erroneously and hence not providing the desired output response (set point). Figure 5 shows the impulse response of the system. The settling time of the system is 60 minutes and 40 minutes for the 17th and 4th column respectively. These values are similar to the settling time for the step response. These high values were reduced largely on application of the MPC as shown in Figure 6.

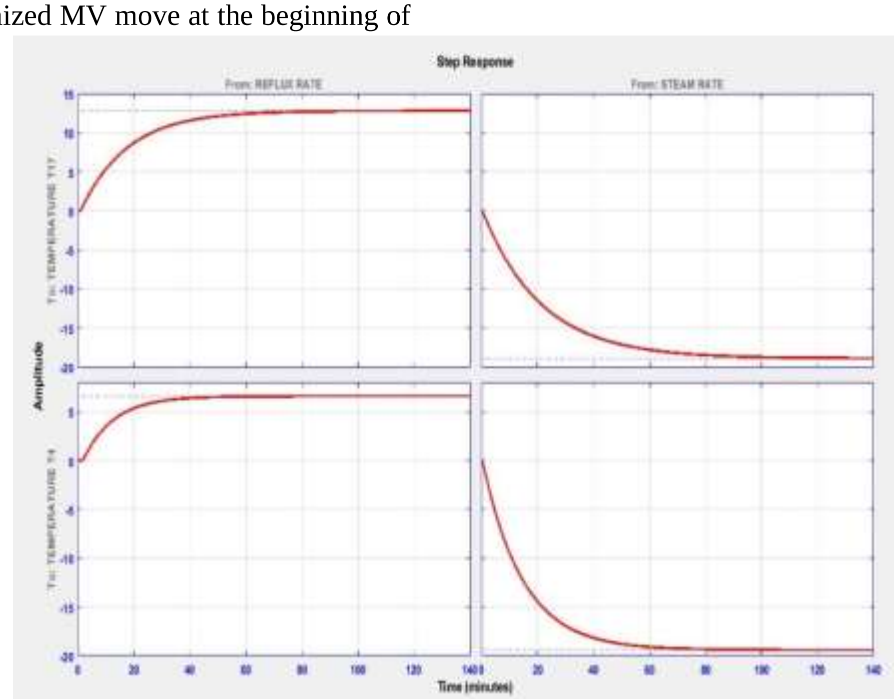

Figure 4: Step Response of Plant without controller

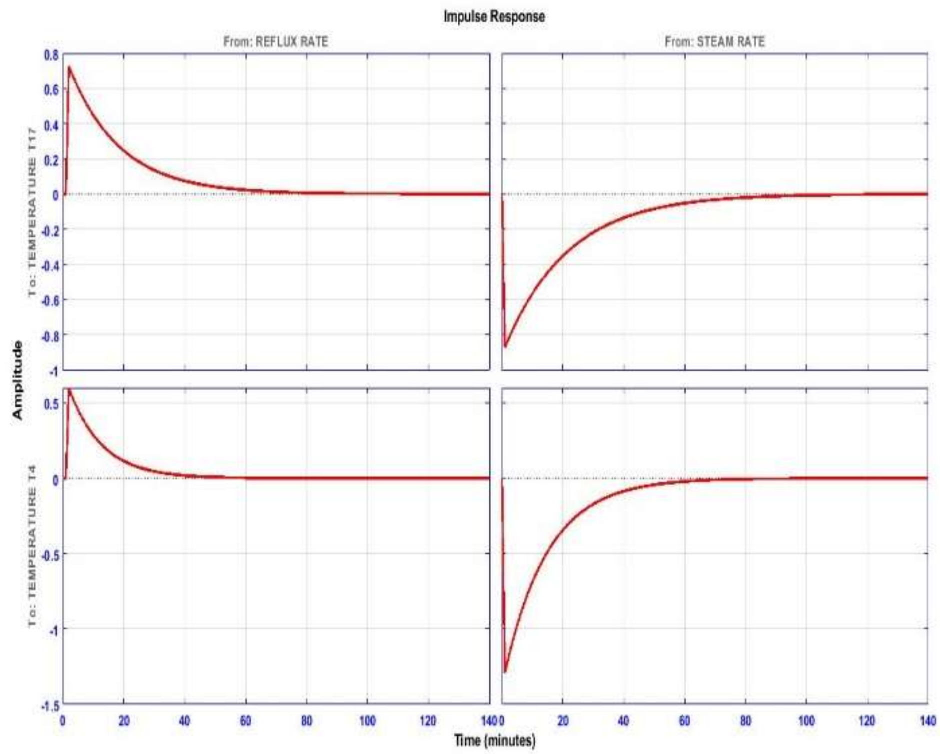

Figure 5: Impulse Response of the System.

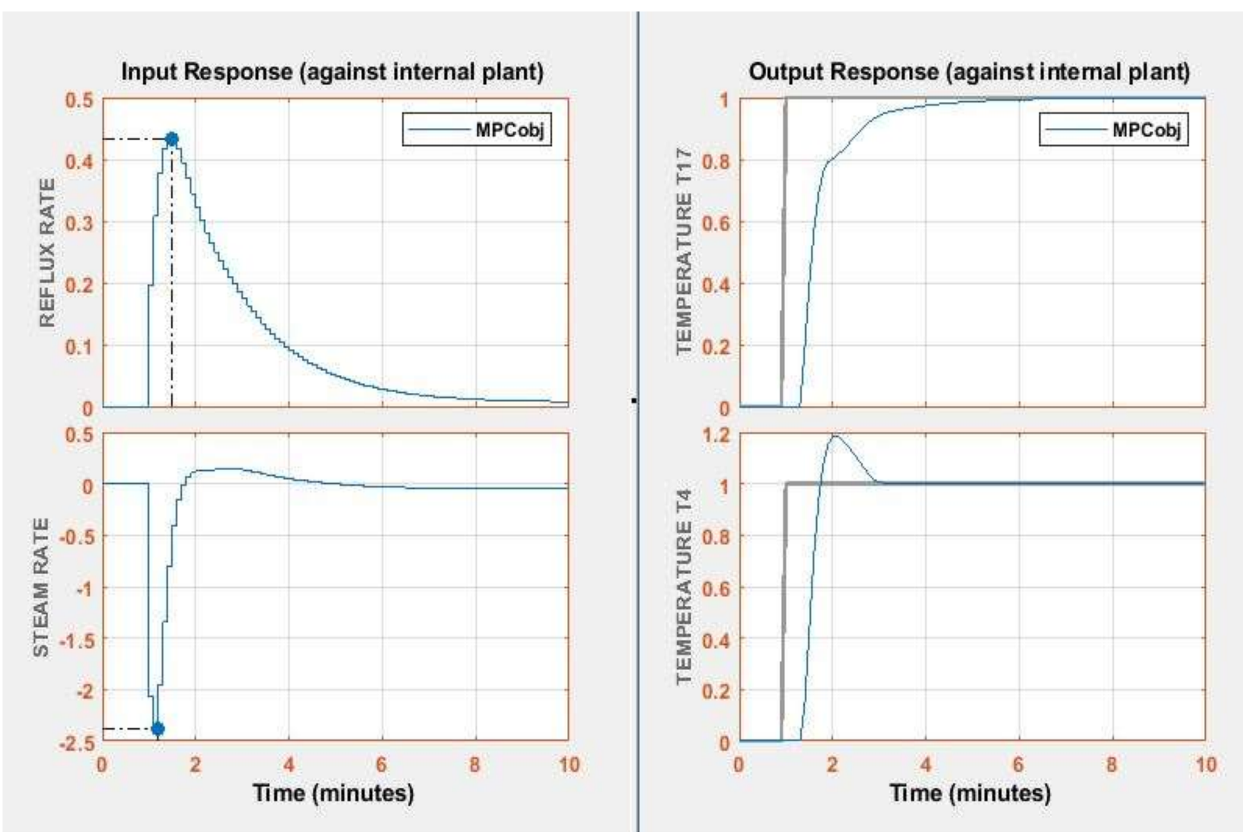

Figure 6: Simulink Output Response of the system.

Results from Figure 6 showed that the system followed the desired set point upon application of the MPC. The system settles at 5 minutes and 3 minutes for the 17^{th} and 4^{th} column respectively. This is a very fast response as compared to the normal response of the distillation column operations, 60 minutes and 40 minutes for the 17^{th} and 4^{th} column respectively.
At steady state, the system has achieved off-set free set point tracking.

6 CONCLUSIONS

This work shows design of MPC using MATLAB and Simulink for TITO Luyben and Vinate distillation system. The design linearized the model and applied a liner MPC for control. Simulation results shows the design is able to achieve offset free tracking (at steady state) and disturbance rejection of about 70%. The controller settling time is 5 minutes and 3 minutes for the 17^{th} and 4^{th} column respectively. The controller is able to effect control relatively fast as compared to the normal response of the plant 60 minutes and 40 minutes for the 17^{th} and 4^{th} column respectively.

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