4.1 Conventional PID Controller

PID controllers are the most effectively used for controlling the systems. The PID controller calculation involves three separate parts: the proportional, the integral and derivative parts. The PID controller produces the control signal as follows.

$$u(t) = u(t-1)+k_p e(t)+k_i \int_0^t e(t)dt + k_d \frac{d(e(t)}{dt} \quad \ldots(15)$$

where u(t) : Plant input or set point ;
e (t) = r(t) − y(t) : Error between the set point and measured value at time instant ;
k_p: Proportional gain; k_i : Integral gain; k_d : Derivative gain.

A PID controller is adopted to control the composition of the bottom product. The controller gains for PID are designed and optimized with simulation model by using Simulink response optimization library block. It is mainly a numerical time domain optimizer developed under MATLAB/Simulink environment There are three parameters to be optimized in order to have satisfactory control performance. Those are corresponding respectively to k_p, T_i and T_d which are the proportional, integral and derivative gains for the composition control. These parameters are tuned using Ziegler-Nichols tuning method.

4.2 Model Predictive Controller (MPC)

Model predictive control (MPC) is an advanced method of process control that has been in use in the process industries in chemical plants and oil refineries. The main advantage of MPC is the fact that it allows the current timeslot to be optimized, while keeping future timeslots in account. This is achieved by optimizing a finite time-horizon, but only implementing the current timeslot. MPC has the ability to anticipate future events and can take control actions accordingly. PID controller do not have this predictive ability. In the control aspect of this work, single controlled variable (bottom composition) and single manipulated variables (reboiler duty (Q)) were selected for the formulation of the model predictive control. It was simulated with the aid of Model Predictive Control Toolbox of MATLAB. MPC has the ability to anticipate future events and can take control actions accordingly.

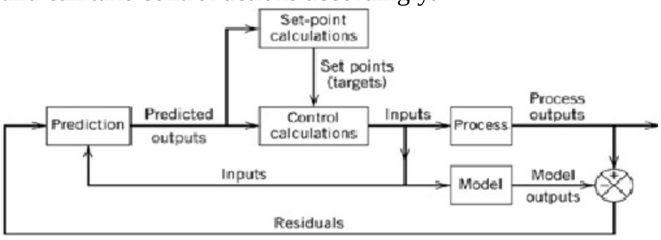

Figure 6: Block Diagram of Model Predictive Control

Basic structure of MPC is given above. Current values of the output variables are calculated using a process model and then difference between predicted and actual outputs are used as a feedback signal to a prediction block. The predicted outputs are used in controlled calculation and setpoint calculation after considering constraints on the input and output variable. MPC configuration is analogues to both internal model control configuration and smith predictor configuration because model and process are parallel acted and difference act as feedback control signal. But coordination of the control and set point calculations makes MPC superior than others. Traditionally economic optimization setpoint for the control calculation, called as target, is calculated from a steady-state model of the process. Economic optimization is depending on maximizing a cost function, or maximizing a production rate. The optimum values of set points are distorted continuously which in turn functions of variations in process conditions, equipment, and instrumentation, as well as economic data such as prices and costs. In MPC, Set points are calculated in each sampling time. Control action is determined based on current measurements and predictions of the future output.

Table 1: Physicochemical properties and nominal steady state operating conditions

Parameter		Value
Number of stages	Rectifying section Reactive section Stripping section	0 13 2
Liquid holdup (mol)	Condenser Column tray Reboiler	30,000.0 1000.0 30,000.0
Reaction Rate (mol$m^{-3}s^{-1}$)	Main reaction Side reaction	$3.15\times10^{15} e^{\frac{-9547}{T}} x_{EO} x_W$ $6.30\times10^{15} e^{\frac{-9547}{T}} x_{EO} x_{EG}$
Feed location	EO W	8 2
Heat of reaction ($Jmol^{-1}$)	Main reaction Side reaction	−80,000.0 −13,100.0
Feed flow rate (mols^{-1})	EO W	7.65 7.31
Latent heat of vaporization ($Jmol^{-1}$)		40,000.0
Bottom product specification (EG, mol %)		94.54

5. Simulation Results

In this project, used ordinary differential equation solver ode23s, Math toolbox and Simulink toolbox of MATLAB software. First lumped parameter model of the RD column is developed and dynamic simulation is done. Simple PID control strategy and the MPC control strategy for bottom composition control by manipulating reboiler steam flow rate are designed and simulated in Simulink. The steady state response shows a stable behaviour in which the bottom product, mole fraction of ethylene glycol reaches a steady state value of 0.945% of purity in composition. The steady state response is shown in fig.7. The system attains steady state response only when no disturbance affects the system. If there is any change occurs in the nominal value of feed flow rate and feed composition the product composition will not attain the steady state value.

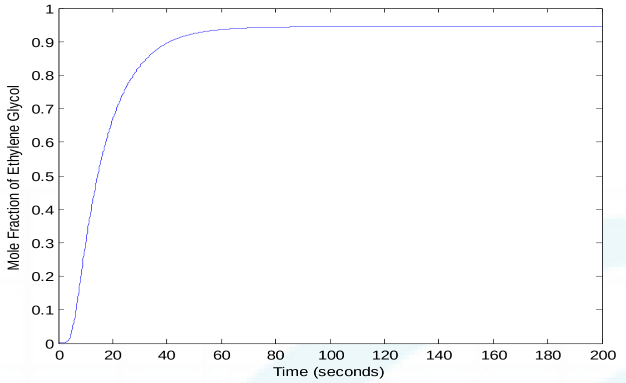

Figure 7: Steady State Response

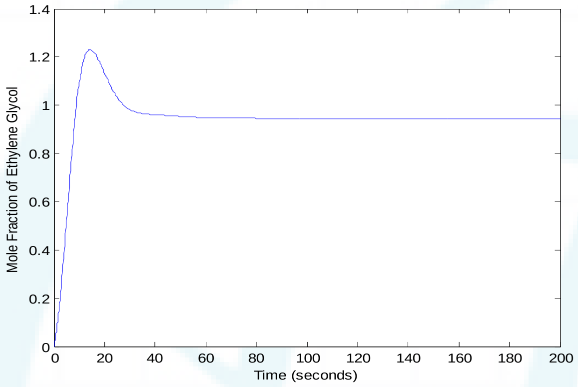

Figure 8: Bottom Composition Control using PID controller

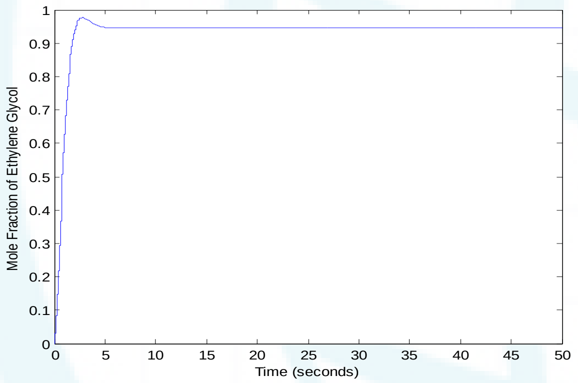

Figure 9: Bottom Composition Control using MPC

Response with two controllers is given in fig.8 & 9 respectively. PID control scheme exhibits large and poor dynamic performance, while MPC control scheme showed perfect dynamic response with the same controller and tuning parameters. The MPC controller shows significant performance improvement with little overshoot and no oscillations.

6. Conclusion

Here in this paper I have presented the modeling and control design of a totally refluxed reactive distillation column without disturbance. Control of the bottom compositions of the column is a difficult task due to presence of process nonlinearities. The structure allows taking into account dynamic variations of the process and adapting the controller parameters to this various conditions. MPC controller achieved a accurate performance in controlling the bottom compositions.

7. Future Scope

From the foregoing analysis, disturbance is not introduced into the system. A Model predictive controller eliminates the disturbance affect into the system. A disturbance is added in the system and compare the response of the system with tuned PID controller.

References

[1] R. Taylor and R. Krishna. “Modelling reactive distillation. Chemical Engineering Science,” 55:5183–5229, 2000.

[2] K. Sundmacher and A. Kienle, "Reactive distillation: status and future directions," Weinheim: WILEY-VCH, 2003.

[3] M. A. Al-Arfaj and W. L. Luyben, "Comparison of alternative control structures for an ideal two product reactive distillation column," Ind. Eng. Chem. Res., vol. 39, pp.3298-3307, 2000.

[4] Al-Arfaj, M. A., and W. L. Luyben, ‘‘Comparative Control Study of Ideal and Methyl Acetate Reactive Distillation,’’ Ind. Eng. Chem. Ž.Res. , in press 2002a.

[5] Kumar,A., and P. Daoutidis, ‘‘Modeling, Analysis and Control of Ethylene Glycol Reactive Distillation Column,’’ AIChE J., 45, 51 Ž.

[6] Al-Arfaj, M. A., and W. L. Luyben, ‘‘Control Study of ETBE ReacŽ.tive Distillation,’’ Ind. Eng. Chem. Res., in press.

[7] Al-Arfaj, M. A., and W. L. Luyben, ‘‘Design and Control of One-Reactant-Two-Products Reactive Distillation: Olefin Metathesis Ž.Case,’’ Chem. Eng. Sci., 57, 715.

[8] R. Khaledi and B. R. Young, “Modeling and model predictive control of composition and conversion in an ETBE reactive distillation column,” Industrial & Engineering Chemistry Research, 44(2005), 3134-3145.

[9] A. Giwa and S. Karacan, “Modeling and simulation of a reactive packed distillation column using delayed neural networks,” Chaotic Modeling and Simulation, 1(2012a), 101-108.

[10] A. Bemporad, M. Morari and N.L. Ricker, Model Predictive Control Toolbox: Getting Started Guide, The Math Works, Inc., Natick, 2012a.

Author Profile

Athira G P was born in Kerala, India in 05/09/1989. She received B.Tech degree in Electrical and Electronics Engineering from Travancore Engineering College, Oyoor, Kollam, Kerala in 2011.Currently, she is pursuing his M Tech degree in Industrial Instrumentation & Control from TKM College of Engineering, Karicode, Kollam. Her research interests include Control Systems.