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System Identification and Conditional Control for an Optimal Operation of a Pilot Plant Binary Distillation Column

Eadala Sarath Yadav[1], Thirunavukkarasu Indiran [1*] and S Shanmuga Priya[2]

[1] Department of Instrumentation and Control Engineering, [2]Department of Chemical Engineering, Manipal Institute of Technology, Manipal Academy of Higher Education, Manipal-576104, Karnataka, India

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Abstract: In this paper a new control structure has been proposed that minimizes the actuator energy by optimally reducing the error at steady state of closed loop response. A dynamic switching phenomena is employed such that effective actuator input manipulates the process variable. Parameter identification using PRBS input signal is carried out for MISO and MIMO processes. Pilot scale binary distillation column is considered for the study with tray temperature as variable of interest, heater voltage and reflux flow rate as its manipulated variables. The usage of proposed control structure with existing controller improves the system performance and reduces the operational cost. To depict the efficacy of the control structure, system is subjected to uncertainty by perturbing plant parameters of 30% from their actual values. Result shows significant improvement in actuator energy consumption. Control methodology has been experimentally validated on both MISO and MIMO schemes.

Keywords: Conditional Control, MIMO Process, MISO Process, PRBS input, Actuator Energy Consumption.

1. Introduction

Distillation column is one of the key application in many process industries. The principle of distillation is to separate the liquid mixture into two or more components through the phenomena of evaporation and condensation by means of relative volatility difference [1]. The effective separation of components takes place when appropriate temperature and pressure is maintained in the column. Control design plays a vital role in maintaining those variables at a desired position. Mathematical model estimation is one task needs to be carried for precise control of process variables. Input signal design is a fundamental objective in data driven based parameter estimation. Input is set to excite in such a way that the effect of output should be larger than those responses caused by sensor noise. Input signal generation should contain amplitude, rate of input change (Frequency), bias and variance. One such signal is Pseudo Random Binary Sequence (PRBS) input signal. Temperature control of distillation column is the major application in many of the process industries. This control problem has been stratified in different ways by researchers with respect to the control objective. Wood and Berry [2] conducted experiments on binary distillation column for composition control while feed flow rate is being disturbed. William L. Luyben [3] had presented an analytical solution to obtain the system model parameters for highly nonlinear distillation columns by using K.J.Astrom and T.Hagglund [4] autotuning method using feedback relay. Skogestad [5] presented a simple PID formula for different types of model structure. He also gave a brief study on Pros and cons of distillation column through different literatures and also recommended the best possible control configurations of distillation column which helps in modelling and control of it [6].

This paper is arranged with system identification of pilot scale distillation column using PRBS and step input excitation for MISO and MIMO processes respectively in section.2. Conditional control structure with anti-reset windup in section.3. Implementation of control scheme on MISO and MIMO structure are presented in section.4 and followed by conclusion in the Section 5.

2. System Identification

A. Distillation process description

The process of distillation is referred as fraction of distillation or fractionation and distillation tower sometimes called fractionators. Fractionation is the distillation that occurs at different levels of the tower. In general if the feed position is at middle of the column, the upper section is called rectifying section and lower section is called stripping section. Figure.1 shows the schematic representation of distillation column.

E-mail: sarath.eadala@gmail.com, it.arasu@manipal.edu and shan.priya@manipal.edu*

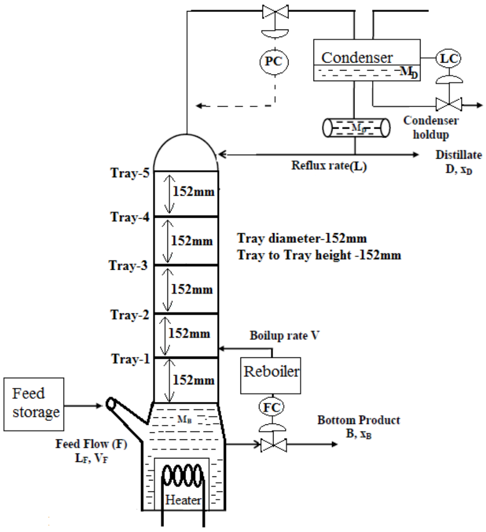

Figure 1. Schematic diagram of distillation column

The condensate will contain more of the volatile components which is accumulated at top of the column and the mixture at bottom end of the column contains less of the volatile components. The tower consist of different trays designed as bubble cap trays which separates vapors and liquid [7]. There are several methods used to maximize the product purity in the distillation column. One among them are refluxing, by sending distillate back to the column which is often referred to as external reflux. Another method used to maximize the product purity is re-boiling. As the bottom liquid from the tower is send to a heater, the re-boiler heats the liquid it receives so that the mixture of vapor and liquid is formed, depending on the system either vapors or the mixture of the vapor and liquid is the reintroduced into the tower via external reflux. It is necessary to understand and model the plant dynamics to control the reflux flow rate and re-boiling rate by which distillate purity is reliant on.

B. Model Parameters Estimation

Parameter estimation of multivariable processes requires thorough knowledge on the dynamics of system, variables of interest and their behavior. However data driven based modeling approach doesn't require prior knowledge of the plant dynamics but Input-Output (IO) data. Random binary signal is generated by passing a random Gaussian signal through the sign function. The input signal should contain the amplitude and rate of change in input (frequency) [8]. As the input excitation is independent and within the operator's boundary to vary, it is important to know what kind of excitation is suitable for particular system. In this paper PRBS input based mathematical modeling has been carried out for MISO process and step input based mathematical modeling for MIMO configuration. The experimental setup to perform modeling is shown in the Figure.2.

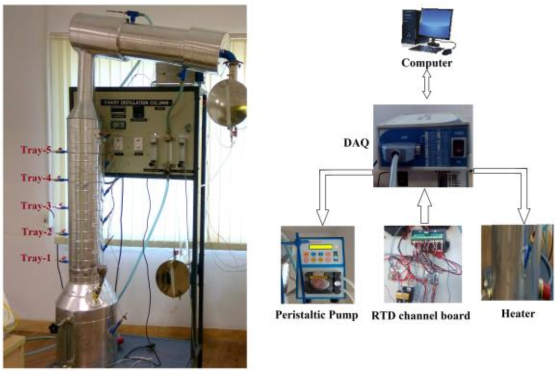

Figure 2. Experimental setup of pilot plant binary distillation column

C. Modeling of Multi Input Single Output (MISO) Process

The amplitude and frequency of PRBS input signal is preferably selected based on the following steps [10].

1. Amplitude of PRBS can be selected based on the operating region of the process.
2. To calculate the frequency, step test needs to be performed around the preferred region of operation with (5-10) % of change in input.
3. The corresponding step response is used to fetch the information of system gain, time constant and dead time around that operating region.
4. By using time constant of the system response, bandwidth (Ω) can be estimated.

$$\Omega = \frac{1}{\tau}$$

5. Sampling frequency F_s which can be selected anywhere in between (10-20) % of Ω

$$F_s = \frac{1}{T_s}$$

Where T_s is sampling time

Simulink design of PRBS signal generator is given in Figure.3. The amplitude of PRBS determines the operating region of input signal. In this paper, the amplitude variation is considered as ±10 from mean value of input. For example, if the heater is manipulated with mean value as 50, the PRBS signal will vary with an amplitude as [40, 60] for ±10 variation.

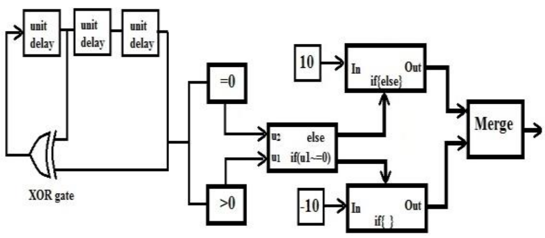

Figure 3. Simulink design of PRBS signal generator with gain of ±10

The variable of interest for MISO case is considered as follows

Process Variables:

Y: tray temperature at bottom tray-3

Manipulated Variables:

U_1: Reflux flow rate. (L)

U_2: Re-boiler heater voltage. (Q)

Mathematical model is assumed as first order plus dead time (FOPDT) structure as shown in the equation.1. Considering the dynamics of temperature of tray-3 for MISO process, manipulating variable U_1 tries to reduce the temperature and manipulating variable U_2 tries to increase the temperature. U_2 is kept constant when U_1 is excited, and vice versa.

$$G(s) = \frac{K}{\tau s + 1} e^{-\theta s} \tag{1}$$

Parameters estimation as FOPDT structure is obtained by using input-output data using above variables configuration. The input-output data from the experimental setup is fetched into system identification tool box (ident) in MATLAB. The process model is selected to obtain FOPDT model. The methodology used in backend of the tool box is regression approach [8].

$$\frac{y}{U_1} = G_1(s) = \frac{-2.24}{14.56s + 1} e^{-0.01s} \tag{2}$$

$$\frac{y}{U_2} = G_2(s) = \frac{2.39}{5.48s + 1} e^{-0.01s} \tag{3}$$

Note: The time constant and dead time of equation.2 and equation.3 are in terms of minutes. While considering those model parameters for control design, they were considered in terms of hours. The control loop structure for non-square process variables has been extended from Kalpana and Thyagarajan [9]

D. Modeling of Multi Input Multi Output (MIMO) Process

The variable of interest for MIMO case is considered as follows:

Process Variables:

Y_1: tray temperature at bottom tray-5

Y_2: tray temperature at bottom tray-1

Manipulated Variables:

U_1: Reflux flow rate. (L)

U_2: Re-boiler heater voltage. (Q)

The challenging issue of MIMO configuration is handing the input-output (IO) interactions. Therefore while modeling system for MIMO configuration, it is important to note that the process output must reach steady state with change in input. It is observed that steady state time of the process response based on frequency selected for PRBS through the methodology [10] is not sufficient for this MIMO application. Whereas same methodology has been validated using interacting level control process [11]. The reason of being batch process, there is an insufficient reflux to maintain steady state on tray-1. From Figure.4 it is observed that the PRBS design of reflux flow rate is not sufficient enough to manipulate tray-1 to reach steady state. Therefore step input is used as an excitation and process output is recorded.

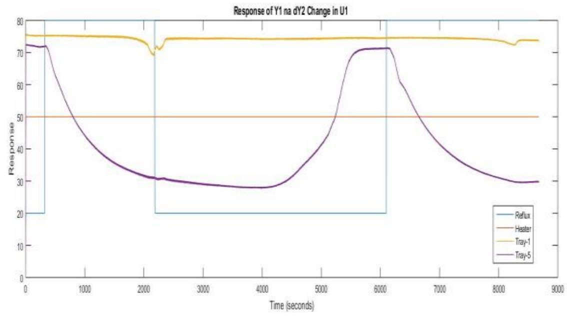

Figure 4. Experimental response of variables when reflux flow rate is manipulated as PRBS input

Once input-output data is obtained from the step response, using least square regression approach [12] [13], FODPT model parameters are obtained by Sarath and Arasu is as follows:

$$G(S) = \begin{pmatrix} \dfrac{-0.436e^{-1.58s}}{1.85s+1} & \dfrac{0.035e^{-1.16s}}{0.262s+1} \\ \dfrac{-0.096e^{-0.989s}}{0.198s+1} & \dfrac{0.088e^{-0.76s}}{0.349s+1} \end{pmatrix} \tag{4}$$

3. CONDITIONAL CONTROL METHODOLOGY

For sensitive processes, small change in manipulated variable (MV) influences larger effect in process variable (PV). Usually once PV reaches set-point (SP), MV gradually stabilizes. MV shows its reverse effect only if PV overshoots with respect to SP. At this point, the accumulation of error in the integrator will be higher which leads to integral windup phenomena, this could be restricted by introducing anti-reset windup [14].

Though anti-reset windup restricts the effect of saturated gain on the process and makes MV to act within the constraint limits, as soon as PV reaches SP, the MV slowly stabilizes. For sensitive processes, even short time taken for MV to stabilize will make PV undershoot in its response (eg little reflux flow rate on tray in the column reduces its temperature). Further MV acts as function of error results in lifting up PV to reach SP. Eventually the process keeps oscillating around the SP and takes more time for the controller to compensate and make stabilize the process.

A new conditional control structure is introduced to overcome the above problem by introducing the conditional switch in the control loop. Conditional control structure eliminates the drawback of overshoot and undershoot in the process by switching the actual gain (PID) loop to other input gain (K_{con}) by using conditional switch. Conditional gain 'K_{con}' can be selected as the minimum gain required for the process. Figure.5

represents the schematic diagram of conditional control structure.

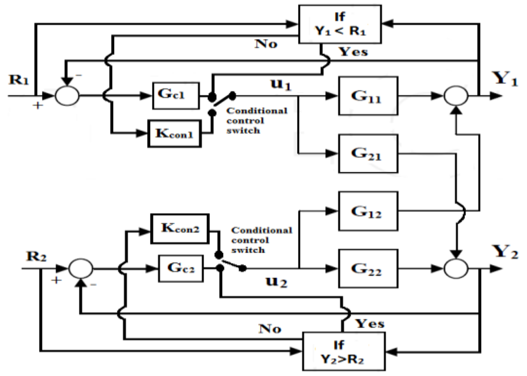

Figure 5. Conditional control structure schematic diagram

This structure helps to reduce overshoot and undershoot problem for the sensitive processes. The mathematical expression for MIMO process with conditional control structure is given by:

$$Y_1 = G_{11}u_1 + G_{12}u_2 \qquad (5)$$

Such that

$$u_1 = G_{c1} \quad \text{for } Y_1 < R_1$$

$$\text{else} \quad u_1 = K_{con1}$$

$$Y_2 = G_{21}u_1 + G_{22}u_2 \qquad (6)$$

Such that

$$u_2 = G_{c2} \quad \text{for } Y_2 > R_2$$

$$\text{else} \quad u_2 = K_{con2}$$

These conditions are not necessary to be same for all processes, it depends on the function of error and sensitivity of the process. Flow charts of the process with conditional control scheme as follows, Figure.6 shows flowchart for controller enables with respect to error effect on the process.

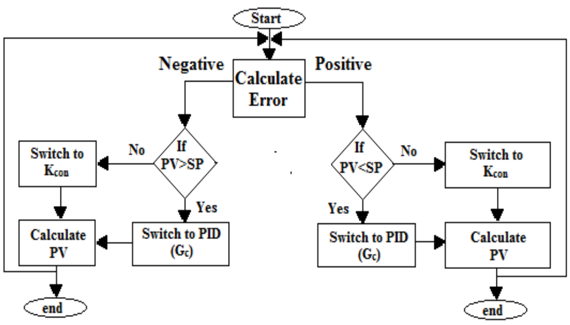

Figure 6. Flowchart of conditional control sequence with respect to error

Conditional gain (K_{con}) in this paper is considered as a function of PID controller gain with product of 'λ'.

$$K_{con} = G_c * \lambda \qquad (7)$$

Where 'λ' varies between (0, 1].

4. Result Anaysis

A. Analysis of MISO control loop

Simulation responses comparison has been carried out using IMC-PI [15], Skogestad's PI and Controller incorporated with conditional control structure using Skogestad's PI. Process variable and manipulated variable comparison simulation responses has been shown in Figure.7 and Figure.8 respectively.

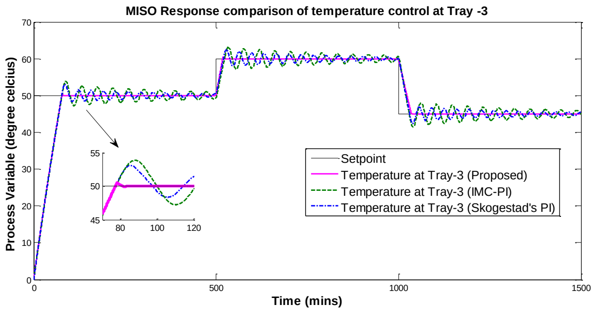

Figure 7. Closed loop simulation response of tray-3 temperature for MISO scheme

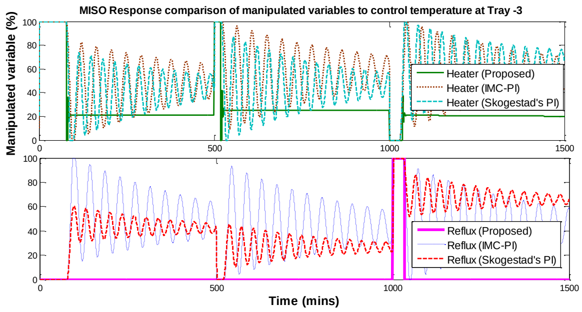

Figure 8. Manipulated Variable (MV) simulation response comparison of MISO scheme

To show the effectiveness of control structure, the experimentation is carried out with and without conditional control structure. Figure.9 depicts the closed loop response of tray-3 temperature without conditional control structure. Sustained oscillations in process variable is observed. Figure.10 represents corresponding manipulated variables response, it can be observed that manipulated variables are oscillating continuously within the band of operation. Therefore through the responses, it is conveyed that maximum actuator energy utilized in this operation.

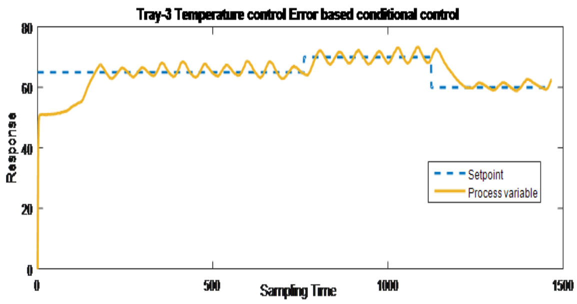

Figure 9. Experimental validation of conventional PI without conditional control scheme (MISO scheme)

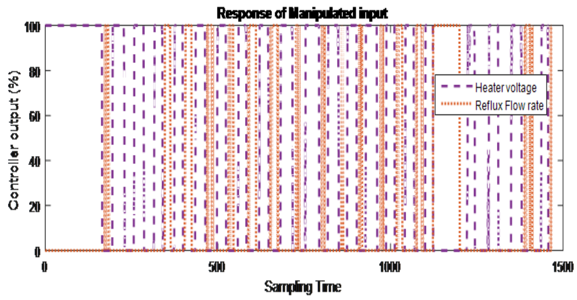

Figure 10. Response of manipulated variables without conditional control scheme

Figure.11 shows the closed loop experimental response with conditional control structure. Error limit to enable switch is set to -1, which means that if there exist negative error of '1', the switch needs to be enabled to pass reflux flow rate into the column. It is observed that mechanism is optimal in utilizing the actuators timely. At 1000sec of run time, servo change of '-15' is given to the process. In this case negative error exist about the gain of 10, it is observed that at this point of time heater is switched to '0' and reflux flow is enabled which is acting 100%. As soon as error reduced to more than '-1', heater is enabled and continued to maintain the process variable at desired level. Similarly real time switch is set to negative error magnitude of '2'.

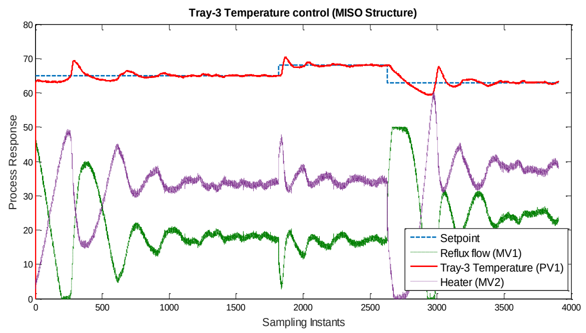

Figure 11. Experimental validation of skogestad's PI with conditional control structure (MISO scheme)

Plant models shown in equation.2 and equation.3 are subjected to 30% uncertainty from actual parameters to execute controller performance in servo and load operations. Table.1, Table.2 and Table.3 shows the performance indices of proposed, IMC-PI and Skogestad's PI controllers respectively.

Note: All the performance indices are recorded by exciting unit step change for both servo and load operations. This way of validating performance has been given by Chidambaram et al [16] but the authors has implemented for 10% uncertainty in plant parameters.

TABLE 1. MISO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR PROPOSED CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	4.93	18.98	116	645.6
	C-2	5.58	20.89	141.7	747.3
	C-3	5.191	21.17	142.9	864.7
	C-4	5.18	21.11	141.9	860.3
Load	C-1	3.16	16.6	79.89	812.7
	C-2	3.84	20.83	133.1	1191
	C-3	3.72	20.98	131	1346
	C-4	3.76	21.12	132.7	1359

TABLE 2. MISO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR IMC-PI CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	35.81	80.25	2376	8716
	C-2	42.42	93.45	3585	11140
	C-3	47.29	99.41	4136	11970
	C-4	47.35	99.48	4144	11980
Load	C-1	33.79	74.79	1847	7742
	C-2	38.57	87.01	2965	10200
	C-3	43.07	92.95	3447	11060
	C-4	43.1	93.01	3454	11070

TABLE 3. MISO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR SKOGESTAD'S PI CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	21.13	58.41	1176	6094
	C-2	26.7	71.76	2011	8254
	C-3	30.65	78.34	2503	9317
	C-4	30.79	78.6	2526	9365
Load	C-1	19.68	55.46	1032	5803
	C-2	24.27	67.72	1777	7790
	C-3	27.87	74.03	2225	8823
	C-4	27.99	74.76	2246	8868

Where,

C-1: Nominal Plant
C-2: 30% uncertainty in 'K'
C-3: 30% uncertainty in 'K' and 'τ'
C-4: 30% uncertainty in 'K', 'τ' and 'θ'
ISE: Integral Square Error
IAE: Integral Absolute Error
ITSE: Integral Time Square Error
ITAE: Integral Time Absolute Error

B. Analysis of MIMO control loop

MIMO process control is difficult and challenging because of variable interaction in the process. Conventional control design for MIMO process requires effective transfer functions for individual loops that defines the whole dynamics of the process. The method to obtain effective transfer function from actual model of equation.4 is given by Yadav et al [12]. The Process variable and manipulated variable simulation responses comparison of proposed, IMC-PI and Skogestad's PI are shown in the Figure.12 and Figure.13 respectively.

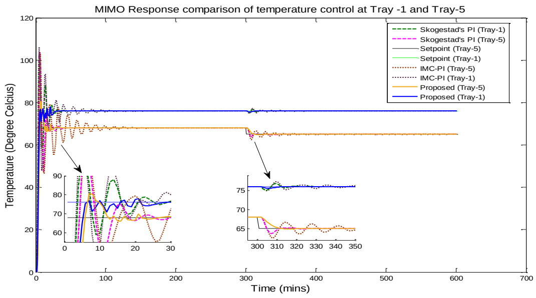

Figure 12. Closed loop simulation response of tray-1 and tray-5 temperatures for MIMO scheme

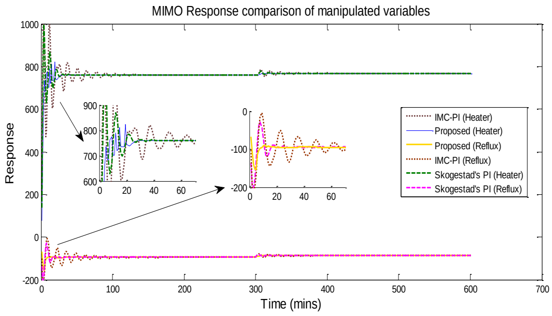

Figure 13. Manipulated Variable (MV) simulation response comparison of MIMO process

Conventional control response of MIMO process without conditional controller is given in the Figure.14. Oscillations of temperature at tray-5 is observed around the applied set-point. Heater is the manipulated input which changes as a function of error with respect to temperature of tray-1 and reflux flow rate changes as a function of error for temperature at tray-5. The increase in the temperature of tray-5 is because of heater voltage and reduce in the temperature is because of reflux flow rate which is noted as an interaction effect.

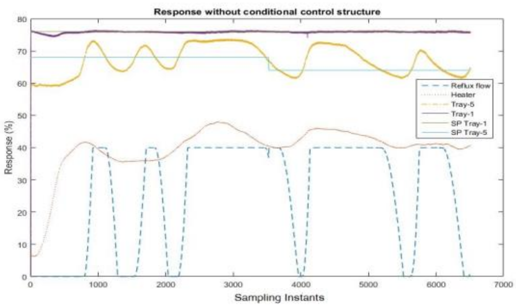

Figure 14. Experimental validation of conventional PI without conditional control scheme (MIMO scheme)

The interaction effect shown in Figure.14 can overcome by introducing condition control structure depicted in the Figure.5. The closed loop MIMO response with conditional controller is shown in the Figure.15. Skogestad's PI controller formula is used in implementation. By using conditional control structure, oscillations in the process variable (tray-5) gradually reduced and closed loop performance has been improved.

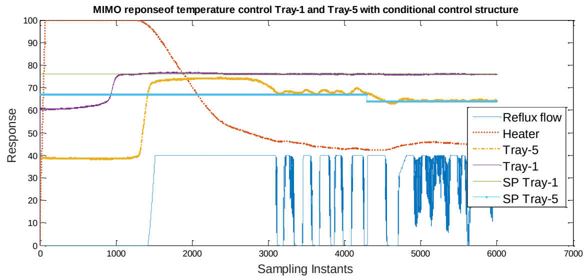

Figure 15. Experimental validation of skogestad's PI with conditional control scheme (MISO scheme)

Performance indices of MIMO process is slightly different from MISO process. As the process includes the variables interaction, the performance of loop-1 and loop-2 are calculated individually and summated to obtain overall performance. 30% parameter uncertainty for MIMO process of equation.4 is given in equation.8 Table.4, Table.5 and Table.6 shows the performance indices of proposed, IMC-PI and Skogestad's PI controllers respectively for MIMO process.

Note: It has been observed that in MIMO process performance of servo and load operations indices remains same. It is also observed that 30% uncertainty in dead time (i.e., C-4: 30% uncertainty in 'K', 'τ' and 'θ') for IMC-PI is making the closed loop response unstable.

$$G(S)=\begin{pmatrix} \dfrac{-0.306e^{-2.05s}}{2.406s+1} & \dfrac{0.0455e^{-1.508s}}{0.34s+1} \\ \dfrac{-0.0672e^{-1.28s}}{0.2874s+1} & \dfrac{0.1144e^{-0.988s}}{0.4537s+1} \end{pmatrix} \qquad (8)$$

TABLE 4. MIMO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR PROPOSED CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	7.16	12.89	17.7	61.03
	C-2	8.02	14.38	24.64	91.78
	C-3	8.55	15.31	26.42	91.81
	C-4	9.69	16.7	31.01	99.53
Load	C-1	7.72	13.19	19.56	59.05
	C-2	8.69	14.96	26.7	82.23
	C-3	9.14	15.55	29.7	85.65
	C-4	10.32	16.89	35.13	95.76

TABLE 5. MIMO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR IMC-PI CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	11.55	32.37	85.95	578
	C-2	19.75	68.14	416.74	3656.9
	C-3	66.12	213.28	6338.1	27460
	C-4	-	-	-	-
Load	C-1	11.55	32.37	85.95	578
	C-2	19.75	68.14	416.74	3656.9
	C-3	66.12	213.28	6338.1	27460
	C-4	-	-	-	-

TABLE 6. MIMO PERFORMANCE INDICES WITH PLANT UNCERTAINTY FOR SKOGESTAD'S PI CONTROLLER

	Cases	ISE	IAE	ITSE	ITAE
Servo	C-1	6.95	14.73	17.65	94.13
	C-2	7.23	14.97	21.57	100.21
	C-3	8.18	17.58	29.41	141.68
	C-4	19.21	53.39	270.99	1841.5
Load	C-1	6.95	14.73	17.65	94.13
	C-2	7.23	14.97	21.57	100.21
	C-3	8.18	17.58	29.41	141.68
	C-4	19.21	53.39	270.99	1841.5

From performance indices, it is observed that the conditional controller performance is significantly improved with respect to IAE and ITAE (ref: Table.4). As the conditional controller disables the controller action if the process variable (PV) overlaps the conditional limit. Therefore overshoot and undershoot of PV can be reduced. IAE and ITAE gives the performance in terms of absolute error, as overshoot and undershoot reduces, its indices has significantly improved.

Energy consumption with respect to actuator action using conditional control structure is observed by comparing the manipulated inputs in both MISO and MIMO schemes at closed loop steady state. In MISO scheme, it is observed that the manipulated variable (controller effort) varies between the minimum and the maximum value (i.e. 0% to 100%, ref: Figure. 10) without conditional control structure. While only 5% variation in control effort is observed on both manipulated variables (Heater and Reflux) at steady state by using conditional control structure (ref: Figure.11). In MIMO scheme, it is observed that the manipulated variable U_2 (Heater) varies between the values of 35% and 50% (15% variation) in Figure.14. Whereas, using conditional control structure the manipulated input reduced to 3% variation (i.e., 39% to 42%, ref: Figure.15). It is also observed that performance indices with conditional control presented in Table.4 is dominated compared to IMC-PI and skogestad's PI in Table-5 and Table-6 respectively. Through the above assertions, the claim of minimum control effort is accompanied.

5. Conclusion

In this paper system modeling using PRBS input for MISO process and step response excitation for MIMO process has been carried. Besides modeling, control design with a new conditional control structure has been presented. Binary distillation column is considered as a case study and implementation has been carried out for MISO and MIMO control configurations. Control scheme is implemented with 30% plant uncertainty in all model parameters. With an objective of actuator energy minimization, Simulation and experimental results has been presented to show the efficiency of control scheme.

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Eadala Sarath Yadav is a research scholar in ICE department, MIT, Manipal. He has 3 years of research experience. He worked as junior research fellow for a DST project, govt of India. His area of interest includes system modeling, advances in PID control and predictive control strategies.

Dr. I. Thirunavukkarasu is an Associate Professor (Senior Scale) in ICE department, MIT, Manipal. He is having 14 years of teaching/research experience. His aread of interest includes robust, H-infinity control, optimal control etc. He published more than 90 research articles in conferences/journals. He organized several workshops/conferences/FDPs, He worked as principle investigator of a DST project, Govt of India. Under the excellence in teaching and research policy conference scheme of MAHE he visited several contries for scientific conferences

Dr. S. Shanmuga Priya is an Associate Professor (Senior Scale) in chemical Engineering department, MIT, Manipal. She is having 10 years of teaching/research experience. Her area of interest includes Energy Engineering, Heat Transfer, Environmental Engineering and Pollution Control, Environmental Studies, Safety in Chemical Industries, Cryogenic Engineering, Bioenergy engineering, Industrial waste water engineering.