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Model Reference Adaptive Control of Binary Distillation Column Composition using MIT Adaptive Mechanism

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***Abstract*-This paper presents a model reference adaptive control for binary distillation column based on the adaptation mechanism calledMIT rule.The goals are: to present a theoretical calculation procedure of a binary distillation column for simulation. Second, the controller design: a reduced-order linear model is derived suchthat it best reflects the dynamics of the distillation process and used as the reference model.Third, selecting an adaptation mechanism suitable for the system.Simulation results using MATLAB shows that the responses are satisfactory and stability is achieved.**

Index Terms-Model Reference Adaptive Control, Adaptation mechanism,MIT Rule, binary distillation column.

I. INTRODUCTION

Distillation is the most common method of separation of two or more components from a mixture using the relative volatility of the components.It is widely used in chemical industry to separate components with desired purity.Distillation columnsare made up of several components, each of which is usedeither to transfer heat energy or enhance mass transfer.

The distillation column model is computed isbased on a real petroleum project gas processingplant.The feed to the column is a pseudo-mixture of ligas and naphthas and the distillate product is gasoline and bottoms product is LPG.The input variables are feed flow rate F and feed compositionx_F. The manipulated variables are reflux flow rate L and vapor boilup rate V. The quality of the outputproducts is the purityof the distillate, x_D, higher/equal than98% and the impurity of the bottoms, x_B, less/equal than 2%.The basic feed stock data and its actual compositions are basedon the reference [1].

The distillation column has 14 trays and each tray is modeled using mass balance equations of LV control configuration. The control of composition of distillate and bottoms product is not easy as the feed flow cannot be constant. Thus proper controlling techniques should be used.

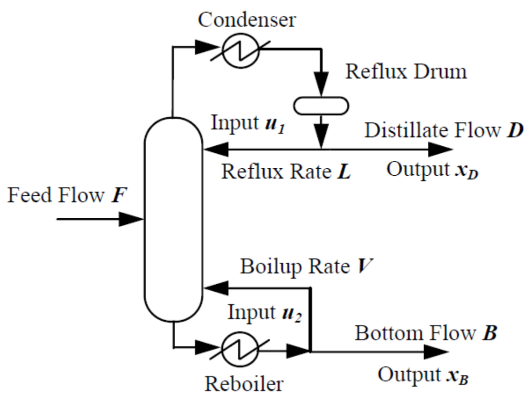

Fig.1: Distillation equipment

The mathematical modelling of distillation column is obtained from Vu Trieu Minh and John Pumwa [2], which is based on physical laws such as mass balance equation. The nonlinear set of equations are linearized and order is reduced for create the reference model for MRAC.The controller design and control theories for adaptive controller proposed by Astrom was found very helpful [4].

II. MATHEMATICAL MODELLING OF BINARY DISTILLATION COLUMN

The purity of distillate or bottom is affected by twofundamental variables: feed split and fractionation. The feed split variable refers to the fraction of the feed that is takenoverhead or out the bottom. The fractionation variable refersto the energy that is put into the column to accomplish theseparation.Feed split variables are used as those have much stronger effect on product compositions than fractionation variables.

The L-V control structure, which is called energy balance structure, used as standard thecontrol structure for dual composition control of distillation [5] has been adopted in this paper. In this control structure,he reflux flow rate L and the boil-up flow rate V are used to control the primary outputsassociated with the product specifications.

The nonlinear system is modelled with the mass balance and component balance equations.The model is simplified under assumptions in [7].Constant relative volatility throughout the column and thevapour-liquid equilibrium relation can be expressed as:

$$y_n = \frac{\alpha x_n}{1+(\alpha-1)x_n} \quad (1)$$

wherex_n : liquid concentrations on n^{th} stage; y_n : vapour concentration on n^{th} stage; α : relative volatility.

The dynamic model can be expressed by the following equations [8]:

Condenser (n=N+2):

$$M_D \dot{x}_n = (V + V_F)y_{n-1} - Lx_n - Dx_n \quad (2)$$

Tray n (n=f+2 to N+1):

$$M\dot{x}_n = (V + V_F)(y_{n-1} - y_n) + L(x_{n+1} - x_n) \quad (3)$$

Tray above the feed flow (n=f+1):

$$M\dot{x}_n = (V)(y_{n-1} - y_n) + L(x_{n+1} - x_n) - (V_F)(y_F - y_n) \quad (4)$$

Tray below the feed flow (n=f):

$$M\dot{x}_n = (V)(y_{n-1} - y_n) + L(x_{n+1} - x_n) - (L_F)(x_F - x_n) \quad (5)$$

Tray n (n=2 to f-1):

$$M\dot{x}_n = (V)(y_{n-1} - y_n) + (L+L_F)(x_{n+1} - x_n) \quad (6)$$

Reboiler (n=1):

$$M_B \dot{x}_1 = (L + L_F)x_2 - Vy_1 - Bx_1 \quad (7)$$

The dynamic model is represented bynonlinear differential equations where x_1=x_B: liquidconcentration in bottom; x_2: liquid concentration in the 1st tray, x_3: liquid concentration in the 2nd tray; … ;x_{15}: liquidconcentration in the 14^{th} tray; and x_{16}=x_D: liquidconcentration in the distillate. The distillation processsimulation is shown in Figure 3.

If there is no disturbance in the operating conditions, thesystem reaches steady state such that the purity of thedistillate product $x_D$is 0.9654 *and* the impurity of the bottomsproduct $x_B$is 0.0375 (Fig 2).

Table. I indicates the steady state values of concentration of$x_n$and $y_n$on each tray.

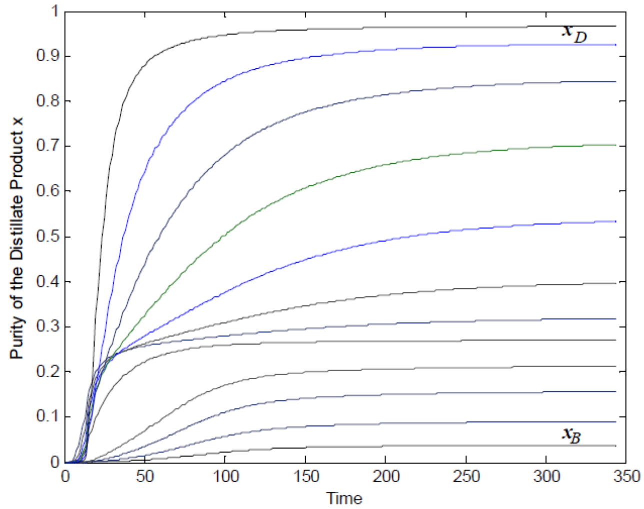

Fig.2: Steady values of concentration of x_n on each tray

Table I: Steady values of concentration of x_n on each tray

Stage	Bottom	Tray 1	Tray 2	Tray 3	Tray 4	Tray 5	Tray 6	Tray 7
x_n	**0.0375**	0.0920	0.1559	0.2120	0.2461	0.2628	0.2701	0.2731
y_n	0.1812	0.3653	0.5120	0.6044	0.6496	0.6694	0.6776	0.6809
Stage	Tray 8	Tray 9	Tray 10	Tray 11	Tray 12	Tray 13	Tray 14	Distillate
x_n	0.2811	0.3177	0.3963	0.5336	0.7041	0.8449	0.9369	**0.9654**
y_n	0.6895	0.7256	0.7885	0.8666	0.9311	0.9687	0.9883	0.9937

The changes in feed flow rates and feedcompositionsof the feed stream can be considered disturbances.Simulations with disturbances indicate that the qualityof the output products deteriorate with more disturbances (Table II).

Table II: Product quality depending on the change in feed flow rate

	Purity of the Distillate Product x_D (%)	Impurity of the Bottoms Product x_B (%)
Normal Feed Rate	96.54	3.75
Reduced Feed Rate 10%	90.23	0.66
Increased Feed Rate 10%	97.30	11.66

The designed system does not achieve the operationalobjective of the product quality (x_D≥ 0.98and x_B≤ 0.02) andthe product quality will get worse dealing with disturbances.Hence we will use an adaptive controller - MRAC to take thesystem from these steady state outputs to the desired outputtargets (x_D≥ 98%andx_B≤ 2%).

III. LINEARISED DISTILLATION COLUMN MODEL

The nonlinear set of differential equations is linearized using Taylors series. In order to obtain a linear mathematical model for a nonlinear system, it is assumed that thevariables deviate only slightly from some operating condition [6].The higher-order terms in theseries $x_n - \bar{x}_n$ can be neglected when the variation $x_n - \bar{x}_n$ is small. The linearization leads to 16 order linear model in state space.

$$\dot{z}(t) = Az(t) + Bu(t)$$

$y(t) = Cz(t)$, where;

$$z(t) = \begin{bmatrix} x_1(t) - \bar{x}_{1\ Steady\ state} \\ x_2(t) - \bar{x}_{2\ Steady\ state} \\ \cdot \\ \cdot \\ x_{16}(t) - \bar{x}_{16\ Steady\ state} \end{bmatrix},$$

$$u(t) = \begin{bmatrix} L(t) - \bar{L}(t)_{Steadystate} & = dL \\ V(t) - \bar{V}(t)_{Steadystate} & = dV \end{bmatrix},$$

$$y(t) = \begin{bmatrix} x_1(t) - \bar{x}_{1\ Steady\ state} & = dx_B \\ x_{16}(t) - \bar{x}_{16\ Steady\ state} & = dx_D \end{bmatrix} (8)$$

The full order model can be reduced to second order as follows:

$$\begin{bmatrix} dx_D \\ dx_B \end{bmatrix} = \frac{1}{1+\tau_c} G(0) \begin{bmatrix} dL \\ dV \end{bmatrix} \ ; (9)$$

where: $G(0)$ is the steady state gain: $G(0) = -CA^{-1}B$,$\tau_c$isthe time constant:

$$\tau_c = \frac{M_I}{I_S \ln s} + \frac{M_D(1-x_D)x_D}{I_S} + \frac{M_B(1-x_B)x_B}{I_S} \qquad (10)$$

whereM_I(kmole): the total holdup of liquid inside the column; M_D(kmole): liquid holdup in the condenser; M_B(kmole): liquidholdup in the reboiler; I_S: the "impurity sum"; and S : theseparation factor.

As the result of calculation, the reduced-order linearmodel of the plant is a first order system with a time constant ofτ_c=1.9588 is: [3]

$$\begin{bmatrix} dx_D \\ dx_B \end{bmatrix} = \frac{1}{1+1.9588s} \begin{bmatrix} 0.0042 & -0.0060 \\ -0.0050 & 0.0072 \end{bmatrix} \begin{bmatrix} dL \\ dV \end{bmatrix} (11)$$

IV. MODEL REFERNCE ADAPTIVE CONTROL SYSTEM

The two main reasons for applying adaptive control widely in petroleum industriesare:First, most of chemical processes arenonlinear and the linearized models are used to design thecontrollers, so that the controller must change and adapt to themodel-plant mismatch; Second, most of the chemical processes arenon-stationary or their characteristics are changed with time,this leads again to adapt the changing control parameters.

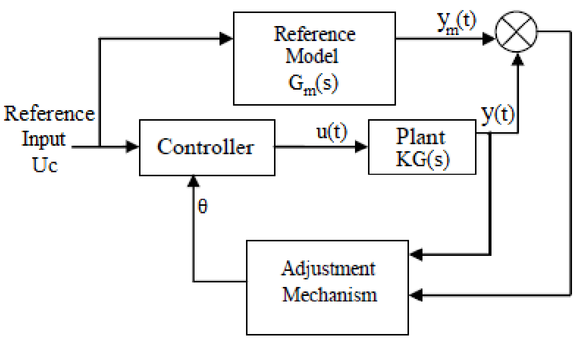

Fig 3: MRAC block diagram

The model reference adaptive control is composed of two loops. The inner loop consists of an ordinary feedback control loop. The outer loop consists of the adaptation mechanism and also looks like a feedback loop. The model output plays the role of the setpoint while the process output is the actual measurement [9].

The mechanism for adjusting the controller parameters in a model reference adaptive system can be obtained in two ways: by using gradient method or by applying stability theory [4].The desired behaviour of the system is specified by the reference model,and the parameters of the controller are adjusted on the error,which is the difference between the output of the closed loop system and the reference model.

We found that the reduced-order linear model in(11) can also maintain the similar steady state outputs as thebasic nonlinear model. Now we use this model as reference model MRAC totake the process plant from these steady state outputs to thedesired targets amid the disturbances and the plant-modelmismatches as the influence of the feed stock disturbances.

The MIT Rule is adopted as the adaptation mechanism is this paper which is also called as the gradient method. According to MIT Rule there will be one controller parameter θ that can be adjusted. The parameter can be adjusted in such a way that the loss function

$$J(\theta) = \frac{1}{2}e^2 \quad (12)$$

is minimized where e is the error between the process output and the model output y_m. To minimize J,the parameters have to be changed in the direction of negative gradient of J, that is,

$$\frac{d\theta}{dt} = -\gamma \frac{\partial J}{\partial \theta} = -\gamma e \frac{\partial e}{\partial \theta} (13)$$

The feedforward control law using MIT Rule is:

$$u = \theta u_c \quad (14)$$

and the parameter update rule is:

$$\frac{d\theta}{dt} = -\gamma y_m e \ (15)$$

The feedback control law using MIT Rule is:

$$u = \theta_1 u_c - \theta_2 y (16)$$

and the parameter update rule is:

$$\frac{d\theta_1}{dt} = -\gamma u_c e \qquad (17)$$

$$\frac{d\theta_2}{dt} = -\gamma y e \qquad (18)$$

V. SIMULATION RESULTS AND DISCUSSIONS

The reduced-order linear model in (11) can also maintain thesimilar steady state outputs as the basic nonlinear model. This model is used as anMRAC totake the process plant from these steady state outputs (x_D =0.9654 andx_B= 0.0375) to the desired targets ($0.98 \leq x_D \leq 1$and $0 \leq x_B \leq 0.02$) amid the disturbancesand the plant-model mismatches as the influence of the feed stock disturbances.

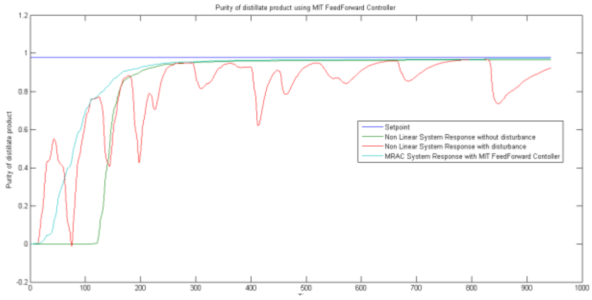

Fig 4: Distillate purity of MRAC FF System

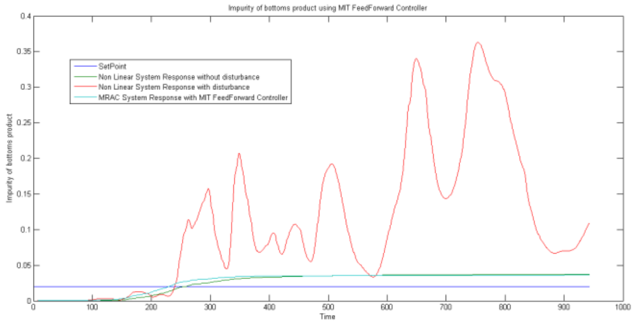

Fig 5: Bottoms impurity of MRAC FF System

Simulation in Fig 4, 5 shows that the controlled outputs of MRAC system with feedforward controller using MIT Rule x_D and x_B are stable and tracking to the model outputs and the reference setpoints amid thedisturbances and the plant-model mismatches.

While the Fig 6, 7 shows that the outputs x_D and x_B are stable and tracking to the model outputs and the reference setpoints are more accurately controlled by using feedback controller. The same can be inferenced from table III.

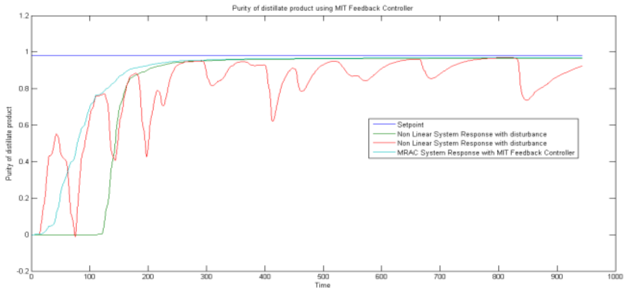

Fig 6: Distillate purity of MRAC FB System

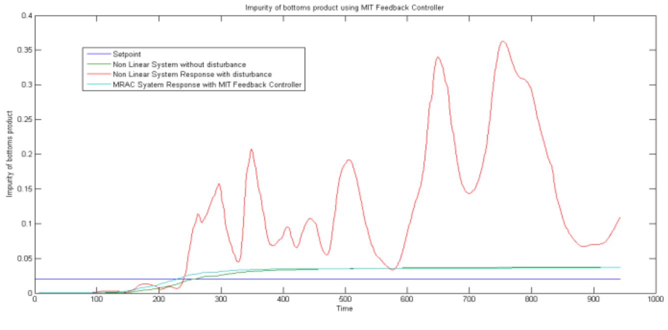

Fig 7: Bottoms impurity of MRAC FB System

Table III: Comparision of MRAC Syatem, Non Linear System, Non Linear System with disturbance

	Purity of distillate Product(mole fraction)	Impurity of bottoms Product(mole fraction)
NonLinear Model	.9654	.0375
NonLinear Model with disturbance	.9225	.1087
Feedforward MIT	.9708	.03612
Feedback MIT	.9719	.03703

VI. CONCLUSION

In this paper, an MRAC controller is designed using the MIT Rule for both feedforward gain and feedback gain even under disturbances. Then both feedback and feedforward controller performance are compared and found feedback controller is more efficient. The empirical evidence shows that the system is stable for small values of adaptation gain in the case of MIT rule but high gains lead to instability.

VII. FUTURE SCOPE

The main drawback of MIT feedforward and feedback controller is both the distillate product and bottoms product composition are not controlled to desired value simultaneously. By designing MRAC PID the performance can be improved so that the desired composition is achieved. Also decoupling can incorporated so that distillate and bottoms product purity can be controlled separately and more accurately.

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