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MULTI-LOOP CONTROL DESIGN BASED ON STEADY STATE SENSITIVITY ANALYSIS

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ABSTRACT

This paper proposes multi-loop control for multivariable distillation processes using analytical design method, based on steady state sensitivity analysis. The design method is aimed to select the best controller pairings for multi-input multi-output (MIMO) processes containing integrators process. Essentially, for keeping the purity of benzene in the top product and impurity in the bottom: under acceptable specification, eliminating the process interaction, and operate the column in safety condition. This method requires step change in the manipulated variables of open loops. The steady state sensitivity of controlled variables to changes in the manipulated variables is analyzed using the relative gains array (RGA). The multi-loops were tuned independently based on the process distinctive behaviours using Ziegler-Nichols and modified Ziegler-Nichols rules to determine the controller gains. The candidate input-output parings are identified and then investigated using simulation and dynamic analysis. The results demonstrate that LV-configuration provided the superior robust performance with fast and well-balanced closed-loop stability for large load disturbances in feed flow rate and feed composition.

Keywords: (Steady state sensitivity analysis, distillation column, control design, relative gains array)

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تصميم التحكم متعدد الدوائر على أساس تحليل حساسية الحالة المستقرة

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ملخص البحث:

تقترح هذه الورقة البحثية نظام تحكم متعدد الدوائر لعمليات التقطير متعددة المتغيرات باستخدام أسلوب التصميم التحليلي، المبني على تحليل الحساسية للحالة المستقرة. تهدف طريقة التصميم هده إلى اختيار أفضل اقتران لعوامل التحكم لعمليات متعددة المدخلات والمخرجات (MIMO)التي تحتوي على عمليات تراكمية. ويهدف هذا الأسلوب أساساً إلى الحفاظ على نقاء البنزين في المنتج العلوي والشوائب في المنتج السفلي: ضمن المواصفات المقبولة، مع القضاء على تداخل البيني في العملية، وتشغيل البرج في ظروف أمانة. تتطلب هذه الطريقة تغييراً تدريجياً في المتغيرات المُتحكّم بها لدوائر المفتوحة. ومنها يتم تحليل حساسية الحالة المستقرة للمتغيرات المُتحكّم للتغيرات الحاصلة في المتغيرات المُتحكّم بها باستخدام مصفوفة الاوزان النسبية .(RGA) وقد تم ضبط الدوائر المتعددة بشكل مستقل بناءً على السلوكيات المميزة للعملية باستخدام قواعد زيجلر-نيكولز وقواعد زيجلر-نيكولز المحسنة لتحديد اوزان المسيطرات. ويتم تحديد اقتران المدخلات والمخرجات المرشحة، ومن ثم يتم التحقق منها باستخدام المحاكاة والتحليل الديناميكي. توضح النتائج أن التكوين (LV) يوفر أداءً صارم متفوقاً مع استقرار سريع ومتوازن للدائرة المغلقة عندما تكون الاضطرابات الكبيرة في معدل تدفق للتغذية ومكونات تيار التغذية

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1. INTRODUCTION

Most of Chemical processes and plants are essentially multivariable and nonlinear systems with multi-loop interactions, making the controller design more complicated. This issue poses numerous challenges in control tasks and attracted many researchers in this area. The selection of an appropriate control structure for unit operation with multi-loops plays the most important role in control system design[1]. Dealing with this problem, a decentralized control is the most broadly used in process industrial[2]. The decentralized control has several advantages such as easy understanding for operators, simple implementation and easy to maintain. While, the centralized control of such large-scale process is often expensive, computational complexity, and time consuming. Hence, a multivariable system is decomposed into a number of independent subsystems. Each output variable is coupled to a single input variable based on dynamical behavior and subsystems interactions[3].

In order to design a successful decentralized control system with excellent performance, two steps are required: control configuration selection and controller tunings[2, 4]. The first step is the main concern of this study; involve the good control configuration selection able to eliminate process interactions and meet control objectives. This is achieved through steady state sensitivity analysis of the system which plays the most dominant role to exam the steady-state variations of all system parameters when deviations or disturbances occur in the process. It is also used to highlight the control structure design and determine the most effective pairings through the relative gains array (RGA). The RGA is a successful systematic approach used to measure interactions and determine the best pairing of controlled and manipulated variables based on steady-state gain information[5, 6]. It is the most commonly used tool and depends only on the plant model. However, the control configuration selection of the process addresses RGA limitation. It has more insightful effect on control performance than whether advanced or conventional control method is applied[7, 8].

The second step is the controller tunings. A well-tuned controllers is a key factor for improving control performance and achieving control objectives[1]. Many tuning-methods have been addressed in literatures, providing controller tuning gains based on the control tasks or control problem solving. Each method has its own special rules for providing acceptable robustness and performance of closed loop and the correct choice method tuning depends on the control objectives of the processes. In practice, disturbance rejection and set-point tracking are the mostly main concerns[9].

In spite of the availability of multi-input multi-output (MIMO) advanced controllers techniques, the decentralized PID/PI controllers using single-input single-output (SISO), designed separately remain the standard method to control MIMO systems with interactions [5, 10]. It is considered a benchmark for many other complex controller designs and a measure for new ideas in control. This is due to its

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simplicity, easy understanding, simple tuning procedure and perform well, dealing with modeling uncertainties and valve and sensor failure [11, 12]. However, there are several simulation examples of multi-loop PI controller design based on RGA design method and some were addressed in[12, 13].

Case study for this investigation is a multivariable distillation column for aromatic components. Control of distillation column is necessary to assure both the quality of the final product as well as the safe operation of the plant. The difficulties in controlling distillation columns lie in their: highly nonlinear characteristics, multiple inputs and multiple outputs (MIMO) structure, and the presence of severe disturbances during operation[8, 14]. The nonlinearity of distillation columns is well known and clearly increases with high purity products in the system.

The objective of this paper is to design a robust and efficient control structure for large scale processes, contained self-regulating loops and integrating loops able to reject any disturbances or deviations in the process and achieve the desired closed-loop response. The designed decentralized control is investigated using simulation and dynamic analysis of the rigorous model which is the most reliable method for evaluating a control structure[1]. The model is simulated in MATLAB® programming environment.

2. Case Study

A distillation column is a typical MIMO system, with strong interactions between the variables. However, the interactions occurring between the inputs and the outputs are difficult to identify. The disturbances to a distillation column can come from many sources such as the feed (feed flow rate, feed composition, feed temperature, feed vapour mole fraction), the pressure inside the column, and the cooling water etc.Multivariable distillation column of a BTX (benzene, toluene, and xylene) was selected as a case study. The schematic diagram of the process is shown in Figure 1. The feed entered the column as a single feed stream and fed as saturated liquid (at bubble point) with molar flow rate (8 kmol) containing 45 % benzene and 20% toluene. It is separated into an overhead product of specified benzene content in excess of 98.53 % and a bottoms product containing less than 3.37 % benzene. The overhead vapour is totally condensed in the condenser and flows into the reflux drum. The contents of the drum are assumed to be perfectly mixed with uniform composition. The liquid product is removed from last tray and the vapour boil-up is generated by reboiler.

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According to Al-Shatri and his co-authors [15], the derived mathematical model of this system was based on nonlinear model. This mathematical model applies the mass and energy balance equations, laws of basic thermodynamics, and algebraic energy equations supported by vapour liquid equilibrium and physical properties to relate all of these variables with each other.

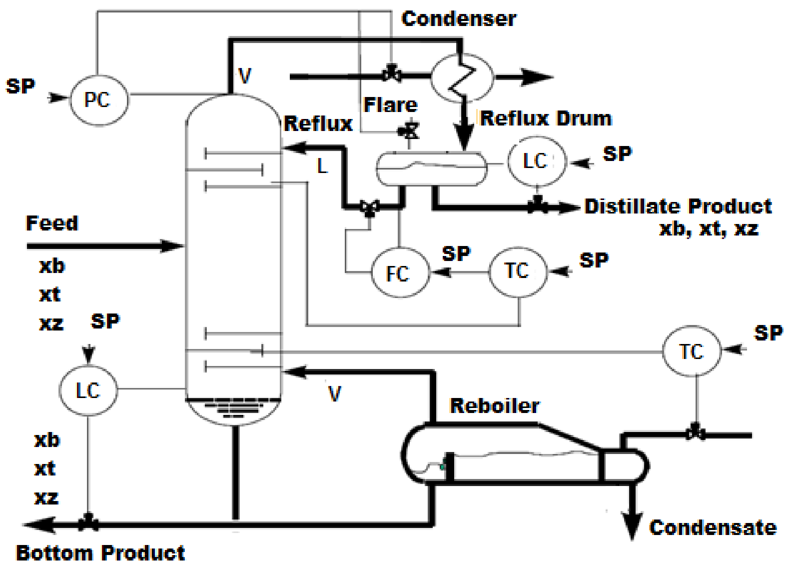

***Figure1*: Case study: Multivariable distillation column**

However, the model has 40 state variables which are represented by a set of nonlinear differential equations that describe the relationship between the various variables of the column.

3. Control configuration selection

A control scheme essentially depends on the designed process goals: ensuring the operation does not exceed its process constrains, keeping process stability, able to suppress and compensate the disturbances, able to alleviate the consequences of designed errors, and optimizes the operation in order to reduce cost and enhance product quality[5].

Achieving this target and find the candidate control strategy, the steady-state sensitivity analysis is used to assess the steady-state variations of all system parameters when deviations or disturbances occur in the process. This procedure reveals the limitations of the process from undesired dynamic behavior, which may drive the process out of its target or require large changes in manipulated variables when disturbances occur.

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There are a number of control configurations used in separation process such as D-B, D-V, L-B, L-V, and (L/D) (V/B) configurations[16]. Selection of the best control configuration is required considering the operating conditions and dynamic modeling of the column. The numbers of independent process variables that can be manipulated play key roles in the control system design. In order to determine the candidate control structures as sets of input and output variables, Seborg *et al.*[5] presented guidelines for selection of the process variables for the purposes of control design. These process variables are classified into input variables and output variables. The input variables represented the manipulated variables that can be adjusted or disturbances that came from the external environmental change. The output variables are all variables associated with exit streams and process condition (composition, temperature, or level) and the important output variables are measurable. The designed multivariable distillation column by [15], is the best example to test high interaction nonlinear system and the sensitivity analysis for the output behavior. Based on the derived modeling of the column, the feed flow rate and feed composition (F, x_f) act as disturbances, the top and bottom product flow rate (D, B), reflux flow rate (L/D), vapor boil up rate (V), and condenser duty (Q_C) act as manipulated variables. Furthermore, bottom and distillate composition, and reflux drum and reboiler level act as controlled variables. The column pressure is assumed constant and trays temperature and composition act as uncontrolled variables.

The steady-state sensitivity analysis is used to screen out poor control structures from further study in more complex processes. The input–output structure bridged the gap between steady state process design and control system synthesis and enhance pairing selection[17]. Step changes in the manipulated variables (the internal and external molar flowrate) of the open-loops were introduced individually. Hence, the process steady state eventually is moved to a new steady state. Consequently, the best controller pairings of multi-loops processes can be found through relative gains array (RGA) as computed from the steady-state gains of the process. This is a more successful approach in selecting the best controller pairings (appropriate control configuration) and very useful to evaluate expected control behaviour [18]. As a result, the case study has twenty-four possible pairings for controlling the top and bottom product composition and reflux drum and reboiler level. Thus, RGA will provide the systematic analysis of alternative pairings of input variables selected from the available manipulated variables for the output variables desired to be controlled.

The second step is controller tunings of the successfully decentralized control system with excellent closed loop performance. Although, advance process controls techniques exist, the PID control algorithm is still extensively used in process industries due to its robustness, easy to understand, and very well in practical application [19, 20]. The PID controller has three main control actions. The proportional (P) term calculates a change in the input proportional to the error between set-point and measurement value. It is more important in initial response of

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control loop. The integral (I) term is called the reset time that adjusts a change in the manipulated variable until error approaches zero and the process steady state offset is eliminated. The derivative (D) term is often used to speed up the response of the process, but sometimes may cause unwanted large changes in the controlled variable. For this reason, a PI controller represents most PID type used in the industrial processes and called decentralized PI. Hence, PID controller is a tool that takes the present, past, and future of the error into consideration as illustrated in Equation (1)

$$u(t) = K_c\left(e(\mathrm{t}) + \frac{1}{T_i}\int_0^t e(\lambda)d\lambda + T_d\frac{de}{dt}\right) \qquad (1)$$

where K_c is the controller gain which acts on the present value of the error. T_i is the integral time constant that represents an average of past errors, and T_d is the derivative time constant which is given a prediction of future errors. Here the digital PID algorithm was used in velocity form as in Equation (2) [21, 22].

$$\mathrm{V}MV_N = K_c\left((e_N - e_{N-1}) + \frac{\mathrm{V}t}{T_i}e_N - \frac{T_d}{\mathrm{V}t}(CV_N - 2CV_{N-1} + CV_{N-2})\right) \qquad (2)$$

$$MV_N = MV_{N-1} + \mathrm{V}MV_N \qquad (3)$$

where CV_N, MV_N, and SP_N are represent the current values of the controlled variable, manipulated variable, set point at the current sample N, with the current error $e_N = SP - CV_N$.

However, the optimum values for the required control response are obtained through a control loop tuning either based on open loop or closed loop method to adjust the control parameters gains. There are many different ways to determine controller parameters such as; Process Reaction Curve, Ziegler-Nichols, Cohen-Coon, and Internal Model Control. In this study, we present the Ziegler-Nichols method based on open loop method, the most reliable and notable tuning rules. A few tuning constants are required to find such as; process gain (K), time constant (τ_c), and time delay (θ) through applying a step change input to the process and control parameters are calculated from Equations (4)-(6);

For P control:

$$K_c = \left(\tau_c / (K * \theta)\right) \qquad (4)$$

For PI control:

$$K_c = 0.9\left(\tau_c / (K * \theta)\right) \; ; \; T_i = 3.33 * \theta \qquad (5)$$

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For PID control:

$$K_c = 1.2\left(\tau_c / (K * \theta)\right) \; ; \; T_i = 2 * \theta \; ; \; T_d = 0.5 * \theta \quad (6)$$

The integrating process response has first-order plus time delay which is different from a non-regulating process. Therefore, it requires different technique tuning rules to identify model parameters. Smuts[23] presented using modified Ziegler-Nichols tuning rules for integrating process to achieve good and fast disturbances rejection response. So, the model parameters identification was carried out based on open loop step change response and the results of modified Z-N tuning rules were deferent from original Z-N tuning, which were more conservative and not a very aggressive loop response and provided nice tolerance for changes in operating conditions. For integrating processes, process integration rate (Kp) and time delay (θ) are only required to determine the control gains. Where, process integration rate (Kp) can be estimated graphically from the slopes of output response resulted from positive and negative step change of the manipulated variable divided by the magnitude of manipulated variable step change as equation (7).

$$K_p = \left(\frac{slope\ 2 - slope\ 1}{\Delta u_2 - \Delta u_1}\right) \quad (7)$$

While, time delay (θ) is calculated from the time interval between processes input change and the intersection between slope 1 and slope 2.

The control tuning parameters are calculated as follows:

For P control:

$$K_c = \left(1 / (SM * K_p * \theta)\right) \quad (8)$$

For PI control:

$$K_c = \left(1 / (SM * \mathrm{k}_p * \theta)\right) \; ; \quad T_i = 3.33 * SM * \theta \quad (9)$$

For PID control:

$$K_c = 1.2\left(1 / (SM * K_p * \theta)\right) \; ; \; T_i = 2 * SM * \theta \; ; \; T_d = 0.5 * \theta \quad (10)$$

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Where the SM is a fine-tuning factor that should set to a value of 2.0 or larger [23]. The too large value may slow the loop response. In this study, the value of SM is selected to be 3.5 to reduce the overshoot and achieve more conservative. Thus, a well-tuned parameter of PID controller provides a good tradeoff between robustness and closed-loop performance.

4. Relative gains array (RGA)

The RGA is the most employed technique for interaction measure in the multivariable systems and input-output pairing selection [2]. It is a systematic tool commonly used in industry and process control due to its simple calculation, requires only the steady state gain matrix of the process obtained from open loop step response tests, and its controller independency and scale invariant property [24]. On the other hand, the steady state RGA is a simple method can be used to screen out unworkable pairing and provide best control structure suggestion for large scale process contained a subset of integrators. The relative gain proposed by Bristol is defined as the ratio of the open-loop gain in an isolated loop to the apparent loop gain in the same loop, when all other loops are under closed-loop tight control. Considered a general square multivariable system has $m \times m$ input-output variables; the steady state gain system matrix and the RGA are defined as;

$$G(0) = \left[g_{ij}(0) \right] \quad \text{i, j =1,.......,m} \qquad (7)$$

$$\Lambda = \left[\lambda_{ij} \right] = \text{G}(0) * \text{G}^{-T}(0) \quad \text{i, j =1,.......,m} \qquad (8)$$

where $g_{ij}(0)$ is the open loop steady state gain of the ith output results from the jth input change, and * indicates the element-by-element.

All chemical plants essentially contain many pure integrators process in the form of material inventory units as shown in Figure 2. These processes are called non-self-regulating or integrating processes, which are not addressed enough in the literature. In fact, these integrating processes raise issues for consideration, when selecting the best control configuration, where the open-loop gain matrix contains one or more zero elements due to pure integrators. In industrial practice, the traditional approach to deal with this situation is the integrators state variables are specified first and put under closed-loop control. Since the inventory loops are closed, the open-loop steady-state gains can be obtained for the rest self-regulating state variables by step response and the RGA approach can be used to select the workable input-output parings. This method is widely used in the literature with systems containing a large number of integrator loops [8, 25, 26], where it is depending on closing the integrator loops firstly and then looking to the remaining candidate control configurations. Fruehauf and Mahoney[27] introduced control strategy design based on material balance of distillation column. Downs and Skogestad[28], Skogestad[29] proposed a systematic analysis procedure for plantwide control design include inventory loops based on economic and optimization considerations, which is sometimes difficult to apply or time consuming.

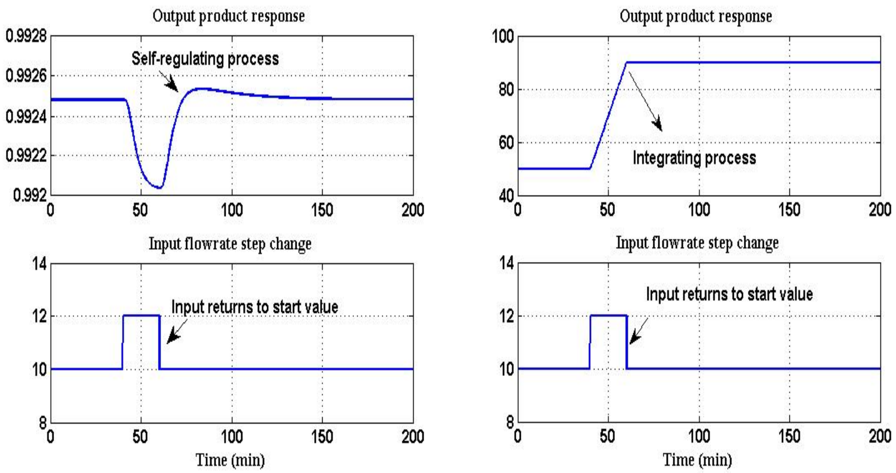

***Figure2*: Self-regulating and non-self-regulating process**

According to this problem, Arkun and Downs[30] proposed a general method to calculate the RGA for multivariable system with integrator variables directly from calculate the state space gains matrix. Thus, the RCA for the integrating process becomes same as the RGA of the non-integrating process such as;

$$RGA\left[G(0)\right] = RGA\begin{bmatrix} g_{NI}(0) \\ g_I(0) \end{bmatrix} \qquad (9)$$

Where g_I is the integrator gain matrix of $G(0)$ and g_{NI} is the non-integrating gain matrix of $G(0)$.

Hence, since the steady state gains of the system can be easily obtained from the step changes of the manipulated variables, this method is applied this study.

5. RESULTS AND DISCUSSION

5.1. Open-loop dynamic response

After providing initial conditions to the column, the next step is 1% changes are introduced individually in all manipulated variables of the open-loops. Hence, the process steady state eventually is moved to a new steady state. The sensitivity to variations of manipulated variables can be characterized by steady-state gains

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$(K = \frac{\Delta y}{\nabla u})$, which is defined as the ratio of the final deviation of a controlled variable from its initial steady-state value to the input step change, time constant (τ_c), and time delay (θ). The results of the sensitivity of the controlled variables to changes in the manipulated variables are given in Table 1. The result show, the top product is more sensitive to change in liquid comes down (LCD) and the impurity of the bottom is sensitive to change in vapor comes up (VCU). On other hand, the reflux drum level is insensitive to the bottom product flowrate change and reboiler level is insensitive to the top product flowrate change. The steady-state gains can be a screening tool at the early design stage of control structure.

Table 1: Steady state model gains and time domain

Manipulated variables	Controlled variables											
	Top Composition (B)			Bottom Composition (B)			Reflux drum level			Reboiler level		
	K	τ_c	θ	K	τ_c	θ	K	τ_c	θ	K	τ_c	θ
Liquid comes down	0.0303	0.013	0.620	0.1513	6.626	0.143	-200	30.27	0.43	200	31.17	0.52
Vapor comes up	-0.002	2.411	0.381	-0.159	5.722	2.139	200	31.04	0.51	-200	31.00	0.35
Top flow-rate	1.9E-07	1.368	0.157	2E-05	1.963	0.335	-200	29.32	1.98	0.00	-	-
Bottom flow-rate	2E-07	1.761	0.017	3E-05	12.591	1.269	0.00	-	-	-200	29.58	1.89

Studying the column sensitivity analysis the system is subjected to disturbances at steady state in order to investigate how large the system may drift away from its nominal operating point. In the same time, determine the impacts of disturbance directions and magnitudes on the stability of the open loop and dynamic response of the column. This can also be used as a tool to analyse how much benefit can be provided by designed control implementation. Thus, two negative and positive direction changes carried out in feed flow rate by (±5 %, ±10 %) in the absence of process control were studied. Figure 3 (a, b, c, d, e, f) shows the responses of open-loop output variables where the effectiveness of disturbance suppression extremely depends on disturbance direction. When increasing the feed flowrate, the system state will drift to another steady-state, the liquid will accumulate in the bottom of the column (flooding) due to the other manipulated variables (heat duty or bottom flowrate) are fixed. This will result in reducing the reboiler temperature which makes the purity of the top product to slightly decrease and the bottom product

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with excess light component. On the other hand, reducing feed flowrate (even slightly -5%) has severe effects on the operation stability as well as dynamic behaviour of the system. With further reduction, it will lose balance and fail to operate. Thus, the availability of effective control structure is necessary to improve the dynamics behaviour of the system, where an open-loop system would not be able to compensate either for disturbances or changes in the process parameters.

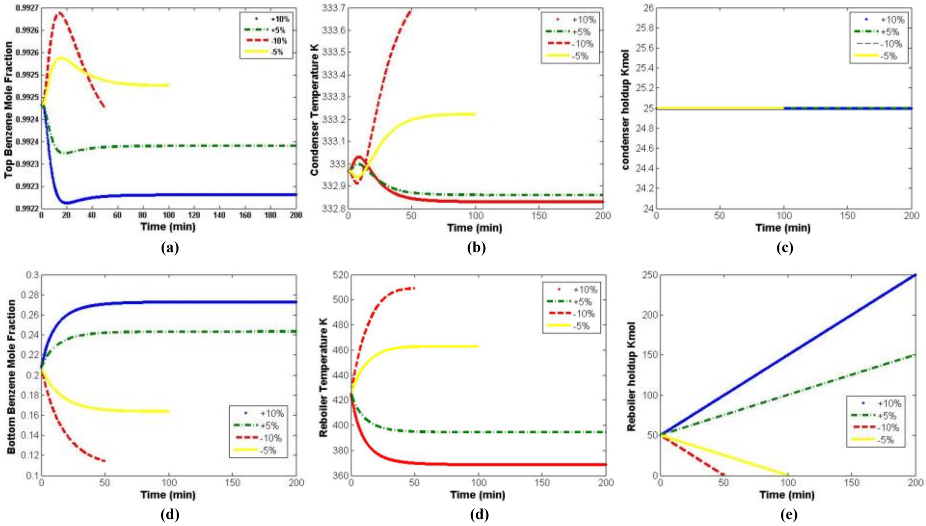

Figure 3: **Sensitivity analysis of feed flow-rate variations on the open-loop**

5.2. Controller pairing

Based on the process constraints and safety, the main objective of the designed control is keeping the purity of benzene in the top product and impurity in the bottom under acceptable specification. Also, the designed control should operate the column in the safe mode though avoiding flooding and drought of the liquid in the reflux drum and reboiler.

Consequently, good initial controller settings are very desirable to stabilize the system. This section addresses the issue of how to pair input-output variables of multivariable process with has large number of alternative control structures.

Based on the results obtained in the Table 1 for step changes in the inputs, the steady-state RGA was computed using the Matlab software for both integrating and non-integrating process (see Table 2). From a quick glance to the results, we can see the positive RGA elements in each column or row close to one are $(\Lambda_{11} = 1.0663,\ \Lambda_{22} = 1.0662,\ \Lambda_{33} = 1.0001,\ \text{and}\ \Lambda_{44} = 0.9998)$. Thus, the only choice that satisfies pairing based on the RGA elements are reflux flow rate to control the distillate composition, vapor boil up rate to control the bottom product composition, top product flow rate to control the reflux drum level, and bottom

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product flow rate to control the reboiler level. This shows a clear preference of the LV-configuration which guaranteed stability of the closed loop response. Whereas both product compositions are mainly affected by changes in the reflux and boil up flow rate, these compositions are weakly dependent on the top and bottom product flow. On the other hand, the reboiler level is independent on the top product changes. Also, the condenser level is independent of the bottom product changes as shown in the Table 2. According to Skogestad *et al.*[18], the LV-configuration is usually recommended to compositions control if their RGA elements is less than 5.

Table 2: steady-state RGA

Description	**Liquid comes down**	**Vapour boil-up**	**Overhead flow-rate**	**Bottom flow-rate**
Overhead Comp. (B)	1.066311	-0.06631	-3.46E-07	3.49E-07
Bottom Comp. (B)	-0.0663	1.066212	-0.00012	0.000214
Reflux drum level	1.74E-06	-0.00012	1.000122	0
Reboiler level	-8.17E-06	0.000223	0	0.999786

5.3. Control-loops tuning

The traditional method for designing the control system for this multivariable system is to use four PID controllers to control each output, which are tuned mainly based on a single-input single-output (SISO). However, the proper tuning of multi-loop PID controllers is quite difficult due to the process interactions in MIMO systems [31].The open-loop tuning procedure was performed using Z-N method for self-regulating loops and modified Z-N for non-self-regulating loops. Thus, PID controller parameters are obtained by observing the dynamic response of the outputs of the system when step changes in inputs are applied in turn for each paired loop. Based on the simulation and the results obtained form process gain (K), time constant (τ_c), and time delay (θ) simultaneously in Table 1. The controller gains value for P, PI, and PID were computed and listed in Table 3. This study used only proportional and integral control (PI) and the controllers implemented are the velocity forms of PID control algorithms.

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Table 3: Summary of controller parameters gains using Ziegler-Nichols method

output	Top Composition (B)			Bottom Composition (B)			Reflux drum level			Reboiler level		
	KC	Ti	Td	KC	Ti	Td	KC	Ti	Td	KC	Ti	Td
P	6.782			16.776			0.144			0.168		
PI	6.104	2.066		15.098	7.131		0.1299	23.08		0.151	22.03	
PID	8.139	1.24	0.31	20.131	4.279	1.07	0.1732	13.86	0.99	0.202	13.23	0.945

5.4. Closed-loop disturbance rejections

Evaluating the robustness of the proposed multi-loop PI controllers and investigate the proper overall control strategy of the column to reject the process disturbances and provide the nominal process (set-points tracking). Two different sources of disturbances are likely to be encountered by this multivariable distillation column. Feed flow-rate changes occur due to variations in feed stream value, and feed composition disturbances originate from the feed source tank. Each of these disturbances has its own peculiar effect on the column. These types of disturbances usually represented the most challenging disturbance face on a regular basis for distillation columns. Therefore, the closed-loops performance was analyzed through the dynamic response of the output variables; the top and bottom product composition, the reflux drum temperature and level, and reboiler temperature and level. For that reason, the disturbances rejection will be presented in this section. since in many practical process control systems, the good disturbance compensation is more concerning than set-point tracking [32, 33].

5.4.1. Feed flow-rate changes

Two steps up and down (±10%, ±20%) disturbances of feed flow-rate were introduced to the process at time (t=0). The closed-loops dynamic behavior of positive and negative step changes are shown in Figure 4 (a, b, c, d, e, f), where the purity and impurity of the top and bottom products were achieved through the process constraints. The temperature of the condenser and reboiler was kept so close to its set-point and the level of reflux drum and reboiler was managed at acceptable range and avoiding any faults resulted from flooding and drought of the liquid in the column. Furthermore, all the control variables moved quickly to their final steady-state values. By comparing the results obtained in Figure 4 with that in Figure 3, there is clear improvement in the dynamic behavior of the system despite the increase in the magnitude of disturbances. It is clearly seen that the PI-control and L-V configuration provide superior performance for disturbance rejection and maintained the process at safe mode successfully.

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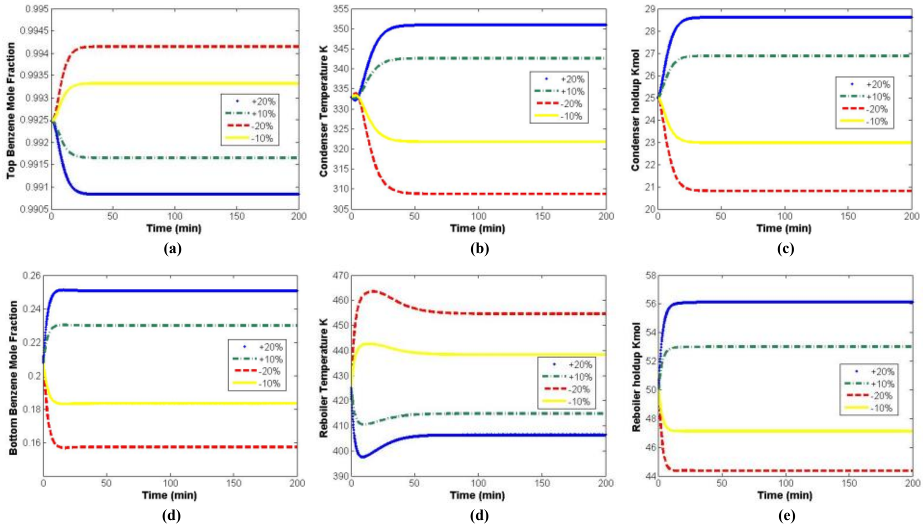

***Figure 4*: Closed-loop performances for top and bottom of the column to step changes in feed flow-rate (±10%, ±20%)**

5.4.2.Feed composition changes

Feed composition changes are commonly considering the most challenging disturbance that distillation columns faces on a regular basis. Two steps up and down (±5%, ±10%) variations of benzene composition were made in the feed flow rate at time (t=0). The dynamic simulation results of the model show the ability of the PI control and L-V configuration to reject these disturbances successfully. As shown in Figure 5 (a, b, c, d, e, f), the purity of benzene in the top product and impurity in the bottom were kept so close to their set points. Also, the temperature and level of the reflux drum and reboiler were kept under acceptable range and the system operated in the safe mode.

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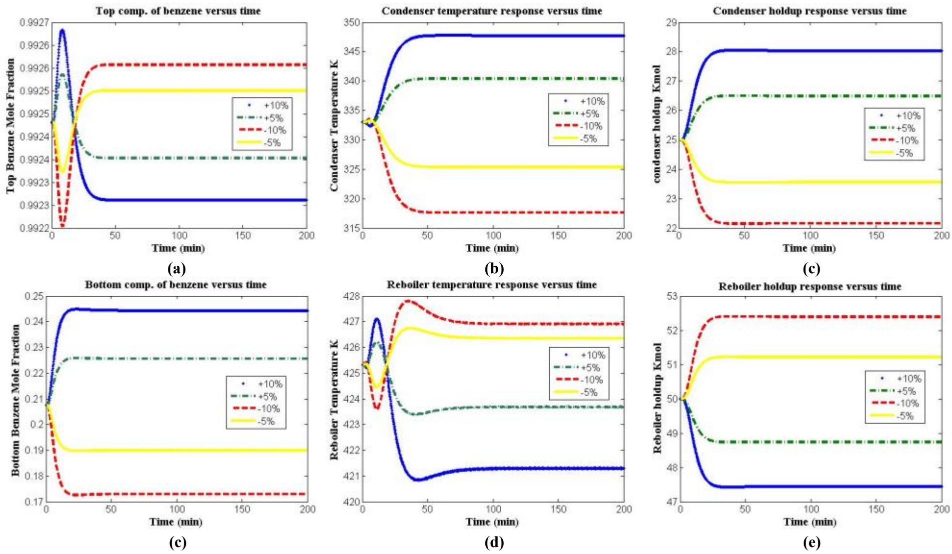

***Figure 5*: Closed-loop performances for top and bottom of the column to step changes in feed composition variations (±5% and ±10%)**

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6. Conclusion

Distillation is considered as a high energy consuming and operating cost process in the petrochemical and refining industries. Thus, robust and efficient control design is one way of energy saving and necessary to ensure stable dynamic performance, correct the process deviations, and operate the system at desirable condition. Despite, there are many alternative control structures proposed for multivariable distillation column. Selecting an appropriate control structure plays the most important role in control system design. Based on the steady state RGA technique obtained from the steady-state sensitivity analysis, the L-V control configuration is the best control strategy. The unworkable or poor control structures were eliminated for large scale process containing a subset of integrators. In this study, the tuning parameters of PID controllers were determined using Z-N rules for self-regulating loops and modified Z-N rules for integrating loops, which provided the best performance of disturbance rejection and fast set-point tracking.

The results obtained from the dynamic simulations of the closed loops revealed the reliability and capability of the control structure to maintain a nominal operating point and eliminate the interactions between the process variables. Furthermore, this designed control scheme has demonstrated the ability to reject negative or positive persistent disturbances in feed flow rate and feed composition in reasonably short times and achieved the desired operation without violating the process constraints comparative with the open loop response**.** In practice, this means that when the plant is subjected to disturbances, it will still operate within an acceptable distance from the optimum, and there is no need to re-optimize when disturbances occur. This case study model and decentralized control structure is currently being used for further studies in ability of existing control strategies to manage and suppress abnormal process situations in the system and faults detection and diagnosis.

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المجلة مفهرسة في المواقع الآتية :

العام	2021	2022	2023	2024	2025
معامل أرسيف	0.2692	0.1954	0.3759	0.3068	0.5978
معامل التأثير العربي	1.60	1.73	1.25	1.55	1.81