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DESIGN OF CENTRALISED PI CONTROLLER FOR MULTIVARIABLE SYSTEM

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ABSTRACT

In this paper various methods for the designing of centralized PI controller for desalination system and distillation columns given in the literatures are discussed. Various tuning methods given by Davison and Maciejowski are used. The performance of the closed loop system servo responses are compared of terms of ISE. The better controller tuning method is chosen by means of least ISE value.

INTRODUCTION

Most process industries are dealing with multivariable process. Multivariable processes usually pose difficulties to design controllers, due to the interactions that occur between the input/output variables. Their control characteristics and tuning requirements change as operating conditions change. In addition, many operations experience some type of control loop interaction. When the process variable of one loop severely interacts with or disturbs the process variables of other loops, centralized control can eliminate the interaction. Two different processes are considered. One is desalination of sea water is by Multi Stage Flash Desalination [11]. Another process is distillation column which separates methanol and water [12].

In this centralized controller, if the transfer function of the process is n x n matrix, then it requires n x n PID controllers. This method reduces the interaction of the process [1-10]. However, the design methods of such Centralized controllers require first pairing of input-output variables, and tuning of controllers requires trial and error steps. Only well experienced operators can tune such control loops. For strongly interacting systems, decentralized controllers will not give satisfactory responses. There are some simpler methods of tuning centralized controllers. The centralized control system requires nxn controllers for controlling n output variables using n manipulated variables, which is an advantage if we use standard PI controllers.

The tuning methods are evaluated based on the simple performance criteria which include the rise time, peak time, settling time, maximum peak overshoot and controller performance metrics which are the Integral of absolute Error (IAE) and the Integral square error (ISE), Integral of Time multiply by Absolute Error (ITAE).

DESIGN OF CONTROLLER

The transfer function matrix of the system with s-domain is denoted by Gp(s) given by

Y(s) = Gp(s). U(s) (1)

Where Y is the vector of output variable and U is the vector of manipulated variables. The centralized controller are assumed to be in the form

Gc(s) = Kc + (KI/s) (2)

Where Kc and KI are matrices of size nxn

$Kc = [Kc, ij]$,

and

$K_I = \left[\frac{Kc, ij}{\tau_I, ij}\right]$, i=1 to n and j=1 to n. (3)

2.1 Controller design for Square Process

2.1.1 Davison Method

Davison has proposed a multivariable PI controller where the matrices K_c and K_I are given by

$K_c = \delta\ [G\ (s=0)]^{-1}$ (4)

$K_I = \varepsilon\ [G\ (s=0)]^{-1}$ (5)

Here $[G(s=0)]^{-1}$ is inverse of steady state gain matrix and δ and ε are the fine tuning parameters. The fine tuning parameters range is from 0 to 1. The recommended values are 0.1-0.3.

2.1.2 Maciejowski Method

In this method the matrix K_c is given by

$K_c = \delta\ [G\ (s = i\omega_b)]^{-1}$ (6)

Should be where ω_b the desired bandwidth of the system is. K_I is the same as given in the Davison method. The matrix will be a complex matrix, and a real approximation of the matrix is required for the controller design and implementation.

2.2 Controller design for Non-Square Process

2.2.1 Davison Method

In the present work, this method is extended to non-square system. As inverse does not exist for non-square system, Moore–Penrose pseudo inverse is used. For matrix A, Moore–Penrose pseudo-inverse is

$$A^{-1} = A^T (A * A^T)^{-1} \qquad (7)$$

A^T is the transports matrix of A. So for non-square system, PID controller gains are:

$$K_c = \delta[G(S=0)]^{-1} \; K_I = \varepsilon[G(S=0)]^{-1} \qquad (8)$$

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For matrix Gp, Moore–Penrose pseudo-inverse is

$$G_P{}^{-1} = G_P{}^T (G_P * G_P{}^T)^{-1} \qquad (9)$$

In this method the matrix K_c and K_I are given by

$$K_c = \delta\, mag[G(s = i\omega_b)]^{-1} \qquad (10)$$

$$K_I = \varepsilon[G(s = i\omega_b)]^{-1} \qquad (11)$$

PROCESS DESCRIPTION

3.1 Multi-stage Flash Distillation (MSF) Process

Transfer Function of MSF Desalination Process is given by

$$G_p(s) = \begin{bmatrix} \dfrac{-0.413}{4.34s+1} & \dfrac{20.64e^{-s}}{5.98s+1} \\ \dfrac{-0.09e^{-s}}{3.9s+1} & \dfrac{7.9}{6.4s+1} \end{bmatrix} \qquad (12)$$

It is a water desalination process that distills sea water by flashing a portion of the water into steam in multiple stages using counter current heat exchangers. The controlled variables are the top brine temperature and distillation production rate. The manipulated variables are the recycle flow rate and steam temperature.

3.2 Wood and Barry distillation column

Transfer Function of Wood and Barry distillation column is given by

$$G_p(s) = \begin{bmatrix} \dfrac{12.8e^{-s}}{16.7s+1} & \dfrac{-18.9e^{-3s}}{5.98s+1} \\ \dfrac{6.6e^{-7s}}{10.9s+1} & \dfrac{-19.4e^{-3s}}{14.4s+1} \end{bmatrix} \qquad (13)$$

The process separates methanol and water. The controlled variables are distillate methanol (top) and bottom methanol and the manipulated variables are reflux flow rate and steam flow rate.

3.3 Two Coupled Distillation Process

Transfer Function of Two Coupled Distillation Process is given by

$$G(s) = \begin{bmatrix} \dfrac{0.052e^{-8s}}{19.8s+1} & \dfrac{-0.03(1-15.8s}{108s^2+63s+1} & \dfrac{0.012(1-47s)}{181s^2+29s+1} \\ \dfrac{0.0725}{890s^2+64s+1} & \dfrac{-0.0029(1-560s)}{293s^2+51s+1} & \dfrac{0.0078}{423s+1} \end{bmatrix}$$

The controlled variables are mole fraction of ethanol in distillate and mole fraction of water in bottoms and manipulated variables are distillate flow rate, steam flow rate and product fraction from the side column.

RESULT AND DISCUSSION

4.1 MSF PROCESS

Fig 1 Servo response of desalination process (Y1)

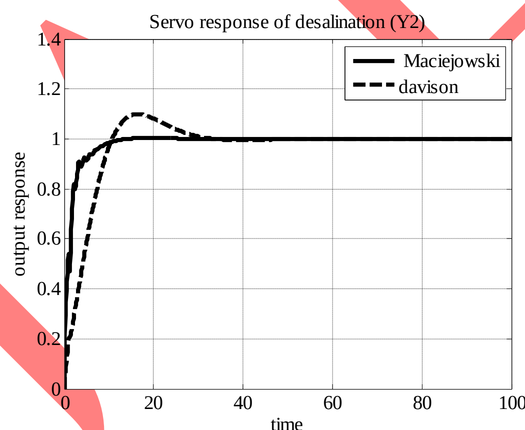

Fig 2 Servo response of desalination process (Y2)

Table 1 ISE values for centralized controllers for a step change in set point

Methods	δ	ε	***ISE in y1***	***ISE in y2***
Davison	**0.8**	**0.3**	**7.897**	**2.767**
Maciejowski	**3.8**	**0.6**	**3.147**	**0.6926**

From Figure1 and 2 and the Table 1 the response of desalination process with Maciejowski method produces better results in terms of less ISE values.

4.2 **WOOD AND BARRY DISTILLATION COLUMN**

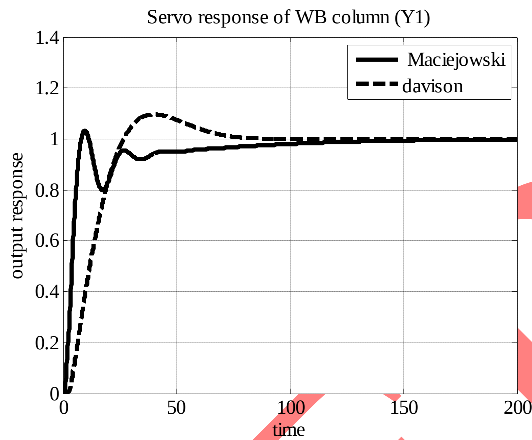

Fig 3 Servo response of wood and Barry distillation column (Y1)

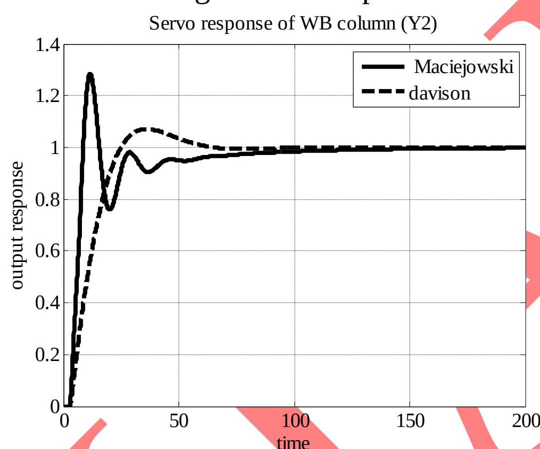

Fig 4 Servo response of wood and Barry distillation column (Y2)

Table 2 ISE values for centralized controllers for a step change in set point

Methods	δ	ε	***ISE in y1***	***ISE in y2***
Davison	**1.0**	**0.1**	**9.312**	**8.134**
Maciejowski	**3.0**	**0.1**	**3.59**	**5.746**

From Figure3 and 4 and the Table 2 the response of wood and Barry distillation column with Maciejowski method produces better results in terms of less ISE values.

4.3 TWO COUPLED DISTILLATION PROCESS

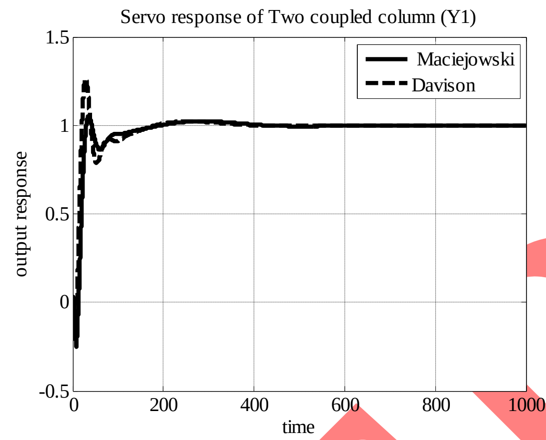

Fig 5 Servo response Two Coupled Distillation Process (Y1)

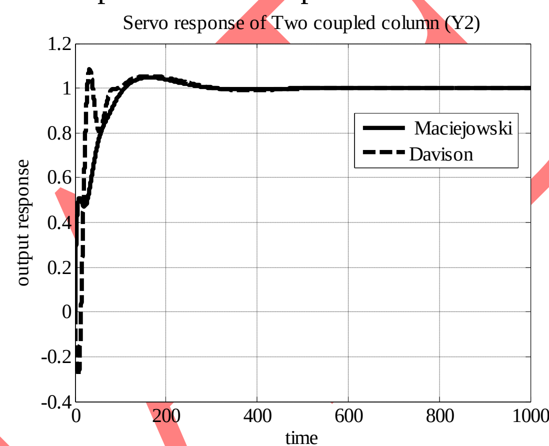

Fig 6 Servo response Two Coupled Distillation Process (Y2)

Table 3 ISE values for centralized controllers for a step change in set point

Methods	δ	ε	***ISE in y1***	***ISE in y2***
Davison	1.8	0.05	17.09	23.95
Maciejowski	0.88	0.04	20.9	12.98

From Figure 5 and 6 and the Table 3 the response of Two Coupled Distillation Process with Maciejowski method produces better results in terms of less ISE values.

CONCLUSION

Multivariable process has multiple inputs and multiple outputs. Centralised controllers are designed and implemented. Two different controllers' tunings are used for three different type processes. The results are analysed. The Maciejowski method produces better result in terms ISE compared with Davison method.

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