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Design and Implementation of Decupled PI Controller for MIMO Process

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*Abstract—***In this paper, PI controller with a simplified decoupled network is developed for controlling the interactive multivariable processes.The controller design is divided into two parts.The first part is designing Relative Gain Array(RGA) matrix in order to measure the amount of interaction among the loops.RGA alsogives a recommendation concerning the most effective pairing of controlled and manipulated variables. The second part is addinga decoupler with PI controller to the control loops to cancel the interaction effects. This proposed design method for multivariable decoupling controller is implemented toMIMO process (Pilot plant).Integral Squared Error (ISE) comparison was carried out to evaluate the performance of the controller. From the ISE analysis conclusions were drawn for the decupled PID controller to be the best performing controller in comparison to PI without decoupler. The controller although complex to understand is very effective in its output performance by providing optimal results.**

Index Terms: **PID controllers, multivariable control systems, decoupling.; PID controller; Pilot plant,Integral Squared Error (ISE)., Process modeling.**

I. INTRODUCTION

These days, MIMO (Multi-Input-Multi-Output) systems have become more and more widely used in industrial applications. A real example of MIMO system is Murdoch university’s pilot plant, which is located in Western Australia.The Murdoch University Pilot Plant was constructed in 1998 through the collaboration of Murdoch University along with industrial companies Alcoa of Australia, Honeywell Ltd, and Control and Thermal Engineering. The plant was initially designed to simulate the Digestion, Clarification, and Precipitation stages of the Bayer Process which is used to refine approximately 90% of Alumina in the world. The Digestion stage is simulated using the Supply tanks, Ball Mill, Ball Mill tank, Cyclone and Cyclone Underflow tank. The overflow of the Cyclone Tank, the Lamella and Needle tank simulate the clarification stage and the Precipitation stage is simulated using three CSTR’s.

The pilot plant’s process media is water so the event of a tank overflow is not going to create a critical safety situation, however, industrial systems such as the Bayer process rely on higher pressures and temperatures as well as corrosive fluids therefore it is good practice to ensure that tanks do not overflow. The scope of this project only required half of the pilot plant to be operated from the non-linear tank so the upstream vessels are isolated by closing the flow control that joins them[1-2].

In a multivariable system, a change in any one of the input variables results in a change in one or more output variables. This isanother context known as cross-coupling or interaction. A control loop design to an interacting MIMO system without considering the cross-coupling between the variables can result in an unsatisfactory result. Control loops designed to control different outputs can affect each other’s performance. Furthermore, a hidden feedback loop can develop leading to the destabilization of the system. The process can be stabilized by tuning controller parameters, but this would corrupt the performance of the controller. The loop configuration is also very sensitive and can be easily driven towards instability [3-4].

The purpose of this paper is to get initial exposure to the plant, analyze the basic behavior of plant variables and implement advanced control schemes. The paper starts by developing an RGA matrix in order to measure the amount of interaction among the loops. Then, implementing two control techniques such as PI controller andPI with decoupler control techniques. Finally, controller performance on the process is examined using ISE analysis.

II. PILOT PLANT BACKGROUND

Figure.1 shows the Murdoch University Pilot Plant which was constructed through the collaboration of Murdoch University along with industrial companies Alcoa of Australia, Honeywell Ltd, and Control and Thermal Engineering[1].

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Figure 1.Overview of the whole pilot plant

The Supply Tanks are the major source for providing the plant with the feed. Alternate tanks can be utilized to provide a source of disturbances to test control strategies. The feed is pumped to the Ball Mill and mixed with the recycled streams from the Cyclone Underflow Tank and the Lamella Separator. The mixture then travels through the Ball Mill to the ball mill pump and is then pumped into the Hydro-Cyclone. Part of the material is separated in the Hydro-Cyclone and the underflow is passed to the Cyclone Underflow Tank. The outflow from this tank is recycled to the supply tanks and the Ball Mill. The overflow from the Hydro-Cyclone goes to the Lamella Separator, where most of the material is removed and pumped back to the Supply Tanks. The overflow goes to the Needle Tank. The second supply of water to the Needle Tank is through Nonlinear Tank which is used to create disturbances in the level control in this tank. The outlet of the Needle Tank flows to the three Heated Tanks.

The pilot plant uses Honeywell's Experion series SCADA system. The HMI software is called Stationand allows for control and trend monitoring.Experion is compatible with the use of Microsoft Excel Data Exchange (MSDE) which allows for data to be shared with programs such as Microsoft Excel. This will be elaborated upon in the next report where controllers to be implemented within Microsoft Excel. The relevant Point IDs for the system are given in the table below:

Table 1. Point IDS of processvariables

Process Variables	**Point ID**
Levels	
Non-Linear Tank	LT_542
Needle Tank	LT_501
CSTR3	LT_667
Pump Speeds	
NonLinear Tank outflow	FDP_521
Needle Tank outflow	NTP_561
CSTR3 Outflow	PP_681
FCV positions	
Raw Water Feed	FCV_622

III. DECOUPLING DESIGN

There is a one-way interaction between the Nonlinear tank system and Needle tank system therefore, a decoupler needed to be implemented. First, the change in needle tank manipulated variable (FDP-521) will affect nonlinear tank system because the needle tank manipulated variable is the output pump of the nonlinear tank. The relationship between FCV-541 with the level in Needle tank and the relationship between FDP-521 versus Needle tank level and Non linear tank level needs to be figured out. From there, the equation can be deduced and the gain of the system can be found so it can be used in RGA matrix.

A. RGA Analysis

Relative Gain Array is an analytical tool used to determine the optimal input-output variable pairings for a multi-input-multi-output (MIMO) system.There are two methods to calculate the RGA, the first, called the open-loop closed-loop method, where once the system is under control, one loop is opened to determine the effect of input variables on the output variables, the experimentally obtained data is then compiled the results into an RGA matrix. By definition, the Relative Gain Array can be calculated by the following.

$$\lambda_{ij} = \frac{(\partial y_i / \partial m_j)_m}{(\partial y_i / \partial m_j)_y} = \frac{\text{open - loop gain}}{\text{closed - loop gain}} \tag{1}$$

Once every combination of λ has been calculated, it is then put in an array in the form shown below.

$$\Lambda = \begin{array}{c} \\ y_1 \\ y_2 \\ \cdots \\ y_3 \end{array} \begin{array}{c} \begin{array}{cccc} m_1 & m_2 & \cdots & m_n \end{array} \\ \begin{bmatrix} \lambda_{11} & \lambda_{12} & \cdots & \lambda_{1n} \\ \lambda_{21} & \lambda_{22} & \cdots & \lambda_{2n} \\ \cdots & \cdots & \cdots & \cdots \\ \lambda_{n1} & \lambda_{n2} & \cdots & \lambda_{nn} \end{bmatrix} \end{array} \tag{2}$$

This method requires controllers to be implemented and is particularly tedious and time-consuming. The second method to obtain the RGA uses the steady-state gain matrix which is again can be determined experimentally; To calculate steady-state gains, the plant is run at steady state (using the operating points found previously) and step changes are conducted on in each manipulated variable, one at a time, and the ultimate change in each process variable is measured to find the open-loop gain array.
Open loop step testing was conducted on the steam FCVs supplying the CSTRs and a 1st order approximation (with time delay) was used to find the transfer functions.However, the tanks levels, which as pure capacity systems, behave slightly different as pure capacity systems do not have an ultimate gain since they ramp in response to a step:

The transfer function of a pure capacity system is given as:

$$y(s) = \frac{K}{s} \qquad (3)$$

When converted to the time domain:

$$y(t) = AK^{*}t \qquad (4)$$

The slope, m, is given by:

$$m = AK^{*} \qquad (5)$$

Where A is the magnitude of the step, the system's gain, K* is:

$$K^{*} = \frac{m}{A} \qquad (6)$$

The following figures show the results of the step change experimentation:

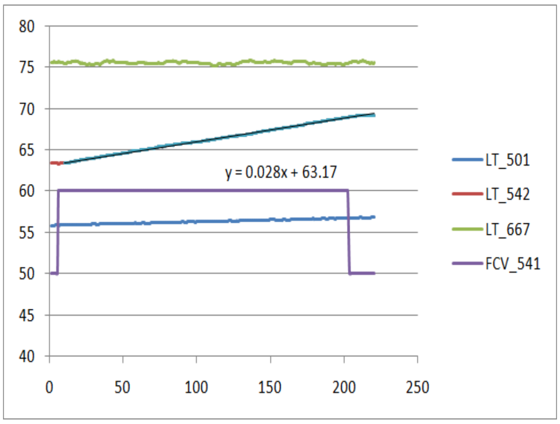


Figure 2 . Raw water valve (FCV_541) step response

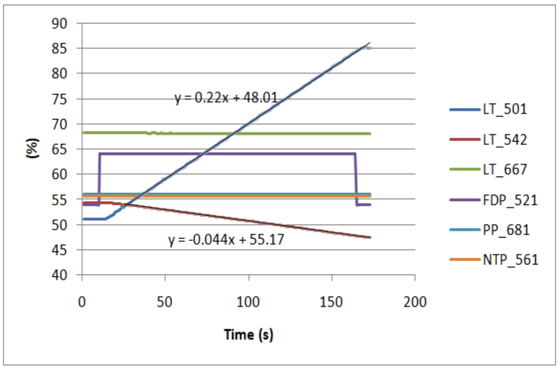


Figure 3. FDP_521 step response

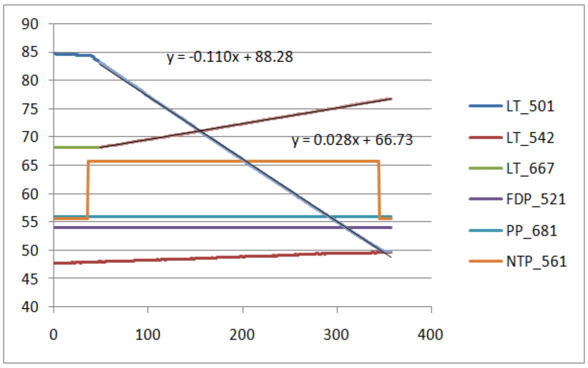


Figure 4. NTP_561 step response

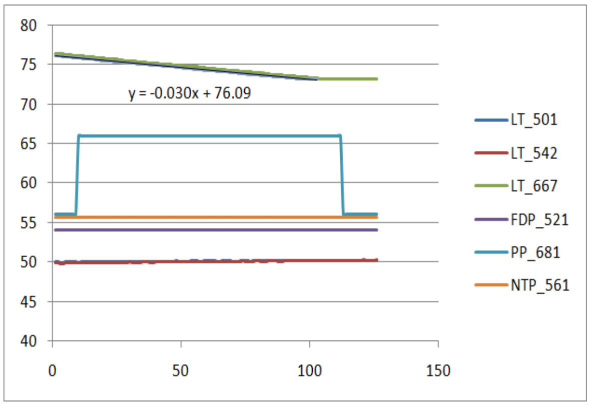


Figure 5. PP_681 step response

By taking gradients of the responses and dividing them by the magnitude of each respective step, the steady State Gain matrix can be generated:

Table 2. StateE gainmatrix

Process Variable		Raw Valve	FDP-521	NTP-561	PP-681
	NLT (%H)	0.00281	-0.00447	0	0
	NT (%H)	0	0.022	-0.0111	0
	CSTR3 (%H)	0	0	0.0028	-0.00303

The pseudo-inverse of this gain matrix was performed using Matlab script. This pseudo inverse was transposed as suggested by the RGA formula. The resultant RGA matrix was calculated by performing element by element multiplication of gain matrix and the inverse transpose of the gain matrix. A resultant matrix shown in the table below was obtained, Which shows one way interaction as expected.

Table 3.RGA analysiswith pairings highlighted

Process Variable		**Raw Valve**	**FDP-521**	**NTP-561**	**PP-681**
	NLT (%H)	0.76598	0.234	0	0
	NT (%H)	0	0.66	0.326	0
	CSTR3 (%H)	0	0	0.31022	0.68977

B. Loop Pairing from RGA

By studying these 3 relative gain arrays the pairing of inputs and outputs was selected yielding the following;

- NLT level controlled by FDP-521
- NT level controlled by NTP-561
- CSTR3 level controlled by PP-681

The final P&ID of the Pilot plant is shown in figure 6.

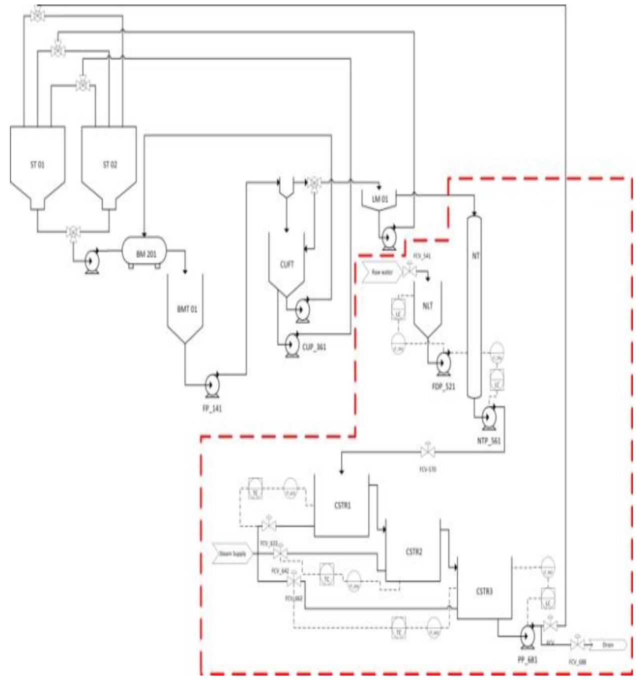


Figure 6. Final P&ID showing controllers

C. Design of PI Controllers with Decouplers

Decouplers can be used to improve the performances of multiloop control systems by compensating for process interactions. Similar in function to a feed-forward controller, decouplers eliminate disturbance by compensating for any interaction created by the cross coupling of the process variables. In an ideally decoupled system, the action of an MV should have no effect on the PV's other than theone which the MV is paired with.
Decouplers require accurate process models otherwise they can degrade overall control performance.

Decoupler design can be significantly simplified by only implementing Steady-State Decouplers which eliminate only steady-state interactions from all loops. Such decouplers only use a gain ratio rather than a full discretized transfer function.Level Decouplers were implemented between both the NLT and the NT and between the NT and CSTR as there was found to be strong interaction.
Figure 7. shows the design of decoupling on non-linear tank and needle tank. The new control action for adding decouple on Non linear tank system can be implemented by using an equation. 8and 9.

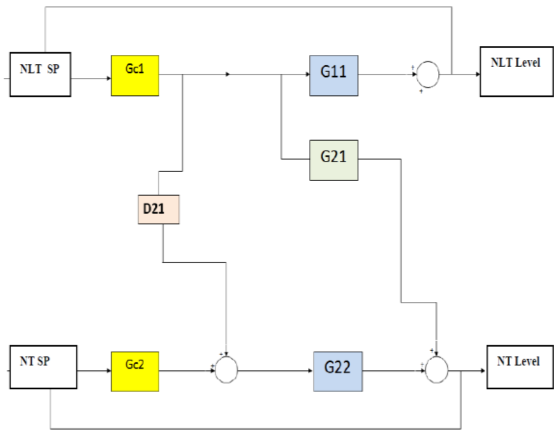


Figure 7.Decoupling Design on Non Linear Tank and Needle Tank

Table 4. Steadystate gain matrix (K)

K_{11}=	-0.00447	K_{12}=	0
K_{21}=	0.022	K_{22}=	-0.0111

Where the K values were found during the RGA development and are taken from the steady state gain matrix.

$$D_{21} = -\frac{G_{21}}{G_{22}} = -\frac{K_{21}}{K_{22}} = -\frac{0.022}{-0.0111} = 2 \quad (7)$$

Control action =Gc1+D21*Gc2 (8)

IV. PI TUNING METHOD

As the control system in the pilot plant is discrete in nature, position or velocity form PI controllers must be used. Controllers were implemented in the pilot plant using the velocity form equation rather than the Position form as it is simpler to implement (not requiring the total sum or error to be calculated, only the previous three errors) and as such, has inherent anti reset windup.
The velocity form equation is given as:

$$\Delta U(k) = \mathrm{K}\left[\left(1+\frac{\Delta t}{Tau_I}\right)\mathcal{E}(k) - \mathcal{E}(k-1)\right] (9)$$

Where:

$$\mathcal{E}(k) = SP - Y(k-1); \ \ \mathcal{E}(k-1) = SP - Y(k-2)$$

The change in MV is then added to the previous MV:

$$U(k) = \Delta U(k) + U(k-1) \quad (10)$$

Where:
$u(k)$ = control action(manipulated variable)
$\Delta u(k)$ = change in control action
Kc = controller gain
Ti = integral time
Δt = sampling time
$e(k)$ = current error
$e(k-1)$ = previous error

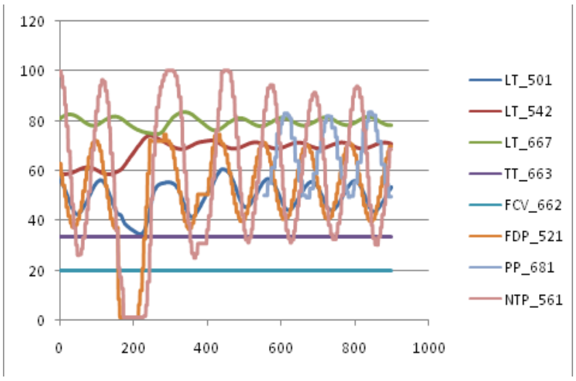
The velocity form of the controller was chosen to be implemented in the plant to allow for better accuracy and to reduce the complexity.It should be noted that before being written to the server, MVs were conditioned such that they saturated at 100% and would not drop below 1% (to prevent the low-flow protection interlocks from turning off the pumps.)In the case of controlling the NLT, testing from the previous report showed that the duty of the outflow pump had to be limited to 75% to avoid the Needle tank potentially overflowing. The final (hand tunned) controller parameters used in testing can be seen in the following graph:

Table 5. Picontroller parametersusing manualtuning

$\Delta t = 5s.$			
PV	***NLT***	***NT***	***CSTR3(L)***
MV	NLT-UFP	NT-UFP	PP
Kc	-5	-5	-5
Taui	100	100	100

IV. EXPERIMENTAL RESULTS

The following plot shows the effect of the unstable NLT controller on the whole plant.



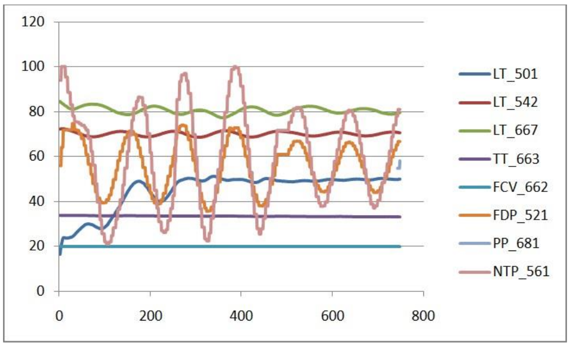


Figure 8. Unstable NLT SP change with and without decouplers

This testing did demonstrate the benefits of decouplers even with the NLT level control on the verge of instability, the decouplers were effective at significantly compensating for disturbance to the Needle tank .The following section illustrates the controller performance comparisons for each tank :

A. Set-Point Tracking in Non-Linear Tank Level

Figure9.shows time response with and without decouplers. Note that the interaction between Non linear tank system and Needle tank decreases. If this fact is quantified by means of some measure as the ISE test, it can be proven that in the response without decouplers, ISE is 6660.811, while when decouplers are applied it is 6276.007.

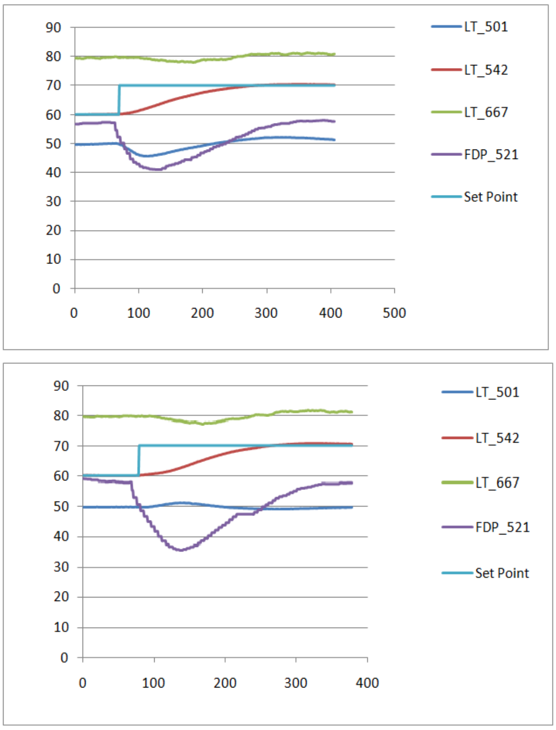


Figure 9. Set-point change in NLT level from 60% to 70% (with and without decoupler)

B. Set-Point Tracking in Needle Tank Level:

The plot in figure 10 shows needle tank response to setpoint changes from 50-60% while under the control of PI with and without decoupler. It can be seen that the decoupler system response smoothly transitions between old and new set-points with little or no associated noise.

Its manipulated variable too moves in a decisive manner smoothly adjusting itself to accommodate for changes in the tank's level.

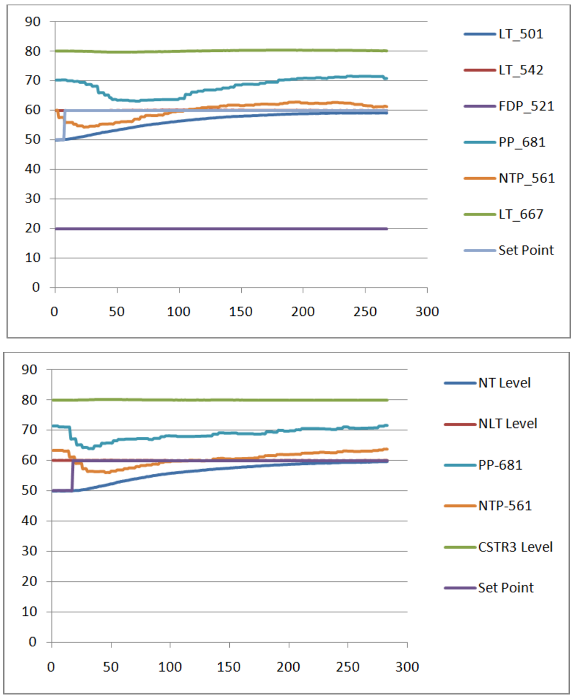


Figure 10. Set-point change in NT level from 50% to 60% (with and without decoupler)

C. Set-Point Tracking in CSTR3 Tank Level:

In this case, just PI controller has been implemented. a set point change was made From CSTR3 level from 80% to 90% .From figure 11 below, it can be seen that the process variable which illustrated in green color has a slight overshoot in tracking of set pint and oscillations in the manipulated variable which was represented by pump pp-681 in orange color.

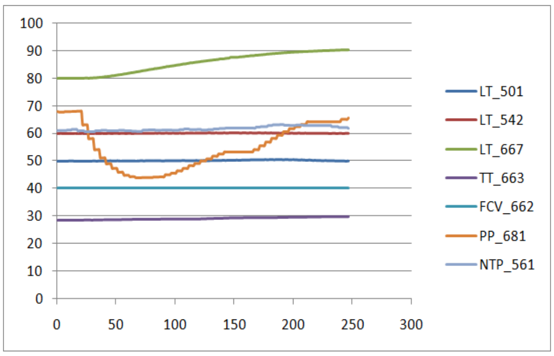


Figure 11. Set-point change in CSTR3 level from 80% to 90% (without decoupler)

D. Disturbance Rejection for Tank Levels:

The robustness of the decuopler is evaluated by disturbing the system by opening the Non-Linear Tank outflow pump. Note that interaction effects decrease not very significantly.The robustness performance of the controller is shown in Figure 12.

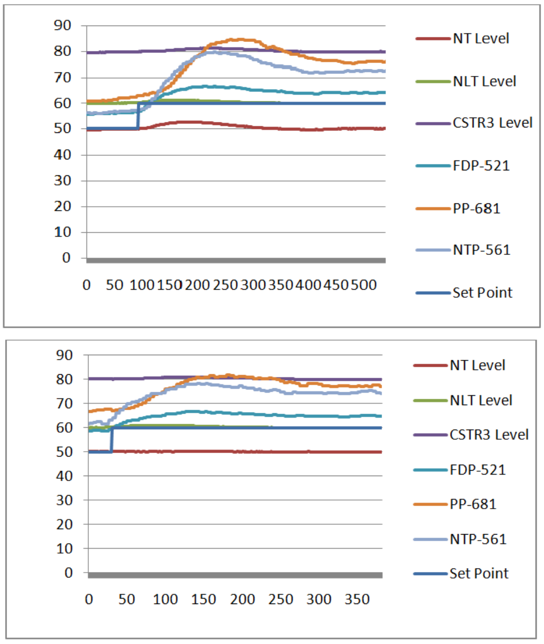


Figure 12.Disturbance rejection test for 3 tanks (with and without decoupler)

V. CONTROLLER PERFORMANCE COMPARISON

To determine what control strategies are more effective, an objective measure of controller performance must be used. Beyond just reaching and maintain the set point, some of the important transient characteristics include:

- Minimum settling time
- Fastest rise time
- Maximum Overshoot?
- Decay Ratio

Beyond these, more sophisticated performance analysis tools that use a function of transient error and time to quantify performance have been developed. Such tool includesSquared Error (ISE):

$$ISE = \int_0^{\infty} e^2(t)dt \qquad (11)$$

Table 6 .Controllerperformancecomparison: ISE valuesfordifferentcontrolstrategies

	PI Controller	PI with Decoupler
NLT Level	6660.811	6276.007 (N/A)
NT Level	5153.455	4774.064(N/A)
CSTR3 Level	5352.57	Not Implemented
Raw water disturbance ∑ error for 3 tank levels	1020.299	157.7953

From the performance criteria table above, the least value of ISE can be observed of PI with decupler controller for the two tank tests. The ISE test for the disturbance rejection using decoupler can be observed almost 10 times less compared to PI controller alone .

VI. CONCLUSION

This paper has detailed MIMO control of the pilot plant using a variety of control strategies.The paper has started with obtaining steady state gain matrix of the process. Then calculating RGA matrix which provided a recommendation concerning the most effective pairing of controlled and manipulated variables. Finlay, PI control was implemented and then performance was improved by designing decouplers. To demonstrate the advantages of PI with decupler over the conventional PI controller, ISE comparisons have been presented. From this comparison, we conclude that decuolers can deliver good performance in terms of setpoint tracking and disturbance rejection.

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