Research Article

Modeling and Control of Distillation Column in a Petroleum Process

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This paper introduces a calculation procedure for modeling and control simulation of a condensate distillation column based on the energy balance (*L-V*) structure. In this control, the reflux rate *L* and the boilup rate *V* are used as the inputs to control the outputs of the purity of the distillate overhead and the impurity of the bottom products. The modeling simulation is important for process dynamic analysis and the plant initial design. In this paper, the modeling and simulation are accomplished over three phases: the basic nonlinear model of the plant, the full-order linearised model, and the reduced-order linear model. The reduced-order linear model is then used as the reference model for a model-reference adaptive control (MRAC) system to verify the applicable ability of a conventional adaptive controller for a distillation column dealing with the disturbance and the model-plant mismatch as the influence of the plant feed disturbances.

1. Introduction

Distillation is the most popular and important separation method in the petroleum industries for purification of final products. Distillation columns are made up of several components, each of which is used either to transfer heat energy or to enhance mass transfer. A typical distillation column contains a vertical column where trays or plates are used to enhance the component separations, a reboiler to provide heat for the necessary vaporization from the bottom of the column, a condenser to cool and condense the vapor from the top of the column, and a reflux drum to hold the condensed vapor so that liquid reflux can be recycled back from the top of the column.

Calculation of the distillation column in this paper is based on a real petroleum project to build a gas processing plant to raise the utility value of condensate. The nominal capacity of the plant is 130 000 tons of raw condensate per year based on 24 operating hours per day and
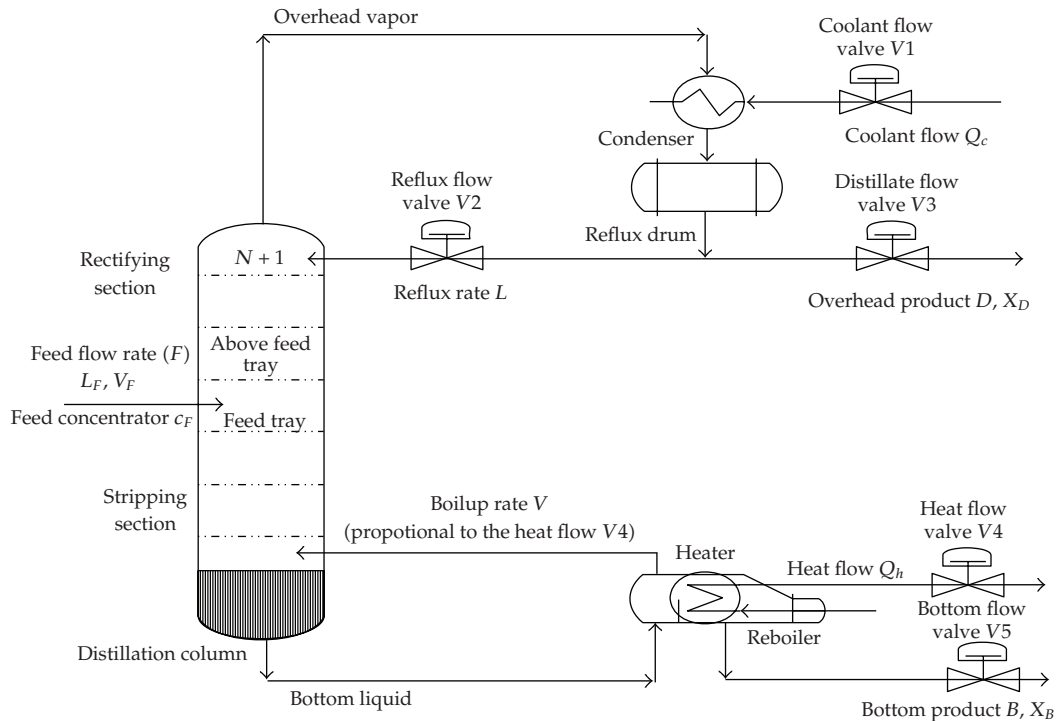
Figure 1: Distillation flowsheet.

350 working days per year. The quality of the output products is the purity of the distillate, x_D, higher than or equal to 98% and the impurity of the bottoms, x_B, less/equal than 2%. The basic feed stock data and its actual compositions are based on [1].

Most of distillation control systems, either conventional or advanced, assume that the column operates at a constant pressure. Pressure fluctuations make the control more difficult and reduce the performance. The L-V structure, which is called energy balance structure, can be considered as the standard control structure for a dual composition control distillation. In this control structure the liquid flow rate L and the vapor flow rate V are the control inputs. The objective of the controller is to maintain the product outputs concentrations x_B and x_D despite the disturbance in the feed flow F and the feed concentration c_F (Figure 1).

The goals of this paper are twofold: first, to present a theoretical calculation procedure of a condensate column for simulation and analysis as an initial step of a project feasibility study, and second, for the controller design: a reduced-order linear model is derived such that it best reflects the dynamics of the distillation process and used as the reference model for a model-reference adaptive control (MRAC) system to verify the ability of a conventional adaptive controller for a distillation process dealing with the disturbance and the plant-model mismatch as the influence of the feed disturbances.

In this study, the system identification is not employed since experiments requiring a real distillation column are still not implemented yet. So that a process model based on experimentation on a real process cannot be done. A mathematical modeling based on physical laws is performed instead. Further, the MRAC controller model is not suitable for handling the process constraints on inputs and outputs as shown in [2] for a coordinator

Table 1: The main streams.

Stream	Condensate	LPG	Raw gasoline
Temperature (°C)	118	46	144
Pressure (atm)	4.6	4.0	4.6
Density (kg/m^3)	670	585	727
Volume flow rate (m^3/h)	22.76	8.78	21.88
Mass flow rate (kg/h)	15480	5061	10405
Plant capacity (ton/year)	130000	43000	87000

model predictive control (MPC). In this paper, the calculations and simulations are implemented by using MATLAB (version 7.0) software package.

2. Process Model and Simulation

The feed can be considered as a pseudobinary mixture of Ligas (iso-butane, n-butane and propane) and Naphthas (iso-pentane, n-pentane, and higher components). The column is designed with N = 14 trays. The model is simplified by lumping some components together (pseudocomponents) and modeling of the column dynamics is based on these pseudocomponents only [3].

For the feed section, the operating pressure at the feed section is given at 4.6 atm. The feed temperature for the preheater is the temperature at which the required phase equilibrium is established. Consulting the equilibrium flash vaporization (EFV) curve at 4.6 atm, the required feed temperature is selected at 118°C corresponding to the point of 42% of the vapor phase feed rate V_F.

For the rectifying section, the typical pressure drop per tray is 6.75 kPa. Thus, the pressure at the top section is 4 atm. Also consulting the Cox chart, the top section temperature is determined at 46°C. Then, we can calculate the reflux flow rate L via the energy balance equation.

For the stripping section, the column base pressure is approximately the pressure of the feed section (4.6 atm) because the pressure drop across this section is neglected. Consulting the EFV curve and the Cox chart, the equilibrium temperature at this section (4.6 atm) is determined at 144°C. Then, we can calculate the reboiler duty or the heat input Q_B to increase the temperature of stripping section from 118°C to 144°C.

Table 1 summarizes the initial calculated data for the main streams of input feed flow rate (Condensate), output distillate overhead product: (LPG) and output bottom product (Raw gasoline).

The vapor boilup V generated by the heat input to the reboiler is calculated as [4]: $V = (Q_B - Bc_B(t_B - t_F))/\lambda$ (kmole/h), where Q_B is the heat input (kJ/h); B is the flow rate of bottom product (kg/h); c_B is the specific heat capacity (kJ/kg·°C); t_F is the inlet temperature (°C); t_B is the outlet temperature (°C); λ is the latent heat or the heat of vaporization (kJ/kg). The latent heat at any temperature is described in terms of the latent heat at the normal boiling point [5] $\lambda = \gamma\lambda_B(T/T_B)$, where λ is the latent heat at the absolute temperature T in degrees Rankine (°R); λ_B is the latent heat at the absolute normal boiling point T_B in degrees Rankine (°R); and γ is the correction factor obtained from the empirical chart.

Major design parameters to determine the liquid holdup on tray, column base and reflux drum are calculated mainly based on [6–8].

Velocity of vapor phase is arising in the column $w_n = C\sqrt{(\rho_L - \rho_G)/\rho_G}$ (m/s), where ρ_L (kg/m³) is the density of liquid phase; ρ_G (kg/m³) is the density of vapor phase; C is the correction factor depending flow rates of two-phase flows.

The actual velocity ω is normally selected at $\omega = (0.80 - 0.85)w_n$ for paraffinic vapor.

The diameter of the column is calculated on the formula: $D_k = \sqrt{4V_m/3600\pi\omega}$ (m), where V_m (kmole/h) is the mean flow of vapor in the column.

The holdup in the column base is $M_B = (\pi H_{NB} D_k^2/4)(\rho_B/(MW)_B)$ (kmole), where H_{NB} (m) is the normal liquid level in the column base; $(MW)_B$ is the molar weight of the bottom product (kg/kmole); ρ_B is the density of the bottom product (kg/m³).

Similarly, the holdup on each tray is $M = (0.95\pi h_T D_k^2/4)(\rho_T/(MW)_T)$ (kmole), where h_T is the average depth of clear liquid on a tray (m); $(MW)_T$ is the molar weight of the liquid holdup on a tray (kg/kmole); ρ_T is the mean density of the liquid holdup on a tray (kg/m³). And the holdup in the reflux drum $M_D = 5(L_f + V_f)/60$ (kmole), where L_f is the reflux flow rate (kmole/h); V_f is the distillate flow rate (kmole/h).

The rate of accumulation of material in a system is equal to the amount entered and generated, less the amount leaving and consumed within the system. The model is simplified under assumptions in [9].

(i) Constant relative volatility throughout the column and the vapor-liquid equilibrium relation can be expressed by

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n}, \tag{2.1}$$

where x_n is the liquid concentration on nth stage; y_n is the vapor concentration on nth stage; α is the relative volatility.

(ii) The overhead vapor is totally condensed.

(iii) The liquid holdups on each tray, the condenser, and the reboiler are constant and perfectly mixed.

(iv) The holdup of vapor is negligible throughout the system

(v) The molar flow rates of the vapor and liquid through the stripping and rectifying sections are constant.

Under these assumptions, the dynamic model can be expressed by the following equations:

(i) condenser $(n = N + 2)$:

$$M_D \dot{x}_n = (V + V_F)y_{n-1} - Lx_n - Dx_n, \tag{2.2}$$

(ii) tray $n(n = f + 2$ to $N + 1)$:

$$M\dot{x}_n = (V + V_F)(y_{n-1} - y_n) + L(x_{n+1} - x_n), \tag{2.3}$$

(iii) tray above the feed flow $(n = f + 1)$:

$$M\dot{x}_n = V(y_{n-1} - y_n) + L(x_{n+1} - x_n) + V_F(y_F - y_n), \tag{2.4}$$

Table 2: The steady state values of concentrations x_n and y_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

Table 3: Product quality depending on the change of the feed rates.

	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

(iv) tray below the feed flow ($n = f$):

$$M\dot{x}_n = V(y_{n-1} - y_n) + L(x_{n+1} - x_n) + L_F(x_F - x_n), \tag{2.5}$$

(v) tray $n(n = 2$ to $f - 1)$:

$$M\dot{x}_n = V(y_{n-1} - y_n) + (L + L_F)(x_{n+1} - x_n), \tag{2.6}$$

(vi) reboiler ($n = 1$):

$$M_B\dot{x}_1 = (L + L_F)x_2 - Vy_1 - Bx_1. \tag{2.7}$$

Although the model is simplified, the representation of the distillation system is still nonlinear due to the vapor-liquid equilibrium relationship between y_n and x_n in (2.1).

The distillation process simulation is done using Matlab Simulink as shown in Figure 2. The dynamic model is represented by a set of 16 nonlinear differential equations: $x_1 = x_B$ is the liquid concentration in bottom; x_2 is the liquid concentration in the 1st tray, x_3 is the liquid concentration in the 2nd tray; ...; x_{15} is the liquid concentration in the 14th tray; and $x_{16} = x_D$ is the liquid concentration in the distillate.

If there are no disturbance in the operating conditions as shown in Figure 3, the system is to reach the steady state such that the purity of the distillate product $x_D$equals 0.9654 and the impurity of the bottoms product $x_B$equals 0.0375.

Table 2 indicates the steady-state values of concentration of x_n and y_n on each tray.

Since the feed stream depends on the upstream processes, the changes of the feed stream can be considered as disturbances including the changing in feed flow rates and feed compositions. Simulations with these disturbances indicate that the quality of the output products gets worse if the disturbances exceed some certain ranges as shown in Table 3.

The designed system does not achieve the operational objective of the product quality ($x_D \geq 0.98$ and $x_B \leq 0.02$) and the product quality will get worse dealing with disturbances.
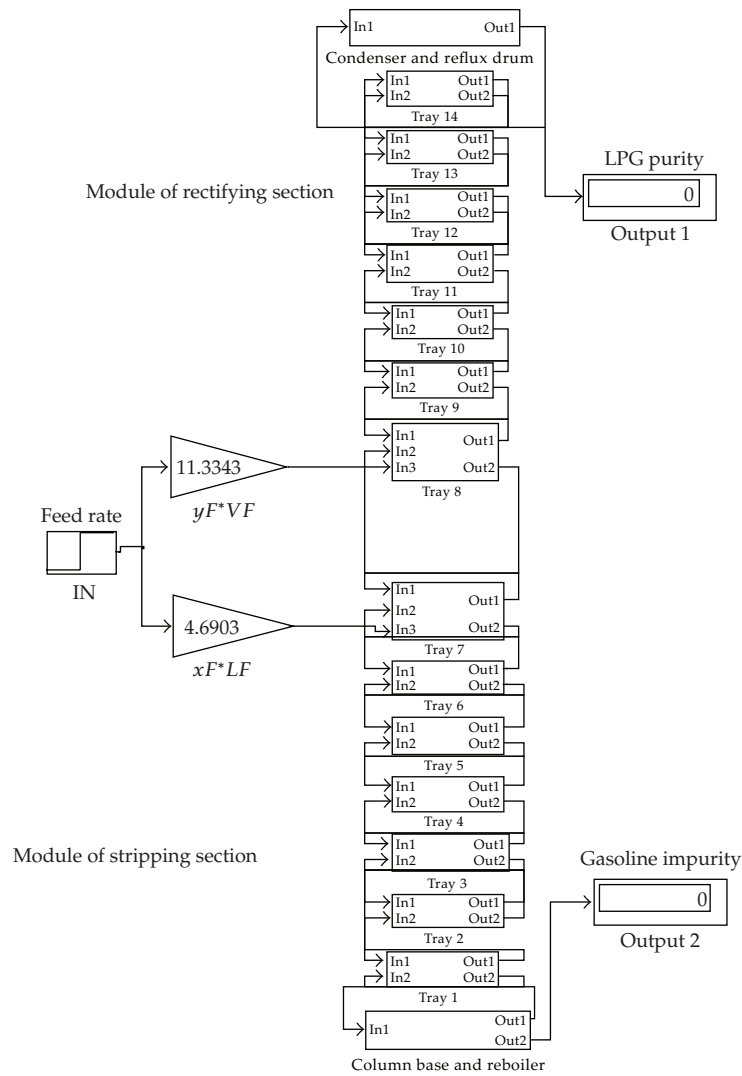
Figure 2: Model simulation with Matlab Simulink.

Hence we will use an adaptive controller—MRAC—to take the system from these steady-state outputs of $x_D = 0.9654$ and $x_B = 0.0375$ to the desired output targets.

3. Linearization of the Distillation Process

In order to obtain a linear control model for this nonlinear system, we assume that the variables deviate only slightly from some operating conditions [10]. Then the nonlinear equation in (2.1) can be expanded into a Taylor's series. If the variation $x_n - \overline{x}_n$ is small,
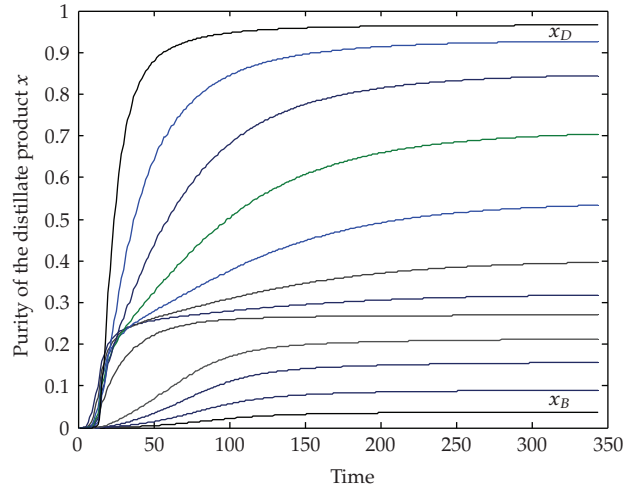
Figure 3: The steady-state values of concentrations x_n on each tray.

we can neglect the higher-order terms in $x_n - \overline{x}_n$. The linearization of the distillation column leads to a 16th-order linear model in the state space form:

$$\dot{z}(t) = Az(t) + Bu(t),$$
$$y(t) = Cz(t), \tag{3.1}$$

where

$$z(t) = \begin{bmatrix} x_1(t) - \overline{x}_{1 \text{ Steady State}} \\ x_2(t) - \overline{x}_{2 \text{ Steady State}} \\ \vdots \\ x_{16}(t) - \overline{x}_{16 \text{ Steady State}} \end{bmatrix}, \qquad u(t) = \begin{bmatrix} L(t) - \overline{L}_{\text{Steady State}} \\ V(t) - \overline{V}_{\text{Steady State}} \end{bmatrix}, \tag{3.2}$$

$$y(t) = \begin{bmatrix} x_1(t) - \overline{x}_{1 \text{ Steady State}} \\ x_{16}(t) - \overline{x}_{16 \text{ Steady State}} \end{bmatrix}.$$

The matrix A elements (n for each stage) are
    (i) reboiler:

$$\text{for } n = 1, \qquad a_{1,1} = -\frac{\left(K_1\overline{V} + B\right)}{M_B}, \qquad a_{1,2} = \frac{\left(\overline{L} + \overline{L}_F\right)}{M_B}, \tag{3.3}$$

(ii) stripping section, tray $1 \div 6$:

$$\text{for } n = 2 \div 7, \qquad a_{n,n-1} = \frac{\left(K_{n-1}\overline{V}\right)}{M}, \qquad a_{n.n} = -\frac{\left(K_n\overline{V} + \overline{L} + L_F\right)}{M}, \qquad a_{n,n+1} = \frac{\left(\overline{L} + L_F\right)}{M},$$

$$(3.4)$$

(iii) feeding section, tray $7 \div 8$:

$$\text{for } n = 8, \qquad a_{8,7} = \frac{\left(K_7\overline{V}\right)}{M}, \qquad a_{8.8} = -\frac{\left(K_8\overline{V} + \overline{L} + L_F\right)}{M}, \qquad a_{8,9} = \frac{\left(\overline{L}\right)}{M},$$

$$\text{for } n = 9, \qquad a_{9,8} = \frac{\left(K_8\overline{V}\right)}{M}, \qquad a_{9.9} = -\frac{\left(K_9\overline{V} + \overline{L}\right)}{M}, \qquad a_{9,10} = \frac{\left(\overline{L}\right)}{M},$$

$$(3.5)$$

(iv) rectifying section, tray $9 \div 14$:

$$\text{for } n = 10 \div 15, \qquad a_{n,n-1} = \frac{\left(K_{n-1}\left(\overline{V} + V_F\right)\right)}{M}, \qquad a_{n.n} = -\frac{\left(K_n\left(\overline{V} + V_F\right) + \overline{L}\right)}{M},$$

$$a_{n,n+1} = \frac{\left(\overline{L}\right)}{M}$$

$$(3.6)$$

(v) condenser:

$$\text{for } n = 16, \qquad a_{16,15} = \frac{\left(K_{15}\left(\overline{V} + V_F\right)\right)}{M_D}, \qquad a_{16,16} = -\frac{\left(\overline{L} + D\right)}{M_D}, \qquad (3.7)$$

where K_n is the linearized Vapor-Liquid Equilibria (VLE) constant:

$$K_n = \frac{dy_n}{dx_n} = \frac{\alpha}{(1 + (\alpha - 1)x_n)^2} = \frac{5.68}{(1 + 4.68x_n)^2}. \qquad (3.8)$$

The matrix B elements are

$$\text{for } n = 1, \qquad b_{1,1} = \frac{(\overline{x}_2)}{M_B}L, \qquad b_{1,2} = -\frac{(\overline{y}_1)}{M_B}V,$$

$$\text{for } n = 2 \div 15, \qquad b_{n,1} = \frac{(\overline{x}_{n+1} - \overline{x}_n)}{M}L, \qquad b_{n,2} = -\frac{(\overline{y}_n - \overline{y}_{n-1})}{M}V, \qquad (3.9)$$

$$\text{for } n = 16, \qquad b_{16,1} = -\frac{(\overline{x}_{16})}{M_D}L, \qquad b_{16,2}\frac{(\overline{y}_{15})}{M_D}V.$$

The output matrix C is

$$C = \begin{vmatrix} 1 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 \\ 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 0 & 1 \end{vmatrix}. \tag{3.10}$$

The full-order linear model which represents a two inputs-two outputs plant in equation in (3.3) can be expressed as a reduced order linear model as in [11, 12]:

$$\begin{bmatrix} x_D \\ x_B \end{bmatrix} = \frac{1}{1 + \tau_c s} G(0) \begin{bmatrix} L \\ V \end{bmatrix}, \tag{3.11}$$

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

$$\tau_c = \frac{M_I}{I_s \ln S} + \frac{M_D(1 - x_D)x_D}{I_s} + \frac{M_B(1 - x_B)x_B}{I_s}, \tag{3.12}$$

where M_I (kmole) is the total holdup of liquid inside the column; M_D (kmole) is the liquid holdup in the condenser; M_B (kmole) is the liquid holdup in the reboiler; I_s is the "impurity sum"; S is the separation factor.

As the result of calculation, the reduced-order linear model of the plant is a first-order system with a time constant of $\tau_c = 1.9588(h)$:

$$\begin{bmatrix} x_D \\ x_B \end{bmatrix} = \frac{1}{1 + 1.9588s} \begin{bmatrix} 0.0042 & -0.0062 \\ -0.0052 & 0.0072 \end{bmatrix} \begin{bmatrix} L \\ V \end{bmatrix}. \tag{3.13}$$

Equation (3.13) is equivalent to the following linear model in state space:

$$\dot{z}_r(t) = \begin{vmatrix} -0.5105 & 0 \\ 0 & -0.5105 \end{vmatrix} z_r(t) + \begin{vmatrix} 1 & 0 \\ 0 & 1 \end{vmatrix} u(t),$$

$$y_r(t) = \begin{vmatrix} 0.0021 & -0.0031 \\ -0.0026 & 0.0037 \end{vmatrix} z_r(t), \tag{3.14}$$

Where $z_r = \begin{bmatrix} z_{r1} \\ z_{r2} \end{bmatrix}$ are state variable, $u = \begin{bmatrix} dL \\ dV \end{bmatrix}$ are two manipulated inputs, and $y_r = \begin{bmatrix} dx_B \\ dx_D \end{bmatrix}$ are two outputs of LPG and gasoline product.

Stability test. The system is asymptotically stable since all eigenvalues of the state matrix are in the left half of the complex plane ($[-0.5105, -0.5105]$).

4. MRAC Building and Simulation

Adaptive control system is the ability of a controller which can adjust its parameters in such a way as to compensate for the variations in the characteristics of the process. Adaptive control
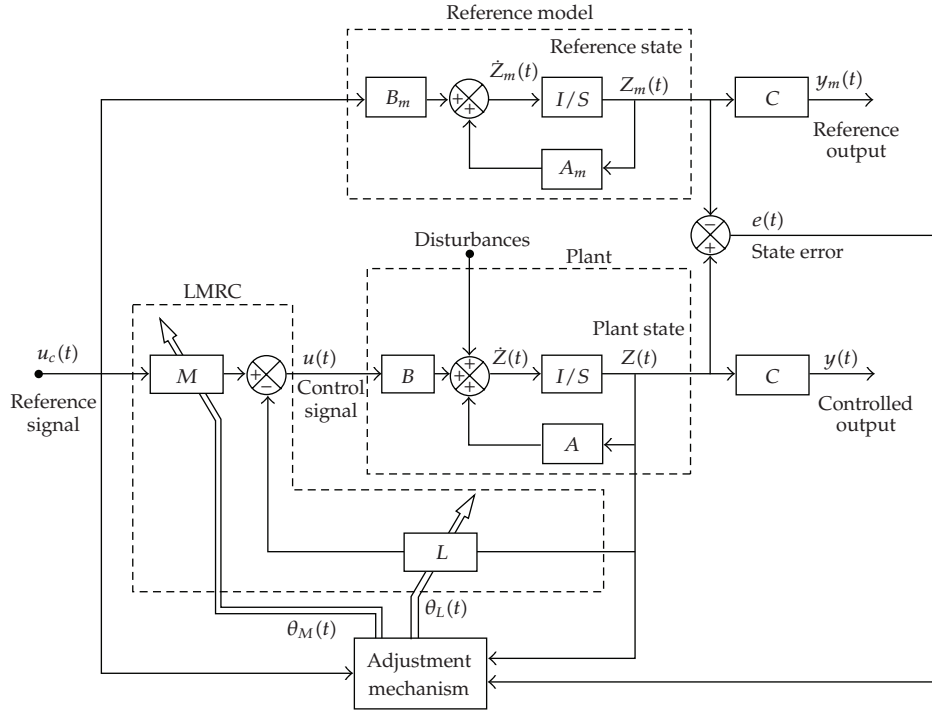
Figure 4: MRAC block diagram.

is widely applied in petroleum industries because of the two main reasons: firstly, most of processes are nonlinear and the linearized models are used to design the controllers, so that the controller must change and adapt to the model-plant mismatch; secondly, most of the processes are nonstationary or their characteristics are changed with time, and this leads again to adapt the changing control parameters.

The general form of an MRAC is based on an inner-loop Linear Model Reference Controller (LMRC) and an outer adaptive loop shown in Figure 4. In order to eliminate errors between the model and the plant and the controller is asymptotically stable, MRAC will calculate online the adjustment parameters in gains L and M by $\theta_L(t)$ and $\theta_M(t)$ as detected state error $e(t)$ when changing A, B in the process plant.

Simulation program is constructed using Maltab Simulink with the following data.

(1) *Process Plant:*

$$\dot{z} = Az + Bu + \text{noise},$$
$$y = Cz, \tag{4.1}$$

where $A = \begin{bmatrix} \alpha_1 & 0 \\ 0 & \alpha_2 \end{bmatrix}$, $B = \begin{bmatrix} \beta_1 & 0 \\ 0 & \beta_2 \end{bmatrix}$, $C = \begin{bmatrix} 0.004 & -0.007 \\ -0.0011 & 0.0017 \end{bmatrix}$, and α_1, α_2, β_1, β_2 are changing and dependent on the process dynamics.

(2) *Reference Model:*

$$\dot{z}_m = A_m z_m + B_m u_c,$$
$$y_m = C_m z_m, \tag{4.2}$$

where $A_m = \begin{bmatrix} -0.2616 & 0 \\ 0 & -0.2616 \end{bmatrix}$, $B_m = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$, $C_m = \begin{bmatrix} 0.004 & -0.007 \\ -0.0011 & 0.0017 \end{bmatrix}$

(3) *State Feedback:*

$$u = M u_c - Lz, \tag{4.3}$$

where $L = \begin{bmatrix} \theta_1 & 0 \\ 0 & \theta_2 \end{bmatrix}$ and $M = \begin{bmatrix} \theta_3 & 0 \\ 0 & \theta_4 \end{bmatrix}$.

(4) *Closed Loop:*

$$\dot{z} = (A - BL)z + BM u_c = A_c(\theta)z + B_c(\theta)u_c \tag{4.4}$$

(5) *Error Equation:*

$e = z - z_m = \begin{bmatrix} e_1 \\ e_2 \end{bmatrix}$ is a vector of state errors,

$$\dot{e} = \dot{z} - \dot{z}_m = Az + Bu - A_m z_m - B_m u_c = A_m e + (A_c(\theta) - A_m)z + (B_c(\theta) - B_m)u_c$$
$$= A_m e + \Psi\left(\theta - \theta^0\right), \tag{4.5}$$

where $\Psi = \begin{bmatrix} -\beta_1 z_1 & 0 & \beta_1 u_{c1} & 0 \\ 0 & -\beta_2 z_2 & 0 & \beta_2 u_{c2} \end{bmatrix}$.

(6) *Lyapunov Function:*

$$V(e, \theta) = \frac{1}{2}\left(\gamma e^T P e + \left(\theta - \theta^0\right)^T \left(\theta - \theta^0\right)\right), \tag{4.6}$$

where γ is an adaptive gain and P is a chosen positive matrix.

(7) *Derivative Calculation of Lyapunov Function:*

$$\frac{dV}{dt} = -\frac{\gamma}{2}e^T Q e + \left(\theta - \theta^0\right)^T \left(\frac{d\theta}{dt} + \gamma \Psi^T P e\right), \tag{4.7}$$

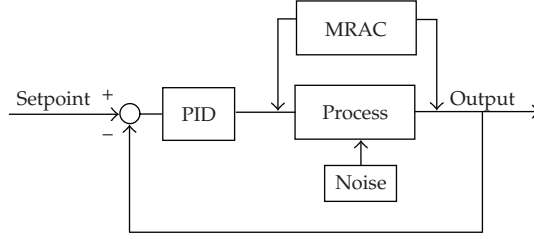where $Q = -A_m^T P - P A_m$.

Figure 5: Adaptive controller with MRAC and PID.

For the stability of the system, $dV/dt < 0$, we can assign the second item $(\theta - \theta^0)^T((d\theta/dt) + \gamma \Psi^T Pe) = 0$ or $d\theta/dt = -\gamma \Psi^T Pe$. Then we always have $dV/dt = -(\gamma/2)e^T Qe$. If we select a positive matrix $P > 0$, for instance, $P = \begin{bmatrix} 1 & 0 \\ 0 & 2 \end{bmatrix}$, then we have $Q = -A_m^T P - PA_m = \begin{bmatrix} 0.5232 & 0 \\ 0 & 1.0465 \end{bmatrix}$. Since matrix Q is obviously positive definite, then we always have $dV/dt = -(\gamma/2)e^T Qe < 0$ and the system is stable with any plant-model mismatches.

(8) *Parameters Adjustment:*

$$\frac{d\theta}{dt} = -\gamma \begin{bmatrix} -\beta_1 z_1 & 0 \\ 0 & -\beta_2 z_2 \\ \beta c_1 u_1 & 0 \\ 0 & \beta_2 u_{2c} \end{bmatrix} [P] \begin{bmatrix} e_1 \\ e_2 \end{bmatrix} = \begin{bmatrix} d\theta_1/dt \\ d\theta_2/dt \\ d\theta_3/dt \\ d\theta_4/dt \end{bmatrix} = \begin{bmatrix} \gamma \beta_1 z_1 e_1 \\ 2\gamma \beta_2 z_2 e_2 \\ -\gamma \beta_1 u_{c1} e_1 \\ -2\gamma \beta_2 u_{c2} e_2 \end{bmatrix}. \tag{4.8}$$

(9) *Simulation Results and Analysis:*

We assume that the reduced-order linear model in (3.14) can also maintain the similar steady-state outputs as the basic nonlinear model. Now we use this model as an MRAC to take the process plant from these steady-state outputs ($x_D = 0.9654$ and $x_B = 0.0375$) to the desired targets ($0.98 \leq x_D \leq 1$ and $0 \leq x_B \leq 0.02$) amid the disturbances and the plant-model mismatches as the influence of the feed stock disturbances.

The design of a new adaptive controller is shown in Figure 5 where we install an MRAC and a closed-loop PID (Proportional, Integral, Derivative) controller to eliminate the errors between the reference setpoints and the outputs.

We run this controller system with different plant-model mismatches, for instance, a plant with $A = \begin{bmatrix} -0.50 & 0 \\ 0 & -0.75 \end{bmatrix}$, $B = \begin{bmatrix} 1.5 & 0 \\ 0 & 2.5 \end{bmatrix}$ and an adaptive gain $\gamma = 25$. The operating setpoints for the real outputs are $x_{DR} = 0.99$ and $x_{BR} = 0.01$. Then, the reference setpoints for the PID controller are $r_D = 0.0261$ and $r_B = -0.0275$ since the real steady-state outputs are $x_D = 0.9654$ and $x_B = 0.0375$. Simulation in Figure 6 shows that the controlled outputs x_D and x_B are always stable and tracking to the model outputs and the reference setpoints (the dotted lines, r_D and r_B) amid the disturbances and the plant-model mismatches.
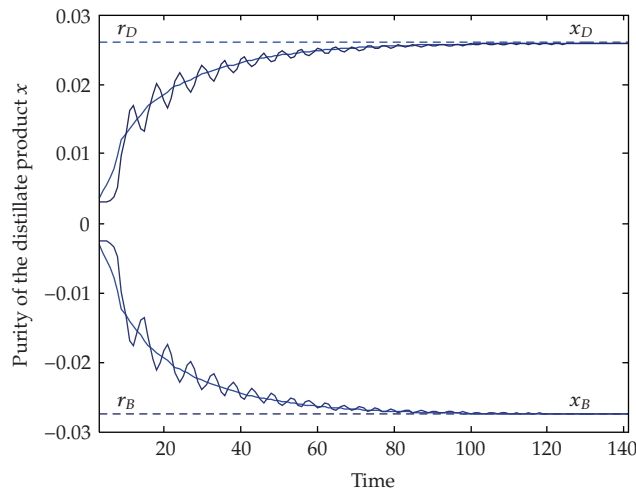
Figure 6: Correlation of plant outputs, model outputs, and reference setpoints.

5. Conclusion

We have introduced a procedure to build up a mathematical model and simulation for a condensate distillation column based on the energy balance (L-V) structure. The mathematical modeling simulation is accomplished over three phases: the basic nonlinear model, the full-order linearized model and the reduced-order linear model. Results from the simulations and analysis are helpful for initial steps of a petroleum project feasibility study and design.

The reduced-order linear model is used as the reference model for an MRAC controller. The controller of MRAC and PID theoretically allows the plant outputs tracking the reference setpoints to achieve the desired product quality amid the disturbances and the model-plant mismatches as the influence of the feed stock disturbances.

In this paper, the calculation of the mathematical model building and the reduced-order linear adaptive controller is only based on the physical laws from the process. The real system identifications including the experimental production factors, specific designed structures, parameters estimation, and the system validation are not mentioned here. Further, the MRAC controller is not suitable for the on-line handling of the process constraints.

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