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Design of Multivariable Control System of a Distillation Tower via Simulation Using MATLAB/Simulink

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A B S T R A C T

This paper deals with the multivariable control system of the distillation tower by applying multi-structures in MATLAB/Simulink for a binary mixture of benzene and toluene. Four structures configurations of the distillation are applied based on the level and temperature variables. The PID controller is used in all structures in multivariable. These structures are compared with different disturbances. The integral absolute error is a criterion to test the controller's performance under step change disturbances. The controller's performance was investigated by recording responses to disturbances in set-point of reflux ratio, flow-rate of top and bottom products. The step testing appears to be the single-ended temperature control with bottom and top-level structures to regulate the flow rate of the bottom and top products. The best structure is top-level, bottom level and condenser temperature because the column made more stable, the integral absolute error is minimum value and fast access to set-point value.

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محاكاة تصميم نظام سيطرة متعدد المتغيرات لبرج التقطير بأستخدام ماتلاب/سيمولنك

دريد فاضل أحمد/ قسم الهندسة الكيمياوية، كلية الهندسة ، جامعة تكريت ، تكريت ، العراق
مهند ياسين نواف/ قسم الهندسة الكيمياوية، كلية الهندسة ، جامعة تكريت ، تكريت ، العراق

الخلاصة

البحث يتناول محاكاة تصميم منظومات سيطرة متعددة المتغيرات لبرج التقطير بتطبيق اشكال متعددة من منظومات السيطرة لمزيج بنزين-تلوين باستخدام برنامج ماتلاب/سيمولنك. تم تطبيق اربعة انظمة سيطرة معتمدة على متغيرات درجة الحرارة ومستوى السائل في اعلى واسفل البرج. كما تم استخدام مسيطر نوع تناسبي-تكاملي-تفاضلي متعدد المتغيرات. استخدم معيار تكامل مطلق الخطأ في تحليل أداء المسيطر وقورنت كفاءة هذه المسيطرات بادخال عدة دوال مؤثرة. تم فحص كفاءة المسيطر من خلال استجابته لتغير درجي بنسبة الراجع ومعدلات الانتاج في اعلى واسفل برج التقطير. اظهرت النتائج ان افضل ربط للسيطرة بوجود متغيرات درجة الحرارة-مستوى السائل اعلى البرج-مستوى السائل اسفل البرج لتنظيم معدلات الانتاج في اعلى واسفل البرج. واظهرت النتائج ايضا ان البرج اكثر استقرارا واقل قيمة لمعيار تكامل مطلق الخطأ وأسرع الوصول الى القيمة المرغوبة كان في شكل نظام السيطرة لدرجة حرارة المكثف ومستوى السائل لأسفل واعلى البرج.

الكلمات الدالة: بناء منظومة السيطرة، برج التقطير، سيطرة متعددة المتغيرات، محاكاة سيمولنك.

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1. Introduction

Two key steps in a successful a multi-loop controller design are the control configuration selection and the controller tunings. Control configuration for a distillation column can be selected out of the knowledge of the thermodynamic, heat and mass parameters. Unsuitable selecting of input-output pairing can result in poor performance of controller. Decouples are introduced into the multi-loop configuration to compensate for the control loop interactions. The distillation column is the most difficult process to control because of long-dead times, long lags, and process nonlinearities inherent to the process. The studies of distillation have focused not only on control aspects but also on the time response or dynamics of distillation. Distillation technology remains at the forefront of control research due to its importance in separating materials. Its powerful tool for a host of studies: for example, to evaluate alternative process and controller structures, tune controllers, alternatively determine the dynamic effects of disturbances and to optimize the operation of the plant, [1].

The simulations were conducted with and without the presence of a degassing system to evaluate the efficiency of this system in the control of the volatile content. To improve the efficiency of the degassing system, a control loop based on a feedback controller was developed. Kano et al. [2] applied predicative inferential control through the predictability of the focus of the process control variables directly, rather than the current estimates to focus the results of the dynamic simulation showed that the outcome control system was supplementing the cascade control to a large extent from other control systems. Fabro et al. [3] developed intelligent predictive controller Continuous distillation tower using the recurrence of neural networks to highlight the process and processing of predictions about the behavior of the reliance on reaction control applied to the system and then this information is used by the control logic to complete the best performance of the control. Ahmed and Mohammad [4] studied the dynamic behavior of continuous distillation columns and implemented different types of control strategies for the separation of binary mixture composed of ethanol and water. They applied PI and PID controllers to control the top and bottom concentration and tuning parameters of this controller by using (Cohen-coon) method and trial and error method. The study also includes designing fuzzy logic controller using MATLAB program and was compared with other controllers by using mean square error (MSE) and integral square error (ISE). They concluded that the performance of the fuzzy controller was better than the other methods through fast access to the desired value and compensate the disturbances. Ahmed and Khalaf, [5] developed a model for crude oil distillation column of Baiji refinery. They used the model as a reference for optimal predicted operation condition for the column and they applied advanced control strategies of crude distillation column. Ahmed and Shakoor [6] derived a dynamic model for batch distillation column. Thus, both theoretical and experimental studies are carried out. They applied the neural network and PID controllers of the top product temperature. The simulation of control methods of the column was built by using MATLAB software to study the responses, comparison between control methods and then select the best method.

Acharya et al. [7] developed the mathematical model of a Wood-Berry distillation column to separate methanol from water. They used PID, decoupled PID and model predictive controllers. Niloofar et al. [8] designed a robust and stable multivariable decoupling based Proportional-Integral-Derivative (PID) like fuzzy scheduling technique to control multi-input, multi-output (MIMO) nonlinear uncertain dynamical distillation column. A fuzzy PID scheduling controller is applied to multivariable filter decoupling method to solve the coupling effect in a distillation column and improve the temperature controllability and observability in distillation column. This approach reduces the variation in output temperature and improves the rise-time in the certain and uncertain condition. The frequency and phase response have been tested and approved the power of stability and robustness in modified multivariable filtered fuzzy PID scheduling control distillation column.

In this study, the optimum control structure for the distillation column will be chosen assuming that we have available flows, level, and temperatures for control. The derived model will be used to simulate the distillation column using Simulink/MATLAB and then design the optimum control structure. The study includes design PID control system to control the top temperature, condenser temperature and level of top and bottom by using MATLAB simulation and choosing the best structure of PID controller at different conditions of feed flow rate, weight fraction of feed, reflux ratio and level of top and bottom through the simulation.

2. Distillation Column

The system of distillation is a binary and non-ideal liquid solution with ideal vapor system. The feed to the tower is a mix at boiling point, saturated liquid and temperature it 92.3 ºC. The vapor hold-up on the tray is neglected compared to the liquid loaded on the tray. The flow rate of liquid to each tray is constant — the liquid and vapor on each tray are in equilibrium state. The vapor output from the top of the distillation tower is intensive entirely for obtaining a fixed pressure of the tower and at 1 bar. The distillation column separates the benzene from toluene which enters as a feed stream with flow rate and composition between the rectifying and the stripping section as shown in Fig. 1. The column consists of 14 trays and the overhead vapor is totally condensed in a water-cooled condenser. The reboiler is heated by steam, and the preheated feed stream enters the column at the feed tray as saturated liquid. The manipulated variables are the flowrates of reflux, top and bottom products.

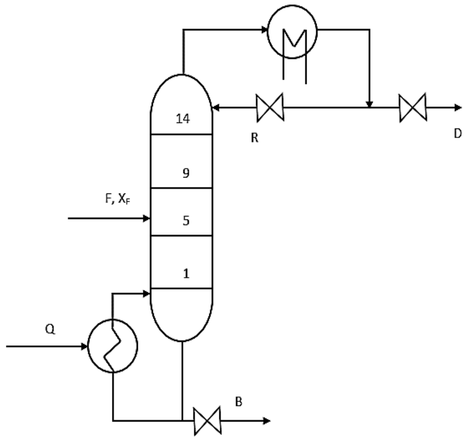

Fig. 1. Flowsheet of the distillation column.

3. Simulation Works

MATLAB /Simulink software is used to simulate the distillation column. The mathematical model is derived for the column in the form of a set of systems, and each system component with a set of subsystems which represents the model equations for distillation column. The model of the distillation column used throughout the paper is developed by Ahmed and Nawaf, [9] composed by the mass, component mass, and energy balance equations used as basis to implement the Simulink diagram. The dynamic model that has been developed and determining the extent of the system in response to some of the changes and the PID controller were used. PID auto tuning can be deployed to embedded software for automatically computing PID gains in real-time. The operating points are found and computed exact linearization of Simulink models at various operating conditions. Simulink control design provides tools that let compute simulation-based frequency responses without modifying model. PID tuner provides a fast and widely applicable single-loop PID tuning method for the Simulink PID controller blocks to achieve a robust design with the desired response time.

3.1 Structures of the Control System

The control structures of the distillation column in MATLAB/Simulink environment are designed and determining the extent of the system's response to some of the changes in the variables and are shown in Figs. (2-5).The four control structures of the distillation column are designed based on the type of the controlled and manipulated variables. Table 1 shows the details of the four control structures.

Table 1
Controller parameters by an auto method in MATLAB/Simulink

Structure No.	Number of Pairs	Pair of Controlled variable and Manipulated variable
1	3	1. Top tray temperature -Top product flowrate 2. Level of top product- Reflux flowrate 3. Level of the bottom column- bottom flowrate
2	3	1. Top tray temperature -Top product flowrate 2. Level of top product- Reflux flowrate 3. Bottom temperature - bottom flowrate
3	3	1. Condenser temperature - Reflux flowrate 2. Level of top product- Top product flowrate 3. Bottom temperature - bottom flowrate
4	3	1. Condenser temperature - Reflux flowrate 2. Level of top product- Top product flowrate 3. Level of the bottom column- bottom flowrate

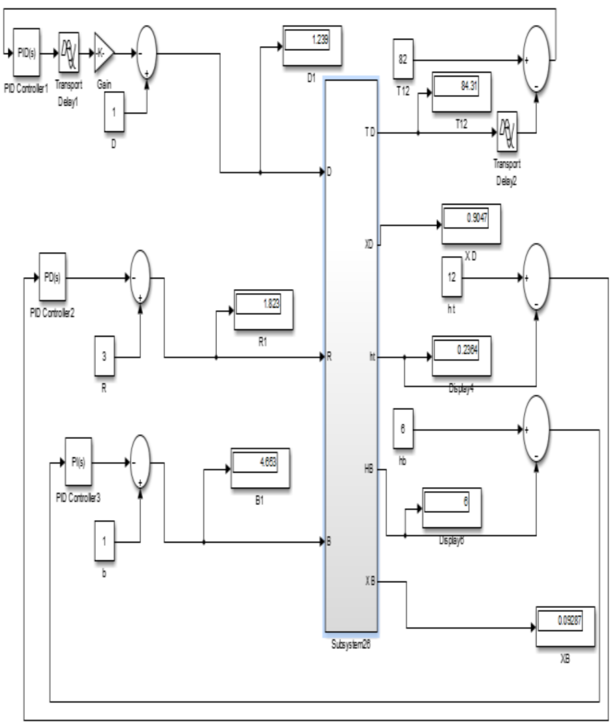

Fig. 2. Control structure No.1.

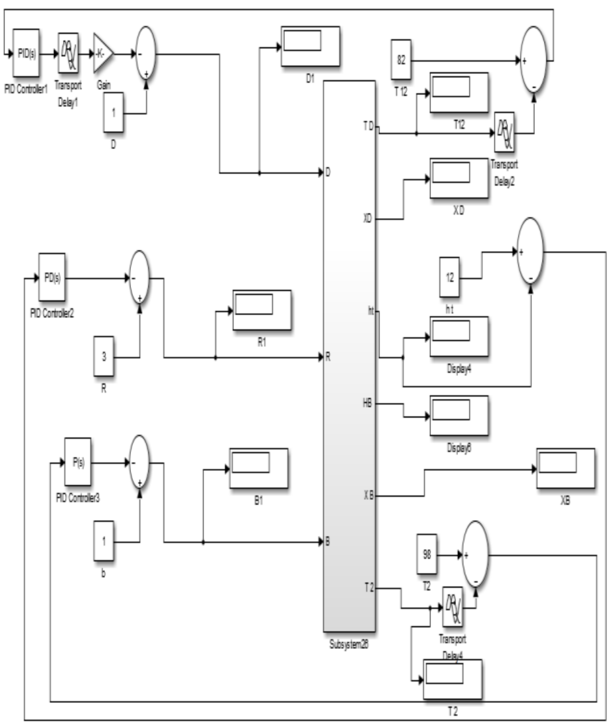

Fig. 3. Control structure No.2.

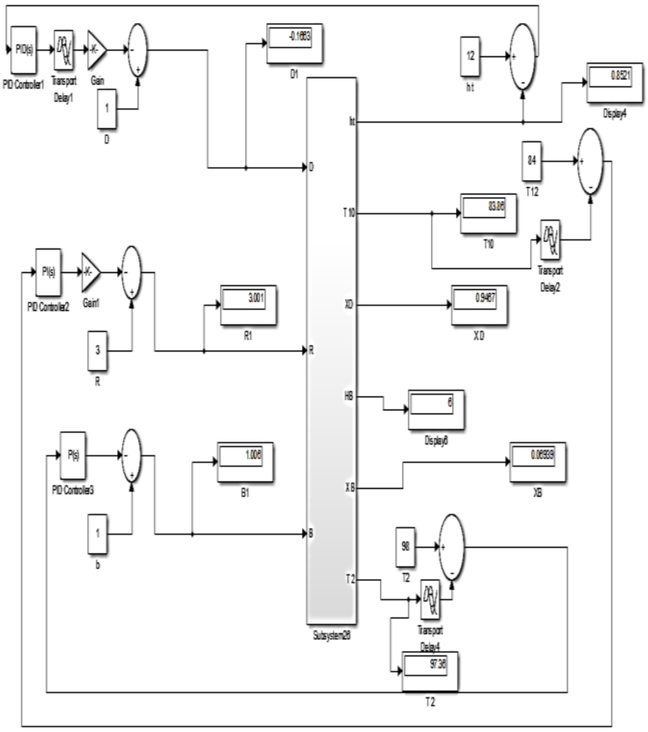

Fig. 4. Control structure No.3.

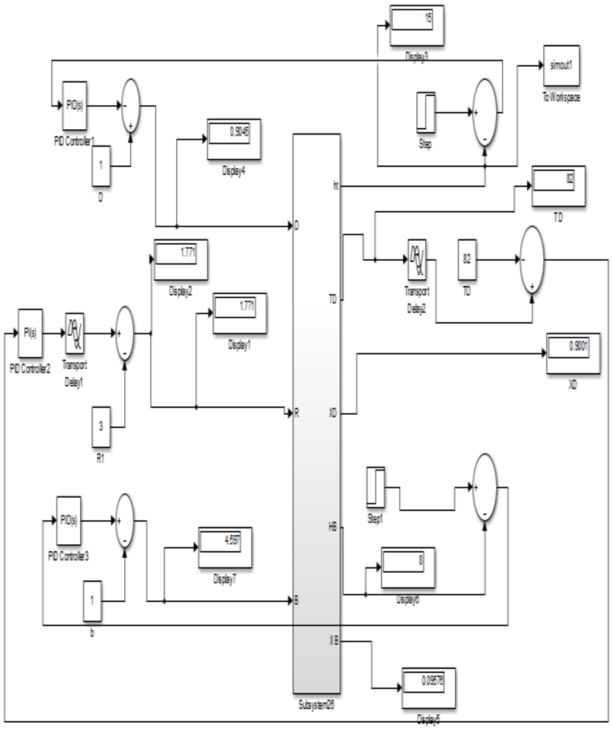

Fig. 5. Control structure No.4.

4. Results and Discussion

The results show the dynamic behavior of the distillation column to separate the benzene from toluene and discussing them, comparing two different control systems PID on the modified dynamic four-structure model of distillation column are made. The reflux flowrate is the basic element to control distillation column. The flowrate and composition of the feed are the disturbance by using step change function. The four structures of multi-loop PID controller are used to control level and temperature of distillation column and to compare between those structures.

4.1 PID Controller

The controller parameters for four control structures are given in Table 2. The controllers' parameters for four control structures through the auto method in MATLAB/Simulink noted that these parameters don't give good results with parameters chosen manually so that the controllers tuning were done by trial and error. Table 3 shows the controllers' parameters in which tuning were manually done by trial and error.

4.2 Level Control

Level controllers are used for maintaining the level in top and bottom products. Figs. (6-11) show the results of the level controller. Loose level control on the top and bottom products has been shown to worsen the composition control problem for material balance control configurations (when either bottom or top flow is used as a manipulated variable for composition control). When top or bottom flowrate is adjusted, the internal vapor/liquid traffic changes only after the corresponding level controller acts because of the changing in top or bottom flow. On the other hand, if a level controller is tuned too aggressively, it can result in oscillations passing back to the column and contributing to erratic operation. When the reboiler duty is set by the level controller there will be two-level loops on a distillation column which must be maintained at an acceptable value. The reflux drum level (top-level) must be maintained at an acceptable value. The performances of these controllers are calculated using the integral absolute error for different disturbances. Table 4 shows performance of these controllers.

Table 2
Controller parameters by an auto method in MATLAB/Simulink

Controller	Structure	K_c	τ_I (min.)	τ_D (min.)
PID	1	0.224	1.27	1.87
PI	1	0.224	1.27	-
PD	1	0.224		1.87
PID	2	0.224	1.27	1.87
PI	2	0.224	1.27	-
P	2	0.224		-
PID	3	0.224	1.27	1.87
PI	3	0.224	1.27	-
P	3	0.224		-
PID	4	0.224	1.27	1.87
PI	4	0.224	1.27	-
PID	4	0.224	1.27	1.87

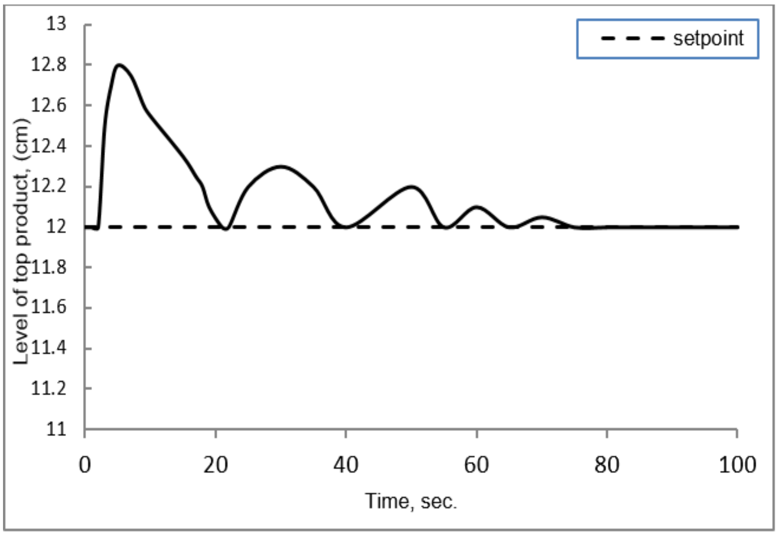

Fig. 6. The response of the level of the top for step-change in bottom flow from (1 to 1.4 kmol/s) at set point 12cm at PID of control structure 4.

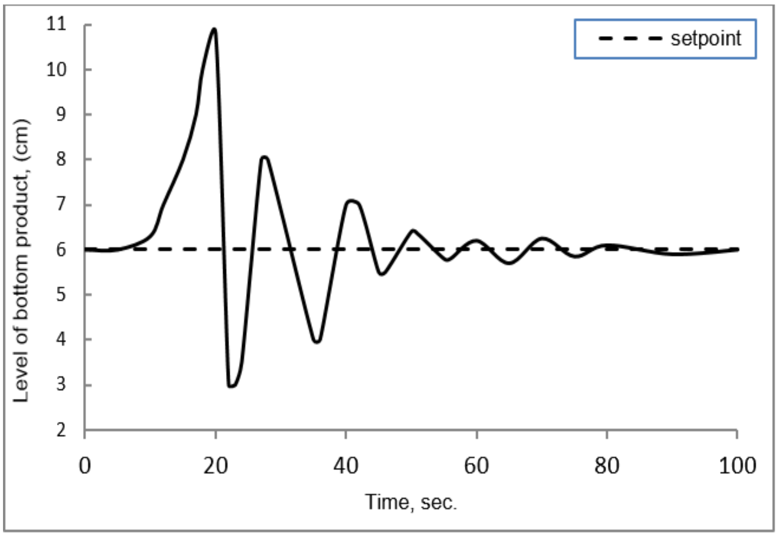

Fig. 7. The response of the level of the bottom for step-change in bottom flow from (1 to 1.4 kmol/s) at setpoint 6cm at PID of control structure 4.

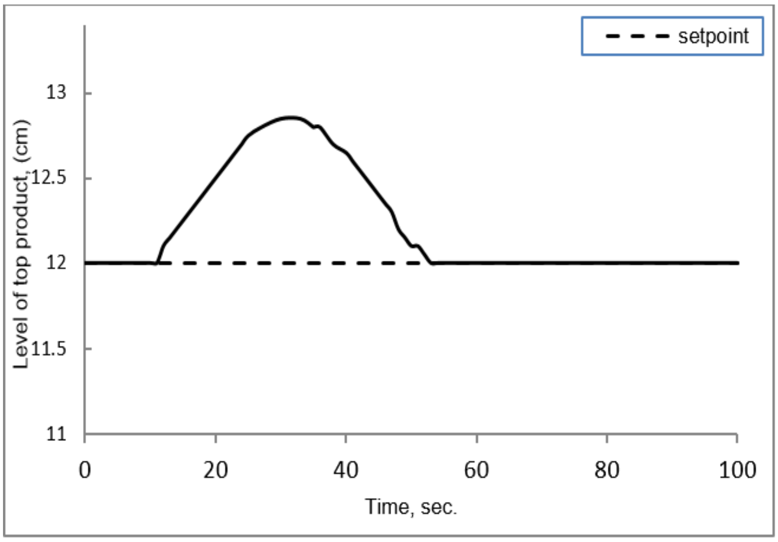

Fig. 8. The response of the level of the top for step-change in top flow from (1 to 1.4 kmol/s) at setpoint 12cm at PID of control structure 4.

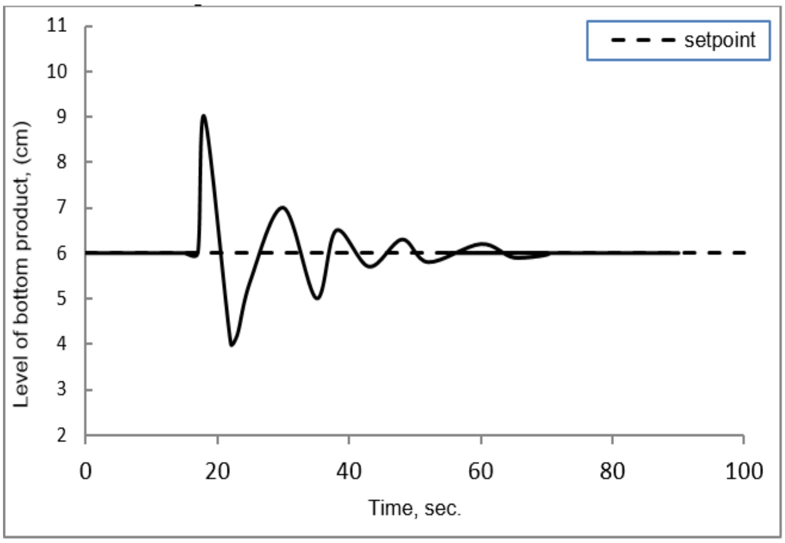

Fig. 9. The response of the level of the bottom for step-change in top flow from (1 to 1.4 kmol/s) at set point 6cm at PID of control structure 4.

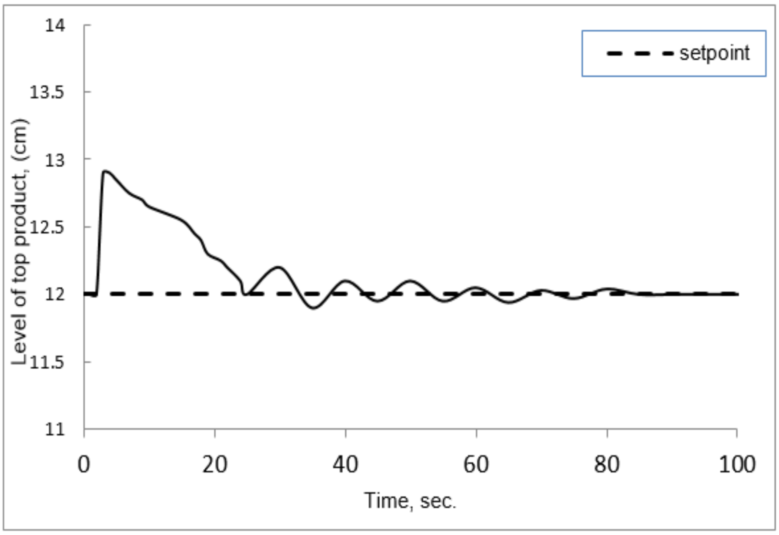

Fig. 10. The response of the level of the top for step-change in reflux ratio from (3 to 4) at set point 12cm at PID of control structure 4.

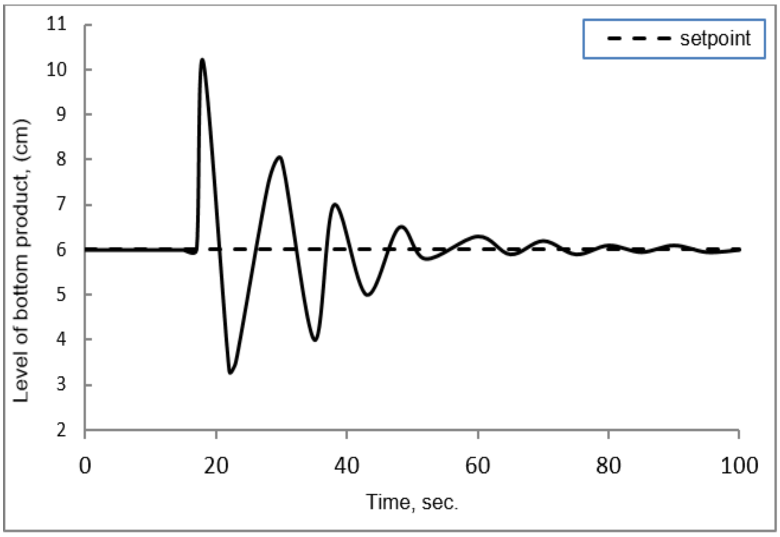

Fig. 11. The response of the level of the bottom for step-change in reflux ratio from (3 to 4) at setpoint 6cm at PID of control structure 4.

Table 3
Controller parameters by trial and error

Controller	Structure	K_c	τ_I (min.)	τ_D (min.)
PID	1	0.5	0.1	1
PI	1	0.1		0.1
PD	1	0.1	1	
PID	2	0.1	0.1	0.1
PI	2	0.1		0.1
P	2	-0.1		
PID	3	0.1	0.1	1
PI	3	-0.01	-0.01	
P	3	-0.01		
PID	4	1	0.04	5
PI	4	10	1	
PID	4	-1	-2.7	-4

4.3 Temperature Control

In this simulation test, the top flow rate step increase from 1 kmol/sec. to 1.4 kmol/sec. In theory, the top tray temperature response should be a monotonic increase to some new steady state. However, the simulation results, shown in Figs. (12-17) are at all monotonic and do appear to be reaching steady-state after different times. Run 5 shown in Fig. 16 after a time of three seconds, the condenser temperature rose extremely quickly to about 82.5°C, then curiously decreased and increased again, followed by another decrease as wave. When tuning the controller parameters, the most effect parameter is the integral time constant; if the integral time constant has increased, the response and the controller performance are increased too. While the controller's gain was less effect compared to the integral time constant and it is found that the settling time of distillation column PID controller was less than 40 min, except running 5,6 those take more time to arrive at steady state but shorter time to arrive to set point. The values of Integral absolute error (IAE) calculation for PID simulation are shown given in Table 5. IAE value for runs 4,5 and 6 was 24.5664, 25.1048, and 25.1543 respectively, and those give the desired performance because those take smaller time to arrive at steady-state and does not appear effected in step-change.

Table 4
The integral absolute error (IAE) of level control

Run no.	Controlled variable	Variable of step-change	Structure	Value of step-change	IAE
1	Top level (Set point 12cm)	Reflux ratio	4	3-4	30.7048
2	Top level (Set point 12cm)	Top flow rate, kmol/sec	4	1-1.4	30.0764
3	Top level (Set point 12cm)	Bottom flow rate, kmol/sec	4	1-1.4	27.7767
4	Bottom level (Set point 6cm)	Reflux ratio	4	3-4	136.0102
5	Bottom level (Set point 6cm)	Top flow rate, kmol/sec	4	1-1.4	127.7697
6	Bottom level (Set point 6cm)	Bottom flow rate, kmol/sec	4	1-1.4	127.6333

Table 5
The integral absolute error (IAE) of temperature

Run no.	Controlled variable	Variable of step-change	Structure	Value of step-change	IAE
1	Temperature of top tray (setpoint 82ºC)	Bottom flow rate, kmol/sec	1	5-7	31.3638
2	Temperature of top tray (setpoint 82ºC)	Top flow rate, kmol/sec	1	1-1.4	31.3682
3	Temperature of top tray (setpoint 82ºC)	Reflux ratio	1	3-4	28.6983
4	Temperature of condenser (setpoint 82ºC)	Reflux ratio	4	3-4	24.5664
5	Temperature of condenser (setpoint 82ºC)	Top flow rate, kmol/sec	4	1-1.4	25.1048
6	Temperature of condenser (setpoint 82ºC)	Bottom flow rate, kmol/sec	4	1-1.4	25.1543

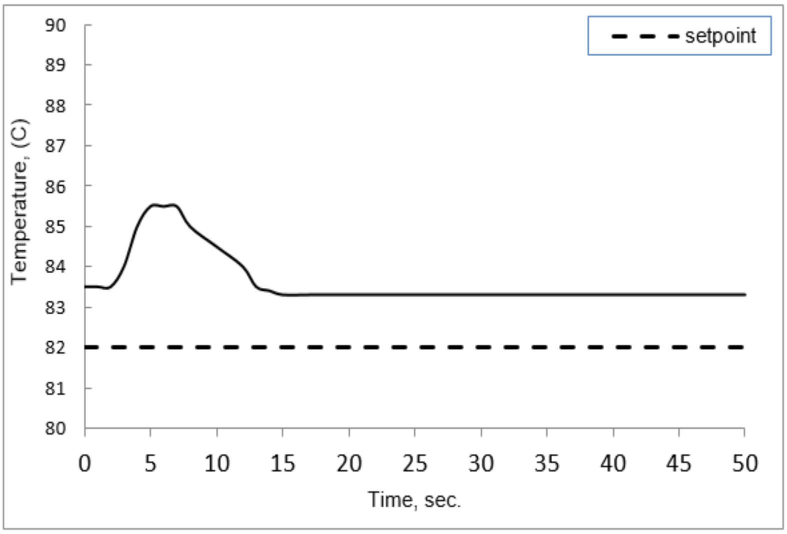

Fig. 12. The response of the top tray temperature for step-change in top flowrate from (1 to 1.4 kmol/s) at setpoint 82C at PID of control structure 1.

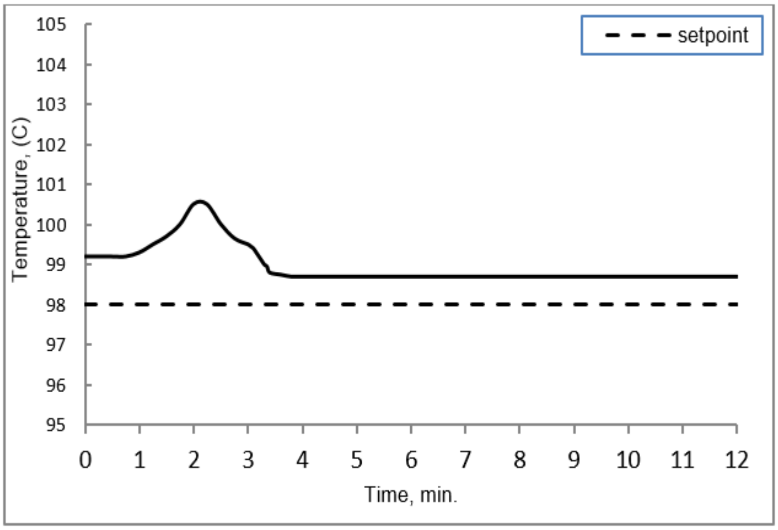

Fig. 13. The response of the temp. of tray 2 for step-change in top flow from (1 to 1.4 kmol/s) at setpoint 98C at PID of control structure 2.

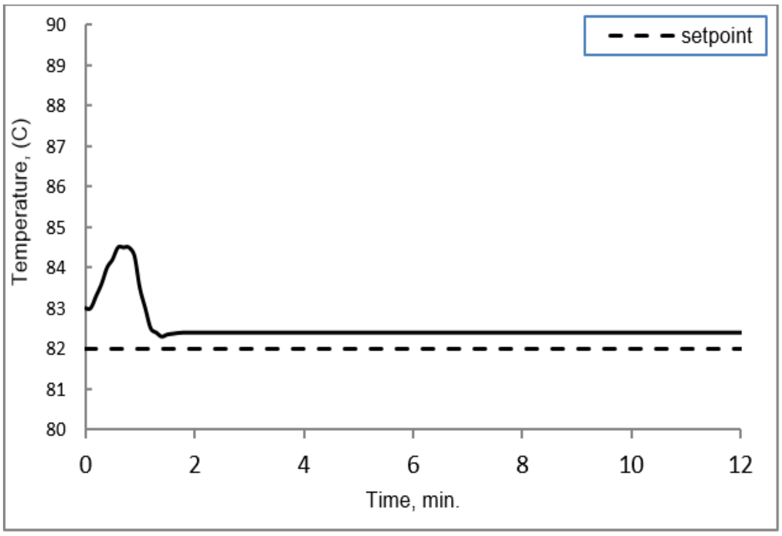

Fig. 14. The response of the temp. of the top tray for step-change in reflux ratio from (3 to 4) at setpoint 82C at PID of control structure 2.

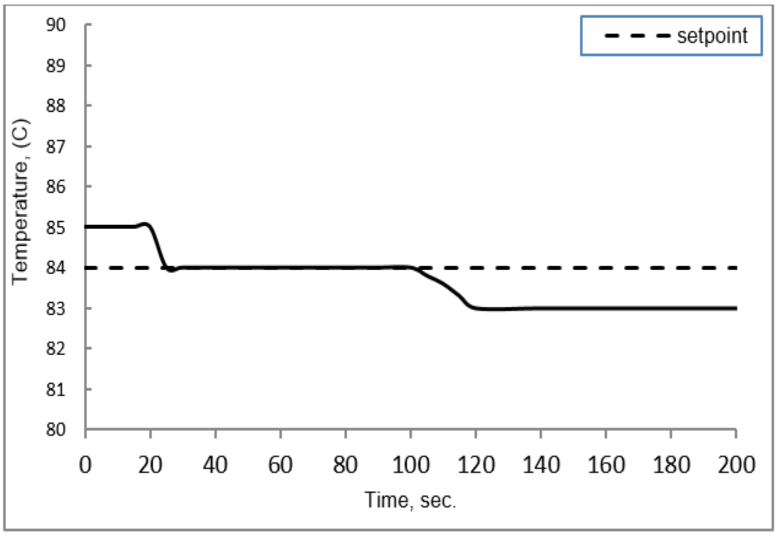

Fig. 15. The response of the temp. of tray 10 for step-change in reflux ratio from (3 to 4) at setpoint 84C at PID of control structure 3.

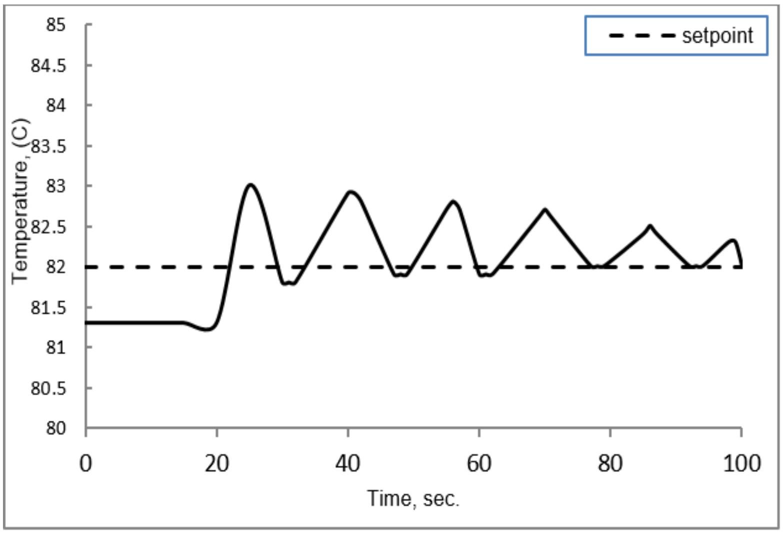

Fig. 16. The response of the temp. of the condenser for step-change in top flow from (1 to 1.4kmol/s) at setpoint 82C at PID of control structure 4.

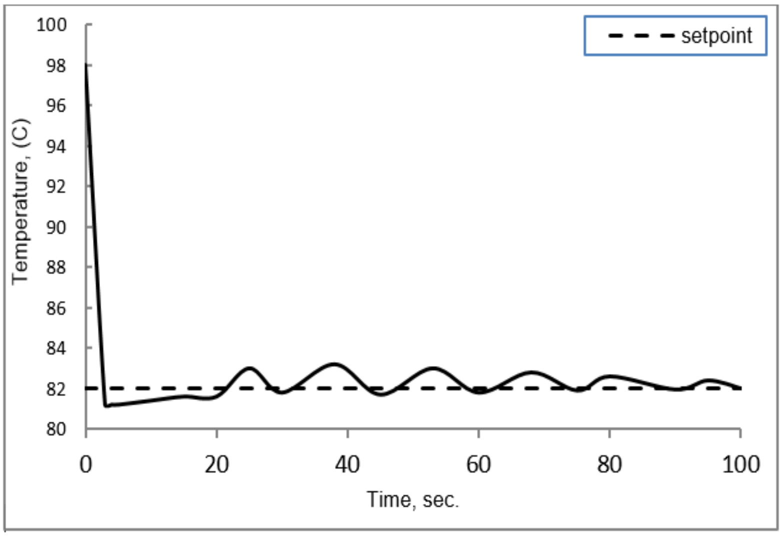

Fig. 17. The response of the temp. of the condenser for step-change in reflux ratio from (3 to 4) at setpoint 82C at PID of control structure 4.

4.4 Comparison between Control Methods

The values of IAE for the PID in four control structures, level, and temperature controllers are given in Tables (4) and (5). The comparisons between the performances of controllers were made using the integral absolute error (IAE). The PID controller of structure 4 for level gave a good performance with an IAE range (27.7767 – 30.7048) for top-level and (127.6333 – 136.0102) for bottom level. While the IAE for PID controller of structure 4 for temperature gave a good performance with an IAE range (24.5664 – 25.15543) for that the best structure of control of the distillation column temperatures and level using PID is control structure 4 and that shown in Fig. 18. and 19 for temperature and level, although we had a lower IAE temperature in structure 2 for level it was higher than others.
The best control structure is structure 4; it has three loops of controller, first loop controlling top level, second loop controlling temperature of condenser, and third loop controlling bottom level. In this section we try to reconfigure structure 4 by testing of set one loop, two loops, and then three loops on distillation column.

In one loop bottom level controlled by top flow rate notice, the top flow rate is still going up without control, as shown in Figs. 18. and 20 that is unlike performance, the same things happen for bottom level when controlling top-level as shown in Figs. 21. and 22. The effect on temperature and mole fraction does not appear when let without controller but it takes more time to arrive at a steady-state, as shown in Fig. 19.
To summarize the above results of close-loop tests, it must be noted that top and bottom level must be controlled; temperature response appeared for a shorter time to arrive steady state under control than without. For that reason the best way for controlling distillation column is single-ended temperature control with the use of top and bottom flow rate to control top and bottom levels.

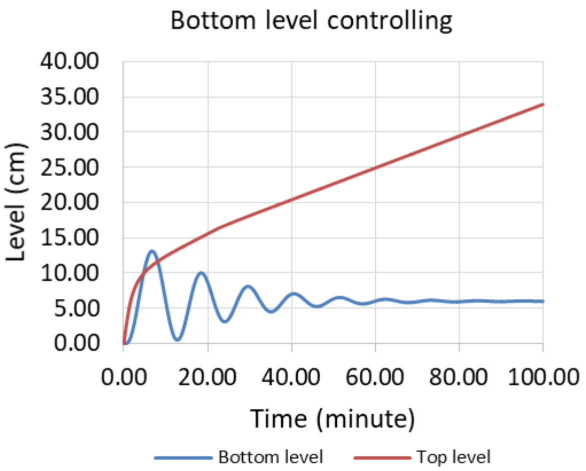

Fig. 18. Comparison of the response of the level if setting by one loop controlling the bottom level of PID of control structure 4.

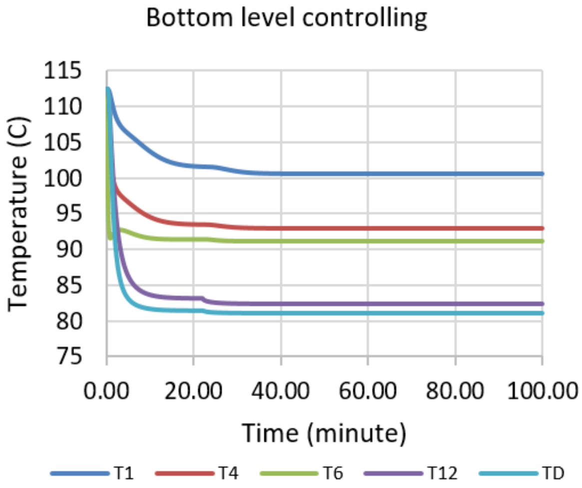

Fig. 19. Comparison of the response of the temperature if setting by one loop controlling the bottom level of PID of control structure 4.

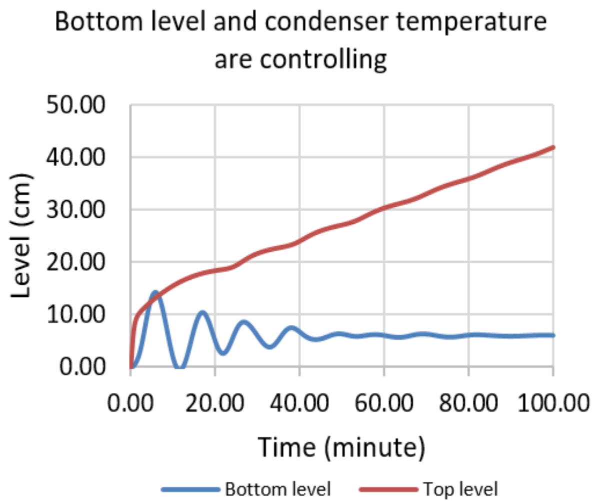

Fig. 20. Comparison of the response of the level if setting by two loops controlling bottom level and condenser temperature of PID of control structure 4.

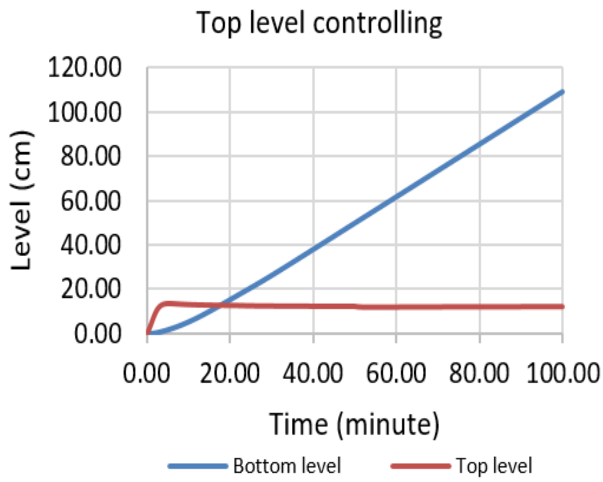

Fig. 21. Comparison of the response of the level if setting by one loop controlling the top level of PID of control structure 4.

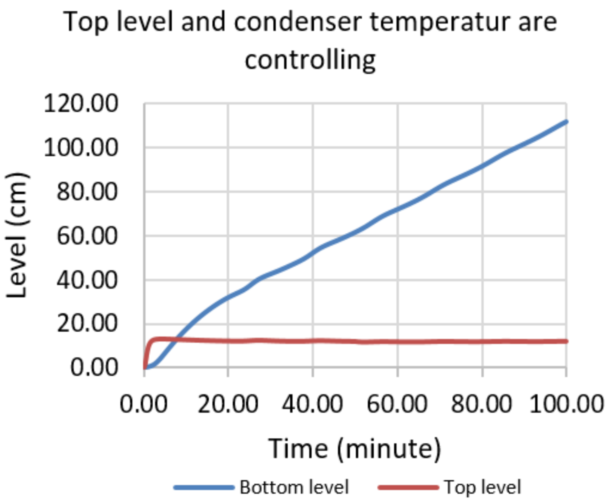

Fig. 22. Comparison of the response of the level if setting by two loops controlling top-level and condenser temperature of PID of control structure 4.

5. Conclusions

The primary purpose of this research was to choose the control structures of distillation columns. A non-linear dynamic distillation model was built, which included the level of both the bottom and the top of the tower. In the system studied here, as the benzene-toluene ratio changed, the effect on column temperature was more significant than the effect of changing the feed rate itself. The column had been separating two critical components with closer boiling points, this effect of composition on bubble point would have been much less, and the temperature control loop might have been less interacting with the composition control loop. The control problem of the distillation column was solved using control configuration. Four control structures were designed and select the best structure. From this study the best structure is top-level, bottom level and condenser temperature because the column made more stable, the integral absolute error is minimum value and fast access to set-point value.

6. References

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