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PLANTWIDE CONTROL STRUCTURE DESIGN OF ETHYL BENZENE USING FIXTURE POINT THEOREM

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A Thesis Submitted in Partial Fulfillment of the Requirements
for the Degree of Master of Engineering Program in Chemical Engineering
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นภาภรณ์ ผลประเสริฐ : วิธีการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ของกระบวนการเอทิลเบนซีนโดยใช้ทฤษฎีจุดตรึง. (PLANTWIDE CONTROL STRUCTURE DESIGN OF ETHYL BENZENE USING FIXTURE POINT THEOREM) อ. ที่ปรึกษาวิทยานิพนธ์หลัก: ผู้ช่วยศาสตราจารย์ ดร.มนตรี วงศ์ศรี, 117 หน้า.

โครงสร้างการควบคุมกระบวนการทางเคมีนั้นมีการศึกษากันอย่างแผ่หลายเพื่อหาจุดที่เหมาะสมทางด้านเศรษฐศาสตร์ภายใต้ข้อจำกัดของกระบวนการและความปลอดภัยด้านสิ่งแวดล้อม

งานวิจัยนี้ได้นำทฤษฎีจุดตรึง" Fixture Point Theorem " ประยุกต์ใช้กับกระบวนการเอทิลเบนซีนเพื่อออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์แบบใหม่และนำเปรียบเทียบกับของ Luyben (2002) หลักการของทฤษฎีจุดตรึงคือตัวแปรที่ถูกรบกวนมากในกระบวนการเมื่อทำการเปลี่ยนแปลงตัวแปรปรับหรือตัวกวนอื่น ตัวแปรนั้นจะถูกนำมาพิจารณาเป็นอันดับแรกในการออกแบบ เราทำการเลือกชุดการควบคุมอีกสองชุดและออกแบบโครงสร้างการควบคุมเป็นสามแบบแล้วทำการเปรียบเทียบ การแสดงพฤติกรรมทางพลวัตรของโครงสร้างการควบคุมที่ออกแบบเมื่อทำการเปลี่ยนแปลงตัวแปรปรับและประสิทธิภาพแสดงโดยค่า IAE พบว่าโครงสร้างที่ออกแบบสามารถรักษาคุณภาพของผลิตภัณฑ์ และลดพลังงานที่ใช้ในกระบวนการ สรุปได้ว่าการออกแบบการควบคุมแบบแพลนท์ไวด์โดยใช้ขั้นตอนของวงศ์ศรี (2009) เป็นวิธีนำไปสู่โครงสร้างการควบคุมที่มีสมรรถนะที่ดี

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Chemical process control structure was studied widely to optimized economic under various process object constrains and environment safety.

This research is applied "Fixture Point Theorem" to ethyl benzene process to design a new plantwide control structures that be compared with Luyben (2002). The principle of the fixture point theorem is the most disturbed variables were affected by manipulated variables or disturbances change should be precedence consideration. We selected two set of controlled variables and three control structures were designed and compared. The dynamic behaviors of designed control structures were illustrated whether manipulated variables change and the performance were presented in IAE value. The design control structure has a good performance and energy use reduction in the process. This research establishes that the Wongsri's procedure lead to a good performance plantwide control system.

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CHAPTER I

INTRODUCTION

This Chapter is an introduction of this research. It consists of importance and reasons for research, research objective, scope of research, contribution of research and research contents.

1.1 Importance and Reasons for research

Control structure design deals with the structural decisions of control system, including what to control and how to pair the variables to the best set of form control loop. Although these are very important issues, these decisions are in most case made in a fashion, base on experience and engineering insight, without considering the details of process variable link with problem.

The general problems of plantwide control design procedure are selection the controlled variables, selection the manipulated variables, selection the measurement, pairing the controlled and manipulated variables for the best control configurations and selection of controller type.

The process had good controllability one reason came from selection appropriate the set of controlled and manipulated variables for the control configuration of chemical plant.

In this paper propose a basic idea of fixture point control is used for screening controlled variable and paring with manipulated variable.

Ethyl Benzene process is widely used because it is a realistically complex chemical process that creates disturbance propagation and the complicated dynamic system behavior. Therefore, this research will design plantwide control structures of Ethyl Benzene process using new plantwide control structure design procedure to select set of controlled variables and simulate them by using HYSYS simulator in order to study about dynamic behavior and evaluate the performance of the designed structures.

1.2 Objective of present research

The objectives of this research are:

1. To simulate Ethyl benzene process both steady state and dynamics condition by using HYSYS simulator.
2. To design plantwide control structures of Ethyl benzene process using Fixture point theorem (Wongsri, 2008).

1.3 Scopes of research

The scope of this research can be listed as follows:

1. The simulator in this research is HYSYS simulator.

2. Description and data of Ethyl benzene process is obtained from William L. Luyben (2002).

3. To provide the vital variables and also to design control structures of Ethyl benzene process.

1.4 Expected Contributions

The expected contribution of this research is:

This research provides the appropriate controlled variables and good efficacious design control structure by using plantwide control structure design procedure, Fixture Point theorem.

1.5 Research procedures

Procedure plans of this research are:

1. Study of general plantwide control structure design procedure.

2. Study of ethyl benzene process and concerned information.

3. Simulations of ethyl benzene process at steady state and dynamic.

4. Screening key process variables for fixture point analysis

5. Pairing manipulated variable with control variables of ethyl benzene process

6. Simulation of ethyl benzene process at dynamic

7. Collection and summarization of simulation results

8. Conclude the thesis

1.6 Research Contents

This thesis matter is classified into six chapters as follows:

Chapter I	provides an introduction, motivation, objectives, scopes, contributions and thesis outline.
Chapter II	presents literature review related to plantwide control structure design procedure and the method of selection set of controlled variables.
Chapter III	purposes new plantwide control structure design procedure and present the fixture point theorem for selection appropriate set of controlled variables.
Chapter IV	shows case study of the ethyl benzene process (EB) by using theory in chapter 3.
Chapter V	presents control structure alternatives.
Chapter VI	the last chapter shows overall conclusions of this research and recommendations for future research.

This is followed by:
Appendix A: Tuning of Control structures
Appendix B: Parameter Tuning
Appendix C: Data of fixture point analysis

CHAPTER II

LITERATURE REVIEW

A. Plantwide control design procedure

2.1 Heuristics- Approach Base

2.1.1 Hierarchical decomposition based on process structure

The hierarchy given in Douglas (1988) for process design starts at a crude representation and gets more detailed:

Level 1: Bath vs continuous
Level 2: Input-output structure
Level 3: Recycle structure
Level 4: General structure of separation system
Level 5: Energy interaction

Fisher et al. (1988) propose to use this hierarchy when performing controllability analysis, and Ponton and Laing (1993) point out that this hierarchy, (e.g. level 2 to level 5) could also be used for control system design.

Ng and Stephanopoulos (1998b) propose to use a similar hierarchy for control structure design. The difference between Douglas (1988) and Ng and Stephanopoulos (1998b)'s hierarchy is that level 1 is replaced by a preliminary analysis and that levels 4 and 5 are replaced by more detailed structures. At each step the objectives identified at an earlier step is translated to this level and new objectives are identified. The focus is on construction of mass and energy balance control. The method is applied to the Tennessee Eastman case.

All these methods have in common that at each step (level), a key point is to check if there remain enough manipulated variables to meet the constrains and to optimize operation.

2.1.2 Hierarchical decomposition based on control objectives

The hierarchy based on control objectives is sometime called the tiered procedure. This bottom-up procedure focuses on the task that the controller has to perform. Normally one starts by stabilizing the plant, which mainly involves placing inventory (mass and energy) controllers.

Price et al. (1993) build on the idea that was introduced by Buckley (1964) and introduce a tiered framework. The framework is divided into four different tasks:

I. Inventory and production rate control
II. Product specification control
III. Equipment & operating constraints
IV. Economic performance enhancement

Their paper does not discuss points III or IV. They perform a large number of simulations with different control structures, controllers (P or PI), and tunings on a simple process consisting of a reactor, separator and recycle of unreacted reactant. The configurations are ranked based on integrated absolute error of the product composition for steps in the disturbance. From this simulation they propose some guidelines for selecting the though-put manipulator and inventory controls (1) Prefer internal flows as through-put manipulator. (2) The though-put manipulator and inventory controls should be self-consistent (self-consistent id fulfilled when a change in the though-put propagation through the process by itself and does not depend on composition controllers). They apply their ideas on the Tennessee Eastman problem (Price et al.1994).

Ricker (1996) comments on the work of Price et al. (1994) and points out that plant is often run at full capacity, corresponding to constraints in one or several variables. If a manipulated variable used for level control structures, one looses a degree of freedom for maximum production. This should be considered when choosing a though-put manipulator.

Luyben et al. (1997) point out three limitations of the approach of Buckley. First, he did not explicitly discuss energy management. Second, he did not look at recycles. Third, he placed emphasis on inventory control before quality control. Their plantwide control design procedure is listed below:

1. Establish control objectives.
2. Determine the control degrees of freedom by counting the number of independent valves.
3. Establish energy inventory control, for removing the heats of reactions and to prevent propagation of thermal disturbances.
4. Set production rate. The production rate can only be increased by increasing the reaction rate in the reactor. One recommendation is to use the input to separation section.
5. Product quality and safety control. Here they recommend the usual 'pair close' rule.
6. Inventory control. Fix a flow in all liquid recycle loops. They state that all liquid levels and gas pressure should be controlled.
7. Check component balances. (After this point it might be necessary to go back to item 4).
8. Unit operations control.
9. Use remaining control degree of freedom to optimize economics or improve dynamic controllability.

They apply their procedure on several test problems; the vinyl acetate monomer process, the Tennessee Eastman process, and the HDA process.

Step 3 comes before determining the throughput manipulator, since the reactor is typically the heart of the process and the method for heat removal are intrinsically part of the reactor design. In order to avoid recycling of disturbances they suggest to set a flowrate in all recycle loops; they suggest in step 6 to control all inventories, but this may not be necessary in all case; e.g. it may be optimal to let the pressure float (Shinskey 1988). Skogestad et al. (2000) recommend combining step 1 and 9, that is, the selection of controlled variables (control objectives) in step 1 should be based on overall plant economics.

McAvoy (1999) presents a method where the control objectives are divided into two categories: variables that 'must' be controlled, and product flow and quality. His approach is to identify the set of inputs that minimizes valve movements. This is first solved for the 'must' variables, then for product rate and quality. The optimization problem is simplified by using a linear stable state model. He gives no guidance into how to identify the controlled variables.

2.1.3 Hierarchical decomposition based on time scales

Buckley (1964) proposed to design the quality control system as high-pass filters for disturbances and to design the mass balance control system as low pass filters. If the resonance frequency of the quality control system is designed to be order of magnitude higher than the break frequency of mass balance system then the two loops will be non-interacting.

McAvoy and Yc (1994) divide their method into four stages:

1. Design of inner cascade loops.
2. Design of basic decentralized loops; expect those associated with quality and production rate
3. Production rate and quality controls.
4. Higher layer controls

The decomposition in stages 1-3 is based on speed of the loops. In stage 1 the idea is to locally reduce the effect of disturbances. In stage 2 there generally are a large number of alternative configurations. These may be screened using simple controllability tools, such as the RGA. One problem of selecting outputs based on a controllability analysis is that one may end up with the outputs that are easy to control, rather than the ones that are important to control. The method applied to the Tennessee Eastman test problem.

Douglas (1988) presents a hierarchy for control system design, based on steady-state, normal dynamic response and abnormal dynamic operation. Zheng et al. (1999) continue this work and place a greater attention on feasibility in face of constraints and on robust optimality (self-optimizing control). Zheng and Mahajannam (1999) propose to use minimum surge capacity as a dynamic cost.

2.2 Mathematical-Approach Base

A chemical plant may have thousands of measurement and control loops. There are some methods that use structural information about the plant as a basis for control structure design. Based on sets of inputs and measurements are classified as viable or non-viable. Although the structural methods are interesting, they are not quantitative and usually provide little information other than confirming insights about the structure of the process that most engineers already have.

The tasks of control structure consist of (1) selection of controlled variables, (2) selection of manipulated variables, (3) selection of measurements and (4) selection of control configuration.

Morari et al (1980), Skogestad and Postlethwaite (1996) and Skogestad (2000) propose to base the selection of controlled variables based on considering the overall operational objective. The overall objective may be formulated as a scalar cost function which should be minimized subject to set of operational constraints.

Morari et al. (1980) propose the basic idea of self-optimizing control propose to select the best set of controlled variables based on minimizing the loss. Skogestad et al (2000) attempt to synthesis a feedback optimizing control structure (self-optimizing control), their main objective is to translate the economic objectives into process control objectives. In other words, they want to find a function c (controlled variables) of the process variables which when held constant, leads automatically to the optimal adjustments of the manipulated variables and optimal operating conditions. This means that by keeping the function $c(u,d)$ at the setpoint c_s, through the use of the manipulated variables u, for various disturbances d, it follows uniquely that the process is operating at the optimal steady-state.

Skogestad (2000) gives four requirements that a controlled variable should meet:

1. Its optimal value should be insensitive disturbance.
2. It should be easy to measure and control accurately.
3. Its value should be sensitive to changes in the manipulated variables.

4. For cases with two or more controlled variables, the selected variables should not be closely correlated.

Yi and Luyben (1995) have studied unconstrained problems, and some of their ideas are related to self-optimizing control. However, Luyben propose to select controlled variables which minimizes the steady-state sensitivity of the manipulated variables (u) to disturbances, i.e. to select controlled variables (c) such that $(\partial u/\partial d)_c$ is small, whereas one should really minimize the steady-state sensitivity of the economic loss (L) to disturbances, i.e. to select controlled variables (c) such that $(\partial L/\partial d)_c$ is small.

B. Description and data of Ethyl Benzene process

The process considered in this thesis proposal was ethyl benzene from ethylene and benzene. Ethyl benzene is an aromatic hydrocarbon which important in petrochemical industry as an intermediate raw material in the production of styrene, a common used plastics material.

The process features large high-capacity equipment. The ethylene and benzene feed stocks are produced in petroleum refineries by the use of catalytic and thermal cracking and reforming. The process description and data of Ethyl Benzene process is obtained from chapter 17 William L. Luyben (2002)which a simply version of the real process of the real process with the process equipment and chemistry trimmed down to make the simulation easier. There are two reactors, two distillation columns and two recycle streams.

The basic chemical reaction is:

Reaction 1: Benzene (B) + Ethylene(E) → Ethyl Benzene(EB)

Further reaction of ethylene with ethyl benzene to produced higher-order components. The first reaction is;

Reaction 2: Ethyl Benzene(EB) + Ethylene(E) → Diethyl Benzene(DEB)

To suppress this reaction, the large excess benzene is used so that the ethylene concentration is small. The excess benzene results in the recycle stream of process. The ratio fed of benzene to ethylene to the reaction section is about 2.5. Despite the large benzene excess, there is a significant amount of diethyl benzene produced in the reactor. It must be separated from the ethyl benzene and recycled back to the reaction section where the following reaction occurs;

Reaction 3: Diethyl Benzene (DEB) + Benzene(B) → 2 Ethyl Benzene(EB)

The kinetic equations used for these three reactions are given below. These parameters are selected to give typical reaction conversations and yields.

$$R_1 = C_E C_B (1.528\times10^6) e^{-17,000/RT}$$

$$R_2 = C_E C_{EB} (2.778\times10^4) e^{-20,000/RT}$$

$$R_3 = C_{DEB} C_B (0.416) e^{-15,000/RT}$$

Where reaction rate are in kg-mol/sec/m3, concentrations are in kg-mol/m3, activation energies are in cal/g-mol and temperature are in Kelvin

The reaction section consists of two CSTR reactors (200 m^3) in series. Fresh benzene feed and a recycle benzene stream are fed to the first reactor (R1), which operates at 160 ^{o}C and 9 bar. The exothermic heat of reaction is removed by generating low- pressure steam. The ethylene undergoes almost complete conversion in the first reactor. The effluent from the first reactor and a recycle DEB stream are fed to a second reactor (R2), which operates adiabatically. The composition of the effluent from the second reactor is about 55 mol% benzene, 35mol% ethyl benzene and 10 mol% diethyl benzene.

The reactor effluent is fed to a 24 tray distillation column (C1), operating at 4.5 bar. Recycle benzene goes overhead and the EB/DEB mixture goes out the bottom. The column has a reflux ratio of 0.9, consumes 15.8 MW in the reboiler and is 3.7m in diameter. The specifications on the stream leaving C1 are 0.1 mol% EB in the distillate and 0.1mol% benzene in the bottom.

The bottom is fed to a second distillation column (C2) that produces distillate product of high purity EB and bottoms with mostly DEB, which is recycled back to the second reactor. This 30 tray column operates at 2 bar, has a reflux ratio of 1, a diameter of 3.7 m and consumes 9.8 MW of energy in the reboiler. The reflux drum temperatur 165°C, so low pressure steam is generated in the condenser. The specifications for column are 0.1mol%EB in the bottom and 0.1 mol% DEB in distillate.

CHAPTER III

THEORY

3.1 Plantwide Control Structure Design Procedure

Plantwide process control involves the systems and strategies required to control an entire chemical plant consisting of many interconnected unit operations. A control engineers is typically presented a process flowsheet containing several recycles streams, energy integration and many different unit operation (distillation columns, heat-exchanger, reactor, etc.). Given such a complex, integrated process, one must devise the necessary logic, instrumentation and strategies to operate the plant safety and achieve its design objectives.

3.1.1 Basic concept of plantwide control

Buckley (1964) was the first to suggest the idea of separating plantwide problem into two parts: material balance control and product quality control. He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loops is established, using flowrate of liquid and gas process streams. The idea is to establish the inventory control system by setting up this "hydraulic" control structure as the first step. Then he proposed establishing the product quality control loop by choosing appropriate manipulated variables.

Jim Douglas (1988) has devised a hierarchical approach to the conceptual design of process flowsheet. He points out that in the typical chemical plant the costs of raw materials and the value of products are usually much greater than the costs of capital and energy. This leads to the two Douglas doctrines:

1. Minimize losses of reactants and products.
2. Maximize flowrate though gas recycles systems.

The first idea implies that need tight control of stream composition exiting the process to avoid losses of reactants and products. The second rests on the principle that yield is worth more than energy.

Jim Downs (1992) pointed out the importance of looking at the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively. He must ensure that all components (reactants, products and inert) have a way to leave or be consumed within the process.

Stephanopoulos et al. (1996) stated that the synthesis of a control system for a chemical plant. He asked: "Which variables should be measured in order to monitor completely the operation of a plant?" Which input should be manipulated for effective control? How measurements should be paired with the manipulations to form the control structure. He noted that the problem of plantwide control is "multi-objective" and there is a need for a systematic and organized approach which will identify all necessary control objectives. Their approaches respect the multi-objective nature of the design problem and take into account the propagation of disturbances in the process.

Luyben (1998) presented three laws have been developed as a result of number of case studies of many types of systems:

1. A stream somewhere in all recycle loops should be flow controlled. This is to prevent the snowball effect.
2. A fresh reactant feed stream cannot be flow controlled unless there is essentially complete one-pass conversion of one of the reactants. This law applies to systems with reaction types such as $A + B \rightarrow products$. In systems with consecutive reactions such as $A + B \rightarrow M + C$ and $M + B \rightarrow D + C$, the fresh fees can be flow controlled into the system because any imbalance in the ratios of reactants is accommodated by a shift in the amounts of the two products (M and D) that are generated. An excess of A will result in the production of more M and less D. An excess of B results in the production of more D and less M.
3. If the final product from a process comes out the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor (Cantrell et al., 1995). Changes in feed flowrate or feed composition have less of dynamic effect on distillate composition than they do on bottoms composition if the feed is saturated vapor: bottom is less affected than distillate. If our primary goal is

to achieve tight product quality control, the basic column design should consider the dynamic implications of feed thermal conditions.

Bob Richardson suggested the heuristic that the largest stream should be selected to control the liquid level in a vessel it is called Richardson rule.

3.1.2 Step of plantwide control structure design procedure

Luyben et al. (1998) presented **nine step** of plantwide control design procedure satisfies the two fundamental chemical engineering principles, namely the overall conservation of energy and mass.

Step 1: Establish control objectives

Assess the steady-state design and dynamic control objectives for the process. 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, i.e., the number of variables that can be controlled to setpoint. The valves must be legitimate (flow though a liquid-filled line can be regulated by only one control valve). The placement of these control valves can sometimes be made to improve dynamic performance, but often there is no choice in their location.

Step 3: Establish energy management system

Mask sure that energy disturbances do not propagate throughout the process by transferring the variability to the plant utility system.

They use the term energy management to describe two functions: (1) they must 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. (2) If heat integration does occur between process streams, then the second function of energy management is provide a control system that prevents the propagation of thermal disturbances and ensures the exothermic reactor heat is dissipated and not recycled.

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.

Throughput changes can be achieved only by altering, either directly or indirectly, conditions in the reactor. To obtain higher production rates must increase overall reaction rates. For example, temperature is often a dominant reactor variable. If reactor temperature is not a dominant variable or cannot be changed for safety or yield reasons, in these cases you must find another dominant variable, such as the concentration of the limiting reactant, flowrate of initiator or catalyst to the reactor, reactor residence time, reactor pressure, or agitation rate.

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.

They want tight control of these important qualities for economic and operational reasons, Hence they should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and deadtimes and large steady-state gains. It should be noted that establishing the product quality loops first, before the material balance control structure, is a fundamental difference between their plantwide control design procedure and Buckley’s procedure.

Step 6: Fix a flow in every recycle loop and control inventory (pressure and levels)

Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.

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. Gas recycle loops are normally set circulation rate, as limited by compressor capacity, to achieve maximum yields (Douglas doctrine). An inventory variable should typically be controlled with the manipulated variables that have the largest effect on it within that unit (Richardson rule).

<u>Step 7</u>: Check component balances

Identify how chemical components enter, leave, and generated or consumed in the process.

What are the methods or loops to ensure that the overall component balances for all chemical species are satisfied at steady-state? They don't want reactant components to leave in the product streams because of the yield loss and the desired product purity specifications. Hence they are limited to the use of two methods: consuming the reactants by reaction or adjusting their fresh feed flow. Product and inert component all must have an exit path from the system. In many systems inert are moved by purging off a small fraction of the recycle stream. The purge rate is adjusted to control the inert composition in the recycle stream so that an economic balance is maintained between capital and operating costs.

<u>Step 8</u>: 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 individual chemical units (Shinskey, 1988). For example, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust the fuel flowrate to a furnace supplying energy to the reactor. Crystallizers require manipulation of refrigeration load to control temperature. Oxygen concentration in the stack gas from a furnace is controlled to prevent excess fuel usage. Liquid solvent feed flow to an absorber is controlled as some ratio to the gas feed.

<u>Step 9</u>: Optimize economics or improve dynamic controllability

Establish the best way to use the remaining control degree 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 setpoints in some controllers that can be adjusted. These can be utilized either to optimize steady-state economic process performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

Truls Larsson and Skogestad (2000) presented 8 steps of plantwide control design procedure. They combine step 1 and 9 of nine steps of Luyben (1998) for selection of controlled variables (control objectives) in step1. The procedure is divided in two main parts:

I: Top-down analysis (step 1-4), including definition of operational objectives and consideration of degree of freedom available to meet these.

II: Bottom-up design of the control system (step 5-8), starting with the stabilizing control layer.

Step 1: Definition of operational objectives.
Step 2: Manipulated variables and degrees of freedom.
Step 3: Primary controlled variables.
Step 4: Production rate.
Step 5: Regulatory control layer.
  5.1 Stabilization
  5.2 Local disturbance rejection
Step 6: Supervisory control layer.
  6a. Decentralized (single-loop) control
  6b. Multivariable control
Step 7: Optimization layer
Step 8: Validation

Definition of operational objectives and constraints (step1)

The operational objectives must be clearly defined before attempting to design a control system. Although this seems obvious, this step is frequently overlooked. Preferably, the operational objectives should be combined into a scalar cost function J to be minimized. In many cases, J may be simply selected as the operational cost, but there are many other possibilities. Other objectives, including safety constraints, should normally be formulated as constraints.

Selection of manipulated variables and degree of freedom analysis (Step2)

They start with the number of dynamic or control degrees of freedom, N_m (m here denotes manipulated), which is equal to the number of manipulated variables. The number of manipulated variables *Nm* is usually easily obtained by process insight as the number of independent variables that can be manipulated by external means (typically, the number of adjustable valves plus other adjustable electrical and mechanical variables). Note that the original manipulated variables are always extensive variables.

Next, they identify the N_{opt} optimization degrees of freedom, that is, the degrees of freedom that affect the operational cost J. In most cases, the cost depends on the steady-state only, and N_{opt} equals the number of steady-state degrees of freedom N_{ss}. To obtain the number of steady-state degree of freedom they need to subtract from N_m:

- N_{om} = The number of manipulated (input) variables with no steady-state effect (or more generally, with no effect on the cost). Typically, these are "extra" manipulated variables used to improve the dynamic response, e.g. an extra bypass on a heat exchanger.

- N_{oy} = The number of (output) variables that need to be controlled, but which have no steady-state effect (or more generally, no effect on the cost). Typically, these are liquid levels in holdup tanks, and they have

$$N_{ss} = N_m - (N_{0m} + N_{0y})$$

The optimization is generally subject to constraints, and at the optimum many of these are usually "active". The number of "free" (unconstrained) degrees of freedom that are left to optimize the operation is then

$$N_{free} = N_{opt} - N_{active}$$

This is an important number, since it is generally for the unconstrained degrees of freedom that the selection of controlled variables.

Self-optimizing control (steps 3)

Self-optimizing control (Skogestad, 2000) is achieved if a constant setpoint policy results in an acceptable loss L (without the need to reoptimize when disturbance occur). Their first control the variables directly related to ensuring optimal economic operation (these are the primary controlled variables $y_1 = c$)

To select controlled variables for self-optimizing control, one may use the stepwise procedure of Skogestad (2000):

Step 3.1 Definition of optimal operation (cost and constraints)

Step 3.2 Determine degrees of freedom for optimization

Step 3.3 Identification of important disturbance

Step 3.4 Optimization (nominally and with disturbances)

Step 3.5 Identification of candidate controlled variables

Step 3.6 Evaluation of loss for alternative combinations of controlled variables (loss imposed by keeping constant setpoints when there are disturbances or implementation errors)

Step 3.7 Evaluation and selection (including controllability analysis)

To identify good candidate controlled variables, c one should look for variables that satisfy all of the following requirements:

- The optimal value of c should be insensitive to disturbance.
- c should be easy to measure and control (so that the implementation error is acceptable).
- The value of c should be sensitive to changes in the manipulated variables (the steady-state degree of freedom). Equivalently, the optimum (J as a function of c) should be flat.
- For case with more than one unconstrained degree of freedom, the selected controlled variables should be independent.

Production rate and inventory control (step 4)

The production rate is commonly assumed to be set at the inlet to plant, with outflows used for level control. They have the following rule: Identify the main bottleneck in the plant by optimizing the operation with the feedrate as a degree of freedom. Set the production rate at this location.

Regulatory layer (step5)

The main objective of regulatory control layer is to stabilize the plant. Usually it consists of single input-single output (SISO) PI control loops. The controlled variables for stabilization are measured output variables y_2, and their setpoint y_2 may be used as degree of freedom by the layers above.

A good secondary controlled variable (measurement) usually has the following properties:

- The variable is easy to measure

- The variable is easy to control using one of the available manipulated variables (the manipulated variable should have a “direct” fast and strong effect on it)
- For stabilization: The unstable mode should be detected “quickly” by the measurement (compute, for example, the pole vectors for a more detailed analysis)
- For local disturbance rejection: The variables is located “close” downstream of an important disturbance (use, for example, a partial control analysis for a more detailed analysis)

The “unstable” modes are very often related to inventory in each unit. This includes both the overall inventory (total mass) as well as the inventory of individual components.

- For liquid phase systems, overall inventory in each unit is stabilized by controlling liquid level.
- For gas phase systems, overall inventory (pressure) is controlled in selected units, but in many units it is left uncontrolled (floating), for example, to minimize pressure drop.
- For both gas and liquid phase systems, the inventory of individual components may need to be stabilized. Usually, this involves controlling a composition, or a derived property such as temperature. For example, in a distillation column, a temperature controller is often used to stabilize its otherwise drifting composition profile.

Supervisory control (step6)

They purpose the supervisor control layer is to keep the (primary) controlled outputs c at their optimal setpoints c_s , using degree of freedom the setpoints y_{2s} in the regulatory layer and any unused manipulated inputs. For the supervisory control layer, the first structural issue is deciding on whether to used decentralized or multivariable control. Decentralized single-loop control is the simplest. It is preferred for non-interacting process and cases where active constraints remain constant. Multivariable control is preferred for interacting processes and for process with changes in active constraint.

Optimization (step7)

The purpose of the optimization is to identify the active constraints and recomputed optimal setpoints c_s for controlled variables.

Validation (step8)

After having determined a plantwide control structure, it may be necessary to validate the structure, for example, using nonlinear dynamic simulation of critical parts

3.2 New Plantwide Control Structure Design Procedure

The book by Luyben et al. (1998) outlines the control structure design procedure based on heuristic and their process engineering insight. Several case studies are given in the book. Luyben's procedure is widely studied and used the plantwide process control community. However, the structural design procedure is not given explicitly. Their case study designs followed the outline and collected heuristic law but need the designer's process engineering insight to pair CVs and MVs. Skogestad (2004) presented the new design procedure mainly based on the mathematical analysis. First, the dynamic and steady state degree of freedom are identified. Then the set of primary controlled variables are determined. They basically are the active constraints and the variables that must be maintained to achieve minimal economic loss when disturbances occur. Then the control variable determining the production rate is selected based on the optimization resulted in the previous step. The pairings of the selected sets of MVs and CVs are done hierarchically: regulatory control, supervisory control (loop enhancement for SISO or constraint handling for MPC), and optimization layers. Several analysis tools are utilized in these steps, e.g. pole vector analysis, RGA, CLDG, linear and nonlinear optimization. However, he did not discuss which controlled variables should take precedence สำคัญกว่า? Normally the plant would have a large number of variables; the precedence of the control variables must be established to assure the optimality of the designs and to avoid ambiguity in the design decision. Second, how to handle the disturbances is not discussed. Third, he placed

This paper, we presented our plantwide control structure design procedure based on heuristics and mathematical analysis. In our procedure, the precedence of control variables is established. The major disturbances are directed or managed explicitly to achieve the minimal interaction between loops by using the extended (thermal) disturbance propagation method (Wongsri, 1990) to cover the material disturbances. The purposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward.

Normally, plantwide control design procedures consider decision about plant control structures in perspective. The plantwide control structure design is complex: hierarchical, structural, having mixed objectives, containing many units and layers, and therefore confusing. One easy way to deal with this complexity is compartmentalizing จัดแบ่ง it. However, the plant is not merely the units combined, it has it own properties. The whole is greater than the sum of its parts. There are properties (or behavior) of a system as a whole emerge out of the interaction and the relationship of the components comprising the system. Therefore, a designer must deal with both parts and system.

Our steps of control structure design are:

1. Establishment of control objectives.
2. Selection of controlled variables to maintain product quality and to satisfy safety operational and environmental constrains. The selected CVs are ranked using the Fixture Point theorem.
3. Selection of manipulated variables and measurements via DOF analysis.
4. Energy management via heat exchanger networks.
5. Selection of control configuration using various tools available.
6. Completing control structure design by checking the component balance.
7. Selection of controller type: single loops or MPC
8. Validation via rigorous dynamic simulation

Step 1: Establishment of control objectives

The control objectives are production goals and process constrains. The production goal is both qualitative and quantitative: product purity and production rate. The process constrains can be resulting from process equipment sizes, safe operating limits, and environmental issues. This step should readily gives the list of some important controlled variables.

Step 2: Selection of controlled variables to maintain product quality and to satisfy safety operational and environmental constrains. The selected CVs are ranked using the Fixture Point theorem.

The variables that set the inventory levels are added to the list of the controlled variables listed in Step 1. They are all liquid levels (except in recycle systems), and all gas pressures.

The order of the controlled variables listed in Step 1 is established according to their sensitivity to the important input loads. The most disturbed variable is given highest priority in assigning best manipulated variables before allocating manipulation to less important controlled variables. That is the variable that magnify the input loads most (having large gains) will be controlled at best effort available so that the response or the disturbance propagated will be minimal. We analyzed how changes in each of the process variables affect other variables by perturbing the manipulated and the disturbance variables.

The criteria used in this selection are: (1) the sum of IAE of the variables caused by changes in selected input variables, (2) the economic of the variable and its down stream variables, (3) the nature of the variable whether it is associated with the 'disturbance absorber' or the 'disturbance repeller'.

The control structures that we developed to satisfy the control objectives are established in order of decreasing important.

Step 3: Selection of manipulated variables and measurements via DOF analysis.

Step 4: Energy management via heat exchanger networks.

Step 5: Selection of control configuration using various tools available.

Step 6: Completing control structure design by checking the component balance.

Step 7: Selection of controller type: single loops or MPC.

Step 8: Validation via rigorous dynamic simulation.

3.3 Steady-State Gain

The availability of accurate steady-state gains for a multivariable process facilitates significantly the control system design procedure. The steady-state gains provide the zero frequency characteristics of the system. This piece of information enables the initial screening and selection of proper manipulated and/ or controlled variables, variable pairing, and initial evaluation of candidate control structures (Grosdidier et al., 1985: Yu and Luyben, 1986; Shinskey, 1988).

The steady-state gains can be determined by using either plant tests (although it has been shown (Luyben, 1987a) that the results might be seriously different from those of a linearized model of the process) or some kind of a rating program (Buckley et al., 1985). A third and more complex alternative is to get the steady-state gains through a transfer function identification procedure, if dynamic plant data or data from a dynamic model of the process are available.

The usual method to determine the gains is an open loop type of test. A specific control structure is assumed. A small perturbation is introduced in one of the manipulated variables. All the remaining manipulated variables are held constant. The rating program is converged. All measurement variables changes are recorded. The steady-state gain between the *i* controlled variable and the *j* manipulated variable is calculated as

$$g_{ij} = \Delta x_{ij} / \Delta m_j \quad i = 1, 2, ..., n; \; j = 1, 2, ..., m \qquad (3.2)$$

where g , is the *ij* element of the gain matrix and Δx_{ij} is the change in the *i* controlled variable because of the Δm_j change in the *j* manipulated variable.

According to this procedure, *m* tests need to be performed for every candidate control structure where m is the number of manipulated variables. For controlled variable selection, this procedure is ideal because it provides gains for all controlled variables.

The problem of the effect of scaling on the steady-state gains process is handled by expressing the gains of all the plant transfer functions in dimensionless form. The gains with engineering units are divided by transmitter spans and multiplied by valve gains. The method for scaling on the steady-state gains are elucidated below.

3.4 Scaling

Scaling is very important in practical applications as it makes model analysis and controller design (weight selection) much simpler. It requires the engineer to make a judgment at the start of the design process about the required performance of the system. To do this, decisions are made on the expected magnitudes of disturbance and reference changes, on the allowed magnitude of each input signal, and on the allowed deviation of each input. Let the unscaled (or originally scaled) linear model of the process in deviation variable be

$$\hat{y} = \hat{G}\hat{u} + \hat{G}_d\hat{d}; \quad \hat{e} = \hat{y} - \hat{r} \qquad (3.3)$$

Where a hat (^) is used to show that the variables are in their unscaled units. A useful approach for scaling is to make the variables less than 1 in magnitude. This is done by dividing each variable by its maximum expected or allowed change. For disturbances and manipulated inputs, we used the scaled variables

$$d = \hat{d} / \hat{d}_{\max}, \quad u = \hat{u} / \hat{u}_{\max} \qquad (3.4)$$

where:

- $\hat{d}_{\max}$ - largest expected change in disturbance
- $\hat{u}_{\max}$ - largest allowed input change

The maximum deviation from a nominal value should be chosen by thinking of the maximum value one can expect, or allow, as function of time. The variables $\hat{y}, \hat{e}$ and $\hat{r}$ are in the same units, so the same scaling factor should be applied to each. Two alternatives are possible:

- $\hat{e}_{\max}$ - largest allowed control error
- $\hat{r}_{\max}$ - largest expected change in reference value

Since a major objective of control is to minimize the control error, we here usually choose to scaled with respect to the minimum control error:

$$y = \hat{y} / \hat{e}_{\max}, \quad r = \hat{r} / \hat{e}_{\max}, \quad e = \hat{e} / \hat{e}_{\max} \qquad (3.5)$$

CHAPTER IV

ETHYL BENZENE PROCESS

4.1 Ethyl benzene process description

In the Ethyl benzene process, fresh feed ethylene (pure), fresh feed benzene (pure) mixed with the recycled benzene and nitrogen as inert (Figure 4.1A)

This reactant mixture is fed to continuous stirred tank reactor (CSTR) in series two reactor. Main chemical reaction taking place inside this reactor is

Benzene(B) + Ethylene(E) → *Ethyl benzene(EB)* (4.1)

Otherwise, there are further two reactions of ethylene, ethyl benzene and diethyl benzene;

Ethyl benzene (EB) + Ethylene (E) → *Diethyl benzene (DEB)* (4.2)

Diethyl benzene (DEB) +Benzene (B) → *2Ethyl benzene (EB)* (4.3)

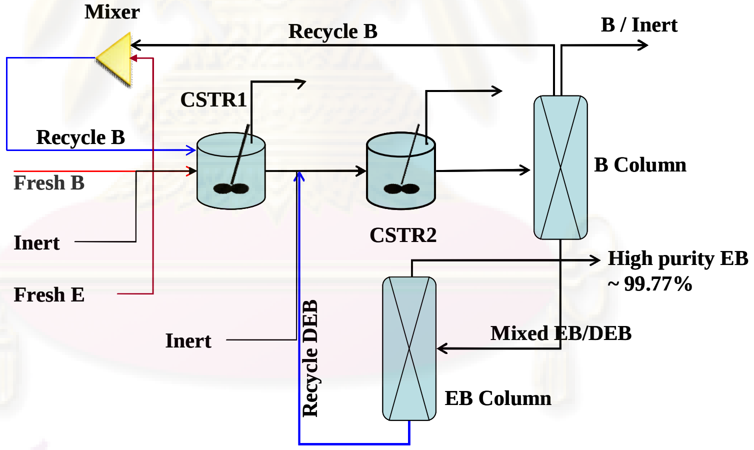


Figure 4.1AThe ethyl benzene process

The kinetic equations used for these three reactions are given below. These parameters are selected to give typical reaction conversations and yields.

$$R_1 = C_E C_B (1.528\times10^6) e^{-17,000/RT} \tag{4.4}$$

$$R_2 = C_E C_{EB} (2.778\times10^4) e^{-20,000/RT} \tag{4.5}$$

$$R_3 = C_{DEB} C_B (0.416) e^{-15,000/RT} \tag{4.6}$$

Where reaction rate are in kg-mol/sec/m3, concentration of benzene C_B, ethylene C_B, ethyl benzene C_{EB} and diethyl benzene C_{DEB} are in kg-mol/m3, activation energies are in cal/g-mol and temperature are in Kelvin.

To suppress this reaction, the large excess benzene is used so that the ethylene concentration is small. The excess benzene results in the recycle stream of process. The ratio fed of benzene to ethylene to the reaction section is about 2.5. Despite the large benzene excess, there is a significant amount of diethyl benzene produced in the reactor. It must be separated from the ethyl benzene and recycled back to the reaction section.

The reaction section consists of two CSTR reactors (200 m^3) in series. Fresh benzene feed and a recycle benzene stream are fed to the first reactor (R1), which operates at 160 ^{o}C and 9 bar. The exothermic heat of reaction is removed by generating low- pressure steam. The ethylene undergoes almost complete conversion in the first reactor. The effluent from the first reactor and a recycle DEB stream are fed to a second reactor (R2), which operates adiabatically. The composition of the effluent from the second reactor is about 55 mol% benzene, 35mol% ethyl benzene and 10 mol% diethyl benzene.

The reactor effluent is fed to a 24 tray distillation column (C1), operating at 4.5 bar. Recycle benzene goes overhead and the EB/DEB mixture goes out the bottom. The column has a reflux ratio of 0.9, consumes 15.8 MW in the reboiler and is 3.7m in diameter. The specifications on the stream leaving C1 are 0.1 mol% EB in the distillate and 0.1mol% benzene in the bottom.

The bottom is fed to a second distillation column (C2) that produces distillate product of high purity EB and bottoms with mostly DEB, which is recycled back to the second reactor. This 30 tray column operates at 2 bar, has a reflux ratio of 1, a diameter of 3.7 m and consumes 9.8 MW of energy in the reboiler. The reflux drum temperature is 165^{o}C, so low pressure steam is generated in the condenser. The specifications for this column are 0.1mol%EB in the bottom and 0.1 mol% DEB in distillate.

4.1.1 Steady-state modeling

The model of the ethyl benzene process used in this paper is a version of the model developed by Luyben (2002).A flowsheet of the Hysys model show in Figure 4.1B. The flowsheet streams data show in table 4.3. The equipment designs follow from Douglas (1988) and Luyben et al. (1998).

The Peng-Robinson (PR) equation of state is selected for the property estimation as it is very reliable for predicting the properties of hydrocarbons over a wide range of conditions and is generally recommended for oil, gas and petrochemical applications.

4.1.2 Operation constraints

1. Minimum production rate

   D2 ethyl benzene $\geq$ 626 kgmole/h

2. Benzene to ethylene feed ratio in reactor1 inlet (Prevent side reactions)

   rB $\geq$ 2.5

3. Maximum ethylene feed rate

   F toluene $\leq$ 630 kgmole/h

4. Operating pressure reactor 1

   P reactor $\leq$ 9.0 bar

5. Operating temperature reactor 1

   T reactor1 $\leq$ 160 C

6. Product purity at the ethyl benzene column distillation

   XD2, ethyl benzene $\geq$ 0.9977

7. Recycle benzene purity at the benzene column distillation

   XD1, benzene $\geq$ 0.9974

Table 4.1 Equipment data of ethyl benzene process

Unit		Specification
RX (Reactor):		
	Diameter (m)	5.537
	Length (m)	8.306
	Vessel volume (m^3)	200
Benzene column:		
	Number of theoretical trays	24
	Feed tray	13
	Diameter (m)	8.201
	Reboiler volume (m^3)	35
	Condenser volume (m^3)	66
Ethyl benzene column:		
	Number of theoretical trays	30
	Feed tray	7
	Diameter (m)	8.128
	Reboiler volume (m^3)	30
	Condenser volume (m^3)	27

4.1.3 Degree of freedom analysis

Table 4.2 Typical number of steady-state degree of freedom for some process units

Process unit	DOF
Each external feed stream	1 (feed rate)
Splitter	n-1 split fractions (n is the number of exit streams)
Mixer	0
Compressor, turbine, and pump	1 (work)
Adiabatic flash tank	0^a
Liquid phase reactor	1 (holdup)
Gas phase reactor	0^a
Heat exchanger	1 (duty or net area)
Column (e.g. distillation) excluding heat exchangers	0^a + number of side streams

Note a = Add 1 degree of freedom for each extra pressure that is set (need an extra valve, compressor, or pump), e.g., in flash tank, gas phase reactor, or column.

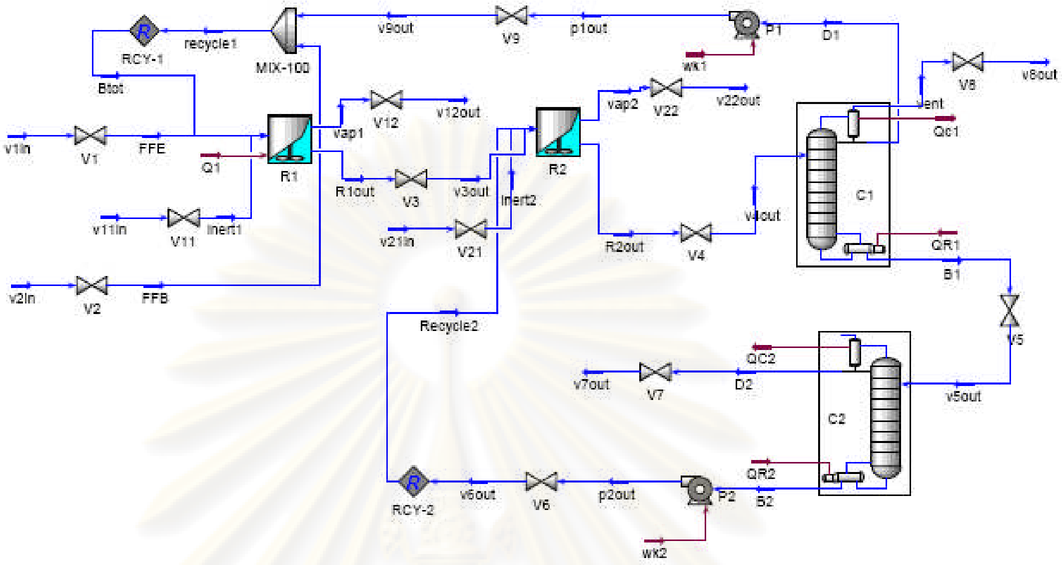


Figure 4.1B Ethyl benzene Hysys process flow sheet.

4.1.4 Manipulated variables

Number	symbol	List of overall manipulatable variables
1	V1	Ethylene fresh feed valve
2	V2	Benzene fresh feed valve
3	Q1	Reactor 1 heat duty
4	V3	Reactor 1 effluent valve
5	V4	Reactor 2 effluent valve
6	Qc1	Benzene column condenser heat duty
7	Qr1	Benzene column reboiler heat duty
8	V5	Benzene column bottom valve
9	V6	Ethyl benzene column bottom valve
10	V7	Ethyl benzene column overhead valve
11	V9	Benzene column overhead valve
12	Qc2	Ethyl benzene column condenser heat duty
13	Qr2	Ethyl benzene column reboiler heat duty

Table 4.3 Stream table for the nominal operating point for the ethyl benzene process (Luyben 2002)

Name	Btot	FFE	R1out	Recycle2	R2out	v3out	D1	B1	v1in	v2in	FFB
Vapour Fraction	0.0000	1.0000	0.0000	0.0000	0.0000	0.0042	0.0000	0.0000	1.0000	0.0000	0.0000
Temperature (°C)	86.10	50.85	157.20	217.60	169.60	156.90	107.20	211.80	51.69 *	49.92 *	49.96
Pressure (kPa)	1000.00	1000.00	1000.00	860.40	860.40	860.40	427.40	442.80	1100 *	1100 *	1000.00
Total Molar Flow (kgmole/h)	1614.00	630.60	1621.00	152.70	1767.00	1621.00	989.90	766.50	630.60	623.90	623.90
Mass Flow (kg/h)	1.26E+05	1.77E+04	1.44E+05	2.05E+04	1.64E+05	1.44E+05	7.73E+04	8.61E+04	1.77E+04	4.87E+04	4.87E+04
Heat Flow(kJ/h)	9.34E+07	3.35E+07	7.93E+07	1.31E+06	7.77E+07	7.93E+07	6.03E+07	1.57E+07	3.35E+07	3.29E+07	3.29E+07
Mole Fraction											
Ethylene	0.0000	1.0000	0.0033	0.0000	0.0000	0.0033	0.0000	0.0000	1.0000	0.0000	0.0000
Benzene	0.9985	0.0000	0.6848	0.0000	0.5622	0.6848	0.9975	0.0007	0.0000	1.0000	1.0000
Ethyl benzene)	0.0004	0.0000	0.2316	0.0004	0.3386	0.2316	0.0007	0.7797	0.0000	0.0000	0.0000
Diethyl benzene	0.0000	0.0000	0.0770	0.9996	0.0953	0.0770	0.0000	0.2197	0.0000	0.0000	0.0000
Nitrogen	0.0011	0.0000	0.0034	0.0000	0.0039	0.0034	0.0018	0.0000	0.0000	0.0000	0.0000
Molar Flow (kgmole/h)											
Ethylene	0.0000	630.6000	5.3493	0.0000	0.0000	5.3493	0.0000	0.0000	630.6000	0.0000	0.0000
Benzene	1611.5790	0.0000	1110.0608	0.0000	993.4074	1110.0608	987.4253	0.5366	0.0000	623.9000	623.9000
Ethyl benzene)	0.6456	0.0000	375.4236	0.0611	598.3062	375.4236	0.6929	597.6401	0.0000	0.0000	0.0000
Diethyl benzene	0.0000	0.0000	124.8170	152.6389	168.3951	124.8170	0.0000	168.4001	0.0000	0.0000	0.0000
Nitrogen	1.7754	0.0000	5.5114	0.0000	6.8913	5.5114	1.7818	0.0000	0.0000	0.0000	0.0000

Table 4.3 (Continued) Stream table for the nominal operating point for the ethyl benzene process (Luyben 2002)

Name	p1out	recycle1	v5out	D2	B2	p2out	v6out	v7out	inert1	vap1	v11in
Vapour Fraction	0.0000	0.0000	0.3030	0.0000	0.0000	0.0000	0.0000	0.1689	1.0000	1.0000	1.0000
Temperature (°C)	107.70	86.10	173.40	162.20	217.10	217.60	217.60	136.90	50.17	157.20	50.49 *
Pressure (kPa)	1159.00	1000.00	196.00	184.90	201.40	1087.00	860.40	100.0 *	1000.00	1000.00	1200 *
Total Molar Flow (kgmole/h)	989.90	1614.00	766.50	613.80	152.70	152.70	152.70	613.80	4.77	4.05	4.77
Mass Flow (kg/h)	7.73E+04	1.26E+05	8.61E+04	6.56E+04	2.05E+04	2.05E+04	2.05E+04	6.56E+04	1.34E+02	2.24E+02	1.34E+02
Heat Flow(kJ/h)	6.04E+07	9.34E+07	1.57E+07	9.70E+06	-1.34E+06	-1.31E+06	-1.31E+06	9.70E+06	3.27E+03	2.40E+05	3.27E+03
Mole Fraction											
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2054	0.0000
Benzene	0.9975	0.9985	0.0007	0.0009	0.0000	0.0000	0.0000	0.0009	0.0000	0.4608	0.0000
Ethyl benzene)	0.0007	0.0004	0.7797	0.9736	0.0004	0.0004	0.0004	0.9736	0.0000	0.0480	0.0000
Diethyl benzene	0.0000	0.0000	0.2197	0.0256	0.9996	0.9996	0.9996	0.0256	0.0000	0.0055	0.0000
Nitrogen	0.0018	0.0011	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.2804	1.0000
Molar Flow (kgmole/h)											
Ethylene	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.8313	0.0000
Benzene	987.4253	1611.5790	0.5366	0.5524	0.0000	0.0000	0.0000	0.5524	0.0000	1.8649	0.0000
Ethyl benzene)	0.6929	0.6456	597.6401	597.5957	0.0611	0.0611	0.0611	597.5957	0.0000	0.1943	0.0000
Diethyl benzene	0.0000	0.0000	168.4001	15.7133	152.6389	152.6389	152.6389	15.7133	0.0000	0.0223	0.0000
Nitrogen	1.7818	1.7754	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	4.7660	1.1348	4.7660

Table 4.3 (Continued) Stream table for the nominal operating point for the ethyl benzene process (Luyben 2002)

Name	v12out	vap2	inert2	v21in	v22out	v4out	vent	v8out	v9out
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	0.0863	1.0000	1.0000	0.0000
Temperature (°C)	153.90	169.60	50.10	50.33 *	167.40	158.70	107.20	105.60	107.70
Pressure (kPa)	600.0 *	860.40	860.40	1000 *	600.0 *	434.00	427.40	250.0 *	1000.00
Total Molar Flow (kgmole/h)	4.05	4.17	2.95	2.95	4.17	1767.00	10.48	10.48	989.90
Mass Flow (kg/h)	2.24E+02	2.66E+02	8.27E+01	8.27E+01	2.66E+02	1.64E+05	5.63E+02	5.63E+02	7.73E+04
Heat Flow(kJ/h)	2.40E+05	2.44E+05	2.04E+03	2.04E+03	2.44E+05	7.77E+07	4.97E+05	4.97E+05	6.04E+07
Mole Fraction									
Ethylene	0.2054	0.0011	0.0000	0.0000	0.0011	0.0000	0.0009	0.0009	0.0000
Benzene	0.4608	0.5309	0.0000	0.0000	0.5309	0.5622	0.5123	0.5123	0.9975
Ethyl benzene)	0.0480	0.1025	0.0000	0.0000	0.1025	0.3386	0.0001	0.0001	0.0007
Diethyl benzene	0.0055	0.0103	0.0000	0.0000	0.0103	0.0953	0.0000	0.0000	0.0000
Nitrogen	0.2804	0.3552	1.0000	1.0000	0.3552	0.0039	0.4867	0.4867	0.0018
Molar Flow (kgmole/h)									
Ethylene	0.8313	0.0046	0.0000	0.0000	0.0046	0.0000	0.0094	0.0094	0.0000
Benzene	1.8649	2.2128	0.0000	0.0000	2.2128	993.4074	5.3689	5.3689	987.4253
Ethyl benzene)	0.1943	0.4272	0.0000	0.0000	0.4272	598.3062	0.0010	0.0010	0.6929
Diethyl benzene	0.0223	0.0429	0.0000	0.0000	0.0429	168.3951	0.0000	0.0000	0.0000
Nitrogen	1.1348	1.4805	2.9530	2.9530	1.4805	6.8913	5.1006	5.1006	1.7818

Table 4.3 (Continued) Stream table for the nominal operating point for the ethyl benzene process (Luyben 2002)

Name	Q1	Qc1	Qr1	wk1	Qc2	Qr2	wk2
Heat Flow (kJ/h)	4.674e+007 *	7.115e+007 *	5.840e+007 *	9.678e+004 *	4.282e+007 *	3.530e+007 *	3.478e+004 *

* specified by Luyben (2002)

CHAPTER V

CONTROL STRUCTURE ALTERNATIVES

This chapter is to present the new control structures of energy integrated process. Moreover, the three designed control structures are also compared between three structures on rigorous dynamic simulation by using the commercial software HYSYS version 3.1.

5.1 New Planwide Control Strategies

Step 1: Establishment of control objectives

The objectives we decomposed into two levels: Plantwide level and Unit level. Plantwide objectives typically include production rate, product quality and minimum economic loss. Unit operation objectives typically include equipment constrains, safety concern and smooth operation (process stability).

Ethyl benzene process:

Plantwide objectives:

The goals of plantwide are to produce ethyl benzene product at 626.0 kgmole/h with 0.9977 purity and minimum loss of economic. We use the optimum value operation follow by Araujo et al. (2006) for minimum utility usage in normal operating point.

Unit operation objectives:

The unit operation targets are avoid drift of process variables (process stability) and control equipment constrains.

Summary:

(1) Production rate: 626.0 kgmole/h

(2) Product quality: ethyl benzene purity ≥ 99.77 %

(3) Process constrains:

1. The reactor temperature should be around 160°C. This is an optimization decision to have better reaction rate.

2. The recycle diethyl benzene temperature should not exceed 222°C to avoid kinetic consider, adiabatically operation.

3. The ratio of benzene to ethylene has to be at least 2.5 at the reactor inlet. Excess benzene suppresses the further second order reaction of ethylene and ethyl benzene.

Step 2: Selection of controlled variables to maintain product quality and to satisfy safety operational and environmental constrains. The selected CVs are ranked using the Fixture Point theorem.

The variables that set the inventory levels are controlled variables list. They are all liquid levels (except in recycle systems), and all gas pressures.

Plantwide level:

Consider material recycle loop it causes a system to be born “snowball effect”. Ethyl benzene processes have two recycle streams: two liquid recycles, a few contaminates in benzene recycle stream and diethyl benzene did not lead to snowball effect.

Unit level:

The reactor is important unit because it has liquid phase exothermic reaction. The reactor can operate at maximum temperature of catalyst damaging or result in a vessel meltdown. The temperature and pressure of reactor should be controlled.

Apply the Fixture Point Theorem to select the appropriated controlled variables. The most disturbed points must be satisfactorily controlled by giving them consideration before other variable. Screening output variables for identification controlled variable by using input variables change (change one percentage of manipulated variable).

Table 5.1 show IAE summation result of flow variables. We select effluent flow rate Reactor 1 stream and outlet stream V4, which are mostly affected.

Table 5.1 IAE summation results of flow variable

Rank	Flow variables	Symbol of position on flowsheet	Sum IAE
1	Effluent Reactor 1 stream	R1out	1.8207
2	Effluent V4 stream	v4out	1.6022
3	Bottom stream of Benzene Column	B1	0.3724
4	Recycle Benzene	Btot	0.3492
5	Fresh feed benzene	FFB	0.3361
6	Distillation stream of E-Benzene Column	D2	0.3243
7	Fresh feed ethylene	FFE	0.2763
8	Distillation stream of Benzene Column	D1	0.2291
9	Bottom stream of E-Benzene Column	B2	0.0414

Table 5.2 IAE summation results of temperature variable

Rank	Temperature variables	Symbol of position on flowsheet	Sum IAE
1	Distillation stream of Benzene Column	D1	0.4197
2	Recycle Benzene	Btot	0.3293
3	Effluent Reactor 1 stream	R1out	0.2420
4	Effluent V4 stream	v4out	0.2260
5	Bottom stream of E-Benzene Column	B2	0.2185
6	Bottom stream of Benzene Column	B1	0.2075
7	Distillation stream of E-Benzene Column	D2	0.2015
8	Fresh feed ethylene	FFE	0.0042
9	Fresh feed benzene	FFB	0.0002

Table 5.3 IAE summation results of pressure variable

Rank	Pressure variables	Symbol of position on flowsheet	Sum IAE
1	Distillation stream of Benzene Column	D1	2.1129
2	Effluent V4 stream	v4out	2.1076
3	Bottom stream of Benzene Column	B1	2.1070
4	Distillation stream of E-Benzene Column	D2	0.9567
5	Bottom stream of E-Benzene Column	B2	0.9552
6	Recycle Benzene	Btot	0.5296
7	Effluent Reactor 1 stream	R1out	0.5198
8	Fresh feed ethylene	FFE	0.4967
9	Fresh feed benzene	FFB	0.4964

Table 5.4 and Figure 5.1 show the IAE summation result of Tray temperature for benzene column from Fixture Point method to find an appropriate controlled of Tray temperature. The temperature Tray 8 benzene column is the appropriate controlled variables for all designed control structure because it is highest sensitive.

Figure 5.2 shows the temperature gradient for benzene column (this value is the slope value of Tray temperature for benzene column from steady state value. The appropriate controlled variables of Tray benzene column from Fixture Point method versus the steady state value is similar result.

Table 5.4 IAE summation result of Tray benzene temperature deviation for benzene column

Tray	Sum IAE	Tray	Sum IAE	Tray	Sum IAE	Tray	Sum IAE
1	0.5130	7	2.7985	13	0.5810	19	0.5516
2	0.6016	8	2.8504	14	0.6190	20	0.5323
3	0.8060	9	2.3856	15	0.6411	21	0.5218
4	1.7092	10	1.6816	16	0.6398	22	0.5157
5	1.1777	11	1.0814	17	0.6157	23	0.5122
6	2.3202	12	0.7089	18	0.5814	24	0.5106

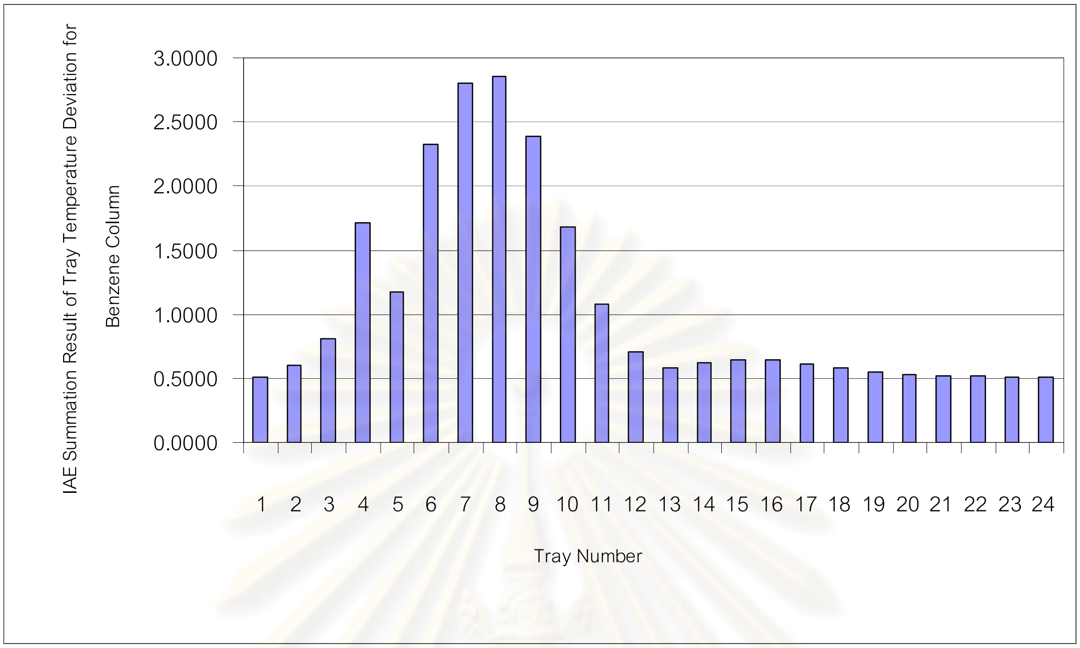


Figure 5.1 IAE summation result of Tray benzene temperature deviation for benzene column

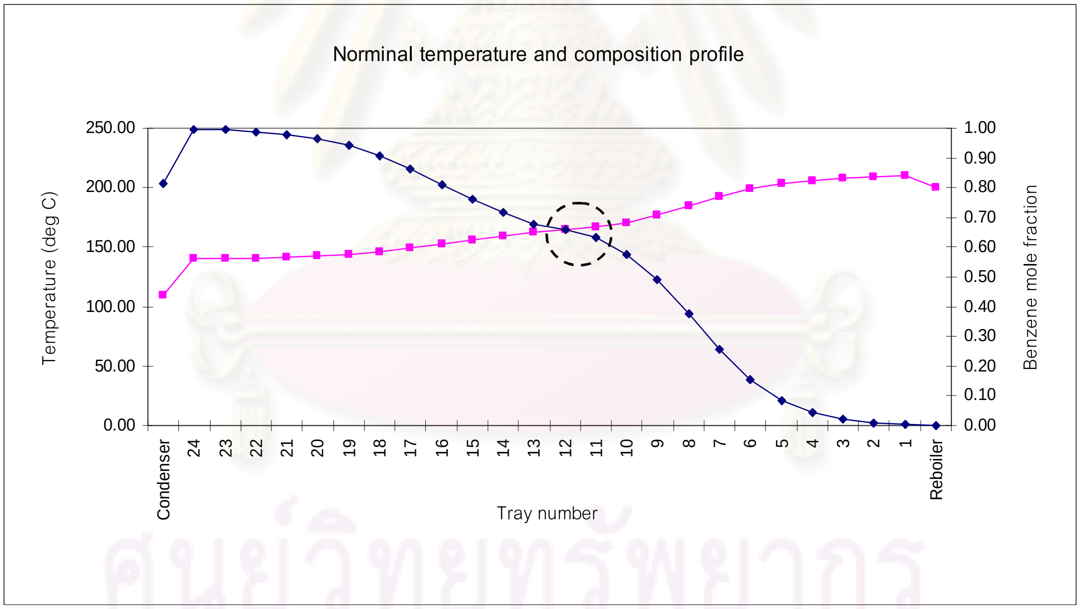


Figure 5.2 Temperature gradient for benzene column from steady state value

Table 5.5 and Figure 5.3 show the IAE summation result of Tray temperature for Ethyl benzene column from Fixture Point method to find an appropriate controlled of Tray temperature. The temperature Tray 8 Ethyl benzene column is the appropriate controlled variables for all designed control structure because it is highest sensitive. Although the Tray

26 Ethyl benzene column has a high sensitive, which meant we should to control temperature at this tray to prove production rate.

Figure 5.5 shows the temperature gradient for Ethyl benzene column (this value is the slope value of Tray temperature for Ethyl benzene column from steady state value. The appropriate controlled variables of Tray Ethyl benzene column from Fixture Point method versus the steady state value is similar result

Table 5.5 IAE summation result of Tray Ethyl benzene temperature deviation for Ethyl benzene column

Tray	Sum IAE	Tray	Sum IAE	Tray	Sum IAE	Tray	Sum IAE
1	0.2170	9	0.2611	17	0.1716	25	0.2476
2	0.2146	10	0.1824	18	0.1729	26	0.2625
3	0.2081	11	0.1792	19	0.1755	27	0.2585
4	0.1955	12	0.1762	20	0.1795	28	0.2324
5	0.1789	13	0.1741	21	0.1859	29	0.2061
6	0.1655	14	0.1723	22	0.1949	30	0.1957
7	0.1905	15	0.1714	23	0.2078		
8	0.2632	16	0.1713	24	0.2259		

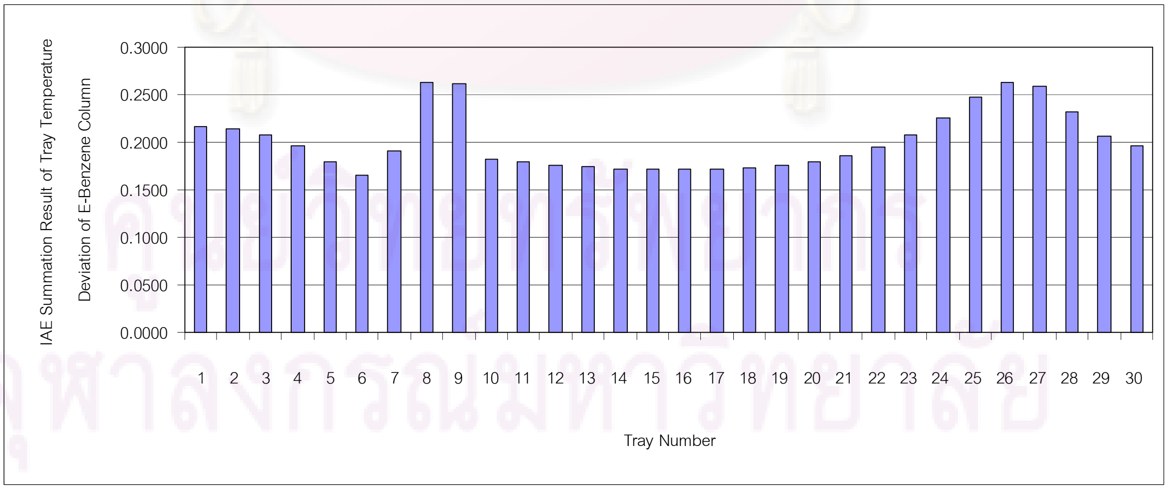


Figure 5.3 IAE summation result of Tray Ethyl benzene temperature deviation for Ethyl benzene column

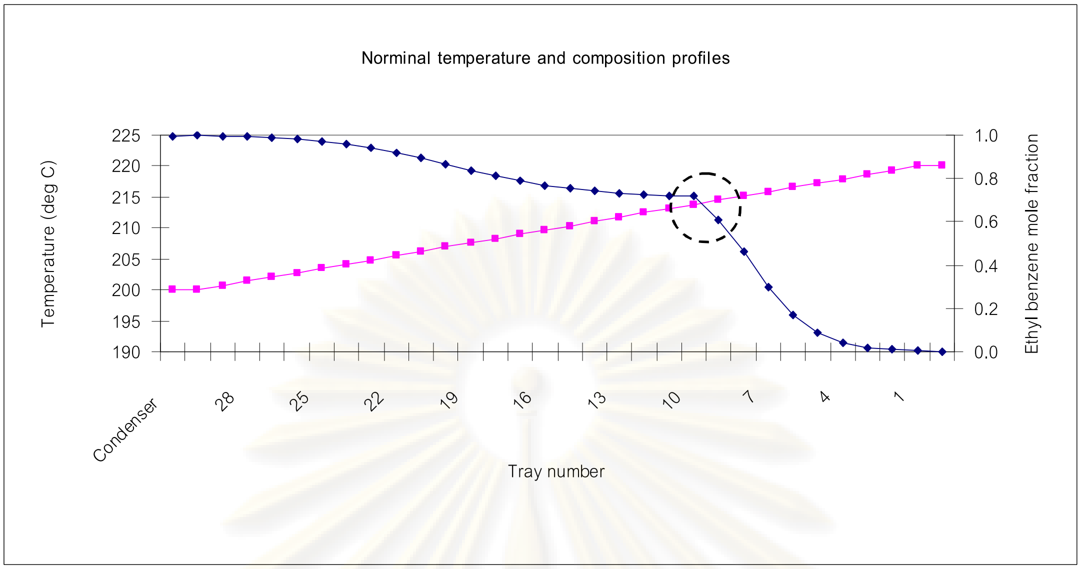


Figure 5.4 Temperature Gradient for Ethyl benzene column from steady state value

Step 3: Selection of manipulated variables and measurements via DOF analysis

We use Ethyl benzene control structure of Buyben(2002) as base case control structure (CS0), has 9 steady-state degrees of freedom and 5 liquid levels that need to be controlled. Totally, we have 15 degree of freedom for control structure design.

For designed control structure I (CS1), designed control structure II (CS2), there are 13 control degree of freedom same as the base case

For designed control structure III (CS3), there is 25 control degree of freedom same as the base case, one more degree of free is Ethyl benzene reflux value

The control degrees of freedom for each control structures are shown in Table 5.6

Table 5.6 The control degree of freedom for each control structures

Position	CS0	CS1	CS2	CS3
FFE fresh feed valve	X	X	X	X
FFB fresh feed valve	X	X	X	X
Reactor 1(R1) liquid effluent valve	X	X	X	X
Reactor 2(R2) liquid effluent valve	X	X	X	X
Reactor 1 cooling duty	X	X	X	X
Condenser duty of Benzene Column	X	X	X	X
Reboiler duty of Benzene Column	X	X	X	X
Distillate valve of Benzene Column	X	X	X	X
Bottom valve of Benzene Column	X	X	X	X
Condenser duty of Ethyl benzene Column	X	X	X	X
Reboiler duty of Ethyl benzene Column	X	X	X	X
Distillate valve of Ethyl benzene Column	X	X	X	X
Bottom valve of Ethyl benzene Column	X	X	X	X
Reflux value of Ethyl benzene Column	_	_	_	_
Reflux value of Ethyl benzene Column	_	_	_	X
Degree of freedom	13	13	13	14

Step 4: Energy management via heat exchanger networks.

In this research, we are not design the heat exchange network

Step 5: Selection of control configuration using various tools available.

Selection of control configuration use heuristic process knowledge. The criteria for selecting an adjustable variable include: causal relationship between the valve and controlled variable, automated valve to influence the selected flow, fast speed of response, ability to compensate for large disturbances and ability to adjust the manipulated variable rapid and with little upset to the reminder of the plant.

For reactor: There are two reactor (CSTR) exothermic reactions. The reactors have no external heat transfer. They operate adiabatically, and the required low temperature. The Reactor 1 has cooling water to remove heat from reaction. The inert streams in two reactors remove some heat to environment as well. The reactor level must be controlled; it can level directly or via flow control.

Consider distillation section: The control schemes are stand-alone manipulated variables are condenser duty, distillate valve, reboiler heat input and bottom product valve.

In CS3 designed control structure, we use reflux flow to control temperature Ethyl benzene column tray.

Step 6: Completing control structure design by checking the component balance.

Ethyl benzene process has five components to be accounted for: There are benzene, ethylene, ethyl benzene, 1,2-diethylbenzene and nitrogen(inert). The composition of two fresh feed are ethylene and benzene with benzene recycle stream. Nitrogen is feed to reactor, almost nitrogen release to environment. The product ethyl benzene, side product 1,2-diethylbenzene, the excess reactant benzene and impurities leave in benzene recovery column, distillated benzene return to the reactor 1. The bottom stream leave in ethyl benzene column, the distillate is product and bottom stream is 1,2- diethylbenzene return to reaction section.

Step 7: Selection of controller type: single loops or MPC.

In this research, all controller type is single-input-single-output loop. There are temperature controller, flow controller and level controller. Temperature controllers are PIDs which are tuned using relay feedback. Flow controllers are PIs. Proportional only level controllers are used and their parameters are heuristic valves. All control valves are half-open at nominal operating condition.

Step 8: Validation via rigorous dynamic simulation.

Using software HYSYS to evaluate performance for ethyl benzene process of all designed control structures and compare with base case control structure (Luyben, 2002) at dynamic simulation

5.2 Design of Plant wide Control Structure

In this research, we apply the new design procedure by Wongsri(2009) to design control structures for ethyl benzene process. The objects were decomposed into level: Plantwide level and Unit level. In all of these control structure (CS0, CS1,CS2 and CS3) the same loops are used as follows:

Plantwide level

- Valve V2 is manipulated to control the make up benzene flow rate

Unit level

Reactor 1 unit:

- Valve V1 is manipulated to control ethylene fresh feed flow rate
- Valve V2 is manipulated to control reactor 1 level
-Cooling duty is manipulated to removal heat from reaction

Reactor 2 unit:

-Valve V3 is manipulated to control reactor 2 level

Benzene column unit:

-Valve v4 is manipulated to control feed rate to benzene column

-Heat duty of condenser (qc1) is manipulated to control the benzene column pressure

-Heat duty of reboiler (qr1) is manipulated to control the benzene column Tray temperature

-Valve V5 is manipulated to control the benzene column reboiler level.

-Valve V9 is manipulated to control the benzene column condenser level.

Ethyl benzene column unit:

-Heat duty of condenser (qc2) is manipulated to control the ethyl benzene column pressure

-Heat duty of reboiler (qr2) is manipulated to control the ethyl benzene column Tray temperature

-Valve V6 is manipulated to control the ethyl benzene column reboiler level.

-Valve V7 is manipulated to control the ethyl benzene column condenser level.

5.2.1 Base case control structure (CS0)

All of refer manipulated controls in unit level are applied to CS0

5.2.2 Design of control structure (CS1)

We considered IAE summation Table 5.1, that show effluent Reactor 1 stream has highest disturbance. For CS1 control structure, we have changed valve V3 to control reactor 1 flow outlet.

5.2.3 Design of control structure (CS2)

Further design in the control structure CS1, the liquid level control is applied to Reactor 2 inventory. The CS2 control structure, we have changed valve V4 to control level reactor 2

5.2.4 Design of control structure (CS3)

The IAE summation table 5.5 show highest IAE that be controlled normally in the control structure CS0. Further consider is temperature tray which has the second IAE high value. We add reflux flow to control ethylene column Tray 26 temperature in control structure CS3

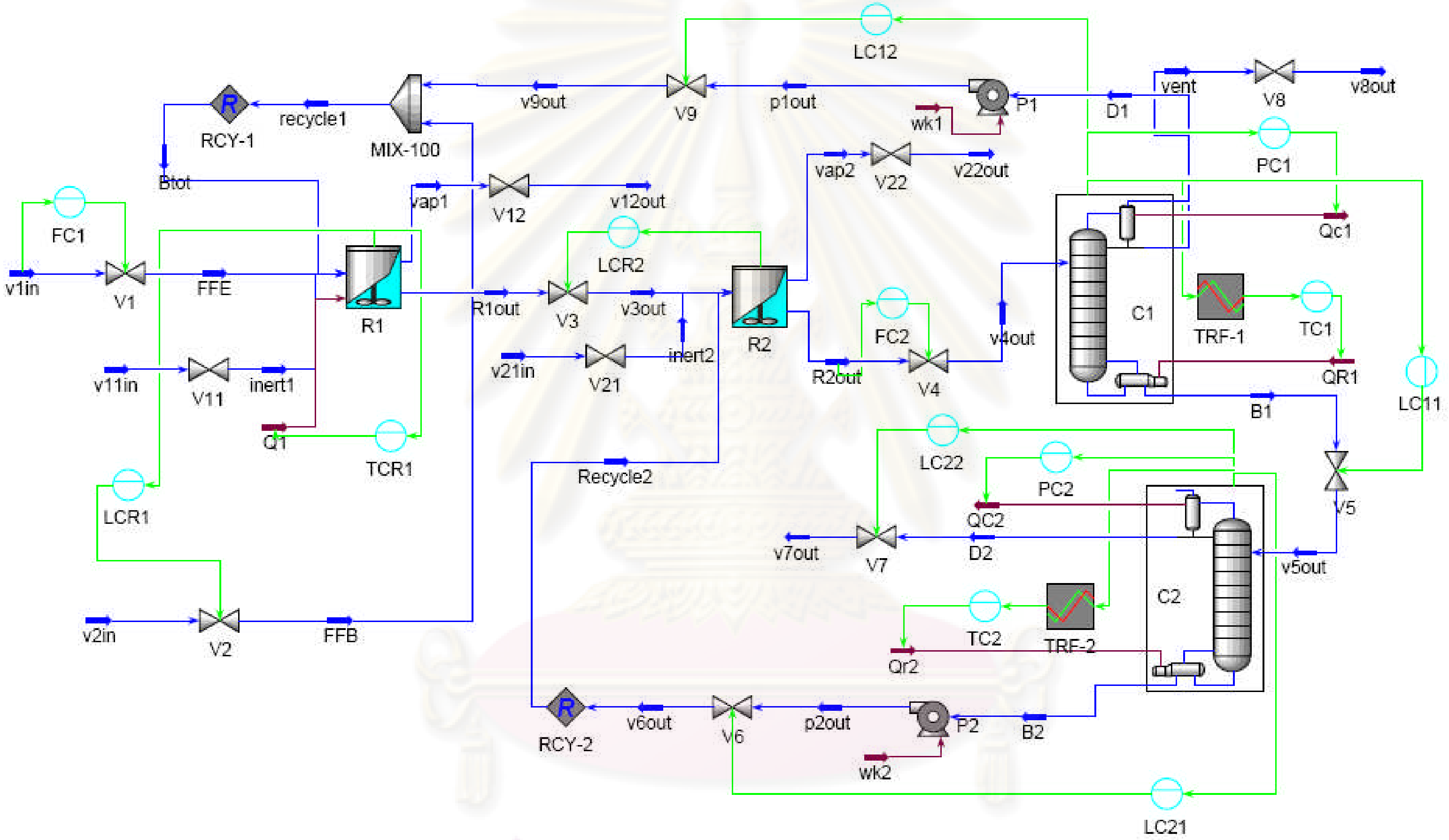


Figure 5.5 Base case control structure (CS0) for ethyl benzene process, Luyben (2002)

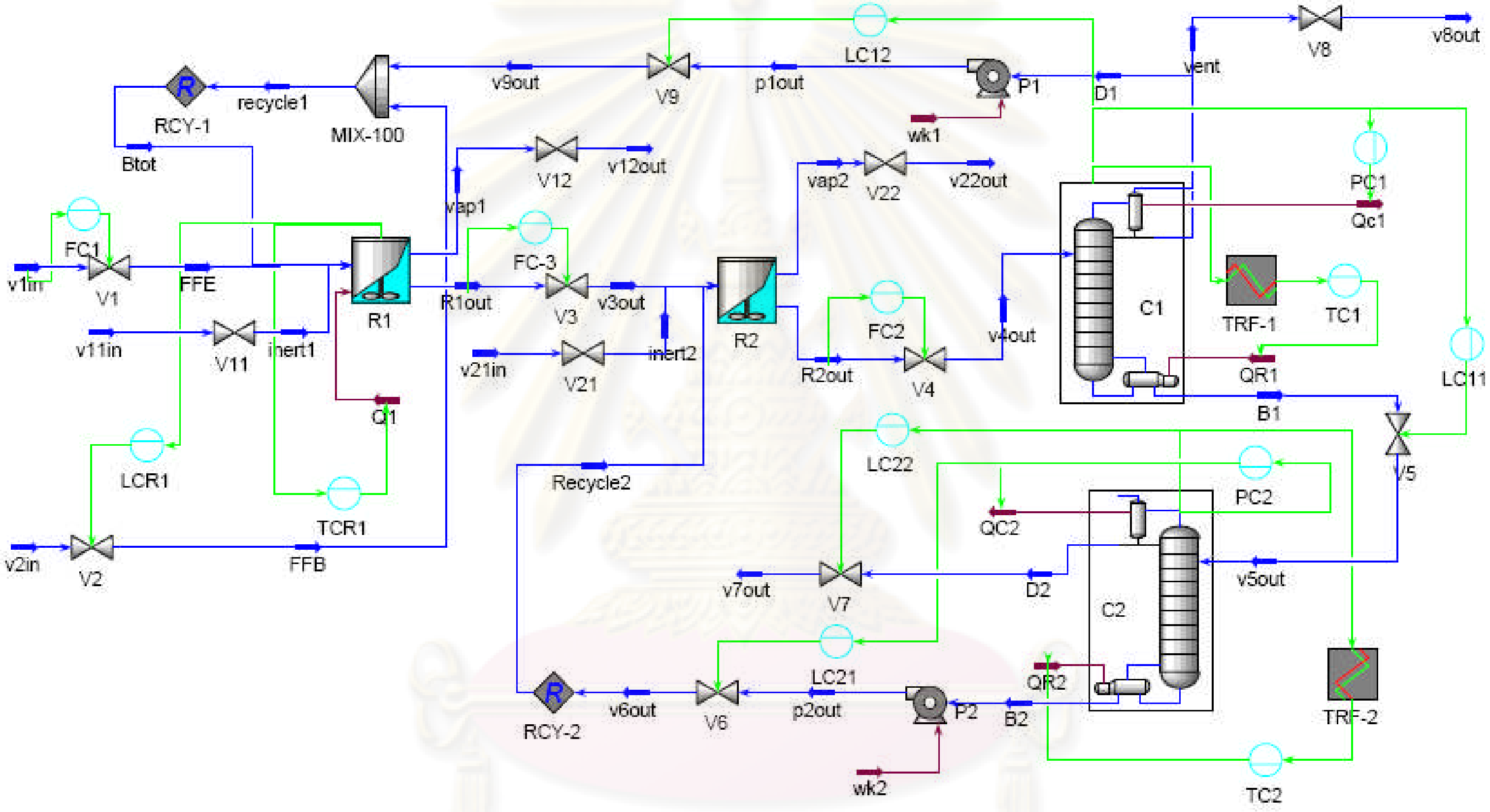


Figure 5.6 Designed control structure I(CS1) for ethyl benzene process

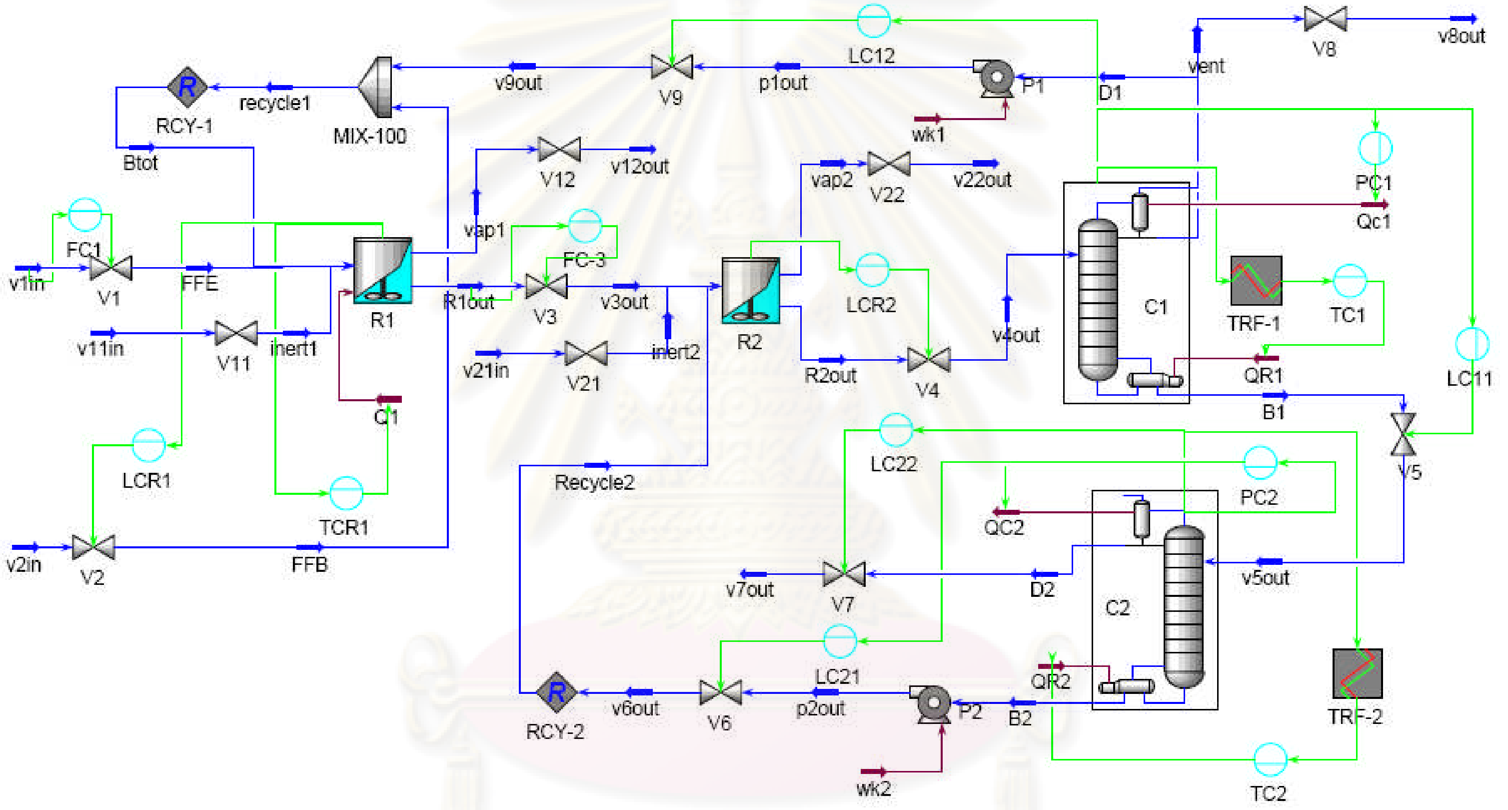


Figure 5.7 Designed control structure II(CS2) for ethyl benzene process

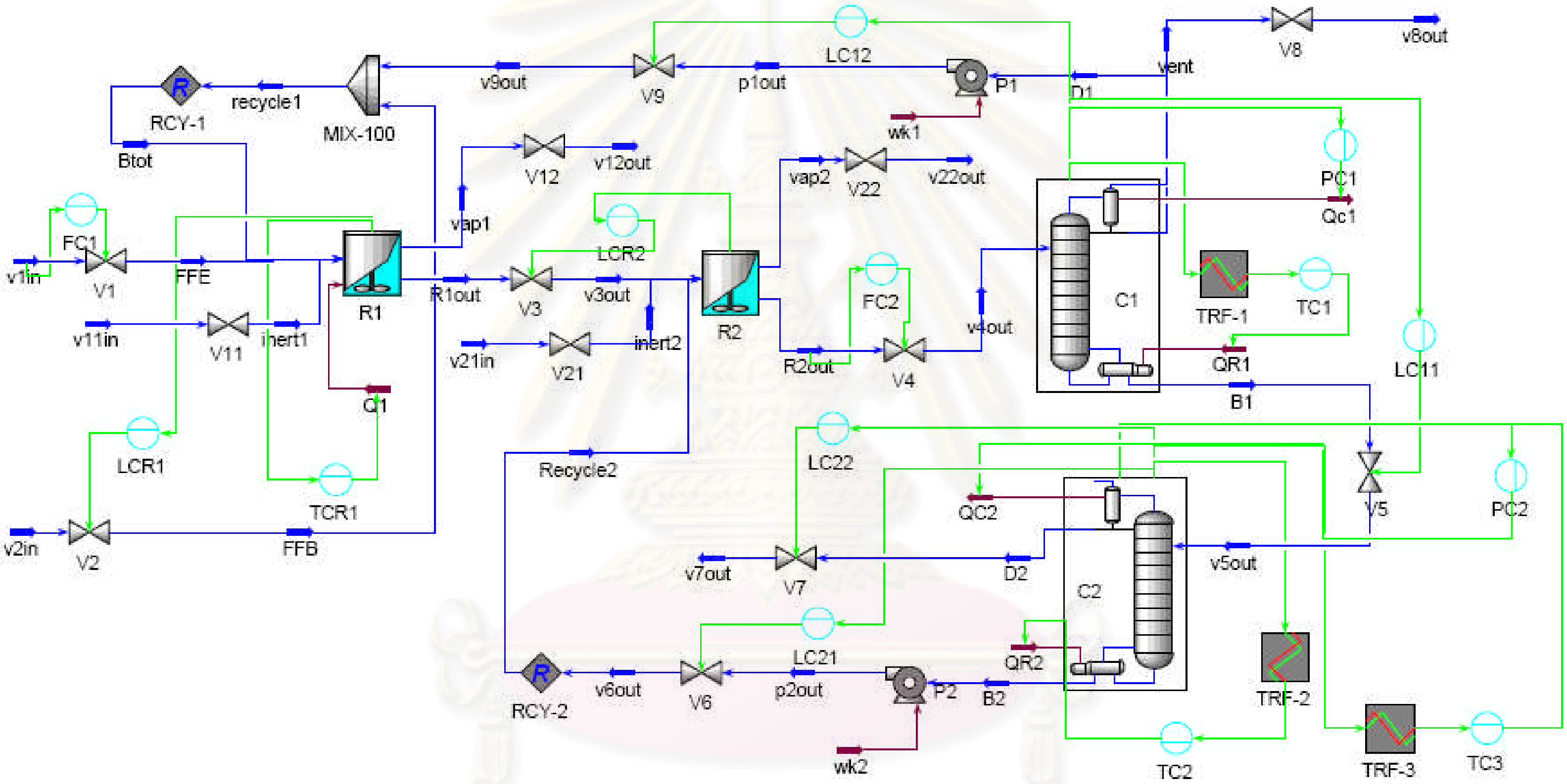


Figure 5.8 Designed control structure III(CS3) for ethyl benzene process

5.3 Dynamic Simulation Results

To illustrate the dynamic behaviors of base case control structure by Luben (2002) and all design control structure using new design procedure of Wongsri (2002), two types of disturbances: material and thermal disturbance are used in evaluation of the base case control structure(CS0) and all new design control structure (CS1,CS2 and CS3)

Two types of disturbance are used to test response of system:

(1) Change in fresh feed molar flow rate by step change of molar flow rate±10% of FFE fresh feed ethylene from 630.60 kg-mole/hr to 693.66 kg-mole/hr at 50 min to 150 min and from 693.66 kg-mole/hr to 567.59 kg-mole/hr at 150 min to 250 min

(2) Change in fresh feed temperature by step change of temperature ±10% of FFE fresh feed ethylene from 51.69 °C to 56.86 °C at 50 min to 150 min and from 56.86 °C to 46.52 °C at 150 min to 250 min

(3) Change in fresh feed temperature by step change of temperature±10% of FFB fresh feed benzene from 49.92 °C to 54.91 °C at 50 min to 150 min and from 54.91 °C to 44.93 °C at 150 min to 250 min

5.3.1 Change in material disturbance of ethylene fresh feed molar flow rate for all control structures (base case control structure (CS0), design control structure I (CS1) to designed control structure III (CS3)

Dynamic response change in fresh feed molar flow rate by step change of molar flow rate±10% of FFE fresh feed ethylene from 630.60 kg-mole/hr to 693.66 kg-mole/hr at 50 min to 150 min and from 693.66 kg-mole/hr to 567.59 kg-mole/hr at 150 min to 250 min

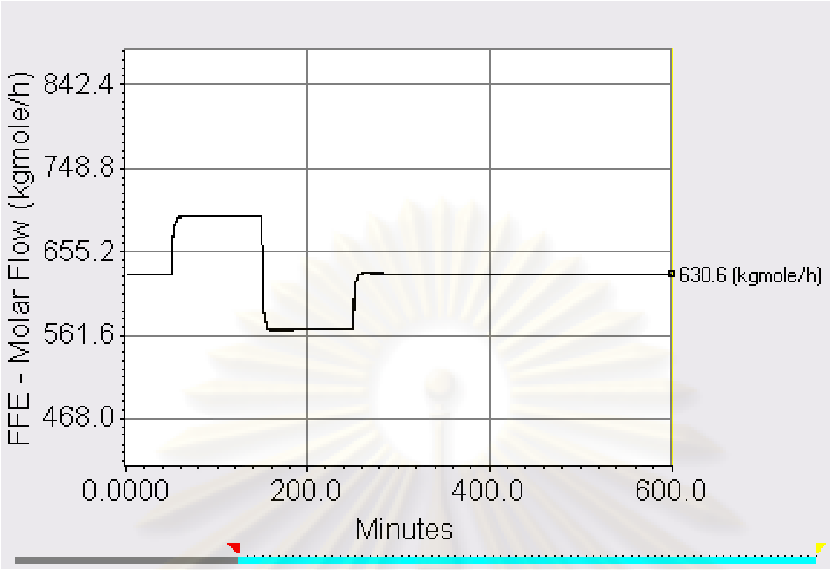
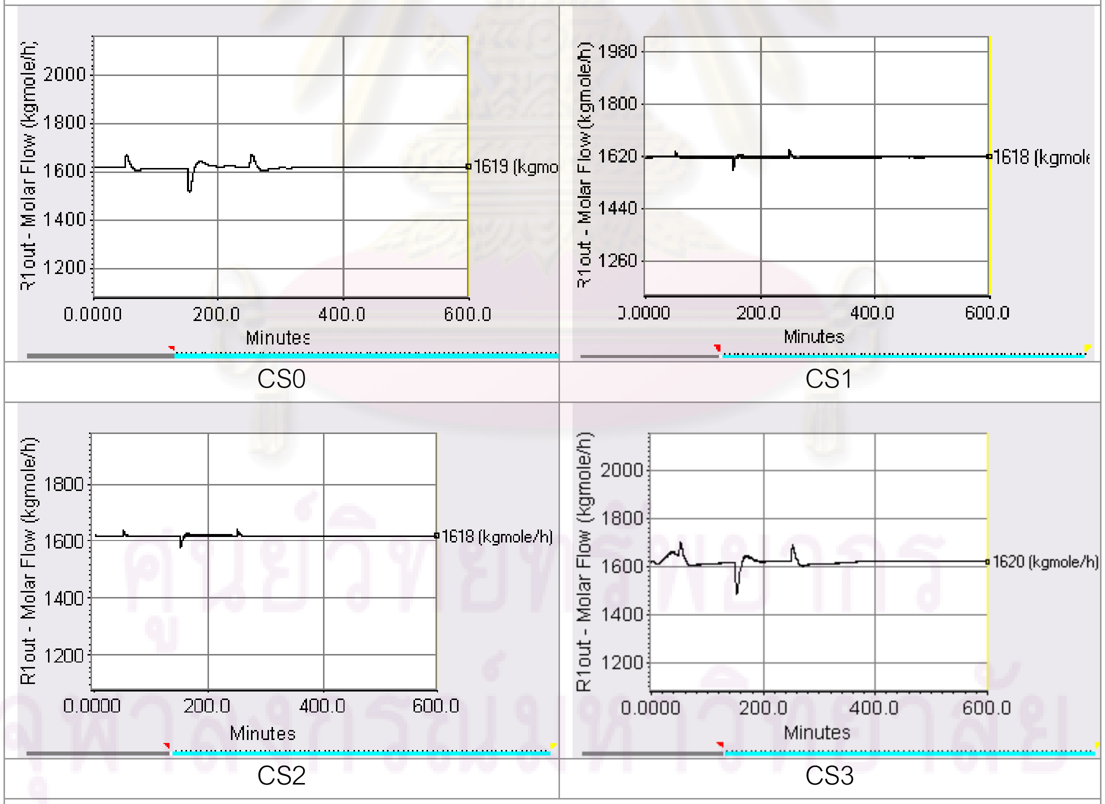


Figure 5.9 Dynamic responses of ethyl benzene process by step change of molar flow rate±10% of FFE fresh feed ethylene

Effluent molar flow reactor1(R1)



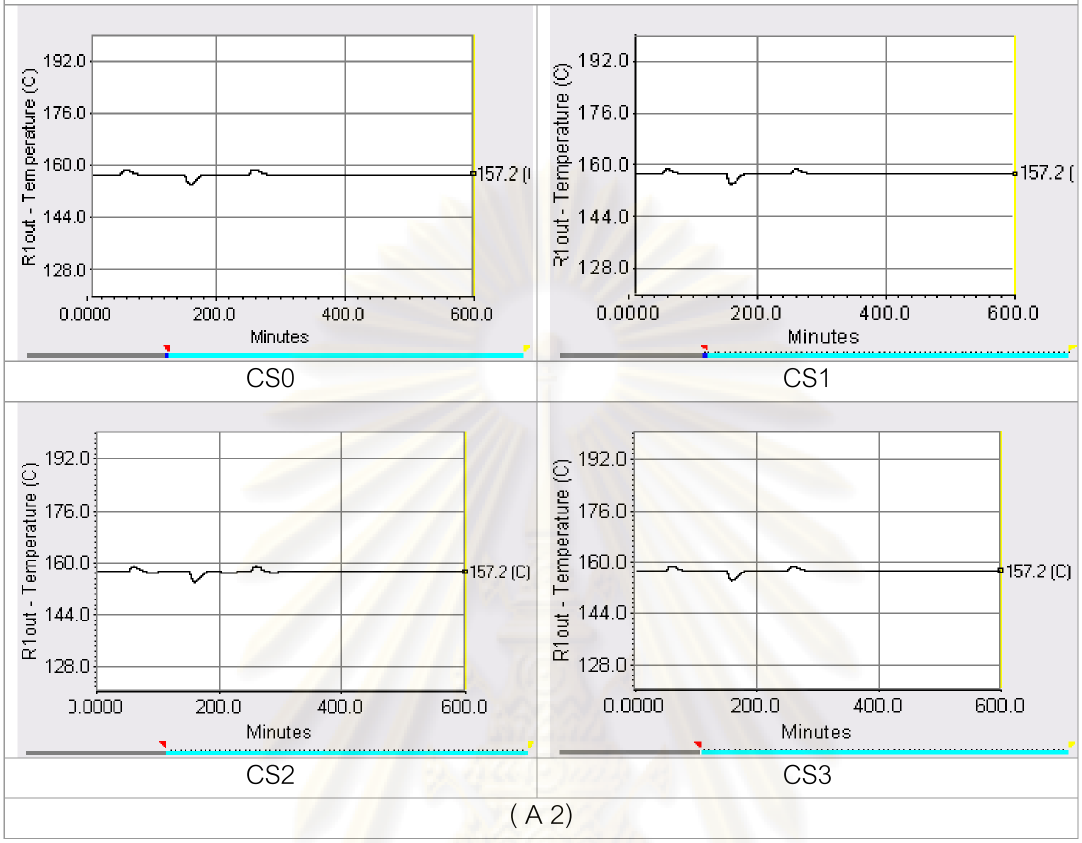
CS0 CS1

CS2 CS3

(A1)

Figure 5.10 Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Effluent Temperature reactor1(R1)



CS0

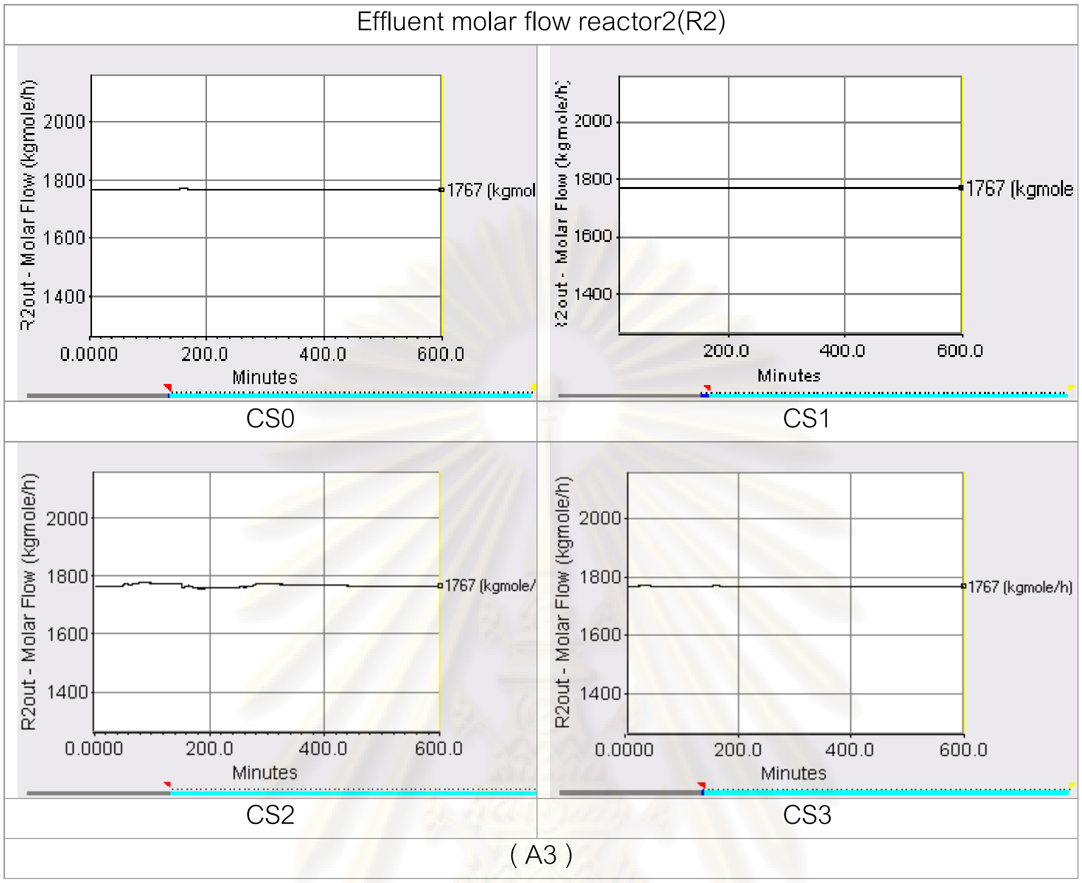
CS1

CS2

CS3

(A 2)

Figure 5.10 (Continued) Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene



Effluent molar flow reactor2(R2)

CS0

CS1

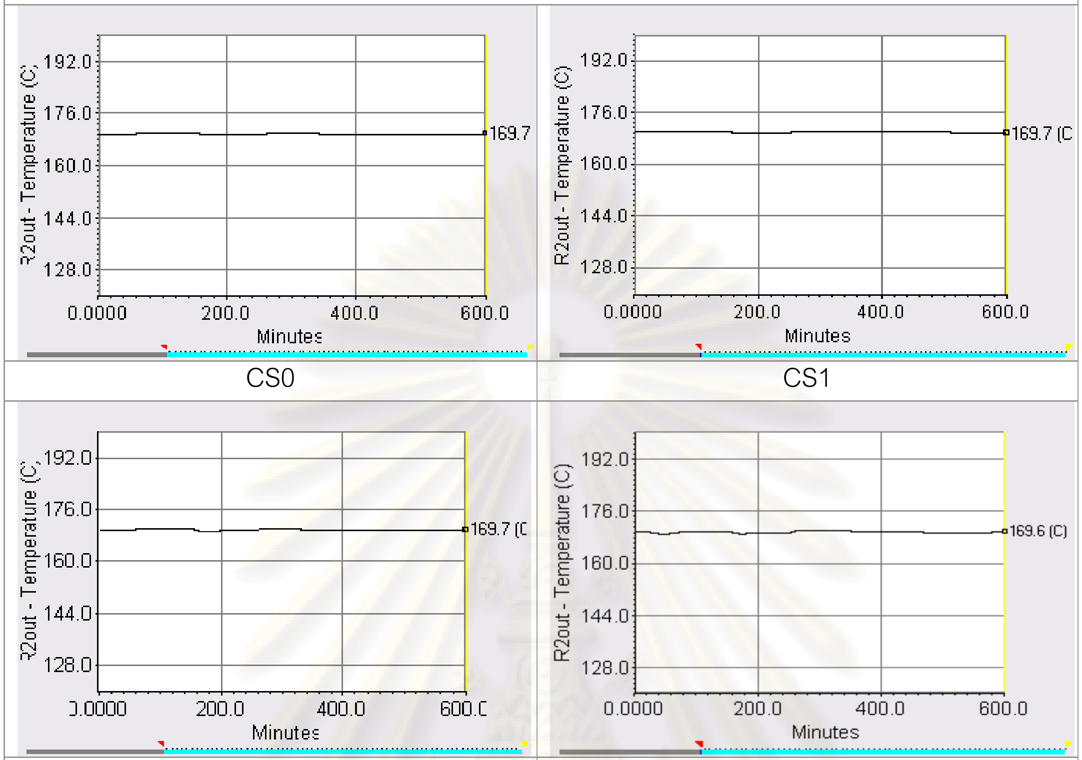
CS2

CS3

(A3)

Figure 5.10 (Continued) Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Effluent temperature reactor 2 (R2)



CS0 CS1

CS2 CS3

(A4)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

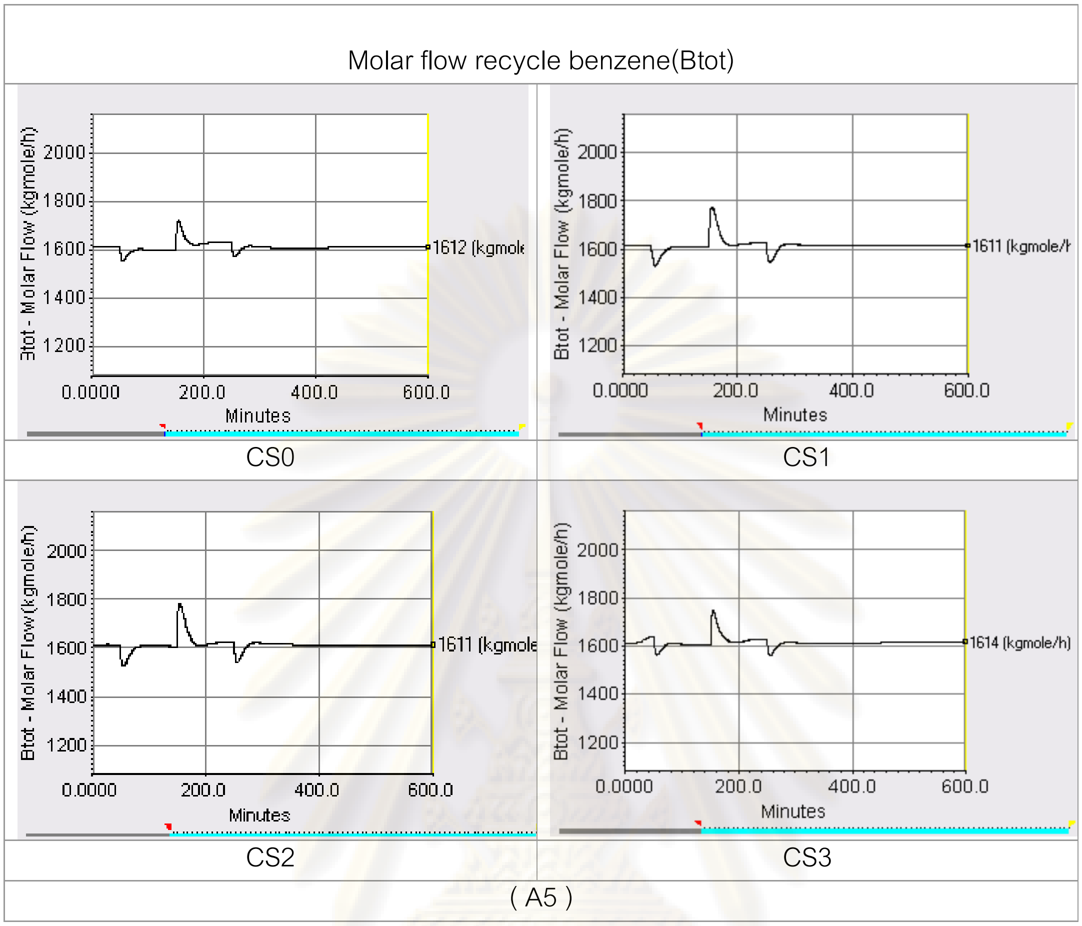
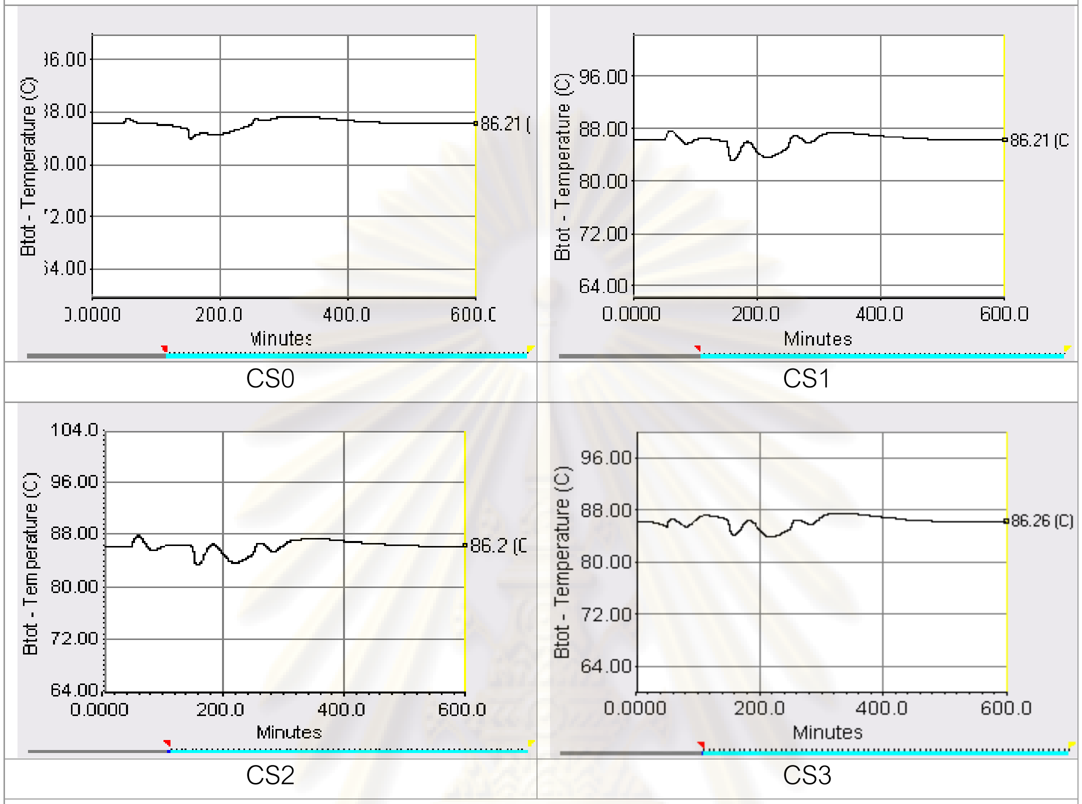


Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Temperature recycle benzene(Btot)



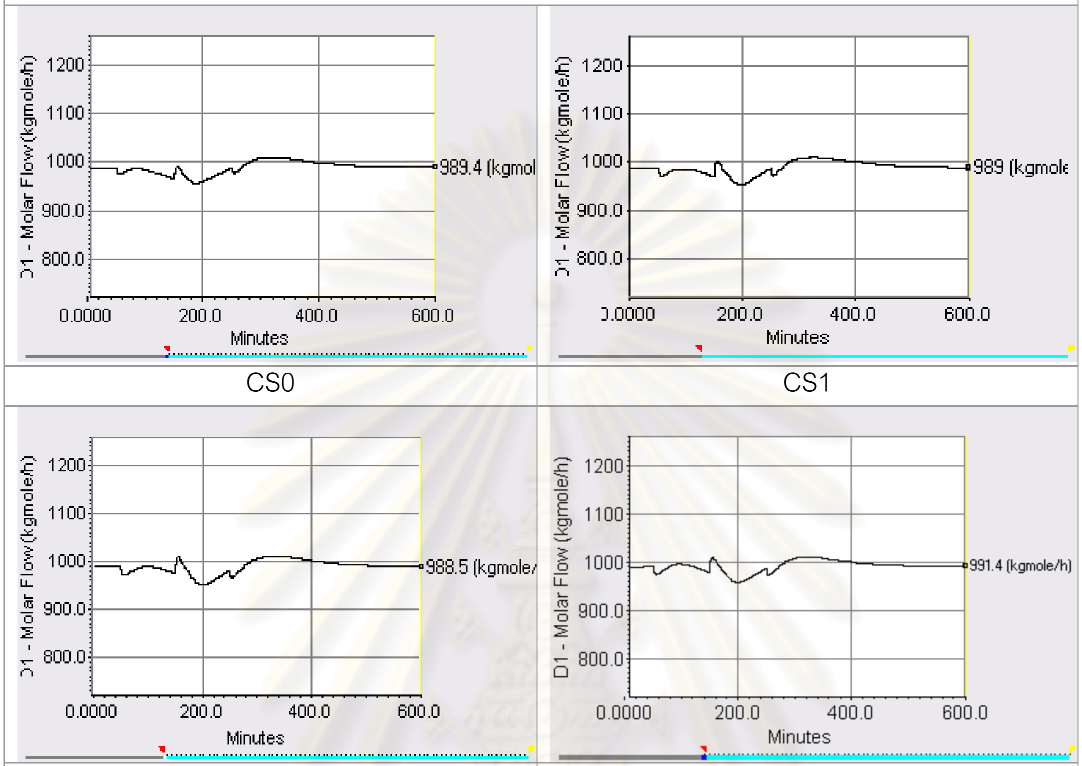
CS0 CS1

CS2 CS3

(A6)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Molar flow distillate benzene column(D1)



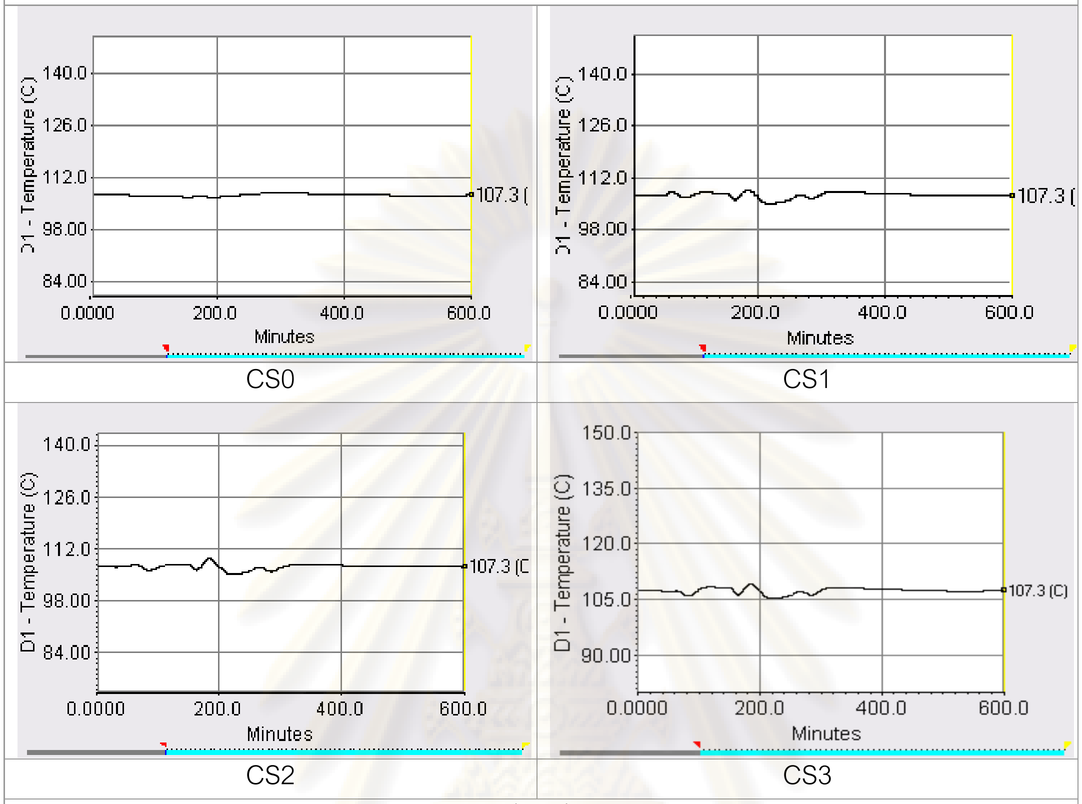
CS0 CS1

CS2 CS3

(A7)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Temperature distillate benzene column(D1)



CS0

CS1

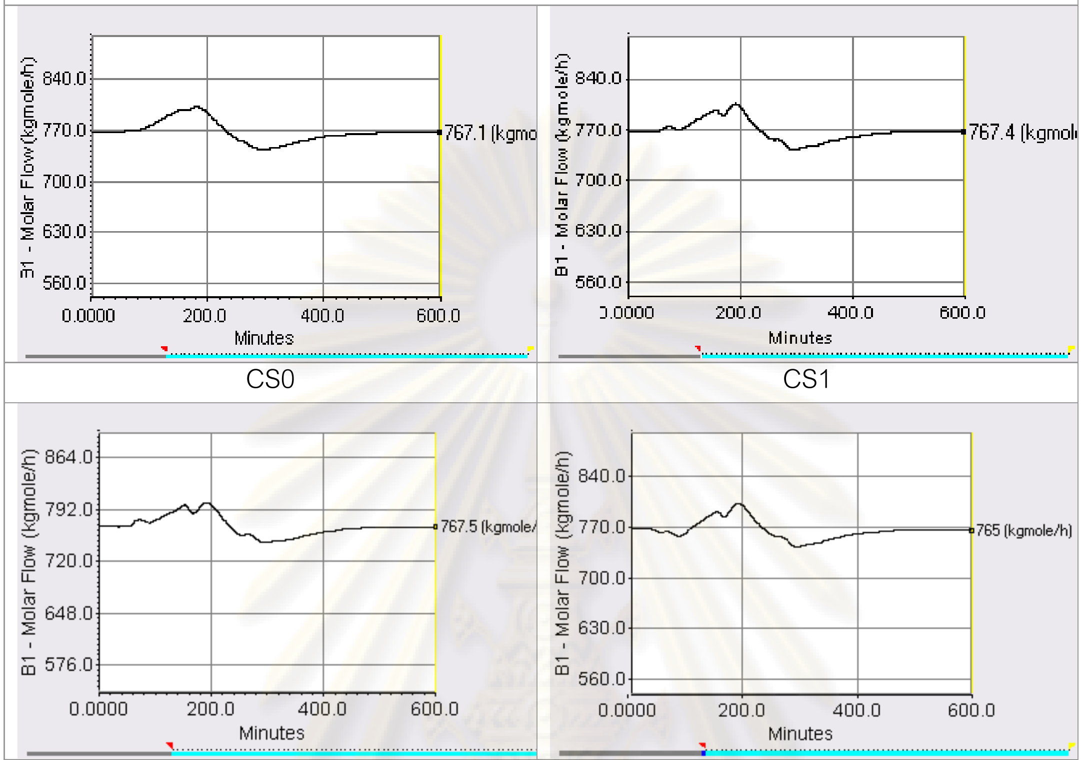
CS2

CS3

(A8)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Molar flow bottom benzene column(B1)



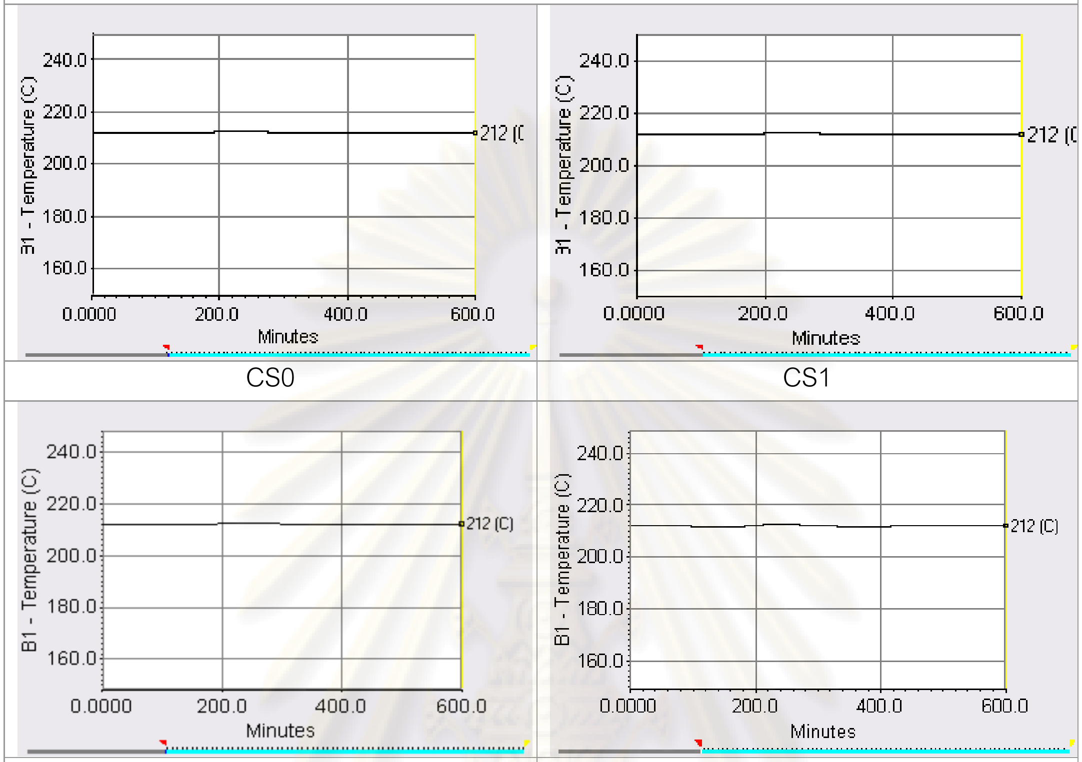
CS0 CS1

CS2 CS3

(A9)

Figure 5.10 (Continued) Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Temperature bottom benzene column(B1)



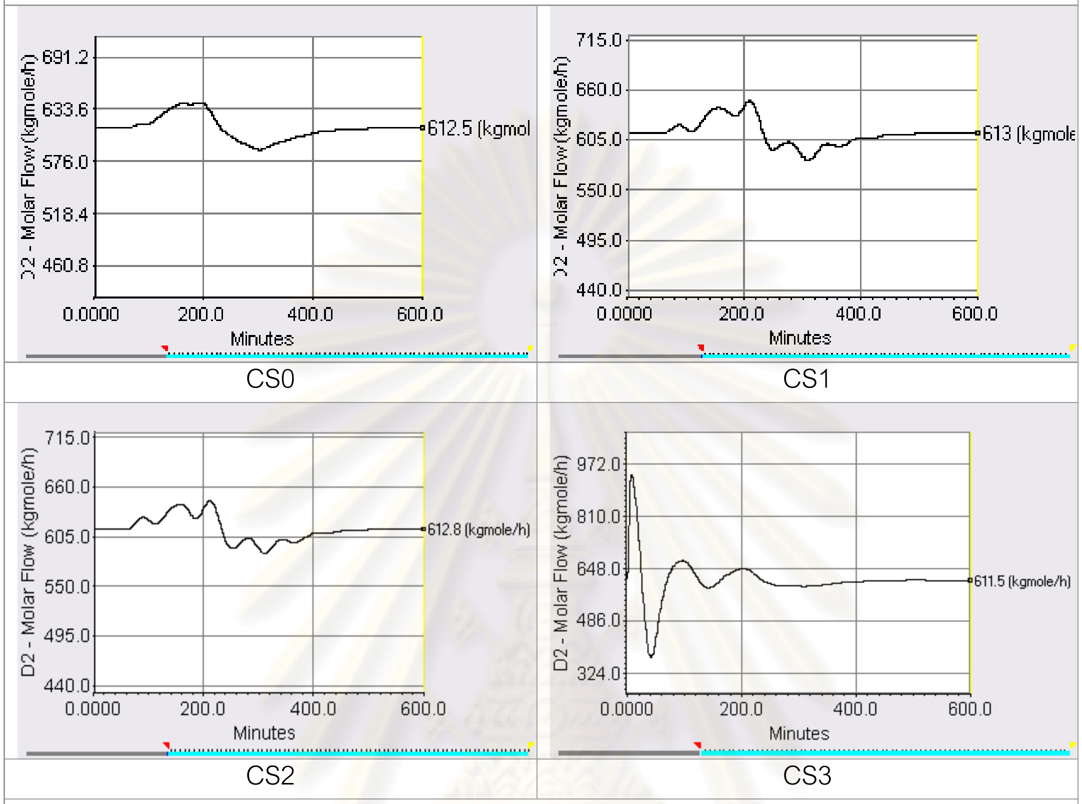
CS0 CS1

CS2 CS3

(A 10)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Molar flow distillate ethyl benzene column(D2)

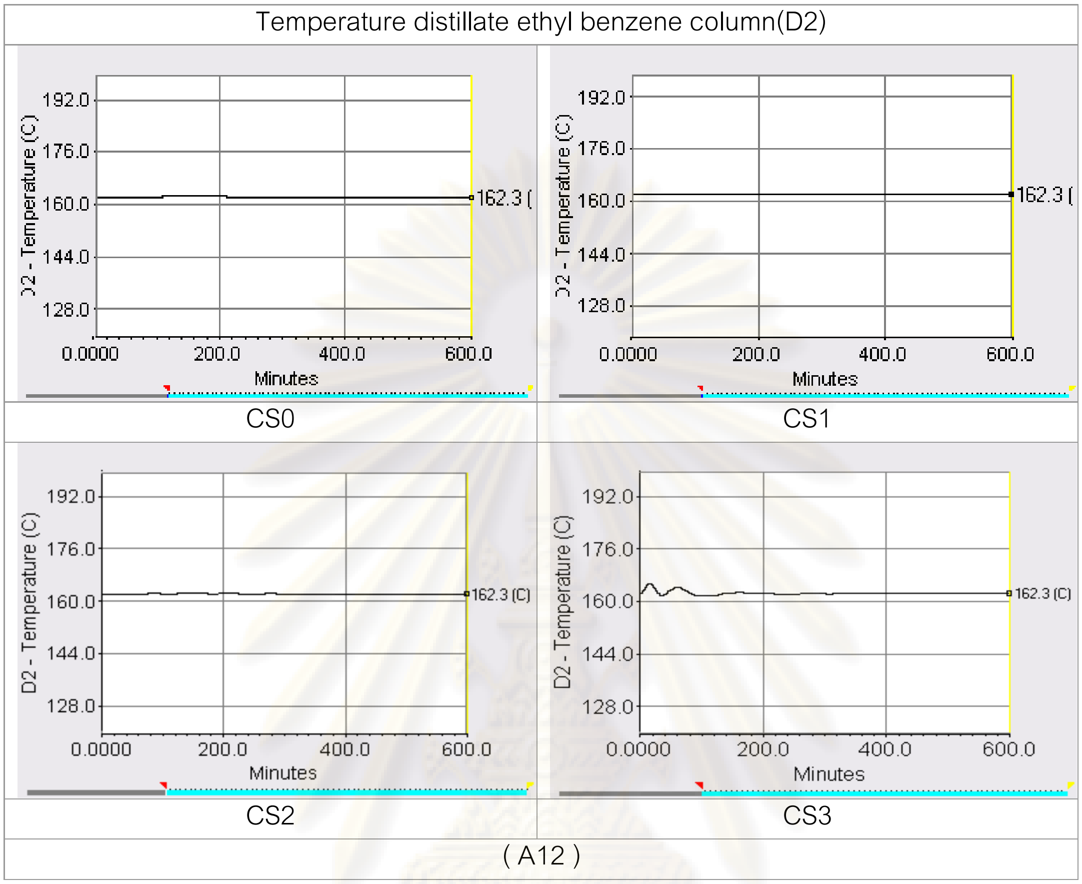


CS0 CS1

CS2 CS3

(A11)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene



Temperature distillate ethyl benzene column(D2)

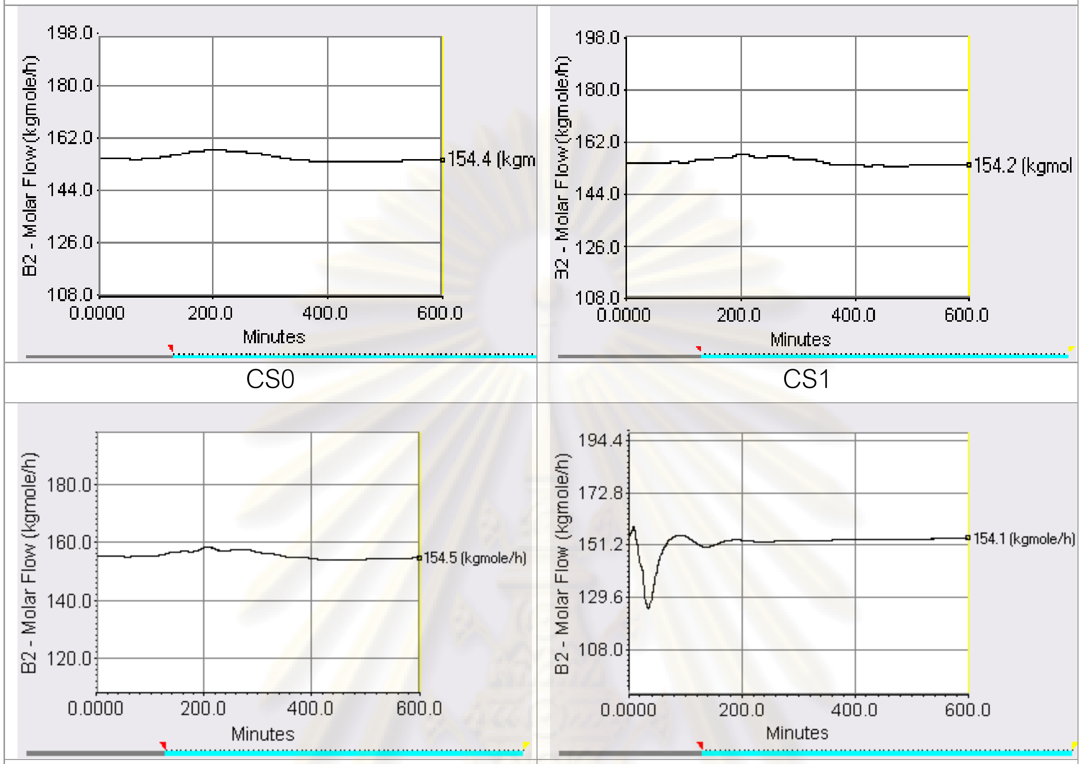
CS0 CS1

CS2 CS3

(A12)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Molar flow bottom ethyl benzene column(B2)



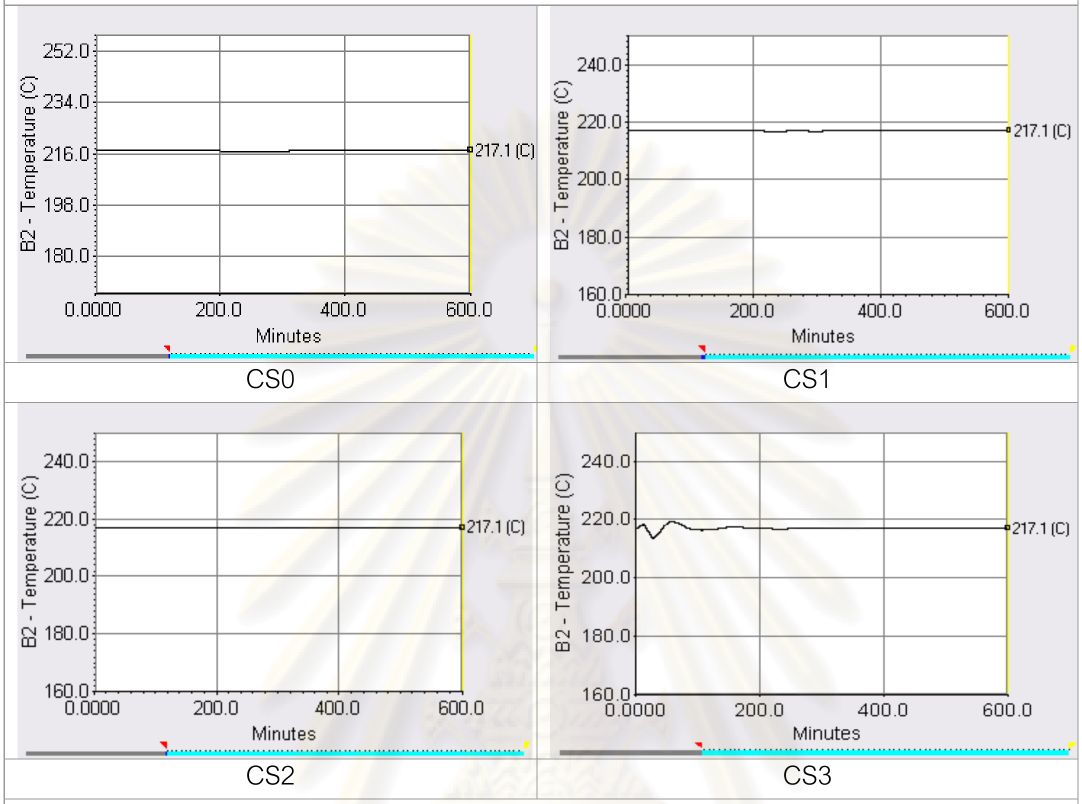
CS0 CS1

CS2 CS3

(A13)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of FFE fresh feed ethylene

Temperature bottom ethyl benzene column(B2)



CS0 CS1

CS2 CS3

(A14)

Figure 5.10 (Continued)Dynamic response of control structure (CS0, CS1,CS2 and CS3) by step change of molar flow rate±10% of FFE fresh feed ethylene

5.3.2 Change in fresh feed temperature by step change of temperature ±10% of FFE fresh feed ethylene for all control structures (base case control structure (CS0), design control structure I (CS1) to designed control structure III (CS3)

Dynamic response change in fresh feed temperature by step change of temperature ±10% of FFE fresh feed ethylene from 51.69 °C to 56.86 °C at 50 min to 150 min and from 56.86 °C to 46.52 °C at 150 min to 250 min

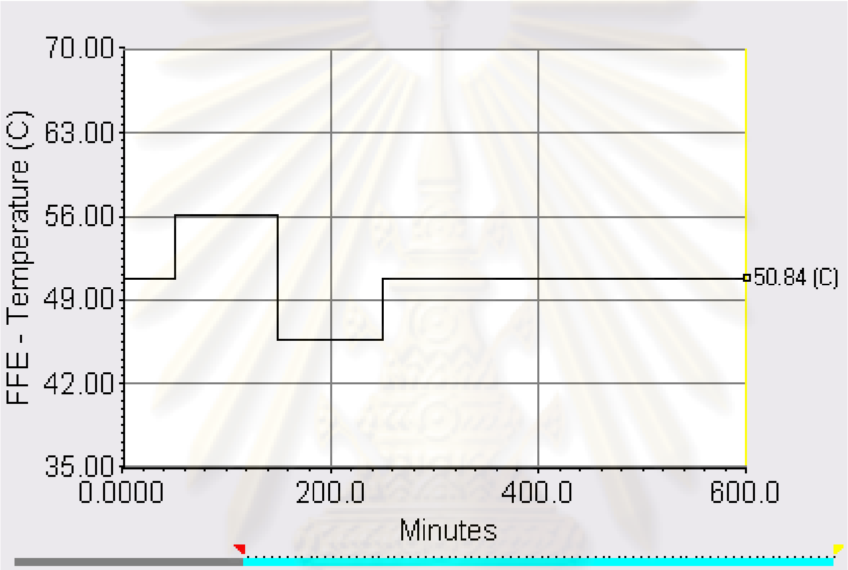
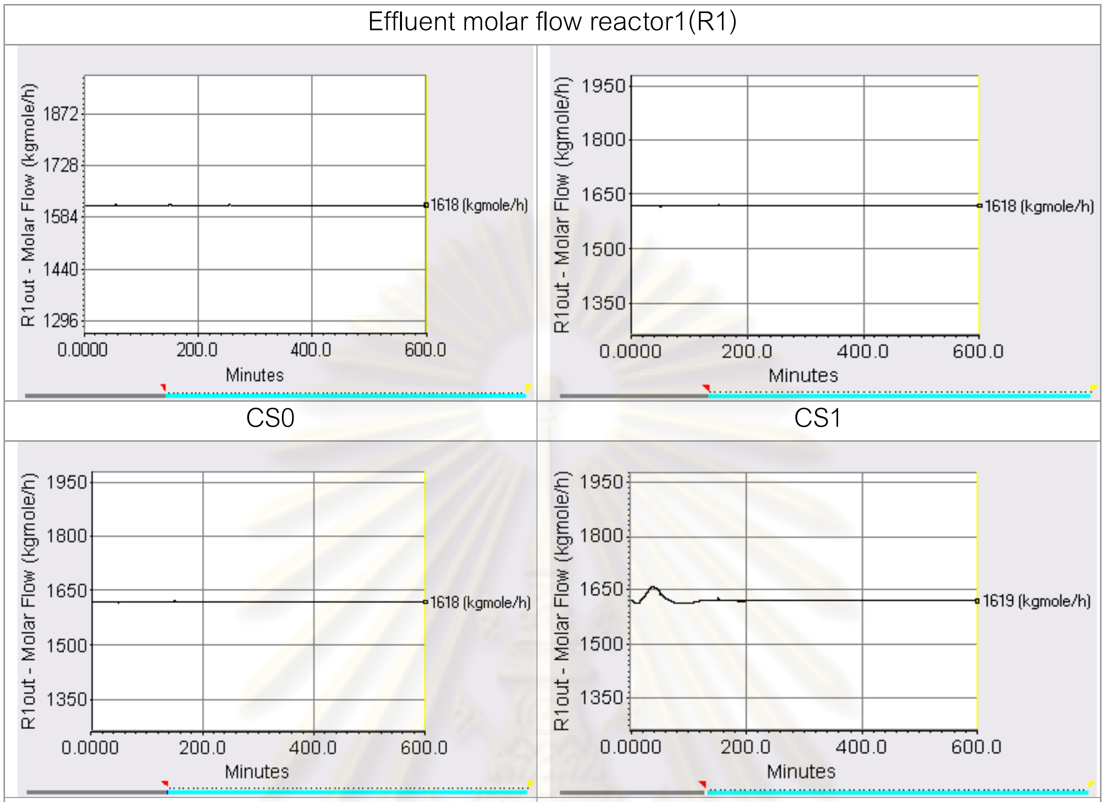


Figure 5.11 Dynamic responses of ethyl benzene process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Effluent molar flow reactor1(R1)



CS0 CS1

CS2 CS3

(B1)

Figure 5.12 Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

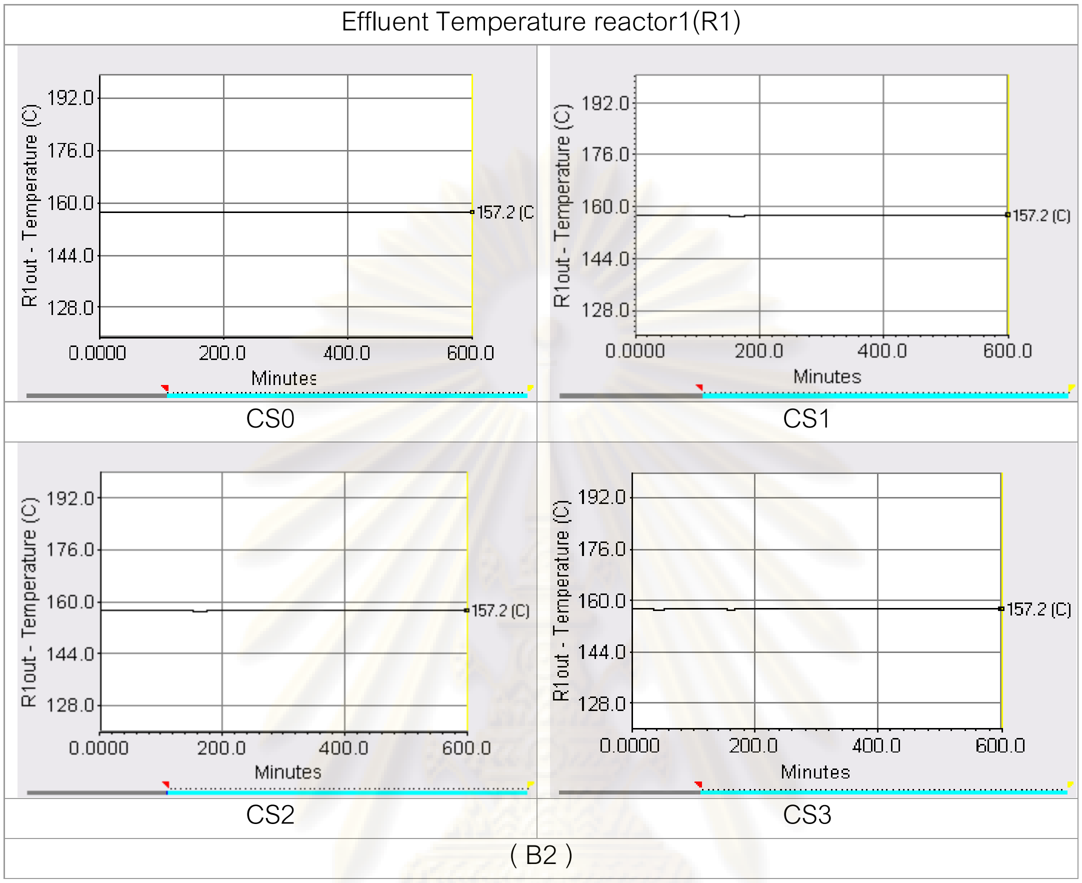
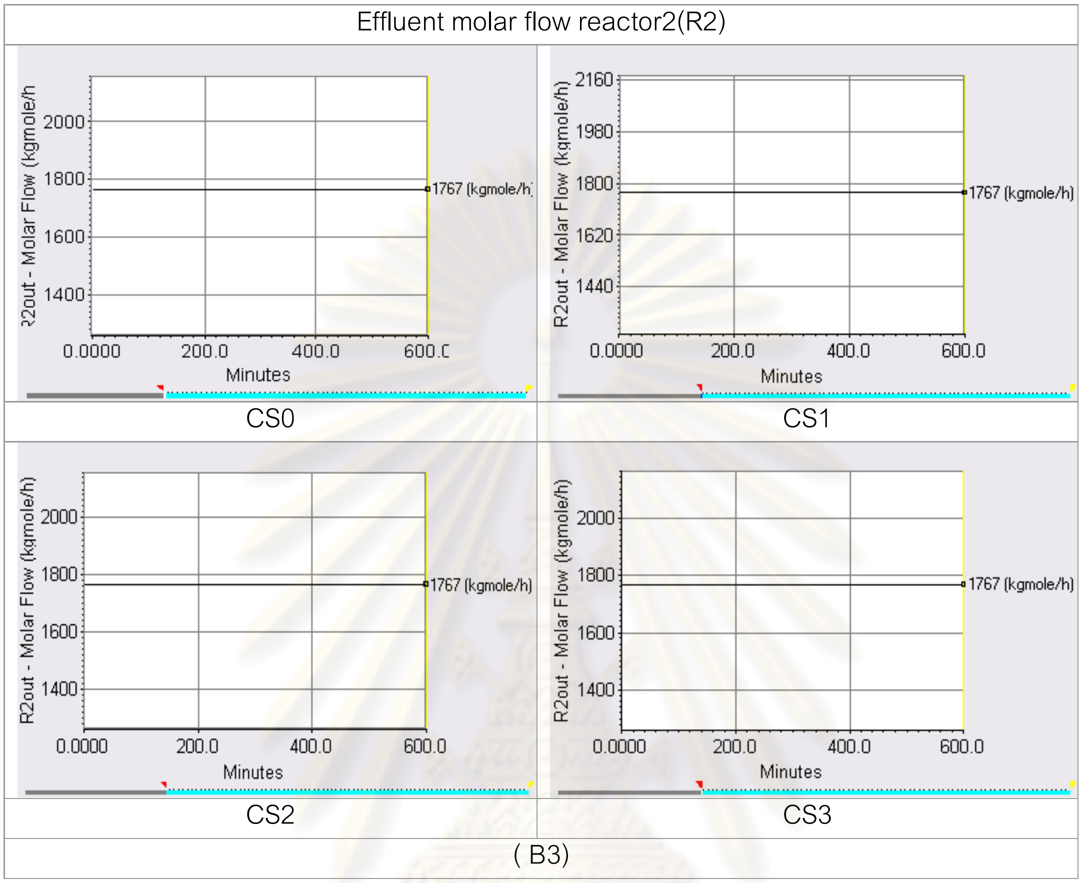


Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature



Effluent molar flow reactor2(R2)

CS0

CS1

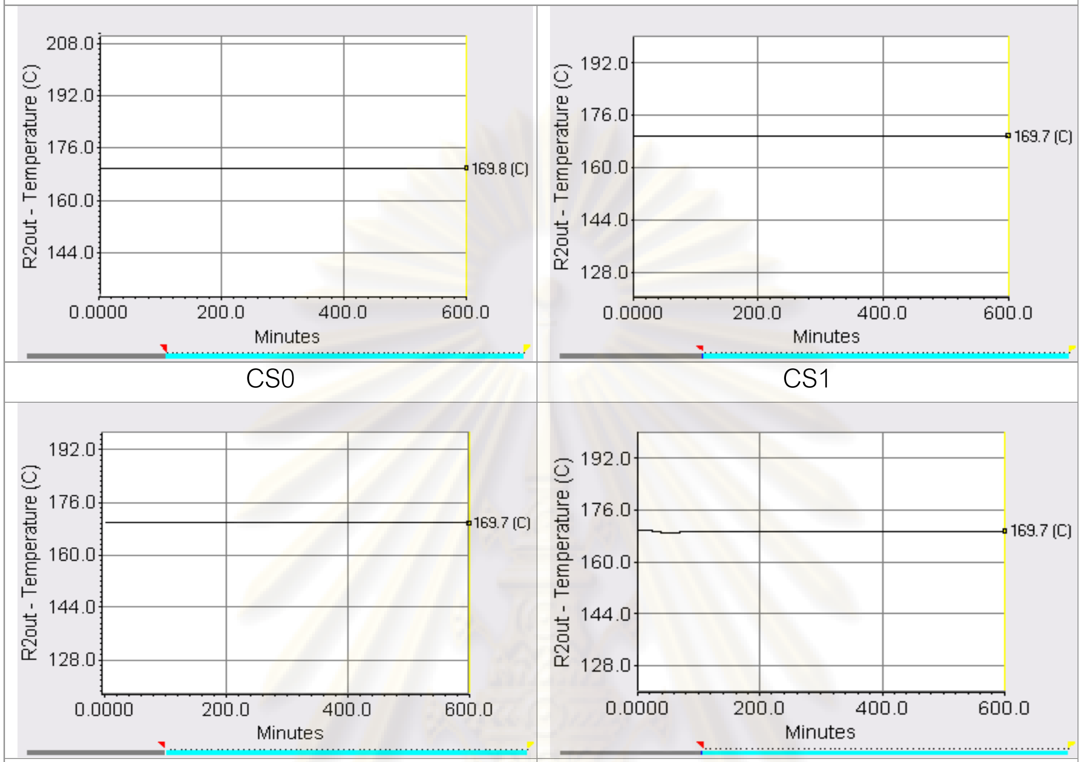
CS2

CS3

(B3)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Effluent temperature reactor2(R2)

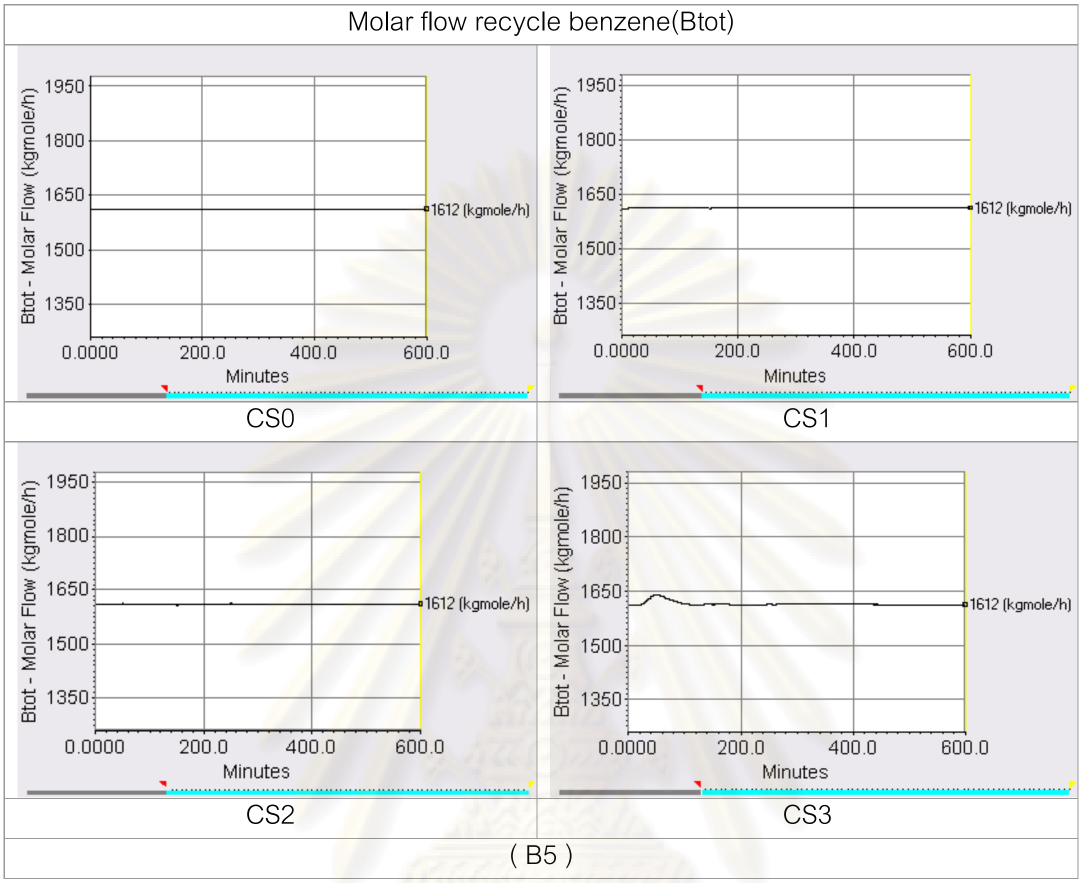


CS0 CS1

CS2 CS3

(B4)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature



Molar flow recycle benzene(Btot)

CS0

CS1

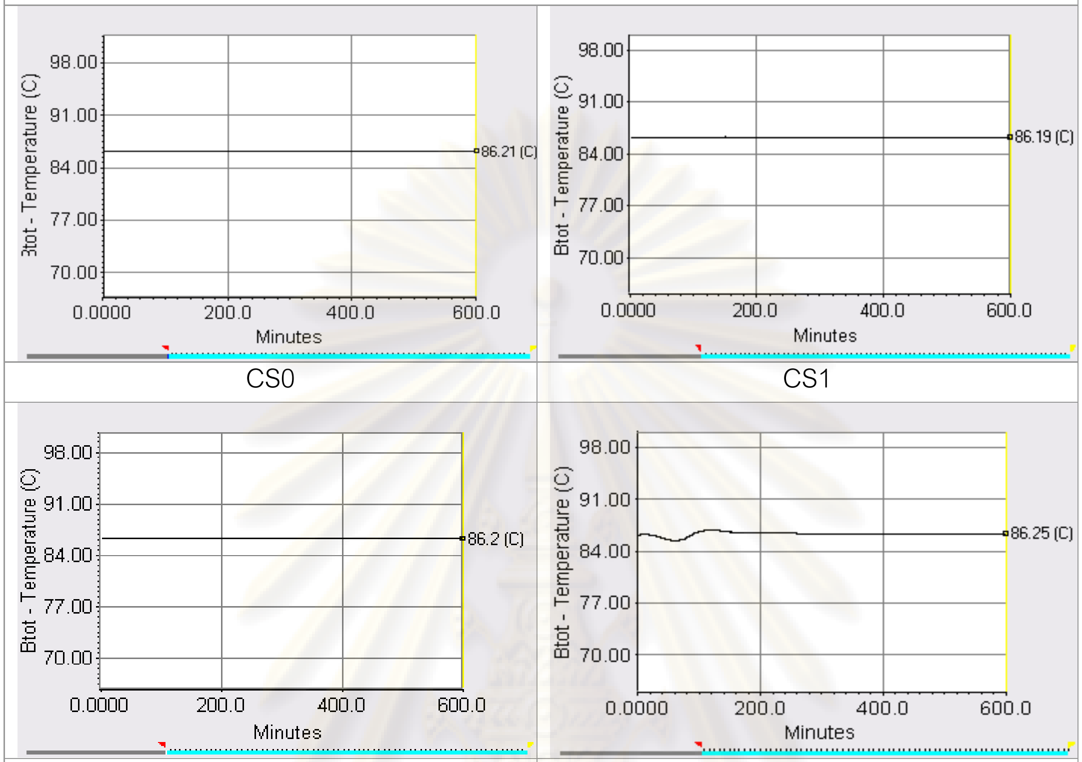
CS2

CS3

(B5)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Temperature recycle benzene(Btot)



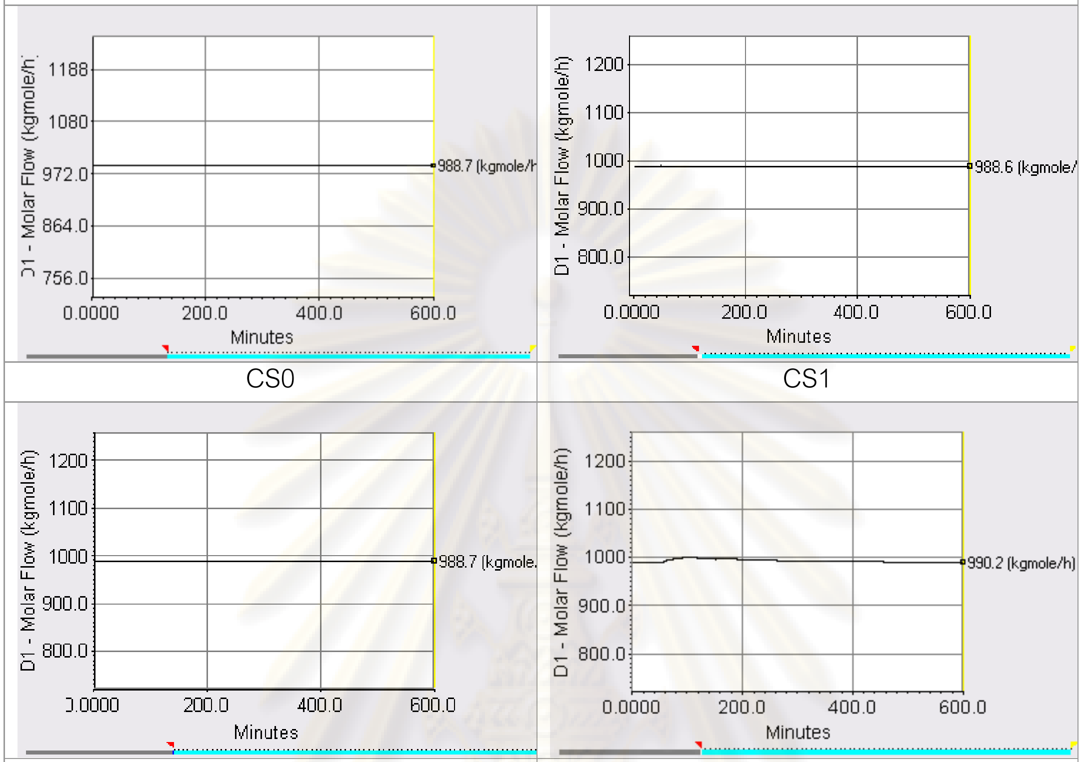
CS0 CS1

CS2 CS3

(B6)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Molar flow distillate benzene column(D1)



CS0 CS1

CS2 CS3

(B7)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

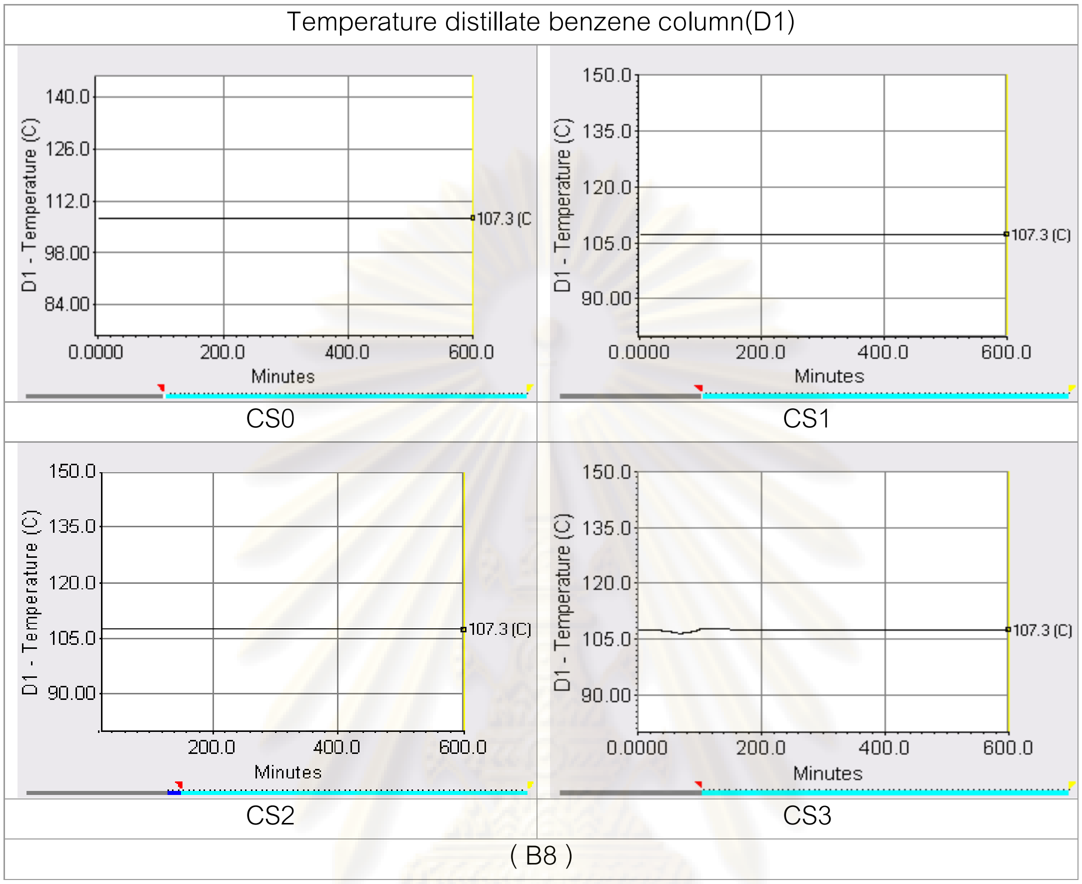
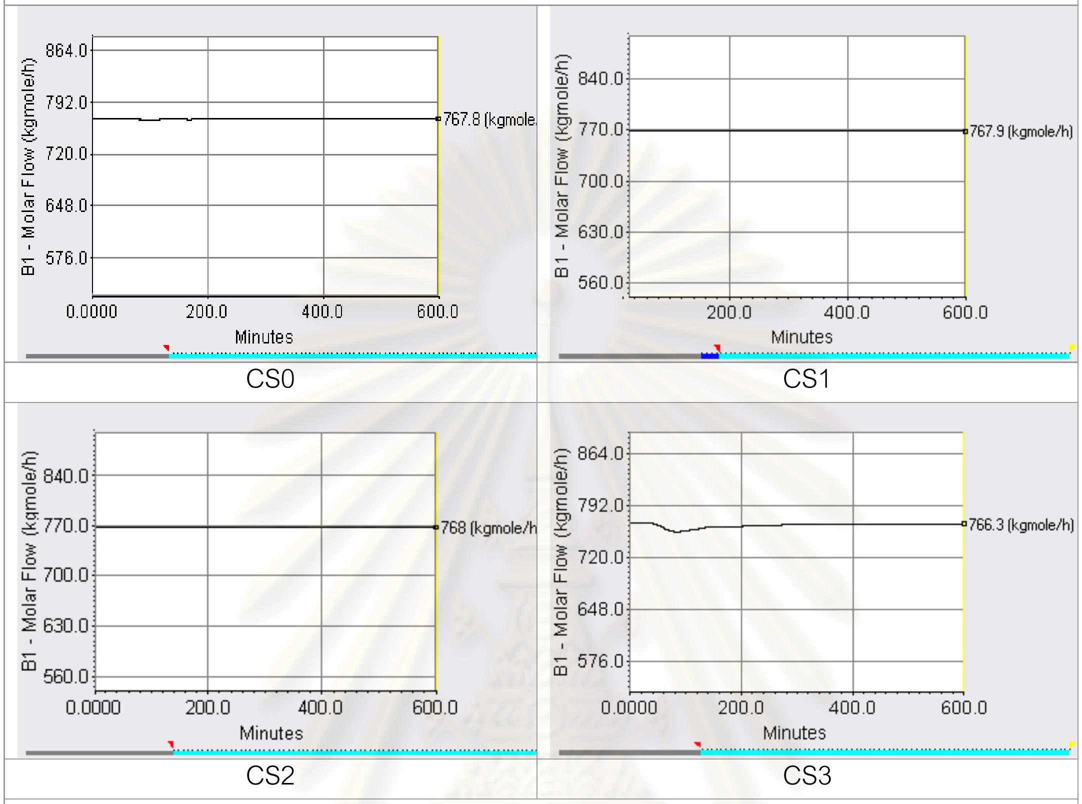


Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Molar flow bottom benzene column(B1)

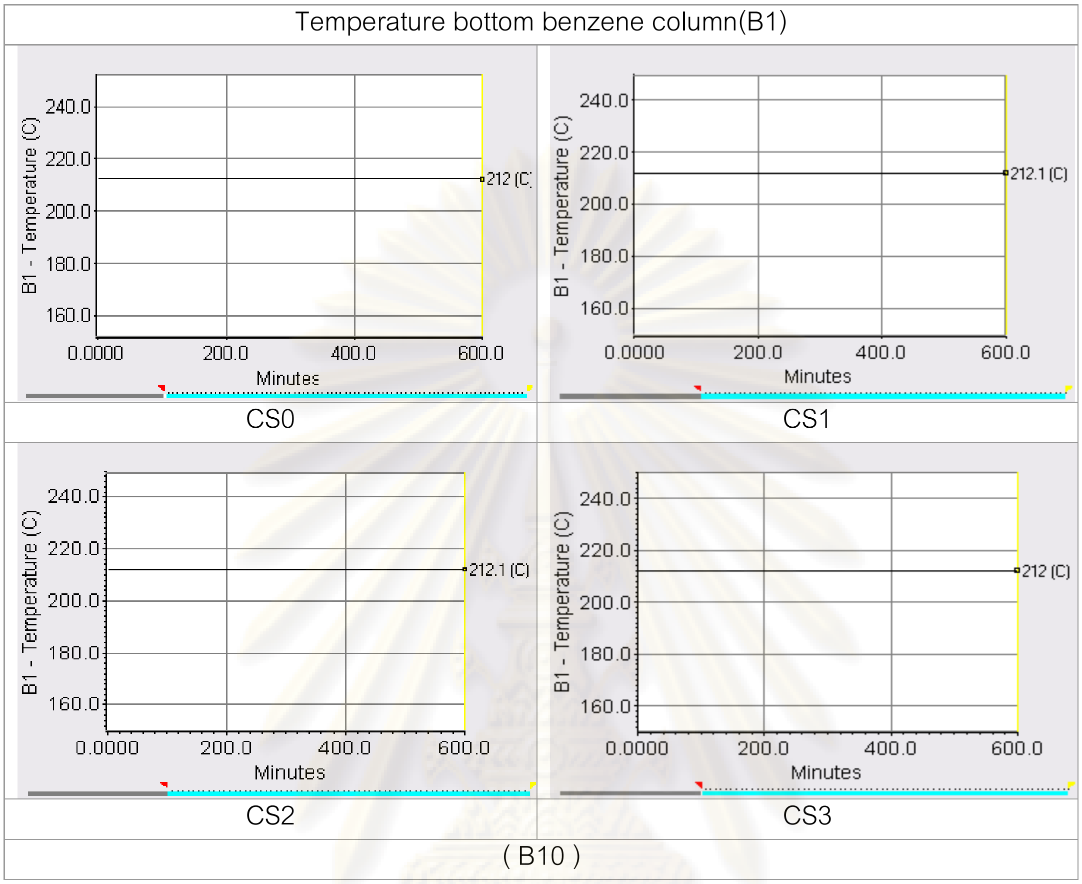


CS0 CS1

CS2 CS3

(B9)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature


Temperature bottom benzene column(B1)

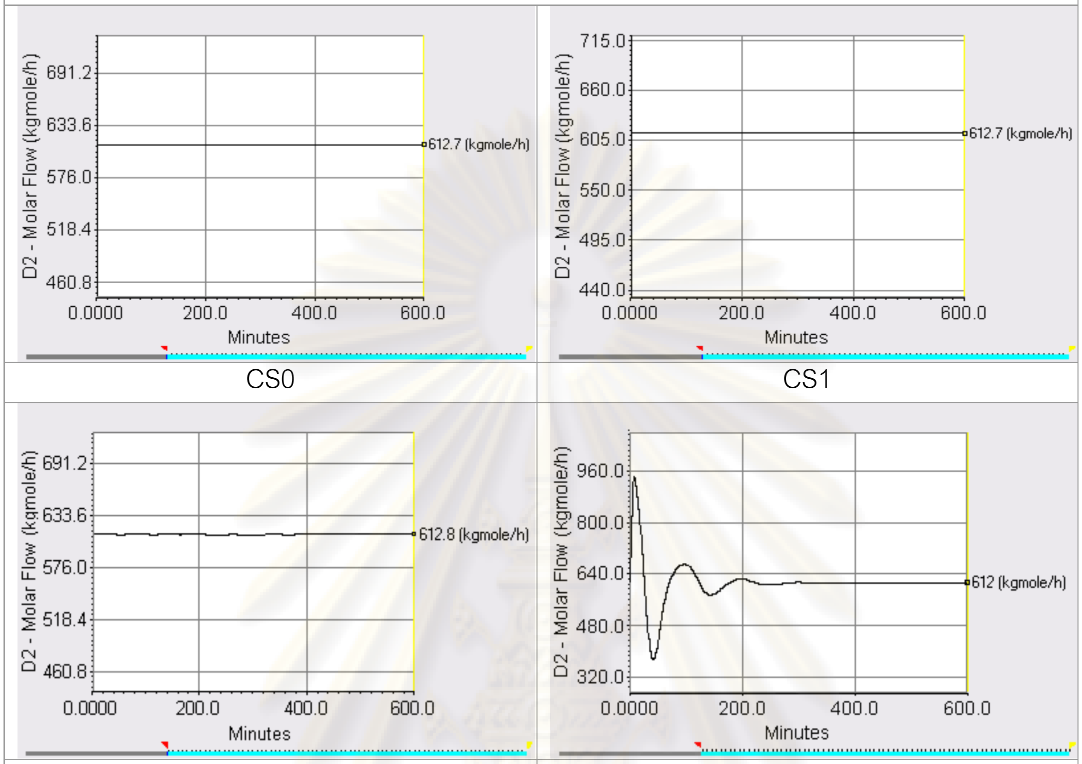
CS0 CS1

CS2 CS3

(B10)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Molar flow distillate ethyl benzene column(D2)



CS0 CS1

CS2 CS3

(B11)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

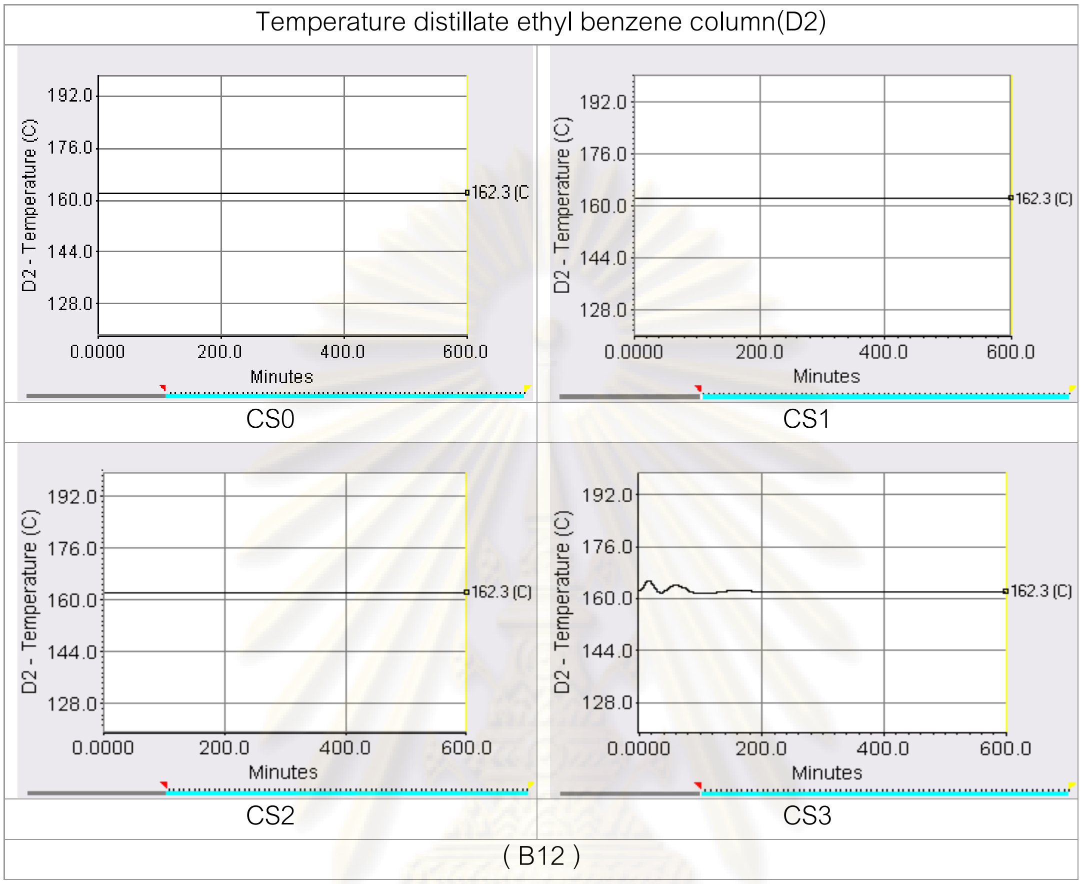
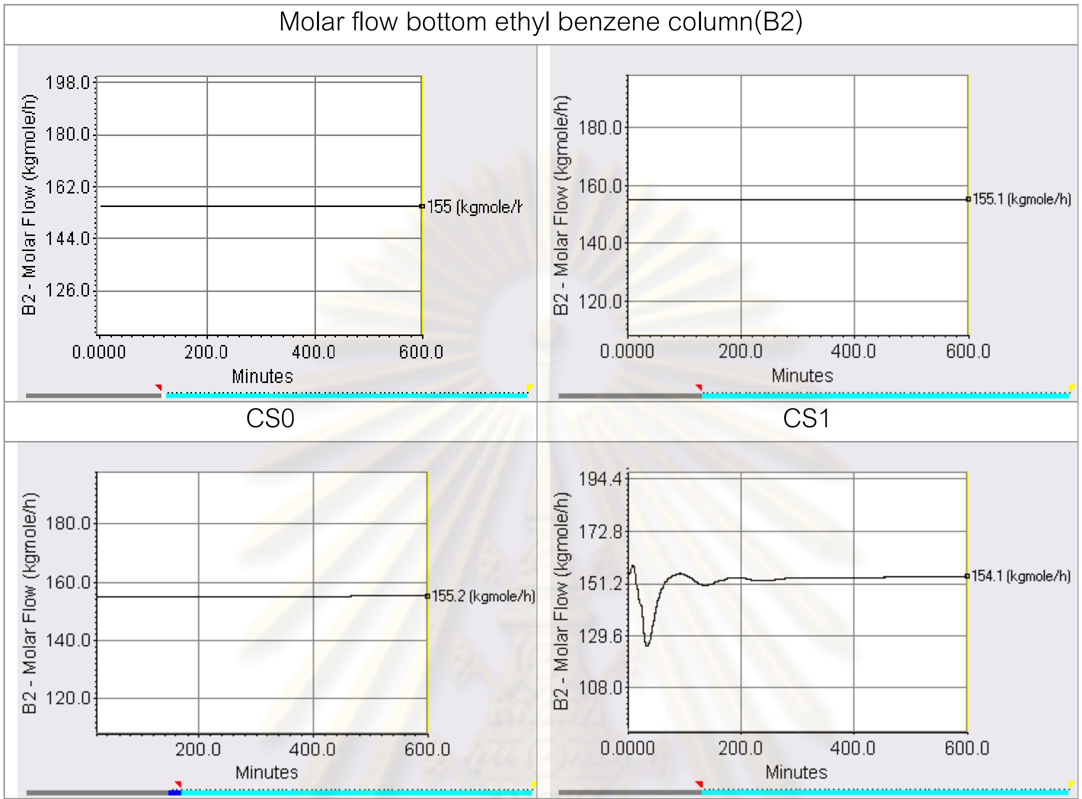


Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

Molar flow bottom ethyl benzene column(B2)



CS0

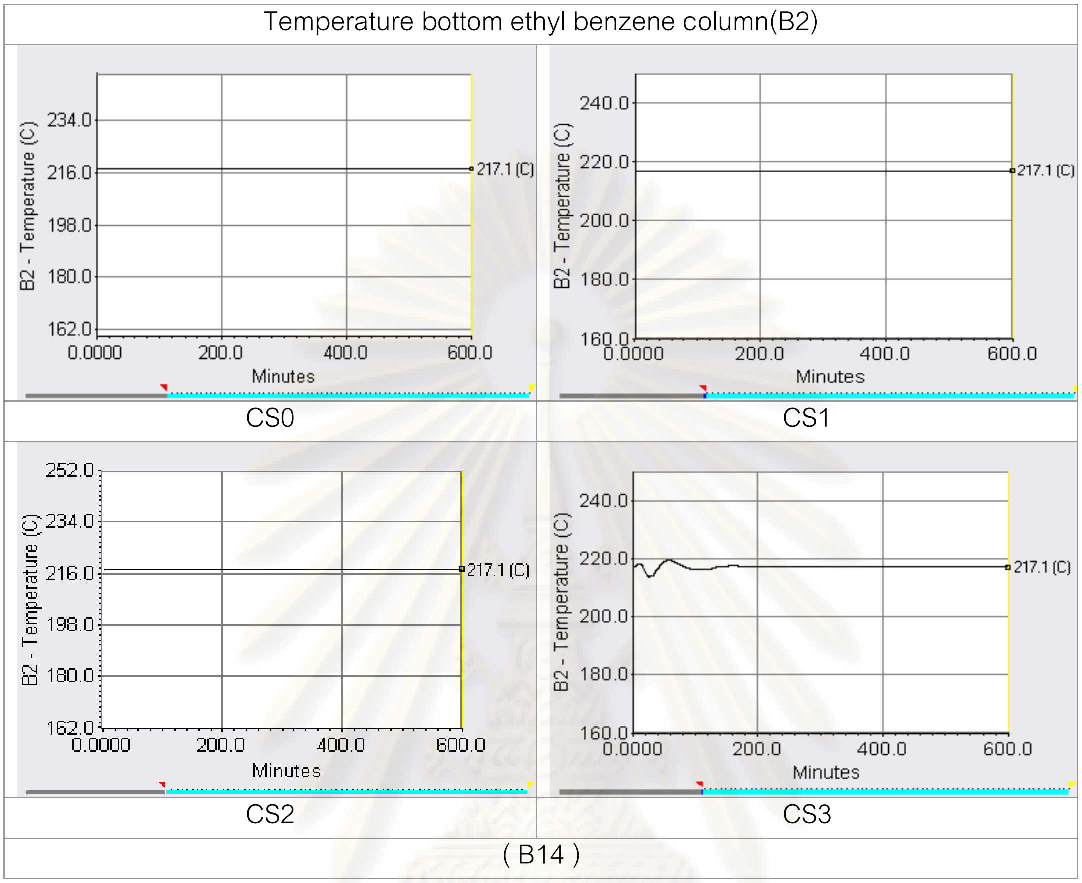
CS1

CS2

CS3

(B13)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) process by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature



Temperature bottom ethyl benzene column(B2)

CS0

CS1

CS2

CS3

(B14)

Figure 5.12(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) by step change of temperature ±10% of FFE fresh feed ethylene in fresh feed temperature

5.3.3 Change in fresh feed temperature by step change of temperature±10% of FFB fresh feed benzene for all control structures (base case control structure (CS0), design control structure I (CS1) to designed control structure III (CS3)

Dynamic response change in fresh feed temperature by step change of temperature±10% of FFB fresh feed benzene from 49.92 °C to 54.91 °C at 50 min to 150 min and from 54.91 °C to 44.93 °C at 150 min to 250 min

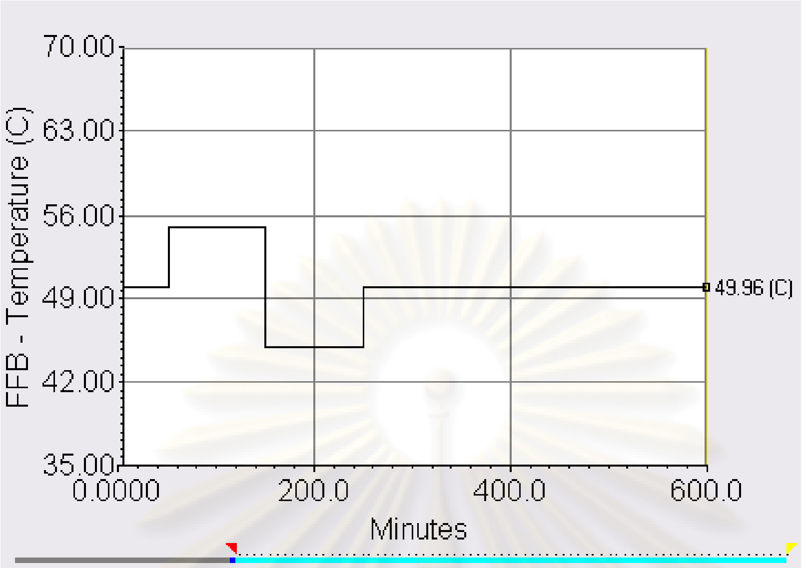
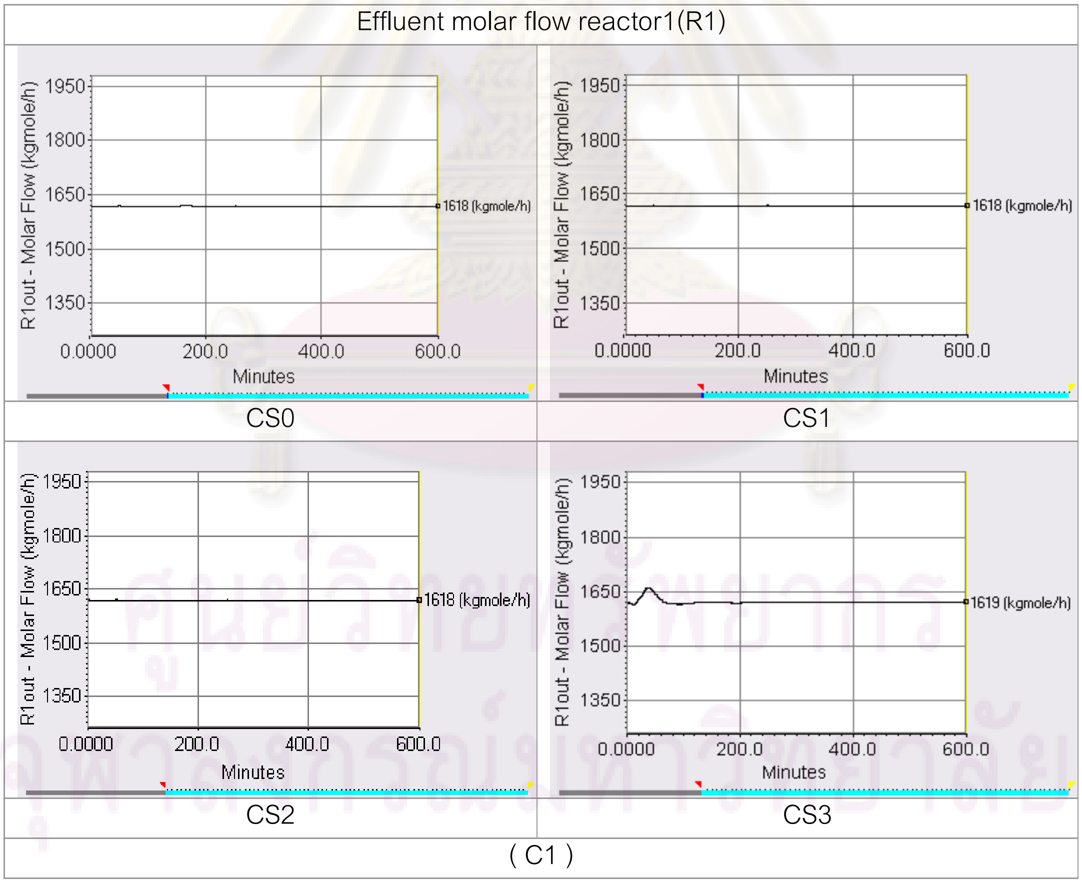


Figure 5.13 Dynamic responses of ethyl benzene process by step change of temperature ±10% of FFB fresh feed benzene



Effluent molar flow reactor1(R1)

CS0

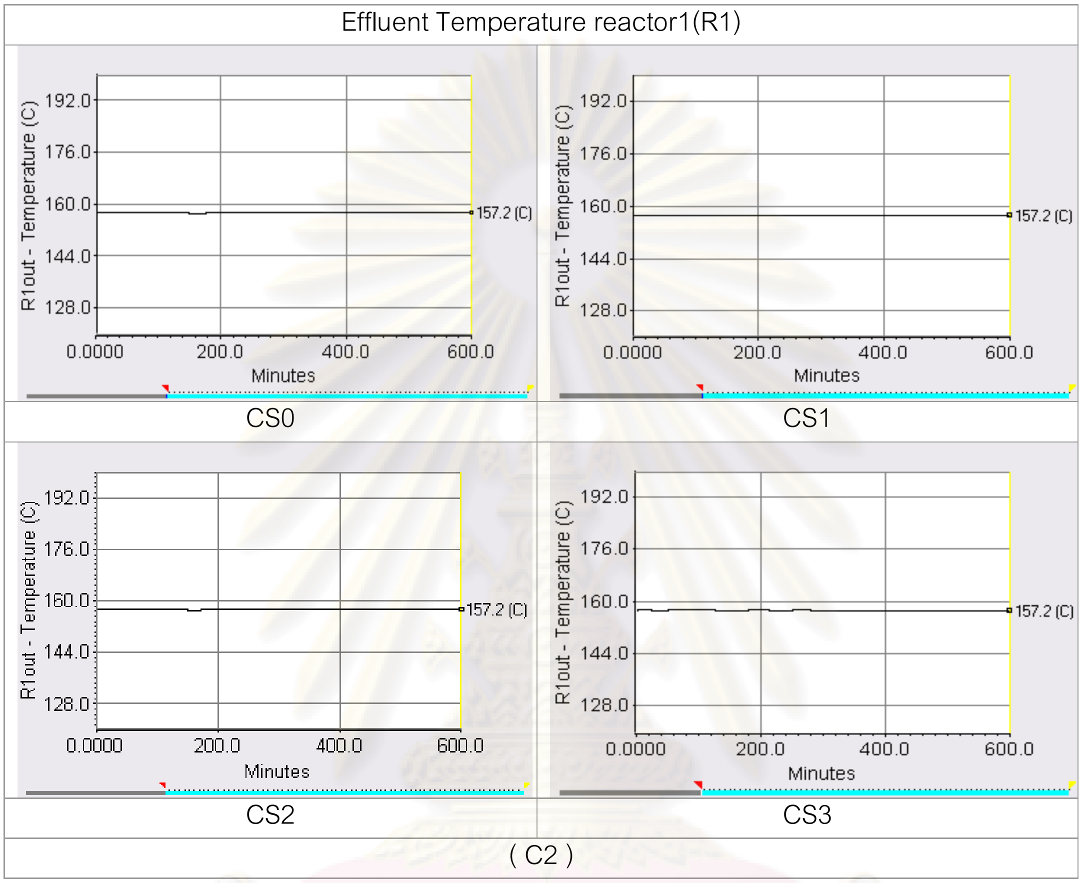
CS1

CS2

CS3

(C1)

Figure 5.14 Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene


Effluent Temperature reactor1(R1)

CS0

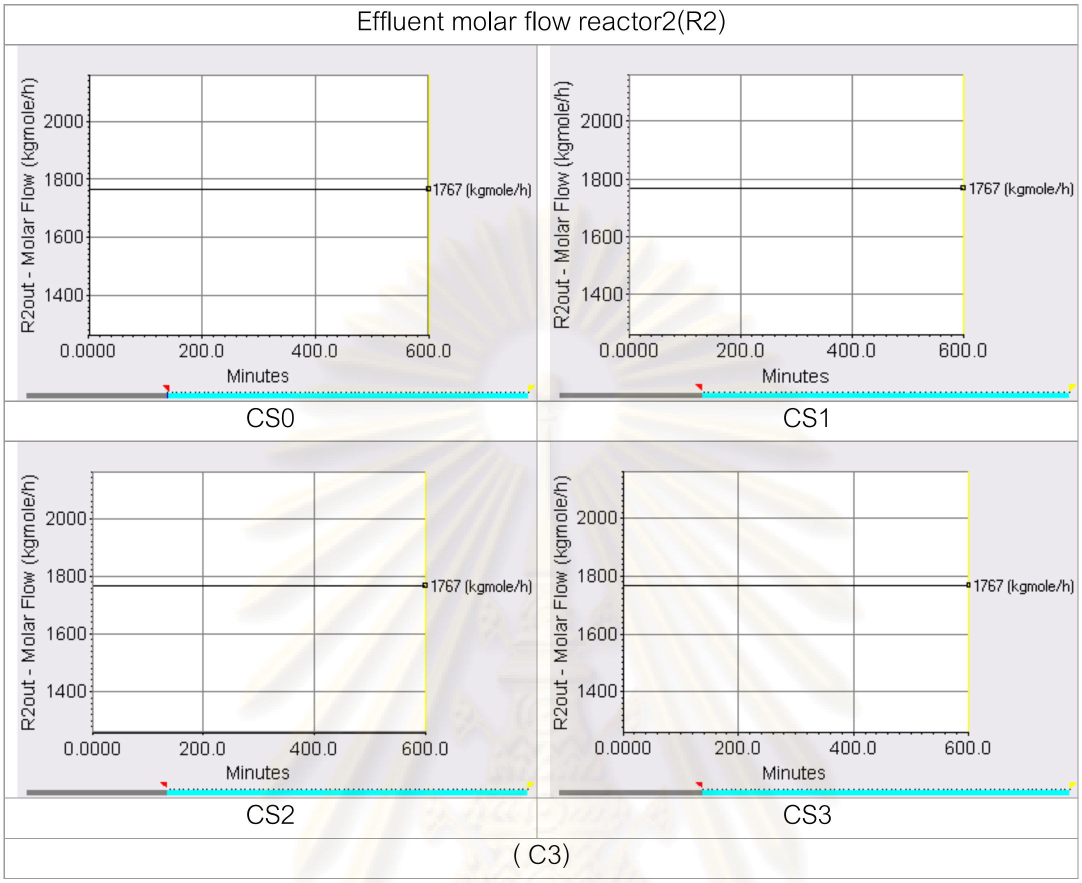
CS1

CS2

CS3

(C2)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene



Effluent molar flow reactor2(R2)

CS0

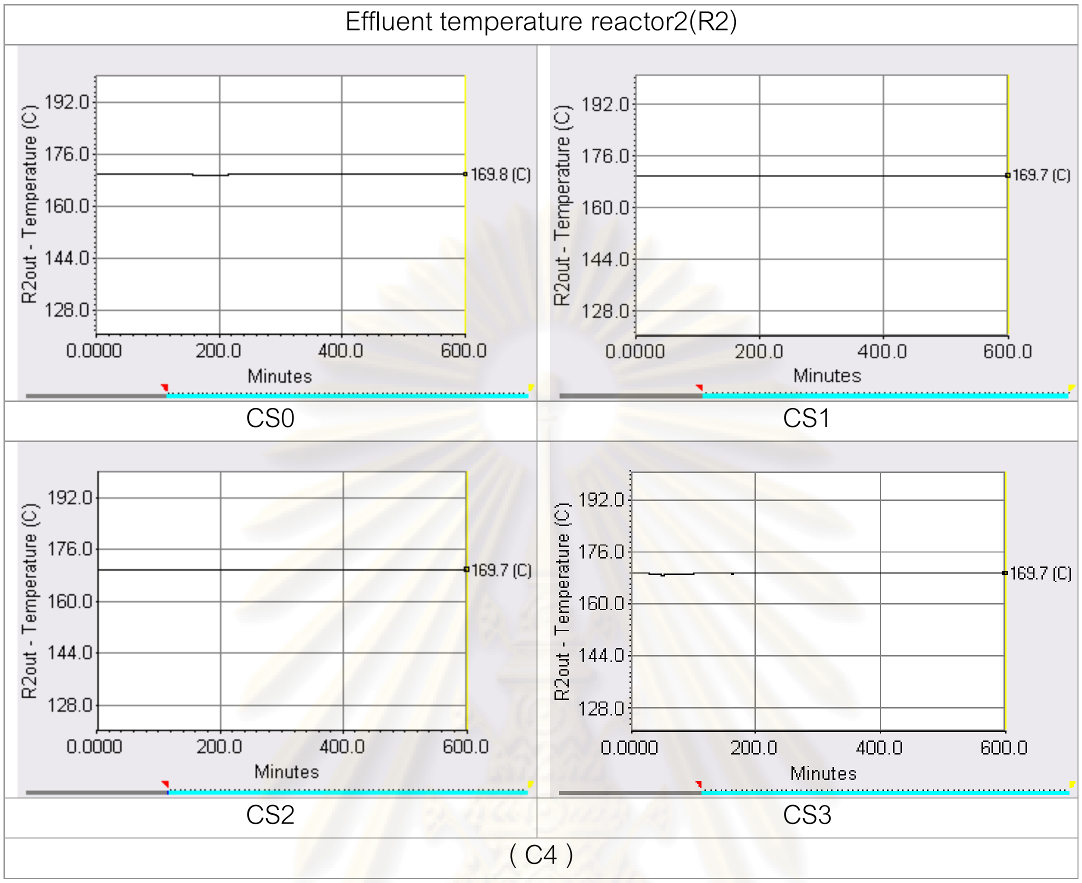
CS1

CS2

CS3

(C3)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene


Effluent temperature reactor2(R2)

CS0

CS1

CS2

CS3

(C4)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

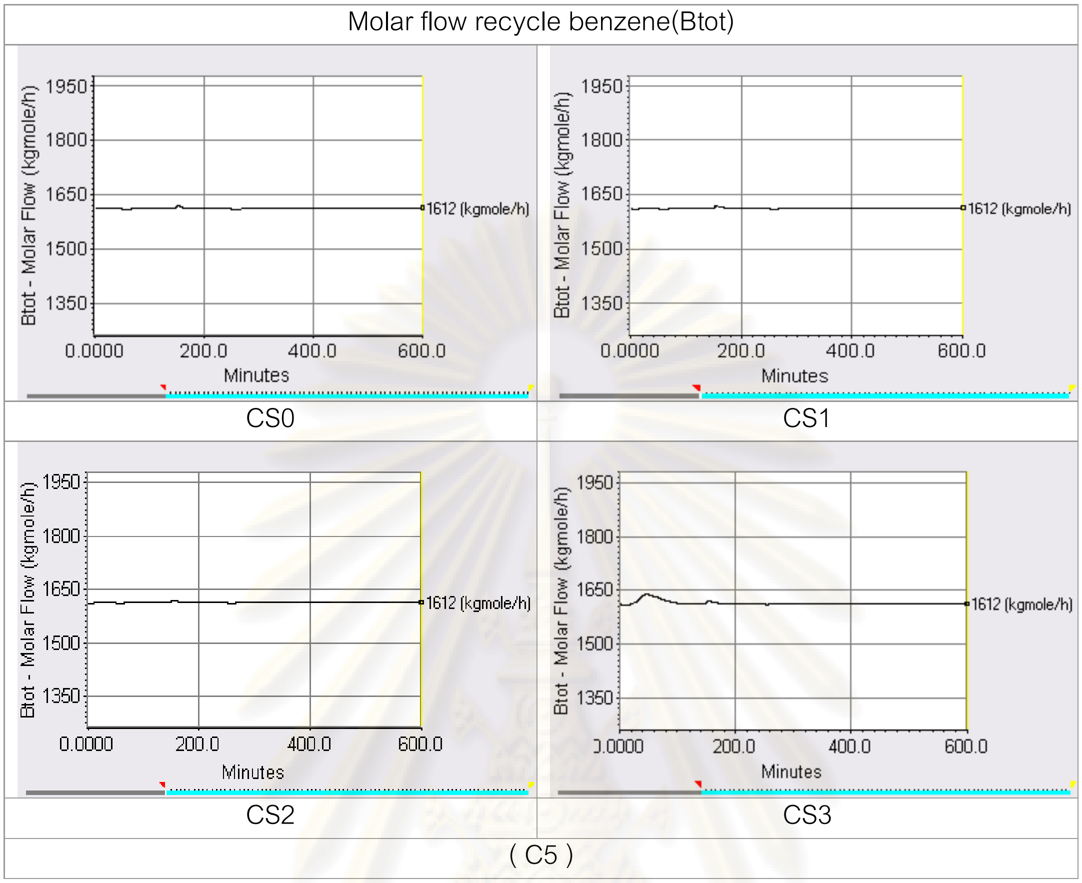
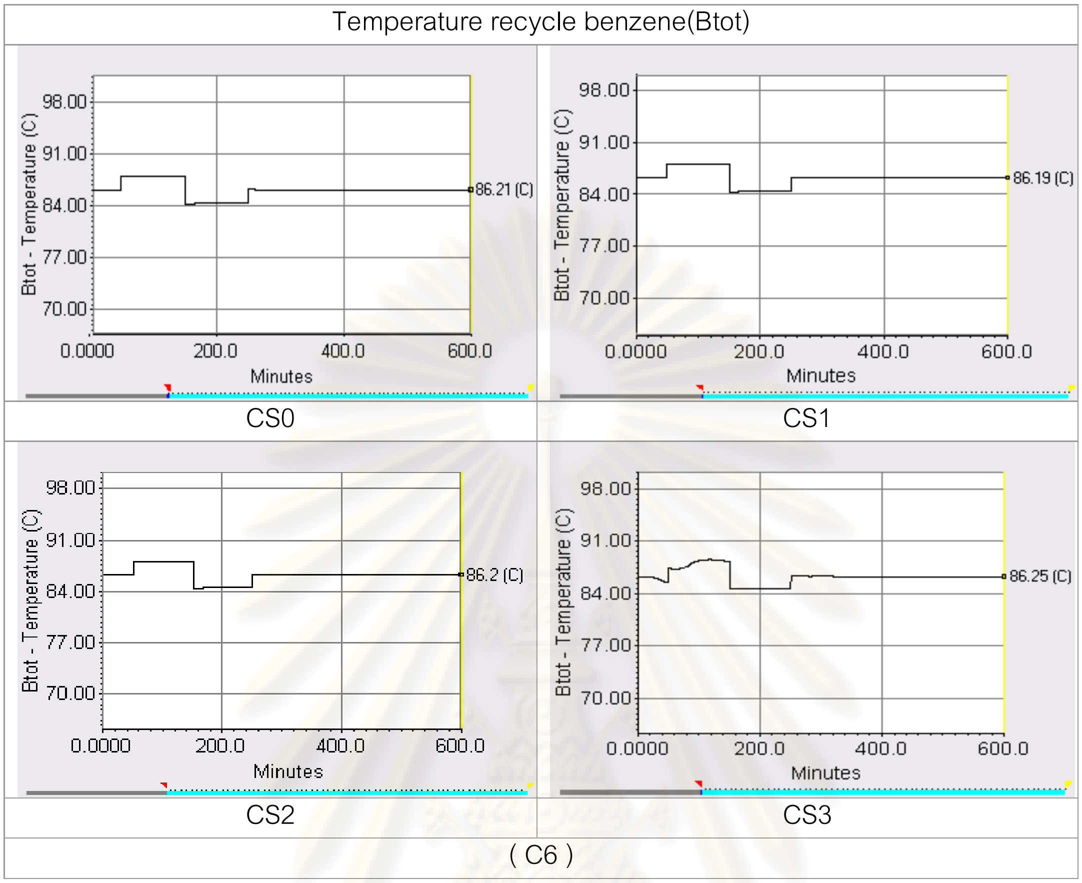


Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

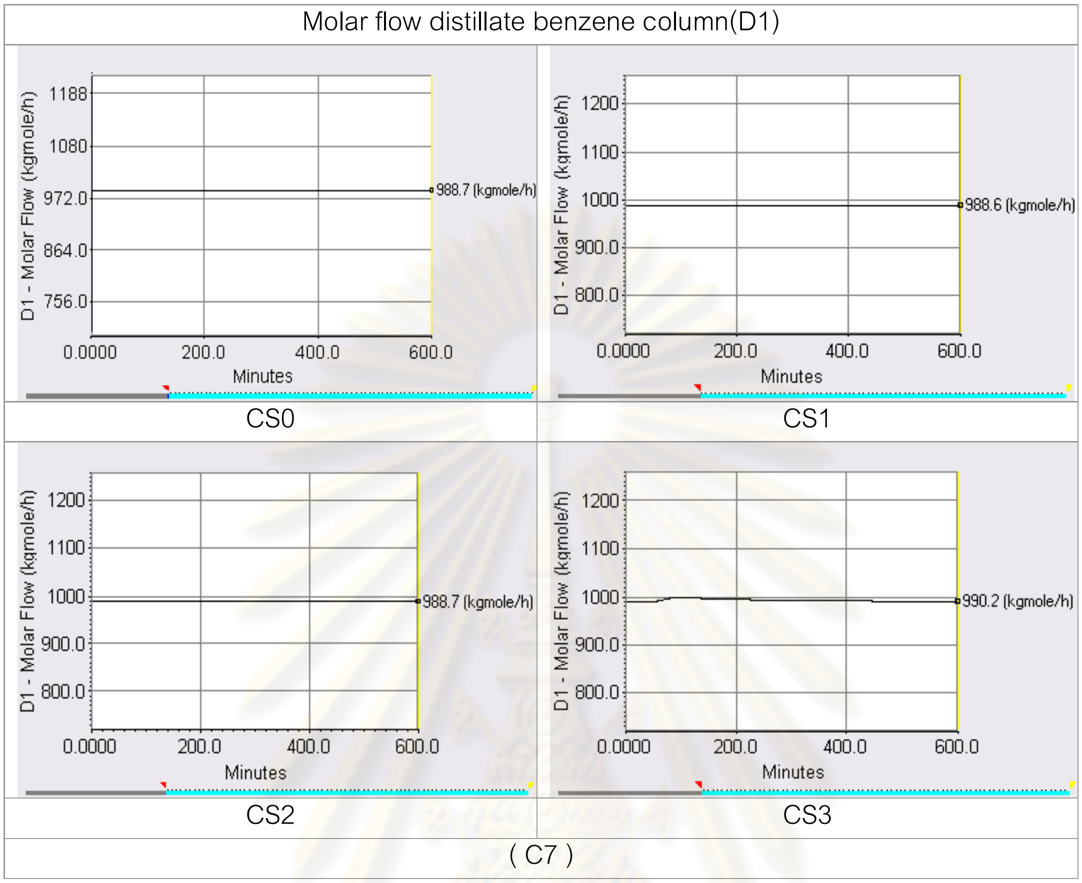


CS0 CS1

CS2 CS3

(C6)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene



Molar flow distillate benzene column(D1)

CS0

CS1

CS2

CS3

(C7)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

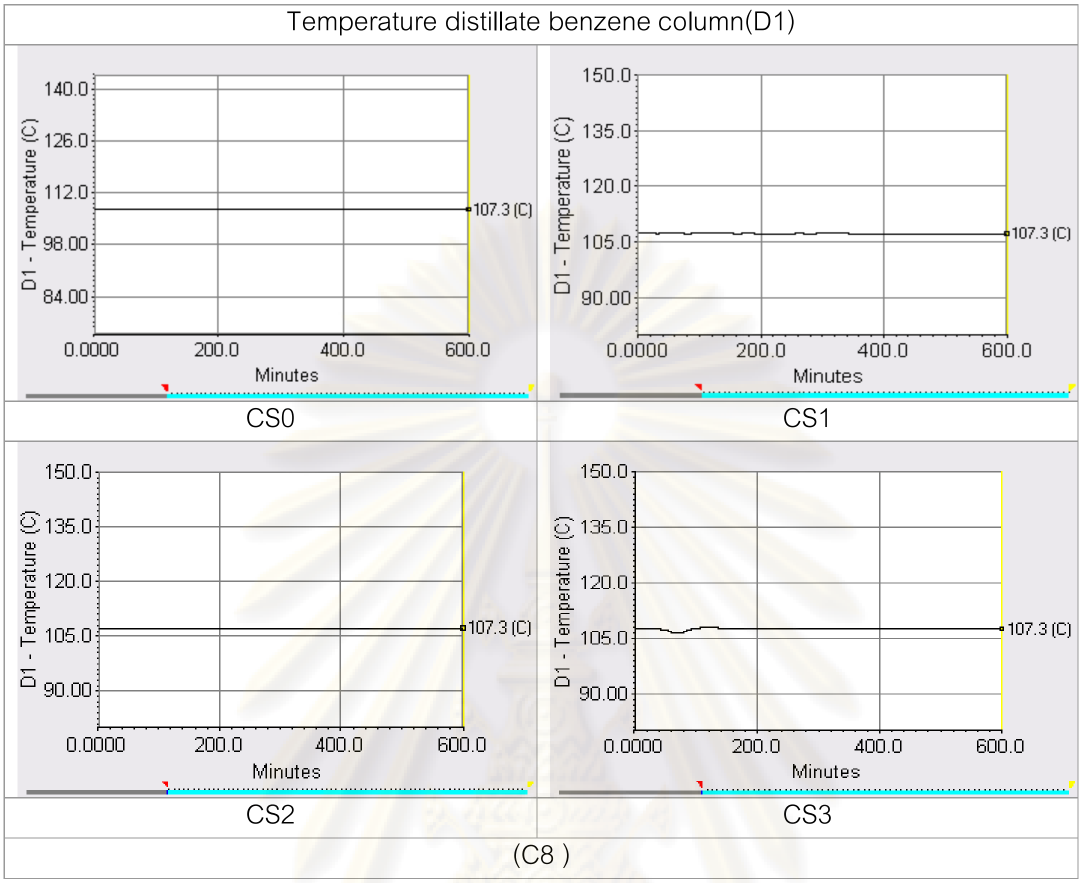
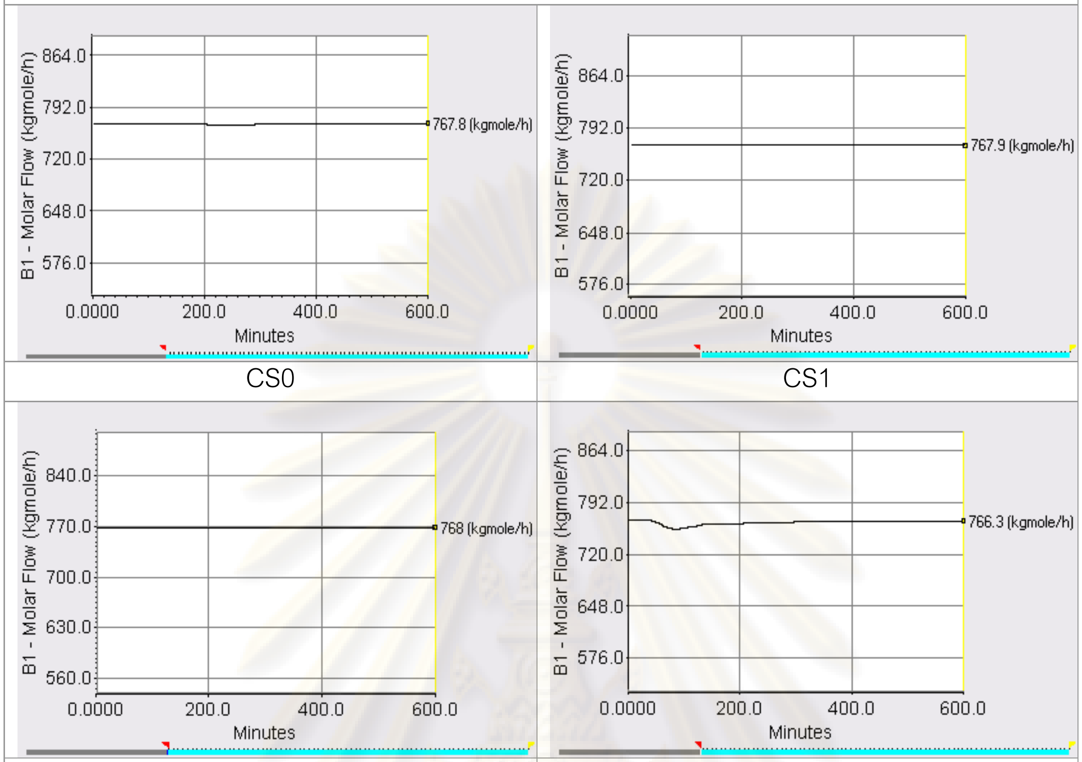


Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

Molar flow bottom benzene column(B1)



CS0 CS1

CS2 CS3

(C9)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

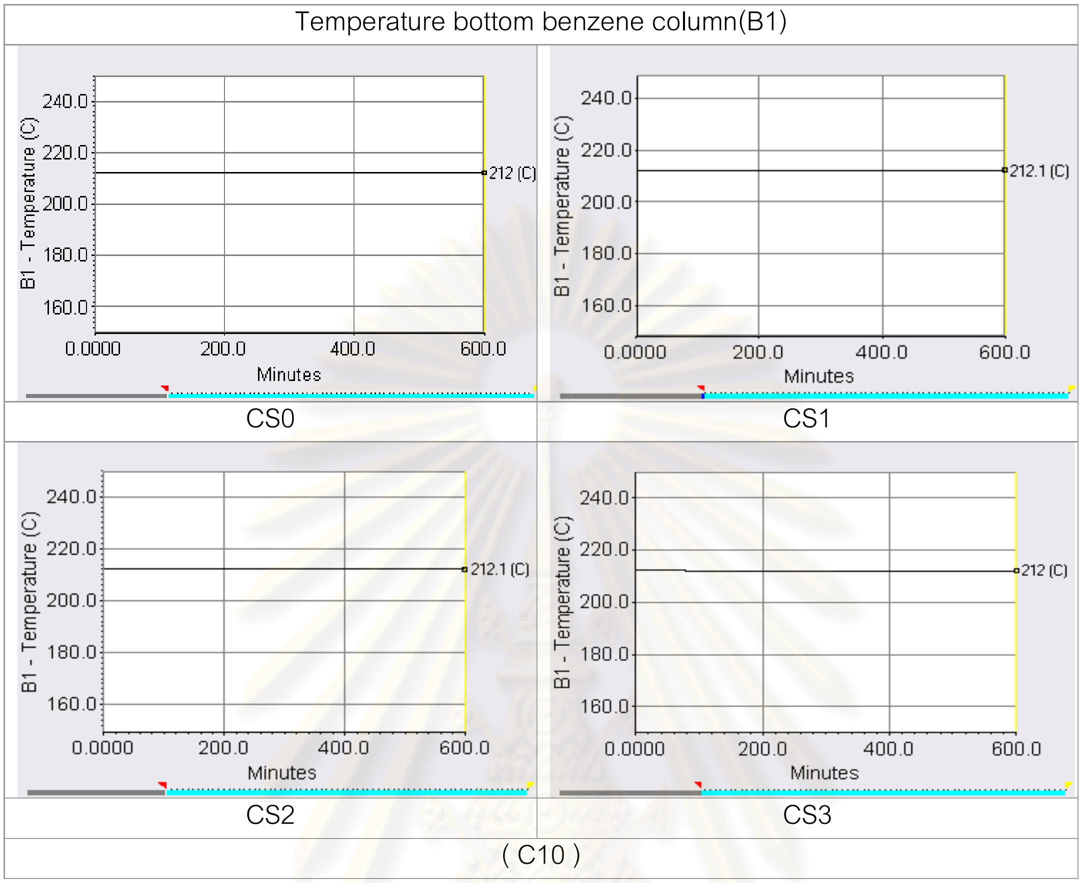
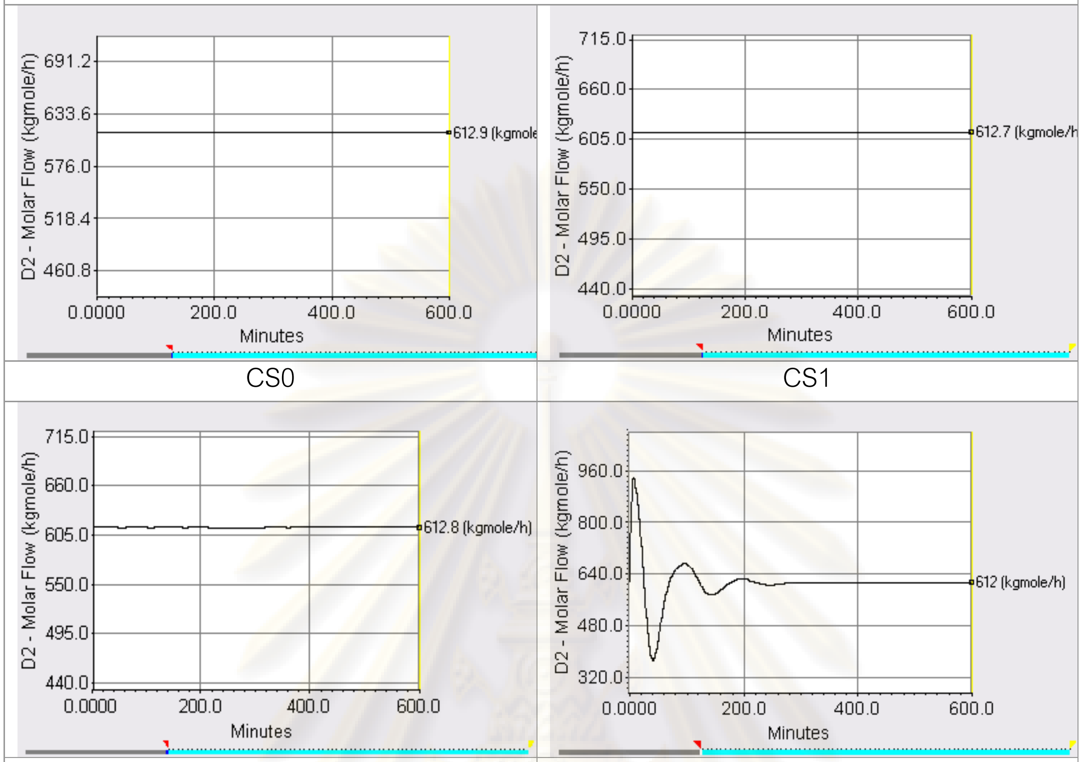


Figure 5.14(Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

Molar flow distillate ethyl benzene column(D2)



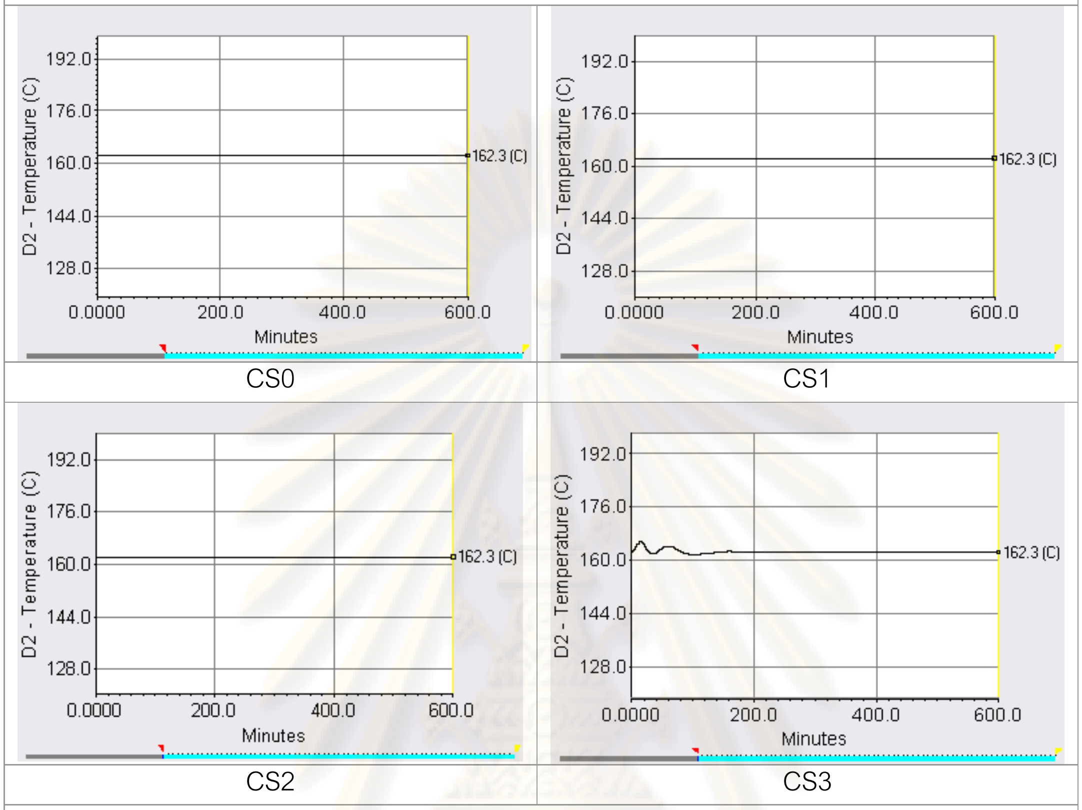
CS0 CS1

CS2 CS3

(C11)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

Temperature distillate ethyl benzene column(D2)



CS0 CS1

CS2 CS3

(C12)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene

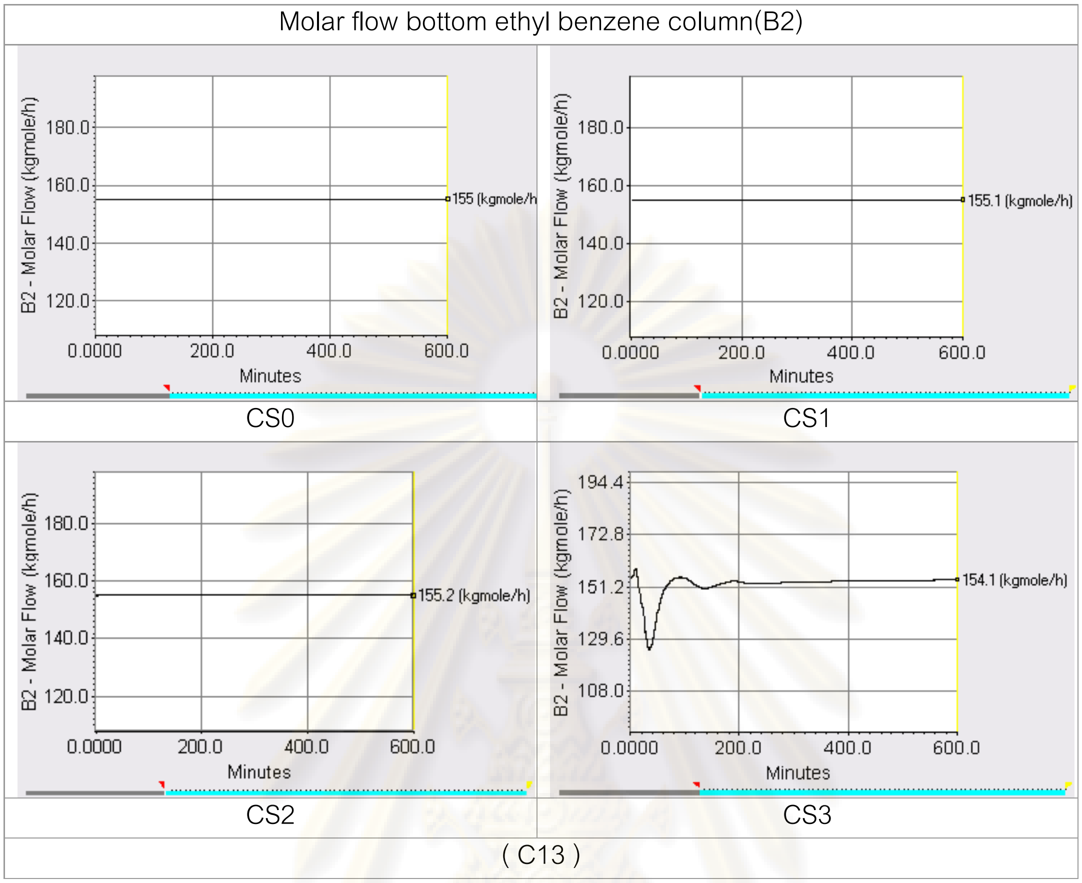
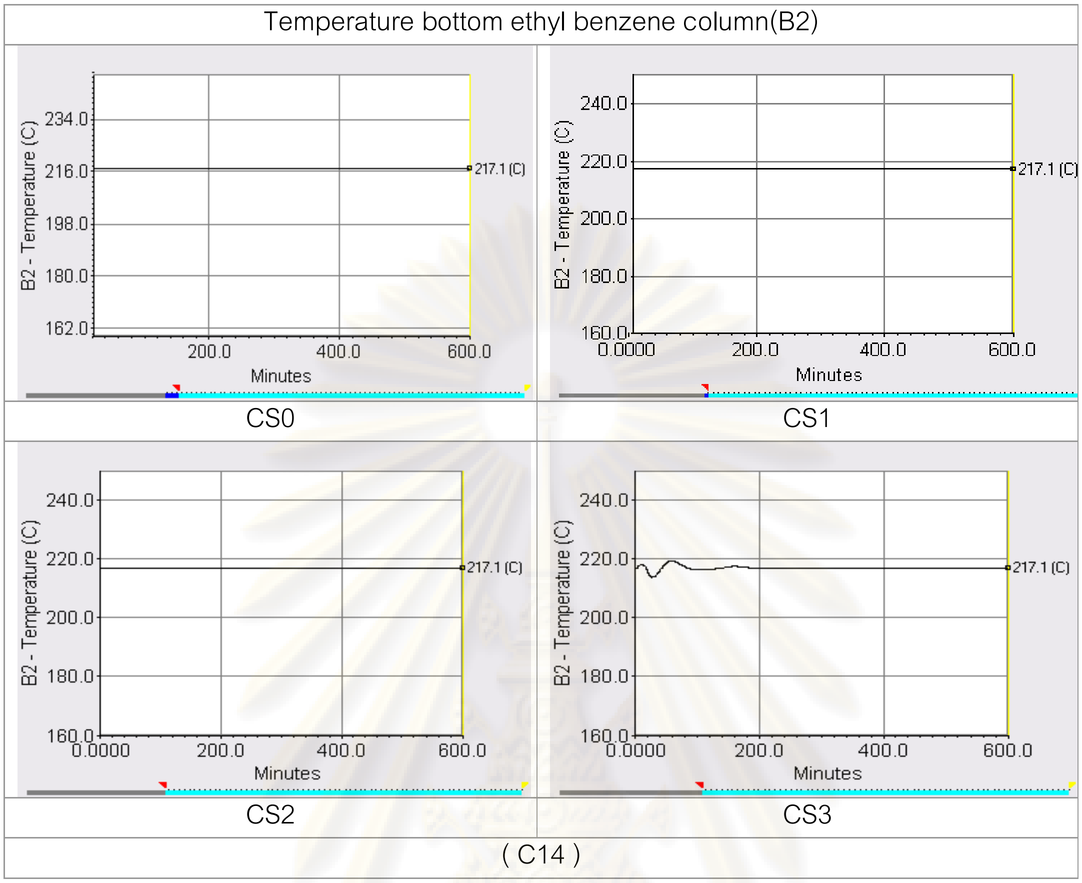


Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3)process by step change of temperature ±10% of FFB fresh feed benzene



Temperature bottom ethyl benzene column(B2)

CS0

CS1

CS2

CS3

(C14)

Figure 5.14 (Continued) Dynamic response of control structure (CS0, CS1, CS2 and CS3) by step change of temperature ±10% of FFB fresh feed benzene

The IAE results of thermal and material disturbance test in four control structures. These results show the control structure 1(CS0) is the best control structure to handle disturbance and CS2 is the best design to maintain product quality and minimize energy use.

CHAPTER VI

CONCLUSION AND RECOMENDATION

6.1 Conclusion

In this research has discussed control structure design for the HDA process, using the design procedure of Wongsri (2008). The process variables should be control are selected by the “Fixture Point theorem”. The procedure is base on the selection of controlled variable that the most disturbances propagate to plantwide, previously using heuristic approach.

The new control structure design use mathematical analysis included heuristic is simple, straightforward and intuitive, the rank of IAE process variables helps process engineer to design the best control structure to handle disturbances.

6.2 Recommendations

This research proves the good control structure Fixture Point theorem to ethyl benzene process is robust to disturbance. Further study on ethyl benzene process, applied Skogestard (2002) plantwide structure control methodology, compare two strategies.

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APPENDICES

APPENDIX A

TUNING OF CONTROL STRUCTURES

A.1 Tuning Controllers

Notice throughout this work uses several types of controllers such as P, PI, and PID controllers. They depend on the control loop. In theory, control performance can be improved by the use of derivative action but in practice the use of derivative has some significant drawbacks:

1. Three tuning constants must be specified.
2. Signal noise is amplified.
3. Several types of PID control algorithms are used, so important to careful that the right algorithm is used with its matching tuning method.
4. The simulation is an approximation of the real plant. If high performance controllers are required to get good dynamics from the simulation, the real plant may not work well.

A.2 Tuning Flow, Level and Pressure Loops

The dynamics of flow measurement are fast. The time constants for moving control valves are small. Therefore, the controller can be turned with a small integral or reset time constant. A value of τ_I = 0.3 minutes work in most controllers. The value of controller gain should be kept modest because flow measurement signal are sometime noisy due to the turbulent flow through the orifice plate. A value of controller gain of K_C = 0.5 is often used. Derivative action should not be used.

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady state offset (the level will not be returned to its setpoint value). However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is K_C = 2.

Most pressure controllers can be fairly easily tuned. The process time constant is estimated by dividing the gas volume of the system by the volumetric flowrate of gas flowing through the system. Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are K_C = 2 and τ_I = 10 minutes.

A.3 Relay- Feedback Testing

The relay-feedback test is a tool that serves a quick and simple method for identifying the dynamic parameters that are important for to design a feedback controller. The results of the test are the ultimate gain and the ultimate frequency. This information is usually sufficient to permit us to calculate some reasonable controller tuning constants.

The method consists of merely inserting an on-off relay in the feedback loop. The only parameter that must be specified is the height of the relay, h. This height is typically 5 to 10 % of the controller output scale. The loop starts to oscillate around the setpoint with the controller output switching every time the process variable (PV) signal crosses the setpoint. Figure A.1 shows the PV and OP signals from a typical relay-feedback test.

The maximum amplitude (a) of the PV signal is used to calculate the ultimate gain, K_U from the equation

$$K_U = \frac{4h}{a\pi} \tag{A.1}$$

The period of the output PV curve is the ultimate period, P_U from these two parameters controller tuning constants can be calculated for PI and PID controllers, using a variety of tuning methods proposed in the literature that require only the ultimate gain and the ultimate frequency, e.g. Ziegler-Nichols, Tyreus-Luyben.

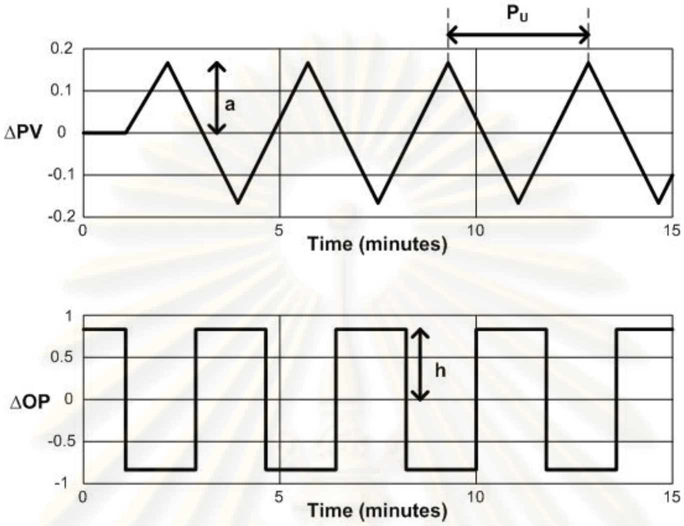

Figure A.1 Input and Output from Relay-Feedback Test

The test has many positive features that have led to its widespread use in real plants as well in simulation studies:

1. Only one parameter has to be specified (relay height).
2. The time it takes to run the test is short, particularly compared to the extended periods required for methods like PRBS.
3. The test is closedloop, so the process is not driven away from the setpoint.
4. The information obtained is very accurate in the frequency range that is important for the design of a feedback controller.
5. The impact of load changes that occur during the test can be detected by a change to asymmetric pulses in the manipulated variable.

These entire features make relay-feedback testing a useful identification tool. Knowing the ultimate gain, K_U and the ultimate period, P_U permits us to calculate controller settings. There are several methods that require only these two parameters. The Ziegler-Nichols tuning equations for a PI controller are:

$$\begin{aligned} K_C &= K_U / 2.2 \\ \tau_I &= P_U / 1.2 \end{aligned} \tag{A.2}$$

These tuning constants are frequently too aggressive for many chemical engineering applications. The Tyreus-Luyben tuning method provides more conservative settings with increased robustness. The TL equations for a PI controller are:

$$\begin{aligned} K_C &= K_U / 3.2 \\ \tau_I &= 2.2 P_U \end{aligned} \tag{A.3}$$

A.4 Inclusion of Lags

Any real physical system has many lags. Measurement and actuator lags always exist. In simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed.

Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint.

Table A.1 summarizes some recommended lags to include in several different types of control loops.

Table A.1 Typical measurement lags

		Number	Time constant (minutes)	Type
Temperature	Liquid	2	0.5	First-order lags
	Gas	3	1	First-order lags
Composition	Chromatograph	1	3 to 10	Deadtime

APPENDIX B

PARAMETER TUNING

Table B.1 Tuning parameters for the reference control structure 0 (CS0)

Controller	Controlled variables	Manipulated variable	Tuning parameter			Action controller	PV Range
			K_C	τ_I	τ_D		
FC1	Fresh ethylene feed flowrate	V1	0.5	0.3	-	Reverse	0-1260 kgmole/hr
LCR1	Reactor 1 liquid level	V2	5.0	-	-	Reverse	0-100 %
TCR1	Temperature control of Reactor 1	Q1	2.0	10.0	-	Direct	100-200 °C
LCR1	Reactor 2 liquid level	V3	5.0	-	-	Reverse	0-100 %
FC2	Reactor 2 effluent feed flowrate	V4	0.5	0.3	-	Reverse	0-3540 kgmole/hr
PC1	Condenser pressure in benzene column	Qc1	2.0	10.0	-	Direct	300-600 kPa
TC1	Temperature at tray 8 in benzene column	Qr1	1.4	15.0	-	Reverse	150-200 °C
LC12	Condenser liquid level in benzene column	V9	2.0	-	-	Direct	0-100 %
LC11	Reboiler liquid level in benzene column	V5	2.0	-	-	Direct	0-100 %
PC2	Condenser pressure in ethyl benzene column	Qc2	2.0	10.0	-	Direct	100-300 kPa
TC2	Temperature at tray 8 in ethyl benzene column	Qr2	3.3	10.0	-	Reverse	150-250 °C
LC22	Condenser liquid level in ethyl benzene column	V7	2.0	-	-	Direct	0-100 %
LC21	Reboiler liquid level in ethyl benzene column	V6	2.0	-	-	Direct	0-100 %

Table B.2 Tuning parameters for the reference control structure I (CS1)

Controller	Controlled variables	Manipulated variable	Tuning parameter			Action controller	PV Range
			K_C	τ_I	τ_D		
FC1	Fresh ethylene feed flowrate	V1	0.5	0.3	-	Reverse	0-1260 kgmole/hr
LCR1	Reactor 1 liquid level	V2	5.0	-	-	Reverse	0-100 %
TCR1	Temperature control of Reactor 1	Q1	2.0	10.0	-	Direct	100-200 °C
FC-R1	Reactor 1 effluent feed flowrate	V3	0.5	0.3	-	Reverse	0-3240 kgmole/hr
FC2	Reactor 2 effluent feed flowrate	V4	0.5	0.3	-	Reverse	0-3540 kgmole/hr
PC1	Condenser pressure in benzene column	Qc1	2.0	10.0	-	Direct	300-600 kPa
TC1	Temperature at tray 8 in benzene column	Qr1	1.4	15.0	-	Reverse	150-200 °C
LC12	Condenser liquid level in benzene column	V9	2.0	-	-	Direct	0-100 %
LC11	Reboiler liquid level in benzene column	V5	2.0	-	-	Direct	0-100 %
PC2	Condenser pressure in ethyl benzene column	Qc2	2.0	10.0	-	Direct	100-300 kPa
TC2	Temperature at tray 8 in ethyl benzene column	Qr2	3.3	10.0	-	Reverse	150-250 °C
LC22	Condenser liquid level in ethyl benzene column	V7	2.0	-	-	Direct	0-100 %
LC21	Reboiler liquid level in ethyl benzene column	V6	2.0	-	-	Direct	0-100 %

Table B.3 Tuning parameters for the reference control structure II (CS2)

Controller	Controlled variables	Manipulated variable	Tuning parameter			Action controller	PV Range
			K_C	τ_I	τ_D		
FC1	Fresh ethylene feed flowrate	V1	0.5	0.3	-	Reverse	0-1260 kgmole/hr
LCR1	Reactor 1 liquid level	V2	5.0	-	-	Reverse	0-100 %
TCR1	Temperature control of Reactor 1	Q1	2.0	10.0	-	Direct	100-200 °C
FC-R1	Reactor 1 effluent feed flowrate	V3	0.5	0.3	-	Reverse	0-3240 kgmole/hr
LC-R2	Reactor 1 liquid level	V4	5.0	-	-	Reverse	0-100 %
PC1	Condenser pressure in benzene column	Qc1	2.0	10.0	-	Direct	300-600 kPa
TC1	Temperature at tray 8 in benzene column	Qr1	1.4	15.0	-	Reverse	150-200 °C
LC12	Condenser liquid level in benzene column	V9	2.0	-	-	Direct	0-100 %
LC11	Reboiler liquid level in benzene column	V5	2.0	-	-	Direct	0-100 %
PC2	Condenser pressure in ethyl benzene column	Qc2	2.0	10.0	-	Direct	100-300 kPa
TC2	Temperature at tray 8 in ethyl benzene column	Qr2	3.3	10.0	-	Reverse	150-250 °C
LC22	Condenser liquid level in ethyl benzene column	V7	2.0	-	-	Direct	0-100 %
LC21	Reboiler liquid level in ethyl benzene column	V6	2.0	-	-	Direct	0-100 %

Table B.4 Tuning parameters for the control structure III (CS3)

Controller	Controlled variables	Manipulated variable	Tuning parameter			Action controller	PV Range
			K_C	τ_I	τ_D		
FC1	Fresh ethylene feed flowrate	V1	0.5	0.3	-	Reverse	0-1260 kgmole/hr
LCR1	Reactor 1 liquid level	V2	5.0	-	-	Reverse	0-100 %
TCR1	Temperature control of Reactor 1	Q1	2.0	10.0	-	Direct	100-200 °C
LCR1	Reactor 2 liquid level	V3	5.0	-	-	Reverse	0-100 %
FC2	Reactor 2 effluent feed flowrate	V4	0.5	0.3	-	Reverse	0-3540 kgmole/hr
PC1	Condenser pressure in benzene column	Qc1	2.0	10.0	-	Direct	300-600 kPa
TC1	Temperature at tray 8 in benzene column	Qr1	1.4	15.0	-	Reverse	150-200 °C
LC12	Condenser liquid level in benzene column	V9	2.0	-	-	Direct	0-100 %
LC11	Reboiler liquid level in benzene column	V5	2.0	-	-	Direct	0-100 %
PC2	Condenser pressure in ethyl benzene column	Qc2	2.0	10.0	-	Direct	100-300 kPa
TC2	Temperature at tray 8 in ethyl benzene column	Qr2	3.3	10.0	-	Reverse	150-250 °C
LC22	Condenser liquid level in ethyl benzene column	V7	2.0	-	-	Direct	0-100 %
LC21	Reboiler liquid level in ethyl benzene column	V6	2.0	-	-	Direct	0-100 %
TC2	Temperature at tray 26 in benzene column	Reflux rate	3.3	10.0	-	Direct	150-250 °C

APPENDIX C

DATA OF FIXTURE POINT ANALYSIS

Table C.1 IAE result of temperature of ethyl benzene

Manipulated Valve

Temperature variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
B1	0.0377	0.0058	0.0081	0.0119	0.0008	0.0007	0.0041	0.0118	0.0467	0.0004	0.0668	0.0058	0.0069	0.2075
B2	0.0191	0.0027	0.0010	0.0025	0.0126	0.0003	0.0367	0.0081	0.0231	0.0021	0.0477	0.0027	0.0599	0.2185
D1	0.0729	0.0101	0.0194	0.0106	0.0015	0.0007	0.0082	0.0217	0.0985	0.0007	0.1519	0.0101	0.0135	0.4197
D2	0.0195	0.0035	0.0020	0.0115	0.0063	0.0003	0.0312	0.0050	0.0239	0.0018	0.0268	0.0035	0.0664	0.2015
Btot	0.0598	0.0188	0.0121	0.0063	0.0010	0.0005	0.0056	0.0131	0.0725	0.0005	0.1110	0.0188	0.0093	0.3293
recycle1	0.0598	0.0188	0.0121	0.0063	0.0010	0.0005	0.0056	0.0131	0.0725	0.0005	0.1110	0.0188	0.0093	0.3293
FFB	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002
FFE	0.0016	0.0003	0.0001	0.0002	0.0000	0.0000	0.0000	0.0000	0.0010	0.0000	0.0007	0.0003	0.0000	0.0042
inert1	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0001	0.0000	0.0000	0.0008
inert2	0.0004	0.0001	0.0001	0.0004	0.0000	0.0000	0.0001	0.0001	0.0005	0.0000	0.0008	0.0001	0.0001	0.0027
vap1	0.0634	0.0097	0.0202	0.0174	0.0007	0.0006	0.0023	0.0054	0.0836	0.0004	0.0250	0.0097	0.0037	0.2420
vap2	0.0466	0.0070	0.0097	0.0122	0.0020	0.0021	0.0059	0.0028	0.0561	0.0005	0.0090	0.0070	0.0098	0.1706

Table C.1 (Continued) IAE result of temperature of ethyl benzene

Manipulated Valve

Temperature variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
p1out	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4199
v9out	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.4200
R1out	0.0634	0.0097	0.0202	0.0174	0.0007	0.0006	0.0023	0.0054	0.0836	0.0004	0.0250	0.0097	0.0037	0.2420
v3out	0.0631	0.0097	0.0201	0.0181	0.0007	0.0006	0.0024	0.0053	0.0835	0.0003	0.0249	0.0097	0.0039	0.2424
R2out	0.0192	0.0027	0.0010	0.0025	0.0126	0.0005	0.0367	0.0081	0.0232	0.0021	0.0475	0.0027	0.0599	0.2187
v4out	0.0472	0.0077	0.0081	0.0073	0.0011	0.0014	0.0048	0.0122	0.0520	0.0004	0.0680	0.0077	0.0081	0.2260
Recycle2	0.0192	0.0027	0.0010	0.0025	0.0126	0.0005	0.0367	0.0081	0.0232	0.0021	0.0475	0.0027	0.0599	0.2187
v6out	0.0192	0.0027	0.0010	0.0025	0.0126	0.0005	0.0367	0.0081	0.0232	0.0021	0.0475	0.0027	0.0599	0.2187
p2out	0.0192	0.0027	0.0010	0.0025	0.0126	0.0004	0.0367	0.0081	0.0232	0.0021	0.0475	0.0027	0.0600	0.2189
v5out	0.0172	0.0034	0.0055	0.0103	0.0120	0.0007	0.0341	0.0047	0.0207	0.0020	0.0240	0.0034	0.0549	0.1930
v7out	0.0009	0.0006	0.0010	0.0070	0.0065	0.0005	0.0050	0.0025	0.0012	0.0003	0.0275	0.0006	0.0087	0.0622

Table C.2 IAE result of pressure of ethyl benzene

Manipulated Valve

Pressure variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
B1	0.3486	0.0062	0.0341	0.0115	0.0047	0.0039	0.0313	0.1432	0.4248	0.0030	0.9984	0.0436	0.0537	2.1070
B2	0.0845	0.0029	0.0043	0.0107	0.0569	0.0013	0.1606	0.0359	0.1012	0.0093	0.2121	0.0119	0.2637	0.9552
D1	0.3488	0.0107	0.0343	0.0113	0.0047	0.0040	0.0313	0.1439	0.4253	0.0030	0.9984	0.0435	0.0537	2.1129
D2	0.0846	0.0037	0.0043	0.0108	0.0567	0.0013	0.1617	0.0359	0.1013	0.0093	0.2120	0.0119	0.2633	0.9567
Btot	0.1926	0.0201	0.0186	0.0242	0.0005	0.0011	0.0039	0.0132	0.1322	0.0004	0.0828	0.0343	0.0057	0.5296
recycle1	0.1926	0.0201	0.0186	0.0242	0.0005	0.0011	0.0039	0.0132	0.1322	0.0004	0.0828	0.0343	0.0057	0.5296
FFB	0.1926	0.0000	0.0186	0.0242	0.0005	0.0011	0.0039	0.0000	0.1322	0.0004	0.0828	0.0343	0.0057	0.4964
FFE	0.1926	0.0003	0.0186	0.0242	0.0005	0.0011	0.0039	0.0001	0.1322	0.0004	0.0828	0.0343	0.0057	0.4967
inert1	0.1926	0.0001	0.0186	0.0242	0.0005	0.0011	0.0039	0.0132	0.1322	0.0004	0.0828	0.0343	0.0057	0.5095
inert2	0.2324	0.0001	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6814
vap1	0.1926	0.0104	0.0186	0.0242	0.0005	0.0011	0.0039	0.0132	0.1322	0.0004	0.0828	0.0343	0.0057	0.5198
vap2	0.2324	0.0074	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6888
p1out	0.1022	0.0000	0.0258	0.0167	0.0029	0.0032	0.0141	0.0980	0.1662	0.0017	0.6507	0.0082	0.0248	0.5437
v9out	0.1022	0.0000	0.0258	0.0167	0.0029	0.0032	0.0141	0.0980	0.1662	0.0017	0.6507	0.0082	0.0248	0.9399
R1out	0.1926	0.0104	0.0186	0.0242	0.0005	0.0011	0.0039	0.0132	0.1322	0.0004	0.0828	0.0343	0.0057	0.5198
v3out	0.2324	0.0104	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6917

Table C.2 (Continued) IAE result of pressure of ethyl benzene

Manipulated Valve

Pressure variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
R2out	0.2324	0.0029	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6843
v4out	0.3481	0.0082	0.0341	0.0112	0.0047	0.0040	0.0312	0.1438	0.4245	0.0030	0.9979	0.0433	0.0536	2.1076
Recycle2	0.2324	0.0029	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6843
v6out	0.2324	0.0029	0.0774	0.2589	0.0174	0.0158	0.0374	0.0867	0.3292	0.0030	0.5245	0.0366	0.0620	1.6843
p2out	0.1778	0.0029	0.0519	0.1738	0.0278	0.0103	0.0698	0.0474	0.2458	0.0045	0.2844	0.0231	0.1157	1.2352
v5out	0.0847	0.0036	0.0042	0.0107	0.0567	0.0013	0.1607	0.0359	0.1013	0.0093	0.2122	0.0118	0.2629	0.9552
v7out	0.0000	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007
vent-C1	0.3488	0.0107	0.0343	0.0113	0.0047	0.0040	0.0313	0.1439	0.4253	0.0030	0.9984	0.0435	0.0537	2.1129

Table C.3 IAE result of molar flow of ethyl benzene

Manipulated Valve

Molar flow variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
B1	0.0572	0.0063	0.0021	0.0058	0.0007	0.0016	0.0046	0.0263	0.0705	0.0004	0.1826	0.0063	0.0079	0.3724
B2	0.0042	0.0009	0.0017	0.0058	0.0004	0.0004	0.0013	0.0025	0.0062	0.0001	0.0150	0.0009	0.0021	0.0414
D1	0.0221	0.0016	0.0053	0.0035	0.0006	0.0005	0.0030	0.0192	0.0346	0.0004	0.1316	0.0016	0.0052	0.2291
D2	0.0265	0.0032	0.0004	0.0018	0.0249	0.0009	0.0573	0.0134	0.0320	0.0032	0.0848	0.0032	0.0727	0.3243
Btot	0.1132	0.0225	0.0168	0.0187	0.0009	0.0007	0.0021	0.0112	0.0557	0.0004	0.0805	0.0225	0.0038	0.3492
recycle1	0.1132	0.0225	0.0168	0.0187	0.0009	0.0007	0.0021	0.0112	0.0557	0.0004	0.0805	0.0225	0.0038	0.3492
FFB	0.1217	0.0213	0.0116	0.0152	0.0005	0.0007	0.0022	0.0026	0.0824	0.0002	0.0525	0.0213	0.0038	0.3361
FFE	0.1011	0.0174	0.0095	0.0124	0.0004	0.0006	0.0018	0.0026	0.0669	0.0002	0.0430	0.0174	0.0031	0.2763
inert1	0.0003	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0001	0.0001	0.0000	0.0009
inert2	0.0004	0.0001	0.0001	0.0004	0.0000	0.0000	0.0001	0.0001	0.0005	0.0000	0.0008	0.0001	0.0001	0.0026
vap1	0.0001	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0002	0.0000	0.0000	0.0009
vap2	0.0001	0.0000	0.0001	0.0003	0.0000	0.0000	0.0000	0.0001	0.0002	0.0000	0.0008	0.0000	0.0000	0.0019
p1out	0.0016	0.0003	0.0002	0.0009	0.0001	0.0001	0.0001	0.0006	0.0012	0.0000	0.0003	0.0003	0.0003	0.2291
v9out	0.0016	0.0003	0.0002	0.0009	0.0001	0.0001	0.0001	0.0006	0.0012	0.0000	0.0003	0.0003	0.0003	0.2291
R1out	0.0911	0.0548	0.1091	0.3093	0.0192	0.0185	0.0392	0.1111	0.2470	0.0035	0.6985	0.0548	0.0646	1.8207
v3out	0.0911	0.0548	0.1091	0.3093	0.0192	0.0185	0.0392	0.1111	0.2470	0.0035	0.6985	0.0548	0.0646	1.8207

Table C.3 (Continued) IAE result of molar flow of ethyl benzene

Manipulated Valve

Molar flow variable of process stream	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
R2out	0.0042	0.0009	0.0017	0.0058	0.0004	0.0004	0.0013	0.0025	0.0062	0.0001	0.0150	0.0009	0.0021	0.0414
v4out	0.0949	0.0382	0.0608	0.0664	0.0152	0.0149	0.0196	0.1402	0.0821	0.0014	1.0000	0.0382	0.0303	1.6022
Recycle2	0.0042	0.0009	0.0017	0.0058	0.0004	0.0004	0.0013	0.0025	0.0062	0.0001	0.0150	0.0009	0.0021	0.0414
v6out	0.0042	0.0009	0.0017	0.0058	0.0004	0.0004	0.0013	0.0025	0.0062	0.0001	0.0150	0.0009	0.0021	0.0414
p2out	0.0042	0.0009	0.0017	0.0058	0.0004	0.0004	0.0013	0.0025	0.0062	0.0001	0.0150	0.0009	0.0021	0.0414
v5out	0.0572	0.0063	0.0021	0.0058	0.0007	0.0016	0.0046	0.0263	0.0705	0.0004	0.1826	0.0063	0.0079	0.3724
v7out	0.0265	0.0032	0.0004	0.0018	0.0249	0.0009	0.0573	0.0134	0.0320	0.0032	0.0848	0.0032	0.0727	0.3243
vent-C1	0.0010	0.0001	0.0001	0.0001	0.0000	0.0000	0.0001	0.0005	0.0011	0.0000	0.0037	0.0001	0.0001	0.0070

Table C.4 IAE result of tray temperature deviation of benzene column

Manipulated Valve

Tray temperature	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
24	0.3266	0.0047	0.0037	0.0015	0.0005	0.0004	0.0033	0.0143	0.0446	0.0003	0.1003	0.0047	0.0057	0.5106
23	0.3266	0.0049	0.0038	0.0023	0.0006	0.0002	0.0034	0.0136	0.0455	0.0003	0.1002	0.0049	0.0059	0.5122
22	0.3265	0.0054	0.0041	0.0038	0.0007	0.0001	0.0036	0.0125	0.0472	0.0003	0.0997	0.0054	0.0063	0.5157
21	0.3265	0.0062	0.0046	0.0064	0.0008	0.0002	0.0040	0.0104	0.0501	0.0004	0.0987	0.0062	0.0071	0.5218
20	0.3264	0.0077	0.0055	0.0110	0.0011	0.0007	0.0047	0.0072	0.0551	0.0005	0.0962	0.0077	0.0083	0.5323
19	0.3263	0.0100	0.0069	0.0184	0.0015	0.0015	0.0057	0.0062	0.0627	0.0007	0.0914	0.0100	0.0102	0.5516
18	0.3262	0.0130	0.0088	0.0287	0.0022	0.0027	0.0071	0.0096	0.0730	0.0008	0.0832	0.0130	0.0129	0.5814
17	0.3262	0.0165	0.0110	0.0411	0.0030	0.0041	0.0087	0.0165	0.0844	0.0010	0.0710	0.0165	0.0158	0.6157
16	0.3261	0.0191	0.0129	0.0516	0.0036	0.0052	0.0100	0.0232	0.0929	0.0011	0.0570	0.0191	0.0180	0.6398
15	0.3261	0.0197	0.0136	0.0563	0.0039	0.0058	0.0103	0.0259	0.0946	0.0011	0.0454	0.0197	0.0187	0.6411
14	0.3260	0.0183	0.0133	0.0532	0.0038	0.0056	0.0097	0.0238	0.0884	0.0010	0.0400	0.0183	0.0176	0.6190
13	0.3260	0.0154	0.0124	0.0431	0.0032	0.0049	0.0084	0.0179	0.0756	0.0009	0.0428	0.0154	0.0151	0.5810
12	0.3259	0.0235	0.0188	0.0866	0.0058	0.0091	0.0127	0.0531	0.0998	0.0014	0.0256	0.0235	0.0231	0.7089
11	0.3259	0.0382	0.0308	0.1706	0.0106	0.0168	0.0205	0.1152	0.1451	0.0023	0.1298	0.0382	0.0374	1.0814
10	0.3258	0.0600	0.0490	0.3104	0.0177	0.0279	0.0321	0.1978	0.2160	0.0036	0.3235	0.0600	0.0579	1.6816

Table C.4 (Continued) IAE result of tray temperature deviation of benzene column

Tray temperature	Manipulated Valve													Summation IAE
	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	
	IAE value													
9	0.3258	0.0818	0.0679	0.4866	0.0244	0.0382	0.0436	0.2624	0.2940	0.0049	0.5980	0.0818	0.0763	2.3856
8	0.3259	0.0885	0.0751	0.6147	0.0260	0.0405	0.0471	0.2612	0.3306	0.0052	0.8676	0.0885	0.0795	2.8504
7	0.3261	0.0743	0.0645	0.6044	0.0212	0.0328	0.0396	0.1969	0.2952	0.0043	1.0000	0.0743	0.0648	2.7985
6	0.3263	0.0509	0.0450	0.4677	0.0140	0.0213	0.0273	0.1177	0.2174	0.0029	0.9349	0.0509	0.0438	2.3202
5	0.3264	0.0185	0.0170	0.1661	0.0042	0.0060	0.0104	0.0214	0.0973	0.0011	0.4743	0.0185	0.0167	1.1777
4	0.3265	0.0310	0.0280	0.2983	0.0080	0.0118	0.0169	0.0575	0.1451	0.0018	0.7262	0.0310	0.0270	1.7092
3	0.3265	0.0117	0.0110	0.0849	0.0022	0.0029	0.0069	0.0040	0.0708	0.0007	0.2617	0.0117	0.0111	0.8060
2	0.3265	0.0082	0.0082	0.0414	0.0013	0.0013	0.0051	0.0073	0.0573	0.0005	0.1279	0.0082	0.0084	0.6016
1	0.3265	0.0065	0.0073	0.0204	0.0009	0.0007	0.0043	0.0112	0.0506	0.0004	0.0703	0.0065	0.0073	0.5130

Table C.5 IAE result of tray temperature deviation of ethyl benzene column

Tray temperature	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	Manipulated Valve													
	IAE value													
30	0.0171	0.0033	0.0030	0.0096	0.0086	0.0007	0.0210	0.0036	0.0201	0.0013	0.0236	0.0033	0.0804	0.1957
29	0.0149	0.0036	0.0045	0.0127	0.0235	0.0015	0.0090	0.0011	0.0165	0.0010	0.0179	0.0036	0.0963	0.2061
28	0.0122	0.0038	0.0061	0.0161	0.0388	0.0022	0.0044	0.0037	0.0127	0.0012	0.0184	0.0038	0.1093	0.2324
27	0.0096	0.0038	0.0072	0.0182	0.0496	0.0027	0.0081	0.0058	0.0092	0.0014	0.0240	0.0038	0.1153	0.2585
26	0.0078	0.0036	0.0077	0.0185	0.0537	0.0028	0.0083	0.0065	0.0077	0.0014	0.0269	0.0036	0.1139	0.2625
25	0.0068	0.0034	0.0076	0.0176	0.0520	0.0026	0.0058	0.0062	0.0068	0.0014	0.0260	0.0034	0.1079	0.2476
24	0.0062	0.0031	0.0073	0.0159	0.0472	0.0025	0.0047	0.0054	0.0058	0.0013	0.0226	0.0031	0.1009	0.2259
23	0.0060	0.0029	0.0069	0.0144	0.0418	0.0022	0.0063	0.0044	0.0056	0.0013	0.0186	0.0029	0.0945	0.2078
22	0.0060	0.0027	0.0066	0.0132	0.0370	0.0020	0.0095	0.0036	0.0056	0.0013	0.0149	0.0027	0.0898	0.1949
21	0.0058	0.0026	0.0064	0.0123	0.0335	0.0019	0.0122	0.0030	0.0057	0.0013	0.0121	0.0026	0.0864	0.1859
20	0.0057	0.0025	0.0063	0.0117	0.0311	0.0018	0.0139	0.0027	0.0055	0.0013	0.0101	0.0025	0.0843	0.1795
19	0.0055	0.0024	0.0064	0.0113	0.0297	0.0017	0.0150	0.0025	0.0053	0.0014	0.0089	0.0024	0.0830	0.1755
18	0.0052	0.0023	0.0064	0.0111	0.0289	0.0017	0.0156	0.0024	0.0050	0.0014	0.0083	0.0023	0.0823	0.1729
17	0.0048	0.0023	0.0065	0.0110	0.0286	0.0017	0.0159	0.0025	0.0045	0.0014	0.0081	0.0023	0.0820	0.1716
16	0.0044	0.0023	0.0067	0.0110	0.0286	0.0017	0.0160	0.0025	0.0041	0.0015	0.0084	0.0023	0.0819	0.1713
15	0.0040	0.0022	0.0069	0.0111	0.0288	0.0017	0.0159	0.0027	0.0036	0.0015	0.0089	0.0022	0.0820	0.1714
14	0.0036	0.0021	0.0070	0.0111	0.0292	0.0018	0.0159	0.0029	0.0033	0.0015	0.0096	0.0021	0.0822	0.1723

Table C.5 (Continued) IAE result of tray temperature deviation of ethyl benzene column

Manipulated Valve

Tray temperature	V1	V2	V3	V4	V5	V6	V7	V9	Q1	Qc1	Qr1	Qc2	Qr2	Summation IAE
	IAE value													
13	0.0031	0.0021	0.0072	0.0112	0.0297	0.0018	0.0160	0.0030	0.0033	0.0015	0.0105	0.0021	0.0824	0.1741
12	0.0026	0.0021	0.0074	0.0113	0.0303	0.0019	0.0162	0.0032	0.0033	0.0015	0.0115	0.0021	0.0828	0.1762
11	0.0024	0.0021	0.0077	0.0115	0.0309	0.0018	0.0164	0.0034	0.0035	0.0017	0.0127	0.0021	0.0830	0.1792
10	0.0024	0.0020	0.0079	0.0116	0.0315	0.0019	0.0167	0.0036	0.0037	0.0017	0.0139	0.0020	0.0835	0.1824
9	0.0079	0.0013	0.0090	0.0126	0.0578	0.0025	0.0129	0.0079	0.0108	0.0024	0.0355	0.0013	0.0993	0.2611
8	0.0080	0.0011	0.0080	0.0116	0.0612	0.0022	0.0130	0.0073	0.0101	0.0026	0.0366	0.0011	0.1003	0.2632
7	0.0020	0.0014	0.0055	0.0086	0.0423	0.0014	0.0153	0.0031	0.0034	0.0021	0.0146	0.0014	0.0892	0.1905
6	0.0076	0.0020	0.0034	0.0058	0.0199	0.0006	0.0217	0.0027	0.0097	0.0018	0.0113	0.0020	0.0770	0.1655
5	0.0132	0.0023	0.0021	0.0040	0.0053	0.0002	0.0288	0.0050	0.0163	0.0018	0.0288	0.0023	0.0687	0.1789
4	0.0164	0.0025	0.0015	0.0031	0.0051	0.0001	0.0329	0.0066	0.0199	0.0019	0.0388	0.0025	0.0642	0.1955
3	0.0179	0.0026	0.0012	0.0027	0.0091	0.0002	0.0350	0.0075	0.0217	0.0020	0.0438	0.0026	0.0619	0.2081
2	0.0186	0.0027	0.0011	0.0026	0.0111	0.0003	0.0361	0.0079	0.0226	0.0021	0.0461	0.0027	0.0607	0.2146
1	0.0190	0.0027	0.0010	0.0025	0.0122	0.0003	0.0364	0.0080	0.0229	0.0021	0.0472	0.0027	0.0602	0.2170

VITA

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