Modeling and Control of Tray Temperature along with Column Pressure in a Pilot Plant Distillation Column

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***Abstract*— In a distillation column pressure is normally regulated by the vapor-condenser cooling water flow rate. If the pressure is high, increase the water flow to condense vapor faster. The reflux rate, feed rate, change in composition set point, boil-up rate are some of the factors that influences the pressure inside the column. The pressure also affects the speed of vapor through the column and internal reflux. The temperature inside the column and pressure are interdependent, so the interaction between these parameters are high. The objective is to control the bottom tray temperature (T_1) and column pressure, nominally by manipulating the reflux flow and boilup rate.**

***Keywords*— Gain Margin, Phase Margin, Reboiler heater power, reflux flow control.**

I. Introduction

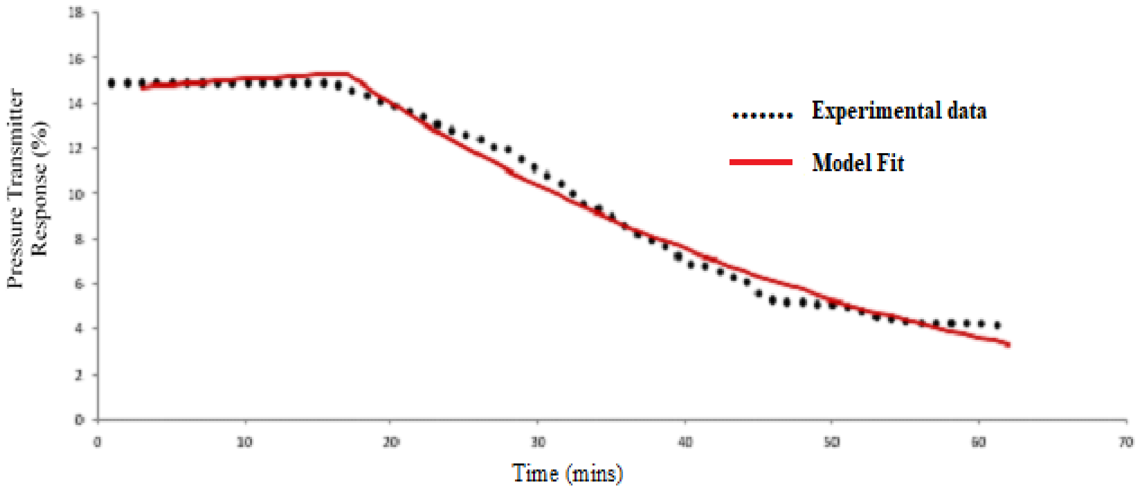
In general many process control industries like petro chemical, oil and gas, Cement plant, the physical process are Multi Input and Multi Output (MIMO). The MIMO process can be controlled by multivariable controllers. In multi-loop control, the MIMO processes are treated as a collection of multi-stage loops. The controller is designed and implemented on each loop by considering loop interactions. Distillation is a process that is used to separate pure liquid from a mixture of liquids. Distillation is commonly used inseparable part of the process industry for the refinement of the products. The column consists of trays, feed section, reboiler, reflux drum, condenser and pressure transmitter. The condenser collects the condensate and stored in the reflux drum. The RTDs are used to record the temperature from the bottom tray continuously with the help of temperature transmitters (4-20mA). The pressure transmitter (0-0.5 bar) is used to record the pressure developed at the bottom of the column. In the present research, Isopropyl alcohol and water mixture in the ratio of 30% and 70%, are considered for the distillation. The current article considers the mathematical model in form of transfer function for simulation. Here the manipulated variable (MV) are, the reflux flow rate (L) and reboiler power rate (Q). The temperature of bottom tray and pressure at the bottom of the column are the controlled variable (PV). Also the article presents the mathematical modeling for the pressure inside the column, simulated control algorithm using MATLAB, later a closed loop response is validated through real-time experimentation

The article is organized as follows: Section 2 is system description, the modeling for the pressure inside the column and Section 3 presents decoupler design and Section 4 gives the expression for Multivariable PI controller. The simulation and implementation results of multivariable PI controller for tray temperature along with column pressure is presented in Section 5.

II. System Descripton

The valid linear model is determined by open loop experimentation, which is achieved by introducing incremental change in the reboiler heater power and reflux flow rate, and then recording the pressure developed in the column near the bottom tray. The reflux flow rate and reboiler power rate is used as manipulated variable. Bottom tray temperature T and Pressure developed P_B are the controllable variables and can be recorded. The open loop response is considered for identifying the model by keeping constant reflux at 20% and giving a step change in the heater from of 50% to 80% and determine the pressure developed in the column.

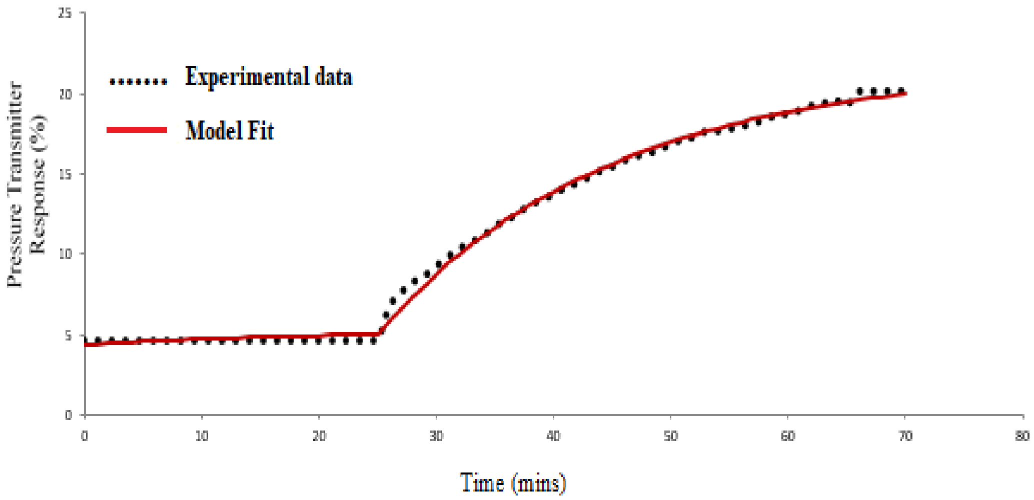
The second step is keeping heater as constant 70% and giving a step change in the reflux from 40% to 70% and the pressure induced inside the column is determined. Fig. (1a) and Fig. (1b) shows the corresponding experimental and model fit response. Modelling in the column is classified as fundamental modelling, empirical modelling and grey box modelling. Most of the process industries rely on empirical modelling, where the model is identified upon the collection analysis of experimental data.



$$FOPTD\ model\ is\ \frac{-0.633e^{-0.0252s}}{0.665s+1}$$

Fig.1. (a) Experimental data and Model fit of Bottom Pressure to a step of 30% in reflux

(Pressure Transmitter is in the range of 0-0.5 bar)



$$FOPTD\ model\ is\ \frac{0.547e^{-0.007s}}{0.328s+1}$$

Fig.1. (b) Experimental data and Model fit of Bottom Pressure to a step of 30% in reboiler heater

This method reduces the root-mean-square deviation between the model and data. In this empirical model input-output relations are built on the data obtained through the open loop test. The response is optimized through the 'Leapfrogging Optimization Technique' which uses all the N data points to fit the model and rejects the noise and disturbances [3]. This method uses nonlinear regression, and provides coefficient of the FOPTD (First Order Plus Dead Time Model) model, based on the best match between the model and the experimental plant output. The FOPTD model obtained is represented as deviation variables as [4].

$$\tau \frac{dy'(t)}{dt} + y'(t) = K_p u'(t-\theta) \tag{1}$$

$$\tilde{y}'(t) = \tilde{y}(t) - \tilde{y}_{base}$$
$$\tilde{u}'(t) = \tilde{u}(t) - \tilde{u}_{base} \tag{1a}$$

t= time after the user defined start point.

The initial condition is

$$y'(t=0) = y" - y_{base} \tag{2}$$

The most commonly used plant transfer matrix model considered here is of the form

$$y(s) = \frac{K_p}{\tau s + 1} e^{-\theta s}\ u(s) \tag{3}$$

where K_p, the process gain is measured in °C/%, τ is the time constant in hours and θ is the delay time measured in hours. The mathematical model of the pilot plant distillation column identified by Vinaya and Arasu for the top and bottom tray temperature is [5][6]

$$G(s) = \begin{bmatrix} \frac{-0.13e^{-0.03s}}{1.14s+1} & \frac{0.18e^{-0.03s}}{0.64s+1} \\ \frac{-0.34e^{-1.22s}}{1.23s+1} & \frac{0.18e^{-0.03s}}{0.32s+1} \end{bmatrix} \quad (4)$$

In the present research the bottom tray temperature and the bottom pressure is considered as the PV. The step change is applied to the heater and reflux, the regression curve is obtained with respect to the output response across the bottom tray (temperature and pressure) with respect to time. The updated process model is given by

$$\begin{bmatrix} y_1(s) \\ y_2(s) \end{bmatrix} = \begin{bmatrix} \frac{-0.34e^{-1.22s}}{1.23s+1} & \frac{0.18e^{-0.03s}}{0.32s+1} \\ \frac{-0.633e^{-0.0252s}}{0.665s+1} & \frac{0.547e^{-0.007s}}{0.328s+1} \end{bmatrix} \quad (5)$$

III. DECOUPLER DESIGN

To eliminate or to minimize the control loop interactions, one of the popular approaches is to design a decoupler. Decoupler decomposes a MIMO process into independent single loop sub-systems [7]. The block diagram of this structure is shown in Fig 2. The generalized MIMO process model with dead time is given by

$$G(s) = \begin{bmatrix} g_{11}(s)e^{-\tau_{11}s} & g_{12}(s)e^{-\tau_{12}s} \\ g_{21}(s)e^{-\tau_{21}s} & g_{22}(s)e^{-\tau_{22}s} \end{bmatrix} \quad (6)$$

The decoupler matrix is given as

$$D(s) = \begin{bmatrix} \vartheta_1(s) & d_{12}(s)\vartheta_2(s) \\ d_{21}(s)\vartheta_1(s) & \vartheta_2(s) \end{bmatrix} \quad (7)$$

with $\vartheta_1(s)$, $\vartheta_2(s)$, $d_{12}(s)$ and $d_{21}(s)$ as given in Eq. (8)

$$\vartheta_1(s) = \begin{cases} 1 & \tau_{21} \geq \tau_{22} \\ e^{(\tau_{21}-\tau_{22})s} & \tau_{21} < \tau_{22} \end{cases}$$

$$\vartheta_2(s) = \begin{cases} 1 & \tau_{12} \geq \tau_{11} \\ e^{(\tau_{12}-\tau_{11})s} & \tau_{12} < \tau_{11} \end{cases}$$

$$d_{12}(s) = -\frac{g_{12}(s)}{g_{11}(s)} e^{-(\tau_{12}-\tau_{11})s}$$

$$d_{12}(s) = -\frac{g_{21}(s)}{g_{22}(s)} e^{-(\tau_{21}-\tau_{22})s} \quad (8)$$

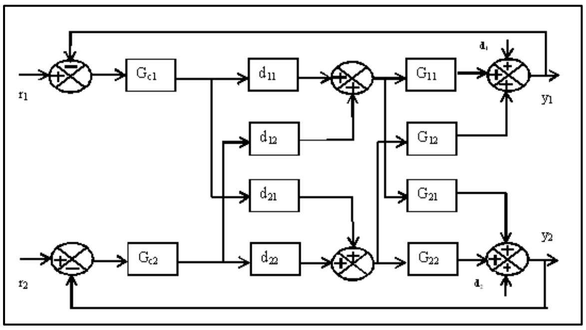


Fig.2. Block Diagram of MIMO System with Decoupler

The diagonal elements of overall open loop transfer function of process with decoupler i.e, $q_{11}(s)$ and $q_{22}(s)$ are approximated into FOPDT model based on frequency response at two points $\omega = 0$ and $\omega = \omega_{pii}$ where ω_{cii} is phase cross over frequency.

$$q_{ii}(s) = \frac{K_{ii}e^{-L_{ii}s}}{T_{ii}s+1} \quad (9)$$

The parameters of the FOPDT model can be calculated using

$$K_{ii} = q_{ii}(0) \quad (10)$$

$$T_{ii} = \sqrt{\frac{K_{ii}^2 - |q_{ii}(j\omega_{pii})|^2}{|q_{ii}(j\omega_{pii})|^2 \omega_{pii}^2}} \quad (11)$$

$$L_{ii} = \frac{\pi + \tan^{-1}(-\omega_{pii}T_{ii})}{\omega_{pii}T_{ii}} \quad (12)$$

IV. CONTROLLER DESIGN METHOD

The Gain margin (GM) and Phase margin (PM) are loop specifications associated with the frequency response. The phase margin is related to damping of the system which serves as important measure for performance. A simple formula is stated to design a multivariable PI controller to meet the user defined gain margin and phase margin. The following set of equations can be used to determine the tuning parameter [1].

$$arg\{q_{ii}(j\omega_{pii})G_{cii}(j\omega_{pii})\} = -\pi \quad (13)$$

$$A_{mii}|q_{ii}(j\omega_{pii})G_{cii}(j\omega_{pii})| = 1 \quad (14)$$

$$\varphi_{mii} = \pi + arg\{(q_{ii}(j\omega_{gii})(G_{cii}(j\omega_{gii})\} \quad (15)$$

Where A_{mii} and φ_{mii} are GM and PM respectively and ω_{gii} and ω_{pii} are gain and phase cross over frequencies. The PI controller is given by

$$G_{cii}(s) = K_{Pii}\left(1 + \frac{1}{sT_{Iii}}\right) \quad (16)$$

Where, K_{Pii} and T_{Iii} are proportional gain and integral time respectively. The PI controller parameters are given as [2]

$$K_{Pii} = \frac{\omega_{pii}T_{ii}}{A_{mii}K_{ii}} \quad (17)$$

$$T_{Iii} = \left(2\omega_{pii} - \frac{4\omega_{pii}^2 L_{ii}}{\pi} + \frac{1}{T_{ii}}\right)^{-1} \quad (18)$$

$$\omega_{pii} = \frac{A_{mii}\varphi_{mii} + \frac{1}{2}\pi(A_{mii}-1)}{(A_{mii}^2 - 1)L_{ii}} \quad (19)$$

V. SIMULATION RESULTS

The controller design method explained under section IV has been carried out for the model given in Equ (5) in section II. Note that the time constants and dead times in the pilot plant distillation column model are measured in terms of hours and the process gain unit is given by ˚C/% [4] [5]. Since for the sampling time of 0.01 sec is used in the DAQ card supplied by Ark Automation, the open loop

data has huge data collection. Hence, the data were converted into hours to get the plant model
The decoupler is

$$D(s) = \begin{bmatrix} 1 & \dfrac{0.65s + 0.529e^{-2.38s}}{0.32s + 1} \\ \dfrac{0.379s + 1.157e^{-0.0182s}}{0.665s + 1} & e^{-1.19s} \end{bmatrix} \quad (20)$$

The diagonal elements of open loop transfer function in FOPDT model is

$$q_{11} = \frac{-0.13e^{-2.44s}}{s + 1}$$
$$q_{22} = \frac{0.212e^{-0.835s}}{2.232s + 1} \quad (21)$$

Assume the gain margin to be 6 and phase margin to be 60 deg. Then resulting PI controller is

$$K_P = \begin{bmatrix} -0.1194 & 0 \\ 0 & 0.4846 \end{bmatrix} \qquad K_i = \begin{bmatrix} -0.1436 & 0 \\ 0 & 0.5041 \end{bmatrix}$$

The response shows that the controller provides good tracking with the given setpoint. Fig. 4(a) and 4(b) shows the closed loop response of Gain Margin-Phase Margin based PI controller with decoupler for a pilot plant binary distillation column.

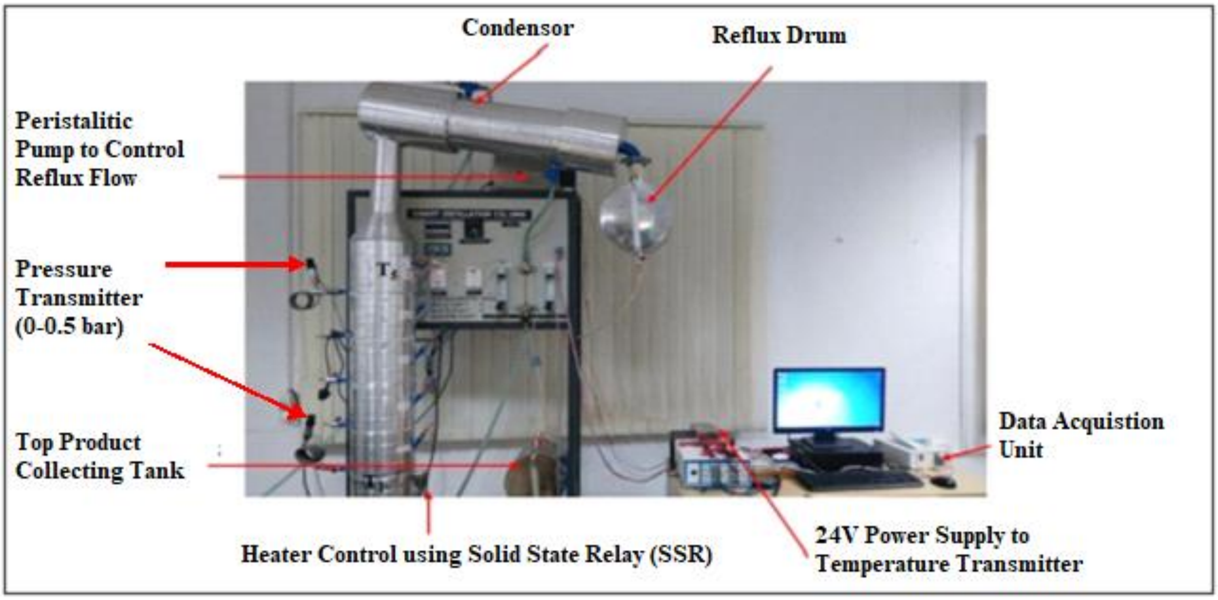
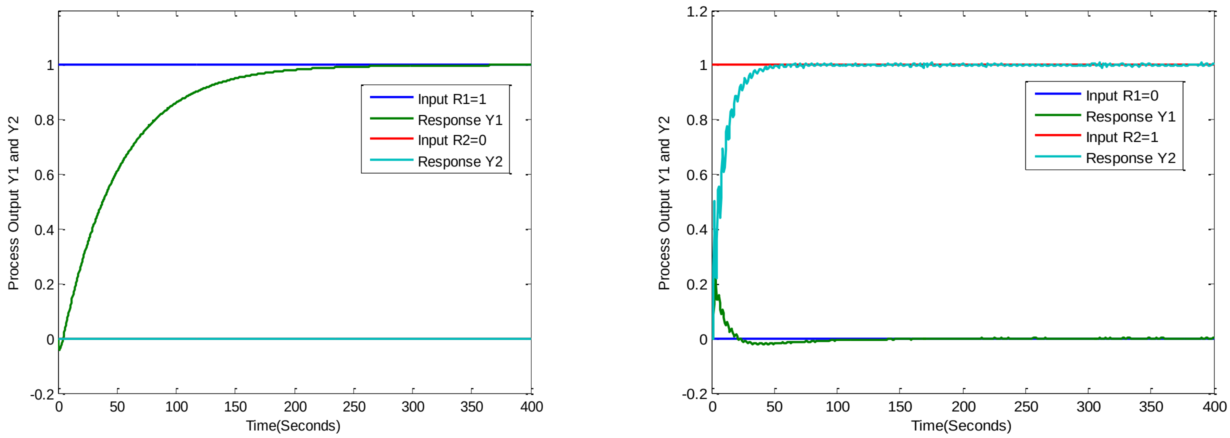


Fig.3 Pilot Plant Distillation Column Setup



(a) (b)

Fig.4 (a) Servo response of GMPM based controller when input R1=1 and input R2=0,
(b) Servo response of GMPM based controller when input R1=0 and input R2=1

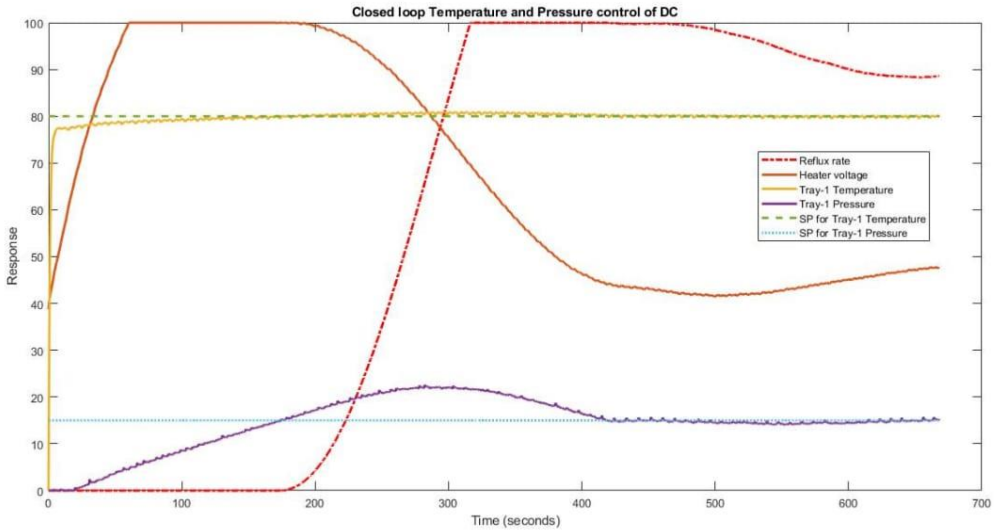


Fig. 5. Implementation of PI controller for Temperature and Pressure Control in pilot plant binary distillation column

VI. CONCLUSION

The GM-PM based PI controller is designed for the pilot plant binary distillation column model shown in Fig. 3. The simulation studies for the obtained controller shows good tracking of the servo response for the given step input signal. The real- time implementation of the control algorithm for bottom temperature and pressure control is presented in Fig. 5.

FUTURE WORK

The pressure transmitter needs to be placed near the top tray (T_5) of the column, and its model can be identified with step change in Boil-up rate and Reflux flow rate. Also, the cold water supply of condenser can be made as the third MV in the process.

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