GAIN SCHEDULING CONTROL DESIGN FOR SHELL HEAVY OIL FRACTIONATOR COLUMN

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ABSTRACT: *Multivariable system is usually characterized with loop interactions which normally have deteriorating effect on closed loop performance. Thus, there is need to decouple the system for efficient performance of the multivariable feedback system. In this work, dynamic and static compensators were used to remove loop interactions. Inverse of the steady state gain was used as static compensator while dynamic compensator elements were obtained using feedforward design technique. These were applied to design feedback control system for Shell Heavy Oil Fractionator (SHOF) using Proportional integral (PI) control settings. PI controllers for the plant were tuned using Ziegler-Nichols, Tyreus-Luyben and PID modules built in MATLAB for different plant parameters. The closed loop system was implemented based on gain scheduling strategy. Good control performance was achieved using settling time, rise time and overshoot as performance metrics for the control strategy*.

KEYWORDS: Compensators, Shell heavy oil Fractionator (SHOF), PI controller, multivariable system, gain scheduling

INTRODUCTION

Distillation is one of separation techniques used in chemical process industries to achieve separation of mixture of two or more components based on differences in their boiling points. It is a separation process which has received significant attention for better integration in chemical plants. Distillation is the principal separation method used in oil refinery to obtain different petroleum products at various boiling ranges from crude oil. Crude distillation unit is used to separate crude oil into different products depending on the difference of boiling temperature. These products can be gases, naphtha, kerosene, diesel, gasoline and other heavy products which can be recovered either as a final product or feedstock to other unit in the plant refinery for further processing. It is highly energy intensive process unit and represents one of the most important accesses for energy integration in a refinery (Mohammed *et al.*, 2007). Thus, efficient performance of the unit is required to minimize energy usage. One of the ways to achieve this is by having the plant under automatic control.

Control is needed for most of industrial chemical processes because of transient process behavior that occurs during start-ups and shut downs, unusual process disturbance and in order to ensure that a process operates at a desired operating condition, safely and efficiently (Luyben, 1996). However, like many industrial processes, distillation is associated with high non-linearity, time-varying dynamic behavior, unpredictable parameter deviations, parameters uncertainties and external disturbances (Nguyen, 2009). In addition, the column is associated with loop interactions during operation because it is a multivariable system in which most of

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the variables are required to be maintained close to their optimum operating values for efficient performance during plant operation. These make the control of the column a challenging task. Process control is about making the process output to behave in the desired way by manipulating the process input in an automatic way. Proportional Integral Derivative (PID) controller is the most widely employed feedback controller types in industries (Vilanova, 2008, Goncalves *et al.,* 2008). This is due to its simplicity (only three parameters to adjust), robustness to uncertainties and performance characteristics and it is used to improve the quality the products while optimizing the utility consumption (Seborg *et al.*, 2004). The simplicity and performance characteristics of PI/PID controller make it easier to understand than most of the advanced controllers (Vilanova, 2008).

Many researchers have applied different advanced control techniques in the control of distillation column. Among them is distributed model predictive controller (MPC) using Nash – optimization technique proposed by Li *et al.,* (2005), in order to reduce the computational complexity while maintaining the satisfactory performance of the plant. Also, Julian *et al.*, (2008) used simplified MPC on the SHOF model to obtain tracking performance and stability characteristics. Michael, (2003) used squared MPC for constraint handling performance and controlling the system. However, advanced controllers are complex, cumbersome with enormous computational burden and require significant effort and skill to tune and also difficult to implement (Zhang *et al.,* 2007, Wahab *et al.,* 2007). In addition, there are some limitations in terms of stability and robustness (Yela, 2009) with high maintenance cost, lack of flexibility and associated complexity (Hugo, 2000).

PID controller consists of proportional, integral and derivative terms and due to this, it is called three –mode control scheme. PID controller parameters (K_p, K_I and K_D) can be tuned to control a system so as to meet a specific process requirement. These parameters can be adjusted using some empirical methods like Ziegler-Nichols method, cohen and cohoon method, Tyreus-Luyben (Doust *et al.*, 2012). Also, the response of the controller can be described in terms of the responsiveness of the controller to an error, the degree to which the controller overshoots the set point and the degree of system oscillation.

However, PI/PID controller mostly employed for single –input single – output (SISO) process but due to their wide control scope and greater accuracy, multivariable PI/PID control methods that takes loop interactions into consideration have received significant attention (Zhang *et al.*, 2007). Furthermore, multivariable process is usually encountered with destabilization of the system due to loop interactions while adjusting the controller parameters (loop interaction) of one loop that affects the performance other loops in which decoupling techniques are used to remove these coupling effects. Wahab *et al.*, (2007) analyzed multivariable PI/PID control schemes for wastewater systems and showed that they are suitable for MIMO control loops that experience interactions. Also, Zhang *et al.*, (2007) used a backstepping-based decentralized PI/PID control scheme for the SHOF problem to reduce loop interaction.

PID controller can only perform satisfactorily for nominal plant however, when the plant is perturbed, the closed loop may have poor response in term of performance and stability because of change in plant parameters. One of the ways to circumvent these problems is through the use of gain scheduling protocol. Gain scheduling controller which ia also called an open-loop adaptive control system is designed to achieve effective controller performance by changing

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the parameters of the controller to eliminate the effect of parameter variations (Landau *et al.*, 2011). Krishna *et al.*, (2012) used a gain scheduling controller to design and simulate a ball and beam system, the performance metrics used showed that gain scheduling controller is faster than the normal PID controller.

METHODOLOGY

Distillation column or fractionators are widely used equipment in petrochemical industries or refineries for initially separation of the crude fractions to different product draws by cooling down the mixed-phase oil feed and also require an intensive energy for their operation (Lee *et al.,* 2011). In Shell Heavy Oil Fractionator (SHOF), heavy oil split into several streams which are further processed downstream. Fractions of this oil are divided to different products before leaving the fractionator with the aid of reflux flows that enhance the separation procedure. The reflux flow helps to control the top product composition while the heat input is used to control the bottom product composition (Zhang *et al.*, 2007). It has three product draws and three side circulating loops as shown in Figure 1. The three circulating loops eliminate heat to achieve the desired product separation. The heat requirement of the system varies, because the streams are reboiled in other parts of the plant. It has three heat exchangers in these loops which are used to recover energy from the recycling streams since vaporization of the feed stream consumes much energy (Morari *et al.,* 2002).

Also, the bottom reflux loop contains an enthalpy controller that regulates heat removal in the loop by adjusting the steam production. This heat duty is a manipulated variable that helps in controlling the column and the heat duties behaves as the disturbances to the entire column (Li *et al.*, 2005). Furthermore, the product specifications for the top and side draw streams are determined by economics and operating requirements and none for the bottom draw but operating constraint on the temperature in the lower part of the column is present (Micheal, 2003; Li *et al.*, 2005).

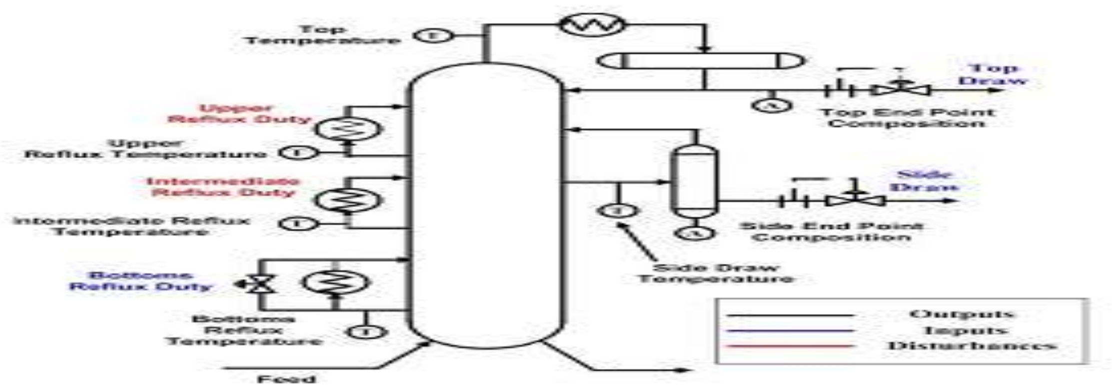

Figure 1: Shell's Heavy oil fractionator (Shead *et al.*, 2007)

Shell presented a process model in 1987, 'SHOF model', to serve as a standard performance test for new control techniques (Prett and Morari, 1988). The process model generally has seven measured outputs, three outputs and two input disturbances as shown in Table 1(Shead *et al.,* 2007). The plant is designed to handle some gain uncertainties and disturbances associated with the plant as shown in Table 2 and Table 3 respectively and to handle the following constraints:

a. $-0.5 \leq u_1$, u_2 and $u_3 \leq 0.5$.

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b. $|\Delta u_1|, |\Delta u_2|, |\Delta u_3| \leq 0.05$
c. $-0.5 \leq y_1 \leq 0.5$, $y_7 \geq -0.5$
d. Sampling time ≥ 1 min.

Input / Output	TD (u_1)	SD (u_2)	BRD (u_3)	URD (d_1)	IRD (d_2)
TEP (y_1)	$\frac{4.05e^{-27s}}{50s+1}$	$\frac{1.77e^{-28s}}{60s+1}$	$\frac{5.88e^{-27s}}{50s+1}$	$\frac{1.20e^{-27s}}{45s+1}$	$\frac{1.44e^{-27s}}{40s+1}$
SEP (y_2)	$\frac{5.39e^{-18s}}{50s+1}$	$\frac{5.72e^{-14s}}{60s+1}$	$\frac{6.90e^{-15s}}{40s+1}$	$\frac{1.52e^{-15s}}{25s+1}$	$\frac{1.83e^{-15s}}{20s+1}$
TT (y_3)	$\frac{3.66e^{-2s}}{9s+1}$	$\frac{1.65e^{-20s}}{30s+1}$	$\frac{5.53e^{-2s}}{40s+1}$	$\frac{1.16}{11s+1}$	$\frac{1.27}{6s+1}$
URT (y_4)	$\frac{5.92e^{-11s}}{12s+1}$	$\frac{2.54e^{-12s}}{27s+1}$	$\frac{8.10e^{-2s}}{40s+1}$	$\frac{1.73}{5s+1}$	$\frac{1.79}{19s+1}$
SDT (y_5)	$\frac{4.13e^{-5s}}{8s+1}$	$\frac{2.38e^{-7s}}{19s+1}$	$\frac{6.23e^{-2s}}{10s+1}$	$\frac{1.31}{2s+1}$	$\frac{1.26}{22s+1}$
IRT (y_6)	$\frac{4.06e^{-8s}}{13s+1}$	$\frac{4.18e^{-4s}}{33s+1}$	$\frac{6.53e^{-1s}}{9s+1}$	$\frac{1.19}{19s+1}$	$\frac{1.17}{24s+1}$
BRT (y_7)	$\frac{4.38e^{-20s}}{33s+1}$	$\frac{4.42e^{-22s}}{44s+1}$	$\frac{7.2}{19s+1}$	$\frac{1.14}{27s+1}$	$\frac{1.26}{32s+1}$

Table 1: Model showing the SHOF control problem

Input / Output	TD (u_1)	SD (u_2)	BRD (u_3)	URD (d_1)	IRD (d_2)
TEP (y_1)	$4.05+2.11\varepsilon_1$	$1.77+0.39\varepsilon_2$	$5.88+0.59\varepsilon_3$	$1.2+0.12\varepsilon_4$	$1.44+0.16\varepsilon_5$
SEP (y_2)	$5.39+3.29\varepsilon_1$	$5.72+0.57\varepsilon_2$	$6.90+0.89\varepsilon_3$	$1.52+0.13\varepsilon_4$	$1.83+0.13\varepsilon_5$
TT (y_3)	$3.66+2.29\varepsilon_1$	$1.65+0.35\varepsilon_2$	$5.53+0.67\varepsilon_3$	$1.16+0.08\varepsilon_4$	$1.27+0.08\varepsilon_5$
URT (y_4)	$5.92+2.34\varepsilon_1$	$2.54+0.24\varepsilon_2$	$8.10+0.32\varepsilon_3$	$1.73+0.02\varepsilon_4$	$1.79+0.04\varepsilon_5$
SDT (y_5)	$4.13+1.71\varepsilon_1$	$2.38+0.93\varepsilon_2$	$6.23+0.30\varepsilon_3$	$1.31+0.03\varepsilon_4$	$1.26+0.02\varepsilon_5$
IRT (y_6)	$4.06+2.39\varepsilon_1$	$4.18+0.35\varepsilon_2$	$6.53+0.72\varepsilon_3$	$1.19+0.08\varepsilon_4$	$1.17+0.01\varepsilon_5$
BRT (y_7)	$4.38+3.11\varepsilon_1$	$4.42+0.73\varepsilon_2$	$7.2+1.33\varepsilon_3$	$1.14+0.18\varepsilon_4$	$1.26+0.18\varepsilon_5$

Table 2: Structure of the variations incorporated in the mode

	d_1	d_2	ε_1	ε_2	ε_3	ε_4	ε_5
Case I	0.5	0.5	0	0	0	0	0
Case II	-0.5	-0.5	-1	-1	-1	1	1
Case III	-0.5	-0.5	1	-1	1	1	1
Case IV	0.5	-0.5	1	1	1	1	1
Case V	-0.5	-0.5	-1	1	0	0	0

Table 3: Cases of disturbances and gain uncertainties

where TEP = Top End Composition (Kmol/hr)
SEP = Side end Composition (Kmol/hr)

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TT= Top Temperature (F); URT =Upper Reflux Temperature (F)
SDT= Side Draw Temperature (F); IRT=Intermediate Reflux Temperature (F)
BRT = Bottom Reflux Temperature (F); TD = Top Draw; SD = Side Draw
BRD = Bottom Reflux Duty; URD = Upper Reflux Duty
d = disturbance ε = gain variation

This work is limited to the three manipulated variables (u_1, u_2 and u_3) that have direct influence on the Top End point Composition (y_1), Side End Point Composition (y_2) and the Bottom Reflux Temperature (y_7) which is in accordance with the product specifications (Jusagemal *et al.*, 2011) which resulted in three manipulate input – three controlled output system as shown in Equation 1.

$$G_p = \begin{bmatrix} \frac{4.05e^{-27s}}{50s+1} & \frac{1.77e^{-28s}}{60s+1} & \frac{5.88e^{-27s}}{50s+1} \\ \frac{5.39e^{-18s}}{50s+1} & \frac{5.72e^{-14s}}{60s+1} & \frac{6.90e^{-15s}}{40s+1} \\ \frac{4.38e^{-20s}}{33s+1} & \frac{4.42e^{-22s}}{44s+1} & \frac{7.2}{19s+1} \end{bmatrix} \qquad (1)$$

The block diagram of a multivariable (3 manipulated input – 3 controlled output) system in shown in Figure 2.

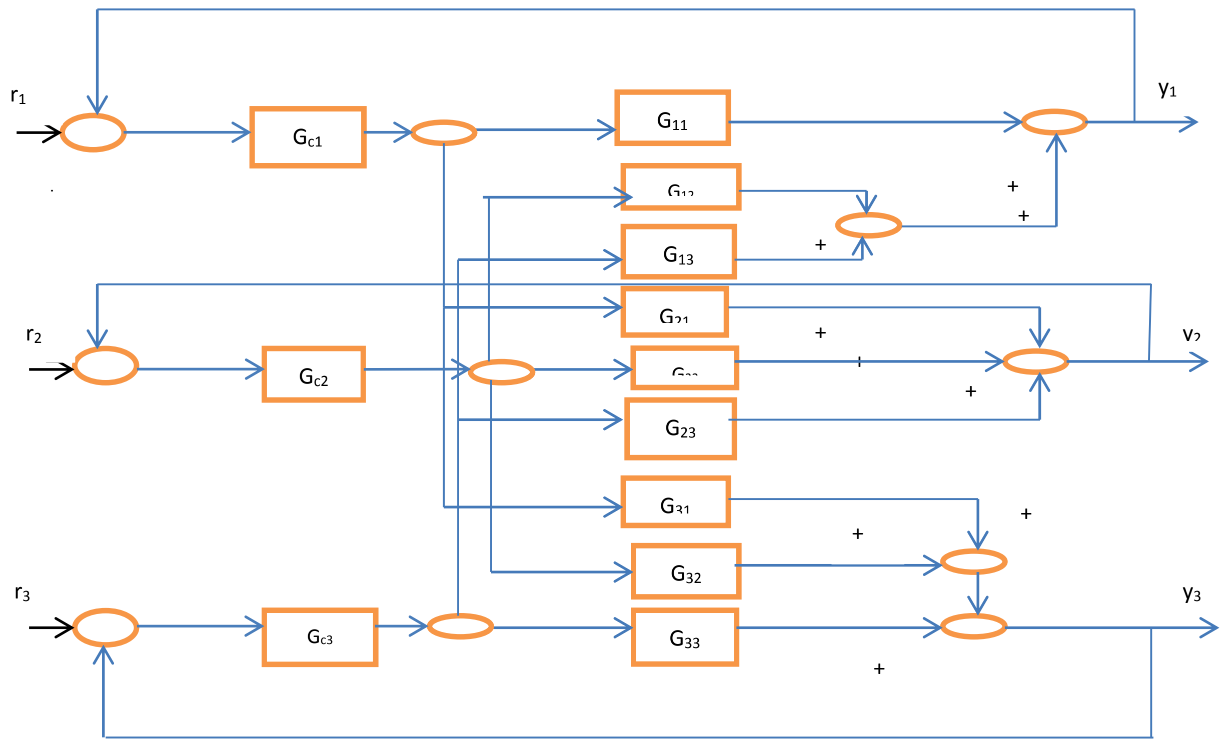

Figure 2: Block diagram of a multivariable (3 input – 3 output) system

Steady state Relative Gain Array (RGA) was used to determine the best input manipulative – out controller pairing. This was calculated as:

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} & \lambda_{13} \\ \lambda_{21} & \lambda_{22} & \lambda_{23} \\ \lambda_{31} & \lambda_{32} & \lambda_{33} \end{bmatrix} \qquad (2)$$

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where λ_{ij}= relative gain which relates the i[th] controlled variable with j[th] manipulated variable.

$$\lambda_{ij} \triangleq \frac{\left({\partial y_i}/{\partial u_j}\right)_u}{\left({\partial y_i}/{\partial u_j}\right)_y} = \frac{open\ loop\ gain}{closed\ loop\ gain} \tag{3}$$

where u= input, y= output

A decoupling technique was designed to compensate for interactions in a process and reduce control loop interactions. Decoupling techniques considered are dynamic decoupling and static decoupling. The block diagram of a decoupler plant is shown in Figure 3.

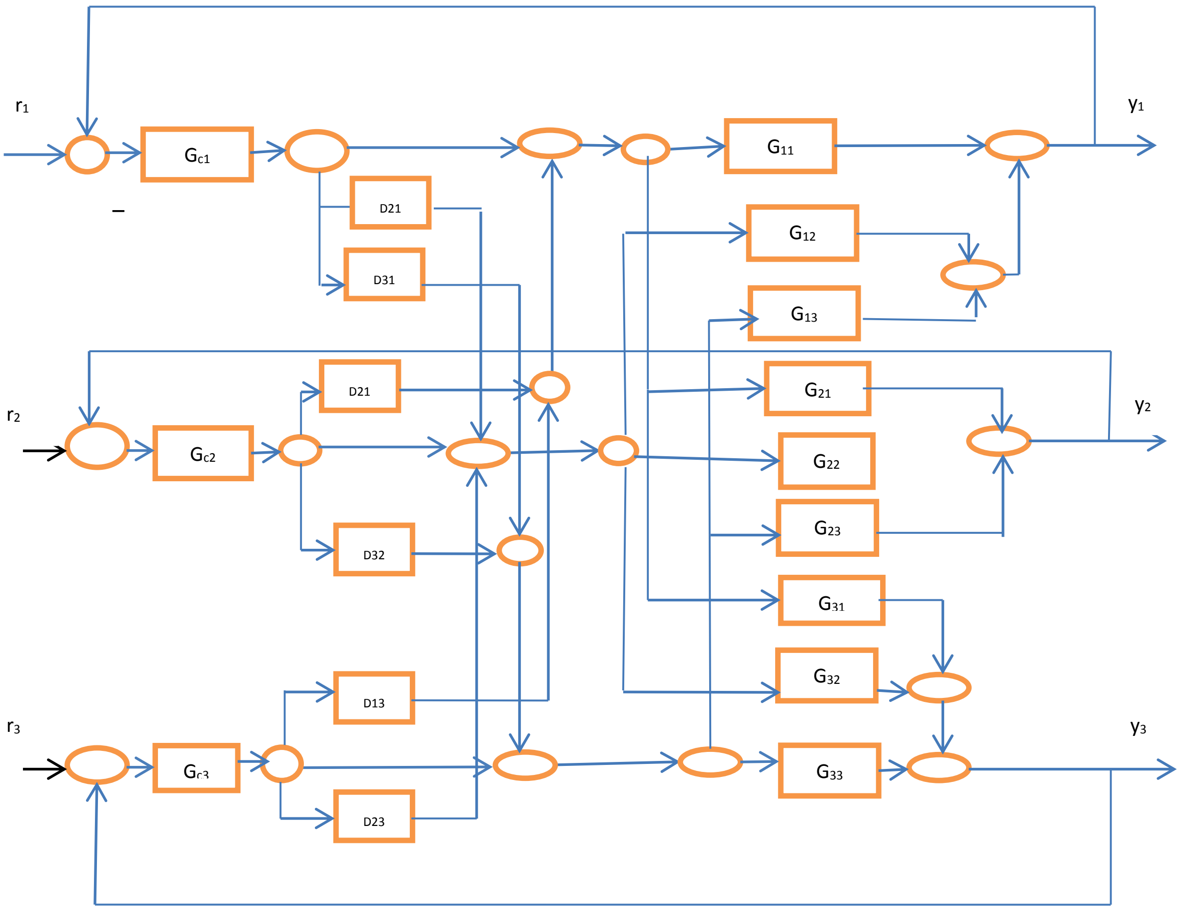

Figure 3: Block diagram of plant with decoupler

In dynamic decoupling design, a compensator matrix was found such that the resulting plant will have zero off-diagonal elements.

Dynamic Decoupler was calculated using Chau (2001) decoupler relations as:

$$\begin{bmatrix} C_1 \\ C_2 \\ C_3 \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} & G_{13} \\ G_{21} & G_{22} & G_{23} \\ G_{31} & G_{32} & G_{33} \end{bmatrix} \begin{bmatrix} G_{c1} & 0 & 0 \\ 0 & G_{c2} & 0 \\ 0 & 0 & G_{c3} \end{bmatrix} \begin{bmatrix} R_1 - C_1 \\ R_2 - C_2 \\ R_3 - C_3 \end{bmatrix} \tag{4}$$

However, in a system with interactions, off diagonal matrix of process transfer matrix are not zero, we therefore need to manipulate the system so that it can be decoupled, we introduced decoupling function and the manipulated variables such that:

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$$\begin{bmatrix} M_1 \\ M_2 \\ M_3 \end{bmatrix} = \begin{bmatrix} d_{11} & d_{12} & d_{13} \\ d_{21} & d_{22} & d_{23} \\ d_{31} & d_{32} & d_{33} \end{bmatrix} \begin{bmatrix} G_{c1} & 0 & 0 \\ 0 & G_{c2} & 0 \\ 0 & 0 & G_{c3} \end{bmatrix} \begin{bmatrix} R_1 - C_1 \\ R_2 - C_2 \\ R_3 - C_3 \end{bmatrix} \quad (5)$$

And the system equation is:

$$\begin{bmatrix} C_1 \\ C_2 \\ C_3 \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} & G_{13} \\ G_{21} & G_{22} & G_{23} \\ G_{31} & G_{32} & G_{33} \end{bmatrix} \begin{bmatrix} d_{11} & d_{12} & d_{13} \\ d_{21} & d_{22} & d_{23} \\ d_{31} & d_{32} & d_{33} \end{bmatrix} \begin{bmatrix} G_{c1} & 0 & 0 \\ 0 & G_{c2} & 0 \\ 0 & 0 & G_{c3} \end{bmatrix} \begin{bmatrix} R_1 - C_1 \\ R_2 - C_2 \\ R_3 - C_3 \end{bmatrix} = \boldsymbol{GdG_c} \begin{bmatrix} R_1 - C_1 \\ R_2 - C_2 \\ R_3 - C_3 \end{bmatrix} \quad (6)$$

In order to decouple this, $\boldsymbol{GdG_c}$ should be a diagonal matrix. Hence,

$$\boldsymbol{G_0} = \boldsymbol{GdG_c} \quad (7)$$

Hence, **C** will be $\begin{bmatrix} C_1 \\ C_2 \\ C_3 \end{bmatrix} = [I + G_0]^{-1} G_0 \begin{bmatrix} R_1 \\ R_2 \\ R_3 \end{bmatrix}$ (8)

In order to find **D,** it is required that matrix **GD** be diagonal since $\boldsymbol{G_c}$ is diagonal, therefore **Gd=H**

i.e.

$$\begin{bmatrix} H_1 & 0 & 0 \\ 0 & H_2 & 0 \\ 0 & 0 & H_3 \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} & G_{13} \\ G_{21} & G_{22} & G_{23} \\ G_{31} & G_{32} & G_{33} \end{bmatrix} \begin{bmatrix} d_{11} & d_{12} & d_{13} \\ d_{21} & d_{22} & d_{23} \\ d_{31} & d_{32} & d_{33} \end{bmatrix} \quad (9)$$

Where **G** is the process transfer matrix
H is the decoupled plant which should be a diagonal matrix and
d is the compensator which serves as the dynamic decoupler
With little algebraic manipulation, the followings are arrived at:

$$D_{11} = D_{22} = D_{33} = 1$$

$$D_{12} = \frac{(G_{13}G_{32}) - (G_{12}G_{33})}{(G_{11}G_{33}) - (G_{13}G_{31})} \quad (10)$$

$$D_{13} = \frac{(G_{12}G_{23}) - (G_{13}G_{22})}{(G_{11}G_{22}) - (G_{12}G_{21})} \quad (11)$$

$$D_{21} = \frac{(G_{23}G_{31}) - (G_{21}G_{33})}{(G_{33}G_{22}) - (G_{23}G_{32})} \quad (12)$$

$$D_{23} = \frac{(G_{13}G_{21}) - (G_{23}G_{11})}{(G_{11}G_{22}) - (G_{12}G_{21})} \quad (13)$$

$$D_{31} = \frac{(G_{22}G_{31}) - (G_{21}G_{32})}{(G_{32}G_{23}) - (G_{22}G_{33})} \quad (14)$$

$$D_{32} = \frac{(G_{11}G_{32}) - (G_{12}G_{31})}{(G_{13}G_{31}) - (G_{11}G_{33})} \quad (15)$$

Static Decoupling technique

For static decoupling technique, inverse of process gain array was used for decoupling (D).

i. e. $D = inv(\boldsymbol{K})$ (16)

where K= process gain

PI/PID controller Design

The control technique for a PI controller is given as:

$$u(s) = K_p + \frac{K_p}{\tau_I s} \quad (17)$$

where K_p= process gain
τ_I = time constant

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PI- controller was tuned using Ziegler-Nichols, Tyreus-Luyben and MATLAB tuning methods. Ziegler-Nichols and Tyreus-Luyben settings are shown in Table 4 while MATLAB pidtuning algorithm for tuning PID controllers is an automatic tuning method which chooses a crossover frequency based on the plant dynamics and designs for a target phase margin of 60°C. pidtune tunes the parameters of the controller for robustness and response time performance of the system. For multi-input multi-output (MIMO) system, Pidtune designs the controller for each linear model which then returns an array for PID controllers (The Mathwork Inc., 2012). PID tuning helps to attain the closed loop stability of the plant which makes the system to be bounded for bounded inputs. It also helps the closed loop system to track set point and suppress disturbances as fast as possible. Tuning PID also allows loop design to have enough phase and gain margins which allows modeling errors in the dynamics of the system.

Tuning method	**K_c**	**τ_I**
Ziegler-Nichols	$0.455K_{cu}$	$P_u/1.2$
Tyreus-Luyben	$K_{cu}/3.2$	$2.2P_u$

Table 4: Tunings considered for the PI Controller parameter

Gain Scheduling

Gain scheduling is a form of advanced feedback control system in which the feedback gains are adjusted with their feedforward compensation (Krishna *et al.*, 2012). It is an approach that can be used to measure the gain and then changes or schedule the controller to compensate for changes in the process gain (Astrom and Wittenmark, 2008). Hence, it is a very useful technique in reducing the effect of parameter variation and to compensate for known nonlinearity in a process (Liptak, 2006). Figure 4 shows configuration of gain scheduling system.

The controller parameter changes quickly in response to change in the process. The best controller parameters for each case in the system were determined as a function of gain scheduling variable. These parameters are shown in Table 5. In this work, an interval is defined around each gain scheduling value in the parameter table. As the gains scheduling variable changes from one interval to another, the controller parameters are changed, a set of controller and decoupler were developed for each set of process gains. The controllers and decouplers were scheduled which change with change in process gains (**ε_1 ε_2 ε_3**) of the plant.

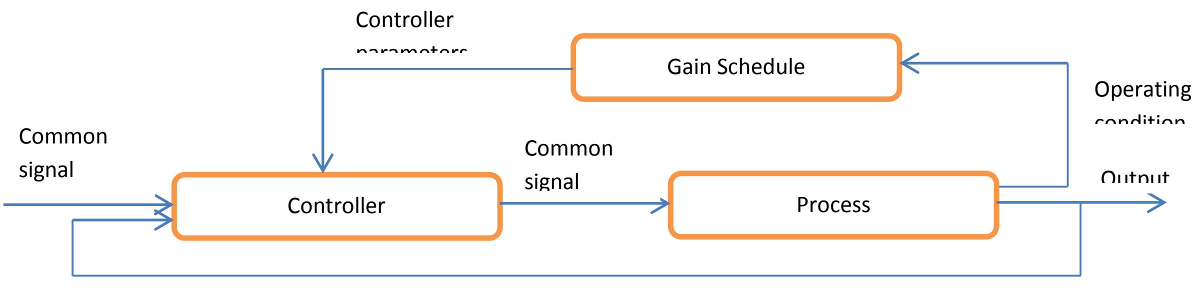

Figure 4: Block diagram of the gain scheduling controller

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Cases (ε_1 ε_2 ε_3)	**Loop 1**		**Loop 2**		**Loop 3**	
	K_p	**τ_I**	**K_p**	**τ_I**	**K_p**	**τ_I**
(0 0 0)	0.139	0.00381	0.291	0.00616	0.0657	0.0117
(-1 -1 -1)	0.290	0.00796	0.323	0.00684	0.0806	0.0144
(1 -1 1)	0.0912	0.00251	0.323	0.00684	0.0555	0.0099
(1 1 1)	0.0912	0.00251	0.264	0.0056	0.0555	0.0099
(-1 1 0)	0.290	0.00796	0.264	0.0056	0.0657	0.0117

Table 5: Optimum values for K_p and τ_I in the three loops

RESULTS/DISCUSSION

Relative gain array (RGA) of the plant was computed as:

$$RGA = \begin{bmatrix} 2.0757 & -0.7289 & -0.3468 \\ 3.4242 & 0.9343 & -3.3585 \\ -4.4999 & 0.7946 & 4.7053 \end{bmatrix} \tag{18}$$

Elements (1,1) and (3,3) which are greater than one indicates that there is strong loop interaction between the loop corresponding to each of these elements and other loops. Also, element (2,2) which is very close to one indicates that there is no much interaction of other loops on loop 2. Figure 5 and 6 show the open loop responses when the dynamic and static decouplers are respectively used to decouple the system. Effects of interactions on each loop were reduced to zero for both decouplers which means that both are able to decouple the plant. Table 6 shows the settling times for main loops and interacting loops of both decouplers. Except the interaction loop (side draw-reflux temperature) with settling time 604secs which is unnecessarily high, dynamic decoupler is faster in removing the interactions as indicated by low settling times of the interaction loop responses recorded for it when compared to those recorded for static decoupler. However, the dynamic decoupler slows down the main loop response as indicated by the large values of settling times recorded for them.

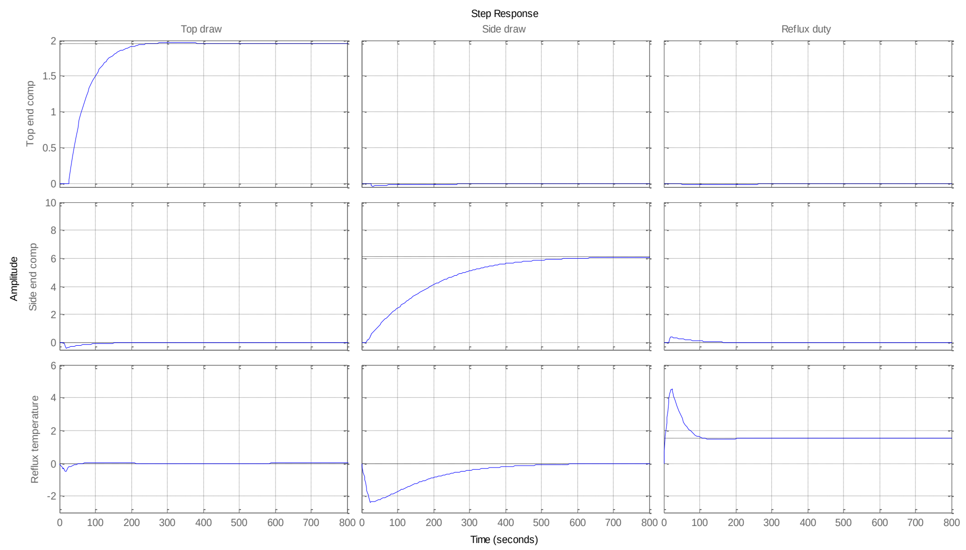

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Figure 5: Dynamic Decoupler of plant

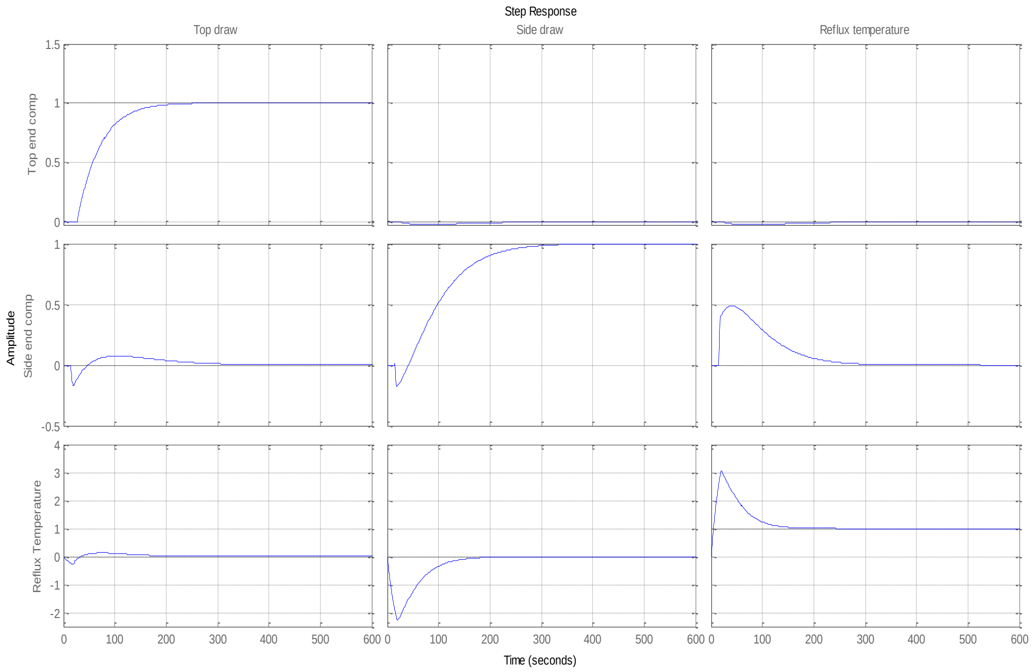

Figure 6: Static Decoupler of plant

	Settling time for Dynamic Decoupler(sec)			Settling time for Static Decoupler (sec)		
	Top draw	**Side draw**	**Reflux duty**	**Top draw**	**Side draw**	**Reflux duty**
Top end composition	201	259	327	185	399	399
Side end composition	216	595	249	376	283	286
Reflux temperature	178	604	101	270	176	151

Table 6: Settling time of decoupler system

A PI controller was designed for a dynamic decoupled plant and tuned using Ziegler-Nichols, Tyreus-Luyben and MATLAB pidtune. The result of the closed loop response of the system is shown in Figure 7. The response of the closed loop system for a dynamic decoupled plant in top end composition output shown in Figure 7a shows that Tyreus-Luyben gave sluggish response as it is unable to reach 0.5set point in time 1000sec but Ziegler-Nichols and MATLAB pidtune showed damped response and were able to reach the 1000 set point with Ziegler-Nichols gave a settling time of 461sec while MATLAB pidtune gave a settling time of 364sec. Response of the closed loop system for a dynamic decoupled plant in side end composition output shown in Figure 7b shows that Ziegler-Nichols and MATLAB pidtune shows underdamped as indicated with the presence of overshoot, however, settling time of Ziegler-Nichols gave 366sec and MATLAB pidtune shows 522 sec. Response of the closed loop system for the dynamic decoupled plant in bottom reflux temperature output is shown in Figure 7c, response of the closed loop system with Ziegler-Nichols, Tyreus-Luyben and MATLAB tuning gave underdamped response as indicated with the presence of overshoot, however, MATLAB pidtune gave a fastest response with settling time of 408sec. MATLAB pidtune gave the fastest settling time in the three outputs.

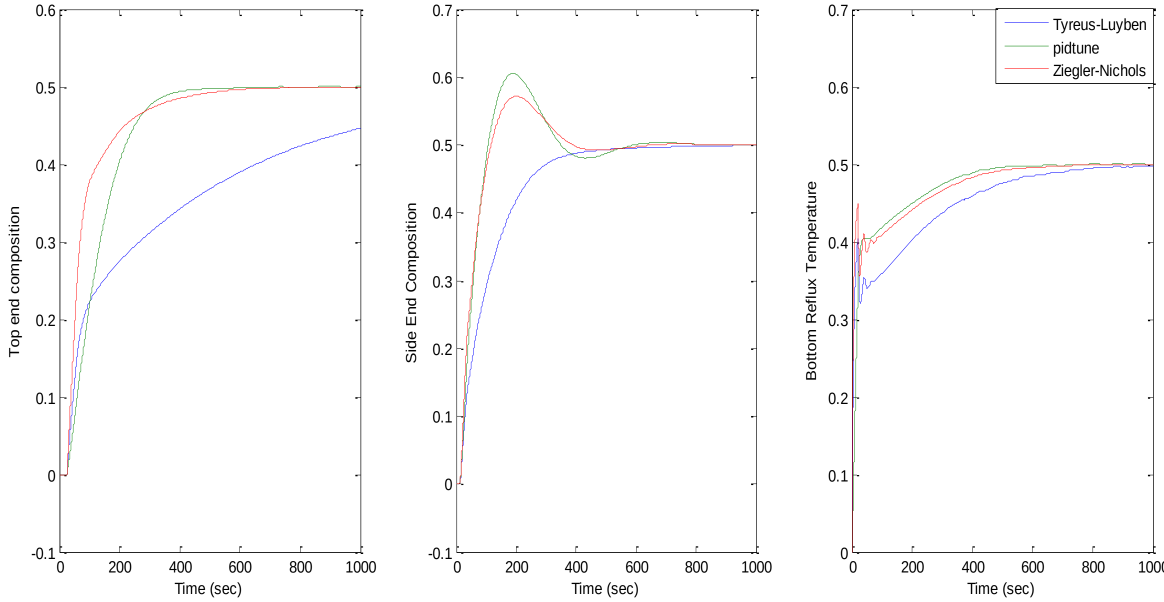

Figure 7: Clos p response of the PI co r for a dynamic decoupl h different tuning

The PI controller was designed for a static decoupled plant and was tuned using Tyreus-Luyben, Ziegler-Nichols and MATLAB pidtune. The response of the closed loop system is shown in Figure 8. Response of the closed loop system for top end composition output is shown in Figure 8a, controller setting with the three tuning methods show sluggish response and were unable to reach the 0.5setpoint at 1000sec, however, Tyreus-Luyben shows the slowest response with settling time 1750sec which shows that set-point was not achieved within the stipulated 1000sec. Closed loop response of the side end composition output is shown in Figure 8b, controller settings for Ziegler-Nichols and MATLAB tune show damped response and were able to reach the 0.5setpoint within 1000sec, however, MATLAB tuning method shows a faster response at settling time of 447sec than Ziegler-Nichols which has the settling time of 569sec. Response of the closed loop system with MATLAB pidtune gave the fastest settling time and best controller performance in the three outputs. Hence, response of the closed loop system with MATLAB pidtune is for a static decoupled plant is compared with response from a dynamic decoupled plant and the result is shown in Figure 9.

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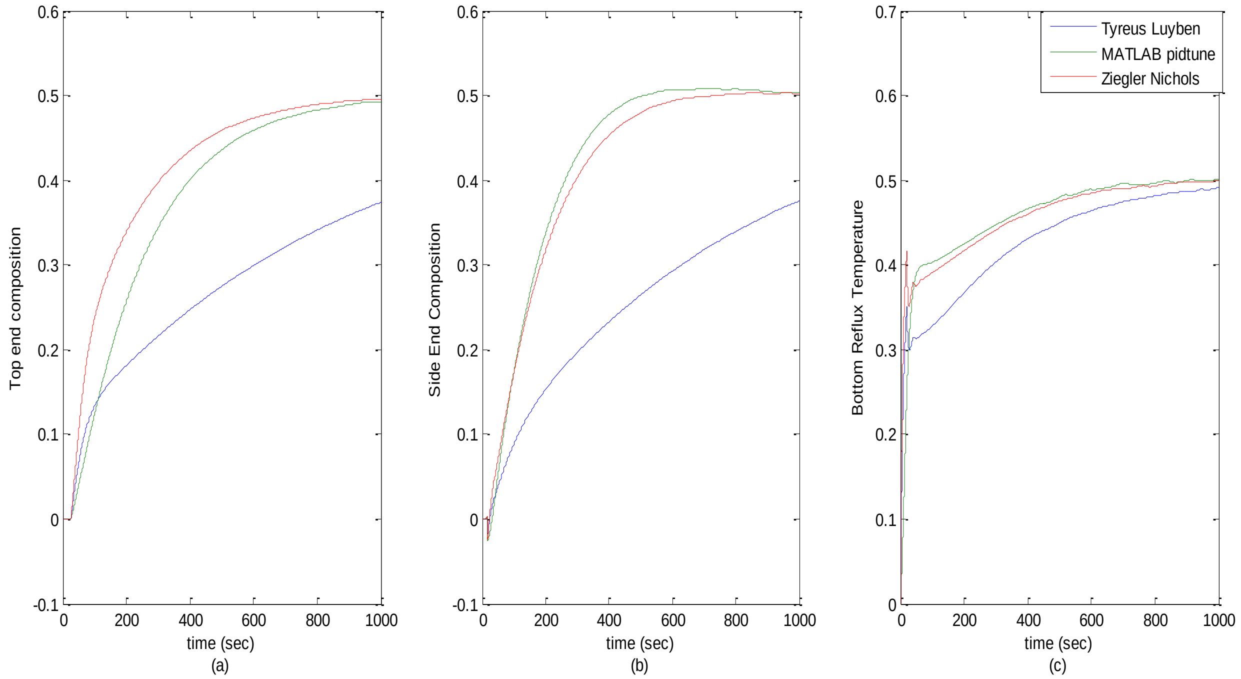

Figure 8: Closed loop response of a static decoupled plant with different tuning methods.

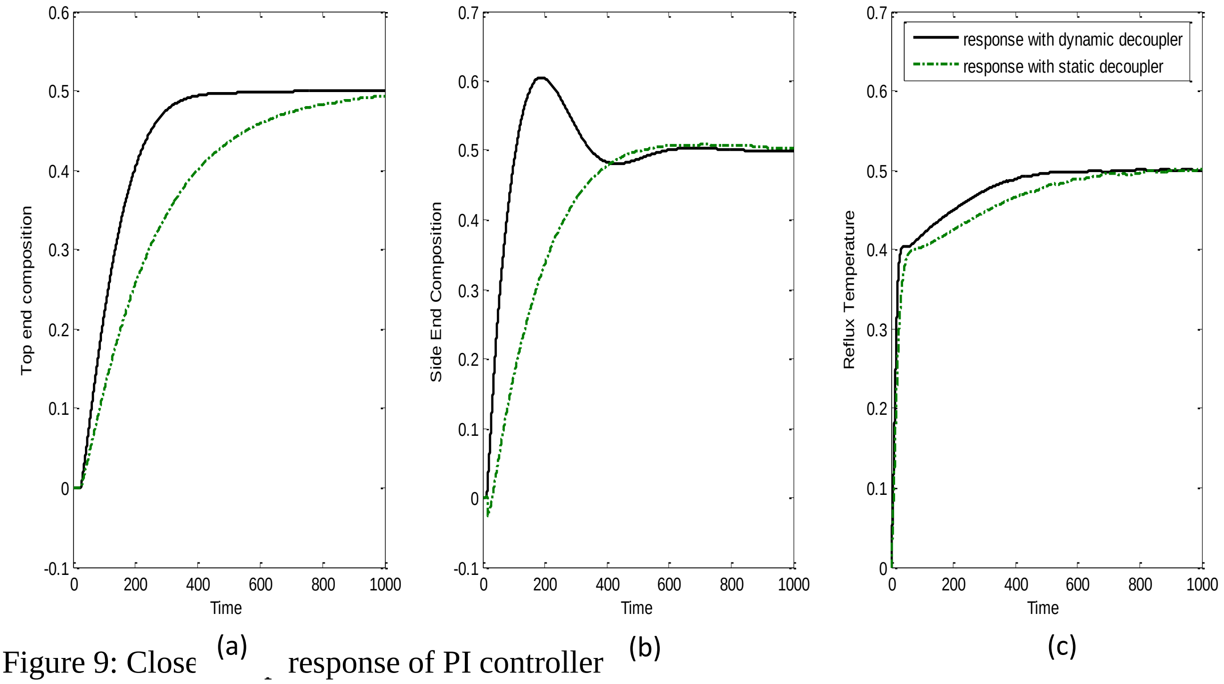

Figure 9: Close response of PI controller

Figure 9 shows the comparison of closed loop response of the dynamic decoupled plant under

PI control with static decoupled under PI control at 0.5 step change in set point. The response of the closed loop system for both controllers in top end composition output is shown in Figure 9a, both decouplers gave overdamped response with static decoupler showing more overdamped response as indicated by sluggish response. The response of the closed loop system in respect of side end composition is shown in Figure 9b, the response of both

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decouplers gave underdamped response as indicated by presence of overshoot. No overshoot is observed in response of the closed loop system in bottom reflux temperature output (Figure 9c). Thus, it can be said that both controllers are able to track set points with the controller settings. The set point tracking for both decouplers is shown in Figure 10.

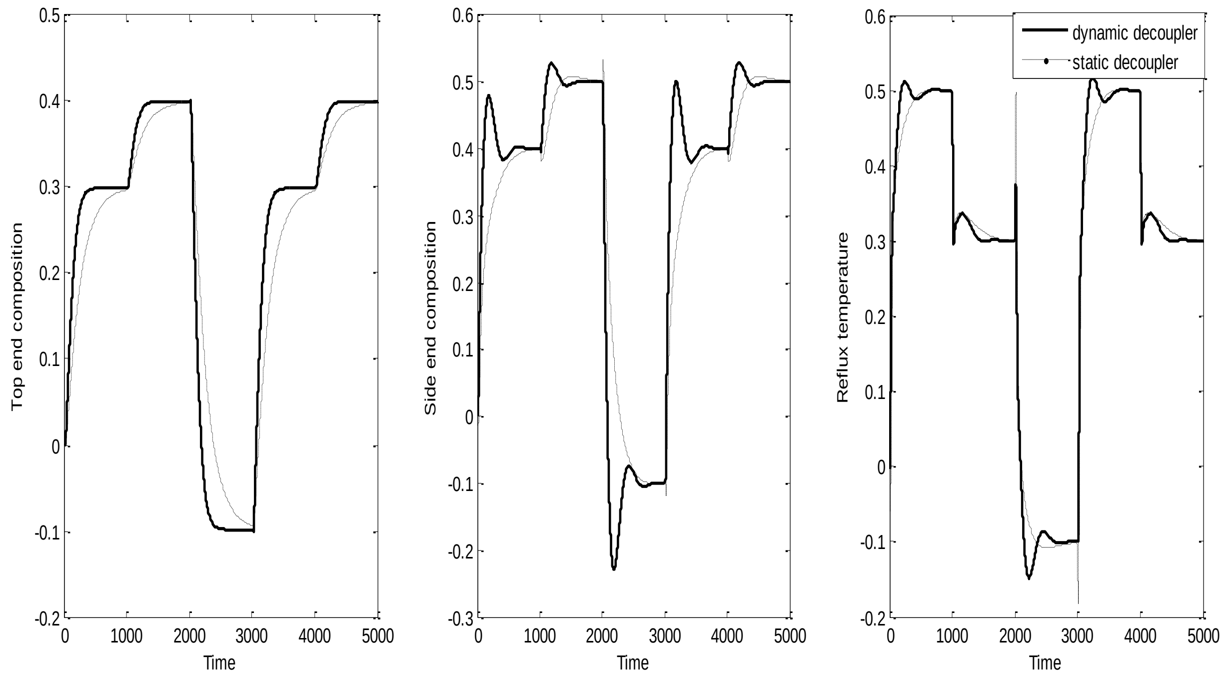

Figure 10: Set point tracking of both decouplers

In Figure 10, a step change of [0.3 0.4 0.5 0.5 0.3 0.4] both decouplers track set point, however, in all the three loops, dynamic decoupler is faster in tracking set point than the static decoupler before proceeding to the next step. It can thus be said that closed loop system with dynamic decoupler is faster in tracking set point than closed loop system with static decoupler. However, in order to improve the controller performance, gain scheduling is used for robust controller performance when there are variations. The gain scheduling controllers was simulated at different points and for different cases of variations using Table 5 and the results is shown in Tables 7, 8 and 9.

Loop 1	**Gain scheduling controller**					**PI controller**				
	Case 1	**Case 2**	**Case 3**	**Case 4**	**Case 5**	**Case1**	**Case 2**	**Case 3**	**Case 4**	**Case 5**
Rise Time	207.64	87.02	401.47	261.16	80.00	207.64	435.84	60.68	76.92	411.16
Settling Time	364.00	467.4	758.33	496.28	495.24	364.0	1.08e+3	2.42e+3	1.20e+3	1.09e+3
Overshoot	0	14.49	0	0	23.40	0	6.38	60.64	21.56	6.44

Table 7: Performance metrics of gain scheduling controller with PI controller in Top end composition output (Loop 1)

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The performance metrics of gain scheduling controller and PI controller on loop 1 (top end composition/top draw) shows that the results for the nominal plant (case 1) does not change after the plant is being scheduled which shows that there is no parameter variation in the nominal plant. The settling time for the conventional PI controller on loop 1 is higher than the gain scheduling controller where the nominal case has the lowest settling time. Furthermore, the gain scheduling controller has less overshoot compared to the conventional PI controller where only the nominal plant does not have overshoot. Cases 2 and 3 have lower rise time for the conventional PI controller which is vice-versa for the gain scheduled plant. This is due to variation in the process. It can thus be said a gain scheduled controller was able to use parameter adjustment mechanism to reduce the variation in the process. This performance metrics was considered for the side end composition loop as shown in Table 7.

Loop 2	Gain scheduling controller					PI controller				
	Case 1	Case 2	Case 3	Case 4	Case 5	Case 1	Case 2	Case 3	Case 4	Case 5
Rise Time	70.98	80.60	62.64	74.08	100.51	70.98	256.99	71.3856	55.73	272.94
Settling Time	522.52	320.31	508.45	566.81	359.03	522.52	836.97	2.67e+3	1.60e+3	863.62
Overshoot	20.98	7.84	22.26	27.09	7.06	20.98	6.02	66.25	59.25	4.45

Table 8: Performance metrics of gain scheduling controller with PI controller in side end composition output (Loop 2)

The result from Table 8 shows that the rise time, settling time and overshoot for a gain scheduled controller is lesser than the conventional PI controller except in nominal case 1 which there is no difference. However, cases 2, 3 and 5 have lower settling time than the nominal plant (case 1) for the gain scheduled plant which shows that case 2 has the lowest settling time. Also, case 3 has a lower rise time than the nominal case for the gain scheduled plant. It can thus be concluded that a gain scheduled controller gave a more effective performance than a PI controller.

Loop 3	Gain scheduling controller					PI controller				
	Case1	Case 2	Case 3	Case 4	Case 5	Case1	Case 2	Case 3	Case 4	Case 5
Rise Time	200.0	94.97	43.85	220.08	88.80	200.0	31.34	287.58	258.48	25.87
Settling Time	408.1	219.7	520.68	433.73	202.24	408.1	1.08e+3	2.40e+03	1.36e+03	1.0971e+3
Overshoot	0	0.17	0	5.6e-4	0.30	0	46.37	26.7146	13.14	40.80

Table 9: Performance metrics of gain scheduling controller with PI controller in Bottom Reflux Temperature output (Loop 3)

Table 9 shows the performance metrics for loop 3 (Bottom Reflux Temperature/Bottom Reflux Duty) and the result shows that the plant has a lower settling time and overshoot for a gain

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scheduled plant. However, case 5 has the lowest settling time while case 2 has the highest rise time. It can thus be concluded that gain scheduled controller is more effective than a PI controller.

IMPLICATION TO RESEARCH AND PRACTICE

The study has shown that for a multivariable system where there is coupling effect, dynamic and static decoupling techniques are methods which can be used to remove these coupling effects. A Proportional Integral controller is able to control the plant, however, parameter variations in the process is controlled using gain scheduling controller. The performance metrics (settling time, rise time and overshoot) show a more effective controller performance for a gain scheduling controller than a PI/PID Controller.

CONCLUSION

Relative Gain Array was used to check for the level of interaction which is significant in the plant. Dynamic and static decouplers were used to remove these coupling effects, however, dynamic decoupler decouples faster but slow in open loop response for the nominal plant but they were sluggish for different variations in the process. A Proportional Integral (PI) controller was used in controlling the decoupled plant with Tyreus-Luyben settings, Ziegler-Nichols settings and MATLAB pidtune, the result shows that closed loop response a dynamic decoupled track set point faster than the closed loop response of the static decoupled plant. Also, MATLAB pidtune showed a better controller performance than Ziegler-Nichols and Tyreus-Luyben settings.

There were slow responses for the controllers in different cases of gain variations incorporated in the plant which shows the need for gain scheduling for robust controller performance. Gain scheduling was able to cater for parameter variations in the process and also shows a faster rise time and lower settling time with little overshoot. This can thus be said that gain scheduling is more efficient than a conventional PI/PID controller.

FUTURE RESEARCH

Gain scheduling mechanism uses an adjustment mechanism that is pre-computed offline but no feedback is provided to compensate for incorrect schedules, therefore, a direct adaptive controller is recommended for further research.

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