

APPLICATION OF MODEL PREDICTIVE CONTROL (MPC) TUNING STRATEGY IN MULTIVARIABLE CONTROL OF DISTILLATION COLUMN

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Abstract

A model predictive control strategy is proposed for multivariable nonlinear control problem in a distillation column. The aim is to provide a solution to nonlinear control problem that is favorable in terms of industrial implementation. The scheme utilizes multiple linear models to cover wider range of operating conditions. Depending on the operating conditions, suitable model is used in control computations. Servo and regulatory controls of the system are examined. Comparisons are made to conventional controllers. The results confirmed the potentials of the proposed strategy.

Keywords : *Tuning, model predictive control, multivariable, nonlinear control, chemical process*

Introduction

Predictive control is now one of the most widely used advanced control methods in industry, especially in the control of processes that are constrained, multivariable and uncertain. A large number of implementation algorithms, including industrial predictive control applications (Qin and Badgwell, 2003) have appeared in the literature.

The cornerstone of MPC is the model (Clarke, 1996). It cause MPC is called MBPC (model-based predictive control). MPC uses models in 2 ways: using a reliable model to predict effect of past control moves on P (prediction horizon) future outputs, assuming no future moves, and using the same model to compute the optimal M (control) horizon moves.

Dynamic matrix control (DMC) (Cutler and Ramaker, 1980) is the most popular MPC algorithm used in chemical process industry today. Over the past decade, DMC has been implemented on a wide range of process applications. A major part of DMC's appeal in industry stems from the use of a linear finite step response model of the process and a simple quadratic performance objective function. The objective function is minimized over a prediction horizon to compute the optimal controller output moves as a least-squares problem.

Tuning a controller is a direct way to reach its optimum performance. Tuning conventional controllers (P, PI, and PID) is related to obtain an optimum setting of controller parameters (controller gain K_c , integral time T_i , and derivative time T_d). Ziegler-Nichols, Lopez, Ciancone, etc. (Marlin, 2000) are some examples of single-loop tuning in P, PI, and PID controllers. Huang, et al. (2003) has proposed a

direct method for multi-loop (multivariable) PI/PID controller design based on FOPDT/SOPDT model of each loop.

An MPC controller has certain parameters setting to achieve its optimum performance. Those parameters are sampling time (T), prediction horizon (P), model horizon (N), control horizon (M), controlled variable weights (γ_s^2), and move suppression coefficients (λ_s^2). During the time, trial-and-error efforts have been done to find out this goal until Shridhar and Cooper (Dougherty and Cooper, 2003) proposed a tuning strategy for unconstrained SISO and multivariable MPC. Dougherty and Cooper (2003) proposed a non-adaptive DMC tuning strategy (see Table 1) based on all of FOPDT models in systems.

OLMR Distillation Column

Consider OLMR (Ogunnaike, Lemaire, Morari, and Ray) distillation column (Ogunnaike, et al, 1983) as shown by Figure 1. The distillation column studied was a 19 plate having variable feed and side stream draw-off locations, with the binary ethanol-water system.

Although this multivariable control strategy may be implemented for either time domain or Laplace transform models, here we shall outline the essential features using a Laplace domain model. For multivariable systems such as distillation columns having multiple delays, a commonly employed linear model takes the form:

$$y(s) = G_p(s) u(s) + G_d(s) d(s) \quad (1)$$

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Table 1. Non-adaptive DMC tuning strategy

Approximate the process dynamics of all controller output to measured process variable pairs with FOPDT models:

$$\frac{y_r(s)}{u_s(s)} = \frac{K_{rs}e^{-\theta_{rs}s}}{\tau_{rs}s+1} \quad (r = 1,2, \dots, R; s = 1,2, \dots, S)$$

1. Select the sample time as close as possible to:

$$T_{rs} = \text{Max}(0.1\tau_{rs}, 0.5\theta_{rs}) \quad (r = 1,2, \dots, R; s = 1,2, \dots, S)$$

$$T = \text{Min}(T_{rs})$$

2. Compute the prediction horizon, P; and the model horizon, N:

$$P = N = \text{Max}\left(\frac{5\tau_{rs}}{T} + k_{rs}\right) \text{ where } k_{rs} = \left(\frac{\theta_{rs}}{T} + 1\right) \quad (r = 1,2, \dots, R; s = 1,2, \dots, S)$$

3. Compute a control horizon, M:

$$M = \text{Max}\left(\frac{\tau_{rs}}{T} + k_{rs}\right) \quad (r = 1,2, \dots, R; s = 1,2, \dots, S)$$

4. Select the controlled variable weights, γ_r^2 , to scale process variable units to be the same.

5. Compute the move suppression coefficients, λ_r^2 :

$$\lambda_r^2 = \frac{M}{10} \sum_{r=1}^R \left[\gamma_r^2 K_{rs}^2 \left\{ P - k_{rs} - \frac{3\tau_{rs}}{2T} + 2 - \frac{(M-1)}{2} \right\} \right] \quad (s = 1,2, \dots, S)$$

6. Implement DMC using the traditional step response matrix of the actual process and the initial values of the parameters computed in steps 1-6.

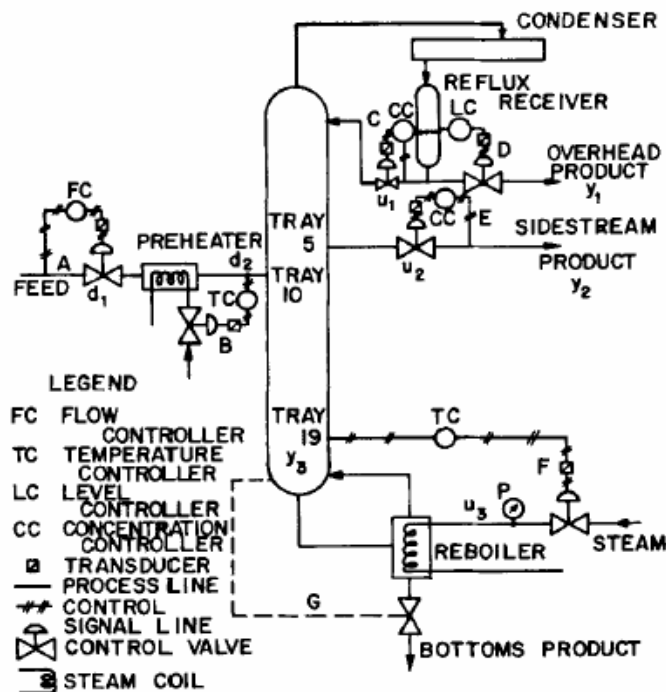


Figure 1. Schematic diagram of OLMR distillation column [7]

where y is a vector of outputs, u a vector of controls, and d a vector of disturbance variables. The outputs are y_1 = overhead ethanol mole fraction, y_2 = side stream ethanol mole fraction composition), and y_3 = tray #19 temperature (°C) (corresponding to bottoms composition). The inputs are u_1 = reflux flow rate, gpm (m³/s), u_2 = side stream product flow rate, gpm (m³/s), and u_3 = reboiler stream pressure, psig (kPa); d_1 = feed flow rate, gpm (m³/s) and d_2 = feed temperature, °C.

The column model takes the Eq.1 form. The transfer functions of manipulated variable ($G_p(s)$) and disturbance variable ($G_d(s)$) are shown in Table 2 (time unit is minutes).

Several methods to get the optimum performance of the plant have been implemented. Ogunnaike et al. (1983) have implemented PI controller with delay compensator to improve the control performance of PI controller without delay. Luyben, et al (1988) have developed the direct method of PID controller called BLT-1 based on Ziegler-Nichols method. Based on Luyben's BLT-1 controller, the new methods were called BLT-2, BLT-3, and BLT-4. The methods have proved their ability to reach a better performance than the previous method. Even, BLT-4 was better than DMC controller. Based on an effective open-loop (EOP) model, Huang et al. (2003) have improved BLTs methods especially in 2x2 processes, but in the 3x3 process the optimum performance is 56% for Huang et al. method and 44% for BLT-4. In the 4x4 process, Huang et al. is better than BLT, while Lee et al. method is worse than BLT. Huang et al. has 69% of the optimum results while Lee et al. has 31%.

Although BLT was better than DMC (Luyben, et al, 1988), Dougherty and Cooper (2003) have proved that the applications of the non-adaptive DMC tuning strategy in the three 2x2 processes (general transfer function, multi-tank, and distillation column) have a satisfactory performance. So, we will implement this strategy to improve the control performance of OLMR distillation column that has 3x3 matrixes of input and output variables.

Results and Discussions

As shown in Table 1, the tuning strategy uses FOPDT (first-order plus dead-time) model to calculate T_s , P , M , γ_s^2 , and λ_s^2 . Because of OR (3x3) model has one SOPDT (second-order plus dead-time), this transfer function has to be changed into FOPDT. By using PRC (process reaction curve) from step change testing of the SOPDT and applying a method developed by Smith (Marlin, 2000), the FOPDT is obtained ($K_p = 0.87$, $\tau_p = 9.12$, $\theta_p = 0.29$). However, this approximation is only to calculate the MPC tuning parameters, not as a plant. The plant to be simulated is the original model. The calculation produces $T_s = 0.71$, $P = 91$, $M = 30$, $\gamma_s^2 = 1$, and $\lambda_s^2 = 433,520$. MPC TOOLS of MATLAB® was used to simulate the plant (OLMR 3x3 process). In the MPC setting, three parameters of T_s , P , and M are used, while the others used the default setting.

Table 2. Transfer function of manipulated variable ($G_p(s)$) and disturbance variable ($G_d(s)$)

$G_p(s)$			$G_d(s)$	
$\frac{0.66e^{-2.6s}}{6.7s + 1}$	$\frac{-0.61e^{-7.5s}}{8.64s + 1}$	$\frac{-0.0049e^{-2s}}{9.06s + 1}$	$\frac{0.14e^{-12s}}{6.2s + 1}$	$\frac{-0.0011(26.32s + 1)e^{-2.66s}}{(7.85s + 1)(14.63s + 1)}$
$\frac{1.11e^{-6.5s}}{3.25s + 1}$	$\frac{-2.36e^{-7s}}{5.0s + 1}$	$\frac{-0.012e^{-12s}}{7.09s + 1}$	$\frac{0.53e^{-10.5s}}{6.9s + 1}$	$\frac{-0.0032(19.62s + 1)e^{-7.44s}}{(7.29s + 1)(8.94s + 1)}$
$\frac{-34.68e^{-9.2s}}{8.15s + 1}$	$\frac{46.2e^{-9.4s}}{10.9s + 1}$	$\frac{0.87(11.61s + 1)e^{-2.6s}}{(3.89s + 1)(18.8s + 1)}$	$\frac{-11.54e^{-0.6s}}{7.01s + 1}$	$\frac{0.32e^{-2.6s}}{7.76s + 1}$

Figure 2 shows IAE of multi-loop control for OMLR (3x3) process based on PID controller performance using BLT-4 and the first Huang et al.'s model. The control performance of y_1 and y_2 show the satisfactory results, except the y_2 performance when y_3 changed. But, all y_3 performances are poor compared by BLT-4. From nine observed IAEs, five minimum IAEs are for Huang et al. and four minimum IAEs are for BLT-4. This result shows that Huang et al. method of PID controller tuning does not give satisfactory result. Application of MPC controller with tuning its parameters in OMLR 3x3 distillation column model will be evaluated.

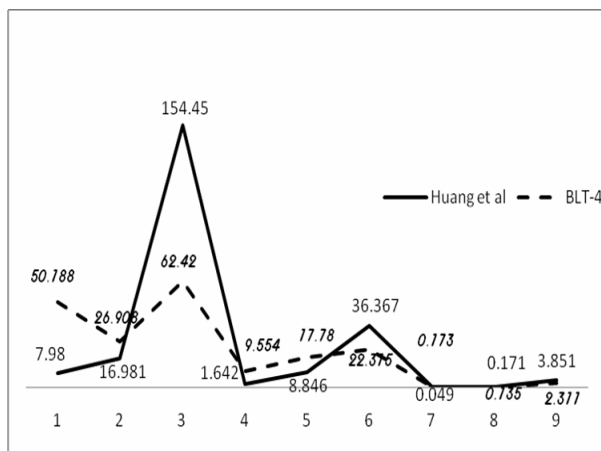


Figure 2. IAE values (in parentheses) of multi-loop control for OR (3x3) process

Figure 3 shows responses of OLMR 3x3 process using MPC with tuning setting. The control performances of this process, compared to application of MPC controller without tuning and application of Huang et al. method are shown in Table 3. Almost performances by MPC tuning setting are better than those the default setting and Huang et al. Only two points where the default setting is better than the tuning setting, they are overshoot of y_2 in y_1 's step change, settling time of y_2 in y_2 's step change, and overshoot of y_1 in y_2 's step change using Huang et al.

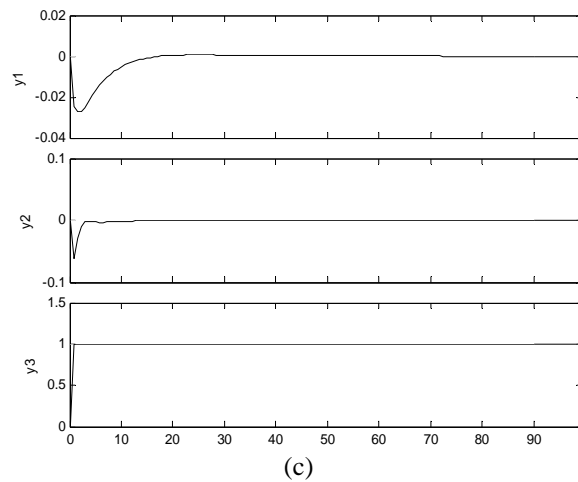
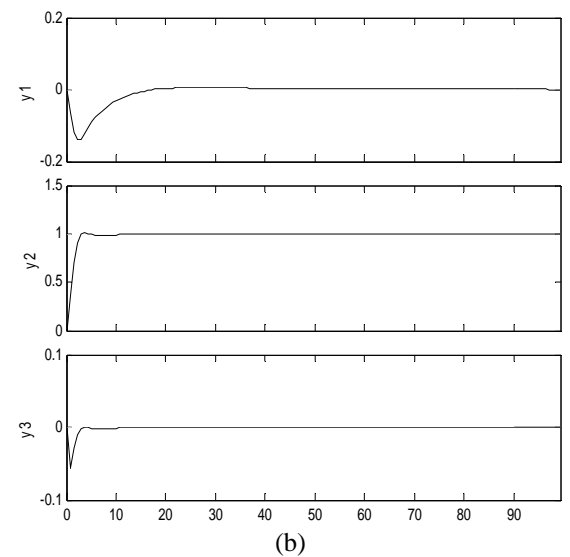
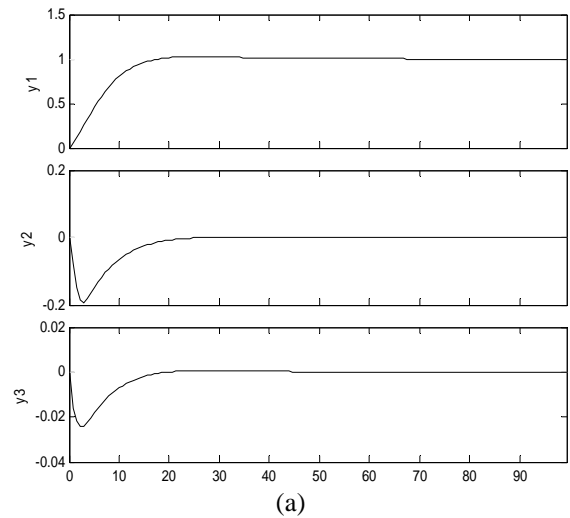


Figure 3. Responses of multi-loop control for OR (3x3) process: step change in y_1 (a), y_2 (b), and y_3 (c) using $T_s=0.71$, $P=91$, $M=30$

Table 4. Performance of MPC controller of OR (3x3 process) model

Step Change	Controlled Variable	Huang et al.		MPC Controller			
		Settling time	Overshoot	Default Setting		Tuning Setting	
				Settling time	Overshoot	Settling time	Overshoot
y1	y1	90.8	0.196	53.4	0.08	47.5	0.02
	y2		1.625		-0.153		-0.193
	y3		-1.557		-0.0386		-0.0241
y2	y1		0.127		-0.156		-0.138
	y2	80.77	0.03	21.7	0.01	31	0.01
	y3		-3.231		-0.151		-0.0552
y3	y1		4.375		-0.0303		-0.0273
	y2		-0.01		-0.0852		-0.0614
	y3	73.08	0.04	6.52	0	3.18	0

Conclusions

MPC controller offers better control performances than PI/PID controller, especially in multivariable processes. Application of MPC controller using tuning strategy in the OLMR (3x3) distillation column model produces a fantastic performance. However, there is still a chance to continue the improvement of MPC controller to solve two bad results in the control performance of OLMR 3x3 process.

References

C. R. Cutler and D. L. Ramaker, (1980), "Dynamic matrix control—a computer control algorithm". *Proceedings of the JACC 1980*. San Francisco.

D.W. Clarke, (1996) "Adaptive Predictive Control". *A Rev. Control*. Vol. 20, pp. 83-94, 1996

Danielle Dougherty and Doug Cooper, (2003), "A Practical Multiple Model Adaptive Strategy for Multivariable Model Predictive Control", *Control Engineering Practice*. (11): 649 – 664, 2003.

Huang et. al. (2003), "A direct method for multi-loop PI/PID controller design". *Journal of Process Control* (13): 769–786.

Ogunnaike, et al., (1983), "Advanced Multivariable Control of a Pilot-Plant Distillation Column". *AIChE Journal* (29/ 4): 632-640.

S.J. Qin and T.A. Badgwell, (2003). "A survey of industrial model predictive control technology". *Control Engineering Practice*. (11): 733 – 764

T. Marlin, (2000), "*Process Control: Designing Processes and Control Systems for Dynamic Performance*", 2nd Edition, McGraw-Hill, New York.

Thomas J. Monica, Cheng-Ching Yu, and William L. Luyben (1988), "Improved Multiloop Single-Input/Single-Output (SISO) Controllers for Multivariable Processes". *Reprinted from I&EC RESEARCH*, 27, 969.