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An Experimental Study on Implementation of Centralized PI Control Techniques on Pilot Plant Binary Distillation Column

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Abstract : The article discusses centralized PI control techniques and its implementation on a pilot plant binary distillation column. More specifically, the paper presents two centralized control methods. The first method is designed based on the Steady State Gain Matrix (SSGM), whereas the second method is designed based on the SSGM along with time delay and time constant of the process transfer function matrix. The interaction and main effects are obtained and analyzed between the process variables along with its performance. Further, the selection of process variables during the closed loop implementation of the control algorithm are studied. Moreover, the effectiveness of the centralized control techniques are discussed both, in simulation and real- time environment. Finally, the control algorithm is implemented in real-time using the MATLAB Simulink platform.
Keywords: Centralized Controller, Pilot Plant, Steady State Gain Matrix, Process Parameter.

I. Introduction

The most commonly used separation technique in the process industry consumes very large amounts of energy[1]. Thus, the need of the hour is to reduce the consumption of energy as much as possible. A basic understanding of the dynamic and control of the distillation column is essential for a process control engineer[2]. The most efficient way of reducing energy consumption is by identifying an accurate model[3]. This will help the process engineer to predict and control the distillation process. There are different modeling techniques in academic literature, such as empirical modeling, fundamental modeling, hybrid modelling, etc. With empirical modeling, the relationship between the input- output is derived from the input- output data of the plant operation[4, 5]. With this method, computation is reduced and there is no necessity to understand the dynamics of the plant[6]. Thus, the empirical modeling technique is used in the present work to identify the pilot plant distillation column. Model identification is a challenging task, since it has: a) a large time constant b) variations due to ambient temperature, c) nonlinearities, d) ill-conditioning of the column, etc[7]. In this article, centralized PI tuning techniques are presented for the identified distillation column model.

The control of a MIMO system is difficult compared to a SISO system. This is due to the cross-coupling and loop interactions between the several variables[8]. The presence of multiple time delays and inverse responses are also important factors to control in a multivariable process. These factors impose limitations and control over the performance. Several control techniques have been reported in academic literature for effective control of the MIMO system. Among these, the centralized PI control techniques are the most promising in the process industry[1, 9]. In the centralized control technique, every input has a significant effect on all controller outputs[10]. Further, it gives a satisfactory result for a strong interacting system in comparison to the decentralized system. Thus, in a centralized approach, the effect of interaction on the process output is effectively suppressed by designing an off-diagonal controller[11]. Different approaches of designing a controller based on the centralized approach have been discussed in recent publication[12, 13]. Two different approaches have been used in academic literature to design the centralized control technique. In the first approach, the decouple is combined with the decentralized controller, whereas the second one is a pure centralized controller[1]. The present article presents pure centralized PI controller techniques. The PI coefficients are designed based on the process parameters and SSGM of the plant transfer function model.

In a brief summary of the article, section 2 gives a concise theory on the experimental setup of the pilot plant distillation column along with the mathematical model of the distillation column. Section 3 discusses the centralized control techniques. Section 4 presents a detailed discussion on the simulation and experimental results, followed by conclusions.

II. Experimental Setup

Distillation is an important commercial process, widely used in petroleum, chemical, and food industries for either separation or purification of two or more components[14]. The separation process is based on the boiling points, and helps to separate the volatile and non-volatile components from the mixture of two or more component[15]. The current research work uses a mixture of isopropyl alcohol and water, which are fed in to the pilot plant binary distillation column. The development of the distillation column model is essential to predict its dynamics and thus, to control. Fig (1) shows the laboratory scale setup of the binary distillation column used in the present research work.

(i) Description of the Plant

The pilot plant distillation column is equipped with 8 bubble cap trays. The column is 15.44cm in diameter and including the reboiler section, column is 172cm in height. A two 3kW heater connected at the bottom of the column supplies heat for the necessary vapourization. The condenser is connected at the top of the column for condensation and to cool the distillate products. The L-V configuration is used here for model identification. The reflux flow rate and reboiler power are the manipulated variables. The controlled variables are the temperatures of tray 6 (T_6) and tray 3 (T_3). The DAQ card is used to monitor and control the manipulated variables.

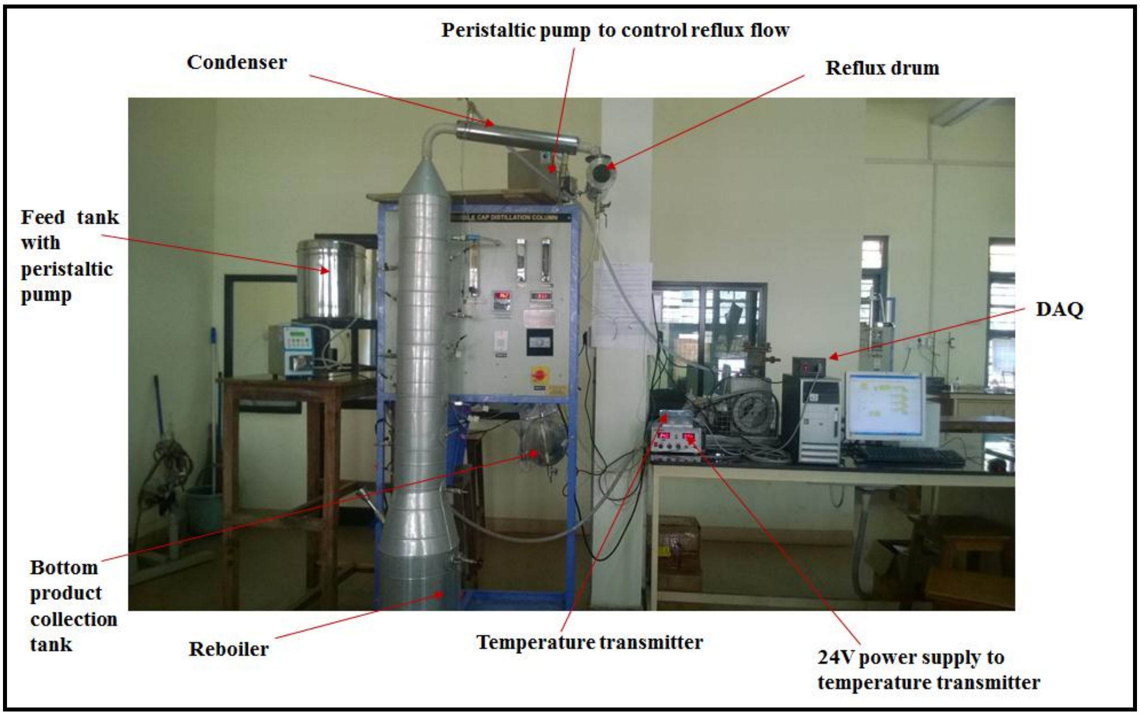


Fig. 1. Laboratory setup of binary distillation column

(ii) Model Identification

The open loop test is performed on the distillation column for model identification[18]. The step change is applied to the manipulated variables, reflux, and the reboiler. Later responses of T_6 and T_3 are recorded. Then, the empirical FOPTD model is identified[19]. The process model can be written as shown in Eq. (1).

$$y(s) = G(s)u(s) \tag{1}$$

i.e.,

$$\begin{bmatrix} y_1(s) \\ y_2(s) \end{bmatrix} = \begin{bmatrix} G_{p11}(s) & G_{p12}(s) \\ G_{p21}(s) & G_{p22}(s) \end{bmatrix} \begin{bmatrix} u_1(s) \\ u_2(s) \end{bmatrix}$$

Where,

$$\begin{bmatrix} y_1(s) \\ y_2(s) \end{bmatrix} = \begin{bmatrix} T_6 \\ T_3 \end{bmatrix}, \quad \begin{bmatrix} u_1(s) \\ u_2(s) \end{bmatrix} = \begin{bmatrix} L \\ Q \end{bmatrix}$$

$$G_p(s) = \begin{bmatrix} \dfrac{-0.16e^{-0.01s}}{0.01s+1} & \dfrac{0.6e^{-1.19s}}{0.05s+1} \\ \dfrac{-0.04e^{-0.01s}}{0.02s+1} & \dfrac{0.49e^{-0.47s}}{0.19s+1} \end{bmatrix} \tag{2}$$

Here, k_p is the process gain measured in $^{o}C/\%$, τ is the time constant, and θ is the time delay, of which the last two are measured in hours.

The control problem of the pilot plant distillation column is characterized by identifying the two important types of process variables:

- Controlled Variables (T_6 and T_3): These variable are to be controlled. The set-point is the desired value of the controlled variable.
- Manipulated Variables (L and Q): These variables are to be adjusted so as to keep the controlled variable close to the set-point.

The specifications of the process variables play a major role in developing the control system in real-time environment. It is a very tedious process because temperature is non-linear in nature. The selection of these is based on experience, process operation, and control objectives.

III. Centralized Pi Techniques

For the multivariable time delay process, $G_p(s)$ is the transfer function matrix of the process with n inputs and n outputs, and is given by Eq. (3):

$$G_p(s) = \begin{bmatrix} g_{11} & g_{12} & \cdots & g_{1n} \\ g_{21} & g_{21} & \cdots & g_{2n} \\ \vdots & \vdots & \ddots & \vdots \\ g_{n1} & g_{n2} & \cdots & g_{nn} \end{bmatrix}_{n\times n} \tag{3}$$

Where, the transfer function is in the form $g_{ij} = g_o e^{-\theta_{p,ij}s}$; here, $g_{oij} = k_{p,ij}/(\tau_{p,ij}s+1)$, is strictly proper and stable transfer function and $k_{p,ij}$, $\tau_{p,ij}$ and $\theta_{p,ij}$ are the corresponding process gain, time constant, and time delay in the process transfer function. $G_c(s)$ is the transfer matrix of the controller with n outputs and n inputs as presented in Eq. (4)

$$G_c(s)=\begin{bmatrix} g_{c11} & g_{c12} & \cdots & g_{c1n} \\ g_{c21} & g_{c21} & \cdots & g_{c2n} \\ \vdots & \vdots & \ddots & \vdots \\ g_{n1} & g_{n2} & \cdots & g_{nn} \end{bmatrix}_{n\times n} \tag{4}$$

Where, $g_{c,ij}=k_{c,ij}+\frac{k_{I,ij}}{s}$ and $k_{c,ij}$ and $k_{I,ij}$ are the corresponding proportional gain and integral gain of the controller transfer matrix.

Method-I

In this approach, the empirical tuning method is used to control the multivariable process, as given by Davison[20], where, k_c and k_I is given by Eqs. (5) and (6).

$$k_c=\delta_1\left[G_p(s=0)\right]^{-1} \tag{5}$$

$$k_I=\delta_2\left[G_p(s=0)\right]^{-1} \tag{6}$$

Here, $[G_p(s=0)]^{-1}$ is called the inverse of the steady state gain matrix $G_p(s)$. The recommended range of the tuning parameters δ_1 and δ_2 is between 0 and 1.

Method-II

In this method, $G_c(s)$ is calculated using the inverse steady state gain matrix and the process parameters as discussed by Vinayambika *et al.*[21]. Here, the time delay and the time constant of the process transfer matrix is considered for the PI controller setting. ε_1 and ε_2 are the tuning parameters with respect to Proportional gain (k_c) and Integral time control (τ_I) setting, respectively[16]. The recommended range of the tuning parameters ε_1 and ε_2 is between 0 to 1 and 1.5 to 3.5, respectively.

IV. Result and Discussion

The identified model of the pilot plant distillation column is validated through simulation, and real-time environment. The Steady State Gain matrix of Eq. (2) is given by,

$$G_p(0)=\begin{bmatrix} -0.16 & 0.6 \\ -0.04 & 0.49 \end{bmatrix}$$

The inverse of the Steady State Gain Matrix is given by,

$$\left[G_p(0)\right]^{-1}=\begin{bmatrix} -8.9286 & 10.7143 \\ -0.7143 & 2.8571 \end{bmatrix}$$

The controller designed based on the Davison method is given by,

$$G_c(s)=\begin{bmatrix} -0.8929-\dfrac{0.4464}{s} & 1.07143+\dfrac{0.5357}{s} \\ -0.7143-\dfrac{0.0357}{s} & 0.2857+\dfrac{0.1428}{s} \end{bmatrix}$$

The controller designed based on Vinayambika method is given by,

$$G_c(s) = \begin{bmatrix} -0.04 - \dfrac{0.11}{s} & 0.048 + \dfrac{0.134}{s} \\ -0.0032 - \dfrac{0.0089}{s} & 0.013 + \dfrac{0.036}{s} \end{bmatrix}$$

The block diagram of the centralized PI controller implementation on a binary distillation column is shown in Fig (2) Even with a well-tuned controller, the actuator saturation was a major issue in the performance, during the real- time implementation of the control technique[17]. The selection of the saturation is based on the saturation limit of the actuators. In the current research work, the saturation value was set within the actual limit for safe plant operation. Moreover, output of the feedback lags, when the limitation in the saturation blocks are achieved. Thus, the controller action is fed to the plant based on its saturation limit. Accordingly, the controller will act on the plant to achieve a satisfactory closed loop stable response.

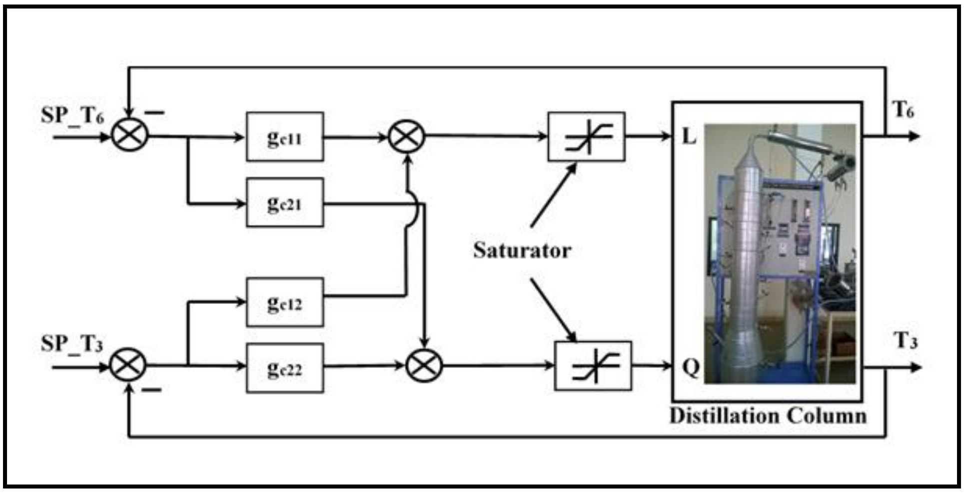
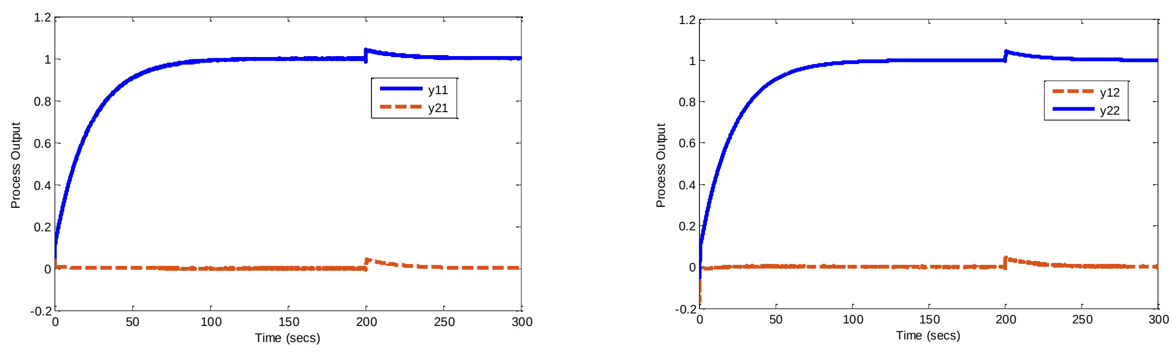


Fig. 2. Block diagram for the centralized controller implementation on a pilot plant binary distillation column

The servo and regulatory responses of the centralized control technique is shown in Fig (3) and Fig (4), respectively. The closed loop response is stable without any overshoot and oscillation. Further, the responses also indicate that the interaction effect is significantly reduced. Both the control techniques are implemented in real- time. Fig (5) and Fig (6) represent the implementation of the centralized control techniques on the pilot plant distillation column. In simulation, 'Method-I' settles fast than Method-II as shown in simulation results Figs (3) and (4). But in the real- time, 'Method II' gives a significantly improved response than Method-I.



(a) (b)

Fig. 3. Servo and regulatory responses of the centralized controller of Method I

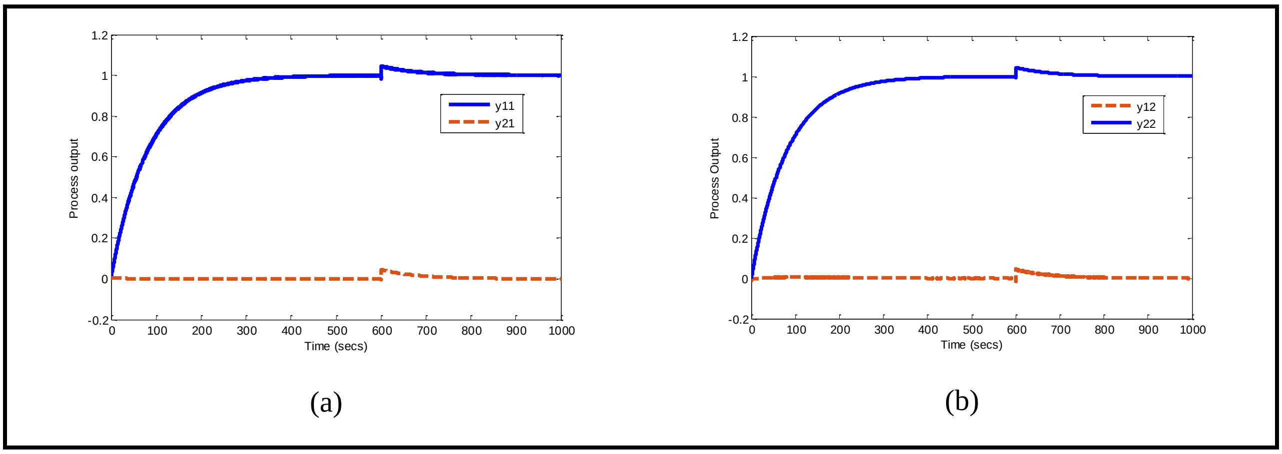


Fig. 4. Servo and regulatory responses of the centralized controller of Method II

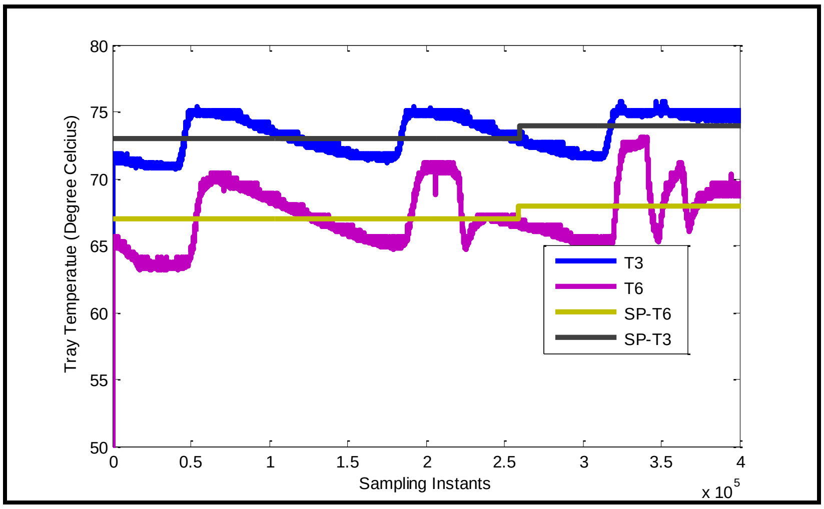


Fig. 5. Implementation of the centralized controller (Method I) on the pilot plant distillation column

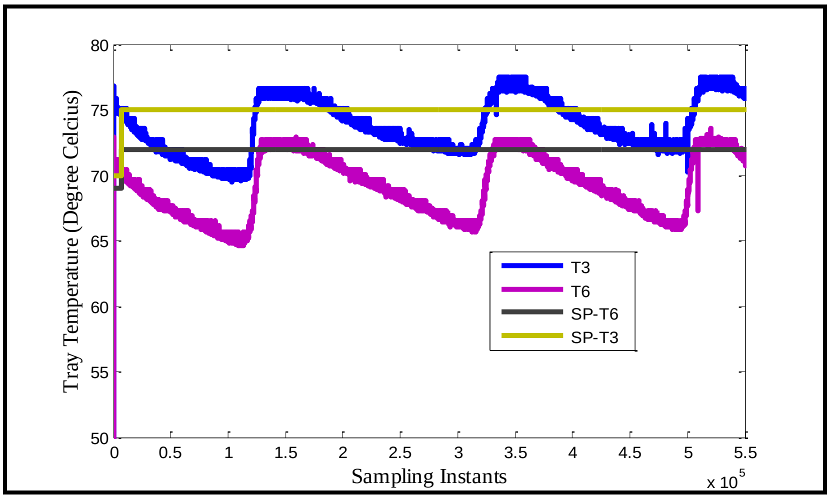


Fig. 6. Implementation of the centralized controller (Method II) on the pilot plant distillation column

V. Conclusion

Two centralized PI control techniques are implemented in MATLAB simulation and experimental real-time process. Both the controllers are tested with the set-point change and disturbance rejection. The results shows that reduced interaction effect and improved performance are achieved with the centralized control technique. The article presents the MATLAB Simulink based on real- time implementation of the control algorithm. The experimentation is performed on a binary distillation column to validate the applicability and effectiveness of the centralized control technique. The selection of the controlled and manipulated variables during the implementation of the control algorithm plays a major role in obtaining a stable response.

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Nomenclatures and Abbreviations

T_3	:	Temperature of Tray-3, near the bottom of the column in Degree Celsius
T_6	:	Temperature of Tray-6, near the top of the column in Degree Celsius
u_1, u_2	:	Manipulated variables
y_1, y_2	:	Controlled variables
L	:	Reflux flow rate in rpm
k_p	:	Process gain $^oC/\%$
τ	:	Time constant in hours
θ	:	Time delay in hours
k_c	:	Proportional gain
τ_I	:	Integral time
k_I	:	Integral gain
δ_1, δ_2, ε_1, ε_2	:	Tuning parameters
$G_p(s)$	:	Process transfer function matrix
$G_c(s)$	:	Controller transfer function matrix
DAQ	:	Data Acquisition
L-V	:	Liquid Vapour
PI	:	Proportional-Integral
FOPTD	:	First Order Plus Time Delay

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