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PLANTWIDE CONTROL STRUCTURE DESIGN OF HYDRODEALKYLATION PROCESS

Mr. Teerapong Juncharoenwongsa

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By	Mr. Teerapong Juncharoenwongsa
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Thesis Advisor	Assistant Professor Montree Wongsri, D.Sc.

Accepted by the Faculty of Engineering, Chulalongkorn University in Partial Fulfillment of the Requirements for the Master's Degree

....Dean of the Faculty of Engineering

(Professor Bundhit Eua-arporn, Ph.D.)

THESIS COMMITTEE

....Chairman

(Associate Professor Muenduen Phisalaphong, Ph.D.)

....Thesis Advisor

(Assistant Professor Montree Wongsri, D.Sc.)

....Examiner

(Assistant Professor Soorathep Kheawhom, Ph.D.)

....External Examiner

(Chaiyapop Siraworakun, D.Eng.)

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กระบวนการไฮโดรดีอัลคิเลชันประกอบด้วยแผนผังกระบวนการที่มีความซับซ้อนกับสายป้อนกลับมวลสารและพลังงานและหน่อยปฏิบัติการที่มีความแตกต่างกันจำนวนมาก ส่งผลให้พฤติกรรมไม่เชิงเส้นของกระบวนการปิโตรเคมีนี้มีความซับซ้อนอย่างมาก ดังนั้นโครงสร้างการควบคุมแบบทั้งกระบวนการที่มีประสิทธิภาพจะเห็นได้ชัดเจนว่าเป็นสิ่งที่สำคัญอย่างมาก ในการทำงานในปัจจุบันนี้ วิธีการออกแบบโครงสร้างการควบคุมแบบทั้งโรงงานของวงศ์ศรีถูกประยุกต์ใช้กับกระบวนการไฮโดรดีอัลคิเลชัน ซึ่งวิธีการออกแบบประกอบด้วย 8 ขั้นตอน ใน 5 ระดับ ซึ่งให้ความสนใจในเรื่องของการทำให้กระบวนการอยู่ในสภาวะคงตัวโดยการปรับอัตราการไหลของมวลสารตามการสะสมของมวลสารด้วยตัวจัดการตามเส้นทางของแต่ละสารและการใช้หลักการจัดการการรบกวน การออกแบบแสดงให้เห็นถึงความชัดเจนและการให้คำแนะนำวิธีการออกแบบโครงสร้างการควบคุมแบบทั้งกระบวนการ ประสิทธิภาพของการออกแบบโครงสร้างการควบคุมถูกประเมินโดยใช้การจำลองเชิงพลวัตที่มีความเข้มงวดและเปรียบเทียบกับโครงสร้างของลูเบน ผลคือประสิทธิภาพของโครงสร้างการควบคุมใหม่ดีกว่าโครงสร้างการควบคุมของลูเบนในเรื่องของการจัดการการรบกวน ความเร็วในการตอบสนอง และการรักษาความบริสุทธิ์ของผลิตภัณฑ์

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The Hydrodealkylation process (HDA) contains a complex flowsheet with material and energy recycles and many different unit operations. The behavior of this nonlinear petrochemical process is considered highly complex. Hence, the effective plantwide control structure is evidently crucial. In this present work, Wongsri’s plantwide control structure design procedure is applied to the HDA plant. The design procedure consists of 5 stages of 8 steps which emphasis on the establishment of a fixture plant by regulating material component flow rates using their quantifiers and handlers on their pathways and employing disturbance management policy. The design is illustrated through transparent and instructive steps resulting in workable plantwide control structures. The performances of the new designs are evaluated using the commercial rigorous dynamic simulator and compared with Luyben’s structure. The performances of the new structures are better than that of Luyben in terms of disturbance handing, response speed, and product purity maintaining.

Department:	Chemical Engineering	Student's Signature	....
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CHAPTER I

INTRODUCTIONS

This chapter presents the importance and reason of research, research objectives, scopes of research, contributions of research, research procedure, research framework, and research plan.

1.1 Importance and Reasons of Research

Nowadays, most industrial processes contain a complex flowsheet with several recycle streams of energy and material and fewer buffer tanks to reduce capital and operation costs. Therefore, the process is highly complex in dynamic behavior and very difficult in control. So the effective plantwide control design procedure is becoming important to develop the control loops in order to operate an entire process to meet its design objective. Several researchers have developed plantwide control structure design procedures, for example, the nine steps heuristic-based guideline by Luyben, Tyreus et al. (1997), self-optimizing control by (de Araújo, Govatsmark et al. 2007) and integrated framework of simulation and heuristics by Konda, Rangaiah et al. (2005). These procedures have their shortcomings for example; the Luyben's design steps are just guideline, no explicit instruction given. While Konda method uses a dynamic simulator to assist designer decision that often the same limitations of Luyben's procedure. The SOC method pairs a selected set of CVs and MVs that considered important. These procedures do not give capable plantwide perspective which is demanded in designing plantwide control structure.

1.2 Research Objectives

To design and evaluate the plantwide control structures of hydrodealkylation process by Wongsri (2012) procedure.

1.3 Scopes of Research

1 Simulation of the hydrodealkylation plant is performed by using Aspen HYSYS simulation.

2 Hydrodealklylation (HDA) process data is given by Luyben, Tyréus et al. (1998) and Luyben (2002)

3 Present new control structure design procedure of Wongsri (2012) for hydrodealklylation process.

4 Compare the new control structures design with the work proposed by Luyben (2002)

1.4 Contributions of Research

The procedure is detailed, instructive, simple, and easy to apply for novice in plantwide control structure design based on process approach.

1.5 Research Procedure

Procedure plants of this research are:

1. Study the plantwide control theory, the hydrodealkylation (HDA) process and concerned information.
2. Simulate steady-state and dynamic of hydrodealkylation (HDA) process by using commercial process simulator.
3. Simulate the reference control structure obtained from Luyben (2002)
4. Study the Wongsri's plantwide control structures and design procedure (2012).
5. Design new plantwide control structures of hydrodealkylation (HDA) process followed the Wongsri's procedure (2012).

6. Simulate the dynamic operation of hydrodealkylation (HDA) process with the new design control structures.
7. Evaluate the dynamic performances of the new control structures based on the external disturbance loads.
8. Analyze and discuss of the design and simulate results.
9. Conclude the research.

1.6 Research Framework

Chapter I: This chapter includes importance and reasons of research, research objectives, scopes of research, contributions of research, and research procedures.

Chapter II: present literature review related to plantwide control structures design procedure.

Chapter III: present the plantwide control structures design procedure.

Chapter IV: displays descriptions of hydrodealkylation process, reaction kinetics and parameter, process description.

Chapter V: described the plantwide control structures design for hydrodealkylation process including dynamic simulation results and discussions compared with control structures of Luyben (2002)

Chapter VI: presents conclusion and recommendations for research.

CHAPTER II

LITERATURE REVIEWS

Price and Georgakis (1993) studied A CSTR/Column process and a procedure of control structure design. They considered throughput, inventory and product quality control in process. The procedure is based on a tiered framework for plantwide control system design and is justified and supported by an extensive set of dynamic simulations. The best-performing structure is “self-consistent”.

Price, Lyman et al. (1994) present principle of a well-designed regulatory control system that it was good performance with management of product rate and regulation of the inventories within the process. They proposed guidelines for the development of production rate and inventory controls in process. The guidelines help a designer to ensure that these parts of the control structure are effectively designed before proceeding to the design of product quality controls. They applied it with the Eastman problem. They compared control structures that it was developed. Result, Simulation testing of control structure makes clear that this decision is an important contributor to the success of the resulting control structure.

Luyben (1994) presented “snowball” phenomenon. The snowball effect is a small change in load causing a very large change in the recycle flow rate around system. He studies numerical case of some typical recycle process. It is important with chemical process that it has recycle stream. They showed case study mathematic analysis. A mathematical analysis of the steady-state snowball effect in several recycle systems has been presented. Results confirm the case-study findings of Luyben that snowball problems can be prevented by using a control structure that fixes the flow rate of one stream somewhere in a liquid recycle loop. In processes with one recycle, the flow rate of the reactor effluent can be set. In processes with two or more recycle streams, the flow rate of each recycle can be fixed.

Luyben (1996) presented principle of finding the design control degrees of freedom. The design degrees of freedom are subtracting the number of equations from the number of variables. He suggested that counting the number of variables and equations are complex industrial processes are impractical and prone to error. He offers counting the number of control valves to find the control degrees of freedom. The number of design degrees of freedom is equal to the number of manipulated variables (the number of control valves in the process)

Ng and Stephanopoulos (1996) synthesized the plantwide control structures with specific control objectives that are consistent with the overall production goals and formulating control procedures. By they consider Hydrodealkylation process. Starting from the input-output level, they develop a control structure for the overall plant. Next, they move onto the next level. The control objectives and the control procedures are being refined. This procedure is repeated until all levels of the plant have been analyzed.

Luyben, Tyreus et al. (1997) presented the nine steps of plantwide control structure design. Plantwide control is the systems and procedure required to control an entire chemical plant. The procedure design of plantwide control consists energy management; production rate; product quality; operational, environmental and safety constraints; liquid-level and gas-pressure inventories; makeup of reactants; component balances; and economic or process optimization. It is applied with the vinyl acetate monomer process, the East-man plantwide-control process, and the HDA process.

Larsson and Skogestad (2000) presented the five tasks of plantwide control that consisted selection of controlled variables, manipulated variables, measurements, control configuration, and controller type, respectively. All this tasks separated two main parts. First, top-down consideration that Degree of freedom is considered in this part. The top-down analysis consist first and second task. The rest of three tasks were a bottom-down design. The first task is useful with steady-state economics. There is link between steady-state optimization and process control.

Skogestad (2004) improved the plantwide control procedure in 2000.By, they extended the procedure from five steps to eight step. This procedure contained an emphasis on degree of freedom analysis, selection of controlled variables, control system complexity, inventory control, and loss in performance by bottom-up design, respectively. The procedure is separated two main parts. First, top-down analysis contained four steps (including of operational objectives and consideration of degrees of freedom). Second, the rest of four steps were a bottom-down design for stabilizing control layer. The fifth and the sixth step are the analysis of control layer using a linear multivariable dynamic model. And, the third and the seventh step are the analysis of optimization layer using a nonlinear steady-state model. This procedure was the analysis based on the mathematical method. However, the mathematical procedure was complex.

Konda, Rangaiah et al. (2005) presented the novel plantwide control (PWC) methodologies that it is important with chemical process and recycle streams. They offered an integrated framework of simulation and heuristics. They applied the plantwide control with Hydrodealkylation (HDA) process. An analysis of results shows that the proposed framework builds synergies between the powers of both the simulation and the heuristics, thereby resulting in a practical PWC methodology that leads to a viable control system.

Konda and Rangaiah (2007) studied dynamic disturbance sensitivity (DDS) that it was then used to assess the performance of plantwide control (PWC) structures (CS1, CS2, CS3). They applied this assessment with three control structure of Hydrodealkylation process. The three control structure had difference of control loop as follows: CS1 used internal control procedure to control throughput (regulated temperature of plug flow reactor), CS2 used on-supply control procedure to control throughput and CS3used on-demand control procedure to control throughput. the rest of the control structure decisions are significantly different. Result, CS3 was observed to be exhibiting the poorest overall dynamic performance. The plantwide dynamic performance of CS2 is better than CS.

Husnil, Yeo et al. (2013) presented plant-wide control for the economic operation of modified single mixed refrigerant (MSMR) process. The NGL recovery and liquefaction units were integrated in the MSMR process to provide an efficient operation. Steady-state optimality analysis is conducted to determine the variable maintaining the economic efficiency of MSMR process. The results showed that the flow rate ratio of heavy and light mixed refrigerant (HK/LK ratio) provides a self-optimizing controlled variable and maintains the MSMR optimality. The plantwide design procedure carries on six steps. The first step is the formulation control objectives and constraints. The second step is to determine the degrees of freedom for identifying the manipulated variables. The third step is to determine the appropriate manipulated variable by sensitivity analysis that gives the relation of input-output. The fourth step is adding the necessary regulatory control loops. The fifth step is the process optimization as well as selecting the optimizing-controlled variable. The sixth step is evaluating the control structures in terms of both dynamic response and steady-state. The control structure with the HK/LK ratio loop provided better performance than the system with the HK flow control loop.

Luppi, Zumoffen et al. (2013) presented decentralized plantwide control strategy for large-scale processes of pulp mill benchmark. Plantwide control strategy consists of five steps, namely define the stabilizing control loops (stabilizing the plant by establishing control loop of level or pressure in tanks and vessels), obtain a reduced process model (estimating steady-state gains and simplified dynamic linear models), select the CVs set together with the optimal CVs–MVs pairing, determine the algorithm and tuning parameters for each control loop, and evaluate the dynamic performance of the designed strategy. The proposed methodology considers tools as the normalized relative gain array (NRGA) and the Hungarian Algorithm (HA). The approach is based on steady-state information, and tries to reduce the use of heuristic considerations. The dynamic simulation results show stable and acceptable dynamic performance under several disturbances and setpoint changes.

CHAPTER III

PLANTWIDE CONTROL STRUCTURES DESIGN PRINCIPLE

Plantwide process control is an important approach that guides control system design for a chemical plant consisting of multiple processing units, interconnecting with each other, such as reactors, furnaces, absorbers, heat exchangers, and distillation columns. Given a complex, integrated processand a diverse assortment of equipment. Thus, in order to achieve the design objectives and plant safety, it is necessary to understand its logic and strategies. In this chapter we present the plantwide control fundamentals and plantwide control structures design procedure.

3.1 Plantwide Control Fundamentals

A typical chemical plant has a mixture of multiple units connected both in series and in parallel. Many measurements and control loops have been used. Plantwide control does not account only for the tuning or behavior of each loops, but for the control philosophy of the overall plant with emphasis on the structural decisions. The structural decisions include the selection/placement of manipulators and measurements as well as the decomposition of the overall problem into smaller sub problems. In conclusion, plantwide control that meant control structure design for complete chemical plant.

The design of a plantwide control strategy becomes a more important part of the overall process control design problem. The interrelation of the plantwide control strategy with the process chemistry and economics requires both control theory and also process knowledge. It has become apparent that the design of plantwide control strategies involves not only the development and application of process control theory, but also in a more fundamental sense, the development of a methodology one uses to approach the plantwide control problem. The need for a plantwide control perspective arises from three important features of integrated processes: the effects of material recycle, of chemical component inventories and of energy integration. However, we did not describe how we arrived at these strategies, and many of our choices may seem mysterious at this point.

The subject of plantwide control deals with unit-to-unit interactions through the choice of measured and manipulated variables in each unit and the selection of a control strategy-namely, how to pair controlled and manipulated variables in individual loops, where to use multi-loop controllers, where to use multivariable controllers. The goals for an effective plantwide process control system include:

1. Safe and smooth process operation.
2. Tight control of product quality in the face off disturbance.
3. Avoidance of unsafe process conditions.
4. Automatic running of control system, requiring minimal operator attention.
5. Rapid rate and product quality transitions.
6. Zero unexpected environmental releases.

3.2 The main function of control system

- To reject disturbance

It is main function in the installation of process control system. The external disturbance is uncertain so the operator cannot monitor the changing in process. As a result, the control system must be installed to follow the process changing and manipulate the process variables to compensate for the external disturbance factors.

- To maintain stability

Every control system must guarantee first the stability. The control system is set to improve the process stability for ensure of product quality and safety to equipment of process and plant.

- To keep the highest performance of process

Besides the rejection of disturbance and maintaining stability, the control system can achieve the great profit because it losses less energy and raw materials during the operating. Furthermore the product will meet the desired specification and have production rate.

3.3 Luyben plantwide control structure design procedure.

The design procedure of Luyben is carried out in nine steps, which contented the two fundamental of the overall conservation of energy and mass. The procedure also accounts for nonconserved entities within a plant such as chemical components (produced and consumed) and entropy (produced). Each of steps is as follows:

Step 1: Establish Control Objectives. *Assess the steady-state design and dynamic control objectives for the process.* This is probably the most important aspect of the problem because different criteria lead to different control structures. These objectives include reactor and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operating conditions.

Step 2: Determine Control Degrees of Freedom. *Count the number of control valves available.* This is the number of degrees of freedom for control, which is the number of variables that can be controlled. The valves must be legitimate (flow through a liquid-filled line can be regulated by only one control valve).

Step 3: Establish Energy Management System. *Make sure that energy disturbances do not propagate throughout the process by transferring the variability to the plant utility system.* We use the term “energy management” to describe two functions:

1. Provide a control system that removes exothermic heats of reaction from the process. If heat is not removed to utilities directly at the reactor, then it can be used elsewhere in the process by other unit operations. This heat, however, must ultimately be dissipated to utilities.

2. If heat integration does title for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relation occur between process streams, then the second function of energy management is to provide a control system that prevents the propagation of thermal disturbances and ensures that the exothermic reactor heat is dissipated and not recycled. Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom for control.

Step 4: Set Production Rate. *Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate.* Determine what valve will be used to set throughput. Often design constraints require that production be set at a certain point. An upstream process may establish the feed flow sent to the plant. A downstream process may require on-demand production, which fixes the product flow rate from the plant. If no constraint applies, then we select the valve that provides smooth and stable production rate transitions and rejects disturbances.

Step 5: Control Product Quality and Handle Safety, Operational, and Environmental Constraints. *Select the "best" valves to control each of the product quality, safety and environmental variables.* We want tight control of these important quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and dead times a large steady-state gains. The former give small closed-loop time constants, and the latter prevents problems with the range ability of the manipulated variable (control-valve saturation). The magnitudes of various flow rates also come into consideration.

Step 6: Control Inventories (Pressures and Levels) and Fix a Flow in Every Recycle Loop. *Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.* Determine the valve to control each inventory variable. These variables include all liquid levels (except for surge volume in certain liquid recycle streams) and gas pressures. In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows that can occur if all flows in the recycle loop are controlled by levels. Two benefits result from this flow control strategy. First, the plant's separation section is not subjected to large load disturbances. Second, consideration must be given to alternative fresh reactant makeup control strategies rather than flow control. In a dynamic sense, level controlling all flows in a recycle loop is a case of recycling disturbances and should be avoided.

Step 7: Check Component Balances. *Identify how chemical components enter, leave, and are generated or consumed in the process.* Ensure that the overall component balances for each chemical species can be satisfied either through reaction or exit streams by accounting for the component's composition or inventory at some point in the process. Component balances can often be quite subtle. They depend upon the specific kinetics and reaction paths in the system. They often affect what variable can be used to set production rate or rate in the reactor.

Step 8: Control Individual Unit Operations. *Establish the control loops necessary to operate each of the individual unit operations.* Many effective control schemes have been established over the years for chemical units Shinskey (1988). For example, Liquid solvent feed flow to an absorber is controlled as some ratio to the gas feed.

Step 9: Optimize Economics or Improve Dynamic Controllability. *Establish the best way to use the remaining control degrees of freedom.* After satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and set points in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

3.4 Wongsri'sPlantwide Control Design Procedure

Wongsri (2012) presented the new plantwide control design procedure carried out in five stages with eight steps, the major steps deal with plant level design; establishing a fixture plant. The component balances are accounted by identifying the material quantifiers that indicate the amounts of the components and using their handlers to control them. The disturbances entering into the process must be directed by using the proposed material and energy disturbance management for avoiding disturbance propagation throughout the plant. Each step is as follows:

Stage 1.Plant Information and Analysis

Step 1: Gather of relevant plant information and control objectives including constraints for control. It is necessary to obtain all information relevant to process control, such as product quality, production rate, smooth operation, process and equipment constraints, plant safety, and environmental regulations.

Step 2: Plant analysis. Several tasks to assist design decision in Step 3 are:

2.1 Control degree of freedom (CDOF). Each single independent stream, physical or virtual, material or energy, must have a handle or one control degree of freedom.

2.2 Heat pathways. The first pathway is heat generated by exothermic reactions and flows out to the environment. A second pathway carries heat from utilities into the process and to the environment. The third pathway is internal to the process. The fourth pathway is accounted for the enthalpies entered and left the plant. Those first three ways proposed by Luyben, Tyréus et al. (1998)

2.3 Material pathways. The pathway is the flow path of a component from an entry point or an originated point to an exit point or an end point.

2.4 Material quantifier. A material quantifier is the place indicating the significant amounts of a chemical component (or a group of components) in the plant which can be handled quite readily by regulating at their handlers. In the case that the quantifier is a flow, it is, but not necessarily, the place that has the highest gain of component flow is the total flow.

2.5 Reaction section. It is necessary to obtain required information for control design of reactor section. In general, what kind of controlled variables used to regulate the reaction yield and where to measure such controlled variables? What is the best control strategy and all? If feeds and recycled streams are fixed, the only places that the material (total or component) flow rates altered are a reactor and also a separator.

2.6 Separation section. The appropriate directions of disturbances are analyzed and specified. A surplus disturbance, D+ is the surplus deviation of the mass load from the nominal load and the deficit disturbance, D- is the deficit deviation of the mass load. The paths of D^+ and D^- in the separation section are analyzed and then designed in order to shift surplus or deficit mass loads to the desired targets to achieve the plant operation objective, e.g. maintaining product quality and avoiding disturbance propagation and recycling. The paths of D+ and D-in the separation section must be shifted to the proper exits. In the case that there is no proper exit for D^+ or shifting it through available exits will disturb the product quality, recycling it would be allowed.

Next, a good location of temperature control is the tray with the largest changes in the temperature from the initial steady state by changing of composition, total flow, temperature, and component flow during keeping the reboiler heat duty and reflux flow or reflux ratio or reflux fraction or boil up ratio constants.

2.7 Production rate control. Throughput changes must be achieved by altering reactor condition. However, reactor temperature, reactant concentration, reactor holdup would be somehow limited.

The production or throughput rate change by increasing/decreasing feed rate should be accompanied by adjusting recycle flow accordingly.

Mode of operation: On-supply, On-demand, and on-internal.

The mode of operation is dictated by a business objective and the mode of operation, such as on-supply (fixed feed rate), on-demand (fixed product rate), and on-internal (fixed internal flow rate) based on throughput manipulator (TPM) decision Price, Lyman et al. (1994). For on-internal control scheme, the throughput manipulator (where the production rate is set) is located inside the plant downstream of this location (normally at the bottleneck), the plant has to process whatever comes in, and upstream of this location the plant has to produce the desired quantity. The selection of on-supply, on-demand or on-internal should depend on the completeness of total control of components.

In some processes, the separation section is placed before the reactor section, there are two locations to fix the material flows into the process: at the entrances of the reactor section or of the separation section. In the case that the reactor influent is fixed, the quantifiers (inventories) prior to this point must be controlled as ‘on-demand production’.

Stage 2.Fixture Plant and Disturbance Management

This stage is a major design stage; plant control structure is created at plant level in two steps: Step 3 and Step 4. The plant control loop design procedure presented in this paper is explicit and systematic while the Luyben design procedure has some shortcomings, Konda, Rangaiah et al. (2005). There are two objectives: the plant nominal material balance is maintained; the heat and material disturbances must be rejected to the nearest exits or directed to less significant streams.

Step 3: Establish fixture plant. The principal idea of establishing a fixture plant is first to have a material-balanced in the plant by controlling each component at its quantifier, i.e. fixture point.

3.1 Keep the materials entered and/or reentered fixed. A fresh feed and/or a combined stream of make-up feed and recycle stream must be kept constant to maintain the plant inventory by flow/composition controls. A recycle flow should not be fixed. This leaves the recycle flow free to be adjusted; one degree of freedom is restored to the plantwide control design process. If the composition of the recycled stream differs from the fresh feed stream significantly, each recycled stream may be flow-controlled. However, in the case that the composition of the recycled reactant can be measured, the composition of the combined stream is controlled to keep the combined reactant flow in check.

3.2 Adjust the flow of exit material streams (products, byproducts, and inert) according to their accumulations. If the flows of the products are controlled (mode of operation is on-demand) the quantifiers of the products, e.g. levels of reflux drums indicating the surplus/deficit will be used to control the feeds.

3.3 Handle the material that not leaving the process. The reactor is the logical place to regulate a component fed or formed in the process and not leaving the process. If there is only one reactor and there is more than one component that not leaving the process, their kinetics must be similar, e.g. increasing the reactor temperature reduces or increases the amount of both components. Handlers of these components must be identified. If their kinetics are not compatible, we must provide exits for the incompatible components.

3.4 Control the amount of the rest of the component at their quantifiers. This step assures the rest of component inventory is regulated from a plantwide perspective. Setting the control at the specified quantifiers is like providing coordination over different sections of the plant to ensure that the rate of accumulation of each component in the overall process is zero.

3.5 Maintain the production rate.

3.5.1 Consume the limiting reactant. Determine the most appropriate manipulate variable to control the limiting reactant for the economic reason, i.e. the reactor temperature, the reactor pressure, or the reactor holdup.

3.5.2 Regulate the production rate. The product rate can be regulated through 3.2.1. If this is done and the production rate does not reach the objective or the production demand, the limiting reactant feed rate must be increased. However, the design constraints may limit this strategy concerning increasing the reactant feed rate.

Step 4: Disturbance management for quality control. The nominal conditions of process streams are maintained by specifying the disturbance shifting directions. The principles of disturbance management are following:

4.1 Heat disturbance management. The heat disturbance is divided into two categories. Heat disturbance of category 1 (HDC1) is the heat disturbance that does not instantly effect on the qualities of process streams, such as heat disturbance in a process stream toward a heater, a cooler, or a process-to-process heat exchanger. Heat disturbance of category 2 (HDC2) is the heat disturbance that will affect the process stream qualities where an additional phase is created or introduced, and the equilibrium is altered; or where chemical reactions are undergoing, such as separators and reactors.

4.1.1 Direct the HDC1 to the environment via the next and nearest exit points, usually heaters or coolers, to keep the thermal conditions of the process stream fixed.

4.1.2 Direct the HDC2 to less significant output streams of separators. This rule is generally apt to a separator using heat as a separating agent.

4.2 Material disturbances management. The configurations of the control loops are decided base on the desired material pathways. As in the case of heat disturbance management, we should direct the material disturbances to the environment via the next and nearest exit points to avoid disturbance recycling and propagation.

Many industrial distillation columns use some type of single-end temperature control because of its simplicity and low maintenance cost. This step presents a procedure to determine the control structure of a distillation column with desired material disturbances (D+ and D-) following step 2.6 by using a dynamics process simulator for various single-end control structures, namely constant reflux flow (R), constant reflux ratio (RR), constant reflux-to-feed ratio (R/F), constant reflux fraction (R/(R+D)), constant boil-up ratio (V/B). Several kinds of material disturbances in feed, such as temperature, flow rate, composition, and component flow rate are generated to test the disturbance shifting ability of these control structures. In addition, the principals of the material disturbance management are as follows:

4.2.1 Direct the material disturbances of byproducts, inerts, and unconverted raw materials to the environment via the next and nearest exit points.

4.2.2 For the main products, the deficit disturbances should follow Rule 4.2.1. However, the main product surplus disturbances should be allowed to propagate to their exits.

4.2.3 MDM rule for the recycled streams: their surplus disturbances of unreacted raw materials are permitted, however, their deficit disturbances must not be allowed to economize the make-ups.

The selection of the distillation control structures is carried out in two steps: preliminary screening using steady-state simulation and the selected candidates are further tested by rigorous dynamic simulation.

Stage 3.Unit Level Design

Control loop design at this stage is solely based on individual unit operations.

Step 5: Design the rest of the control loops. Normally, the rest of the control loops is inventory loops which are self-regulating and less crucial. They can be designed using unit-based approach.

5.1 Design the control loops for the remaining control variables, i.e. the rest of the inventory.

5.2 Adding simple enhanced controls, e.g. cascade, feed forward controls.

Stage 4.Energy Management and Optimization

The supplementary design activities involve heat exchanger network design and control, and plant operation and design optimization.

Step 6: Energy management via heat exchanger networks.

In the case that the exothermic heat of reaction is large enough to heat some process cold streams, i.e. potential heat exchanger networks or alternative heat integrated processes (HIPs) exist, a heat exchanger network must be designed and a HEN must be resilient, i.e. delivering the exchange streams to their target temperature. The resilient heat exchanger network with specified load disturbances can be designed using Wongsri's method. The design of a control system that prevents the propagation of the heat disturbance of Wongsri and Hermawan (2005) is recommended.

Step 7: Optimize economics or improve control performance. The design and control issue remains an open research area regarding the plantwide control design, so the opportunity to alter the process design is possible.

Stage5. Design Validation.

The validation of the design control structures using rigorous nonlinear simulation is inevitable; whatever may be the design procedure.

Step 8: Validate the designed control structures by rigorous dynamic simulation.

The measures would be costs, raw material and energy consumptions, control performances of the total plant or some selected loops, etc. Expected disturbances must be listed to perform the disturbance test on the plant with designed control structures.

Table 3. 1 Plantwide control design procedure of Wongsri and Luyben in comparison

Plantwide Control Design Procedure	
Wongsri (2012)	**Luyben et al., (1997)**
Stage 1. Plant Information and Analysis	
Step 1: Gather of relevant plant information and control objectives, including constraints for control	Step 1: Establish Control Objectives
Step 2: Plant analysis	
2.1 Control degree of freedom (CDOF) - Each single independent stream must have a handle or one control degree of freedom	Step 2: Determine Control Degrees of Freedom - Count the number of control valves available
2.2 Heat pathways	
2.3 Material pathways	
2.4 Material quantifier	
2.5 Reaction section	
2.6 Separation section - The appropriate directions of disturbances are analyzed and specified - Sensitivity test for selecting temperature control tray location	
2.7 Production rate control - Mode of operation: On-supply, On-demand, and on-internal	Step 4: Set Production Rate - Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate

Table 3.1 Plantwide control design procedure of Wongsri and Luyben in comparison.(Continued)

Plantwide Control Design Procedure	
Wongsri (2012)	**Luyben et al., (1997)**
Stage 2. Fixture Plant and Disturbance Management	
Step 3: Establish fixture plant 3.1 Keep the materials entered and/or reentered fixed 3.2 Adjust the flow of exit material streams (products, byproducts, and inert) according to their accumulations 3.3 Handle the material that's not leaving the process 3.4 Control the amount of the rest of the component at their quantifiers 3.5 Maintain the production rate - 3.5.1 Consume the limiting reactant - 3.5.2 Maintain the production rate	Step 7: Check Component Balances - Identify how chemical components enter, leave, and are generated or consumed in the process
Step 4: Disturbance management for quality control 4.1 Heat disturbance management - The heat disturbance that does not instantly affect on the qualities of process streams	Step 3: Establish Energy Management System - Transferring the variability to the plant utility system

Table 3.1 Plantwide control design procedure of Wongsri and Luyben in comparison.(Continued)

Plantwide Control Design Procedure	
Wongsri (2012)	**Luyben et al., (1997)**
- The heat disturbance that will affect the process stream qualities 4.2 Material disturbance management - The configurations of the control loop are decided based on the desired material pathways	Step 5: Control Product Quality and Handle Safety, Operational, and Environmental Constraints - Select the best valves to control each of the product quality, safety and environmental variables
Stage 3. Unit Level Design Step 5: Design the rest of the control loops 5.1 Design the control loops for the remaining control variables, i.e. the rest of the inventory 5.2 Adding simple enhanced controls, e.g. cascade, feed forward controls	Step 6: Control Inventories (Pressures and Levels) and Fix a Flow in Every Recycle Loop Step 8: Control Individual Unit Operations
Stage 4. Energy Management and Optimization Step 6: Energy management via heat exchanger networks Step 7: Optimize economics or improve control performance	Step 9: Optimize Economic or Improve Dynamic Controllability
Stage 5. Design Validation Step 8: Validate the designed control structures by rigorous dynamic simulation	

CHAPTER IV

HYDRODEALKYLION (HDA) PROCESS

4.1 HDA Process Description.

In the HDA process, there are two fresh reactant feed streams: liquid toluene (pure) and gaseous hydrogen (97% hydrogen and 3% methane) are mixed with recycled toluene and hydrogen. This reactant stream is heated in a feed-effluent heat exchanger (FEHE) using the reactor effluent stream and then to the reaction temperature in a furnace before being fed to the plug flow reactor. Two main reactions taking place inside this reactor are

$$C_7H_8 + H_2 \longrightarrow C_6H_6 + CH_4$$

$$\text{Toluene + Hydrogen} \longrightarrow \text{Benzene + Methane}$$

$$2C_6H_6 \longleftrightarrow C_{12}H_{10} + H_2$$

$$\text{2Benzene} \longleftrightarrow \text{Diphenyl +Hydrogen}$$

In this process, the reaction kinetic rate express are function of the partial pressure (psia), with Arrhenius temperature dependence. The reaction kinematic equations used in hydrodealklation process are:

$$R_1 = 3.6858 \times 10^6 \exp(-25{,}616/T)\, P_T P_H^2$$

$$R_2 = 5.987 \times 10^4 \exp(-25{,}616/T)\, P_B^2 - 2.553 \times 10^5 \exp(-25{,}616/T)\, P_D P_H$$

Where

R_1 and R_2= reaction rate (lb-mole/min.ft^3)

T = temperature (Kenlvin)

P_i = partial pressure (psia)

Two vapor-phase reactions are considered to generate benzene, methane, and diphenyl from reactants toluene and hydrogen. The reactor is adiabatic and must be run with an excess of hydrogen to prevent coking. The reactor effluent is quenched with liquid from the separator to prevent fouling in the FEHE. The gas stream from the overhead of the separator comprised of unconverted hydrogen plus methane is purged (to avoid accumulation of methane within the process) and then compressed and recycled back to the reactor. The liquid from the separator is separated in the separation section consisting of three distillation columns. The first column called stabilizer column removes hydrogen and methane as the overhead product. Product benzene leaves as the distillate from the second column. Finally, in the third or recycle column, toluene is separated from diphenyl, as the distillate, and recycled back to the reactor.

4.2 Steady-State Modeling.

In this study, improved Peng Robinson (PR) equation of state is selected for property estimation. The steady-state simulation model of the HDA process has been developed according to the flow-sheet topology (Figure 4.1) and the process information from Douglas (1988) and Luyben, Tyréus et al. (1998)

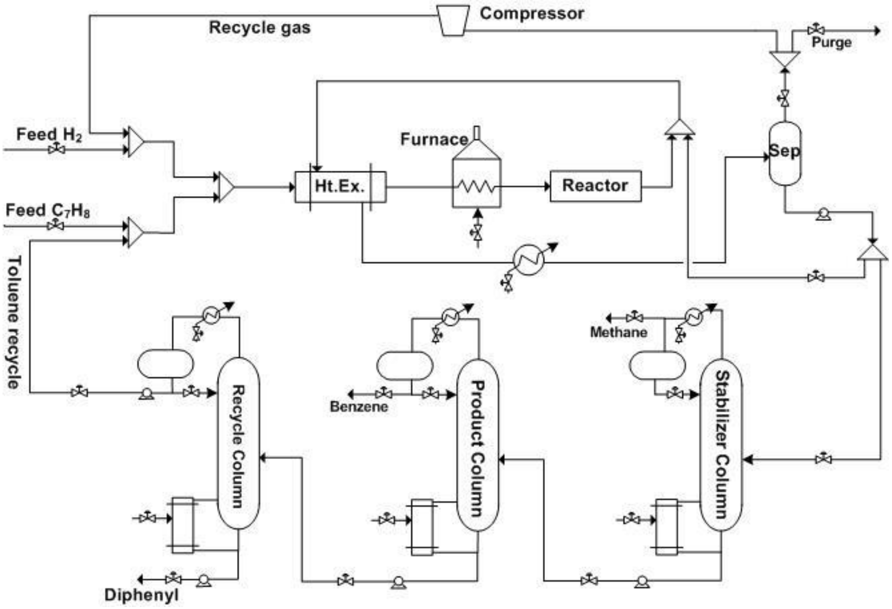


Figure 4. 1 Hydrodealkylation (HDA) process flowsheet

CHAPTER V

CONTROL STRUCTURES DESIGN

In this chapter, we describe a hierarchical design procedure that can be used to develop multiple loop and multivariable measurement and control strategies for plant wide control system. The procedure assists the engineer in determining how to choose the best controlled, manipulated, and measured variables in the plant and how to select appropriate multiple loop control structure. The proposed design procedure is based on the hierarchy of process control activities described in chapter 3. It is important to realize that the design of plantwide control systems an art as well as a science. Typically, more than one design will be satisfactory: thus, there is no single solution to the design problem. Furthermore, a design procedure generally involves iteration of individual step until a satisfactory design result. Thus, the application of a systematic design procedure produces preliminary designs that are subject to further exploration and refinement.

5.1 Control Structure Design Procedures of hydrodealkylatio process.

Plantwide control structure design procedure of Wongsri (2012) is applied to the hydrodealkylation process. It consist eight steps of five stages of plantwide control structure design. It is developed after many years of work and research in the field of process control and process design. The design of a plantwide control system consists as follows:

Step 1: Gather plant information and control objectives.

(i) Plant topology and unit operations: 9 standard unit operations. They are reactor, furnace, vapor-liquid separator, recycle compressor, cooler, heat exchanger, and 3 distillation columns.

(ii) Production rate: 267 lb-mole of benzene/hr

(iii) Product quality: benzene purity 99.96%

(iv) Process constraints:

- The temperature at the reactor inlet is 1150 °F.

- The ratio of hydrogen to toluene has to be at least 5 at the inlet to provide a thermal sink.
- The temperature at the reactor effluent must not exceed 1300 °F to avoid coking.
- The outlet stream from the reactor must be quenched to 1150 °F.

Step 2: Plant Analysis.

2.1 Control degree of freedom (CDOF). List manipulated variables (control degree of freedom, CDOF). There are 23 independent streams indicated in Table 5.1. The CDOF is determined to be 23.

Table 5.1 The control degree of freedom for the hydrodealkylation process.

Independent Streams	Quantity	CDOF
Feed flows	2	2
Purge flow	1	1
Separator top flow	1	1
Separator bottom flow	1	1
Cooler duty	1	1
Liquid quench flow	1	1
Furnace duty	1	1
Distillate flows, Bottom flows, Reflux flows, Reboiler duties, Condenser duties.	3	15
Total degrees of freedom	23	

2.2 Heat pathways.

From the plantwide overall image, four different "heat pathways" are introduced in the process. The first pathway is heat of exothermic reactions ($9.066x10^6$kJ/h). The second pathway is send from the utilities into the process (-$4.57x10^6$kJ/h) MW). The third pathway is internal to the process ($6.652x10^7$kJ/h and $1.232x10^6$kJ/h). The fourth pathway is the enthalpies entered ($1.362x10^6$kJ/h) and left the plant (1.810×10^6 kJ/h) via process stream. Heat pathways is presented by Figure5.1

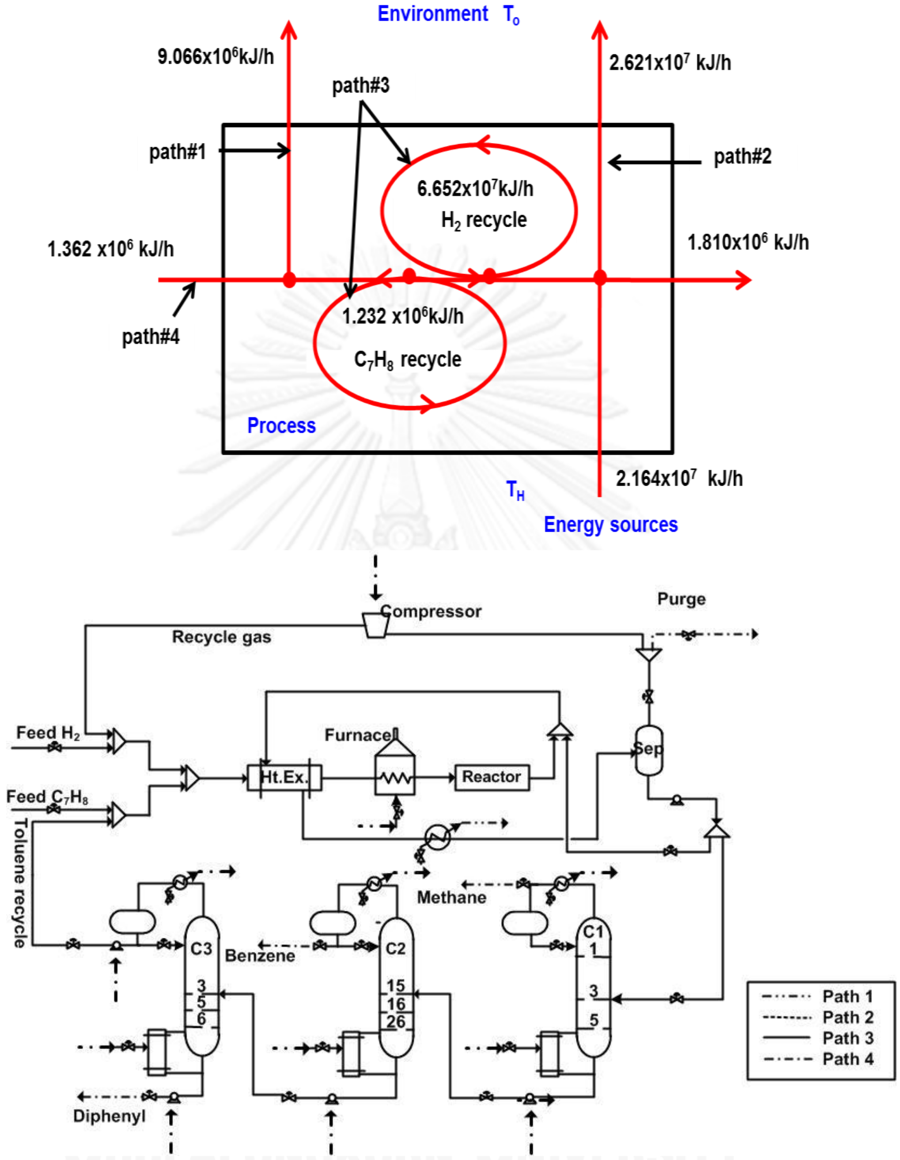


Figure 5.1 Heat pathways

2.3. Material pathway.

The material pathways, showing the flow paths of single of group of components, are useful to identify the material quantifier as discussed in step 2.4. The material pathways are depicted in Table 5.3

2.4. Material quantifier

The material quantifiers are useful to design control loops for component balance as discussed in step 3. The quantifiers and their handlers are shown in Table 5.2 and Table 5.3

The places indicating the significant amounts of hydrogen or its quantifier is the combined flow rate of hydrogen fresh feed and recycle gas. Similarly, the quantifier of toluene is the combined flow rate of toluene fresh feed and recycles toluene. Thirdly, the methane quantifier is the pressure of the top separator stream. Then, the quantifiers for benzene and diphenyl are level of reflux drum in column C2 and level of reboiler in column C3 respectively.

Table 5.2 Quantifiers and handlers of components

Component	Quantifier	Handler
H_2	Combined feed H_2 flow rate	Fresh feed H_2 flow rate
CH_4	Separator top stream pressure	Purge flow rate
C_6H_6	Column C2 reflux drum level	C2 distillate flow rate
C_7H_8	Combined feed C_7H_8 flow rate	Fresh feed C_7H_8 flow rate
$C_{12}H_{10}$	Column C3 base level	C3 bottoms flow rate

Table 5.3 Material pathways, Material quantifiers and Handles of each component

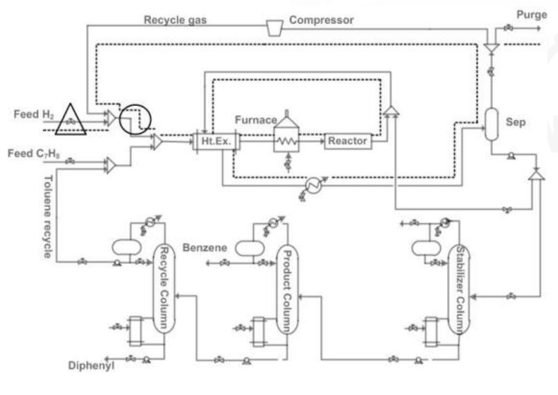
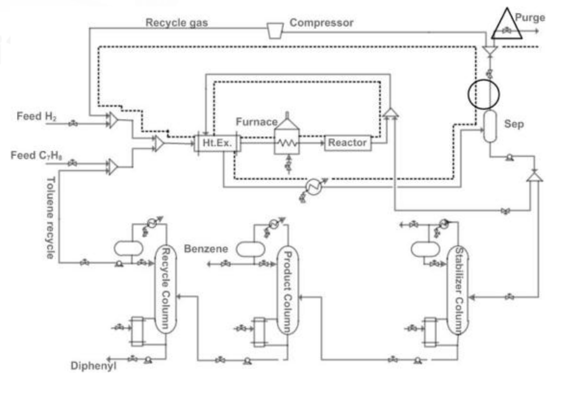
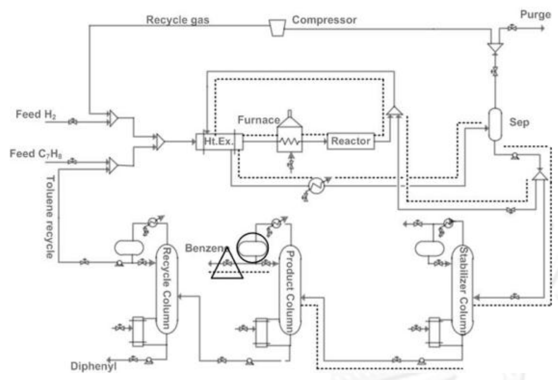
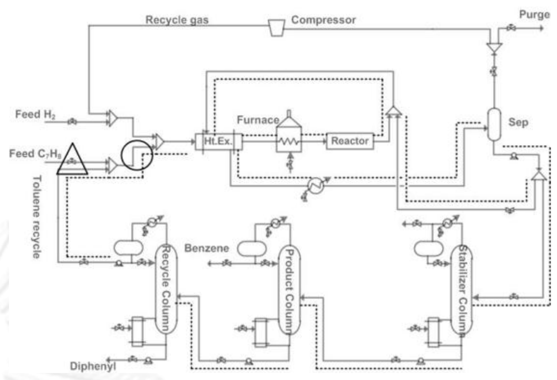
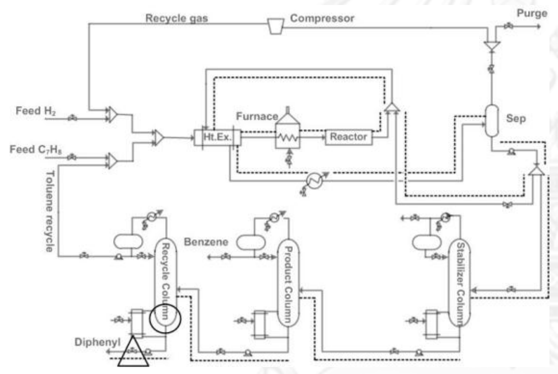
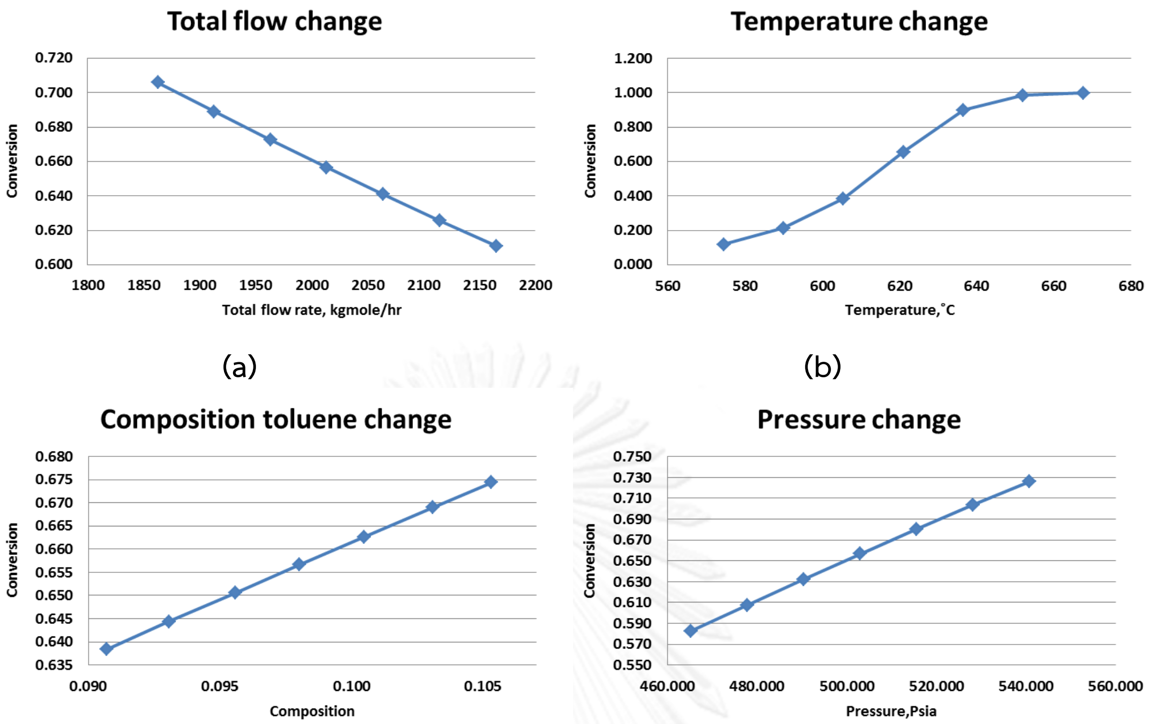
	
H_2 pathway	CH_4 pathway

Table 5.3 Material pathways, Material quantifiers and Handles of each component. (Continue)

	
C_6H_6 pathway	C_7H_8 pathway
	— **Material pathway** ○ **Quantifiers** △ **Handlers**
$C_{12}H_{10}$ pathway	Legend

2.5. Reaction section.

For the HDA process, inlet temperature is a dominant variable since the reaction is exothermic having reasonably high activation energy and the reactor is plug flow. However, we would like to see how other variables affect the reaction. Using steady-state simulation, we change component feed flow rate, total feed flow rate, and temperature in reactor inlet stream and the results are shown in Figure 5.2



(a) (b)

(c) (d)

Figure 5.2 Reactor analysis

2.6. Separation section analysis.

In this step, desire directions of disturbances are determined and column temperature sensibilities are examined. Column C1 separates CH_4 and C_6H_6. The plus and minus disturbances of CH_4 and minus disturbance of C_6H_6 disturbance (CH_4+, CH_4- and C_6H_6-) must be shifted to the top of C1 column in order to maintain the quality of C1 bottoms. C_6H_6+ must be shift to C1 bottoms. Next, column C2 separates C_6H_6 and C_7H_8. C_6H_6+ must be shifted to C2 top product for obvious reason. Both C_7H_8+ and C_7H_8- must be shifted to the bottom of C2 to maintain the benzene product purity. Lastly, Column C3 separates C_7H_8 and $C_{12}H_{10}$. C_7H_8+ goes to the top of C3 since we do not want to lose it. C_7H_8-, $C_{12}H_{10}$+ and $C_{12}H_{10}$ destine to C3 bottom to keep the recycle stream undisturbed. Table 5.4 summarizes the analysis and Figure 5.3 shows the directions of material disturbances.

Table 5.4 Surplus and deficit disturbances shifting direction

Distillation column	Distillate	Bottoms
C1	$CH_4\pm$, C_6H_6-	C_6H_6+
C2	C_6H_6+	$C_7H_8\pm$
C3	C_7H_8+	C_7H_8-, $C_{12}H_{10}\pm$

In the second test, selecting Temperature/Composition Control Tray Location has several methods. They are “Open-end sensibility” and “Effect of disturbance”. To select the temperature control tray location of each column, the tray sensitivities to important disturbances are performed. Two methods are tested in steady state mode.

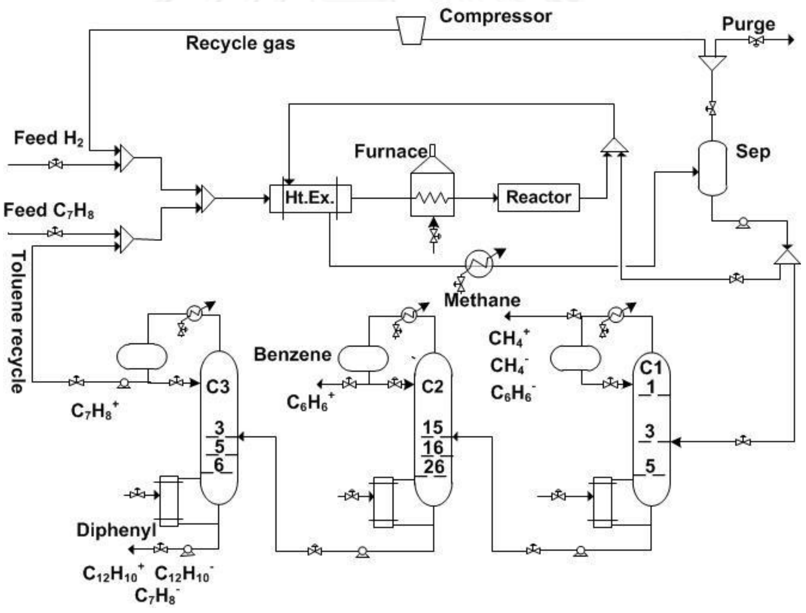


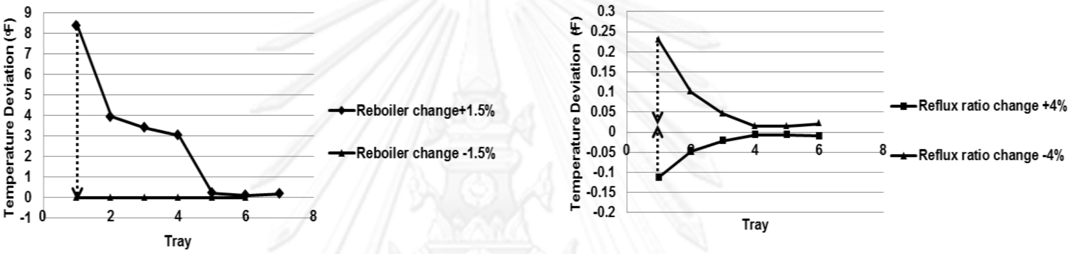
Figure 5.3 The directions of material disturbances of hydrodealkylation process.

In open-end sensibility method, the reflux ratio (RR) and feed condition to column are controlled. The heat duty in reboiler must be increased/decrease. In the end, the temperature profile of column more deviated from steady-state must be analyzed.

In Effect of disturbance method, the sensitivity test is suggested to be done on the changing of composition, total flow rate, temperature, and component flow rate. Nevertheless, the reboiler duty and reflux flow rate or reflux ratio are controlled.

Column C1

Figure 5.4 presents the temperature profile of column 1 after test increase/deceases reboiler duty by change ±1.5% to column 1. Figure 5.4 (a), the temperature profile of +1.5% of reboiler duty the largest changes in temperature from the initial steady state at 1st tray similar to the temperature profile of reflux ratio change ±4 % in Figure 5.4 (b). In summary, the 1st tray may be good location to control temperature at column C1.



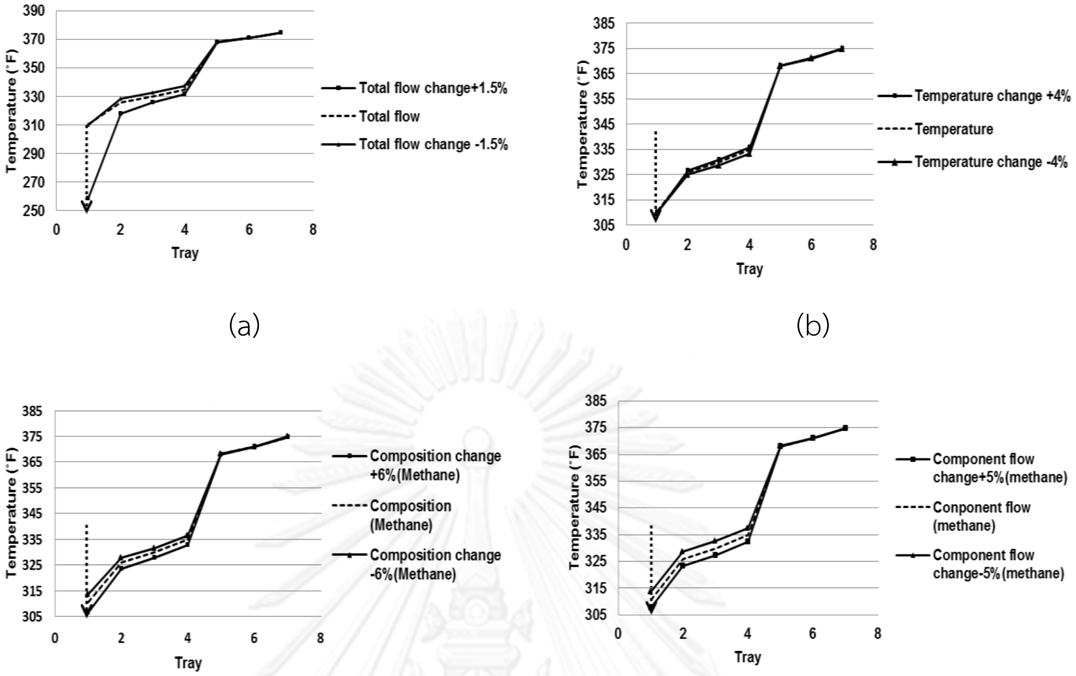
(a) (b)

Figure 5.4 The open loop sensitivity test of column C1

(a) Reboiler change

(b) Reflux ratio change.

Figure 5.5 presents the different temperature profiles of the column C1 by Effect of disturbance method. Form the steady state mode. We tested column C1 by test disturbance 4 case (Total flow rate test ±1.5%, Temperature test ±4%, Composition test ±6% and Component flow rate test ±5%). Figure 5.5 (a) show temperature profile of total flow rate test ±1.5%. It is the largest changes in temperature from the initial steady-state at 1st tray like Figure 5.5 (c) and Figure 5.5 (d). In figure 5.5 (b), ±4% of temperature test is not quite different. In conclusion, The 1st tray of column C1 has appropriate management temperature for maintain composition.



(a) (b)

(c) (d)

Figure 5.5 Effect of disturbances on the steady-state temperature profiles of the column C1

(a) Total flow rate test

(b)Temperature test

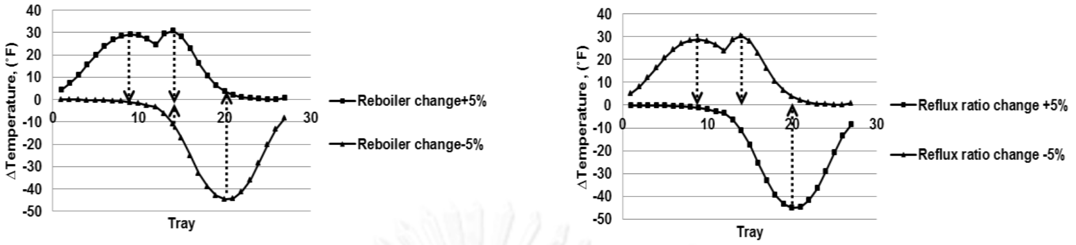
(c) Methane composition test

(d) Component flow rate test

From analytical sensibility test of column C1, 1st tray appropriate controlled the temperature in column C1 since 1st tray deviate the temperature more than the other trays in 2 analyses and it must be used to control impurity of methane in bottom flow rate of column C1.

Column C2

Figure 5.6 presents the temperature profile of change reflux ratio and reboiler duty by Open Loop Sensibility Method. ±5% of reboiler duty and ±5% of reflux flow rate is tested. As a consequence, the temperature profiles of ±5% of reboiler duty have the largest deviated from steady state at 8th, 16th and 23th trays similar to reflux ratio test ±5%. Therefore, three spots of temperature profile can be used in control temperature of column C2.



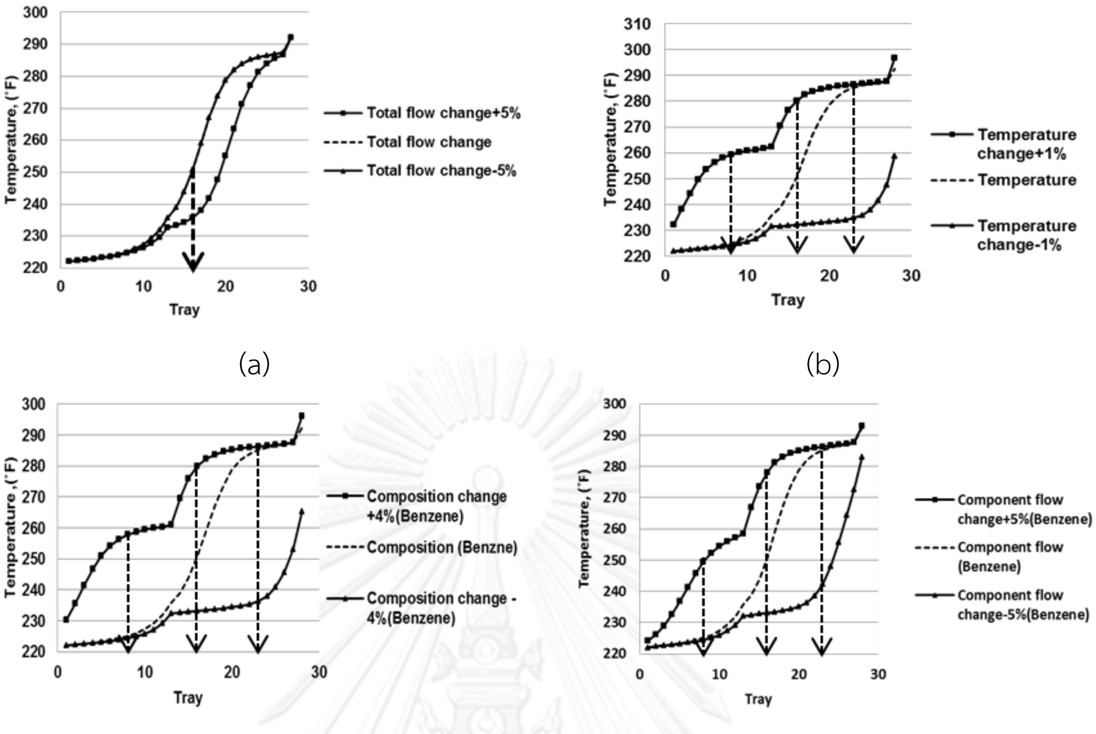
(a) (b)

Figure 5.6 The open loop sensitivity of column C2

(a) Reboiler duty test

(b) Reflux ratio test

Figure 5.7 presents the temperature profiles of change disturbance in condition feed by Effect of Disturbance Method. Four disturbances are ±5% total flow rate test, ±1% temperature test, ± 4% composition test and ± 5% component flow rate test. The temperature profile of total flow change +5% has the biggest deviated from steady state at 16st tray. In the same way, Figure 5.7 (b), Figure 5.7 (c) and Figure 5.7 (d) are the greatest deviated from steady state at 8th, 16th and 23th trays. Three points in columnC2 are impossible to control the temperature tray.



(a) (b)

(c) (d)

Figure 5.7 Effect of disturbances on the steady-state temperature profiles of the column C2

(a) Total flow rate test

(b) Temperature test

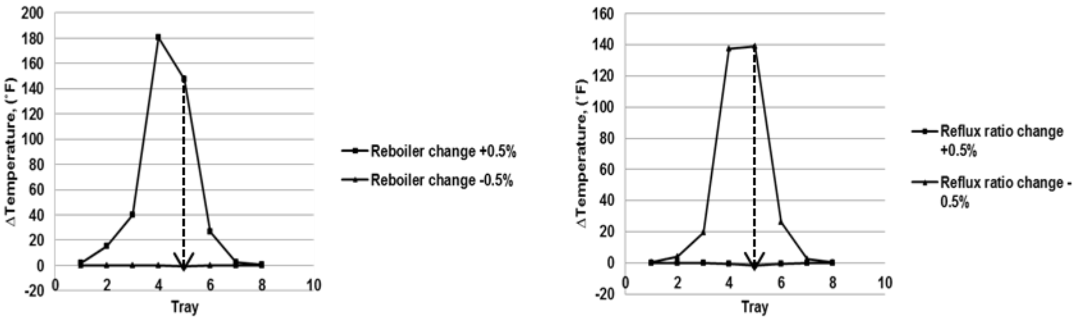
(c) Benzene composition test

(d) Benzene component flow rate test

Form analytical Open loop sensibility method and Effect of Disturbance method. In summary, 8th, 16th and 23th trays can be used to control the temperature in column C2 for quality of product.

Column C3

In Open Loop Sensibility Method, the temperature profile of +0.5% of reboiler duty has the highest deviated from steady state at 5th tray similar to reflux ratio test +0.5%. The open loop sensitivity analyses of Column C3 is presented by Figure 5.8



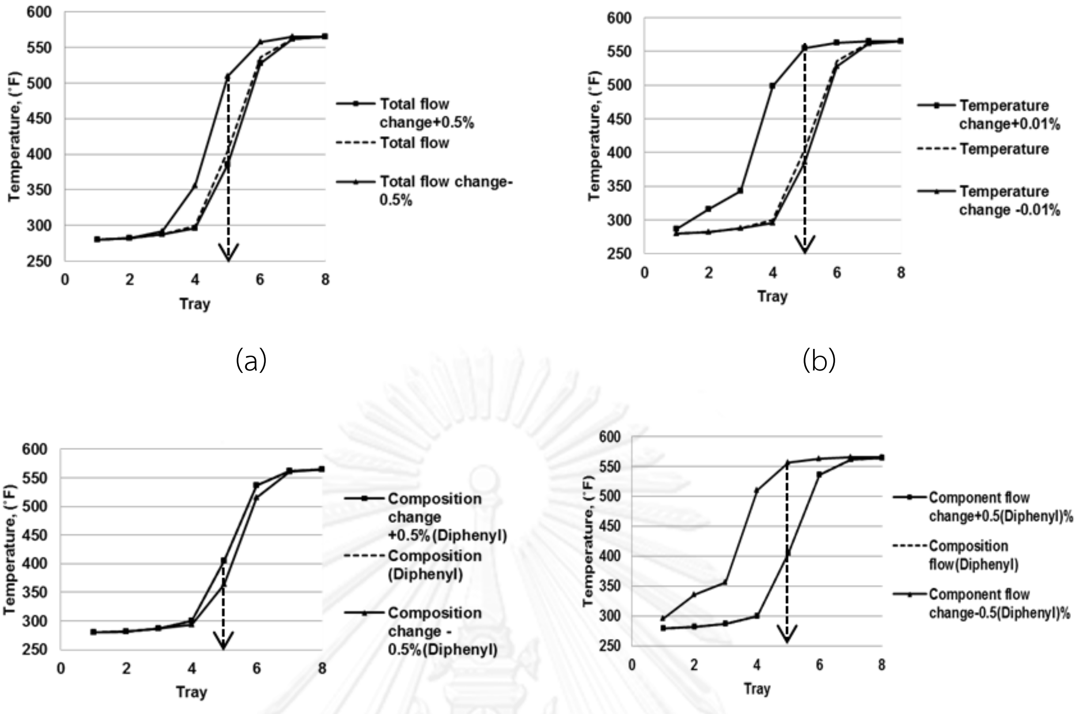
(a) (b)

Figure 5.8 The open loop sensitivity analyses of Column C3

(a) Reboiler duty test

(b) Reflux ratio test

In Effect of Disturbance Method, figure 5.9 (a)-5.9 (d) presents the temperature profile. There have the largest deviated from steady state at 5th tray.



(a) (b)

(c) (d)

Figure 5.9 Effect of disturbances on the steady-state temperature profiles of the column C3

(a) Total flow rate test

(b) Temperature test

(c) Diphenyl composition test

(d) Diphenyl component flow rate test

Form analytical sensibility test of column C3 by Open loop sensibility and effect of disturbance. In short, we will use 5th in control the temperature in column C3 for control composition of toluene in recycle toluene stream.

Stage2. Fixture Plant and Disturbance Management

Step 3: Establish fixture plant.

Creating a material balances in an entire plant by control each component at its quantifier or its fixture point.

3.1 Keep the materials entered and reentered fixed.

In this step, the combined flow of hydrogen fresh feed and gas recycle is controlled. The combined flow of toluene fresh feed and toluene recycle is controlled. Figure 5.10 presents keep the materials entered and reentered fixed.

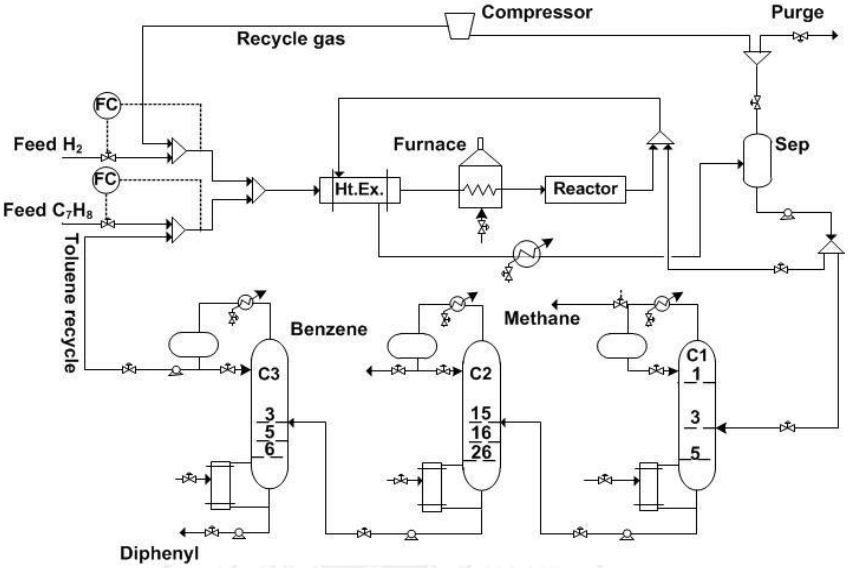


Figure 5.10 Keeps the materials entered of hydrodealkylation process

3.2 Adjust the flow of exit material streams.

The three exit product streams of CH_4, C_6H_6, and $C_{12}H_{10}$ are regulated according to their accumulation in C1 top, C2 reflux drum and C3 base. Figure 5.11 presents adjust the flow of exit material streams to their accumulation.

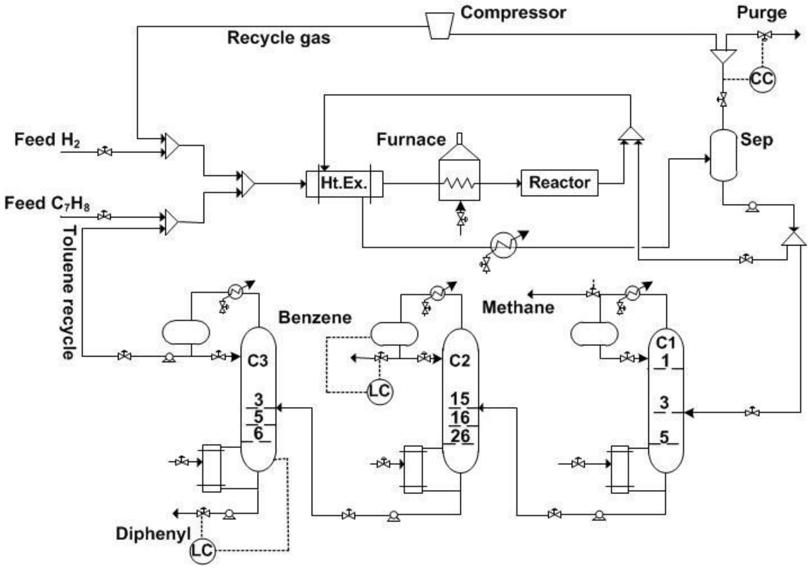


Figure 5. 11 Adjust the flow of exit material streams (methane, benzene, diphenyl)

3.3 Handling the material that is not leaving the process.

There is no component not leaving the process.

3.4 Control the amount of the rest of the component at their quantifiers.

There is no component left to be considered in this step.

3.5 Maintain the production rate.

For throughput change, we change the toluene feed rate. To maintain production rate, we adjust the reactor inlet temperature.

Step 4.Disturbance management for quality control.

In this step, the design of the control loops to manage with both thermal and material disturbances. An analysis of disturbances confronted in the hydrodealkylation process. The disturbances are handled by principle of disturbances management:

4.1 Heat Disturbance Management.

4.1.1 Direct the heat disturbances that are not directly related to quality.

According to the information available in Steps 1.1, the temperatures of reactor effluent are quenched to 1150 °F. According to the information derived in Steps 2.2, and 3.5, the temperatures of reactor inlet stream must be heated to 1150 °F using FEHE and furnace. Direct the heat disturbances that are not directly related to quality is presented by figure 5.12

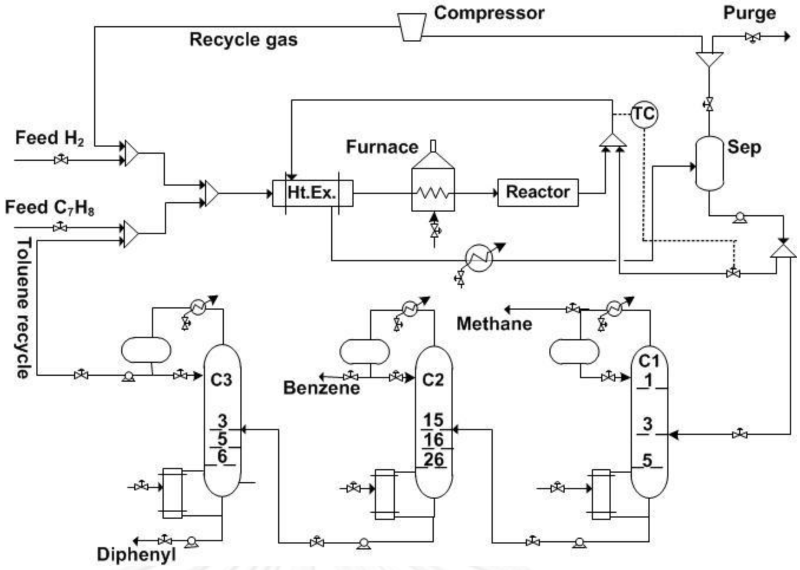


Figure 5. 12 Direct the heat disturbances that are not directly related to quality.

4.1.2 Manage the heat disturbance that related to quality in order to maintain the product constraints.

The temperature of separator liquid is regulated at 113 °F by adjusting the cooler duty. The temperatures of C1, C2 and C3 at trays 1, 16 and 5 are controlled by adjusting their reboiler duties. Figure 5.13 is shown the places at thermal disturbance management.

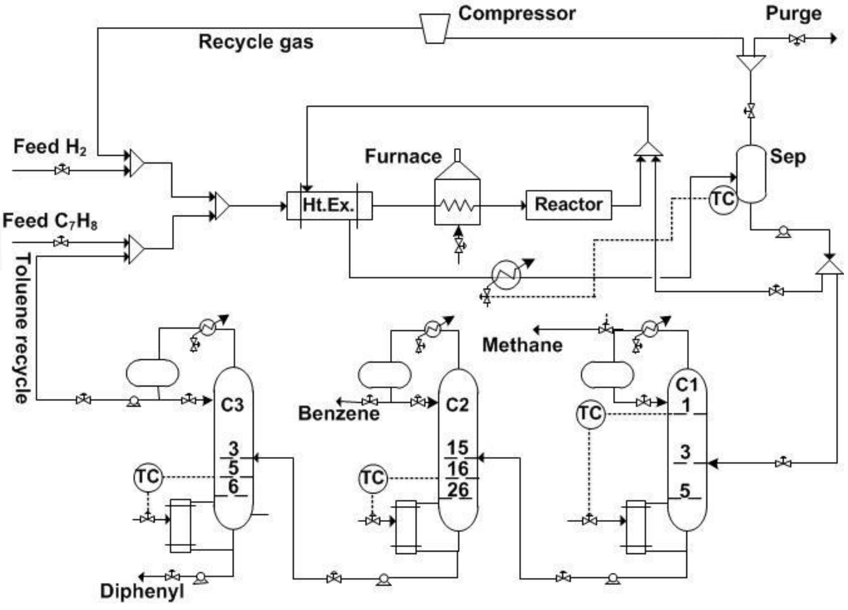


Figure 5. 13 The control structure obtained in step 4.1.2

4.2 Material Disturbance Management.

Material disturbance is one of importance for purity of benzene product so the design control loop for distillation column is importance to handle material disturbance. It separates two sections. Firstly, we run in steady state simulation that we consider possibility of appropriate control loops in separation section. Secondly, we run in dynamic mode simulation that we select the appropriate control loops for separation section.

Table 5.5 presents directions of material disturbances. We tested material direction with control loop of distillation by increase/decrease of light-key/heavy-key component flow rates. We used six control loops into material disturbance test such as reflux flow rate (R), reflux ratio (RR), reflux-to-feed (R/F), reflux fraction (R/R+D), boil up ratio (V/B) and dual temperature control .

Table 5.5 Surplus and deficit disturbances shifting test results for light and heavy keys.

		R	RR	R/F	R/(R+D)	V/B	Dual temp.	Desire shifting
C1	D1	CH_4+(4.21) CH_4-(4.21)	CH_4+(4.21) CH_4-(4.21)	CH_4+(4.21) CH_4-(4.21)	CH_4+(4.21) CH_4-(4.21)	-	-	CH_4±, C_6H_6-
	B1	C_6H_6+(26.81) C_6H_6-(26.81)	C_6H_6+(26.82) C_6H_6-(26.82)	C_6H_6+(27.67) C_6H_6-(27.69)	C_6H_6+(27.51) C_6H_6-(27.50)	-	-	C_6H_6+
C2	D2	C_6H_6+(26.79) C_6H_6-(26.77)	C_6H_6+(26.79) C_6H_6-(26.78)	C_6H_6+(26.79) C_6H_6-(26.78)	C_6H_6+(26.79) C_6H_6+(26.79)	C_6H_6+(26.79) C_6H_6-(26.77)	C_6H_6+(26.79) C_6H_6-(26.78)	C_6H_6+
	B2	C_7H_8+(8.37) C_7H_8-(8.37)	C_7H_8+(8.37) C_7H_8-(8.37)	C_7H_8+(8.38) C_7H_8-(8.38)	C_7H_8+(8.38) C_7H_8-(8.38)	C_7H_8+(8.39) C_7H_8-(8.39)	C_7H_8+(8.37) C_7H_8-(8.37)	C_7H_9±
C3	D3	C_7H_8+(25.11) C_7H_8-(25.12)	C_7H_8+(25.11) C_7H_8-(25.12)	C_7H_8+(25.11) C_7H_8-(25.12)	C_7H_8+(25.11) C_7H_8-(25.11)	C_7H_8+(25.11) C_7H_8-(25.12)	-	C_7H_8+
	B3	$C_{12}H_{10}$+(1.83) $C_{12}H_{10}$-(1.83)	$C_{12}H_{10}$+(1.83) $C_{12}H_{10}$-(1.83)	$C_{12}H_{10}$+(1.83) $C_{12}H_{10}$-(1.83)	$C_{12}H_{10}$+(1.84) $C_{12}H_{10}$-(1.83)	$C_{12}H_{10}$+(1.83) $C_{12}H_{10}$-(1.83)	-	C_7H_8-, $C_{12}H_{10}$±

Column C1

The control gold of column C1 is to shift the plus/minus CH_4 and minus C_6H_6 to the top of the column C1 because the C1 bottom flow must be maintained composition of methane. The benzene plus must be shift to the C1 top flow since the product benzene must be lost in top column C1. All control loops are possible to be chosen to handle disturbances of component flow. However, all control loops cannot be handled minus benzene in distillate of column C1 because the quantity of benzene at top of column C1 is very low. We cannot decrease quantity of benzene at top of column C1.

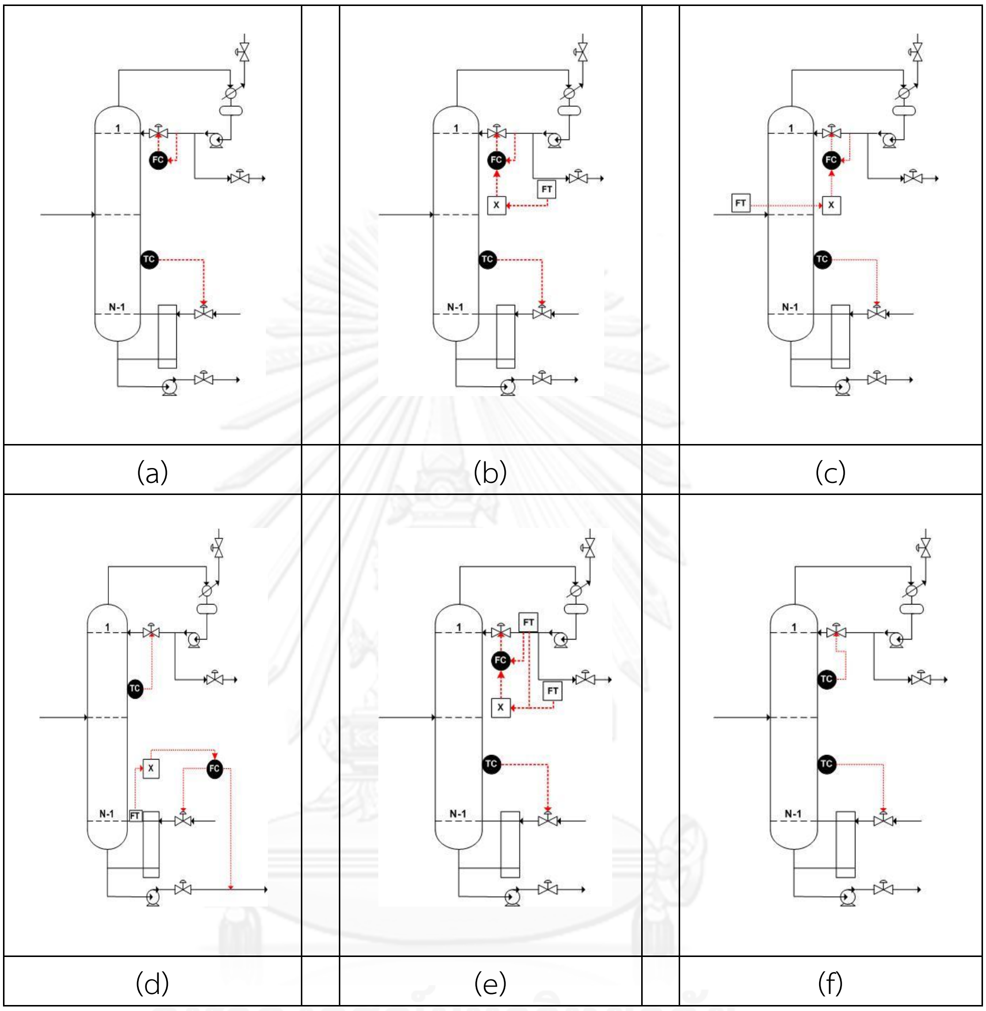
Column C2

The control gold of column C2 cannot avoid shifting the plus benzene component to the top of the column C2 because C_6H_6 should be not lost at top of column C1 and disturb the purity in the recycle stream. The plus/minus toluene are able to shift to the C2 bottom flow. All control loops are possible to be chosen to handle disturbances of component flow. Nevertheless, all control loops cannot be adjusting the minus benzene since the quantity of benzene at bottom of column C2 is very low.

Column C3

The control gold of column C3 is to shift the plus/minus diphenyl and minus toluene component to the bottom of the column C3, because compositions of dipheny in recycle toluene is kept. The plus toluene is shifted to the top column C3 because the toluene must be loss in C3 bottom flow. All control loops are possible to be chosen to handle disturbances of component flow. But, all control loops cannot be managing the deficit of Toluene, for the quantity of Toluene in bottom of distillation C3 is very low.

Next, material direction is tested in dynamic mode. The control structures are tested such as reflux flow rate, reflux ratio, reflux-to-feed, reflux fraction and boil up ratio and dual temperature control. The disturbance of changes in component flow rate, total flow rate, composition and temperature at feed column are tested, with the distillate and bottoms compositions held at their specified values. The configurations of the control loops are decided base on the desired material pathways. Figure 5.14 presents the control structure for handle disturbances in separation section.



(a) (b) (c)

(d) (e) (f)

Figure 5. 14 The control structure for handle disturbances in separation section.

(a) Reflux Control loop
(b) Reflux Ratio Control loop
(c) Reflux to Feed Control loop
(d) Boil up ratio control loop
(e) Reflux Fraction Control loop
(f) Dual Temperature Control loop

Column C1

The light-key component is methane and heavy-key is other components (benzene, toluene, diphenyl). We test disturbances by change feed in distillation column C1. The disturbances include light-key/heavy-key component flow rate, total flow rate, composition of light-key and temperature. Figure 5.15 presents result of simulation in dynamic mode with control loops (reflux flow, reflux ratio, reflux-to-feed). All control loops can regulate the composition in distillate and bottom.

For change in methane feed flow, all structures yield the desired methane disturbance shifting. However, the RR and R/(R+D) structures give the best performance in maintaining methane composition (xD1) in distillate close to the specified values. See Figure 5.15

For plus disturbance of benzene flow in feed, RR and R/(R+D) give the desire result. However, no structures yield the desired minus benzene disturbance shifting due to the separation capability of column C1 and the availability of benzene in distillate. Therefore RR and R/(R+D) structures are the best candidates in maintaining benzene composition (xD1) in distillate. See Figure 5.16.

In addition, for change in total feed flow, the desired control performance is maintaining methane composition in distillate and bottom. The RR, R/(R+D) and R/F structures are the best structures. Furthermore, RR and R/(R+D) structures are able to settle methane compositions (xD1) faster than that of R/F. See Figure 5.17

For change in methane composition in feed, the RR and R/(R+D) structures are the best for maintaining methane compositions (xD1 and xB1). See Figure 5.19

For change in feed temperature, all structures are able to settle methane composition. Besides, RR and R/(R+D) structures yield faster responses See Figure 5.19

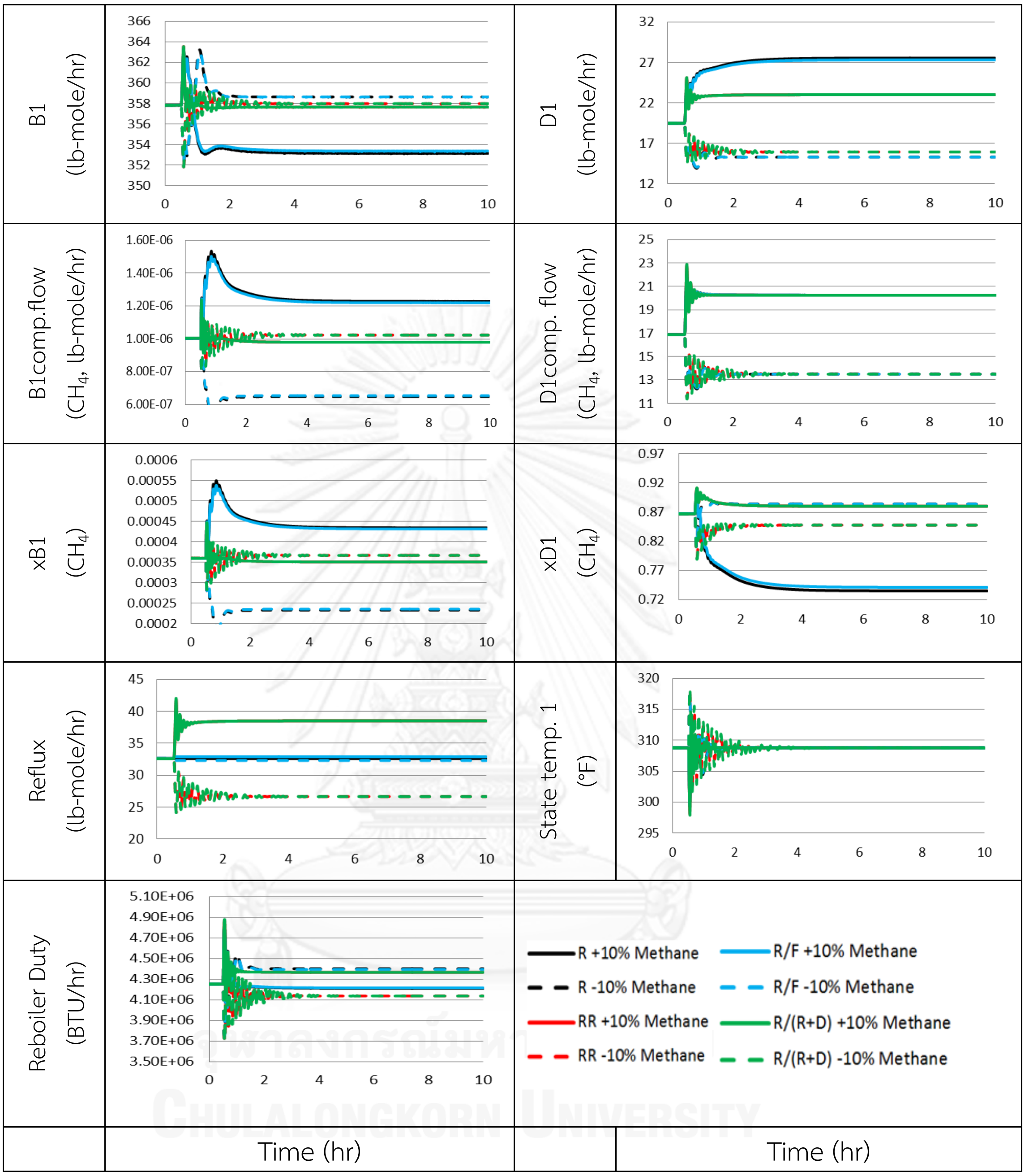


Figure 5. 15 Column C1 dynamic results for ±10% changes of methane feed flow.

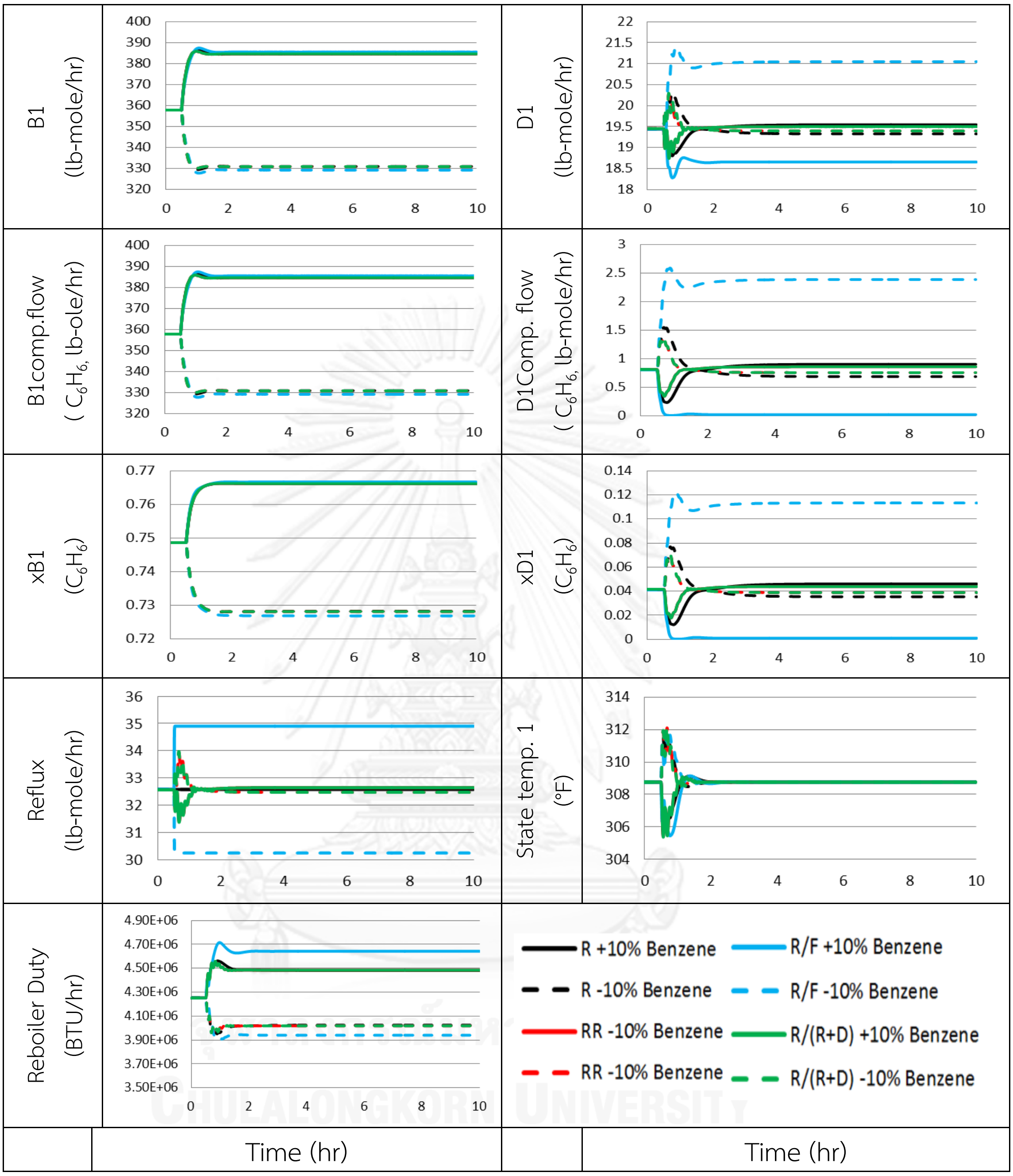


Figure 5. 16 Column C1 dynamic results for ±10% changes of benzene feed flow.

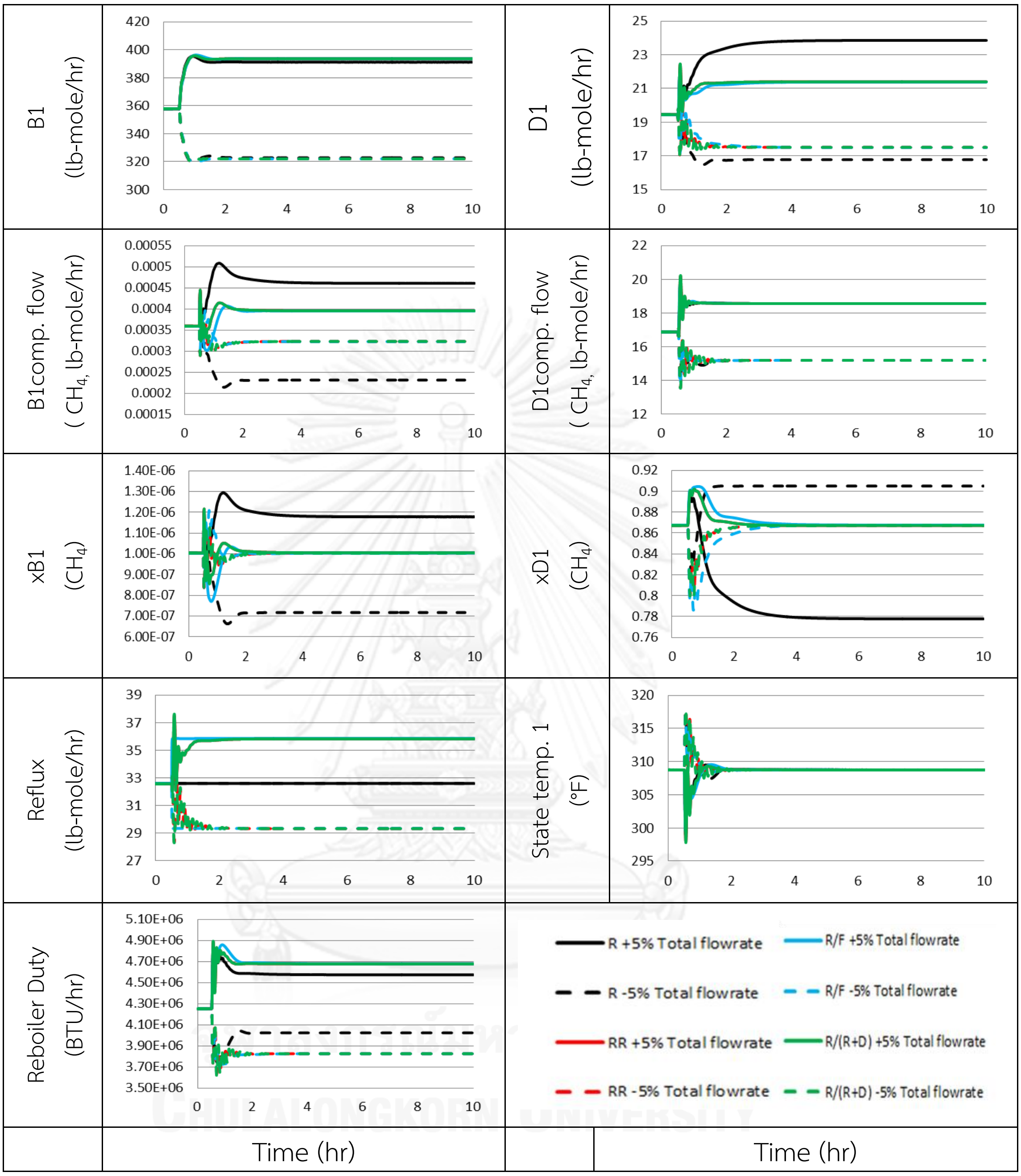


Figure 5. 17 Column C1 dynamic results for ±5% changes of total flow in column feed

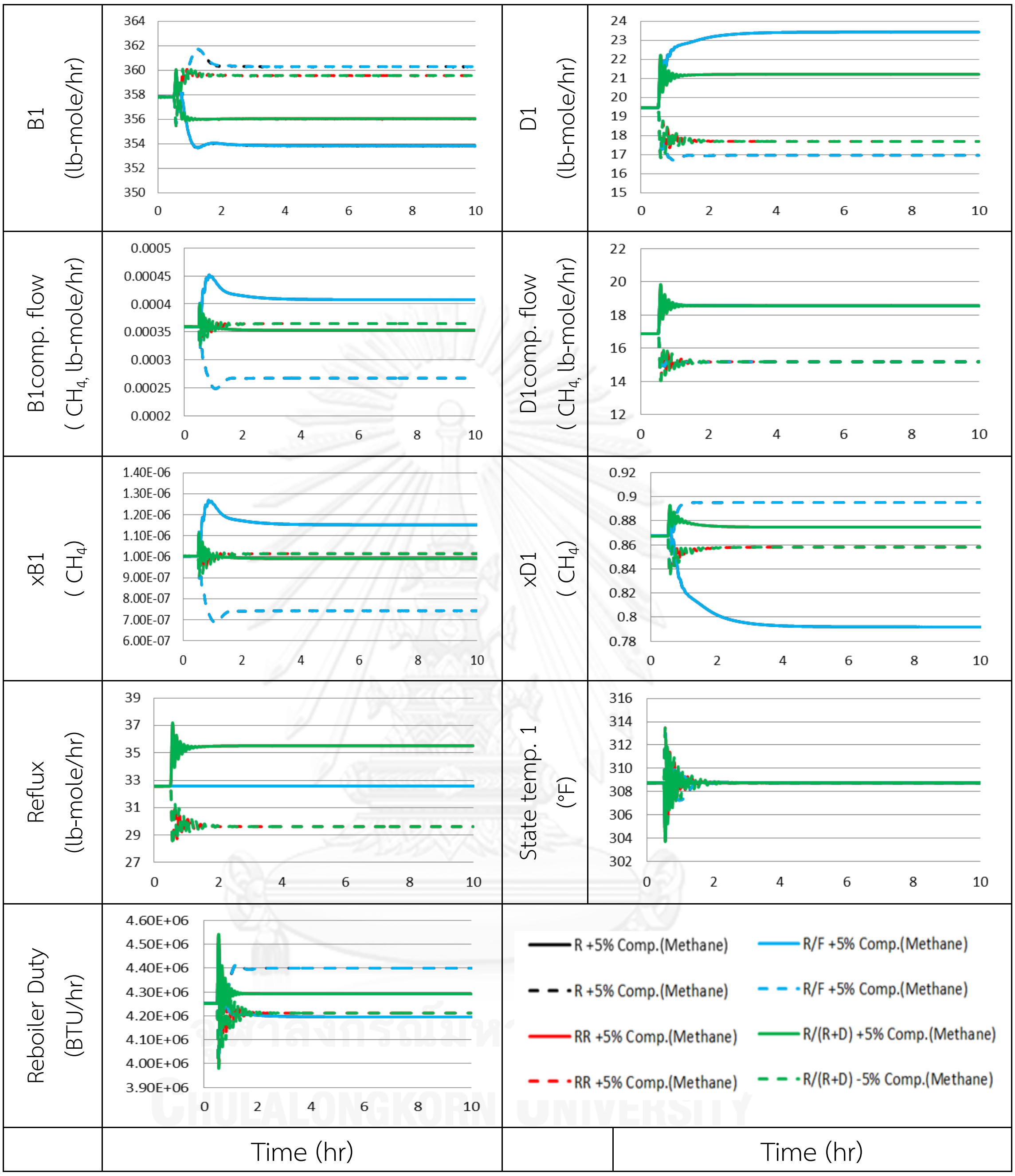


Figure 5. 18 Column C1 dynamic results for ±5 composition methane changes in column feed

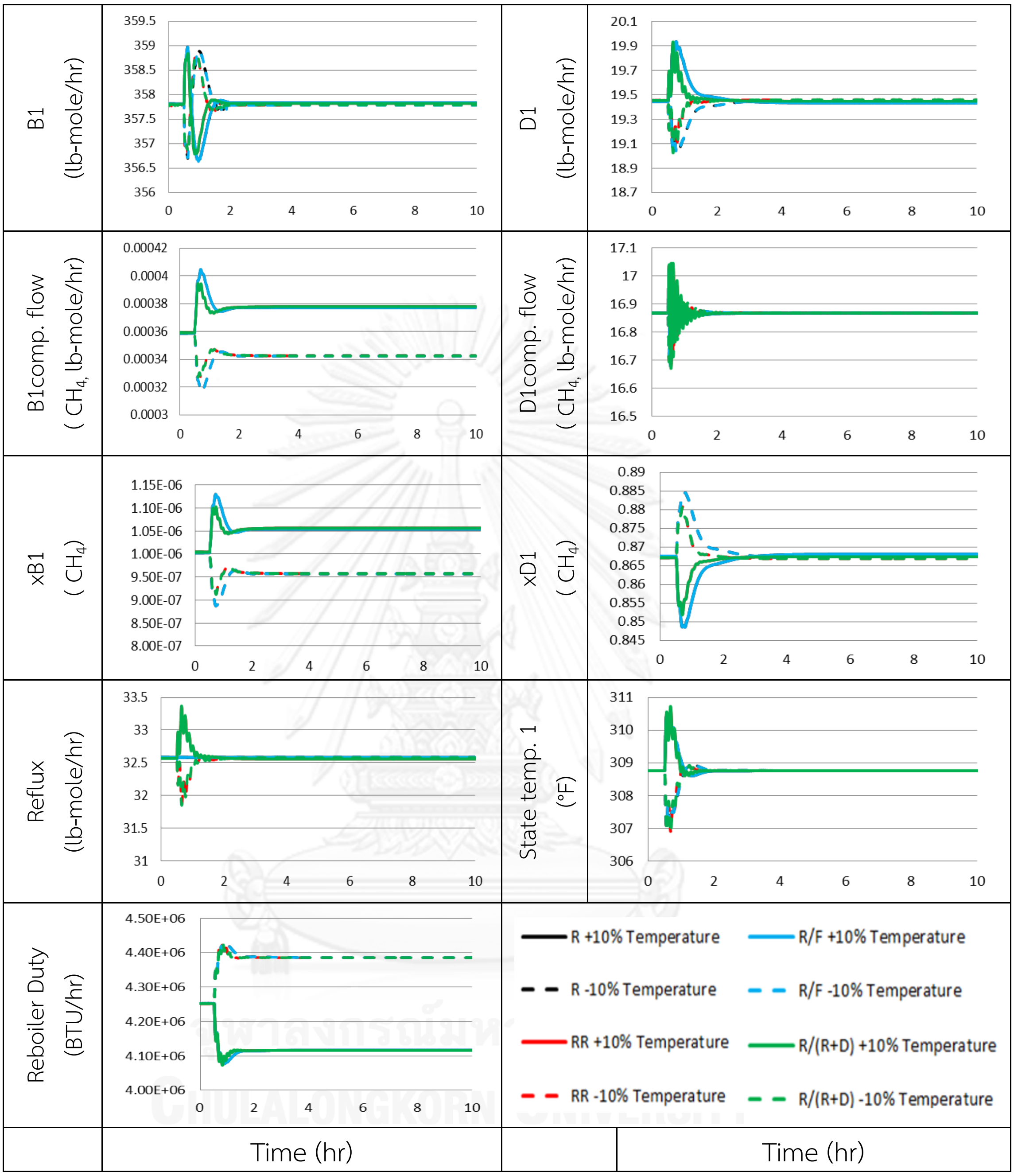


Figure 5. 19 Column C1 dynamic results for ±10% changes of temperature feed flow.

Column C2

The light-key is benzene and the rests of components are heavy-key (toluene and diphenyl) Benzene is product of the hydrodealkylation process. Toluene is substrate of process.

Figure 5.20 presents result of ±10% changes of the component flow rate of benzene. Notice that the benzene flow rate in top (D2) increases and decreases for increases and decreases in Benzene feed flow rate, respectively. The performance results of all control structures are not quite different in D2 component flow rate. However, The R/F structure is the best candidate in maintaining benzene composition in distillate.

Figure 5.21 presents result of ±10% changes of the component flow rate of toluene. Notice that the toluene flow rate in top (B2) increases and decreases for increases and decreases in toluene feed flow rate, respectively. The performance results of all control structures are not quite different in B2 component flow rate (B2 comp. flow) and composition (xD2). However, R/F is the best candidate in maintaining benzene composition in distillate.

Figure 5.22 presents result of ±10% changes of total feed column C2 .Notice that the bottom flow rate (B2) and distillate flow rate (D2) increases and decreases for increases and decreases in toluene feed flow rate. R/F is the best candidate in maintaining benzene composition (xD2) in distillate and benzene composition (xB2) in bottom flow rate.

Figure 5.23 presents result of ±10% changes of the composition of benzene. Notice that the distillate (D2) increases and decreases for increases and decreases in composition change. R/F is the best candidate in maintaining benzene composition (xD2) in distillate.

Figure 5.24 presents results for ±10% change of temperature in Column C2. The performance results of all control loops are not quite different in all graphs. Therefore, they are possible to use in column C2.

The R/F structure is chosen because it are good performance to manage with the disturbances (component flow of benzene and toluene, total feed column C2, composition) while the others structures tend to keep the benzene product composition in the distillate fixed.

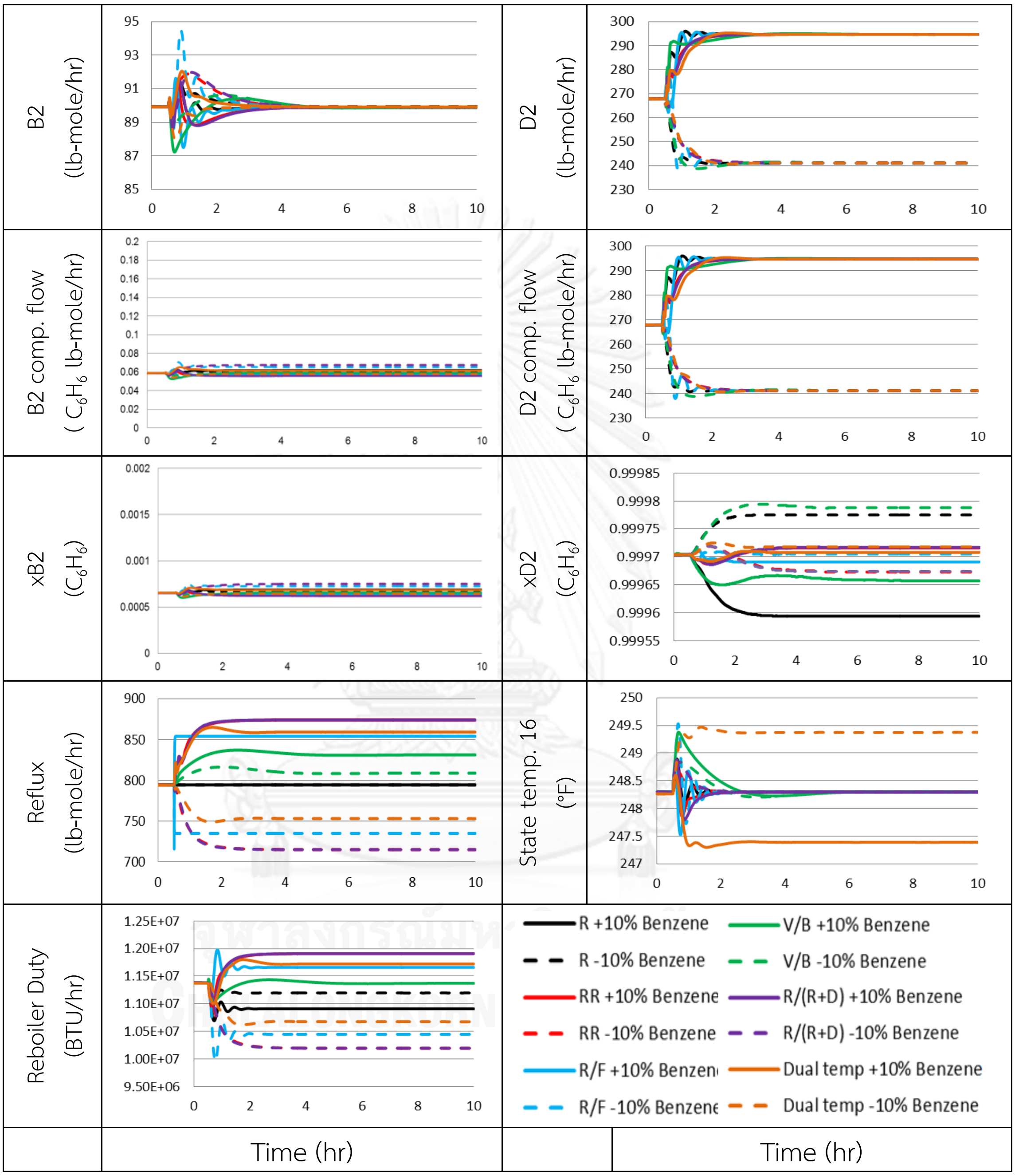


Figure 5. 20 Column C2 dynamic results for ±10% changes of benzene feed flow.

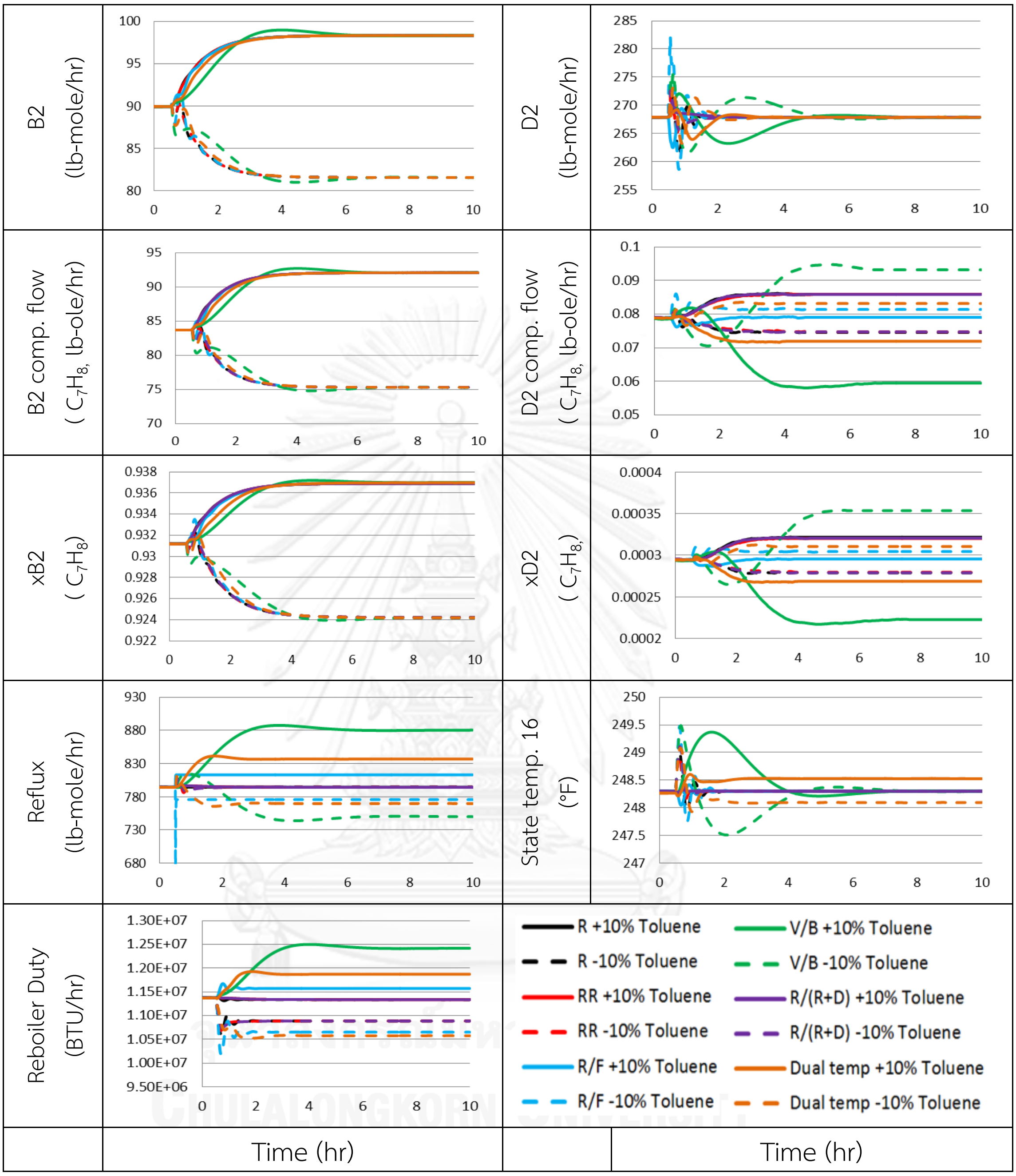


Figure 5. 21 Column C2 dynamic results for ±10% changes of toluene feed flow.

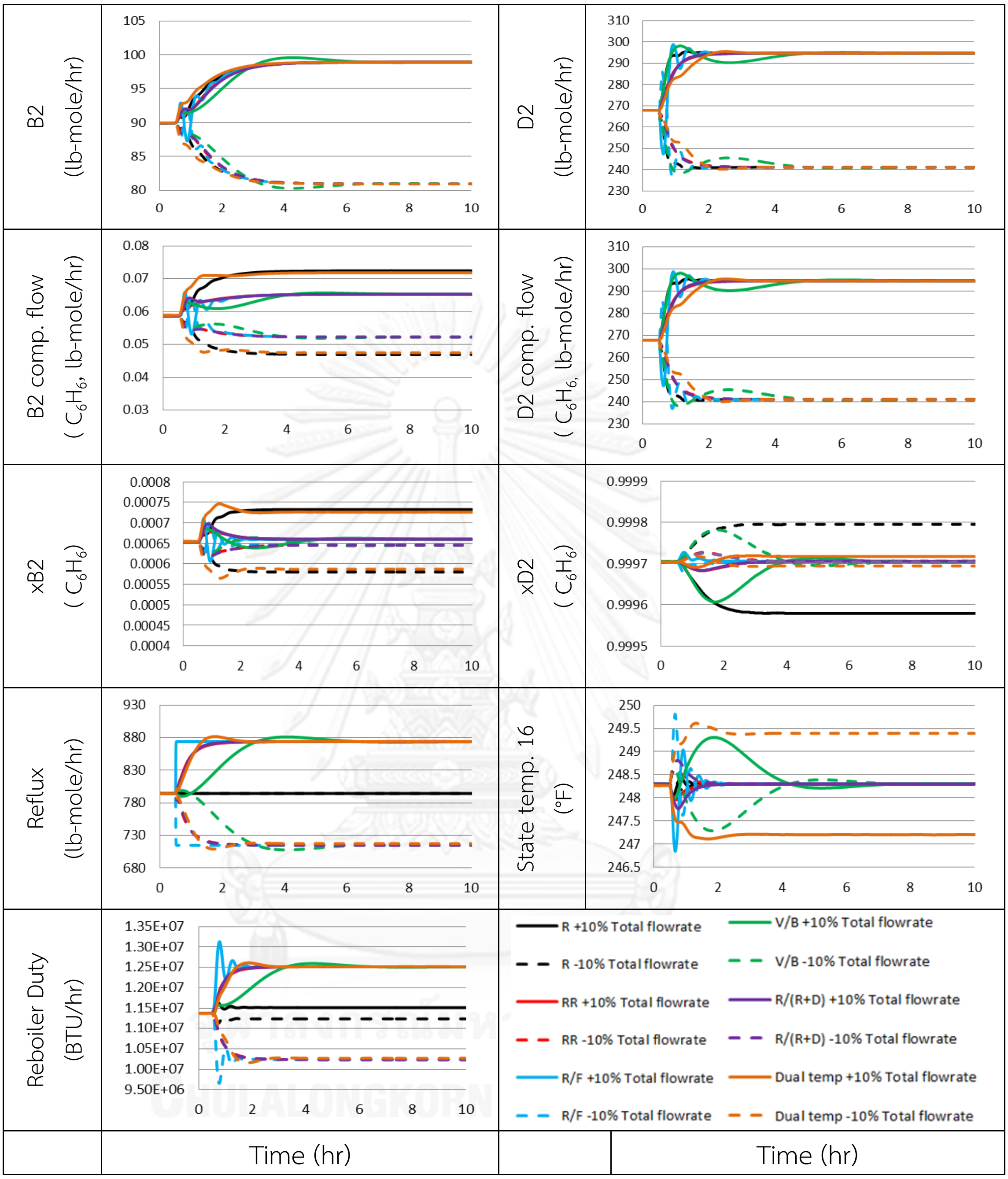


Figure 5. 22 Column C2 dynamic results for ±10% changes of total flow in column feed

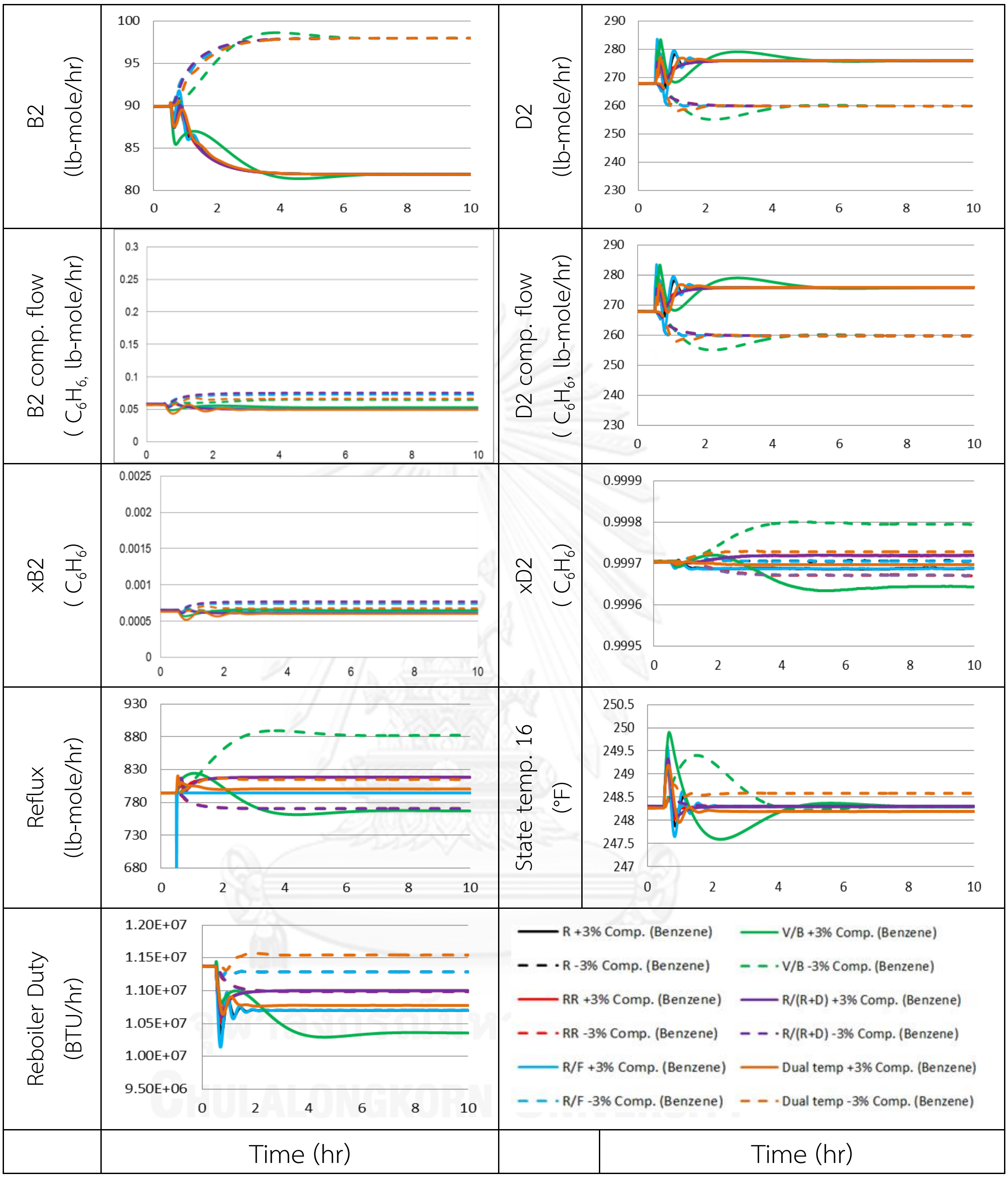


Figure 5. 23 Column C2 dynamic results for ±3% composition benzene changes in column feed

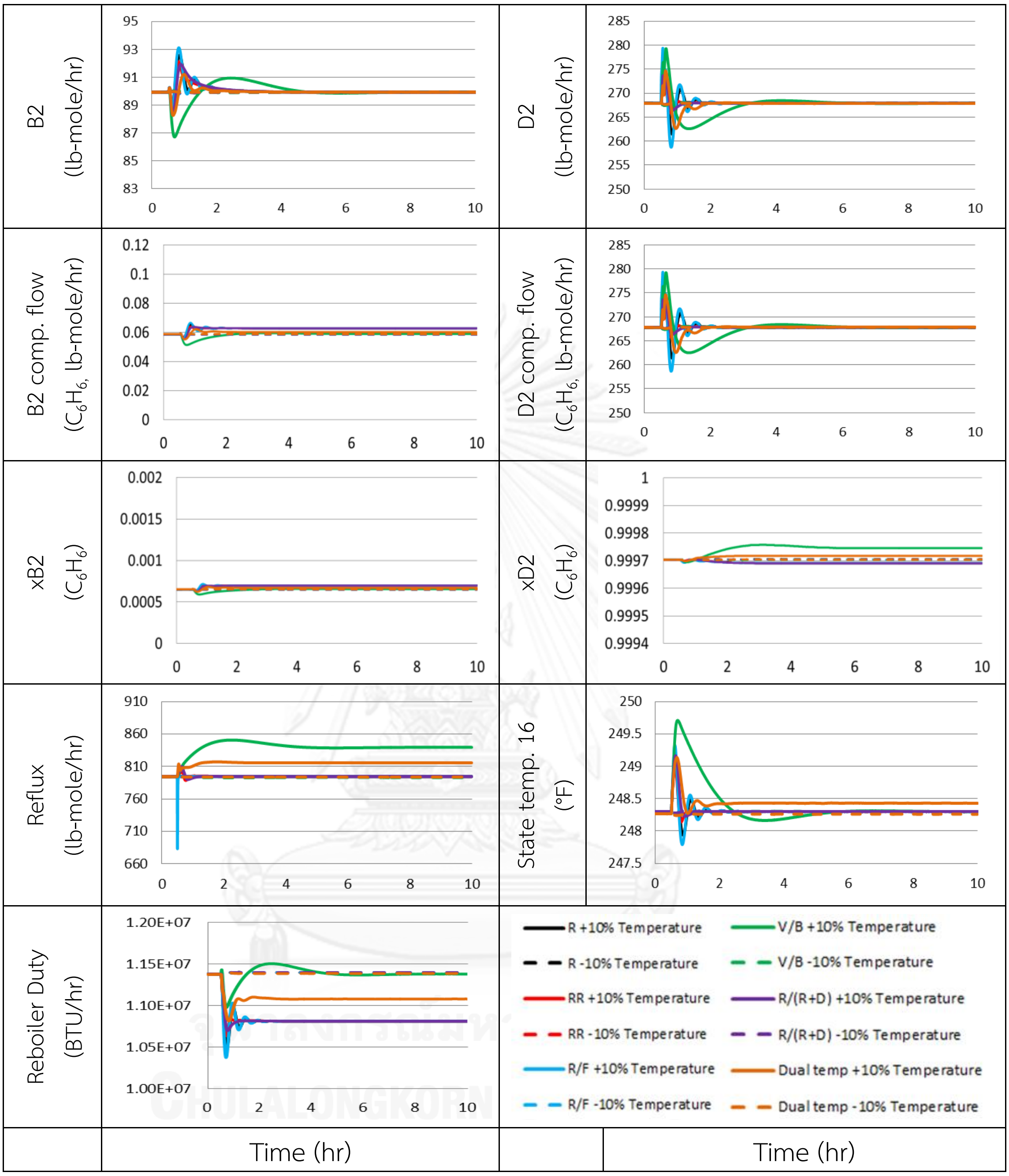


Figure 5. 24 Column C2 dynamic results for ±10% changes of temperature feed flow.

Column C3

Toluene is the light-key component and substrate. Diphenyl is the heavy-key component and by-product.

Figure 5.25 presents result of ±5% changes of the component flow rate of toluene. The performance results of all control structures are not quite different in all graphs. Therefore, all control loops is possible to control column C3.

Figure 5.26 presents result of ±5% changes of the component flow rate of diphenyl. The performance results of all control structures are not quite different in all graphs. Therefore, all control loop possible to control column C3.

Figure 5.27 presents results of ±5% changes of total feed column C3. The performance results of all control structures are not quite different in B3 component flow rate. However, the R/F, RR and R/(R+D) structures are held very close to purity of recycle in xD3 while R/F can keep the impurity more than RR and R/(R+D) in total feed test. In addition, the R/F structure is the quickest converge to setpoint in composition test. Thus, RR and R/F structures are possible in managed the material disturbance.

Figure 5.28 presents result of ±5% changes of composition toluene of column C3. The R/F structure littler shifts diphenyl to top of column C3 than other control loops.

Figure 5.29 presents result of ±5% changes of the temperature. The performance results of all control structures are not quite different in all graphs. Therefore, all control loops possible to control column C3.

The performance results of all control structures are not quite different in component flow test and temperature test. R/F can keep the impurity more than RR and R/(R+D) in total feed test. In the same way, the R/F structure is the quickest converge to setpoint in composition test. Thus, the RR and R/F structure are selected in managed the material disturbance.

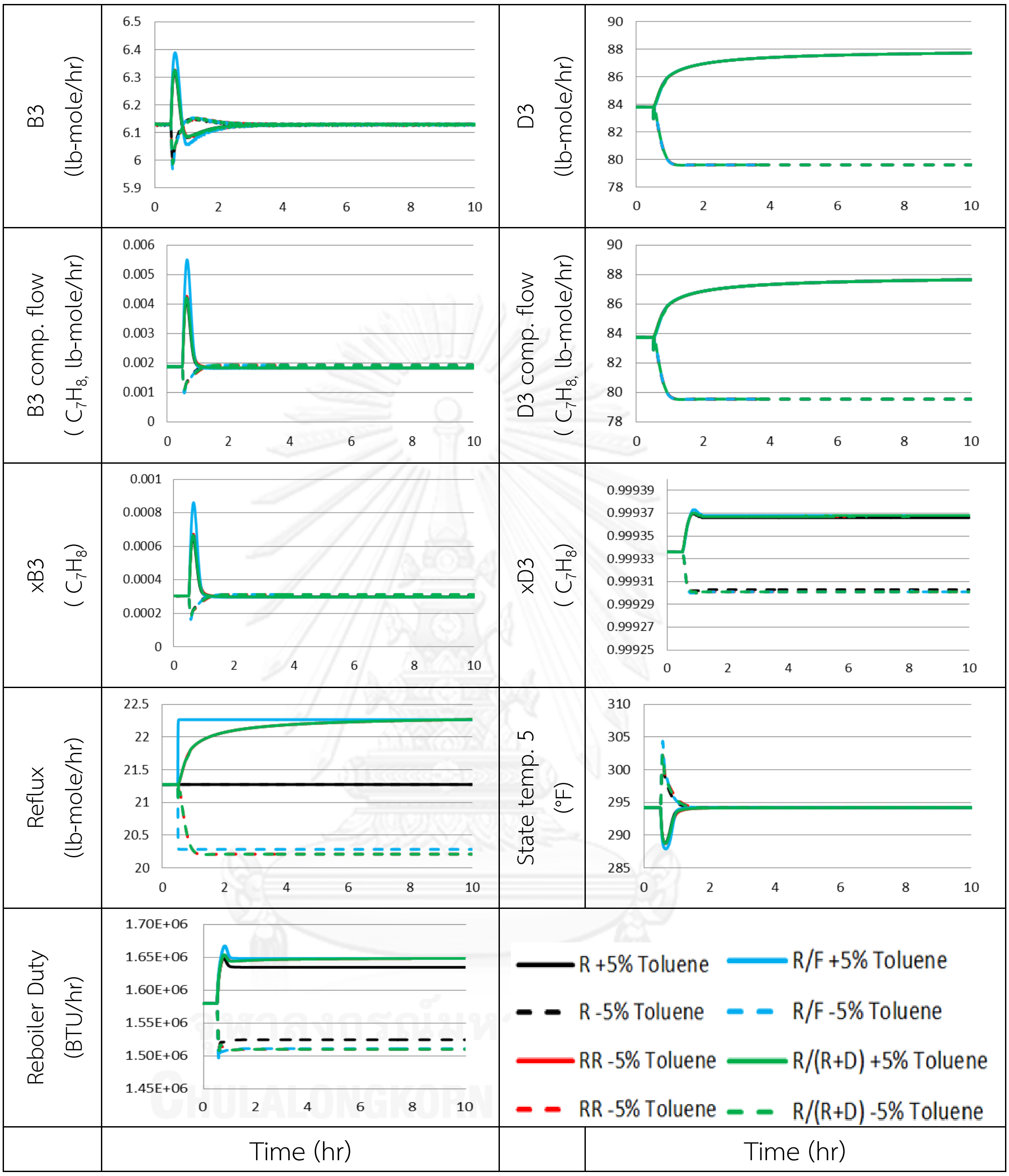


Figure 5. 25 Column C3 dynamic results for ±5% changes of toluene feed flow.

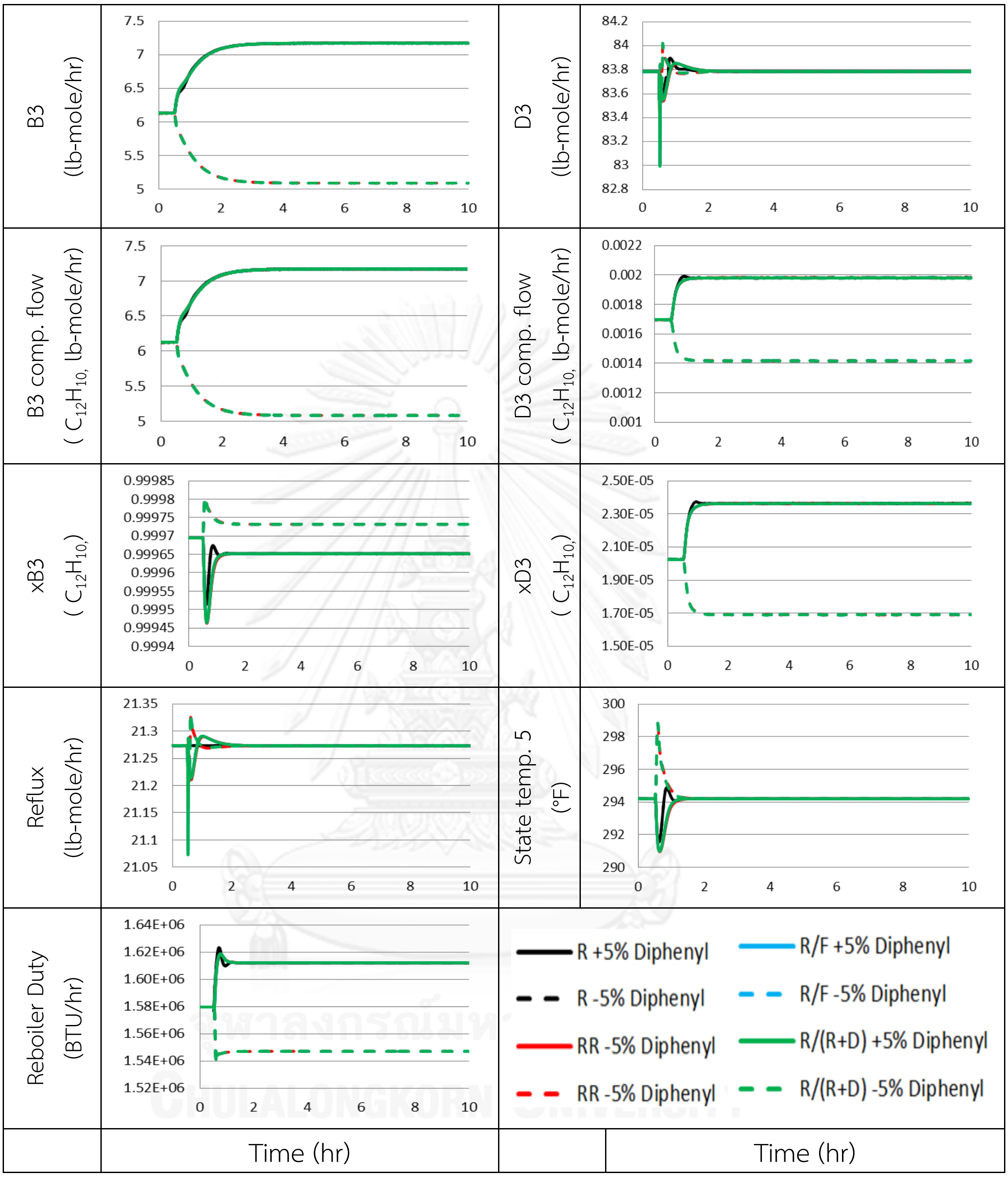


Figure 5. 26 Column C3 dynamic results for ±5% changes of diphenyl feed flow.

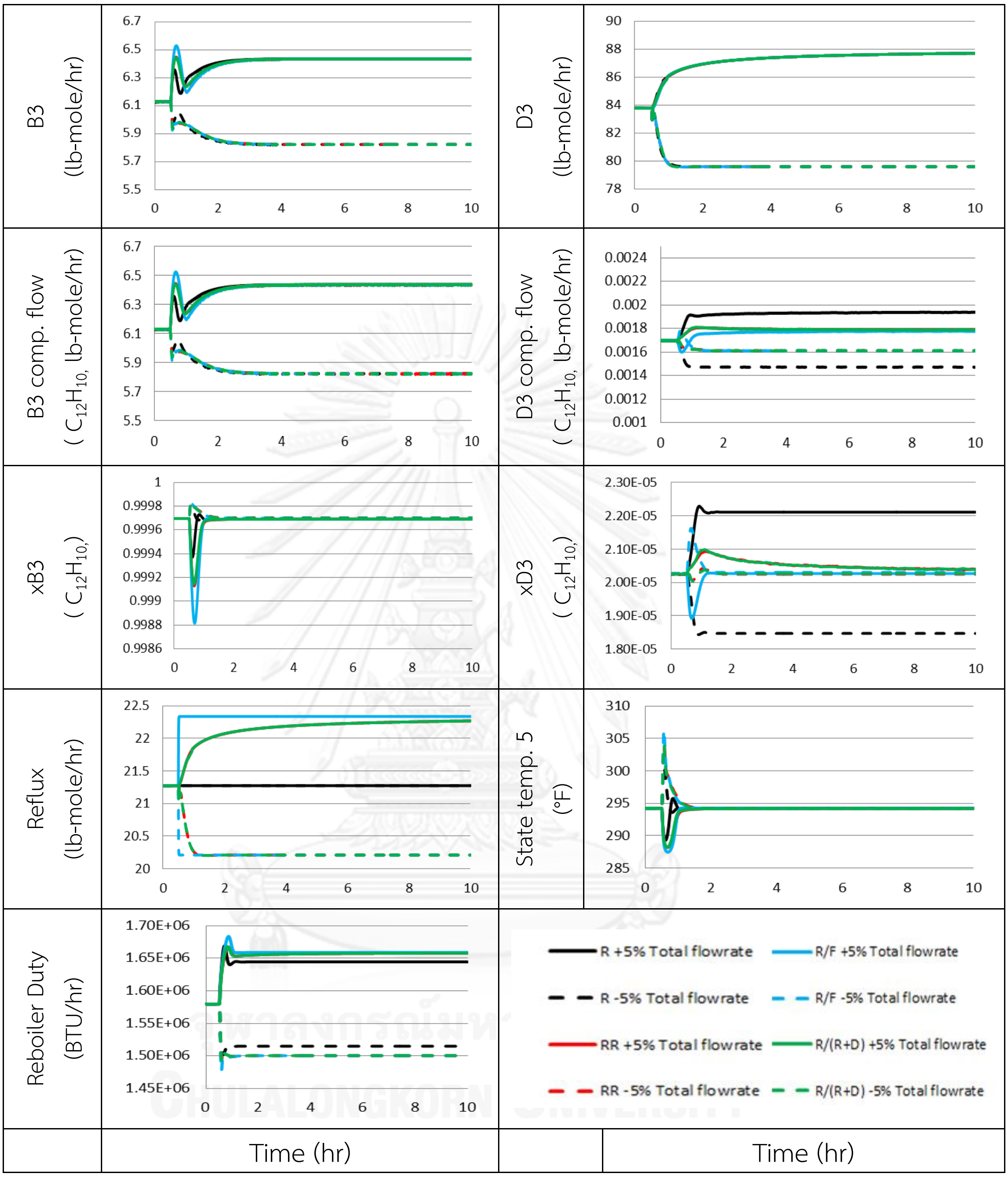


Figure 5. 27 Column C3 dynamic results for ±5% changes of total flow in column feed

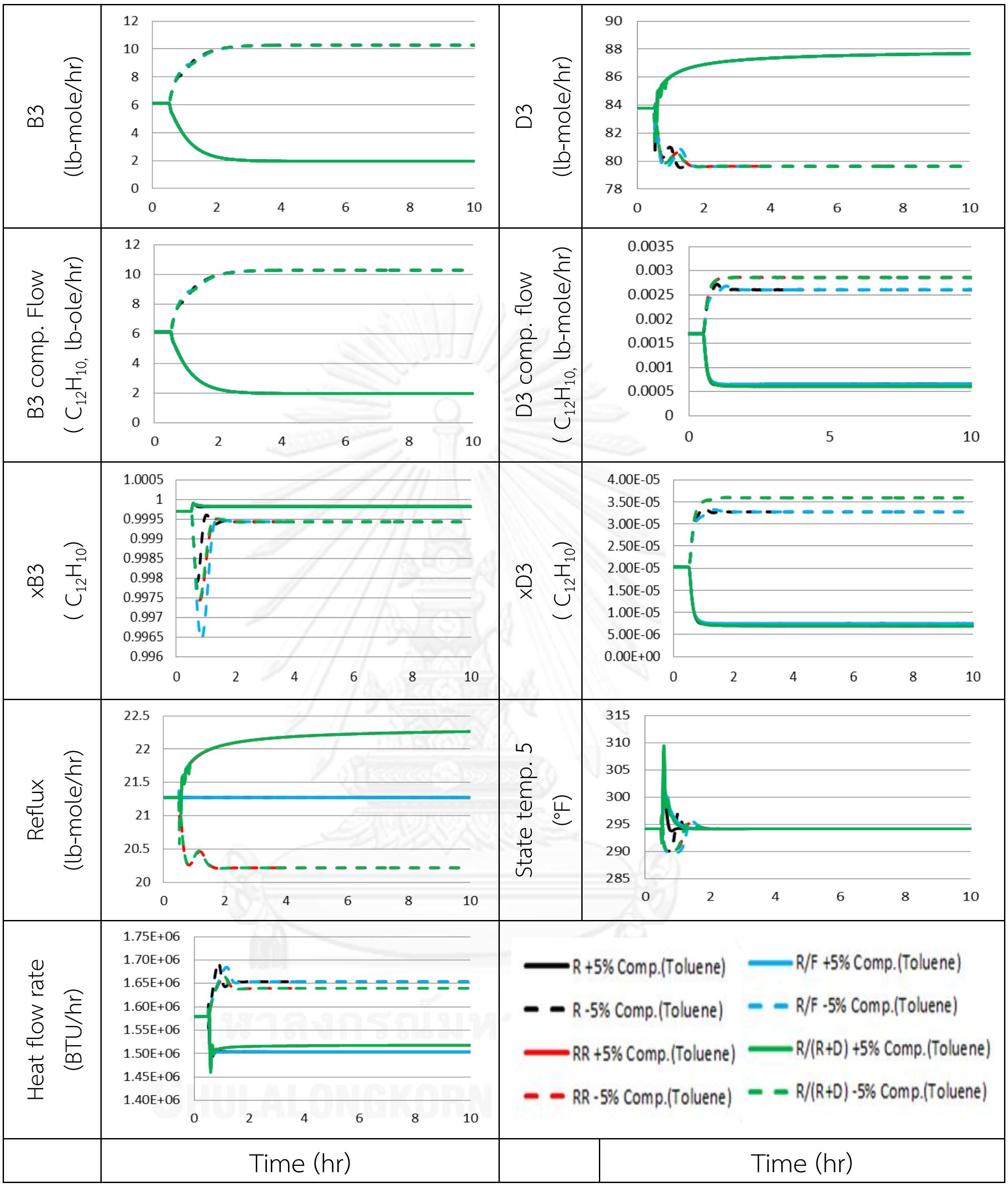


Figure 5. 28 Column C3 dynamic results for ±5 composition toluene changes in column feed

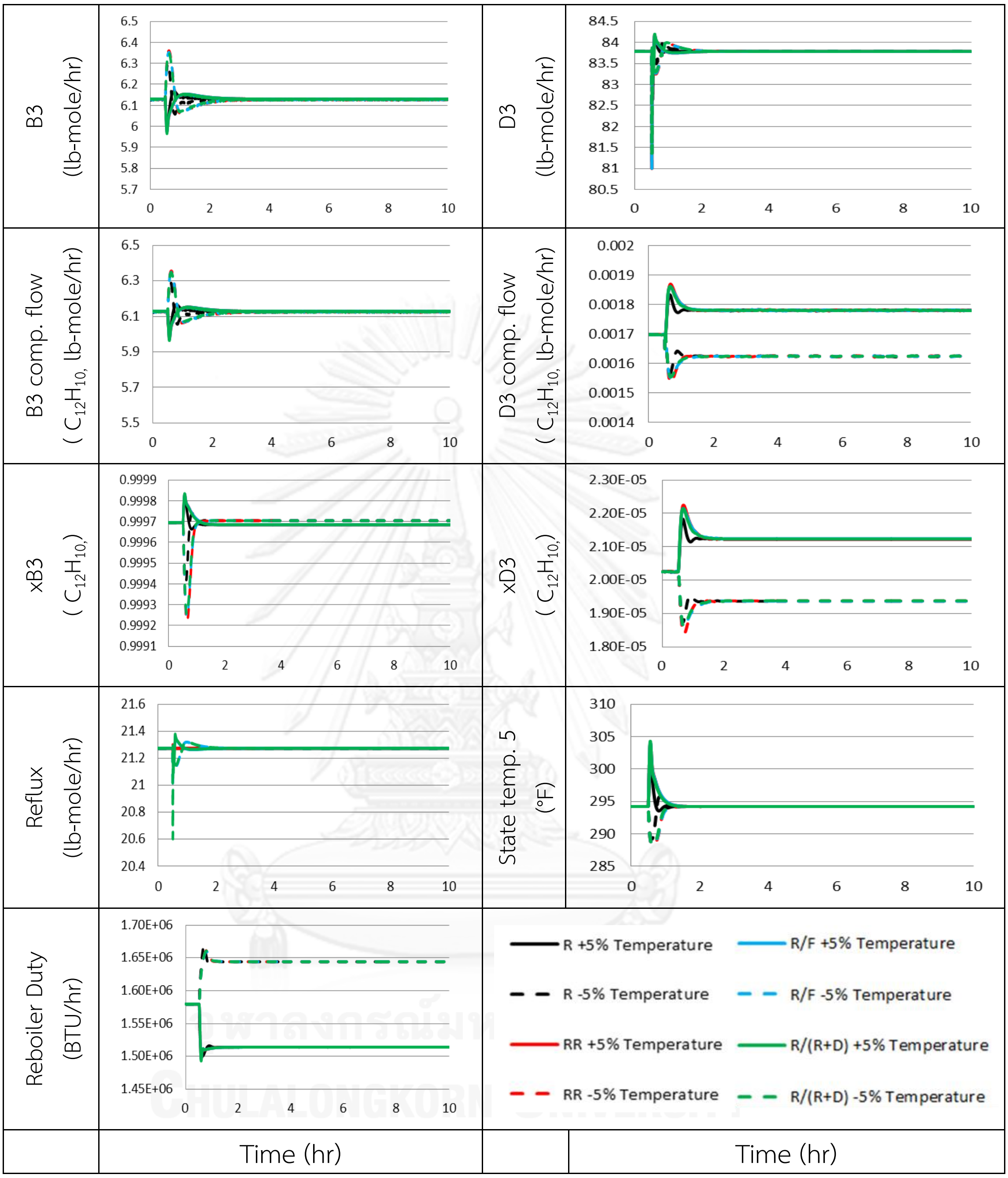


Figure 5. 29 Column C3 dynamic results for ±5% changes of temperature feed flow.

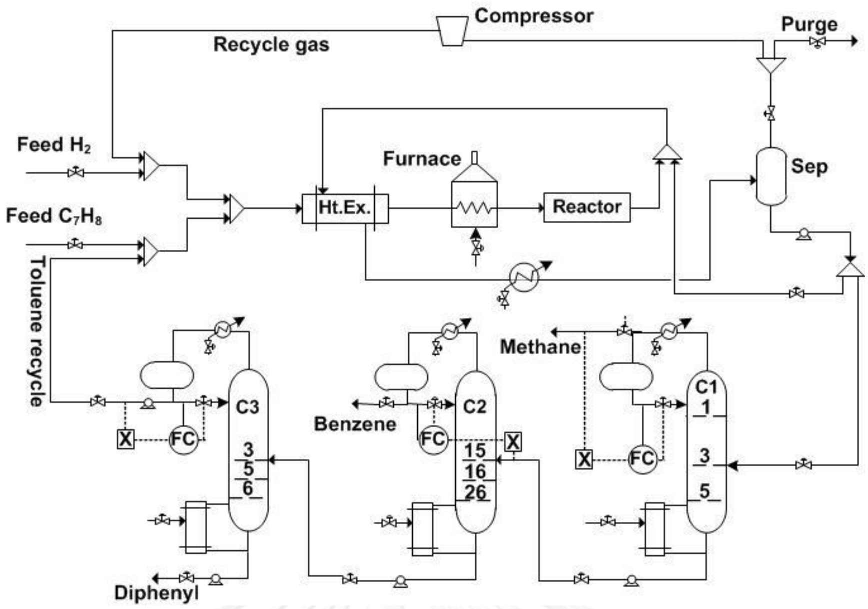


Figure 5. 30 The control structure obtained in step 4.2 with RR column control structure.

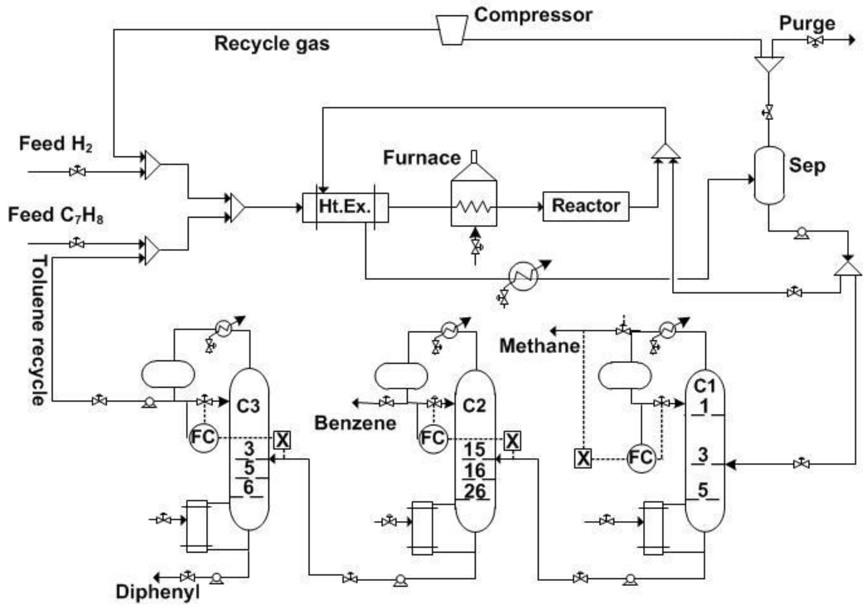


Figure 5. 31 The control structure obtained in step 4.2 with R to F column control structure.

Figure 5.30-5.31 present the control loops for material disturbance management. Three control loops is used to handle material disturbance. Both control structure 1 and 2 are similar in reflux ratio of column C1 and reflux to feed of

C2. However, they differ in column are C3reflux ratio of CS1 and reflux to feed of CS2.

Stage3. Unit Level Design

Step 5: Design the rest of the control loops.

Step 5.1 The remaining loops to be designed in this step are in the inventory loops of separator and 3 columns. Normally, the pressure of separator is controlled by manipulating its top flow rate and the liquid level is controlled by manipulating its bottom flow. Since C1 top product is vapor, its pressure is controlled by adjusting its vapor product flow. C1 reflux drum level is therefore adjusted by its condenser duty. C1 base level is controlled by adjusting C1 bottoms flow. The pressures of C2 and C3 are controlled by manipulating their condenser duties and the base levels are controlled by manipulating the bottoms. Next, the C3 condenser level is controlled by C3 distillate flow. The control structures obtained in this step is shown in Figure 5.32

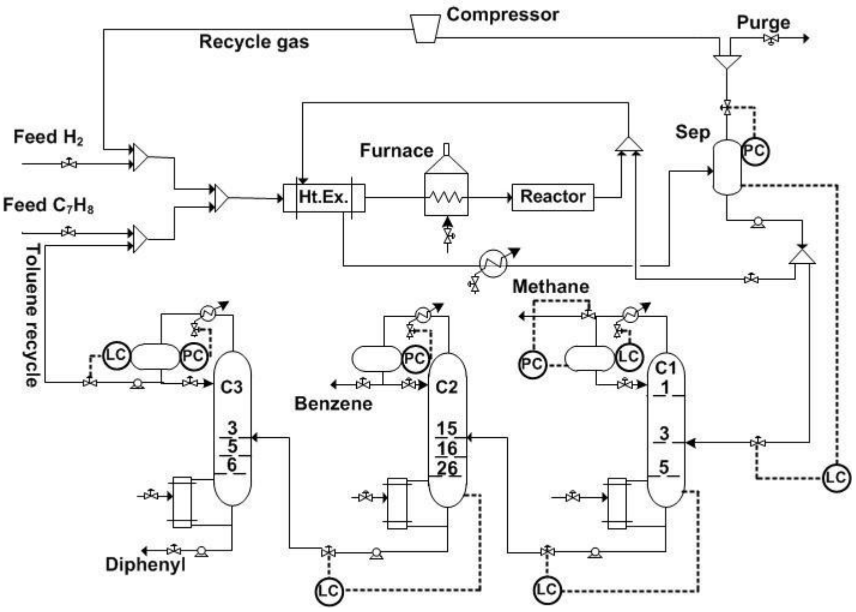


Figure 5. 32 The control structure obtained in step 5.1

Step 5.2 Adding simple enhanced controls. The reactor feed ratio of hydrogen to aromatics must be greater than 5: l. To accomplish this, a ratio control is added as show in Figure 5.33

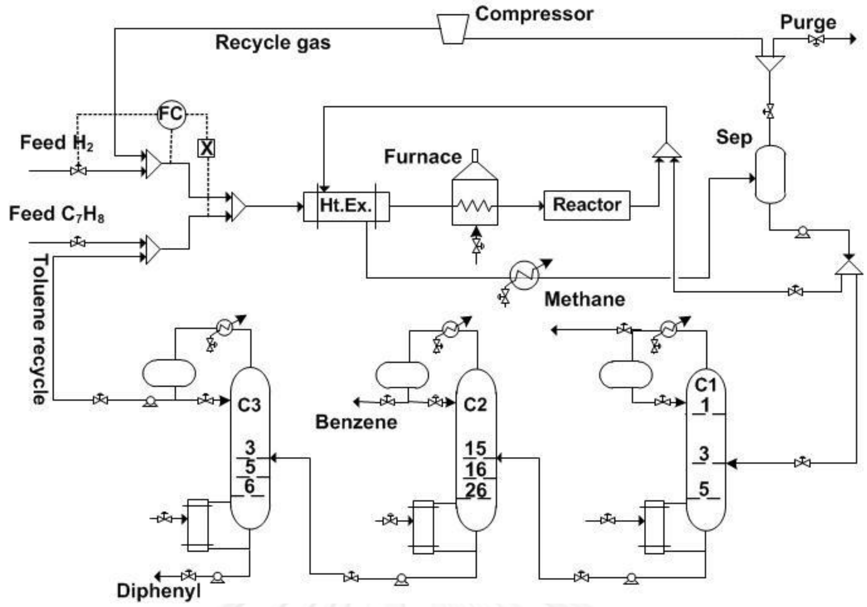


Figure 5. 33 The control structure obtained in step 5.2

4.Energy Management and Optimization

The design steps in this stage are omitted due to the scope of study.

Stage5. Design Validation

Step 8: Validate the designed control structures by rigorous dynamic simulation via HYSYS process simulation software.

The step change and pulse disturbances are used to test the response of the system: reactor inlet temperature, combine feed toluene, and hydrogen composition in makeup feed. The step change disturbances are changed at 20 min until 420 min and the pulse disturbances are changed at 20 min until 40 min. For step change disturbances, we are interested in how effectively the control structures designed cope with them, in particular the product qualities are able to be maintained. For pulse disturbance, we are interested in response speed of handle disturbances. The four control structures: Luyben’s design CS0 and our designs CS1, CS2, CS3 shown in Figure 5.34, 5.35, 5.36 and 5.37 CS1 and CS2 differ in C3 reflux control. Reflux ratio is controlled in CS1 and reflux-to-feed is control in CS2. Table 5.6 presents summation of detail of disturbance. Table 5.7 presents controlled variable (CV) and manipulated variable (MV) for hydrodealkylation process.

Table 5.6 Summation of detail of disturbance.

Disturbance	Change
Combine feed toluene flow rate	-15%
Reactor inlet temperature	+5˚F
Fresh feed hydrogen composition	-0.025

Table 5.7 Controlled variable (CV) and Manipulated variable (MV) for hydrodealkylation process.

Equipment	CS0		CS1	
	CV	MV	CV	MV
Fresh feed H_2	Pressure in separator	Fresh feed H_2 flow rate	Combine feed H_2 flow rate	Fresh feed H_2 flow rate
Fresh feedC_7H_8	Combine feed C_7H_8 flow rate	Fresh feed C_7H_8 flow rete	Combine feed C_7H_8 flow rate	Fresh feed C_7H_8 flow rate
Reactor	Inlet reactor temperature	Furnace duty	Reactor inlet temperature	Furnace duty
Liquid quench	Reactor effluent temperature	Liquid quench flow rate	Reactor effluent temperature	Liquid quench flow rate
Separator			Separator pressure	Separator top stream
	Separator liquid	Bottom flow rate	Separator Liquid	Bottom flow rate
	Separator liquid temperature	Cooler duty	Separator liquid temperature	Cooler duty
Purge	Recycle gas CH_4 composition	Purge flow rate	Separator top CH_4 composition	Purge flow rate
C1	Reflux drum level	Condenser duty	Reflux drum level	Condenser duty
	Reflux drum pressure	Vapor flow rate	Reflux drum pressure	Vapor flow rate
	State 1 temperature	Reboiler duty	State 1 temperature	Reboiler duty
	Reflux flow rate	Reflux flow rate	Reflux ratio	Reflux flow rate
	Base column level	Bottom flow rate	Base Column level	Bottom flow rate
C2	Reflux drum level	Distillate flow rate	Reflux drum level	Distillate flow rate
	Reflux drum pressure	Condenser duty	Reflux drum pressure	Condenser duty
	State 16 temperature	Reboiler duty	State 16 temperature	Reboiler duty
	Reflux flow rate	Reflux flow rate	Reflux to feed	Reflux flow rate
	Base column level	Bottom flow rate	Base column level	Bottom flow rate
C3	Reflux drum level	Distillate flow rate	Reflux drum level	Distillate flow rate
	Reflux drum pressure	Condenser duty	Reflux drum pressure	Condenser duty
	State 5 temperature	Bottom flow rate	State 5 temperature	Reboiler duty
	Reflux flow rate	Reflux flow rate	Reflux ratio	Reflux flow rate
	Base Column level	Reboiler duty	Base column level	Bottom flow rate

Table 5.7 Controlled variable (CV) and Manipulated variable (MV) for hydrodealkylation process. (Continued)

Equipment	CS2		CS3	
	CV	MV	CV	MV
Fresh feed H_2	Combine feed H_2 flow rate	Fresh feed H_2 flow rate	Ratio H_2 to C_7H_8	Fresh feed H_2 flow rate
Fresh feedC_7H_8	Combine feed C_7H_8 flow rate	Fresh feed C_7H_8 flow rate	Combine feed C_7H_8 flow rate	Fresh feed C_7H_8 flow rate
Reactor	Reactor inlet temperature	Furnace duty	Reactor inlet temperature	Furnace t duty
Liquid quench	Reactor effluent temperature	Liquid quench flow rate	Reactor effluent temperature	Liquid quench flow rate
Separator	Separator pressure	Separator top stream	Separator pressure	Separator top stream
	Separator Liquid	Bottom flow rate	Separator Liquid	Bottom flow rate
	Separator liquid temperature	Cooler duty	Separator liquid temperature	Cooler duty
Purge	Separator top CH_4 composition	Purge flow rate	Separator top CH_4 composition	Purge flow rate
C1	Reflux drum level	Condenser duty	Reflux drum level	Condenser duty
	Reflux drum pressure	Vapor flow rate	Reflux drum pressure	Vapor flow rate
	State 1 temperature	Reboiler duty	State 1 temperature	Reboiler duty
	Reflux ratio	Reflux flow rate	Reflux ratio	Reflux flow rate
	Base Column level	Bottom flow rate	Base Column level	Bottom flow rate
C2	Reflux drum level	Distillate flow rate	Reflux drum level	Distillate flow rate
	Reflux drum pressure	Condenser duty	Reflux drum pressure	Condenser duty
	State 16 temperature	Reboiler duty	State 16 temperature	Reboiler duty
	Reflux to feed	Reflux flow rate	Reflux to feed	Reflux flow rate
	Base column level	Bottom flow rate	Base column level	Bottom flow rate
C3	Reflux drum level	Distillate flow rate	Reflux drum level	Distillate flow rate
	Reflux drum pressure	Condenser duty	Reflux drum pressure	Condenser duty
	State 5 temperature	Reboiler duty	State 5 temperature	Reboiler duty
	Reflux to feed	Reflux flow rate	Reflux ratio	Reflux flow rate
	Base column level	Bottom flow rate	Base column level	Bottom flow rate

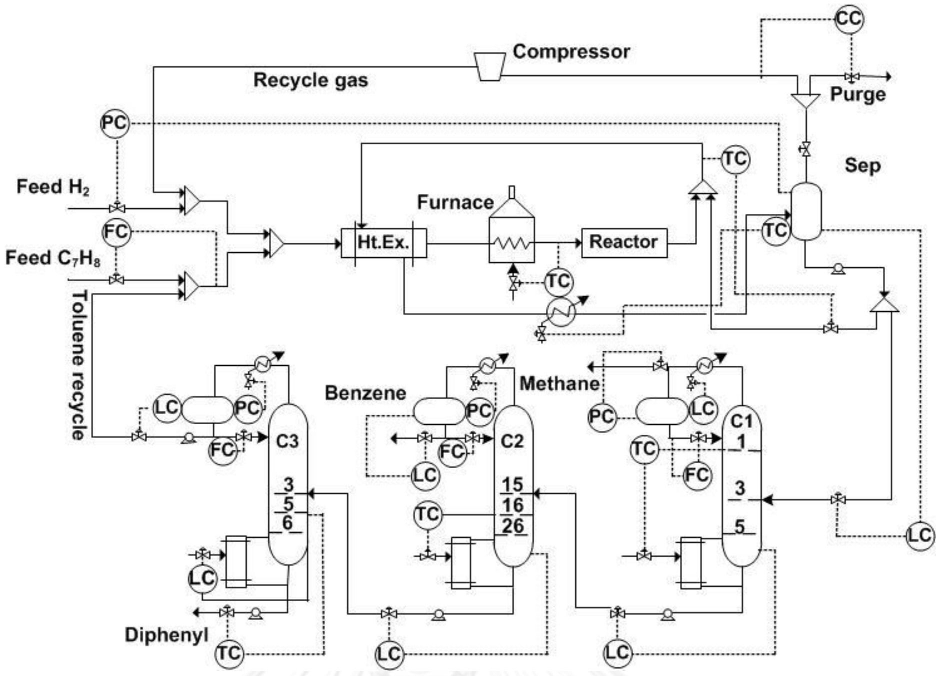


Figure 5. 34 Control structure 0 (CS0).

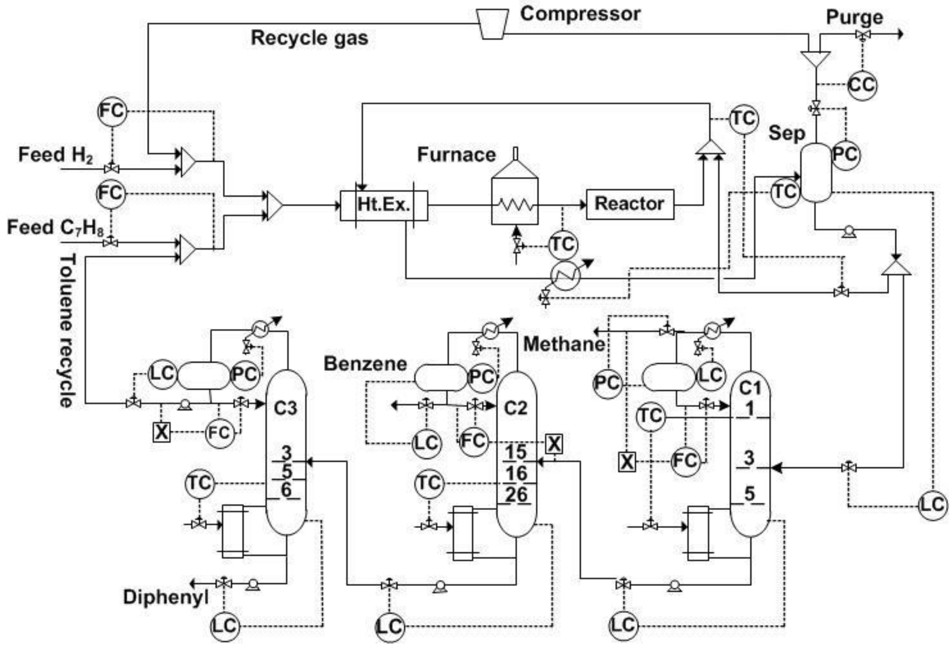


Figure 5. 35 Control structure 1 (CS1)

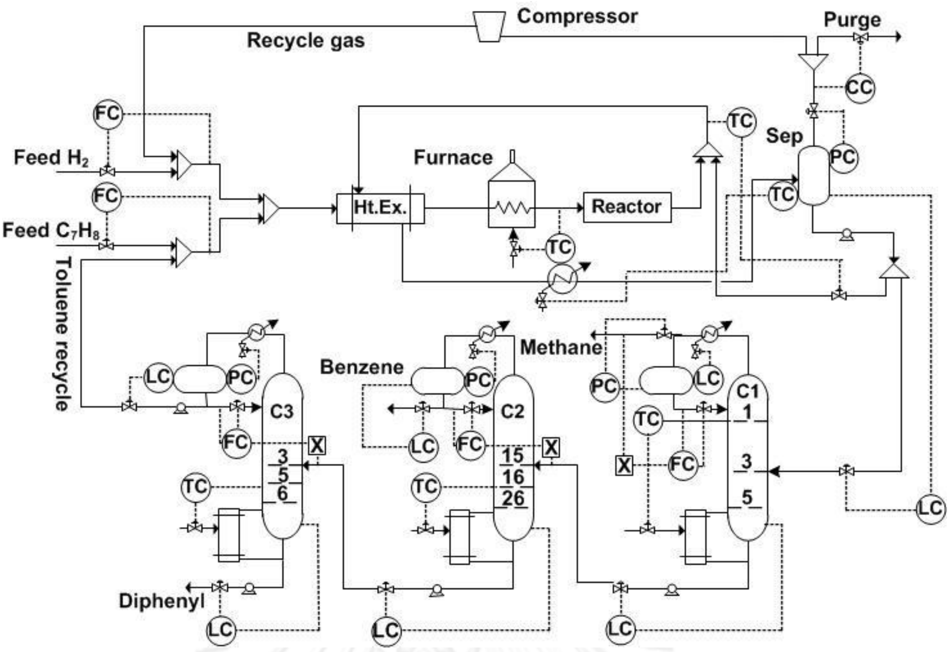


Figure 5. 36 Control structure 2 (CS2)

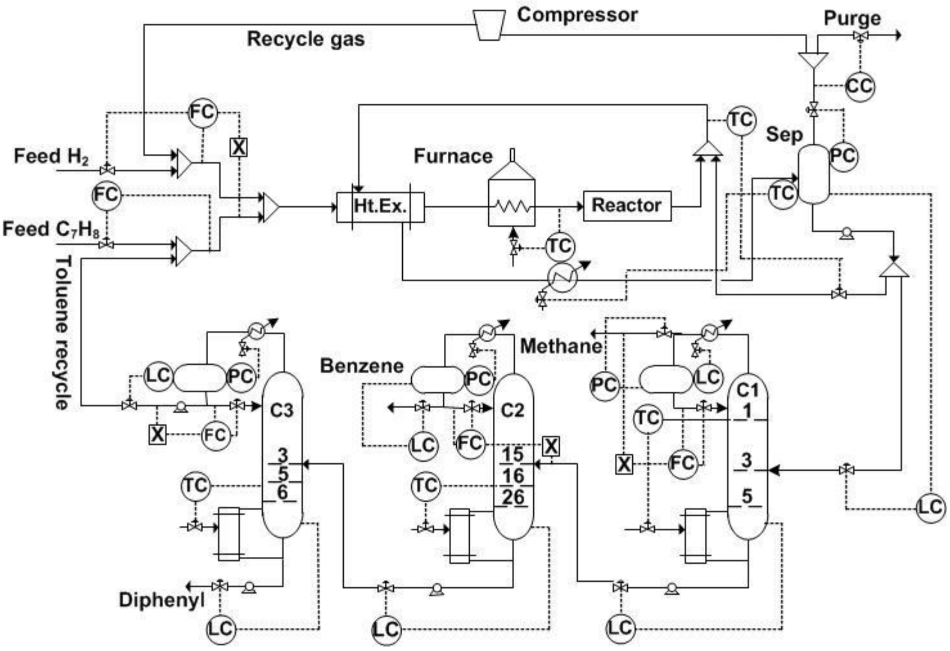
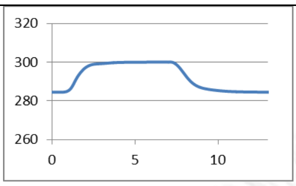
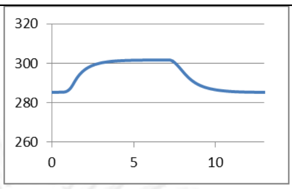
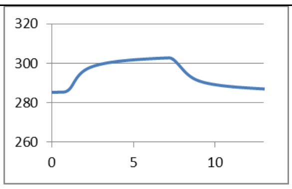
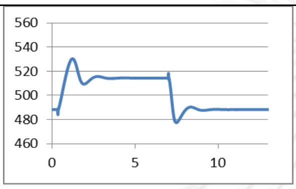
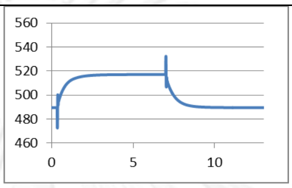
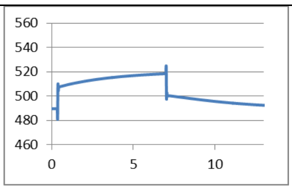
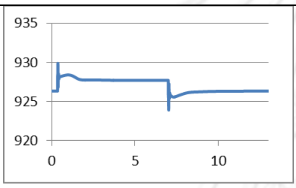
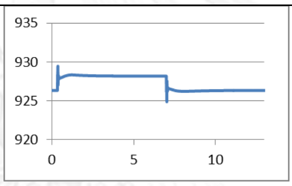
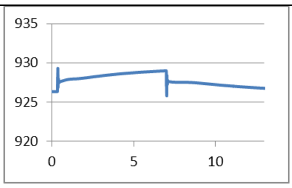
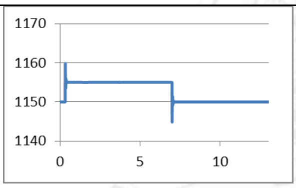
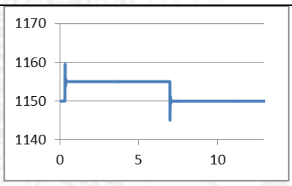
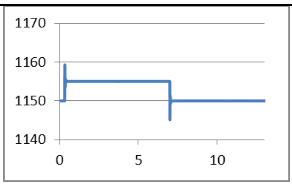
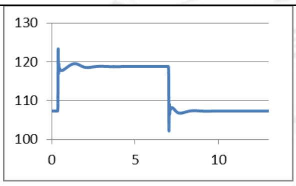
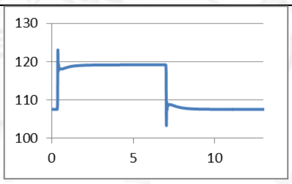
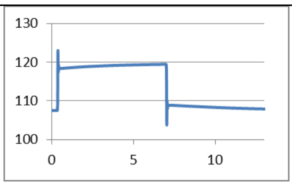
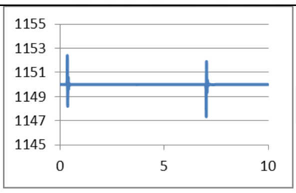
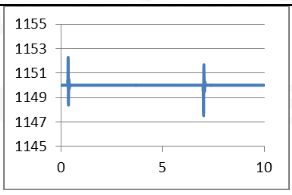
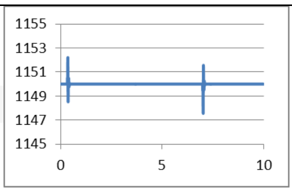
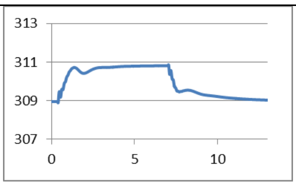
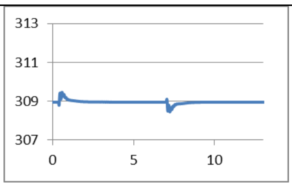
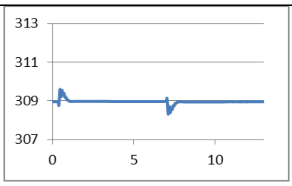
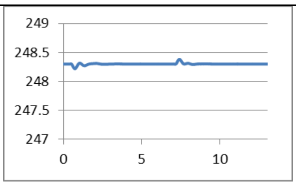
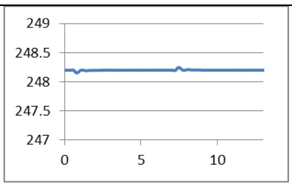
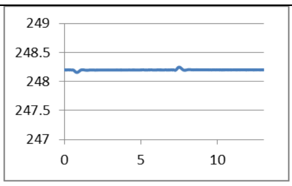


Figure 5. 37 Control structure 3 (CS3).

5.2 Dynamic simulation results

Table 5.8 The dynamic responses for handle step change disturbances to the change reactor inlet temperature.

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace Iinet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb- mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			
(h) C2 Tray 16 (°F)			

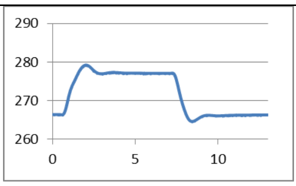
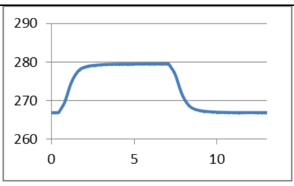
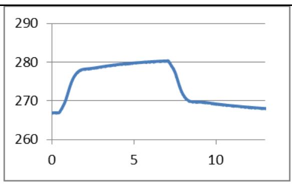
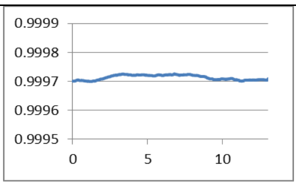
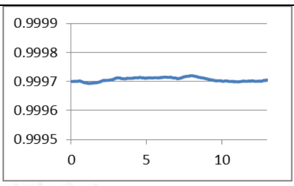
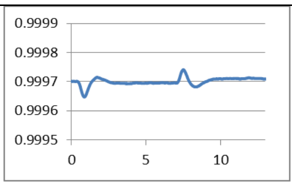
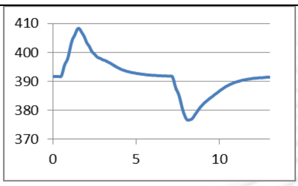
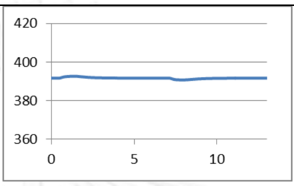
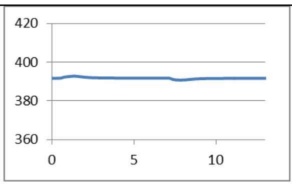
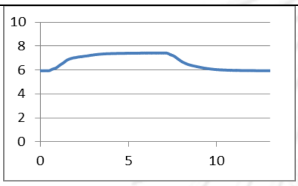
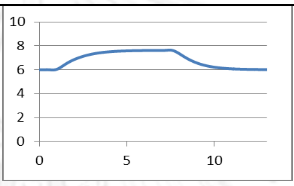
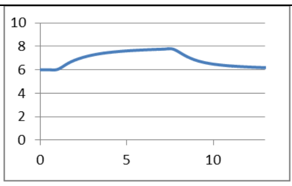
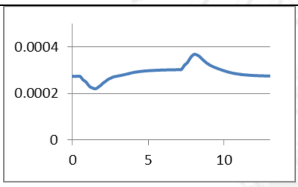
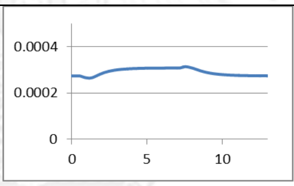
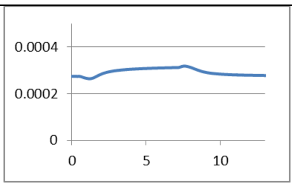
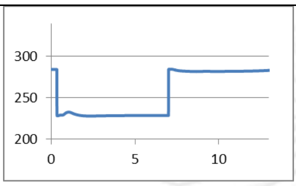
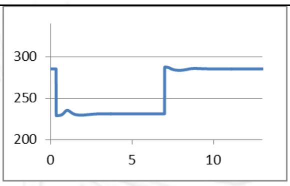
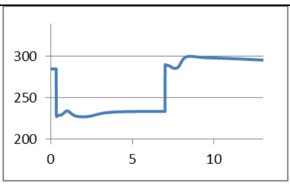
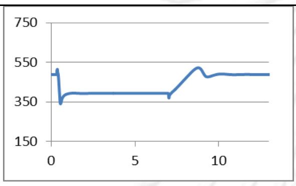
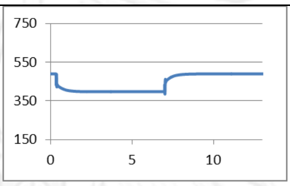
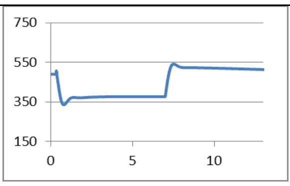
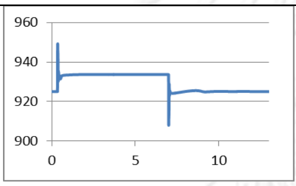
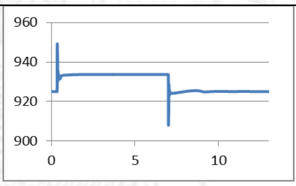
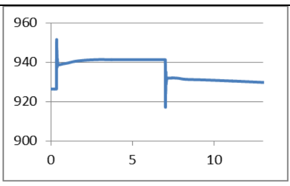
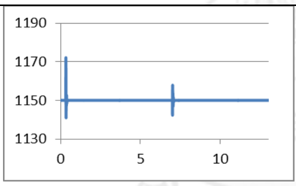
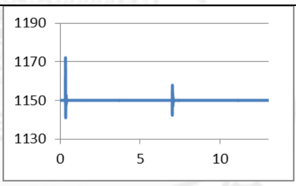
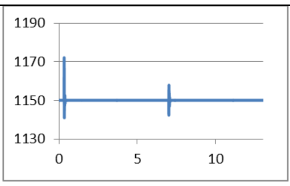
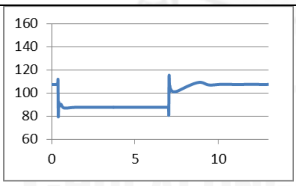
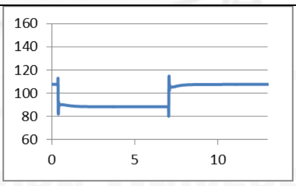
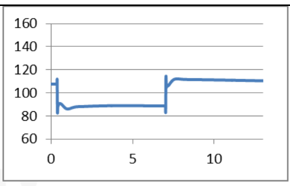
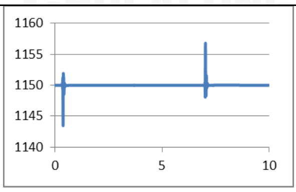
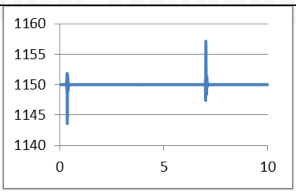
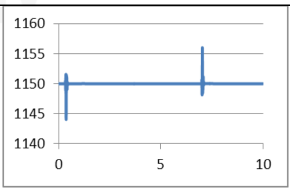
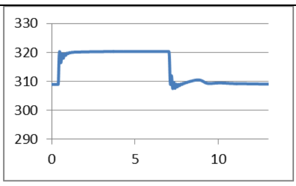
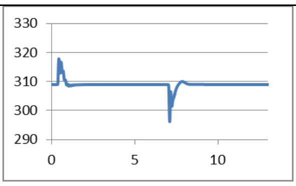
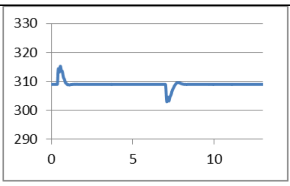
(i) D2 (lb-mole/hr)			
(j) xD2 (Benzene)			
(k) C3 Tray 5 (°F)			
(l) B3 (lb-mole/hr)			
(m) xB3 (Toluene)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.8 gives results for +5 °F changes in reactor inlet temperature at 20 min to 420 min. Fresh feed hydrogen (b) and liquid quench flow rate form separator (e) of CS0 oscillate more than CS2 and CS3 because CS0 is controlled separator pressure by manipulate fresh feed hydrogen which is slowly speed response than CS2 and CS3 that is controlled flow rate of combine feed hydrogen by manipulate fresh feed hydrogen. In basecase structure, column C1 use P controller. Therefore, the 1st tray temperature (g) of CS0 changes from 309 °F to 311 °F. The production rate (i) of CS0 is smaller than CS2 and CS3 since benzene product of CS0 is disappeared in column C1. Additionally, The benzene product quality in distillated (j) of CS2 and CS3 are quite close their setpoint than CS0. Furthermore, CS2 and CS3 are able to control composition toluene in by-product (m) and column C3 tray temperature (k) better than CS0 since CS2 and CS3 are controlled column C3 tray

temperature by manipulate reboiler duty but CS0 is controlled temperature column C3 tray temperature by bottom flow that it is slowly adjust temperature.

Table 5.9 The dynamic responses for handle step change disturbances to the change in combine feed toluene flow rate.

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace Iinet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb-mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			

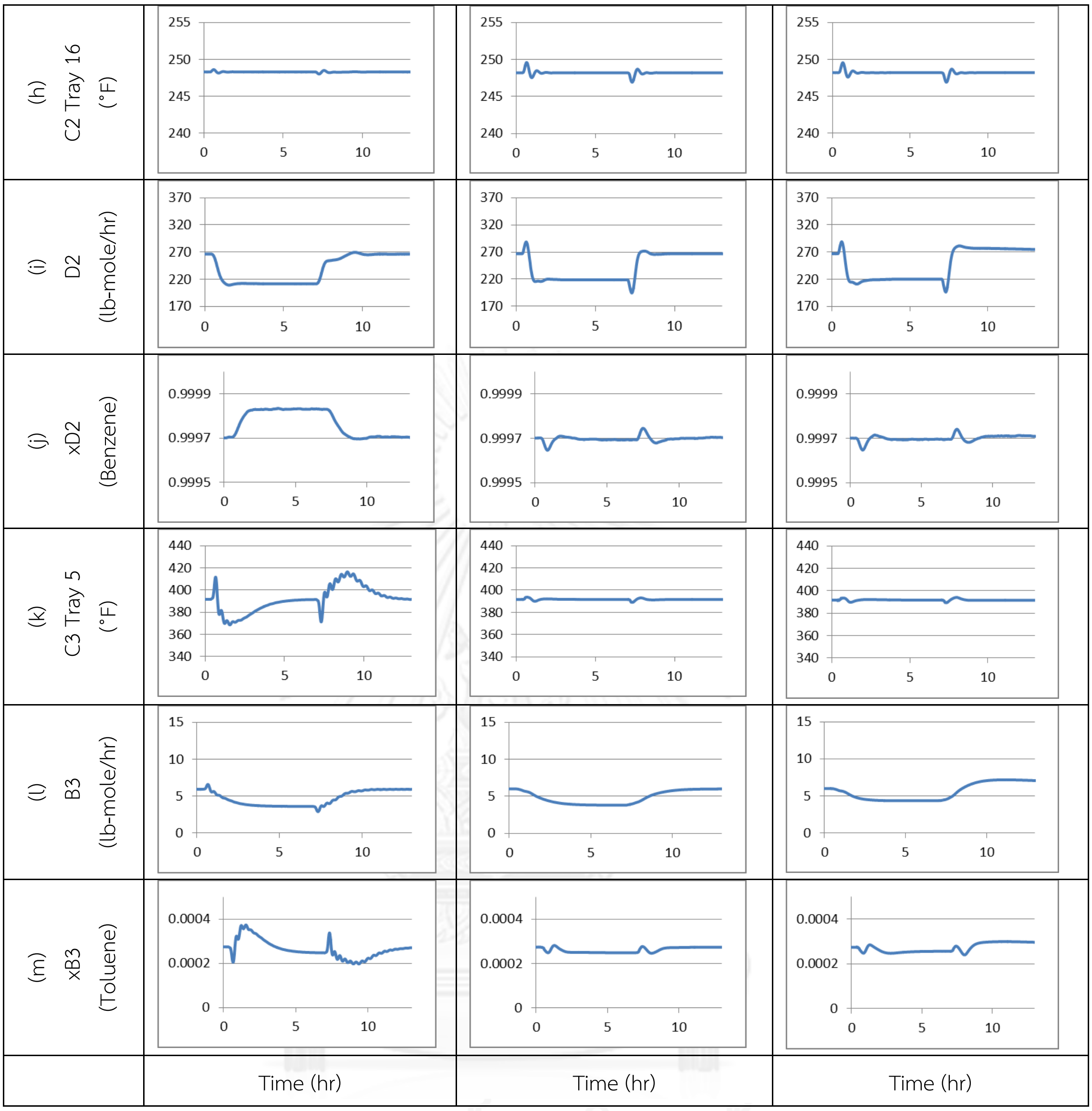
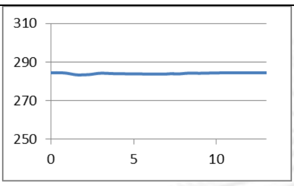
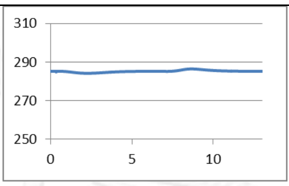
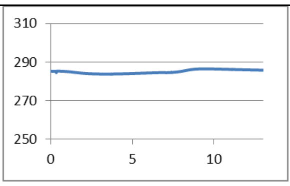
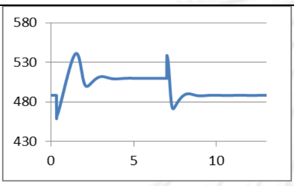
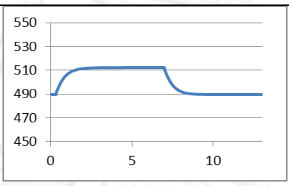
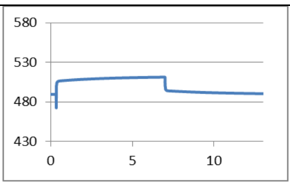
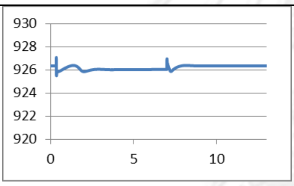
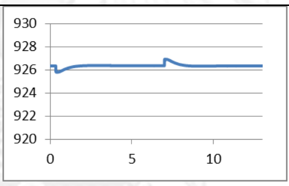
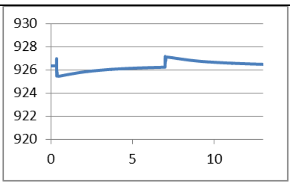
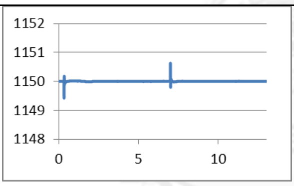
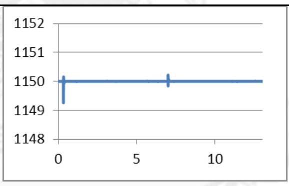
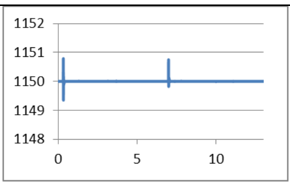
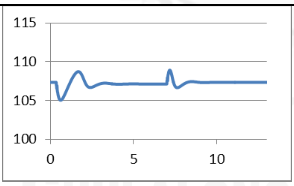
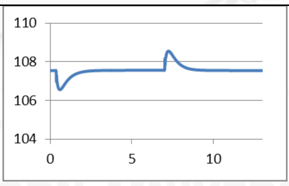
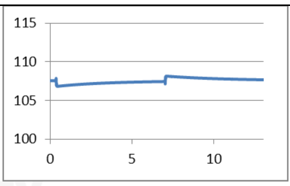
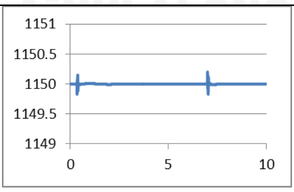
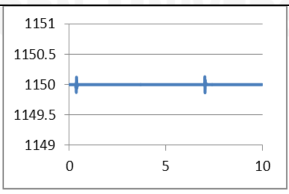
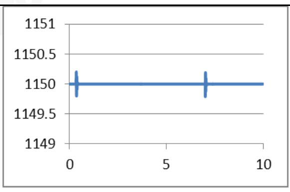
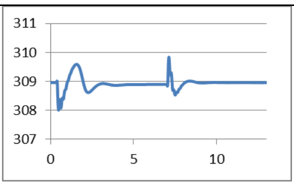
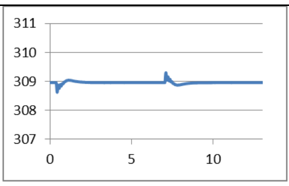
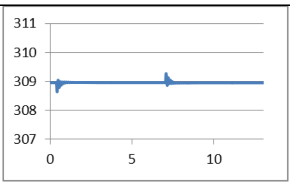


Table 5.9 presents results for -15% changes in combined feed toluene at 20 min to 420 min. Fresh feed hydrogen flow rate (b), Furnace inlet temperature (c) and liquid quench flow rate (e) form separator of CS0 oscillate more than CS2 and CS3 because of same reason with handle step change disturbances to the change reactor inlet temperature. In basecase structure, the 1st tray temperature (g) of CS0 changes from 309 °F to 320 °F. The production rate (i) of CS0 is smaller than CS2 and CS3 because benzene product of CS0 is disappeared in column C1. Additionally, The benzene product quality in distillated (j) of CS2 and CS3 are quite close their setpoint than CS0. Furthermore, CS2 and CS3 are controlled composition toluene in by-product (m) and column C3 tray temperature (k) better than CS0 since because of

same reason with handle step change disturbances to the change reactor inlet temperature.

Table 5.10 The dynamic responses for handle step change disturbances to the change in fresh feed hydrogen composition.

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace Iinet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb-mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			

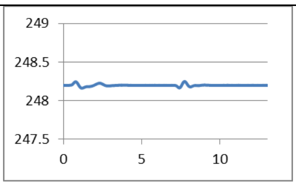
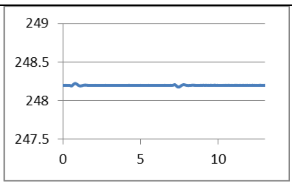
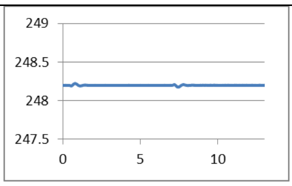
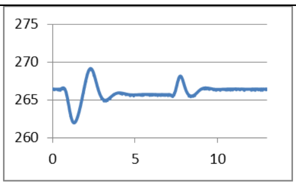
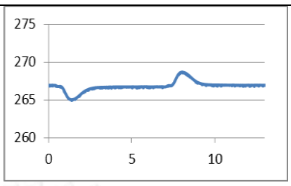
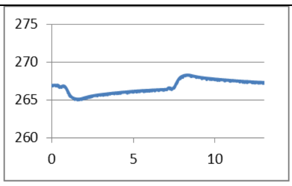
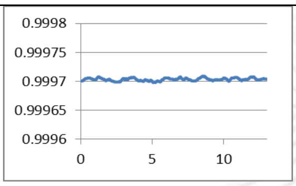
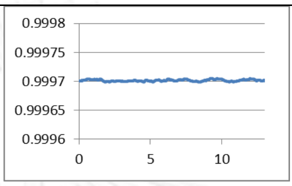
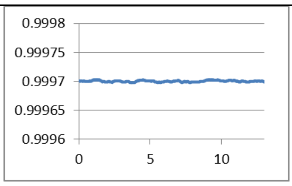
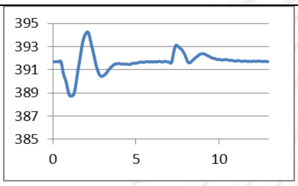
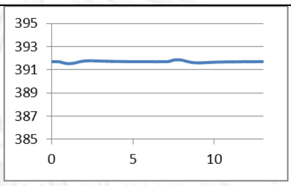
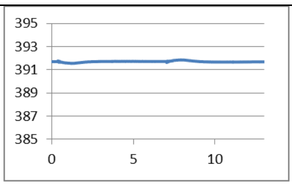
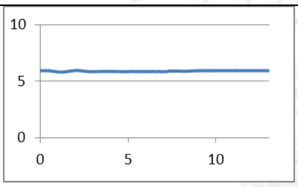
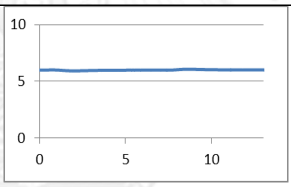
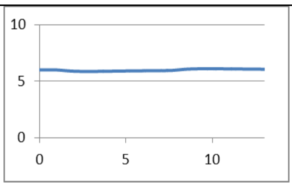
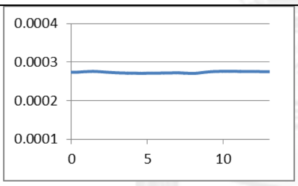
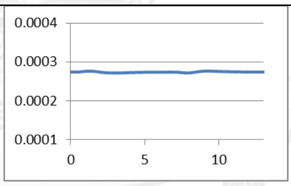
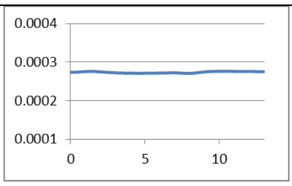
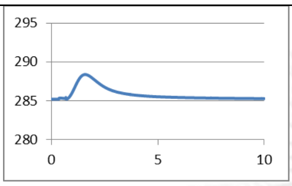
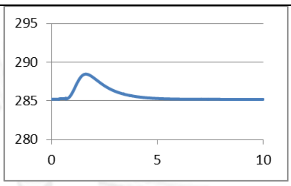
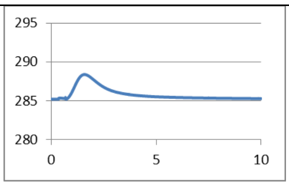
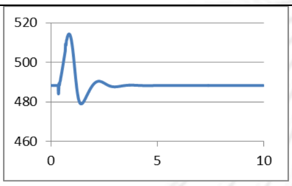
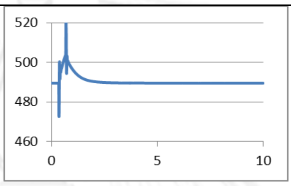
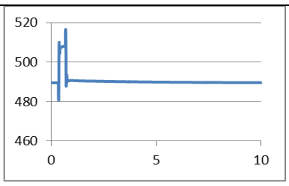
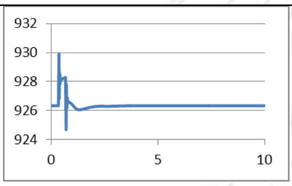
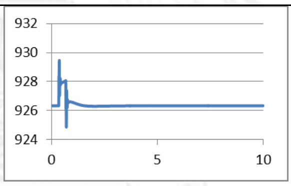
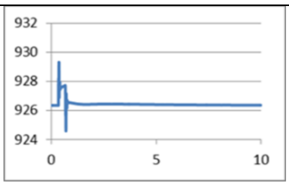
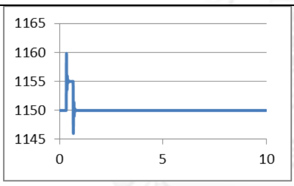
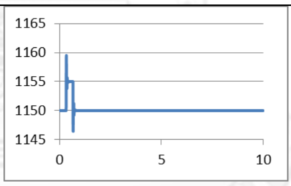
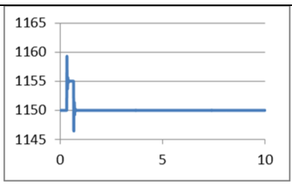
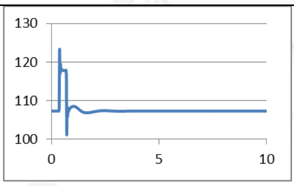
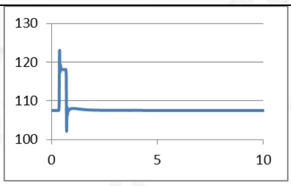
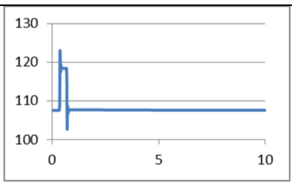
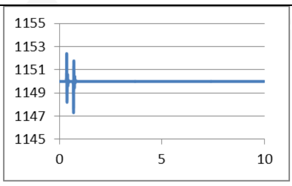
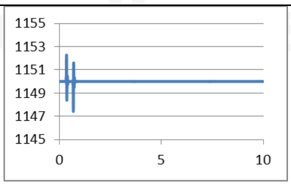
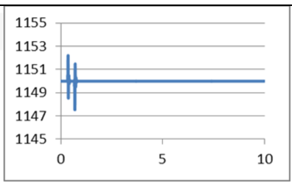
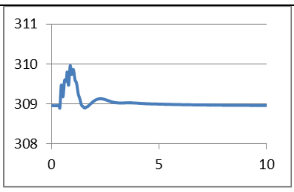
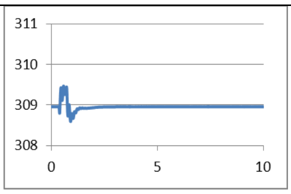
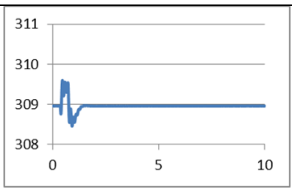
(h) C2 Tray 16 (°F)			
(i) D2 (lb-mole/hr)			
(j) xD2 (Benzene)			
(k) C3 Tray 5 (°F)			
(l) B3 (lb-mole/hr)			
(m) xB3 (Toluene)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.10 shows results for -0.025 changes in fresh feed hydrogen composition at 20 min to 420 min. Fresh feed hydrogen flow rate (b), Furnace inlet temperature(c) and liquid quench flow rate form separator(e) of CS0 oscillate more than CS2 and CS3 because of same reason with handle step change disturbances to the change reactor inlet temperature. The column C1 tray temperature (g) of CS0 is oscillating more than CS2 and CS3 since liquid quench flow rate form separator is decreased/increased for maintain outlet reactor temperature. Therefore, the column C1 feed flow rate is change following separator level. Additionally, The product purity in distillated (j) of all control structures are quite close their setpoint. Furthermore, CS2 and CS3 are controlled composition toluene in by-product(m) and

column C3 tray temperature(k) better than CS0 because of same reason with handle step change disturbances to the change reactor inlet temperature

Table 5.11 The dynamic responses for handle pulse disturbances to the change reactor inlet temperature

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace Iinet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb-mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			

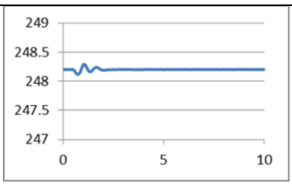
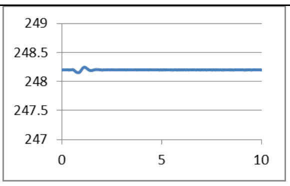
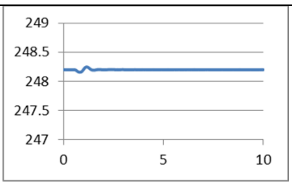
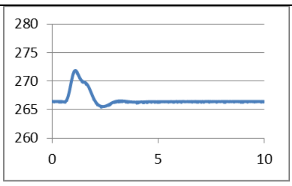
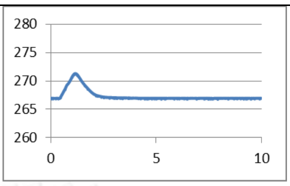
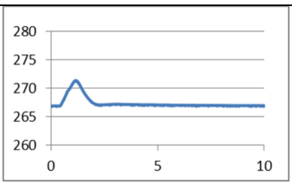
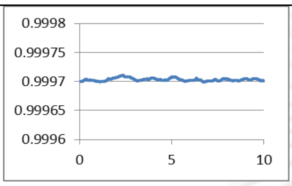
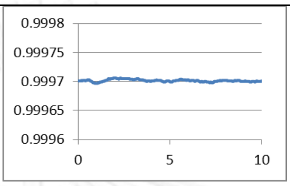
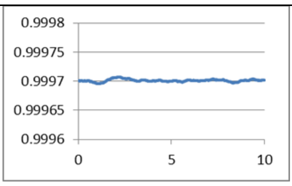
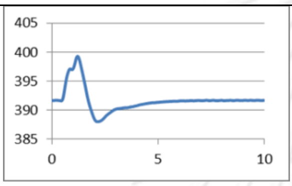
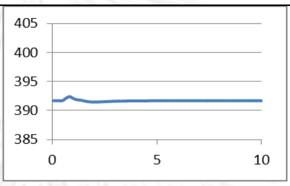
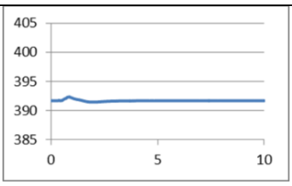
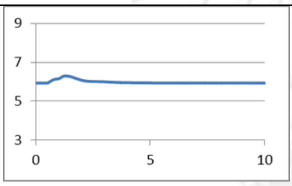
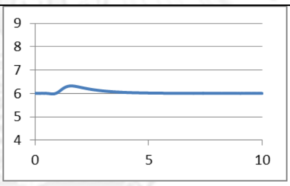
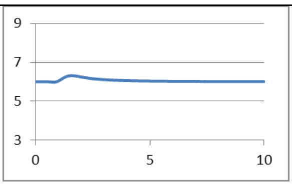
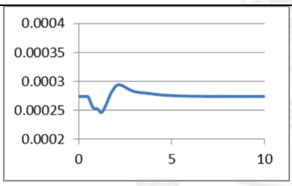
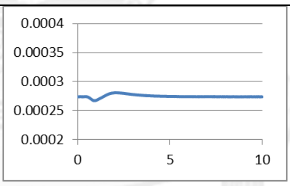
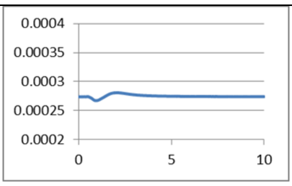
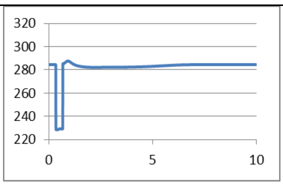
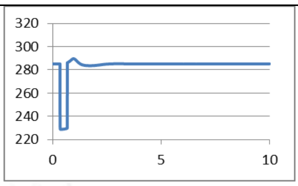
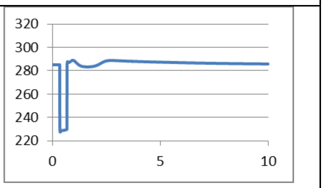
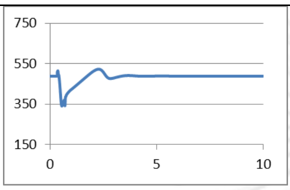
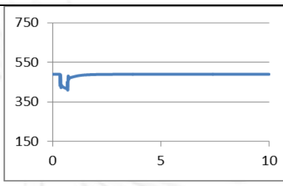
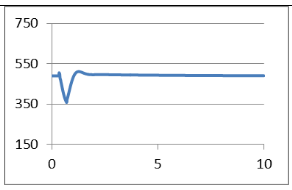
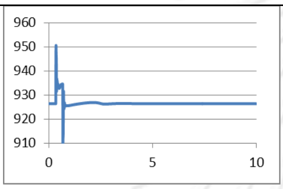
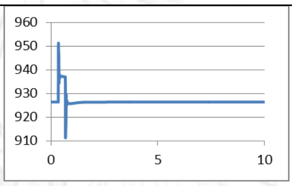
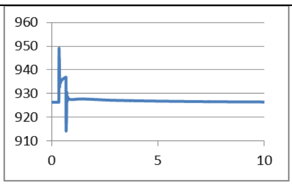
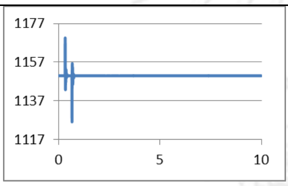
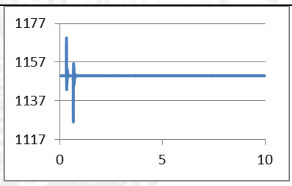
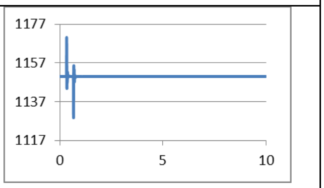
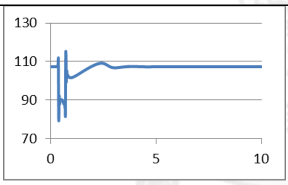
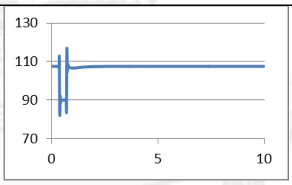
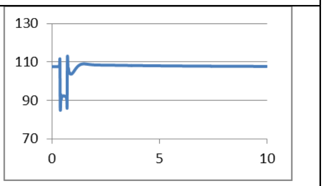
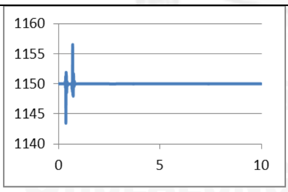
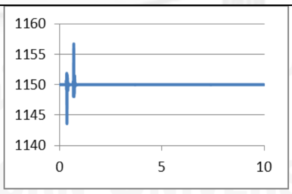
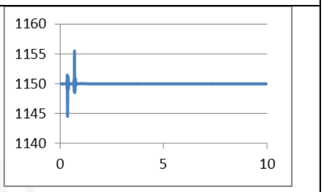
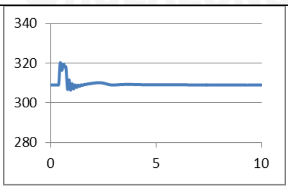
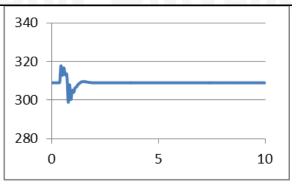
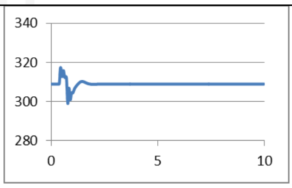
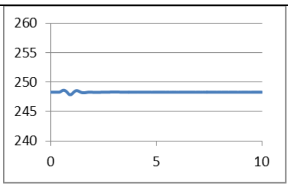
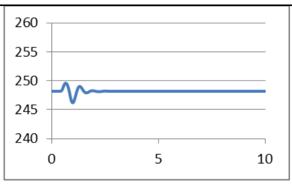
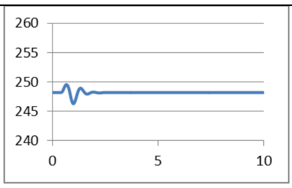
(h) C2 Tray 16 (°F)			
(i) D2 (lb-mole/hr)			
(j) xD2 (Benzene)			
(k) C3 Tray 5 (°F)			
(l) B3 (lb-mole/hr)			
(m) xB3 (Toluene)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.11 presents results for +5 °F changes in reactor inlet temperature at 20 min to 40 min. CS2 and CS3 are quickly able to settle setpoint than CS0 because establish the fixture plant and disturbance management. Fresh feed hydrogen (b), furnace inlet temperature (c) and separator liquid quench (e), column C1 tray temperature (g), benzene production rate (i), column C3 tray temperature (k) and toluene composition in C3 bottom flow (m) of CS3 and CS2 able to return the original value quicker than CS0.

Table 5.12 The dynamic responses for handle pulse disturbances to the change in combine feed toluene flow rate.

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace Iinet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb-mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			
(h) C2 Tray 16 (°F)			

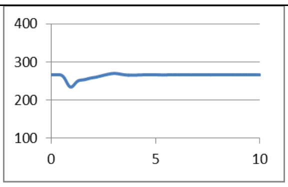
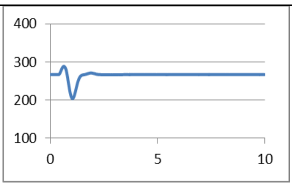
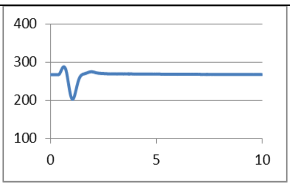
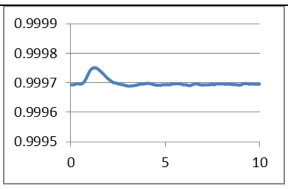
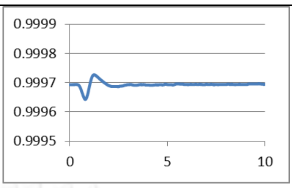
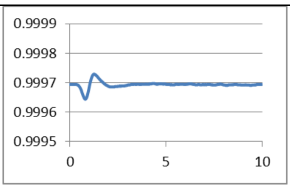
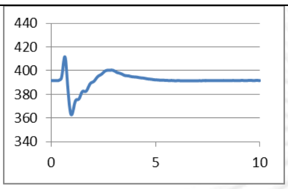
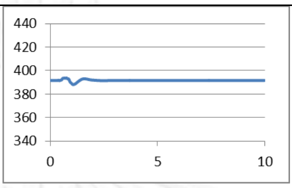
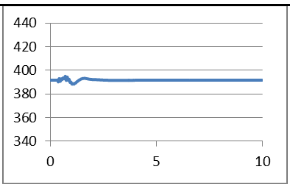
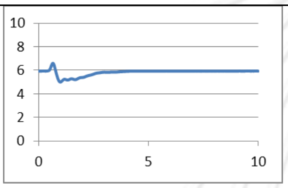
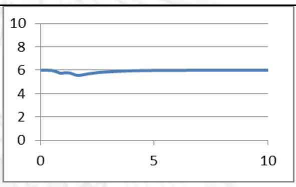
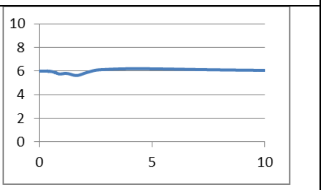
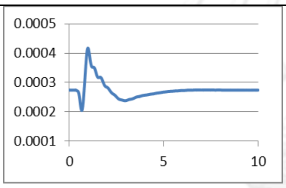
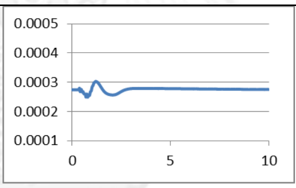
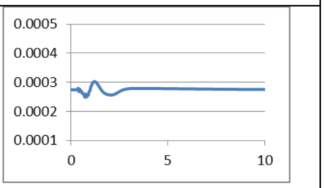
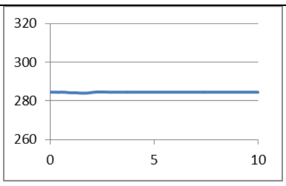
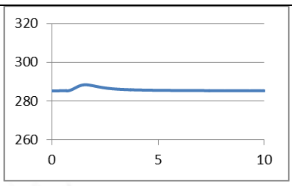
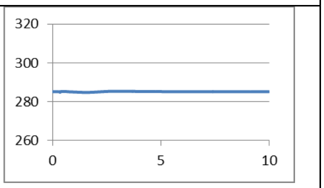
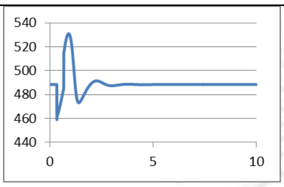
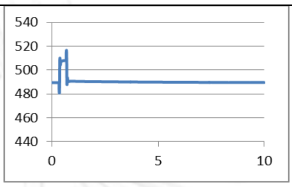
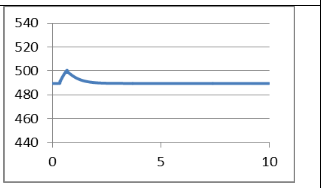
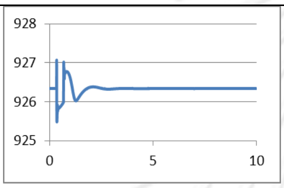
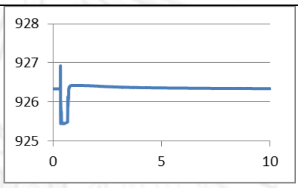
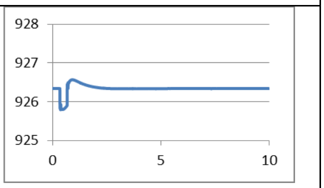
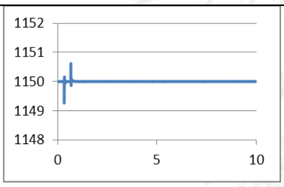
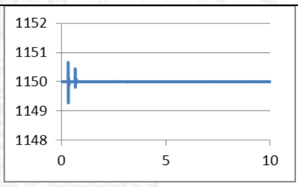
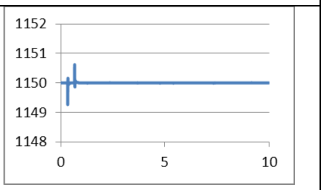
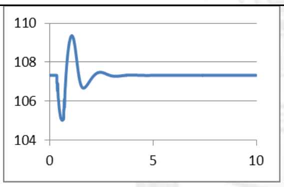
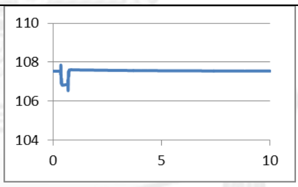
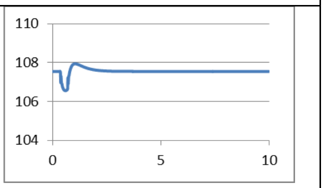
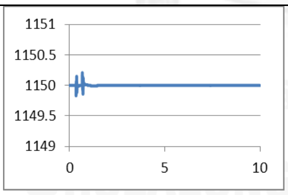
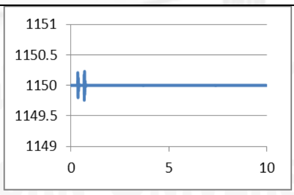
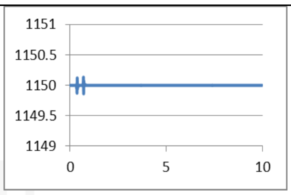
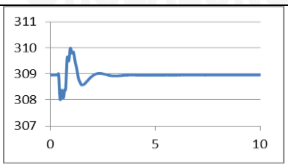
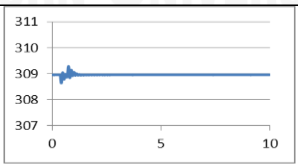
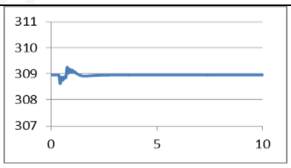
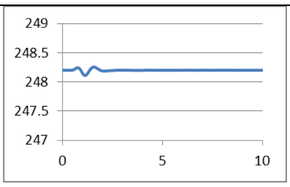
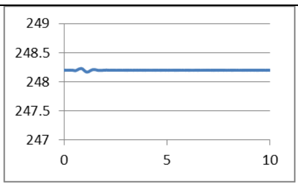
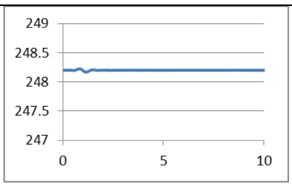
(i) D2 (lb-mole/hr)			
(j) xD2 (Benzene)			
(k) C3 Tray 5 (°F)			
(l) B3 (lb-mole/hr)			
(m) xB3 (Toluene)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.12 gives results for -15% changes combined feed toluene at 20 min to 40 min. CS2 and CS3 are able to settle setpoint than CS0. Fresh feed hydrogen (b), furnace inlet temperature (c), liquid quench flow rate (e), column C1 tray temperature (g), benzene production rate (i), column C3 tray temperature (k) and toluene composition in C3 bottom flow (m)of CS3 and CS2 able to return the original value quicker than CS0.

Table 5.13 The dynamic responses for handle pulse disturbances to the change in fresh feed hydrogen composition.

	CS0	CS2	CS3
(a) FFtot (lb-mole/hr)			
(b) FFH2 (lb-mole/hr)			
(c) Furnace linet (°F)			
(d) Temp. Rin (°F)			
(e) Liquid quench (lb-mole/hr)			
(f) Temp. Quench (°F)			
(g) C1 Tray 1 (°F)			
(h) C2 Tray 16 (°F)			

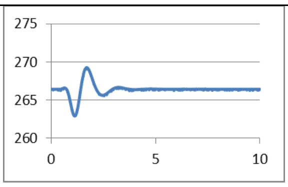
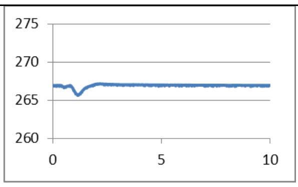
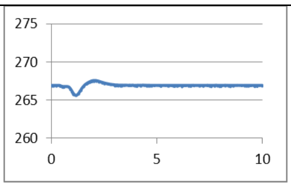
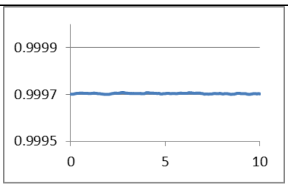
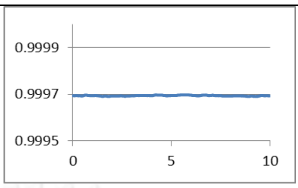
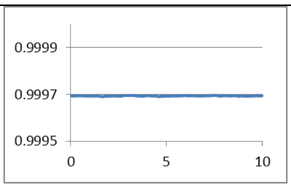
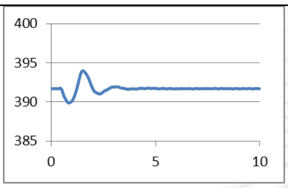
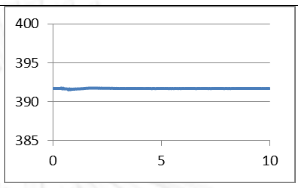
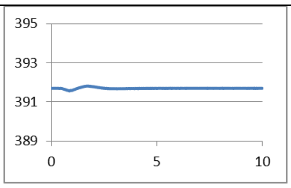
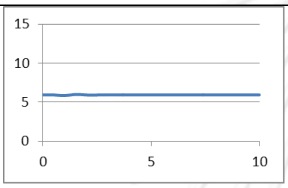
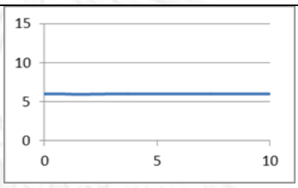
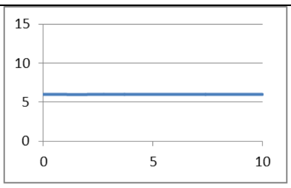
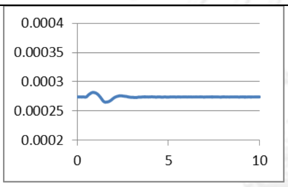
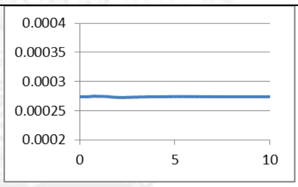
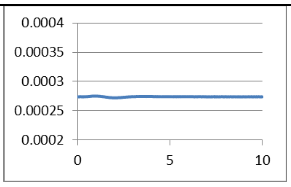
(i) D2 (lb-mole/hr)			
(j) xD2 (Benzene)			
(k) C3 Tray 5 (°F)			
(l) B3 (lb-mole/hr)			
(m) xB3 (Toluene)			
	Time (hr)	Time (hr)	Time (hr)

Table 5.13 shows results for -0.025 changes fresh feed hydrogen composition at 20 min to 40 min. Fresh feed hydrogen (b), furnace inlet temperature (c), liquid quench flow rate (e), column C1 tray temperature (g), benzene production rate (i), column C3 tray temperature (k) and toluene composition in C3 bottom flow (m) of CS3 and CS2 able to return the original value quicker than CS0.

Dynamic responses of CS1 and CS2

Responses of CS1 and CS2 are similar in all disturbances because CS1 and CS2 differ in column C3 and dynamic behaviors of CS1 and CS2 for handle material disturbances in step 4.2 are similar in all disturbances.

5.3 Control Structure Performance Evaluation

Control structure performance evaluation is important to select control structure. It used IAE to compare the control structures by effective control structure will give minimum IAE. We considered IAE result with temperature control loops, pressure control loops and composition control loops. Information of IAE results were presented in appendix c.

Table 5. 14 Summation of the IAE results for handle reactor inlet temperature step change disturbance

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	62.9156	6.2382	0.4410
CS1	3.3930	2.0826	0.0542
CS2	3.2601	2.0529	0.0542
CS3	3.4455	1.8467	0.0694

Table 5. 15 Summation of the IAE results for handle combine feed toluene flow rate step change disturbance.

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	111.6454	18.2913	1.5031
CS1	4.9254	6.8028	0.1176
CS2	5.2948	6.8390	0.1176
CS3	6.8864	7.5797	0.4745

Table 5. 16 Summation of the IAE results for handle fresh feed hydrogen composition step change disturbance.

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	6.4628	4.7491	0.0678
CS1	0.8113	0.6840	0.0211
CS2	0.5878	0.6850	0.0211
CS3	0.5831	0.6359	0.0189

The control structures of hydrodealkylation process are assessed by IAE (Integral Absolute Error). The summations of IAE of CS1, CS2 and CS3 have better efficient than CS0 in all step change disturbances. The IAE values between CS1 and CS2 are not significantly different in all disturbances. The IAE of CS3 gives effective performance for handle fresh feed hydrogen composition disturbance. However CS0 and CS1 yield effective performance for handle reactor inlet temperature disturbance and handle combine feed toluene disturbance. CS2 is the best performance for handle reactor inlet temperature step change disturbance. CS1 is the best performance for handle combine feed toluene flow rate step change disturbance. Finally, CS3 is the best performance for handle fresh feed hydrogen composition step change disturbance.

Table 5. 17 Summation of the IAE results for handle reactor inlet temperature pulse disturbance

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	10.9920	1.9067	0.0407
CS1	0.8022	0.8144	0.0096
CS2	0.7904	0.8119	0.0096
CS3	0.7939	0.7236	0.0060

Table 5. 18 Summation of the IAE results for handle combine feed toluene flow rate pulse disturbance.

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	33.9169	9.7879	0.1845
CS1	3.9485	4.3082	0.0575
CS2	4.2514	4.3369	0.0575
CS3	4.7118	4.9519	0.0579

Table 5. 19 Summation of the IAE results for handle fresh feed hydrogen composition pulse disturbance.

Control Structures	Control loops		
	Temperature	Pressure	Composition
CS0	2.5987	2.2107	0.0340
CS1	0.2041	0.2172	0.0089
CS2	0.2172	0.2186	0.0089
CS3	0.1894	0.1361	0.0023

The summations of IAE of CS1, CS2 and CS3 have batter efficient than CS0 in all pulse disturbances. The IAE values between CS1 and CS2 in composition loop are not significantly different in all disturbances. CS3 is the best performance for handle reactor inlet temperature step change disturbance and handle fresh feed hydrogen composition step change disturbance. CS1 is the best performance for handle combine feed toluene flow rate step change disturbance.

CHAPTER VI

CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusion

The problem of plantwide control is to develop a control strategy for an entire complex that satisfies the plant's design objectives. hydrodealkylation process is very complex dynamic behavior. The plantwide control structure is important for hydrodealkylation process. The new design procedure of Wongsri provides a simple, detailed systematic and clear methodology to design plantwide control structure which emphasis on the establishment of a fixture plant by regulating material component y rates using their quantifiers and handlers on their pathways and executing disturbance management.

The results of establishing a fixture plant and disturbance management give control structures of process different with Luyben procedure such as stabilizer column, product column, recycle column, combined feed hydrogen and separator top flow. All new control structure is tested by three disturbances able to handle disturbance well. CS1, CS2 and CS3 can fast and well handle disturbances; keep product quality well, maintain operating condition smoothly when compared with basecase structure. The performances of each control structures are evaluated by analyzing the IAE value. Therefore, CS1, CS2 and CS3 are appropriate control structure for the hydrodealkylation process.

6.2 Recommendations

Study and control structure design should be applied to other process via design procedure of Wongsri (2012)

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APPENDIX

APPENDIX A

INFORMATION OF PROCESS STREAMS AND EQUIPMENTS DATA

In this section, the detail information of streams and equipment are presented in table A.1-A.3. They have important to convert a steady-state simulation into a dynamic simulation.

Table A. 1 Equipment data of hydrodealkylation process

Unit	Information	value
Reactor	Diameter (ft)	9.53
	Length (ft)	57
FEHE	Shell volume (ft^2)	500
	Tube volume (ft^2)	500
Furnace	Tube volume (ft^2)	40
Separator	Liquid volume (ft^3)	40
Stabilizer Column	Total theoretical trays	6
	Feed tray	3
	Pressure (psia)	150
	Diameter (ft)	1
	Reboiler vol. (ft^3)	250
	Condenser vol. (ft^3)	7.5
	Benzene mole fraction in overhead	0.0042
	Methane mole fraction in bottoms	0.000001
Product Column	Total theoretical trays	27
	Feed tray	15
	Pressure (psia)	30
	Diameter(ft)	5.7
	Reboiler vol. (ft^3)	316
	Condenser vol. (ft^3)	293
	Toluene mole fraction in overhead	0.0003
	Benzene mole fraction in bottoms	0.0006

Table A.1 Equipment data of hydrodealkylation process. (Continued)

Unit	Information	value
Recycle Column	Total theoretical trays	7
	Feed tray	5
	Pressure (psia)	30
	Diameter (ft)	2.5
	Reboiler vol. (ft^3)	46
	Condenser vol. (ft^3)	36
	Diphenyl mole fraction in overhead	0.00002
	Toluene mole fraction in bottoms	0.00026

Table A. 2 Stream table for hydrodealkylation process

Name	FFH2	V1out	Rgas	TotH2	FFtot
Stream Number	1	2	3	4	8
Temperature [F]	86.0000	85.9947	158.9996	150.8358	86.0000
Pressure [psi]	635.0000	605.0000	605.0000	605.0000	635.0000
Molar Flow[lbmole/hr]	489.4770	489.4770	3558.6777	4048.1547	286.5980
H_2 mole fraction	0.9700	0.9700	0.4035	0.4720	0.0000
CH_4 mole fraction	0.0300	0.0300	0.5854	0.5182	0.0000
C_6H_6 mole fraction	0.0000	0.0000	0.0100	0.0088	0.0000
C_7H_8 mole fraction	0.0000	0.0000	0.0012	0.0011	1.0000
$C_{12}H_{10}$ mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000

Name	V2out	Toltot	cHEin	cHEout	Rin
Stream Number	9	7	8	9	38
Temperature [F]	86.2070	136.7902	146.6993	926.3377	1150.0000
Pressure [psi]	605.0000	605.0000	605.0000	543.0000	503.0000
Molar Flow [lbmole/hr]	286.5980	372.5197	4420.6744	4420.6744	4420.6744
H_2 mole fraction	0.0000	0.0000	0.4322	0.4322	0.4322
CH_4 mole fraction	0.0000	0.0000	0.4745	0.4745	0.4745
C_6H_6 mole fraction	0.0000	0.0001	0.0080	0.0080	0.0080
C_7H_8 mole fraction	1.0000	0.9998	0.0852	0.0852	0.0852
$C_{12}H_{10}$ mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000

Table A.2 Stream table for hydrodealkylation process. (Continued)

Name	Rout	Quench	hHEin	hHEout	Coolout
Stream Number	11	12	13	14	15
Temperature [F]	1229.7679	113.8333	1149.8414	420.1943	113.0000
Pressure [psi]	486.0000	486.0000	486.0000	480.0000	476.8000
Molar Flow[lbmole/hr]	4420.6744	108.0254	4530.8487	4530.8487	4530.8487
H_2 mole fraction	0.3690	0.0047	0.3605	0.3605	0.3605
CH_4 mole fraction	0.5391	0.0447	0.5271	0.5271	0.5271
C_6H_6 mole fraction	0.0699	0.7074	0.0851	0.0851	0.0851
C_7H_8 mole fraction	0.0207	0.2275	0.0256	0.0256	0.0256
$C_{12}H_{10}$ mole fraction	0.0014	0.0157	0.0017	0.0017	0.0017

Name	Gas	V12out	Purge	V4out	Grecycle
Stream Number	16	17	18	19	20
Temperature [F]	113.0000	112.9760	112.9760	112.9568	112.9760
Pressure [psi]	476.8000	475.8000	475.8000	475.0000	475.8000
Molar Flow[lbmole/hr]	4042.7261	4042.7261	484.0484	484.0484	3558.6777
H_2 mole fraction	0.4035	0.4035	0.4035	0.4035	0.4035
CH_4 mole fraction	0.5854	0.5854	0.5854	0.5854	0.5854
C_6H_6 mole fraction	0.0100	0.0100	0.0100	0.0100	0.0100
C_7H_8 mole fraction	0.0012	0.0012	0.0012	0.0012	0.0012
$C_{12}H_{10}$ mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000

Name	Liq	P1out	Toquench	ToC1	V5out
Stream Number	21	22	23	24	25
Temperature [F]	113.0000	113.4569	113.4569	113.4569	114.7686
Pressure [psi]	476.8000	551.6000	551.6000	551.6000	152.0000
Molar Flow [lbmole/hr]	488.1226	488.1226	108.0254	380.0972	380.0972
H_2 mole fraction	0.0047	0.0047	0.0047	0.0047	0.0047
CH_4 mole fraction	0.0447	0.0447	0.0447	0.0447	0.0447
C_6H_6 mole fraction	0.7069	0.7069	0.7069	0.7069	0.7069
C_7H_8 mole fraction	0.2280	0.2280	0.2280	0.2280	0.2280
$C_{12}H_{10}$ mole fraction	0.0157	0.0157	0.0157	0.0157	0.0157

Name	D1	V6out	B1	V7out	D2
Stream Number	26	27	28	29	30
Temperature [F]	123.9057	122.2856	375.1323	242.4253	221.9899
Pressure [psi]	150.0000	120.0000	153.0000	32.0000	30.0000
Molar Flow [lbmole/hr]	19.6151	19.6151	360.4821	360.4821	267.9070
H_2 mole fraction	0.0911	0.0911	0.0000	0.0000	0.0000
CH_4 mole fraction	0.8666	0.8666	0.0000	0.0000	0.0000
C_6H_6 mole fraction	0.0420	0.0420	0.7431	0.7431	0.9997
C_7H_8 mole fraction	0.0003	0.0003	0.2403	0.2403	0.0003
$C_{12}H_{10}$ mole fraction	0.0000	0.0000	0.0165	0.0165	0.0000

Name	V8out	B2	P2out	V9out	D3
Stream Number	31	32	33	34	35
Temperature [F]	177.5349	292.1605	292.3059	292.1643	279.7480
Pressure [psi]	15.0000	33.0000	53.0000	33.0000	30.0000
Molar Flow [lbmole/hr]	267.9070	92.5751	92.5751	92.5751	86.6152
H_2 mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4 mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000
C_6H_6 mole fraction	0.9997	0.0006	0.0006	0.0006	0.0006
C_7H_8 mole fraction	0.0003	0.9350	0.9350	0.9350	0.9993
$C_{12}H_{10}$ mole fraction	0.0000	0.0644	0.0644	0.0644	0.0000

Name	P3out	Rtol	B3	V10out
Stream Number	36	37	38	39
Temperature [F]	284.3927	284.4901	564.9709	498.9063
Pressure [psi]	635.0000	605.0000	33.0000	16.0000
Molar Flow [lbmole/hr]	86.6152	85.9217	5.9599	5.9599
H_2 mole fraction	0.0000	0.0000	0.0000	0.0000
CH_4 mole fraction	0.0000	0.0000	0.0000	0.0000
C_6H_6 mole fraction	0.0006	0.0006	0.0000	0.0000
C_7H_8 mole fraction	0.9993	0.9993	0.0003	0.0003
$C_{12}H_{10}$ mole fraction	0.0000	0.0000	0.9997	0.9997

Table A. 3 Energy table for hydrodealkylation process

Name	Qfur	Qcooler	Wkcomp	Wkp1	Qc1
Heat flow [BTU/hr]	15578834	21769436	1315969	13623	645121

Name	Qr1	Qc2	Qr2	Wkp2	Qr3
Heat flow [BTU/hr]	4351827	13625131	11592859	924	1621607

Name	Qc3	Wkp3
Heat flow [BTU/hr]	1482986	25385

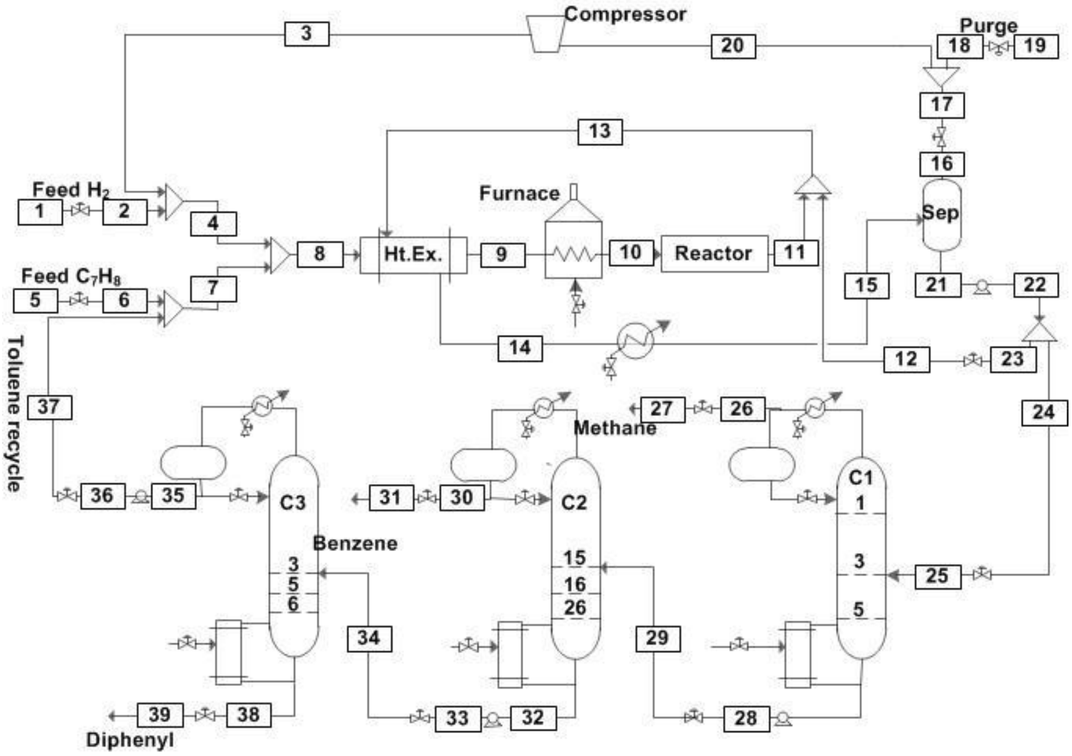

Figure A. 1 Hydrodealkyltion process flowsheet for steady-state simulation

APPENDIX B

CONTROLLER TYPE AND TUNING PARAMETERS FOR PROCESS

Modern complex control structure systems usually require more performance criteria. A performance index is a single measure of a system's performance that emphasizes those characteristics of the response that are deemed to be important. The notion of a performance index is very important in estimator performance of control structure system of hydrodealkylation process.

In this thesis, IAE method and IAE results are used to evaluate the dynamic performance of whole control structures.

$$\mathrm{IAE} = \int |e(t)|\, dt \qquad \text{(B.1)}$$

Where $e(t) = y_{sp}(t)$ is the deviation (error), $y_{sp}(t)$ is setpoint of controller and $y(t)$ is process variable of process.

The absolute error which is involved: straightforward integration of the error would produce zero result, even if the system response was a constant amplitude oscillation. IAE is often used where digital simulation of a system is being employed.

Table B. 1 Parameter tuning for hydrodealkylation process of CS0

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
Fresh feed H_2	PC	Pressure in separator	Fresh feed H_2 rate	PI	Reverse	476.80	376.8-576.8 psia	2.00	10.00
Fresh feed C_7H_8	FC	Combine feed toluene flow rate	Fresh feed C_7H_8 rete	PI	Reverse	370.20	170.2-570.2 lb-mole/hr	0.12	2.58e-3
Reactor	TC	Reactor inlet temperature	Furnace duty	PI	Reverse	1150.00	950-1350 °F	1.11	0.16
Liquid quench	TC	Reactor effluent temperature	Liquid quench flow rate	PI	Direct	1150.00	950-1350 °F	3.47	0.15
Separator	LC	Separator liquid	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
	TC	Separator liquid temperature	Cooler duty	PI	Direct	113.00	0-300 °F	1.61	0.07
Purge	CC	Recycle gas methane composition	Purge flow rate	PI	Direct	0.59	0.00-1.00	0.35	4.41
C1	LC	Reflux drum level	Condenser duty	P	Reverse	50.00	0-100%	2	-
	PC	Reflux drum pressure	Vapor flow rate	PI	Direct	150.00	50-250 psia	2.00	10.00
	TC	State 1 temperature	Reboiler duty	PI	Reverse	308.96	200-400 °F	1.00	-
	FC	Reflux flow rate	Reflux flow rate	PI	Reverse	-	-	-	-
	LC	Base Column level	Bottom flow rate	P	Direct	50.00	0-100%	2.00	-

Table B.1 Parameter tuning for hydrodealkylation process of CS0 (Continued)

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
C2	LC	Reflux drum level	Distillate flow rate	P	Direct	49.86	0-100%	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-60 psia	2.00	10.00
	TC	State 16 temperature	Reboiler duty	PI	Reverse	248.20	147.9-347.9 °F	2.76	3.01
	FC	Reflux flow rate	Reflux flow rate	PI	Reverse	-	-	-	-
	LC	Base Column level	Bottom flow rate	P	Direct	50.84	0-100%	2.00	-
C3	LC	Reflux drum level	Distillate flow rate	P	Direct	30.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-100 psia	2.00	10.00
	TC	State 5 temperature	Bottom flow rate	PI	Direct	391.70	297.2-497.2 °F	1.00	39.00
	FC	Reflux flow rate	Reflux flow rate	PI	Reverse	-	-	-	-
	LC	Base Column level	Reboiler duty	P	Direct	49.96	0-100 %	3.00	-

Table B. 2 Parameter tuning for hydrodealkylation process of CS1

equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
Fresh feed H_2	FC	Combine feed hydrogen flow rate	Fresh feed H_2 flow rate	PI	Reverse	4048.15	3711-4311 lb-mole/hr	0.26	1.76e-3
Fresh feed C_7H_8	FC	Combine feed toluene flow rate	Fresh feed C_7H_8 flow rate	PI	Reverse	370.20	170.2-570.2 lb-mole/hr	0.12	2.58e-3
Reactor	TC	Reactor inlet temperature	Furnace duty	PI	Reverse	1150.00	950-1350 °F	1.11	0.16
Liquid quench	TC	Reactor effluent temperature	Liquid quench flow rate	PI	Direct	1150.00	950-1350 °F	3.47	0.15
Separator	LC	Separator Liquid	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
	TC	Separator liquid temperature	Cooler duty	PI	Direct	113.00	0-300 °F	1.61	0.07
	PC	Separator pressure	Separator top stream	PI	Direct	476.80	376.8-576.8 psia	0.82	1.89e-3
Purge	CC	Separator top stream methane composition	Purge flow rate	PI	Direct	0.59	0.00-1.00	0.35	4.41
C1	LC	Reflux drum level	Condenser duty	P	Reverse	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Vapor flow rate	PI	Direct	150.00	50-250 psia	2.00	10.00
	TC	State 1 temperature	Reboiler duty	PI	Reverse	308.96	200-400 °F	1.00	10.00
	FC	Reflux ratio	Reflux flow rate	PI	Reverse	32.68	0-200 lb-mole/hr	0.09	0.04
	LC	Base Column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-

Table B.2 Parameter tuning for hydrodealkylation process of CS1 (Continued)

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
C2	LC	Reflux drum level	Distillate flow rate	P	Direct	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-60 psia	2.00	10.00
	TC	State 16 temperature	Reboiler duty	PI	Reverse	248.20	147.9-347.9 °F	2.76	3.01
	FC	Reflux to feed	Reflux flow rate	PI	Reverse	786.76	0-2000 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
C3	LC	Reflux drum level	Distillate flow rate	P	Direct	30.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-100 psia	2.00	10.00
	TC	State 4 temperature	Reboiler duty	PI	Direct	391.70	297.2-497.2 °F	0.50	3.01
	FC	Reflux ratio	Reflux flow rate	PI	Reverse	24.59	0-200 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	49.96	0-100%	2.00	-

Table B. 3 Parameter tuning for hydrodealkylation process of CS2

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
Fresh feed H_2	FC	Combine feed hydrogen flow rate	Fresh feed H_2 flow rate	PI	Reverse	4048.15	3711-4311 lbmole/hr	1.76e-3	1.76e-3
Fresh feed C_7H_8	FC	Combine feed toluene flow rate	Fresh feed C_7H_8 flow rate	PI	Reverse	370.20	170.2-570.2 lbmole/hr	2.58e-3	2.58e-3
Reactor	TC	Reactor inlet temperature	Furnace duty	PI	Reverse	1150.00	950-1350 °F	1.11	0.16
Liquid quench	TC	Reactor effluent temperature	Liquid quench flow rate	PI	Direct	1150.00	950-1350 °F	3.47	0.15
Separator	LC	Separator Liquid	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
	TC	Separator liquid temperature	Cooler duty	PI	Direct	113.00	0-300 °F	1.61	0.07
	PC	Separator pressure	Separator top stream	PI	Direct	476.80	376.8-576.8 psia	0.82	1.89e-3
Purge	CC	Separator top stream methane composition	Purge flow rate	PI	Direct	0.59	0.00-1.00	0.35	4.41
C1	LC	Reflux drum level	Condenser duty	P	Reverse	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Vapor flow rate	PI	Direct	150.00	50-250 psia	2.00	10.00
	TC	State 1 temperature	Reboiler duty	PI	Reverse	308.96	200-400°F	1.00	10.00
	FC	Reflux ratio	Reflux flow rate	PI	Reverse	32.68	0-200 lb-mole/hr	0.09	0.04
	LC	Base Column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-

Table B.3 Parameter tuning for hydrodealkylation process of CS2 (Continued)

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
C2	LC	Reflux drum level	Distillate flow rate	P	Direct	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-60 psia	2.00	10.00
	TC	State 16 temperature	Reboiler duty	PI	Reverse	248.20	147.9-347.9 °F	2.76	3.01
	FC	Reflux to feed	Reflux flow rate	PI	Reverse	786.76	0-2000 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
C3	LC	Reflux drum level	Distillate flow rate	P	Direct	30.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-100 psia	2.00	10.00
	TC	State 4 temperature	Reboiler duty	PI	Direct	391.70	297.2-497.2 °F	0.50	3.01
	FC	Reflux to feed	Reflux flow rate	PI	Reverse	24.59	0-200 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	49.96	0-100 %	2.00	-

Table B. 4 Parameter tuning for hydrodealkylation process of CS3

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
Fresh feed H_2	FC	Ratio H_2 to C_7H_8 flow rate	Fresh feed H_2 flow rate	PI	Reverse	1909.98	910 -2910 lb-mole/hr	0.44	0.03
Fresh feed C_7H_8	FC	Combine feed toluene flow rate	Fresh feed C_7H_8 flow rate	PI	Reverse	370.20	170.2-570.2 lb-mole/hr	0.12	0.00
Reactor	TC	Reactor inlet temperature	Furnace heat duty	PI	Reverse	1150.00	950-1350 °F	1.11	0.16
Liquid quench	TC	Reactor effluent temperature	Liquid quench flow rate	PI	Direct	1150.00	950-1350 °F	3.47	0.15
Separator	LC	Separator Liquid	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
	TC	Separator liquid temperature	Cooler heat duty	PI	Direct	113.00	0-300 °F	1.61	0.07
	PC	Separator pressure	Separator top stream	PI	Direct	476.80	376.8-576.8 psia	0.82	1.89e-3
Purge	CC	Separator top stream methane composition	Purge flow rate	PI	Direct	0.59	0.00-1.00	0.35	4.41
C1	LC	Reflux drum level	Condenser duty	P	Reverse	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Vapor flow rate	PI	Direct	150.00	50-250 psia	2.00	10.00
	TC	State 1 temperature	Reboiler duty	PI	Reverse	308.96	200-400 °F	1.00	10.00
	FC	Reflux ratio	Reflux flow rate	PI	Reverse	32.68	0-200 lb-mole/hr	0.09	0.04
	LC	Base Column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-

Table B.4 Parameter tuning for hydrodealkylation process of CS3 (Continued)

Equipment	Controllers	Controlled variables	Manipulate Variables	Controller Types	Action	Setpoint	PV Range	Tuning parameter	
								K_C	τ_I
C2	LC	Reflux drum level	Distillate flow rate	P	Direct	50.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-60 psia	2.00	10.00
	TC	State 16 temperature	Reboiler duty	PI	Reverse	248.20	147.9-347.9 °F	2.76	3.01
	FC	Reflux to feed	Reflux flow rate	PI	Reverse	786.76	0-2000 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	50.00	0-100 %	2.00	-
C3	LC	Reflux drum level	Distillate flow rate	P	Direct	30.00	0-100 %	2.00	-
	PC	Reflux drum pressure	Condenser duty	PI	Direct	30.00	0-100 psia	2.00	10.00
	TC	State 4 temperature	Reboiler duty	PI	Direct	391.70	297.2-497.2 °F	0.50	3.01
	FC	Reflux ratio	Reflux flow rate	PI	Reverse	24.59	0-200 lb-mole/hr	0.12	0.04
	LC	Base column level	Bottom flow rate	P	Direct	49.96	0-100%	2.00	-

APPENDIX C
IAE RESULTS DATA

Table C. 1 The IAE results for handle reactor inlet temperature step change disturbance

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	-	0.1251	0.0015	-	0.0670	62.7220	62.9156
CS1	-	0.1112	0.0100	-	0.0385	3.2333	3.3930
CS2	-	0.1112	0.0100	-	0.0386	3.1003	3.2600
CS3	-	0.1075	0.0105	-	0.0392	3.2883	3.4455

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	1.5932	3.7337	0.2198	0.6915	6.2382
CS1	0.0007	0.5393	0.1663	1.3763	2.0826
CS2	0.0007	0.5411	0.1663	1.3448	2.0529
CS3	0.0011	0.3532	0.1693	1.3232	1.8467

Control Structures	Composition loop controls				IEA	IEA	SUM IAE
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0200	0.4208	1.69E-04	1.24E-05	0.4407	0.0002	0.4410
CS1	0.0198	0.0343	1.09E-4	5.97E-5	0.0541	0.0001	0.0542
CS2	0.0198	0.0343	1.01E-4	5.01E-05	0.0541	0.0001	0.0542
CS3	0.0315	0.0377	9.5E-05	6.37E-05	0.0693	0.0001	0.0694

Table C. 2 The IAE results for handle combine feed toluene flow rate step change disturbance.

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	0.3166	0.3166	0.0140	-	0.2934	110.7048	111.6454
CS1	0.5239	0.2913	0.0419	-	1.0086	3.0596	4.9254
CS2	0.5242	0.2913	0.0418	-	1.0091	3.4283	5.2948
CS3	0.5227	0.2998	0.0432	-	1.0037	5.0169	6.8864

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	4.9588	9.9398	0.8213	2.5714	18.2913
CS1	0.0052	4.1255	1.2847	1.3874	6.8028
CS2	0.0052	4.1250	1.2847	1.4242	6.8390
CS3	0.0038	4.2627	1.4258	1.8874	7.5797

Control Structures	Composition loop controls				IEA	IEA	SUM IEA
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0833	1.4188	0.0009	5.14E-05	1.5022	0.0009	1.5031
CS1	0.0282	0.0892	0.0001	3.76E-05	0.1175	1.07E-04	0.1176
CS2	0.0282	0.0892	0.0001	3.45E-05	0.1175	0.0001	0.1176
CS3	0.3243	0.1500	0.0001	3.40E-05	0.4743	0.0001	0.4745

Table C. 3 The IAE results for handle fresh feed hydrogen composition step change disturbance.

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	0.0309	0.0218	0.0016	-	0.0756	6.3328	6.4628
CS1	0.0131	0.0078	0.0013	-	0.0208	0.5245	0.8113
CS2	0.0130	0.0078	0.0013	-	0.0210	0.5446	0.5878
CS3	0.0194	0.0127	0.0044	-	0.0192	0.5274	0.5831

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	3.0426	1.2446	0.1533	0.3087	4.7491
CS1	0.0001	0.1992	0.0538	0.4309	0.6840
CS2	0.0001	0.1993	0.0539	0.4316	0.6850
CS3	0.0008	0.0349	0.0437	0.5564	0.6359

Control Structures	Composition loop controls				IEA	IEA	SUM IAE
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0366	0.0313	2.69E-05	1.42E-06	0.0678	2.69E-05	0.0678
CS1	0.0147	0.0064	1.70E-05	1.34E-06	0.0211	1.70E-05	0.0211
CS2	0.0147	0.0064	1.70E-05	1.34E-06	0.0211	1.70E-05	0.0211
CS3	0.0112	0.0076	1.60E-05	3.32E-06	0.0189	1.60E-05	0.0189

Table C. 4 The IAE results for handle reactor inlet temperature pulse disturbance.

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	-	0.1289	0.0017	-	0.0602	10.8011	10.9920
CS1	-	0.1153	0.0087	-	0.0278	0.6504	0.8022
CS2	-	0.1155	0.0087	-	0.0278	0.6384	0.7904
CS3	-	0.1127	0.0105	-	0.0265	0.6442	0.7939

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	0.8205	0.7429	0.0839	0.2595	1.9067
CS1	0.0010	0.3414	0.0580	0.4140	0.8144
CS2	0.0011	0.3414	0.0579	0.4116	0.8119
CS3	0.0016	0.2605	0.0613	0.4002	0.7236

Control Structures	Composition loop controls				IEA	IEA	SUM IAE
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0133	0.0274	3.39E-05	9.63E-07	0.0406	3.39E-05	0.0407
CS1	0.0074	0.0022	1.55E-05	2.72E-06	0.0096	1.55E-05	0.0096
CS2	0.0074	0.0022	1.57E-05	2.38E-06	0.0096	1.57E-05	0.0096
CS3	0.0033	0.0027	1.56E-05	3.45E-06	0.0060	1.56E-05	0.0060

Table C. 5 The IAE results for handle combine feed toluene flow rate pulse disturbance.

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	0.6225	0.3225	0.0185	-	0.2751	32.6782	33.9169
CS1	0.5911	0.2862	0.0461	-	1.0064	2.0186	3.9485
CS2	0.5909	0.2862	0.0461	-	1.0068	2.3213	4.2514
CS3	0.6339	0.3121	0.0620	-	0.9759	2.7279	4.7118

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	4.3093	3.5122	0.4513	1.5152	9.7879
CS1	0.0044	2.5578	0.8617	0.8842	4.3082
CS2	0.0044	2.5567	0.8618	0.9140	4.3369
CS3	0.0126	2.7270	0.9217	1.2906	4.9519

Control Structures	Composition loop controls				IEA	IEA	SUM IAE
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0680	0.1165	7.59E-05	2.49E-06	0.1844	7.59E-05	0.1845
CS1	0.0119	0.0456	4.47E-05	2.84E-06	0.0574	4.47E-05	0.0575
CS2	0.0119	0.0455	4.62E-05	2.13E-06	0.0574	4.62E-05	0.0575
CS3	0.0271	0.0307	4.65E-05	7.69E-06	0.0578	4.65E-05	0.0579

Table C. 6 The IAE results for handle fresh feed hydrogen composition pulse disturbance.

Control Structures	Temperature loop controls						Sum
	TC1	TC2	TC3	TC4	TC5	TC6	IEA
CS0	0.0257	0.0184	0.0018	-	0.0621	2.4907	2.5987
CS1	0.0156	0.0078	0.0014	-	0.0171	0.1623	0.2041
CS2	0.0156	0.0077	0.0014	-	0.0168	0.1756	0.2172
CS3	0.0195	0.0139	0.0041	-	0.0184	0.1334	0.1894

Control Structures	Pressure loop controls				Sum
	PC1	PC2	PC3	PC4	IEA
CS0	1.3092	0.7072	0.0909	0.1033	2.2107
CS1	0.0001	0.0793	0.0279	0.1098	0.2172
CS2	0.0001	0.0790	0.0279	0.1116	0.2186
CS3	0.0011	0.0211	0.0192	0.0947	0.1361

Control Structures	Composition loop controls				IEA	IEA	SUM IAE
	xP	xD1	xD2	xD3	xP, xD1 ,xB3	xD2	Comp.
CS0	0.0219	0.0120	2.97E-05	3.54E-07	0.0339	2.97E-05	0.0340
CS1	0.0056	0.0033	9.38E-06	3.44E-07	0.0089	9.38E-06	0.0089
CS2	0.0056	0.0033	8.21E-06	3.56E-07	0.0089	8.21E-06	0.0089
CS3	0.0012	0.0011	1.36E-05	7.45E-07	0.0023	1.36E-05	0.0023

Temperature	
TC1	Reactor inlet temperature
TC2	Liquid quench temperature
TC3	Separator liquid temperature
TC4	Column C1 1st tray temperature
TC5	Column C2 16th tray temperature
TC6	Column C3 5th tray temperature

Pressure	
PC1	Separator pressure
PC2	Reflux drum C1 pressure
PC3	Reflux drum C2 pressure
PC4	Reflux drum C3 pressure

Composition	
xP	Recycle gas methane composition
xD1	Distillate column C1 methane composition
xD2	Distillate column C2 benzene composition
xB3	Bottom column C3 Diphenyl composition

VITA

Mr. Teerapong Juncharoenwongsa was born in Chachoengsao, Thailand on October, 1989. He graduated at Kasetsart University, Thailand and received the degree of Bachelor of Engineering in the field of Chemical Engineering in 2012. After that he entered the Graduate School of Chulalongkorn University to propose the degree of Master of Engineering in Chemical Engineering and completed in 2013.