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CONTROL STRUCTURES DESIGN APPLIED TO ALKYLATION PROCESS PLANTWIDE CONTROL

Miss Panisara Khamanarm

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By	Miss Panisara Khamanarm
Field of Study	Chemical Engineering
Thesis Advisor	Assistant Professor Montree Wongsri

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

B. Lerdhirunwong .... Dean of the Faculty of Engineering

(Associate Professor Boonsom Lerdhirunwong, Dr.Ing.)

THESIS COMMITTEE

Piyasan Praserthdam .... Chairman

(Professor Piyasan Praserthdam, Dr.Ing.)

Montree Wongsri .... Thesis Advisor

(Assistant Professor Montree Wongsri, D.Sc.)

Soorathep Kheawhom .... Examiner

(Assistant Professor Soorathep Kheawhom, Ph.D.)

.... External Examiner

(Veerayut Lersbamrungsuk, D.Eng.)

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งานวิจัยนี้เป็นการศึกษาการจำลองการควบคุมระบบแพลนท์ไวด์สำหรับการออกแบบโครงสร้างการควบคุมของกระบวนการไอโซบิวเทน ในการออกแบบโครงสร้างนี้ เป็นการออกแบบกระบวนการที่สามารถควบคุมและลดสิ่งรบกวนของระบบให้ได้มากที่สุด โดยใช้ขั้นตอนการออกแบบของวงศ์ศรี (2009) เปรียบเทียบกับการออกแบบโครงสร้างพื้นฐานของลูเบน (2002) โดยกำหนดสิ่งรบกวน 2 ประเภท คือ อัตราการไหลของสารที่เข้าสู่ระบบการผลิตและการเปลี่ยนสภาวะของอุณหภูมิของสารที่เข้าสู่กระบวนการผลิต ผลของการทดสอบแสดงให้เห็นว่าโครงสร้างการควบคุมของวงศ์ศรีให้สมรรถนะของการควบคุมที่ดีกว่าโครงสร้างการควบคุมของลูเบนทั้งในด้านการคงคุณภาพของผลิตภัณฑ์และในด้านการประหยัดพลังงาน ทำการจำลองโครงสร้างการควบคุมดังกล่าวโดยใช้ระบบปฏิบัติการ HYSYS

ภาควิชา.......วิศวกรรมเคมี.............. ลายมือชื่อนิสิต.........Panisara K..................

สาขาวิชา......วิศวกรรมเคมี.............. ลายมือชื่อ อ.ที่ปรึกษาวิทยานิพนธ์หลัก..............

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Design a process control structure for a complex process, such as the process having material or energy recycle, is a complicate task. The design control loop would effect the operation significantly.

The study is using plantwide control strategies for designing control structure of Isobutane process control. This structure process design is to achieve the efficiency in control and reduce disturbances in process by using the new process structure design by Wongsri (2009) compare to the basic heuristic structure design by Luben (2002). Disturbances are set up by two categories; flow rate of substances and temperature changed in fresh feed. The result shows that control structure designs based on Wongsri heuristic are better performance than Luben in both sustain in quality and minimize the power efficiency. Evaluate such simulation by HYSYS commercial software.

Department: ......Chemical Engineering.........Student's Signature: ......Panisara K.......

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CHAPTER I

INTRODUCTIONS

This Chapter is an introduction describing the importance and reasons for research, research objective, scope of research, contribution of research, procedure plan and research framework for general point of view the this research.

1.1 Importance and Reasons for research

Nowadays, process plants have been designed in various ways and strategies to capable handling the large changes in throughput, product specifications and feedstock quality.The process should be capable to move between various operating points with smooth and short transitions. These issues are critical also in the case of plants with recycles where strong nonlinearities could manifest by unwanted dynamic phenomena such as unstable steady, side reaction, external disturbances or even snowball effect behavior. A good design should predict and avoid the occurrence off unstable behavior when the process is faced with flexibility and switch-ability requirements.

Important issues are feed policy and reactor design, with regard to desired conversion and selectivity. These should comply with the requirement of a stable operating point faced to large disturbances.

Isobutane process is widely used because it is a realistically complex chemical process and the complicated system's dynamic behavior so, it have been selected to be the model for new control structure design. This research will design plantwide control structures of Isobutane process using new plantwide control structure design procedure based on Fixture point theorem (Wongsri, 2008) to select alternative set of controlled variables comparative the performance of plantwide to handle the disturbance by both internal and external with control structure based case of Luben. The simulation was performed by HYSYS software.

1.2 Research Objective

The general research objective of this thesis has been to study the plantwide dynamics and control behavior of the Isobutane process plant wide control structure design. Specific focus of this research has been to investigate the controllability of the Isobutane process as given in Luyben and Tyreus (1998) for the research based case comparative the plantwide control structure performance to a new alternatives plantwide control strategy design apply from the Fixture point theorem of Wongsri, 2008. Details of key unit operations and external disturbances are investigated to the extent that they impact on the plantwide behaviour and control performance. The specific aims are as follows:

1. To design the Isobutane plantwide control structure design simulate both steady state and dynamics condition by using commercial HYSYS software based on Fixture point theorem (Wongsri, 2008). To compare the structure with Luben and Tyreus (1998) control strategy as based case.
2. To establish and test the alternative control strategies by external disturbances in an effort to provide better control performance and comparative.

1.3 Scopes of research

1. Simulation of the Isobutance alkylation process is operated by HYSYS commercial software.

2. Description and detailed understanding of the of Isobutane process obtained from Douglas, J. M. (1998), William L. Luyben (2002).

3. To establish the 3 alternative control structure designs for Isobutane plantewide process using Fixture point method of Wongsri, 2008 by compare the performance to handle disturbances to the based case of Luyben et al., (1999).

1.4 Contributions of research

The new Isobutane control structure designs with heuristic and mathematic strategy for an industrial alternative with effective operation and maximum the profit.

1.5 Procedure Plan

1. Steady state modeling and simulation of Isobutane process
2. Dynamic modeling simulation and practice
3. Design of control structure of Isobutane
4. Establish the new control structure design follow by Fixture point theorem (Wongsri, 2008).
5. Assessment of performance of the control structure design
6. Analysis the design and simulation results.
7. Conclusion of the thesis.

1.6 Research Framework

This thesis is classified into six chapters as follows:

Chapter I	is an introduction of the research, this chater consist of research objective, scope of research, contribution of research, and procedure plan.
Chapter II	reviews the literature and previous work related to plantwide control structure design procedure.
Chapter III	covers some background information of plantwide and theory concerning with plantwide control fundamentals, plantwide control design procedure, and control structure evaluation.

Chapter IV Isobutane Plantwide process description. results comparative with the control structures of Luyben based case.

Chapter V describes the new designed control structures and dynamic simulation results comparative with the control structures of Luyben based case.

Chapter VI presents the conclusion of this research and makes the recommendations for future work.

This is followed by:

Reference

Appendix A: Isobutane Process Stream and Equipment Data

Appendix B: Parameter Tuning and control structure

Appendix C: Data of Integrate Absolute Error analysis

CHAPTER II

LITERATURE REVIEW

A. Plantwide control design procedure

2. Design and Control of Integrated Plants

Wongsri and Amornchai (2006) were presented the 3 control structures with 3 difference energy integration illustrating a dynamic behavior of plantwide control structures for Hydroalkylation. They were designed to change the disturbance of heat of the cold stream and toluene recycle flow rate. Comparative the performance of those three control structures, the 1st control structure was modified from control structure of Luyben by manipulated the valve position to control the temperature of column. The 2nd control structure was modified from the 1st control structure by adding cooling for controlling the outlet temperature replacing using the internal process flow. For the 3rd control structures, there was the ratio control adding to the 2nd control structure in order to control the hydrogen and toluene ratio during the process. The results obtained by IAE vales have been shown that the 3rd control structures gave a less setting time and had ability to handle the disturbances much better than others. Comparative of utility consumption for all control structures, the result were shown that the 1st control structure was less consumption than the others. Moreover, the control structure 3 was the slowest response comparing to the other alternatives

Wongsri and Kietawarin (2002) presented a research of the comparative of 3 control structure designs of Hydroalkylation plantwide by evaluated the performance to handle the disturbances of production flow rate changed. The load disturbances were designed by changing the feed flow rate and temperature of toluene fresh feed. There was the toluene fresh feed change for control structure 1, the 2nd control structure was modified from the 1st structure by adding the cooling controlling the reactor outlet temperature. The 3rd control structure was modified from the 2nd control structure by ratio changed for controlling the toluene and hydrogen in process. Comparative all control structures with the based case reference control structure of Luyben, the results obtained from the IAE values were shown that those three control structures achieved the good performance better than the based case reference.

P. R. Lyman and C. Georgak (1993) studied the performance of 4 plantwide control structures by developing from the Tennessee Eastman problem. Their control structures have been developed to a tiered fashion without using the quantitative steady state or dynamic process. Their method was 1st chooses the production manipulated located on the main process route. The inventory control loops was an outward direction. Those production and inventory control have been ensured to self consistent as mentioned by Georgakis (1993). The result was shown that the 2nd control structure was the best performance for this studied. One of their structure provides effective control to handle all circumstances for 50 hours of production time which supporting the strength of the tiered plantwide control methodology applied.

J.J. Castroa, F.J. Doyle (2004) presents a heuristic design for plantwide control applied to the pulp mill benchmark. Comparison of 2 control strategies (decentralized control and model prediction control for unit-based model) evaluating by their capacity to reduce all total error and to maximize the production profits. The control strategies were studied through several closed loop simulations included several disturbances and the change of setpoint in the many unit operations. Simulations results showed that pulp mill processes can be controlled with conventional decentralized control also with a combination of decentralized control and unit-based MPC. MPC was able to show a reduction in the total error more than the decentralized control strategy for most of the outputs, also maximized the 30% profits above the decentralized strategy.

Gonzalo Molina and David Zumoffen (2009) present a new systematic methodology for synthesizing the plant-wide control structure, evaluation the process by test a reactor/separator with recycle plant. They were considered the optimum energy consumption by avoiding any heuristic concepts. The approach consists of 3 steps: optimization, stabilization, and final pairing between manipulated and controlled variables, evaluation on the performance to handle and reject the most critical disturbances. Their result achieves a good decentralized structure considering the most normal objectives of any plant, which are trying to optimum energy and good regulatory behavior. Even though their decision making technique was based mainly on steady-state information, the good results were achieved as they were demonstrated the simulations performing by dynamic model. In addition, a systematic multivariable tuning method, thought to be enough robust for handling well the most crucial disturbances, was very useful . Furthermore, their approach was helpful for deciding whether an equipment sizing suitability to assure the plant wide controllability.

Busara Kunajitpimol (2006) presents the heat exchanger networks and new plant wide control structure design for achieving a dynamic maximum energy recovery for Butane Isomerization plant. The 2 control structure of heat exchanger network were used the heat from the reactor stream providing the heat to the column reboiler. The result have been shown that an energy was save 24.88% from the existing design without the heat integration but the production cost was raised by 0.67% because of exchange added to the process. Designed of 4 control structure based on the control structure of Luyben. Comparative all performance shown that the control structures without heat integration were effective than the others.

Luyben and Yi (1995) provided a initial evaluation of plant-wide control structures comparative to some structures which gave a poor result. Equation based algebraic equation solvers were used to find the change in steady state occurring in all manipulated variables for a control structure when the load has been changed. Some variables were fixed for all control structure such as temperatures, flows, compositions etc. Set no. of variables equal to DOF of the closed loop system. When the control structure required the large changes in manipulated variables, such control structure was a poor one since, a valve saturation and/or overloading equipment will occurred. Thee effectiveness of the other structures were simulated by dynamic simulation. Some control structures were found that they had a multiple steady states and produced the closed loop instability.

Suchada Suntisrikomal (2008) presents the "Fixture point theorem" for the new plantwide control structure design by selecting an appropriate set of controlled variables. She was used the maximum (scaled) gain to select and pare the controlled variables with manipulated variables. The control variables were set based on Luben Luyben (1998) heuristic control structure. The controlled variables 3 sets were selected from the rank of fixture point. Those performance of all designed control structures were evaluated by IAE value comparative with reference structure.

Shih-Wen Lina, Cheng-Ching Yu (2004) analyzed the tradeoff between steady-state economics and dynamic controllability for heat-integrated recycle plants. The process consists of one reactor, two distillation columns, and two recycle streams first studied by Tyreus and Luyben which the two distillation columns was heat integrated. Results show that the steady-state controllability deteriorates gradually as the degree of heat integration increases. However, if the recycle plant is optimally designed, acceptable turndown ratio is observed and little tradeoff between steady state economics and dynamic operability may result. The results reveal that improved control can be achieved for well-designed heat-integrated recycle plants (compared to the plants without energy integration). More importantly, better performance is achieved with up to 40% energy saving and close to 20% saving in total annual cost.

Skogestad (2004) proposed the control structure design dealing with the decisions making of the control system, including what should be control and how to pair those variables forming the control loops. The systematic procedure for control structure design was presented for complete chemical plants by starting with defining an operational and economic objectives, and the degrees of freedom available to fulfill them. The other include an inventory and production rate control, decentralized versus multivariable control, the loss in performance by bottom up designing, and a definition of the "complexity number" for control the system.

Busara Kunajitpimol (2006) presents the heat exchanger networks and new plant wide control structure design for achieving a dynamic maximum energy recovery for Butane Isomerization plant. The 2 control structure of heat exchanger network were used the heat from the reactor stream providing the heat to the column reboiler. The result have been shown that an energy was save 24.88% from the existing design without the heat integration but the production cost was raised by 0.67% because of exchange added to the process. Designed of 4 control structure based on the control structure of Luyben. Comparative all performance shown that the control structures without heat integration were effective than the others.

CHAPTER III

THEORY

3. Basic Knowledge for Plantwide Control

The key in chemical process economic realization, one is based on the design of well established set objectives of control structure system. By this point, to select the set of controlled variables by maintain the operation closing to an optimum steady state performance during the transients caused by disturbances or operational parameter changes have decreased. A balance between design and operation is a crucial evaluation in the success of a chemical process. Another matter to achieve the successful in process operation is the understanding of process development. To obtained the clearly picture of plantwide control is the best understanding of the process dynamics. It is very important for engineer of chemical process plants to analyze an operating data in order to maintain and improve the understanding of how was the actual plant action.

Present research is fast developing in plantwide control structure design, operation and control since, it is identified an important criteria for improving the performance of existing processes. The benefits of this studies including to improve an understanding of the process dynamics and response with potentially leading to identify in design and control configuration.

Without performing a plantwide dynamic study it is difficult for the engineer to fully understand the complex behavior of the plantwide design possibly leading to suboptimal control system, process and safety systems design. This analysis is a key item for manufacturers to use in achieving their gods of optimizing the dynamic behavior and performance of the overall plantwide control configuration.

To meet the need of process design information, Luyben and Tyreus (1998) proposed an industrial design for the Isobutane process giving the research community an integrated chemical process to study.

3.1 Multiple unit operations

- Interactions
- Disturbances
- Overall time constants are altered
- Mass recycle
- Energy recycle

3.2 Plantwide Issues

- Process plant is made up processing units in series, the interaction between variables will be evaluated the equipment performances
- If the plant is "once-through", process control and operation issues are quite straightforward.
- Recycle often required leading to serious interaction
- Heat integration adds to the complexity

3.3 Integrated Process

3.1.1 Material Recycles

Material is recycled for six basic and important reasons.

1. Increase conversion.

   Especially for reversible reaction where conversion is limited by equilibrium.

2. Improve economics.

   Lower conversion reactor with recycle is cheaper compared to single or series of reactors with the necessary conversion

3. Improve yields.

In reaction system such as A -> B -> C, where B is the desired product, the per pass conversion of A must be kept low so that less C being produced.

4. Provide thermal sink.

   For highly exothermic reaction with difficulties in providing sufficient cooling. Need more material to absorb the heat

5. Prevent side reactions.

   Maintain excess of one reactant to suppress side reaction due to other reactants, this excess reactant must be separated at reactor effluent and recycled

6. Control properties.

   Important for polymer reactors, conversion of some monomers is limited on purpose to achieve the desired properties and to control the increase in viscosity

3.3.2 Effects of Recycle

1. Alters plant's dynamic and steady state behavior
2. Time constants : Overall time constants can be very different than the sum of time constants of individual units
3. "Snowball Effect" : Small change in throughput or feed composition can lead to large change in steady state recycle stream flow rates, it is better to control the flow of recycle stream

3.3.3 Energy Integration

To improve thermodynamic efficiency of the process, leading to reduction of utility costs

1. Recovery of heat from exothermic reaction
2. Recovery of heat from overhead vapor of distillation column
3. Complex structure, leads to nontrivial control problem.
4. Process to Process Heat Exchanger

3.3.4 Chemical Component Inventories

In chemical processes can characterize a plant's chemical species into three types: reactants, products, and inert. The real problem usually arises when we consider reactants (because of recycle) and account for their inventories within the entire process. Every molecule of reactants fed into the plant must either be consumed or leave as impurity or purge. Because of their value so we prevent reactants from leaving. This means we must ensure that every mole of reactant fed to the process is consumed by the reactions.

This is an important, from the viewpoint of individual unit; chemical component balancing is not a problem because exit streams from the unit automatically adjust their flows and composition. However, when we connect units together with recycle streams, the entire system behaves almost like a pure integrator in terms of reactants. If additional reactant is fed into the system without changing reactor conditions to consume the reactants, this component will build up gradually within the plant because it has no place to leave the system.

3.3.5 Effects of Recycle

This phenomena must be observe and evaluation since, its tendency of able to exhibit the large variation in magnitude of recycle flow.

Moreover, the recycle leads to the "snowball" effect. A small change in throughput or feed composition can lead to a large change in steady-state recycle stream flow rates.

The large swings in recycle flow rates are undesirable in plant because they can overload the capacity of separation section or move the separation section into a flow region below its minimum turndown. Therefore it is an necessary to select a plantwide control structure and evaluate in recycle inventory in order to avoids such effect.

3.4 Plantwide Control Design Procedures

Luyben et al. (1998) described the 9 steps of design procedure in details as follow

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 quantity of an available control valves.

There was number control variable of degrees of freedom, 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 control valves sometimes able to improve the dynamic performance, but often in no choice in their location.

Step 3: Establish energy management system

They use the energy management to describe by two functions: (1) Provide a control structure to removes an exothermic heats of reaction from the process. If heat disable to remove directly at the reactor so, apply such heat to another process with also required such heat. (2) If heat integration occurs between process streams, so, provide the control system to prevent 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).

Step 7: 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.

Step 8: Control individual unit operations

Establish the control loops necessary to operate each of the individual unit operations.

Step 9: 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.

3.5 Fixture point theorem

<u>Fixture points theorem analysis</u>

1. Identify key material variables (KMV) and key energy variables (KEV) from nature of process variables (large magnitude, high frequency and no quickly exit to surrounding).
2. Consideration at dynamic mode (open loop control) of Hysys simulation.
3. Steps of fixture point

<u>Step1</u>. Output (key process variables) screening technique: The candidate output (key process variables) can be considered by selecting the best in term of having the most disturbances from the selected inputs, i.e. having the largest IAE value. Change Input variables (MV and D) for ranking output
sensitivity (The most sensitivity = key process variables)

Step2.Test load disturbance (Load disturbance = key process variables from step1) Rank the key process variables.

Step3.Normalize the IAE value which different units by maximum variation of each typical variable for bring it to summarize (Type of process variables are temperature, pressure, flowrate and composition)

Step4.Consider the first ranking of controlled variables: Considering key process variables that gives the most disturbance of other process variables (sum of normalize IAE value is large) should be controlled.

$$IAE = \int_0^{\infty} |e(t)| dt$$

As equation (3.1), $e(t) = \Delta y_i(t)$ is the change in i process variable on time t when key process j change ((y_{K_j}), when i and j=1, 2, 3 ...n)

CHAPTER IV

PROCESS DESCRIPTION AND CASE STUDY

This chapter presents an overview of the plantwide control structure design starting with a discussion of the reaction kinetics of Isobutene followed by a schematic illustration of the process along with the key unit operation rating and process parameters. In addition, the overall control objectives are presented following by a description of the plantwide control proposed strategy given in Luyben et al., (1999).

4.1 Isobutane process description

The alkylation of isobutane with butane to form iso-octane is a widely used method for producing a high –octane blending component for gasoline.

The main reaction is to combine of isobutene and butane to form iso-octane.

$$iC_4 + C_4^{=} \longrightarrow iC_8$$

However, there is an undesirable consecutive reaction of butene with iso-octane to form dodecane.

$$iC_8 + C_4^{=} \longrightarrow C_{12}$$

The actual chemistry is more complex than these two simple reactions, but for our purpose, they capture the essence of the overall chemistry. The kinetic data is taken from a case study given by Mahajanam et al. (Ind. Eng. Chem. Res. 2001, 40, 3208). These exothermic reactions are irreversible and occur in the liquid phase. Luyben and Tyreus (1998)

The kinetic expressions assumed to be valid for the system are

$$r_1 = 9.6 \times 10^{13} \exp(-28000 / RT)(C_{iC4})(C_{C4})$$

$$r_2 = 2.4 \times 10^{17} \exp(-35000 / RT)(C_{C_8})(C_{C_4^{=}})$$

Where reaction rates are lb-mol/hr-ft^3, activation energies have units of Btu/lb-mole and concentrations are lb-mol/ ft^3.

Note that the activation energy of the second reaction is larger than the first, therefore low temperature favors the desired first reaction. This is why the reaction is carried out at low temperature (50°F)

The second undesirable reaction is also suppressed by keeping the concentration of butane low. This is achieved in two ways. First, there is a large excess of isobutene (ten to one) fed to the reaction section. Second, the butane feed is not all fed into the first reactor, but the stream is split between the first two reactors.

The reaction uses sulfuric acid as a catalyst is conducted in a heterogeneous two-phase liquid mixture of organic and acid phases. We will ignore the acid phase in our simulations.

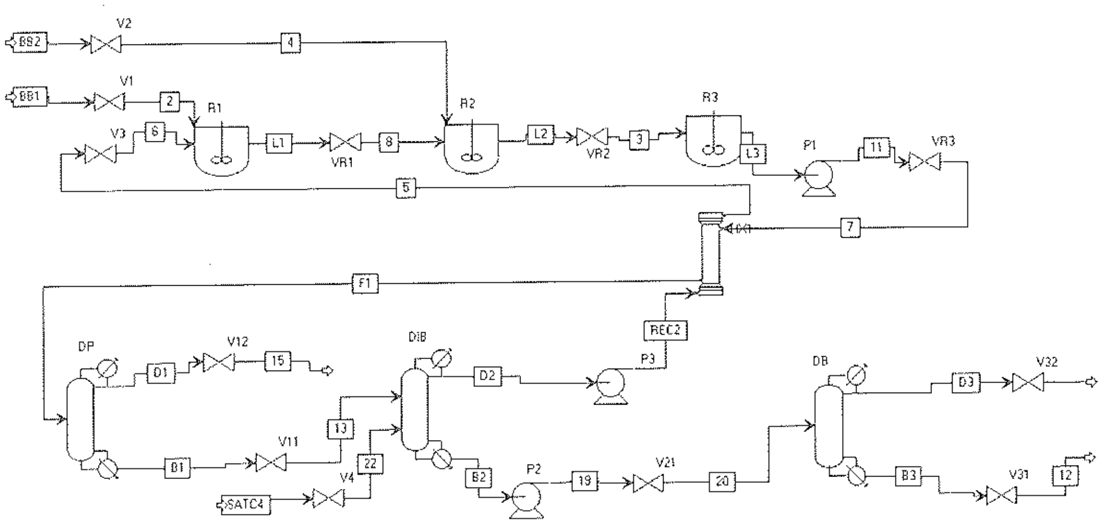

Figure 4.1 The Isobutane process (Luyben 1998)

4.2 Process General view

The process design used in this study focuses on the central production units and excludes the feed storage and product purification stages. To provide an overall view of the process design Figure 4.1 shows a block flow diagram that highlights the major process sections and recycle streams. The design itself represents the classic engineering case study as discussed in Douglas (1988) involving a feed mix area, a reaction section, a liquid vapour separation section, a vapour reactants recovery section, a liquid reactants recovery section (azeotropic distillation column), both a gas and a liquid recycle stream and finally product streams. Figure 4.2 provides the entire process flowsheet schematic showing the individual unit operations, flow lines and control valves. Note, there are 22 degrees of freedom in the process design each represented by the valves given in the diagram. To maximize conversion, the liquid recycle flow is to be maintained at as high a value as possible (Fisher et al., 1988). To do this the valves on the gas exit streams of the vapourizer, separator and absorber are either left filly open or are completely removed to save the cost of the valve (Luyben et al., 1999).

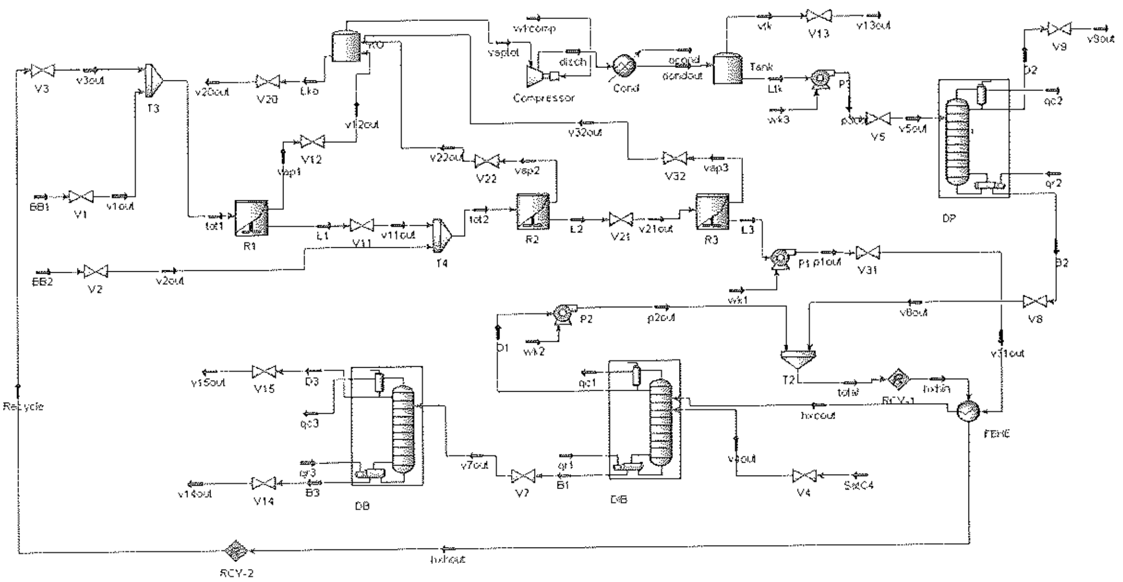

Figure 4.2 Isobutane Hysys process flow sheet.

CHAPTER V

CONTROL STRUCTURE ALTERNATIVES

The goal of control studies is to not only to analyze the performance of existing schemes, but also to investigate the ways to improve them. Therefore, in this section alternative control structure strategies are presented and compared to the performance of the base control scheme.

5.1 Control Objective

For this study, the process has 45 measurements and 22 manipulated variables as listed in Appendix A. prerequisite for most studies on this problem is a process control strategy for operating the plant. The control objectives for this process are typical for a chemical process:

1. Maintain process variables at desired values and satisfactory.
2. Keeps smooth process operating conditions within equipment constraints.

5.1.1 Product variability

1. Minimize the variability of product rate and product quality during disturbances
2. Minimize the movement of valves which able to affect to the other processes.
3. Recover quickly and smoothly from disturbances, production rate changes or product mix changes.

5.1.2 Steady-State Modeling

First, a steady-state model is built in HYSYS PLANT, using the flowsheet and equipment design information, mainly taken from Luyben et al.(1998). The data and specifications for the equipment employed other than the three columns specified in Appendix A. For this simulation, Chao Seader model is selected for fluid package property because of its reliability in predicting the properties of most hydrocarbon-based fluids over a various range of operating conditions. The reaction kinetics of both reactions is modeled with standard Arrhenius kinetic expressions available in HYSYS., and the kinetic data are taken from Luybenet al. (1998).

5.2 Plantwide control design procedure

Step 1. Establish Control Objectives.

For this process, the essential is to produce pure Isobutane recycle back while minimizing yield losses of Isobutane.

1. Production rate: 24.51 lb-mole/h
2. Product quality: Isobutane purity ≥ 90 %
3. Product quality: propane purity ≥ 90 %

Step 2. Determine Control Degree of Freedom.

There are 22 control degrees of freedom for Isobutane three alternatives structure, lists in Table 5.1.

Step 3. Determine Control Degree of Freedom

The control degrees of freedom for each control structures are shown in Table 5.1

Table 5.1 Degree of freedom of Isobutane production process

Unit Operation	Position of Valves	Qty
Reaction sections	Fresh feed valve for butane and recycle of Isobutane (V1, V2, V3), vapor valves (V12, V22, V32) and liquid valves (V11, V31)	8
DIB column	Fresh feed valve for DIB column (V4), bottom valve (V7), Condenser duty DIB (qc1), Reboiler duty DIB (qr1)	4
DP column	Fresh feed valve to DP column (V5), distillate vale (V9), bottom valve (V8), Condenser duty DP (qc2), Reboiler duty DP (qr2)	3
DB column	Bottom valve (V14), Condenser duty DIB (qc3), Reboiler duty DIB (qr3)	4
Compressor power	Wkcomp	1
Cooling duty	Qcond	2
	Degree of freedom	22

Step 3. Establish Energy Management system

In this work, we have no designed the new heat exchanger network, since this Isobutane is required the low temperature for reactor operation. The exothermic heat of reaction also must be removed, reactor feed must be heated to a high enough temperature to initiate the reaction.

Step 4. Set Production Rate

There are not constrained to set production either by supply or demand to set production rate at a certain point in process. Will be determining which variables effect the reactor productivity. Considering in temperature, the relative change in reaction rate depends on temperature through the activation energy, therefore we choose the reactor inlet temperature set point as the production rate handle for irreversible case.

Step 5. Control Production Quality and Address Safety

The achievement of the research is the Isobutane product in distillate from the DIB column must be more than 90% purity and to keep normal butane impurity at 2%, propane more than 90% of distillate

Step 6: Fix a flow in every recycle loop and control inventories (pressure and level)

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 three reactors (CSTR). 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 reactor level must be controlled; it can level directly or via flow control. Level control design in Isobutane production process can be controlled by the drown stream of each vessel. The most direct way to control the remaining levels would be with the exit valves. Recycle loop control, all of the flows around the liquid recycle would be set on the basis of level which would lead to undesirable propagation of disturbances. So a flow somewhere in this loop should be controlled.

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

Step 7: Check component balance.

Isobutane process have a 6 composition such as propane, isobutane, 1-butene, n-butane. The composition of two fresh feed are Isobutane and 1-butene with Isobutane recycle stream.

Component balances control loops consists of:

• Isobutane is purged from the gas recycle loop to prevent it from accumulating and its composition can be controlled with the purge flow.

• Normal-butane is removed in the bottom stream from the DIB column, where bottom stream controls base level. And control temperature (or concentration) with the reboiler steam.

• Isobutane inventory is accounted for via level control in the recycle column overhead receiver.

Step 8: Control Individual Unit Operations and Validation via rigorous dynamic simulation.

The rest degrees of freedom are assigned for control loops within individual units. These include: cooling water flow to the cooler controls process temperature to the separator and refluxs to the stabilizer, product, and recycle columns are flow controlled.

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

5.3 Fixture Point Theorem

For new control structure design CS1 -CS3, we follow the step of use the Fixture point theorem (Wongsri, 2008). The fixture point theorem is provided by Wongsri, 2008 to determine the control variable that the most sensitivity. Dead time control variable should consider to control and pairing with manipulate variable (MV) in the first.

Fixture point theorem analysis

1. Consideration in dynamic mode of simulation until process set up to steady state.
2. Control variable (CV) can be arranged to follow the most sensibility of the process variable by step change the MV (change only one MV, the other should be fixed then alternate to other until complete). Study the magnitude of integral absolute error (IAE) of all process variables that deviates from steady state.
3. Consider CV that give the most deviation from steady state (high IAE score) to match with MV. CV and MV should be directing interactive together, after that will consider the next CV to match with other MV.

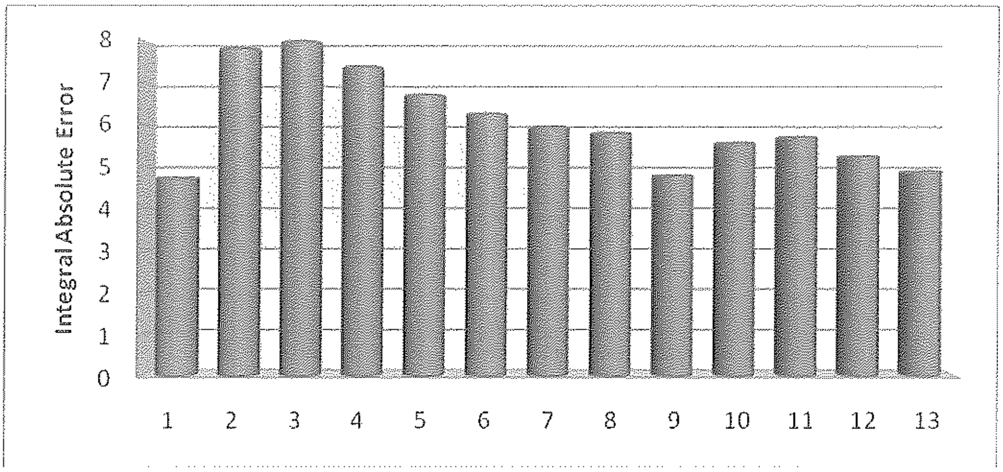

Figure 5.1 IAE summation result of tray temperature deviation of DB column

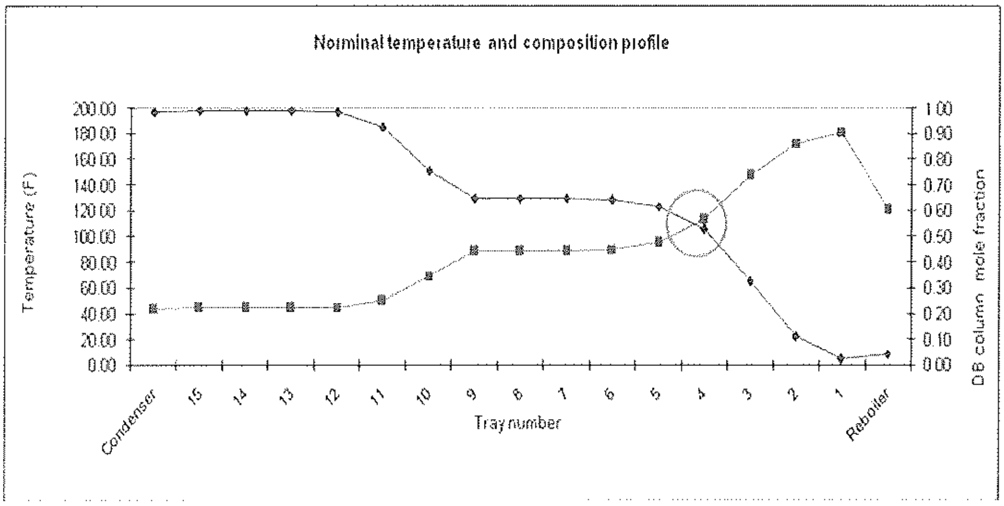

Figure 5.2 Temperature gradient of DB column at steady state

Figure 5.1 show the IAE summation results of Tray temperature of DB column. The temperature at tray 3th had a high value of IAE which shown that this tray is the most sensitive in the column so, we use this tray for temperature control.

Figure 5.2 shows the temperature gradient for DB column from steady state value

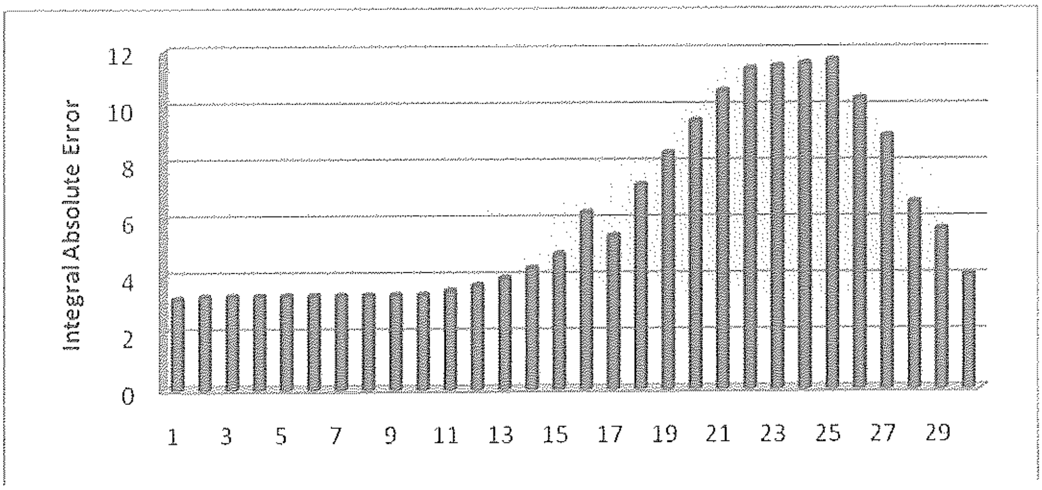

Figure 5.3 IAE summation results of Tray temperature deviation of DP column

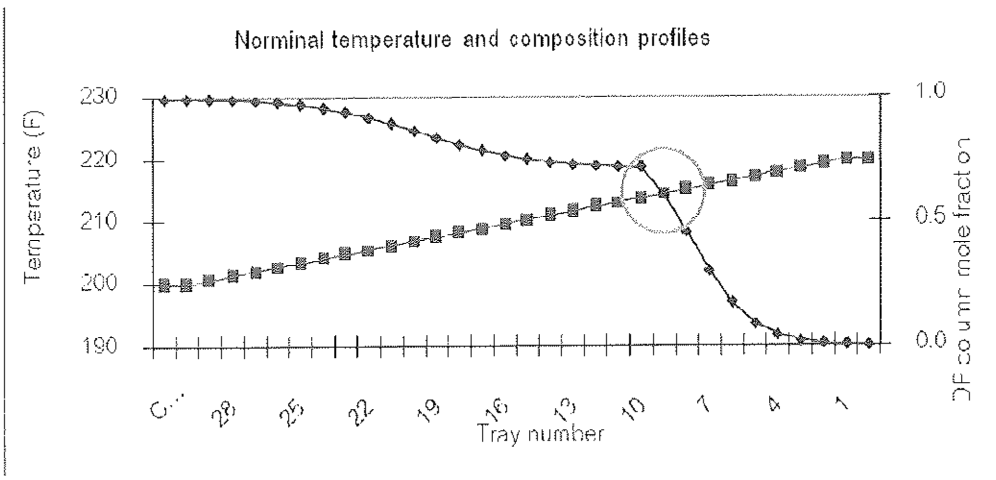

Figure 5.4 Temperature Gradient of DP column at steady state

Figure 5.3 show the IAE summation results of Tray temperature of DP column. The temperature at tray 25th had a high value of IAE which shown that this tray is the most sensitive in the column so, we use this tray for temperature control.

Figure 5.4 shows the temperature gradient for DP column from steady state value

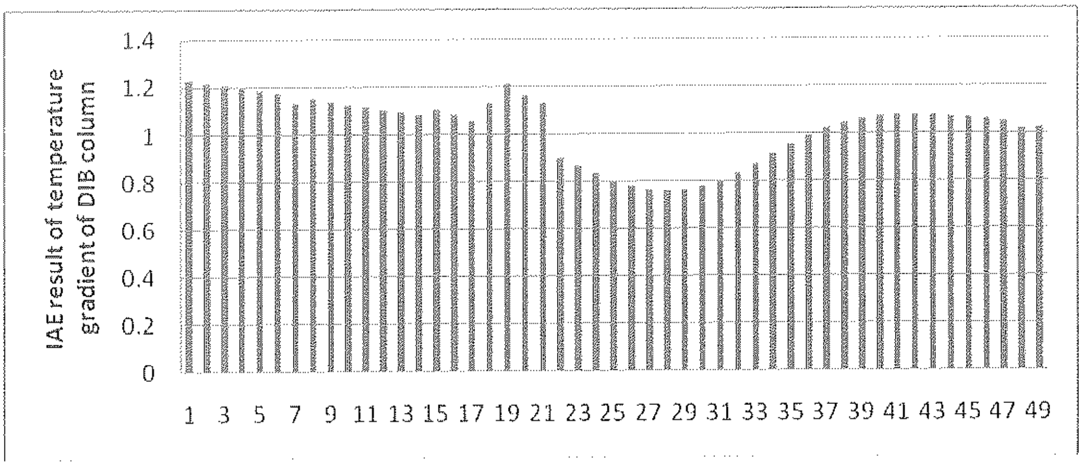

Figure 5.5 IAE summation result of tray temperature deviation of DIB column

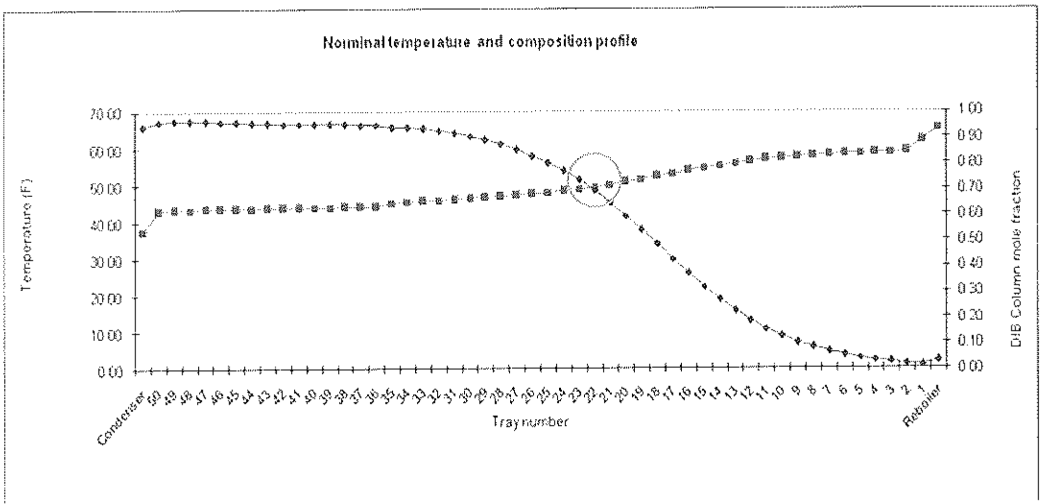

Figure 5.6 Temperature Gradient for DIB column from steady state value

Figure 5.5 show the IAE summation results of Tray temperature of DIB column. The temperature at tray 19th had a high value of IAE which shown that this tray is the most sensitive in the column so, we use this tray for temperature control.

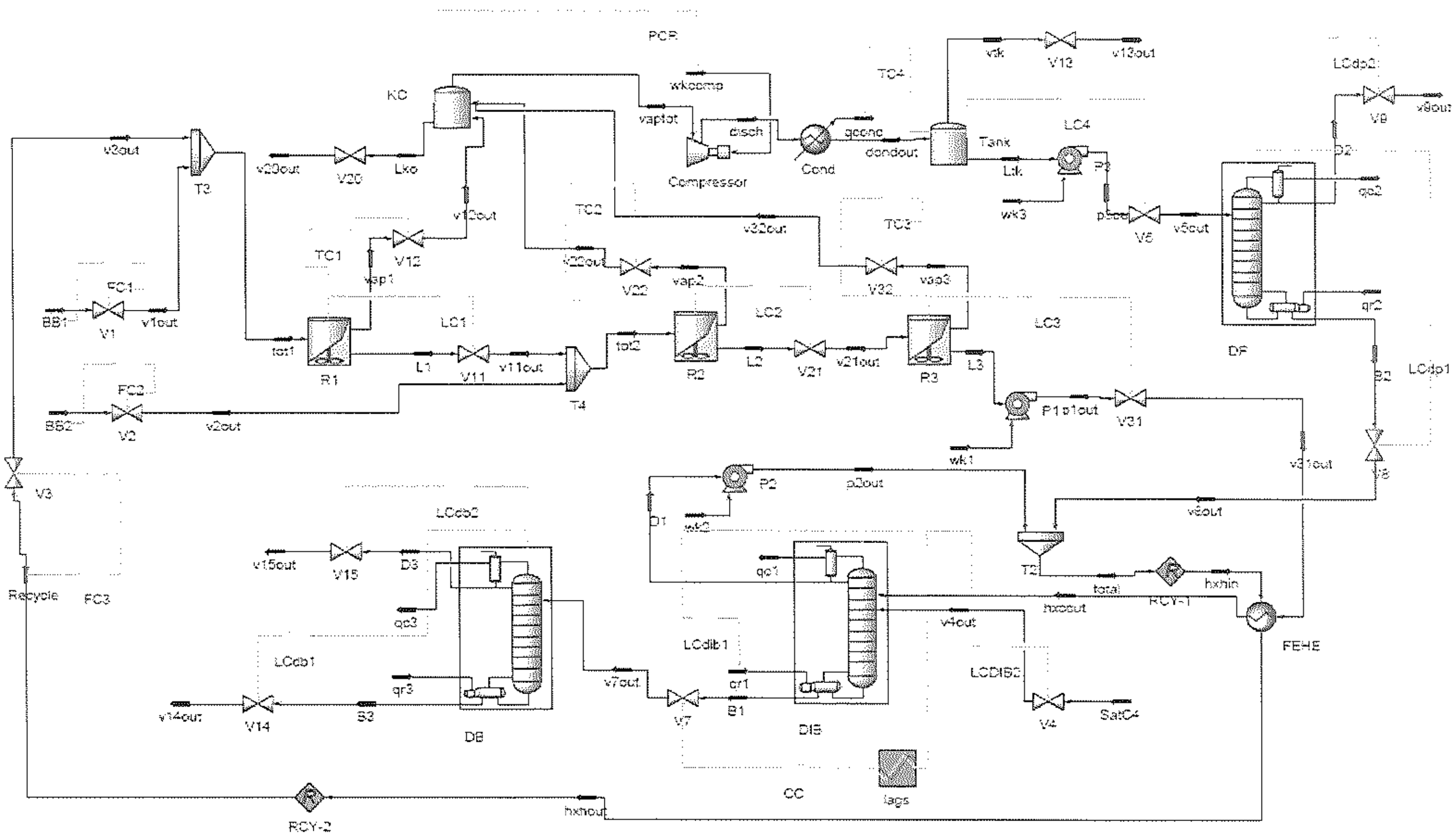

Figure 5.7 Base case control structure (CS0) for Isobutane process, Luyben (2002)

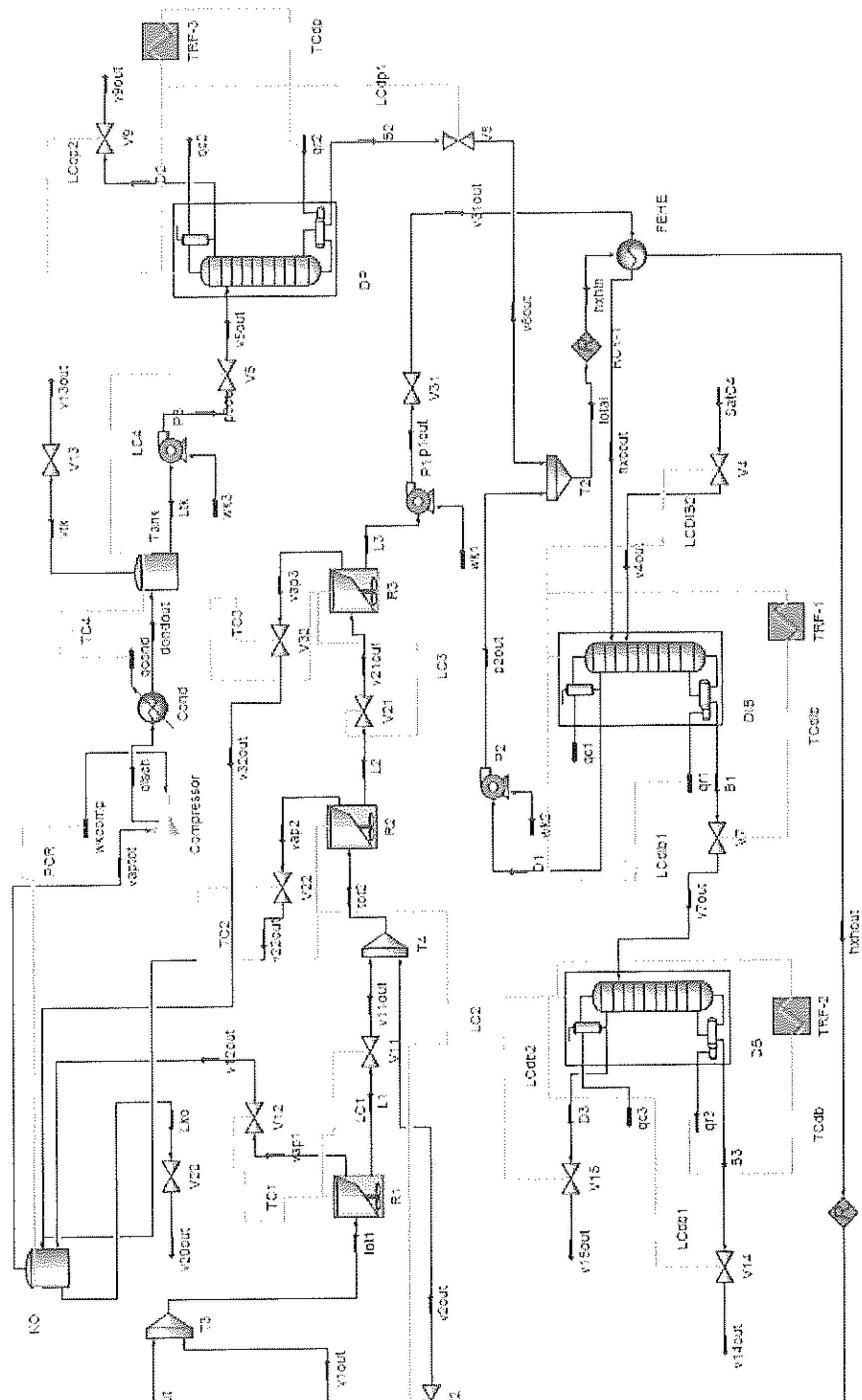

Figure 5.8 Designed control structure I(CS1) for Isobutane process

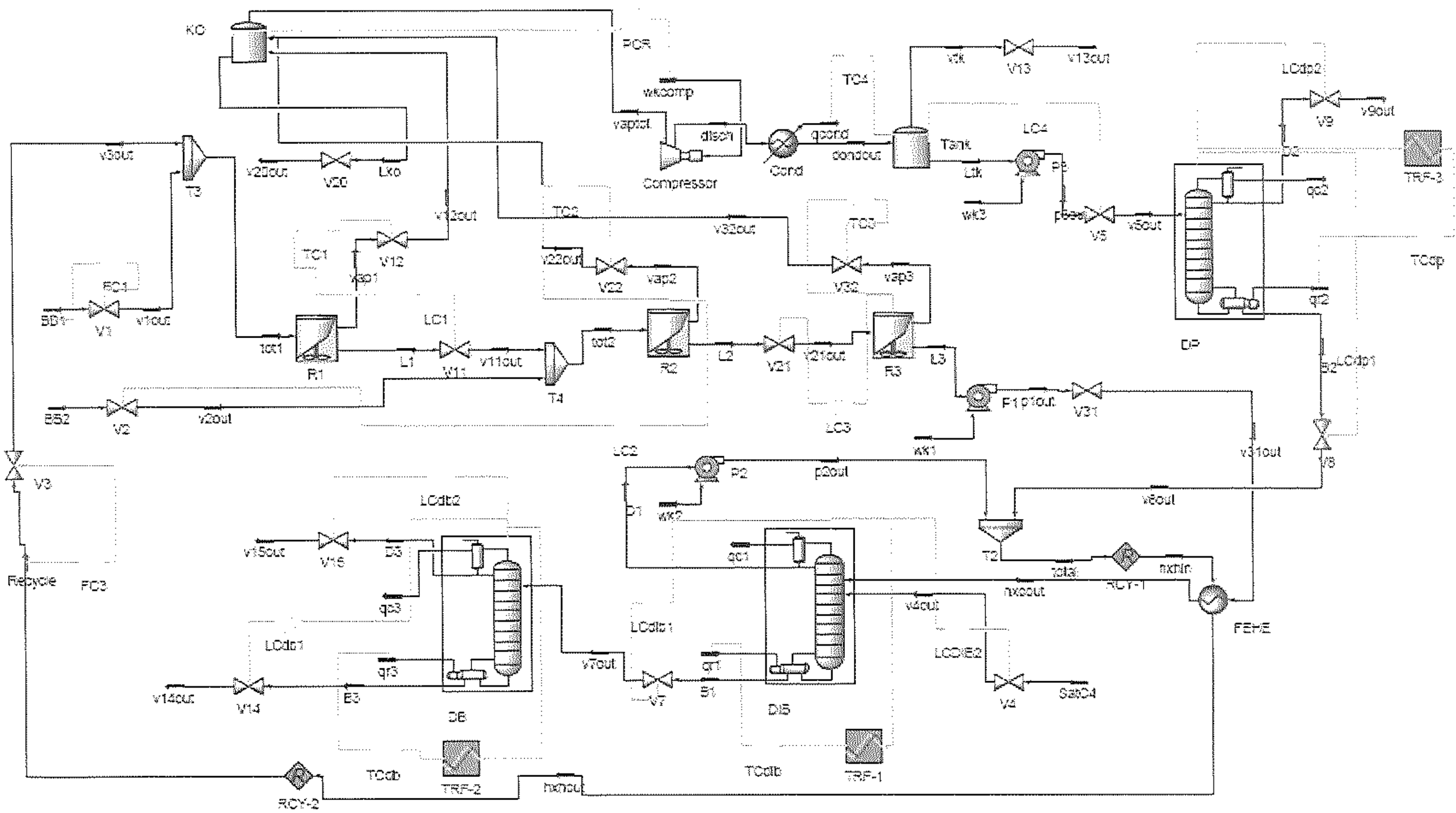

Figure 5.9 Designed control structure II(CS2) for Isobutane process

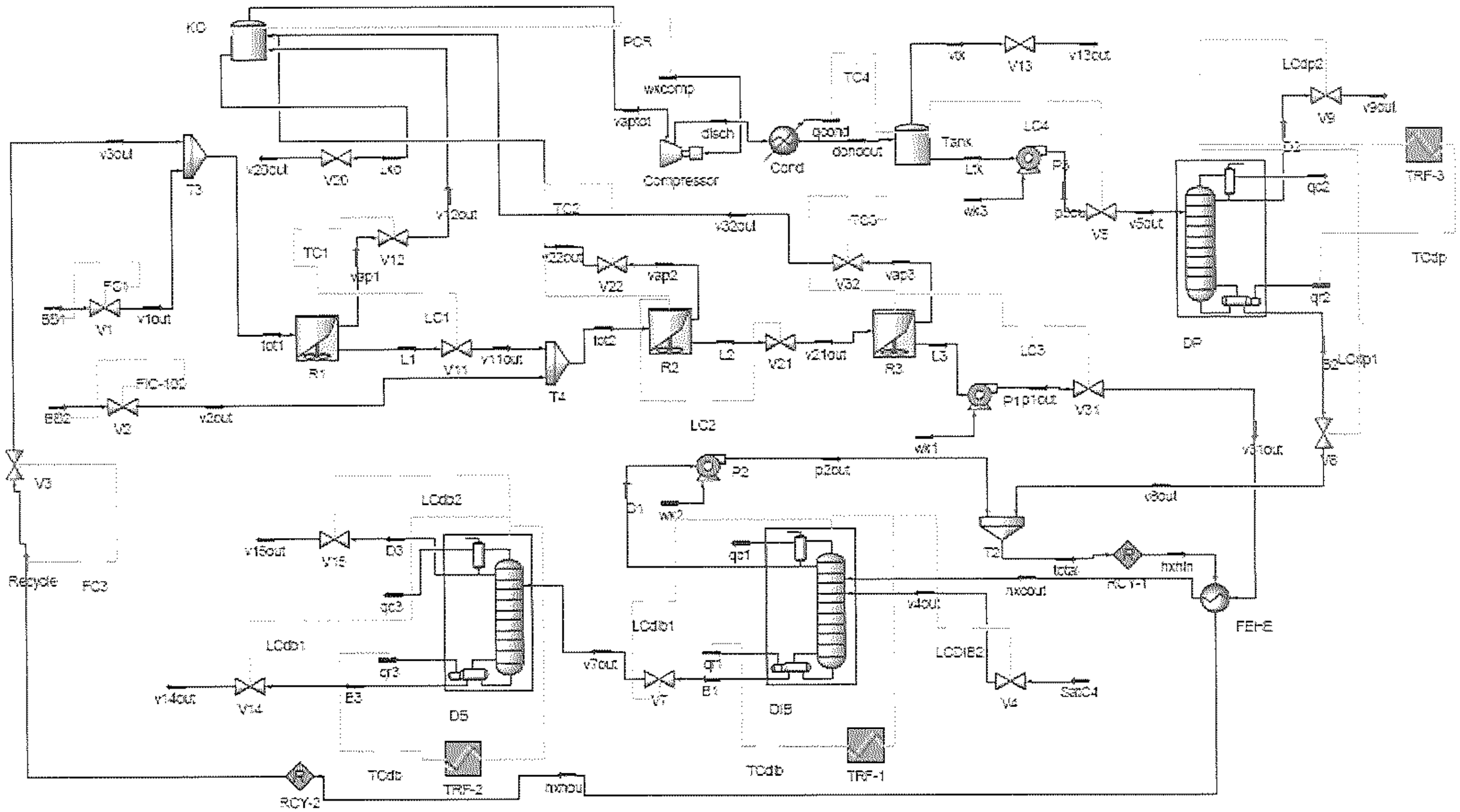

Figure 5.10 Designed control structure III (CS3) for Isobutane process

In all of these control structures (CSO, CSI, CS2, CS3) the difference loops are use as follows

Base case control structure (CS0)

For Isobutane process new 3 alternative structures, we present 3 control structures to reduce the effects of disturbances in order to achieve a desired production flow rate. The first control scheme to flow control of butene flow rate to reactor 2 and control level of reactor 3 revert by manipulated valve 21by inlet stream, the second was modified from the first scheme by change the level control of DIB distillation by manipulated V 7 and composition control by reboler duty (qr1). The last one keep the same structure as Luben but change change the level control of DIB distillation by manipulated V 7 and composition control by reboler duty (qr1) instead.

Design of control structure (CS1)

Considered IAE summation for flow variation, we found that there were a high disturbance in the reaction section which we will take in to consider and determine the effect of disturbance to the inlet and outlet of reactor effluent stream in order to prove the wording of Luben about the "the relative change in reaction rate depend on temperature through the activation energy", therefore we choose the reactor inlet temperature setpoint as the production rate handle for this case.

Therefore we have changed the valve V21 to control level revert of reactor 3 in order to maximize the production rate and valve V2 to control revert level of reactor 2 and to control the flow of BB2 to not exceed than maximum required since, excess in BB2 might lead the side reaction of dodecane which unsatisfied of process. and V7 to control Level of DIB column

For de-isobutanner (DIB) column, the 19th Tray iso-butane composition of de-Isobutanizer(DIB) column is control by manipulate of V7. The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1).

Design of control structure (CS2)

For control structure CS2, we always based on control structure design which would like to evaluate the production rate of Isobutane but we have change a bit in level control of DIB column to manipulated control by V7 instead and composition control by manipulated control by reboiler duty (qr1).

Design of control structure (CS3)

For control structure 3, we based on Luben control structure design but just changed in level control of DIB column to manipulated control by V7 instead and composition control by manipulated control by reboiler duty (qr1).

5.4 Dynamic Simulation Results

Comparison dynamic responses between this work with reference

The disturbance testing is used to compare the dynamic response of this simulation (HYSYS) with with 3 new structure design control and Luyben based case (1998).

By step change in BB1 fresh feed which increase ±10% of BB1 fresh feed from 24.51 lb-mole/hr to 25.98 lb-mole/hr at 50 min to 150 min and from 25.98 lb-mole/hr to 23.03 lb-mole/hr at 150 min to 250 min and step change of fresh feed temperature ±10% of BB1 fresh feed from 90 °F to 94 °F at 50 min to 150 min and from 94 °F to 86 °F at 150 min to 250 min.

Temperature controllers are PIDs which are tuned using relay feedback. Two temperature measurement lags of 0.5 minute are included in the two temperature loops (25th and 3th tray temperature of DP and DB Column accordingly). Flow and pressure controller are PIs and their parameters are heuristic values. Proportional only level controllers are used and their parameters are heuristics values. Isobutane mole composition is measured and controlled using PID controller. All control valves are half-open at nominal operating condition.

Figure show the dynamic responses of the control systems of the Isobutane base case to a change in the BB1 fresh feed molar flow. In order to make this disturbance, first the BB1 fresh feed flow rate is increased from 24.51 lb-mole/hr to 25.98 lb-mole/hr at time equal to 50 min to 150 min and the flow decreased from 25.98 lb-mole/hr to 23.03 lb-mole/hr at time equal to 150 min to 250 min, then the feed flow rate is returned to its nominal value of 24.51 lb-mole/hr

Dynamic response change in fresh feed molar flow rate by step change of molar flow rate±10% of BB1 fresh feed from 24.51 lb-mole/hr to 25.98 lb-mole/hr at 50 min to 150 min and from 25.98 lb-mole/hr to 23.03 lb-mole/hr at 150 min to 250 min

Effluent molar of fresh feed BB1 step change in ±10% of BB1 fresh feed

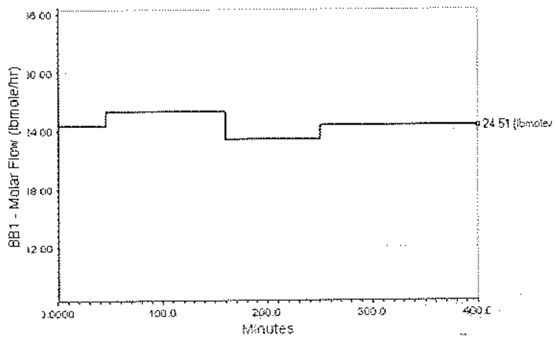

Figure 5.11 Dynamic responses of the Isobutane plant to a change in the BB1 fresh feed rate.

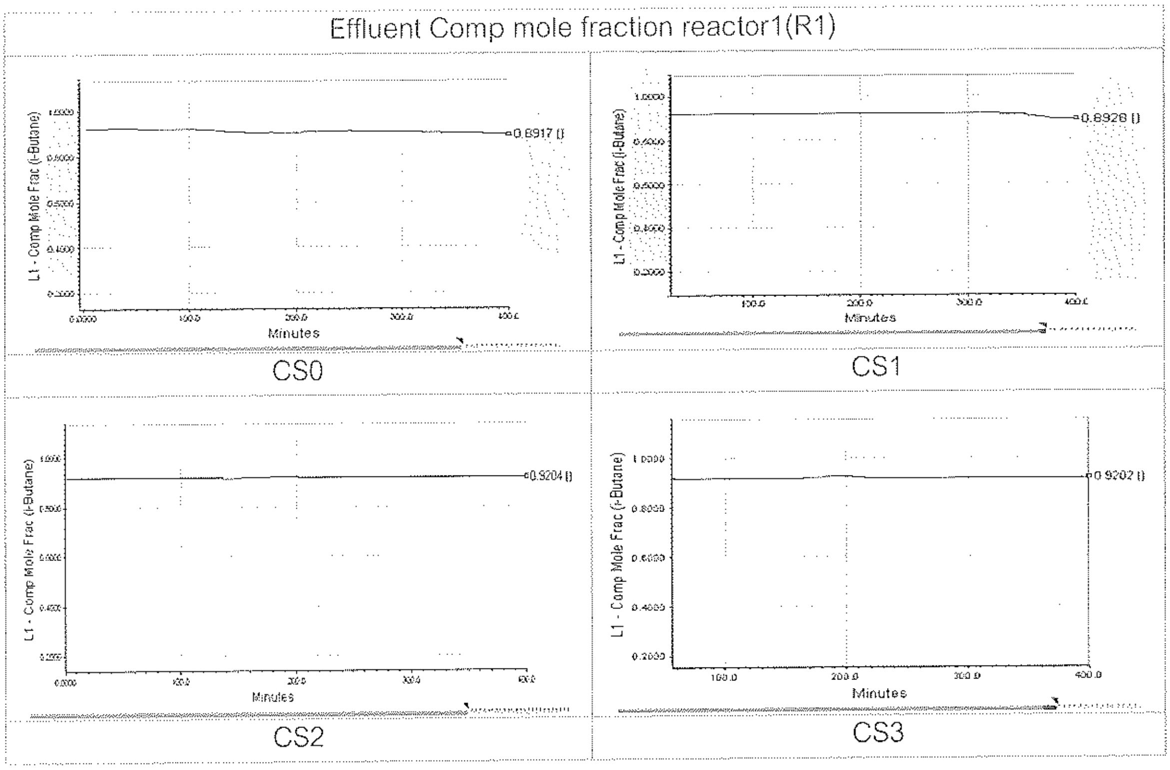

Figure 5.12 Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to control structure design CS1, CS2, CS3.

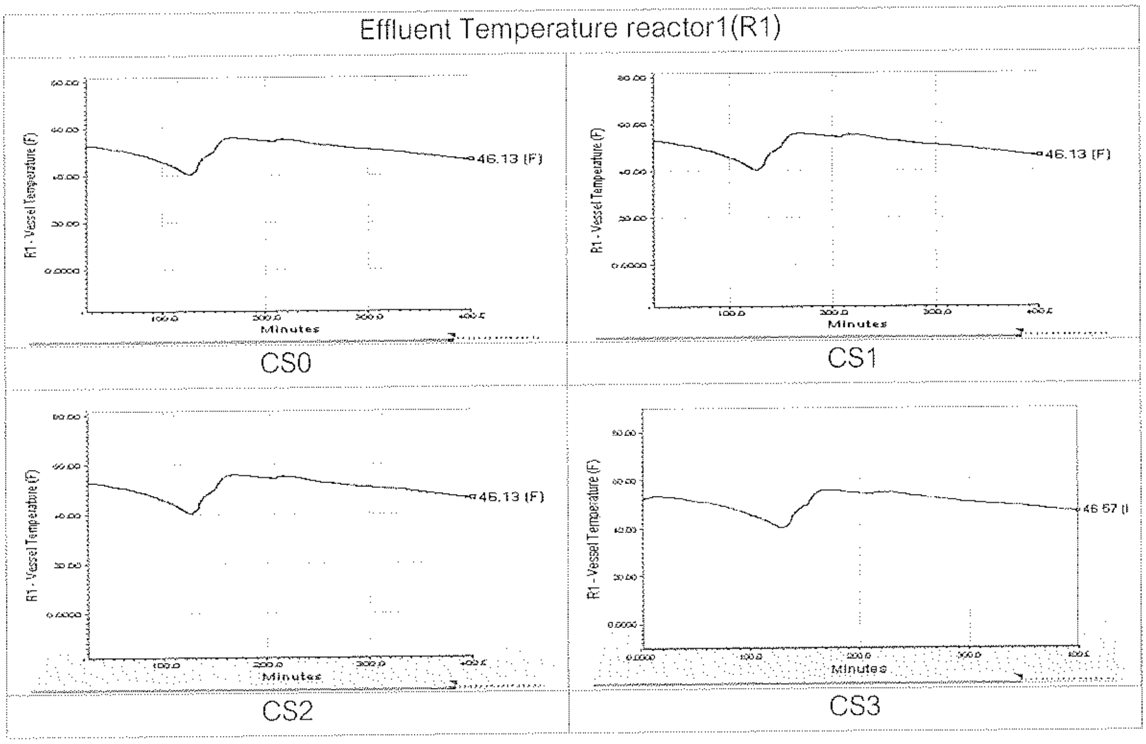

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

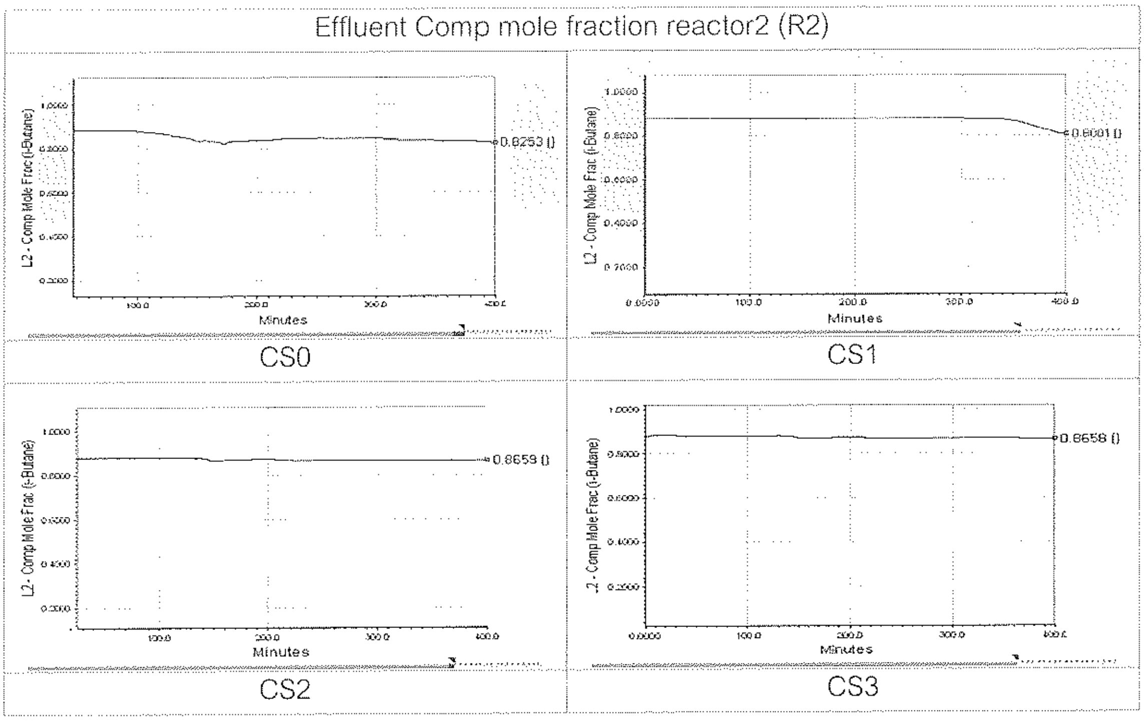

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

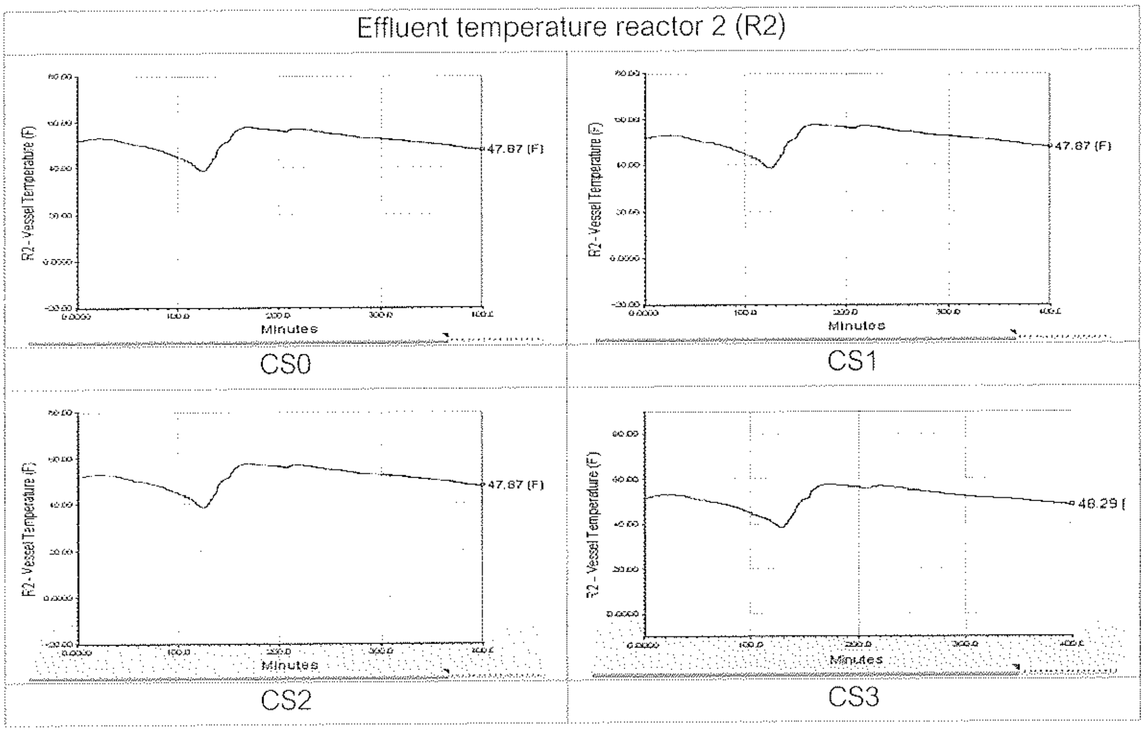

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

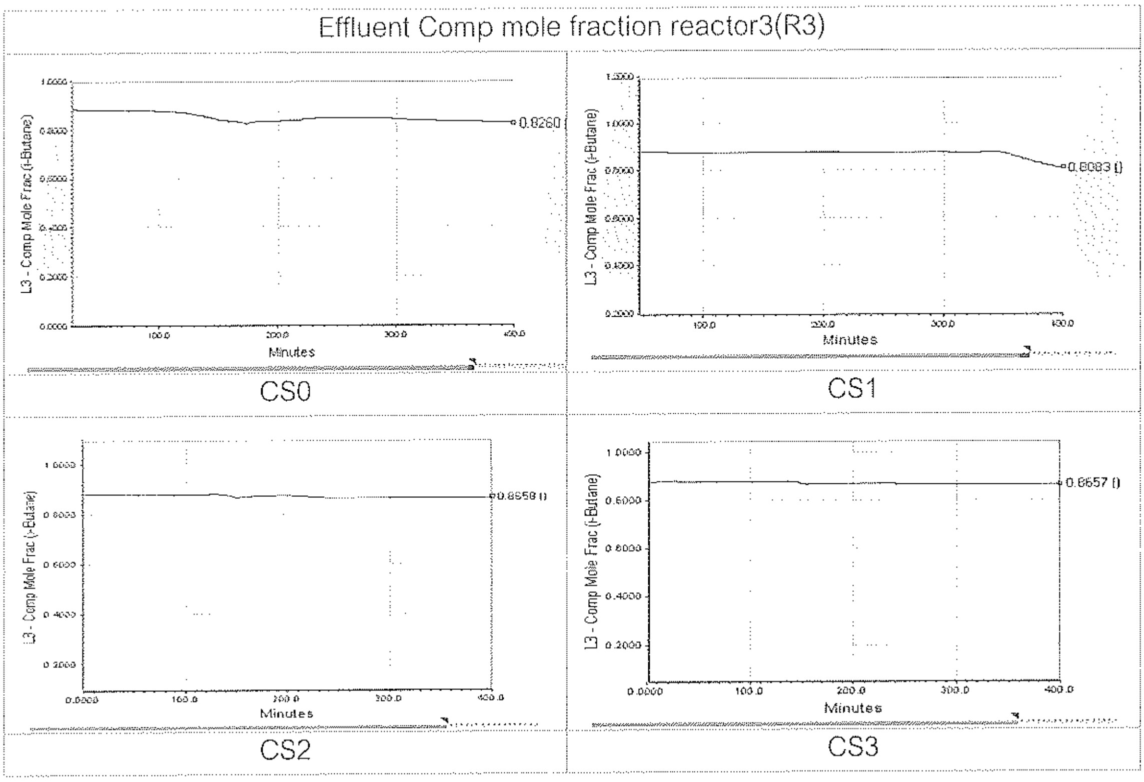

Figure 5.12 (Continued) Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of BB1 fresh feed.

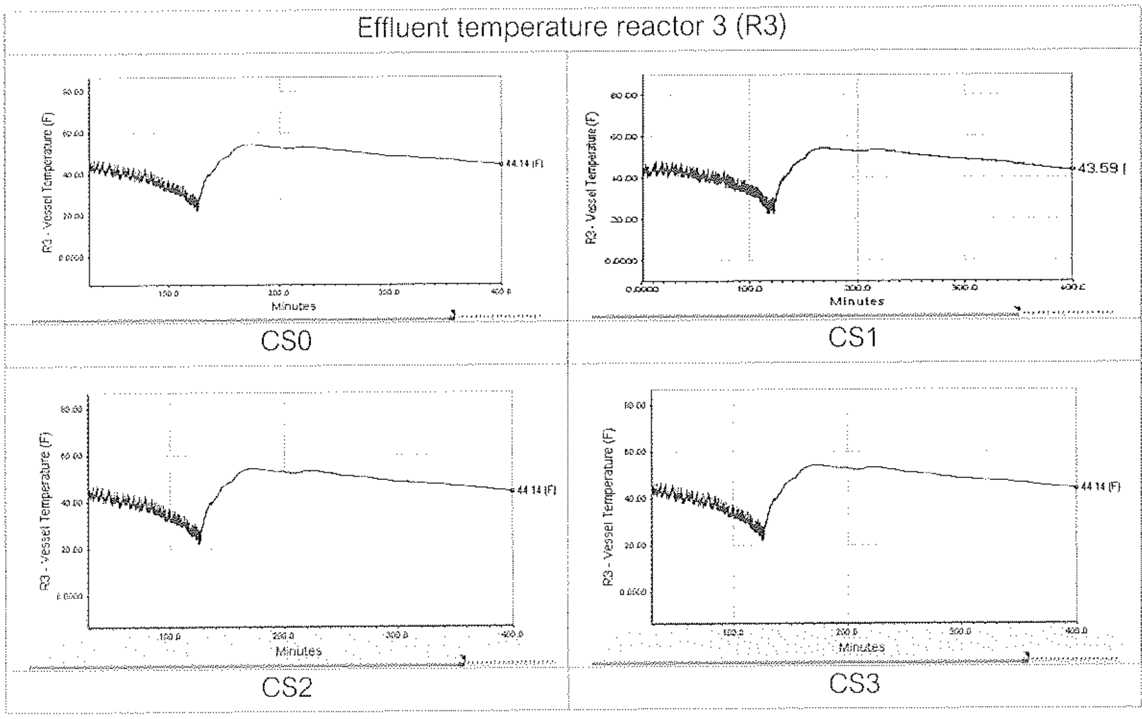

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

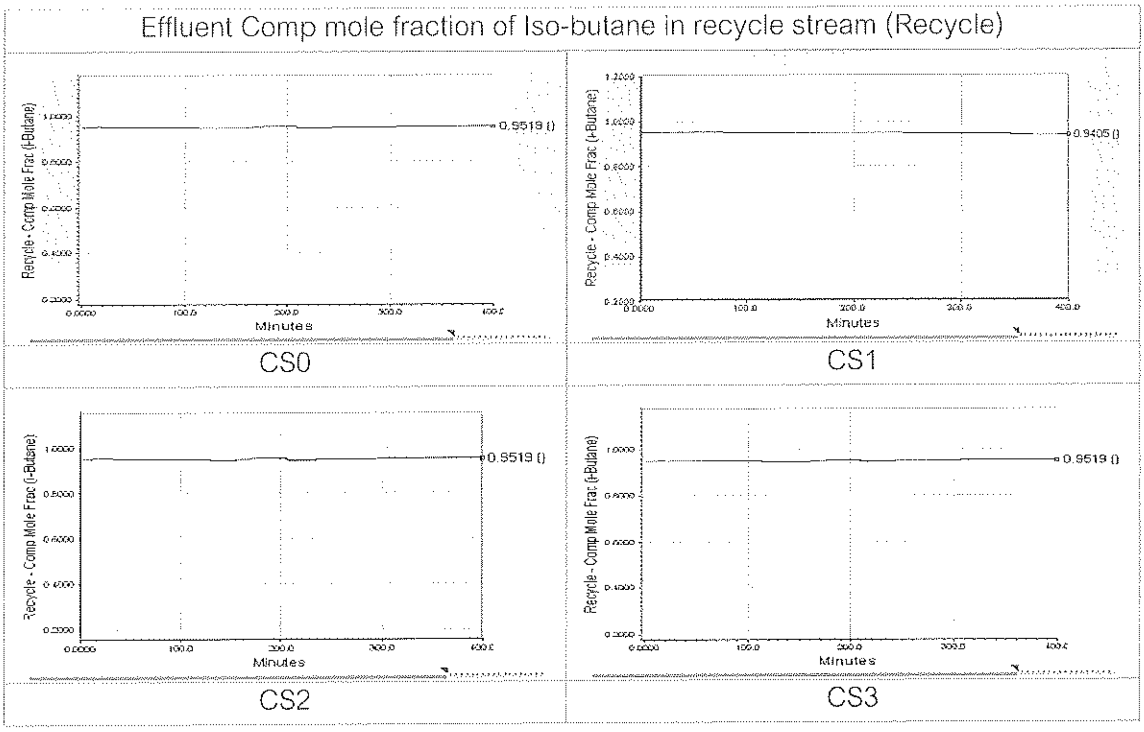

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

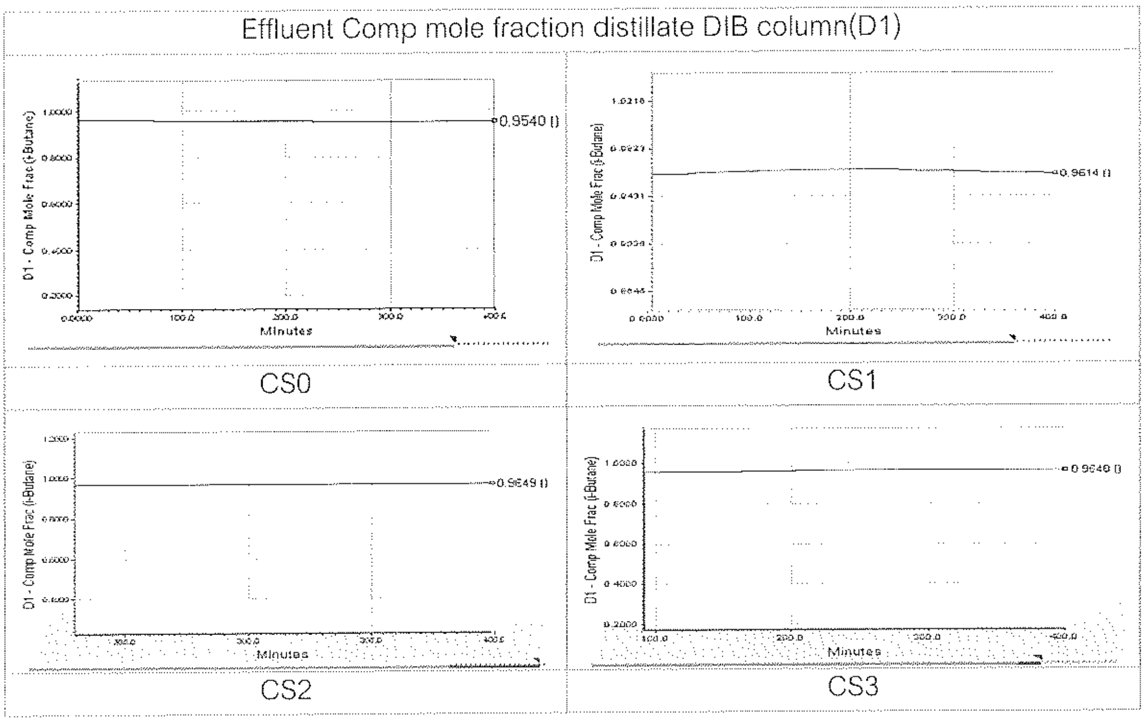

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

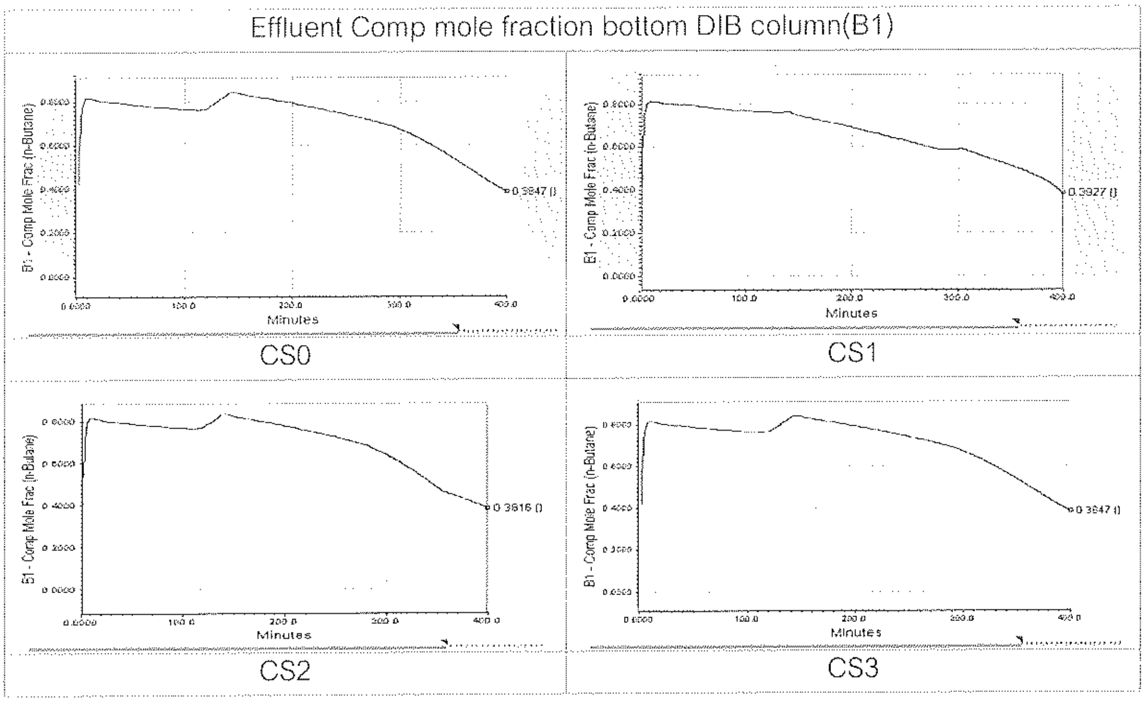

Figure 5.12 (Continued) Dynamic response of control structure (CS0, CS1,CS2 and CS3) process by step change of molar flow rate±10% of BB1 fresh feed, B1 is n-butane composition in bottom stream.

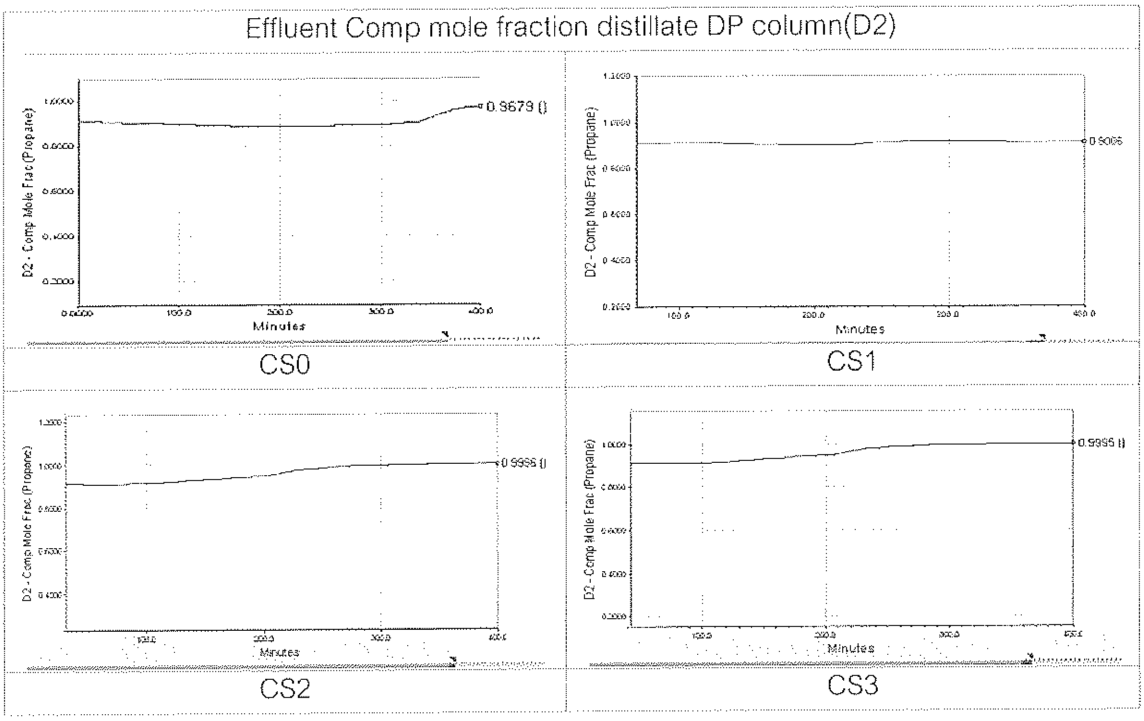

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

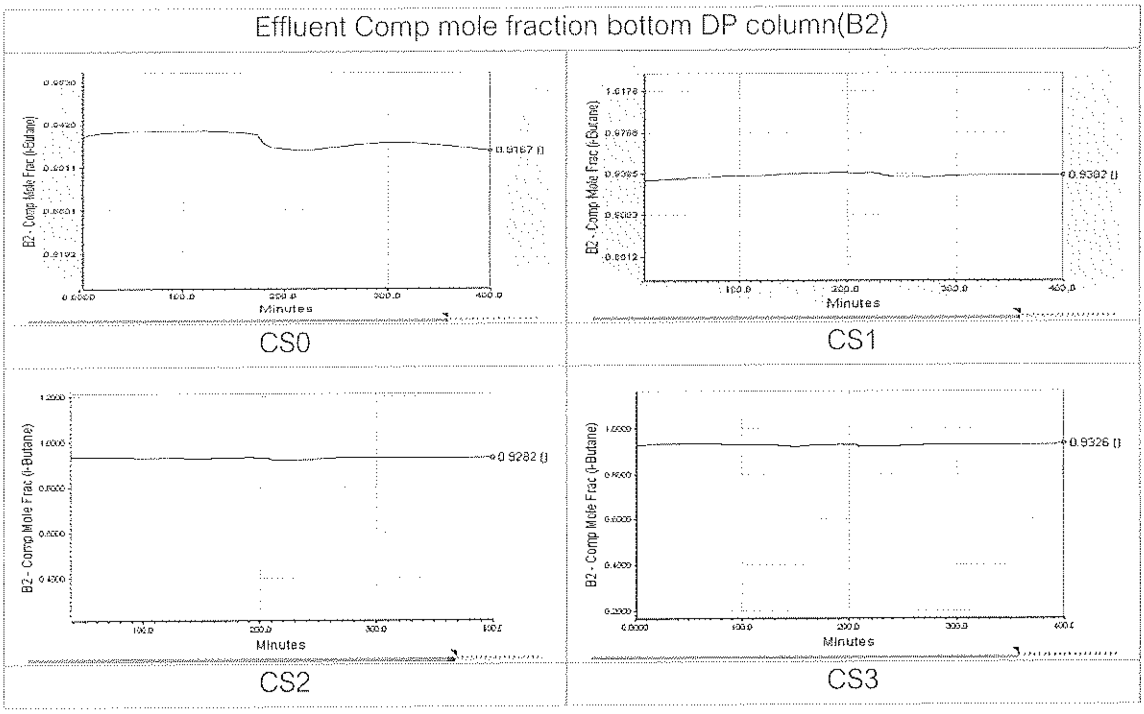

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

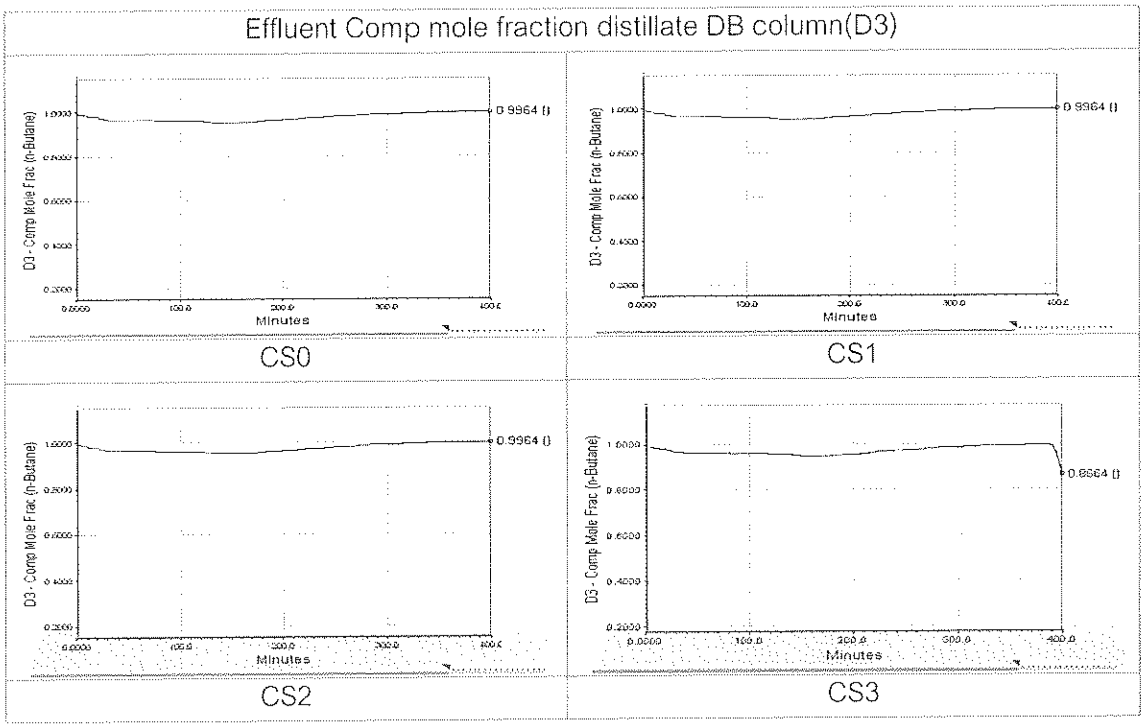

Figure 5.12 (Continued) Comparison dynamic responses of step change in BB1 flow rate (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3.

Figure 5.10a gives simulation result for increasing and decreasing (±10%) the BB1 fresh feed . The molar flow rate response is oscillatory and it comes to new setpoint within a few minutes. Effect from the step change in flow rate, the rate of reaction and mole fraction of Isobutane in reactor 1, 2, 3 and DIB distillate increase from the base case but not much difference in temperature change.

5.3.2 Figure show the dynamic responses of the control systems of the Isobutane base case to a change in the BB1 (±10%) fresh feed temperature. In order to make this disturbance, first the BB1 fresh feed flow rate is increased from 90 °F to 94 °F at time equal to 50 min to 150 min and the flow decreased from 94 °F to 86 °F at time equal to 150 min to 250 min, then the feed flow rate is returned to its nominal value of 90 °F

Step change of ±10% of BB1 fresh feed temperature to all control structures :base case control structure (CS0), control structure I (CS1) to control structure III (CS3)

The result of dynamic response of the Step change of ±10% of BB1 fresh feed temperature from 90 °F to 94 °F at 50 min to 150 min and from 94 °F to 86 °F at 150 min to 250 min

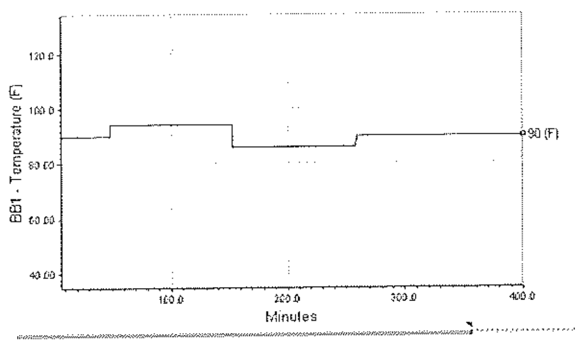

Figure 5.13 Dynamic responses of Isobutane process by step change of ±10% of BB1 fresh feed temperature

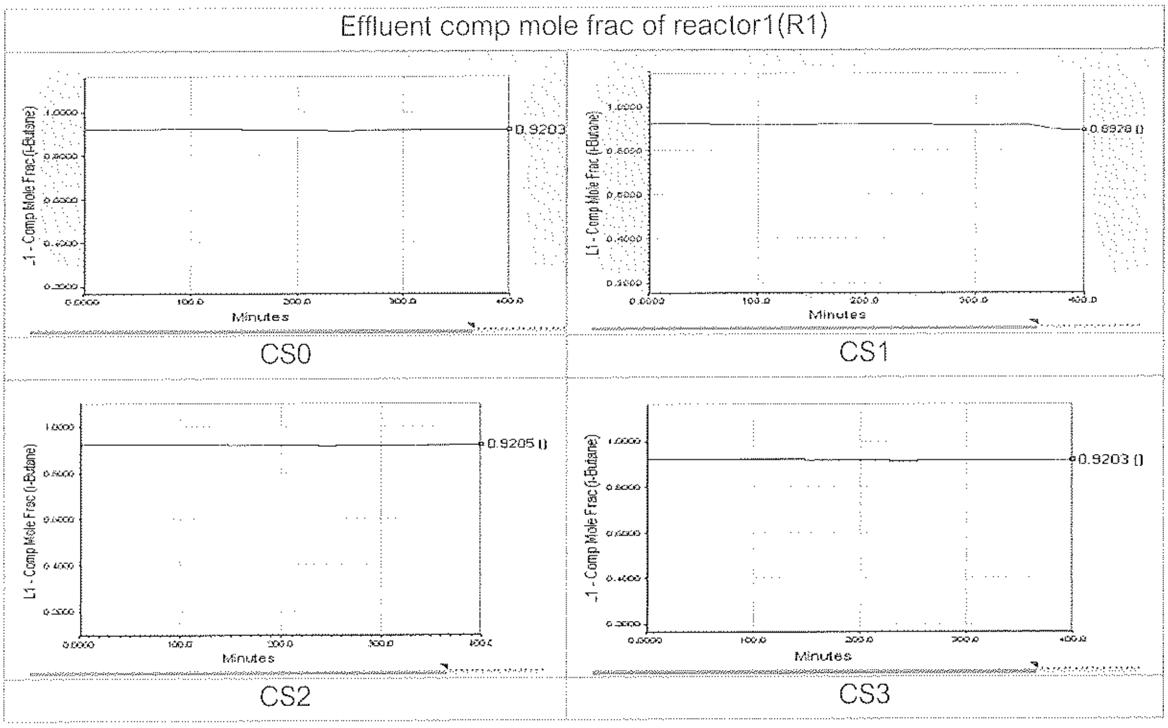

Figure 5.14 Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

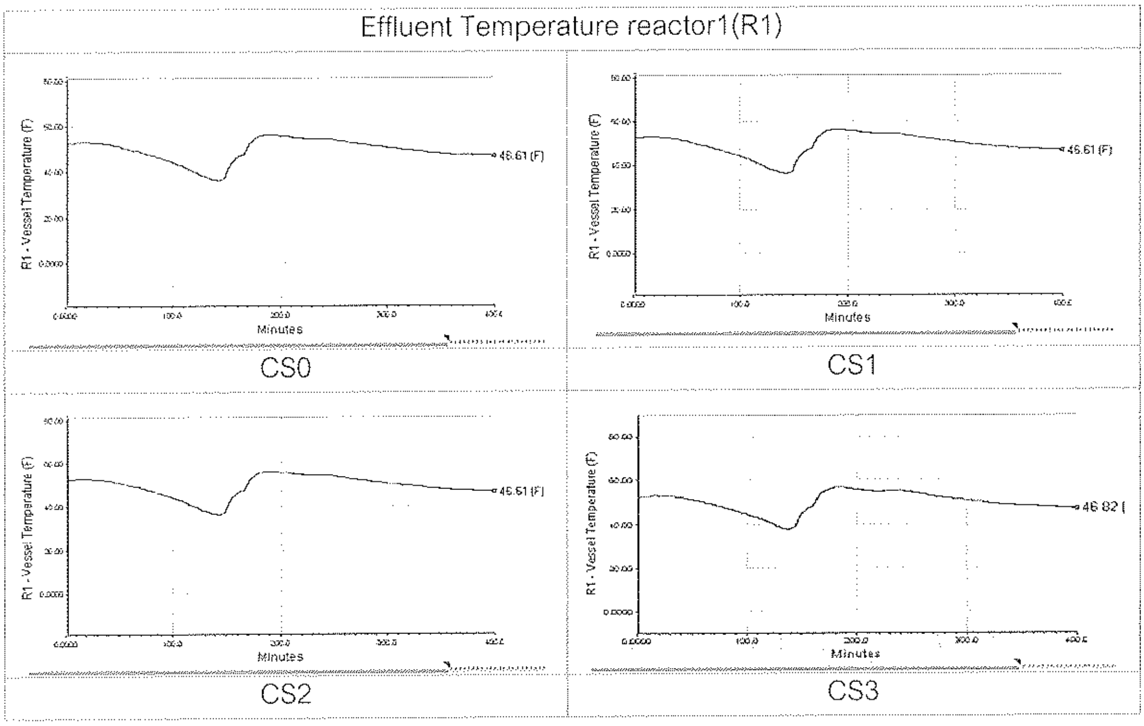

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

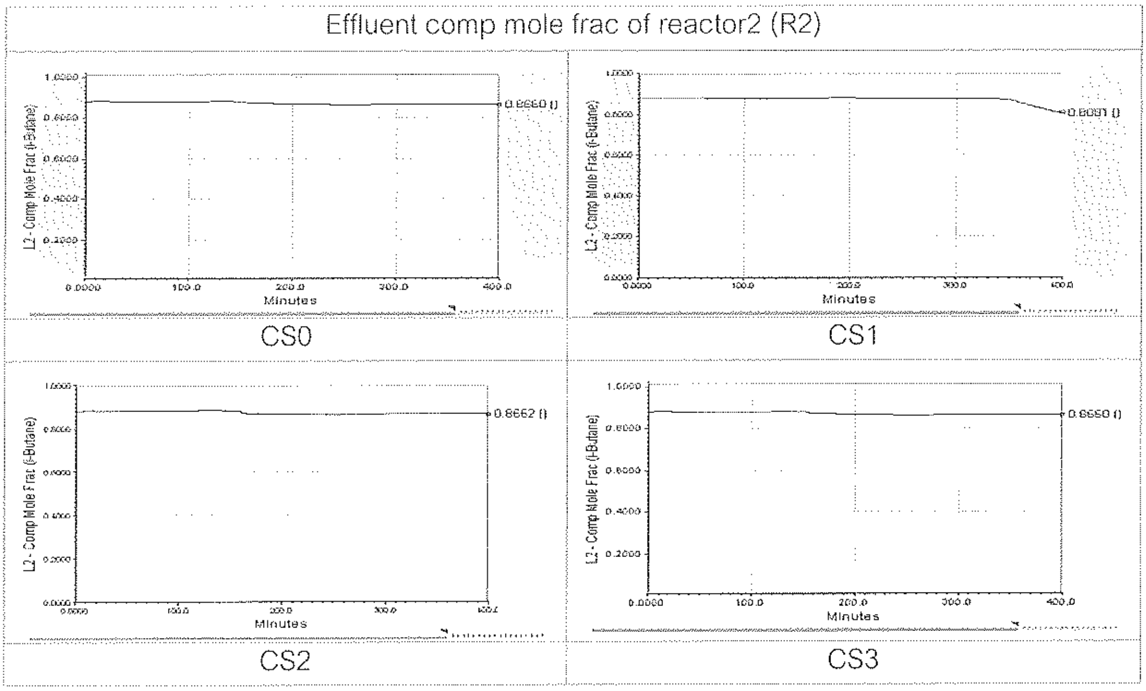

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

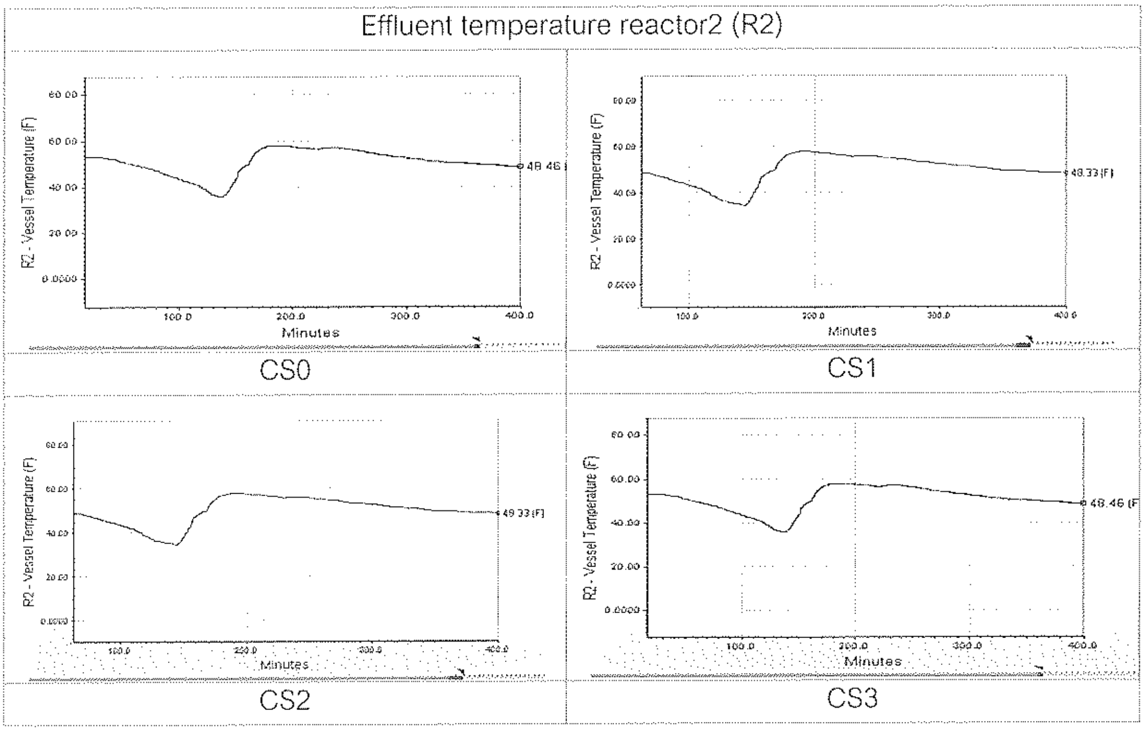

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

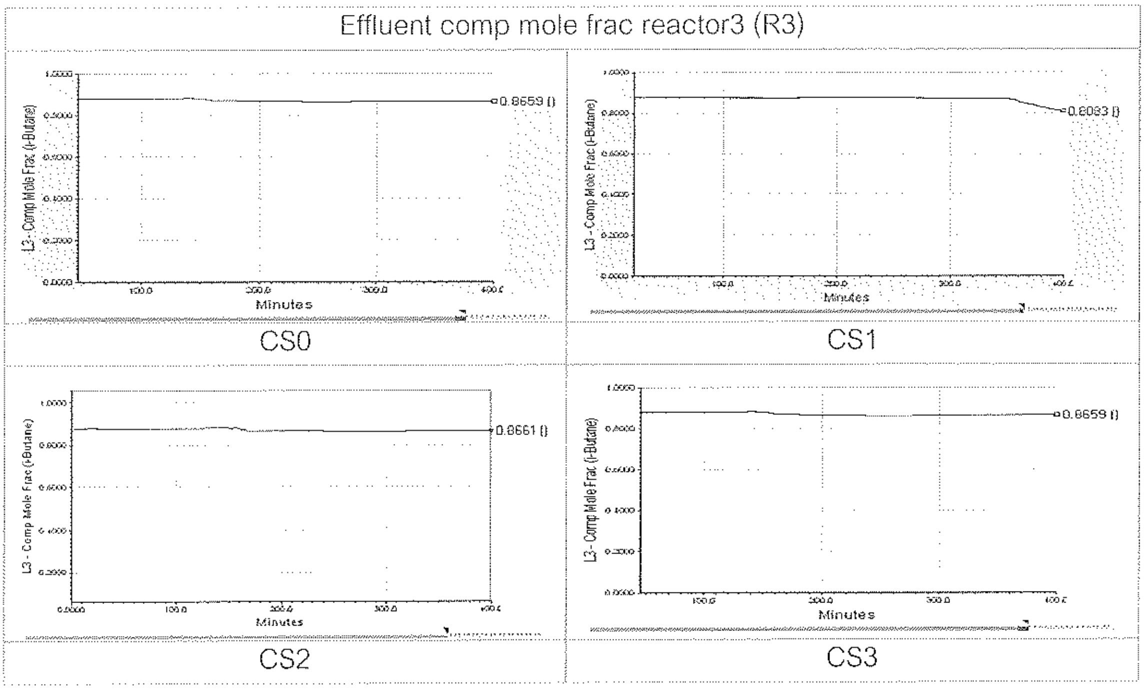

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

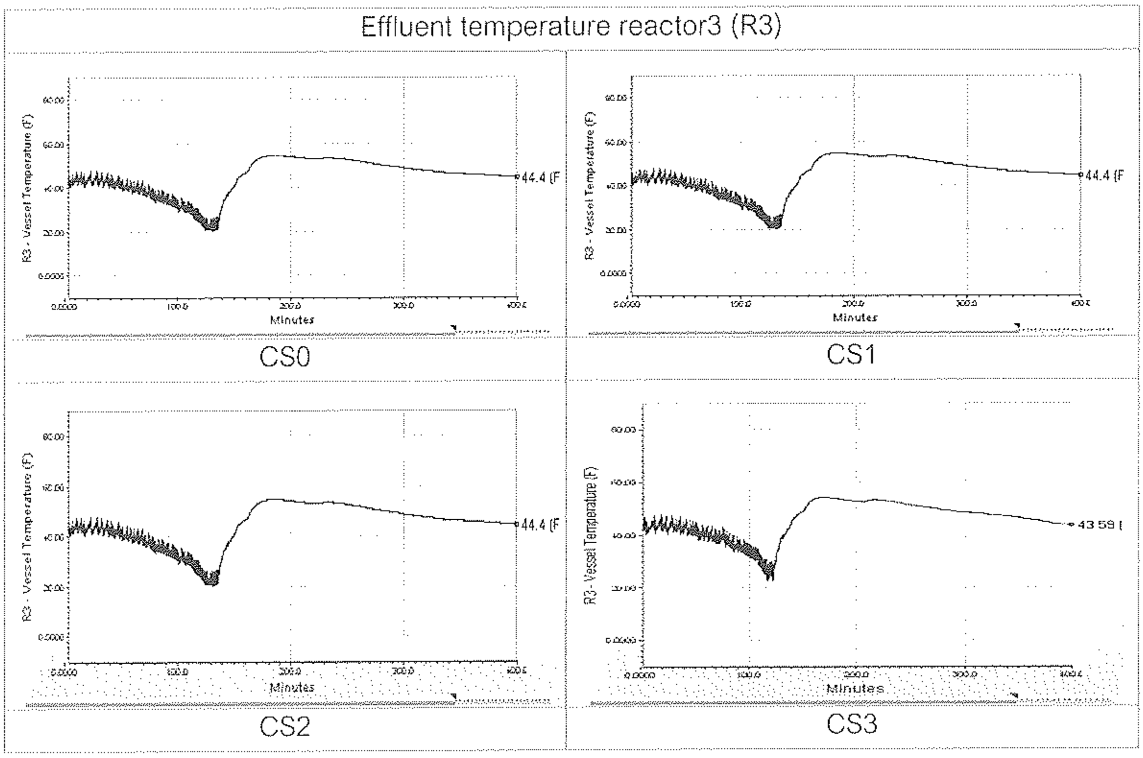

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

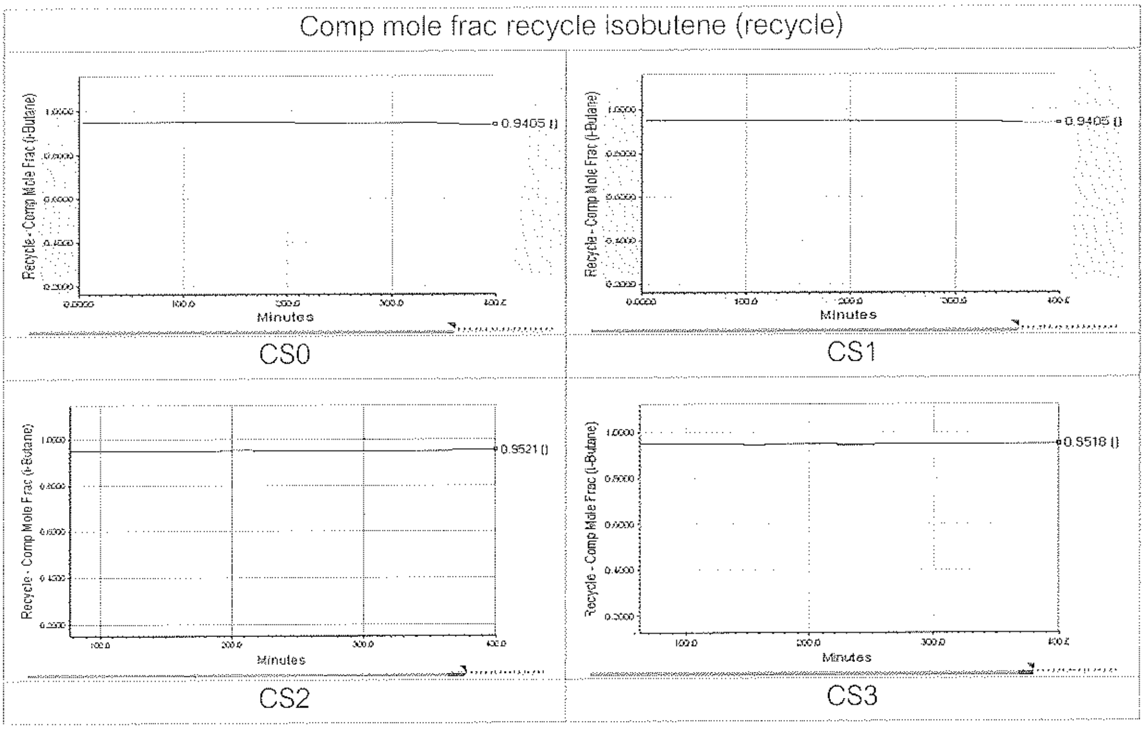

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

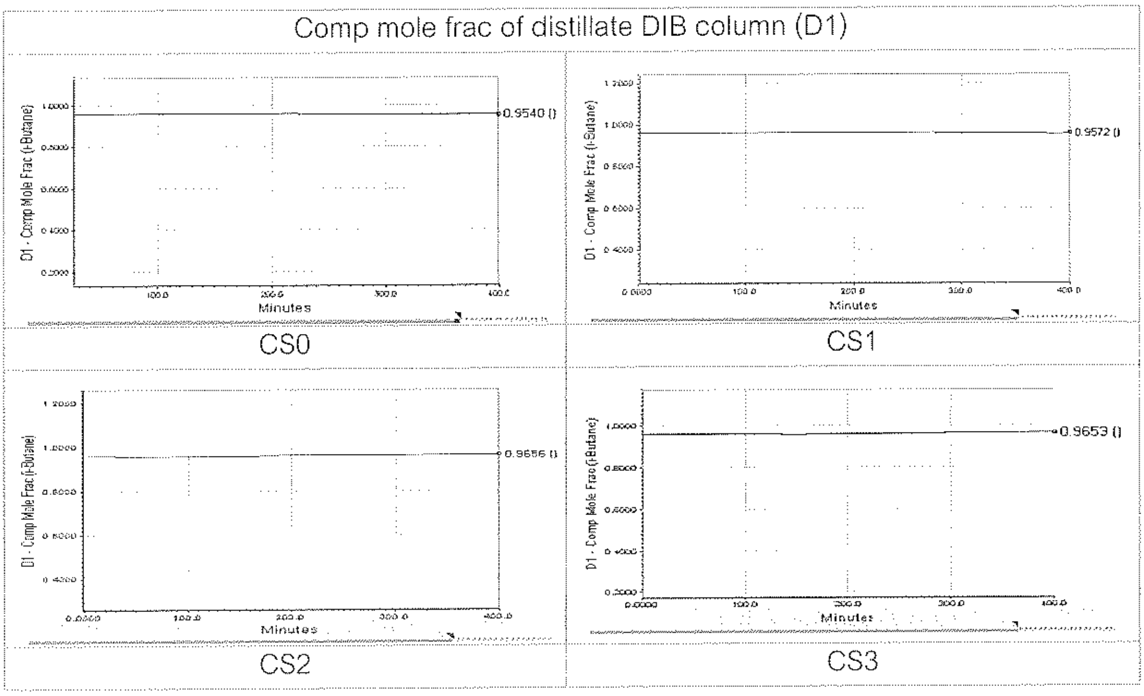

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

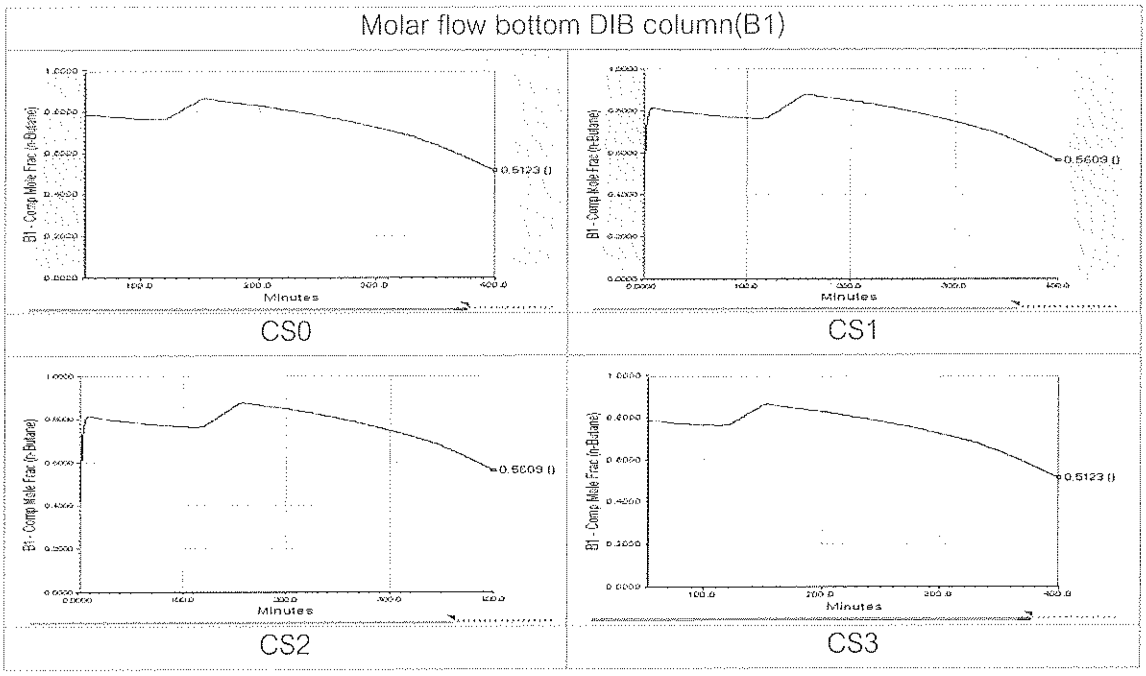

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

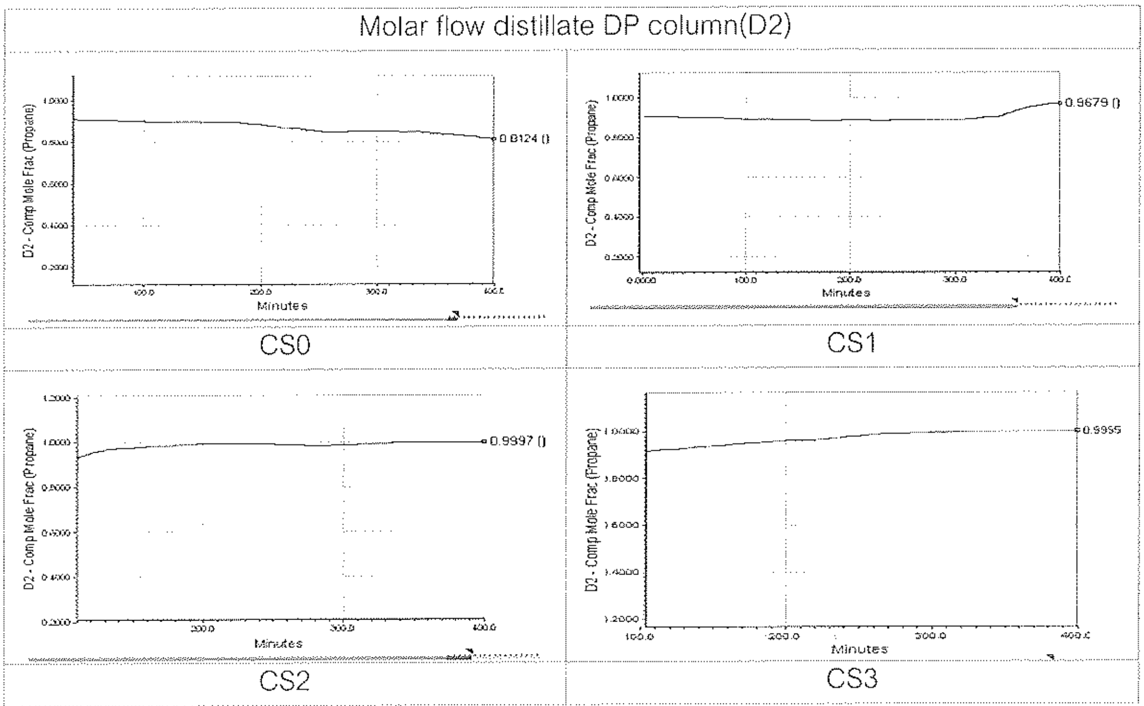

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

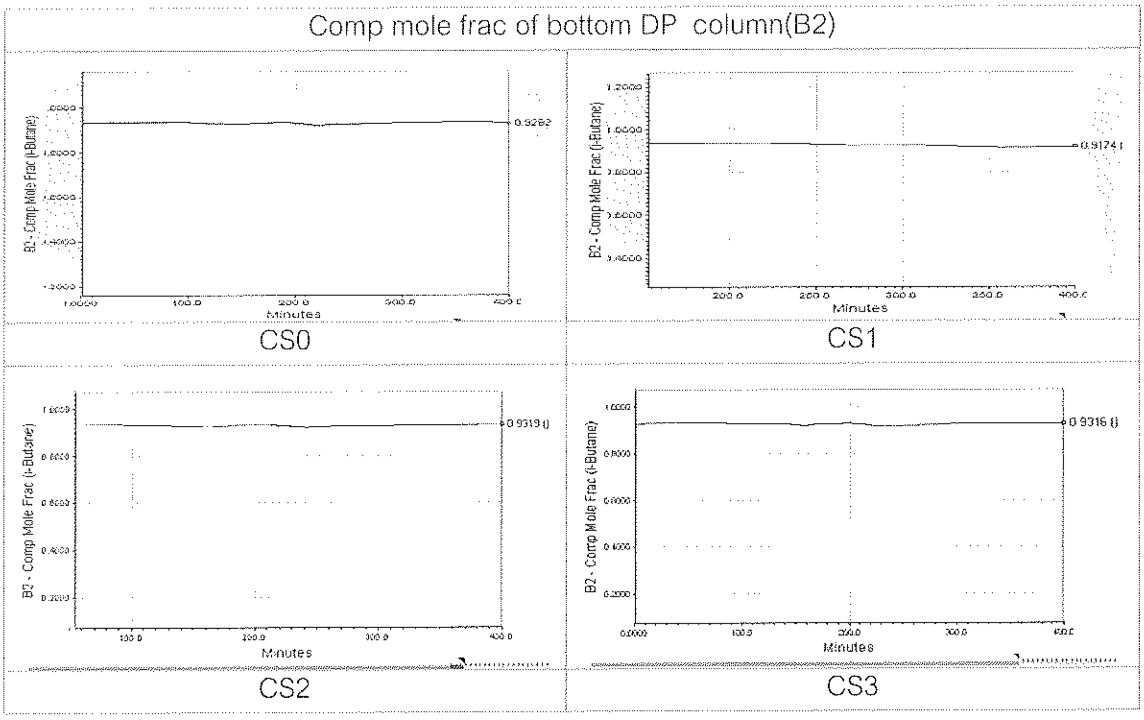

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

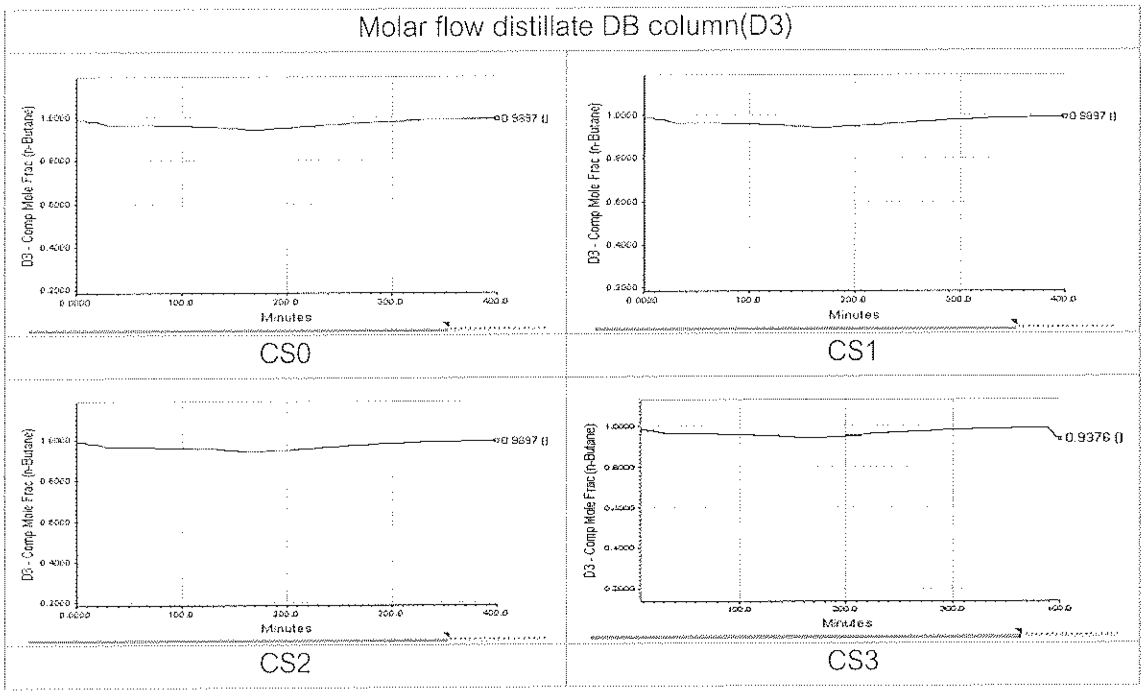

Figure 5.14 (Continued) Comparison dynamic responses for step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

Figure 5.14 simulation results for increasing and decreasing by step change of ±10% of BB1 fresh feed temperature between CS0 of Luben based case compare to new control structure CS1-CS3.

. The result of dynamic responses of temperature is oscillatory and its come to the new setpoint within a few minutes. Effect from the temperature increase, the rate of reaction of Isobutane increase because of an increasing temperature can raise the reaction rate. But this control structure fix production rate, so this disturbance does not affect to the product flow.

The IAE summation results of control structure, CS1-CS3 compare to CS0 of Luyben based cased about how to handle the disturbance when fresh feed flow rate and temperature have been changed by ±10%, it shown that the control structure 2(CS2) is the best control structure to handle disturbance and CS2, also to maintain the product quality and minimize energy consumption.

Table 5.2 The IAE results of the 4 control structures to a change of ±10%BB1 fresh feed flow rates

	Molar flow rate changes ±10% of BB1 fresh feed			
	CS0	CS1	CS2	CS3
LC1	7.339737	6.97935651	8.10686043	6.85478924
LC2	19.71125808	10.19314076	12.30553245	10.19268933
LC3	21.89328723	24.72751744	2.8642217	24.5441139
LC4	26.16481001	19.04538331	25.65637941	19.06375521
LCdp1	16.24507097	13.94496908	16.44566585	13.72057054
LCdp2	0.41924268	0.96346807	0.50240448	0.91831718
LCdib1	17.31079919	20.26097541	20.27488953	20.30739511
LCdib2	0.43416742	0.2988214	0.12070721	0.23623101
LCdb1	20.13525572	18.91310593	20.67962197	20.93836707
LCdb2	2.14201389	1.1295107	0.88370167	0.84007466
PCR	10.05902582	2.86393671	1.07828955	2.57258315
TC1	15.55549624	1.56463775	2.00655345	1.4842041
TC2	16.72252609	1.57522442	2.09163411	1.56462584
TC3	19.25484719	2.75823327	0.8981459	3.10893511
TC4	5.4062154	16.32853072	3.1555757	15.3838403
TCdb	0.0342988	0.087816651	0.001316644	0.020699359
TCdib	0.026551889	0	0.015329142	0
TCdp	0.000188679	0.02963562	1.66069E-05	0.033095001
Total	198.8547923	141.6642637	117.0868458	141.7842861

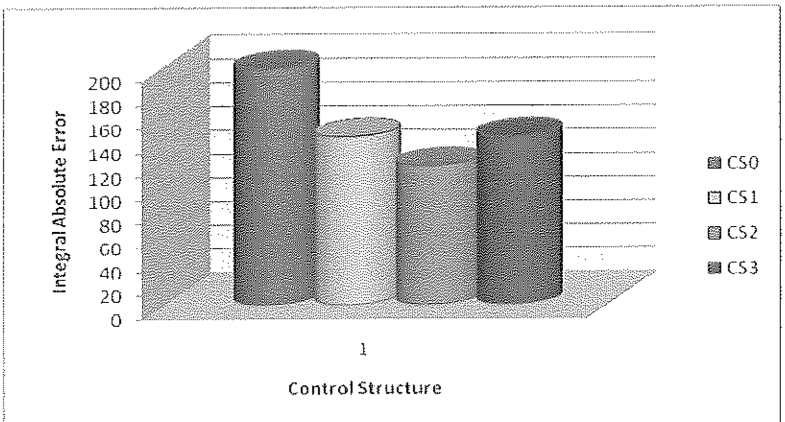

Figure 5.15 The IAE summation results of four control structures Isobutane process to response the change of ±10 of fresh feed flow rate.

Table 5.3 The IAE results of the 4 control structures to a change of ±10%BB1 fresh feed temperature

	Temperature changes ±10% of BB1 fresh feed			
	CS0	CS2	CS1	CS3
LC1	7.24841	6.86131669	8.035212	6.7513314
LC2	19.845733	10.23173795	12.436978	10.20131166
LC3	2.069813	23.98261246	2.744869	24.0667307
LC4	27.245678	18.55549399	23.06086142	18.86672915
LCdp1	14.3578	12.90726329	14.24650864	13.11959574
LCdp2	0.9927	0.89496761	0.4338	0.88281646
LCdib1	19.03878358	20.30901831	20.30658145	20.30878323
LCdib2	0.35798797	0.26839895	0.10031	0.24944873
LCdb1	20.3614	21.00171471	20.90585824	20.97368296
LCdb2	2.2902	0.78919566	0.6784	0.7869177
PCR	10.254168	3.009492459	1.247896	2.756692251
TC1	15.964212	1.97252251	2.478523	1.86983051
TC2	16.924687	2.0632844	2.574311	1.9689748
TC3	19.374847	3.43648677	0.9745022	3.3895632
TC4	2.6922607	15.79833299	3.041323	15.46679101
TCdb	0.0349052	0.017396752	0.001375314	0.018828697
TCdib	0.0305149	0	0.029825352	0
TCdp	0.001672607	0.032788227	0.000130003	0.032707619
Total	179.085773	142.1320237	113.2972646	141.7107358

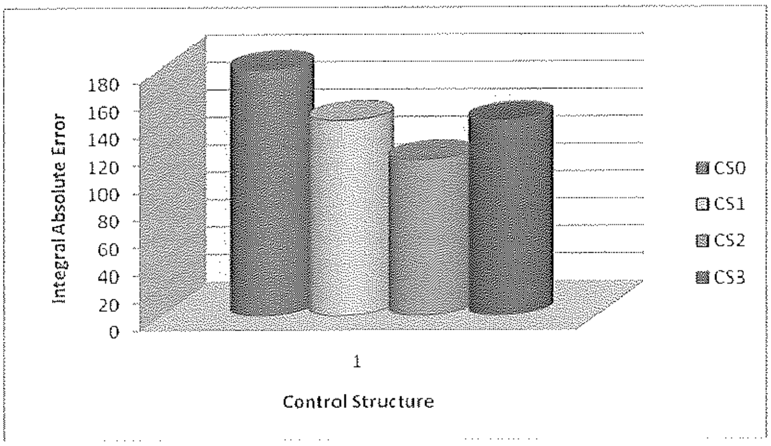

Figure 5.16 The IAE summation results of 4 control structures Isobutane process to response the change of ±10 of fresh feed temperature .

Table 5.4 The IAE results of the 4 control structures to an energy consumption

Heat Flow (Btu/hr)	CS0	CS1	CS2	CS3
Compressor duty	1.94E+06	1.94E+06	1.94E+06	1.94E+06
Cooler duty	1.82E+06	1.82E+06	1.82E+06	1.82E+06
Column 1 Condenser Duty	1.18E+07	1.18E+07	1.18E+07	1.18E+07
Column 1 Reboiler Duty	1.01E+07	1.01E+07	9.12E+06	9.12E+06
Column 2 Condenser Duty	3.12E+05	3.12E+05	3.12E+05	3.12E+05
Column 2 Reboiler Duty	8.55E+05	8.55E+05	8.55E+05	8.55E+05
Column 3 Condenser Duty	2.29E+05	2.29E+05	2.29E+05	2.29E+05
Column 3 Reboiler Duty	4.91E+05	4.91E+05	4.39E+05	4.39E+05
Total	2.76E+07	2.76E+07	2.65E+07	2.65E+07

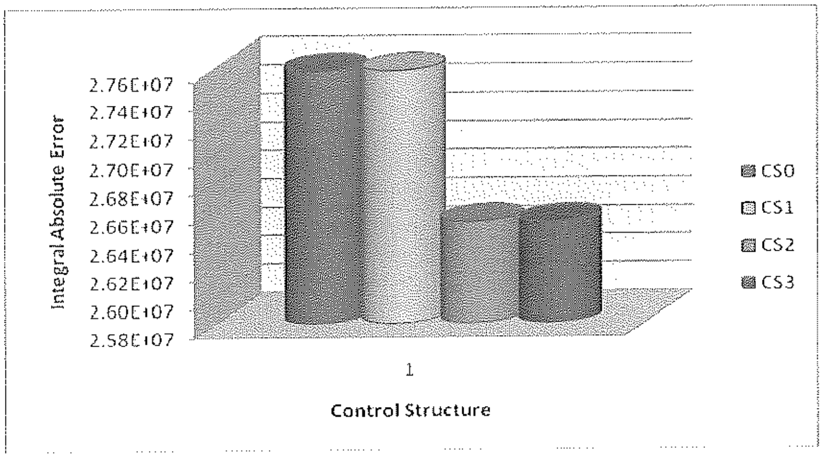

Figure 5.17 The IAE summation results of four control structures Isobutane process showing the value of all energy use to change response to the change of all disturbances.

CHAPTER VI

CONCLUSION AND SUGGESTION FOR FURTHER WORK

6.1 Conclusion

In this work an alternative Isobutane plantwide control structure has been designed to achieve the maximum desired yield and minimum energy consumption, moreover, to handle the disturbance from an external comparative to the reference based cased of Luyben (2009).

Four plantwide control structure configuration of Isobutane process have been designed following by Wongsri heuristic method comparative to the Luyben heuristic existing plantwide process. The based case control structure (CS0) is the Luyben Isobutane process heuristic design, control structure 1 (CS1) is the modification from the control structure based case by applied the Fixture point theorem of Wongsri to the process, considered in the IAE summation value for flow variation, we found that there were a high disturbance in the reaction section which we will take in to consider and determine the effect of disturbance to the inlet and outlet of reactor effluent stream in order to prove the wording of Luben about the "the relative change in reaction rate depend on temperature through the activation energy", therefore we choose the reactor inlet temperature setpoint as the production rate handle for this case. Therefore we have changed the valve controller to revert level control of reactor 3 in order to maximize the production rate and revert level level of reactor 2 in order to control the flow of BB2 to not exceed than maximum required since, excess in BB2 might lead the side reaction of dodecane which unsatisfied of process. In the second control structure (CS2), we always based on control structure CS1, we have change a bit in level control of DIB column to manipulated control by bottom valve instead and composition control by manipulated control by reboiler duty. The third control structure (CS3), is modification a bit from Luben control structure design by changing the level controller of DIB column to manipulated bottom valve control instead and composition control by manipulated control by reboiler duty.

The dynamic simulation of this process to handle the various disturbances is made to evaluated the performance of all control structure; increasing and decreasing the fresh feed flow rate and temperature, the results were successfully performed to identify the key control aspects of the control structure.

Figure from dynamic response have been shown the stability of the plantwide control to handle the external disturbance during the process which can performed as well.

The IAE results of temperature and molar flow rate change for disturbance test in four control structures. These performance of these control structure can be arranged from the best to lowest performance (error of controllability point of view) as following sequence; CS2, CS3, CS0, CS1.

Moreover, the dynamic response obtained from graph has been shown that the mole fraction of Isobutene achieved the high yield at control structures two. The simulation result for increasing and decreasing (±10%) the BB1 fresh feed . The molar flow rate response is oscillatory and it comes to new setpoint within a few minutes. Effect from the step change in flow rate by (±10%) shown that the rate of reaction and mole fraction of Isobutane in reactor 1, 2, 3 and DIB distillate increase from the base case but not much difference in temperature change

Comparison dynamic responses of step change in BB1 flow rate temperature (±10%) between CS0 of Luben based case compare to new control structure design CS1, CS2 and CS3., the simulation result of temperature response in oscillatory, moreover the rate of reaction of Isobutane also increase because of increasing temperature can raise the reaction rate.

6.2 Recommendations

1. Study on the other control structures of Isobutane Alklyation process.
2. Improve the strategy accelerating the dynamic performance of the complex chemical plantwide process.
3. Study and improve the strategy of MPC plantwide control of Isobutane Alklyation process.
4. Study and design the new control atructure of heat exchanger networks applied to the Isobutane plantwide process.
5. Study the self optimizing control structure of the Isobutane Alklyation process.

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APPENDICES

APPENDIX A

TABLE A.1 EQUIPMENT SIZES

UNIT		SPECIFICATION
DIB	COLUMN ID	6 FT
	REFLUX DRUM	360 FT^3
	BASE	300 FT^3
DP	COLUMN ID	1.2 FT
	REFLUX DRUM	20 FT^3
	BASE	100 FT^3
DB	COLUMN ID	1 FT
	REFLUX DRUM	9 FT^3
	BASE	40 FT^3
KO DRUM	VOLUME	100 FT^3
DRUM	VOLUME	62 FT^3
FEHE	SHELL VOLUME	11 FT^3
	TUBE VOLUME	11 FT^3
REACTOR (3)	TOTAL VOLUME (80% LIQUID FULL)	100 FT^3

A1 Manipulated Variables

Number	Symbol	List of overall manipulabel variables
1	V1	BB1 fresh feed valve
2	V2	BB2 fresh feed valve
3	V3	Recycle valve
4	V4	SatC4 feed valve
5	V5	Tank valve
6	V7	DIB column bottom valve
7	QCOND	Cooler heat duty
8	Wkcomp	Compressor power
9	V8	DP bottom valve
10	V9	DP column overhead valve
11	V11	R1 liquid valve
12	V12	R1 vapor valve
13	V13	Tank vapor valve
14	V14	DB column bottom valve
15	V15	DB column overhead valve
16	V21	R2 liquid valve
17	V22	R2 vapor valve
18	V31	R3 liquid valve
19	V32	R3 vapor valve
20	QR1	DIB column reboiler heat duty
21	QC1	DIB column condenser heat duty
22	QR2	DP column reboiler heat duty
23	QC2	DP column condenser heat duty
24	QR3	DB column reboiler heat duty
25	QC3	DB column condenser heat duty

TUNING OF CONTROL STRUCTURES

A.2 Tuning Controllers

This work uses the several types of the controllers such as P, PI, and PID which depend on each control loop. In theory, the control performance able to improve by using of the derivative action but in practice, there is some significant drawbacks for the use of derivative as follow:

1. To specify the 3 tuning constants.
2. Signal noise is amplified.
3. Using a several types of PID control algorithms, therefore an important to be careful that matching tuning method of using the right algorithm.
4. An approximation simulation of the real plant. If additional high performance controllers are required to get better dynamics of the simulation, it is not not work well for the real plant.

A.3 Tuning Flow, Level and Pressure Loops

Since, the very fast of the dynamics of flow measurement and 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 the most consideration for controllers. Value of controller gain always keep modest due to flow measurement signal is sometime noisy due to the flow as turbulent throughout an orifice plate. A value of controller gain of K_C = 0.5 is recommended. Derivative action should not be used.

The most level controllers should use proportional-only action with a gain of 1 to 2. Since, it provides the maximum amount of smooth flow. Proportional control means it will be steady state offset (the level will be not returned to its setpoint value). However, to maintain a liquid certain value level is not necessary when the liquid capacity is being used as surge volume. So the recommended tuning of a level controller is K_C = 2.

Most pressure controllers able to easily tune. The constant of process time is estimated by dividing system volume of gas by gas volumetric flow rate flowing through the system. Integral time is setting equal to 2 to 4 times of process time constant and using a reasonable controller gain which always gives satisfactory pressure control. Typical the tuning constants of pressure controller for columns and tanks are K_C = 2 and τ_I = 10 minutes.

A.4 Relay- Feedback Testing

The relay-feedback test is a tool which serves a quick and simple method to identify the dynamic parameter which is important for feedback controller designing. The results of the test are an ultimate gain and an ultimate frequency. This information is always sufficient for us the permission to calculate the reasonable controller tuning constants.

The method consists of an inserting on-off relay in the feedback loop. Parameter which must be specified is the height of the relay, h. This height is typically 5 to 10 % of the controller output scale. The loop is starting to oscillate around the setpoint with the controller output switching every time the process variable (PV) signal crosses the setpoint.

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

$$K_U = \frac{4h}{a\pi}$$

The period of the output PV curve is an ultimate period, P_U from these two parameters controller tuning constants able to calculate the PI and PID controllers, using a various tuning methods proposed in literature that required only the ultimate gain and the ultimate frequency, e.g. Ziegler-Nichols, Tyreus-Luyben.

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

1. Only one parameter has to be specified (relay height).
2. The time to run the test is short, comparative to the extended periods required for methods like PRBS.
3. The test is closed loop, so the process will be not run away from the setpoint.
4. Information achievement is very accurate in the frequency range that is important for the design of a feedback controller.
5. An impact of the change of load which occurred during the test able to detect by a change to asymmetric pulses in manipulated variable.

The entire features make a useful identification of the relay-feedback testing. To know an ultimate gain, K_U and an ultimate period, P_U was permitted us to calculate controller settings. There are several methods that require only for such two parameters. The Ziegler-Nichols tuning equations for a PI controller are:

$$K_C = K_U / 2.2$$
$$\tau_I = P_U / 1.2$$

These tuning constants are frequently usage 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:

$$K_C = K_U / 3.2$$
$$\tau_I = 2.2 P_U$$

A.5 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 tuning is often possible on the simulation than the real plant. Thus the predictions of dynamic performance able to overly optimistic.

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.2 summarizes some recommended lags to include in several different types of control loops.

Table A.2 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 1 (CS1)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
FC1	BB1 flow rate	V1	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC2	BB2 flow rate	V2	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC3	Recycle flow rate	V3	PI	0.5	0.3	-	Reverse	0-100lbmole/hr
PCR	KO vessel pressure	Wkcomp	PI	3	-	-	Direct	0-50 psia
TC1	Reactor 1 temperature	V12	PID	0.5	10	-	Direct	0-100 F
TC2	Reactor 2 temperature	V22	PID	0.5	10	-	Direct	0-100 F
TC3	Reactor 3 vessel temperature	V32	PID	0.5	-	-	Direct	0-100 F
TC4	Tank temperature	qcond	PID	0.5	0.274	0.069	Direct	50-150 F
LC1	Reactor 1 level	V11	P	3.23	-	-	Direct	0-100 %
LC2	Reactor 2 level	V31	P	4	-	-	Direct	0-100 %
LC3	Reactor 3 level	V10	P	4	-	-	Direct	0-100 %
LC4	Tank level	V5	P	4	-	-	Direct	0-100 %

Table B.1 (Continuous) Tuning parameters for the reference control structure 1 (CS1)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
TCdib	Tray 25 iso-butane comp of DIB column	V7	PID	2.5	10	-	Reverse	0-100 %
PCdib	Pressure condenser of DIB column	qc1	PI	2	-	-	Reverse	0-100 %
PCdp	Pressure condenser of DP column	qc2	PI	2	-	-	Reverse	0-100 %
PCdb	Pressure condenser of DIB column	qc3	PI	2	-	-	Reverse	0-100 %
TCdp	Tray 25 temperature of DP column	qr2	PID	4	2	-	Reverse	100-200 F
TCdb	Tray 3 temperature of DIB column	qr3	PID	2	10	-	Reverse	200-400 F
LCdib2	Level condenser of DIB column	V4	P	2		-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	Qr1	P	2		-	Reverse	0-100 %
LCdp2	Level condenser of DP column	V9	P	2		-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	P	2		-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	P	2		-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	P	12		-	Direct	0-100 %

Table B.2 Tuning parameters for the reference control structure 2 (CS2)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
FC1	BB1 flow rate	V1	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC2	BB2 flow rate	V2	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC3	Recycle flow rate	V3	PI	0.5	0.3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	Wkcomp	PI	3	-	-	Direct	0-50 psia
TC1	Reactor 1 temperature	V12	PID	0.5	10	-	Direct	0-100 F
TC2	Reactor 2 temperature	V22	PID	0.5	10	-	Direct	0-100 F
TC3	Reactor 3 temperature	V32	PID	0.5	-	-	Direct	0-100 F
TC4	Tank temperature	qcond	PID	0.5	0.274	0.069	Direct	50-150 F
LC1	Reactor 1 level	V11	P	3.23	-	-	Direct	0-100 %
LC2	Reactor 2 level	V31	P	4	-	-	Direct	0-100 %
LC3	Reactor 3 level	V10	P	4	-	-	Direct	0-100 %
LC4	Tank level	V5	P	4	-	-	Direct	0-100 %

Table B.2 (Continuous) Tuning parameters for the control structure 2 (CS2)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
TCdib	Tray 19 iso-butane comp of DIB column	V7	PID	2	48	-	Reverse	0-100 %
PCdib	Pressure condenser of DIB column	qc1	PI	2	-	-	Reverse	0-100 %
PCdp	Pressure condenser of DP column	qc2	PI	2	-	-	Reverse	0-100 %
PCdb	Pressure condenser of DIB column	qc3	PI	2	-	-	Reverse	0-100 %
TCdp	Tray 25 temperature of DP column	qr2	PID	4	2	-	Reverse	100-200 F
TCdb	Tray 3 temperature of DIB column	qr3	PID	2	10	-	Reverse	200-400 F
LCdib2	Level condenser of DIB column	V4	P	2		-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	Qr1	P	2		-	Reverse	0-100 %
LCdp2	Level condenser of DP column	V9	P	2		-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	P	2		-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	P	2		-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	P	12		-	Direct	0-100 %

Table B.3 Tuning parameters for the control structure 3 (CS3)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
FC1	BB1 flow rate	V1	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC2	BB2 flow rate	V2	PI	0.5	0.3	-	Reverse	0-50 lbmole/hr
FC3	Recycle flow rate	V3	PI	0.5	0.3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	Wkcomp	PI	3	-	-	Direct	0-50 psia
TC1	Reactor 1 temperature	V12	PID	0.5	10	-	Direct	0-100 F
TC2	Reactor 2 temperature	V22	PID	0.5	10	-	Direct	0-100 F
TC3	Reactor 3 temperature	V32	PID	0.5	-	-	Direct	0-100 F
TC4	Tank temperature	qcond	PID	0.5	0.274	0.069	Direct	50-150 F
LC1	Reactor 1 level	V11	P	3.23	-	-	Direct	0-100 %
LC2	Reactor 2 level	V31	P	4	-	-	Direct	0-100 %
LC3	Reactor 3 level	V10	P	4	-	-	Direct	0-100 %
LC4	Tank level	V5	P	4	-	-	Direct	0-100 %

Table B.3 (Continuous) Tuning parameters for the control structure 3 (CS3)

Controller	Controlled variables	Manipulated variable	Type	Tuning parameter			Action controller	PV Range
				K_C	τ_I	τ_D		
TCdib	Tray 19 iso-butane comp of DIB column	V7	PID	2.5	10	-	Reverse	0-100 %
PCdib	Pressure condenser of DIB column	qc1	PI	2	-	-	Reverse	0-100 %
PCdp	Pressure condenser of DP column	qc2	PI	2	-	-	Reverse	0-100 %
PCdb	Pressure condenser of DIB column	qc3	PI	2	-	-	Reverse	0-100 %
TCdp	Tray 25 temperature of DP column	qr2	PID	4	2	-	Reverse	100-200 F
TCdb	Tray 3 temperature of DIB column	qr3	PID	2	10	-	Reverse	200-400 F
LCdib2	Level condenser of DIB column	V4	P	2		-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	Qr1	P	2		-	Reverse	0-100 %
LCdp2	Level condenser of DP column	V9	P	2		-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	P	2		-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	P	2		-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	P	12		-	Direct	0-100 %

APPENDIX C

DATA OF FIXTURE POINT ANALYSIS

Table C.1 IAE result of temperature of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
BB1	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
B1	0.0433	0.0440	0.1376	0.2194	0.1871	0.1424	0.1046	0.2322	0.5282	0.1201	0.1201
B2	0.0743	0.0771	0.0248	0.0810	0.0873	0.1914	0.0927	0.0815	0.0644	0.1114	0.1114
B3	0.0529	0.0479	0.6204	0.0536	0.0503	0.3633	0.1180	0.0578	0.0953	0.1308	0.1308
D1	0.0094	0.0099	0.0147	0.0691	0.0683	0.0924	0.0532	0.0714	0.0921	0.0673	0.0673
D2	0.0778	0.0859	0.0262	0.0915	0.0969	0.1937	0.1020	0.0920	0.0786	0.1196	0.1196
D3	0.0208	0.0220	0.0002	0.0718	0.0635	0.1075	0.0515	0.0768	0.3323	0.0641	0.0641
disch	0.1277	0.1052	0.1277	0.0661	0.0635	0.0653	0.0519	0.0660	0.0779	0.0592	0.0592
dondout	0.1295	0.1009	0.1295	0.0779	0.0749	0.0713	0.0604	0.0779	0.0935	0.0691	0.0691
hxcount	0.1503	0.1011	0.1503	0.1250	0.1243	0.0977	0.0931	0.1294	0.1637	0.1105	0.1105
hxhin	0.1488	0.1316	0.1488	0.1311	0.1322	0.1625	0.1205	0.1319	0.1515	0.1393	0.1393
hxhout	0.1298	0.1097	0.1298	0.0901	0.0987	0.3402	0.1154	0.0922	0.0682	0.1372	0.1372

Table C.1 (Continue) I AE result of temperature of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
L1	0.0573	0.0512	0.0573	0.0391	0.0388	0.1028	0.0409	0.0390	0.0460	0.0406	0.0406
L2	0.1230	0.0999	0.1742	0.0901	0.0998	0.3410	0.1177	0.0924	0.0701	0.1395	0.1395
L3	0.1530	0.1422	0.1406	0.1413	0.1330	0.0865	0.0996	0.1403	0.2037	0.1159	0.1159
Lko	0.0122	0.0165	0.0122	0.0474	0.0476	0.0908	0.0395	0.0488	0.0579	0.0494	0.0494
Ltk	0.0874	0.0768	0.0874	0.0621	0.0614	0.1076	0.0579	0.0617	0.0630	0.0608	0.0608
tot1	0.1376	0.1103	0.1376	0.1393	0.1313	0.0885	0.1000	0.1371	0.1732	0.1181	0.1181
tot2	0.0095	0.0099	0.0095	0.0691	0.0683	0.0924	0.0532	0.0713	0.0919	0.0672	0.0672
total	0.1302	0.1102	0.1302	0.0852	0.0945	0.3272	0.1109	0.0874	0.0645	0.1322	0.1322
SatC4	0.0874	0.0768	0.0874	0.0621	0.0614	0.1076	0.0579	0.0617	0.0630	0.0608	0.0608
p1out	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
p2out	0.1211	0.0997	0.1211	0.0644	0.0617	0.0643	0.0507	0.0642	0.0760	0.0576	0.0576
p3out	0.1289	0.1051	0.1289	0.0733	0.0705	0.0681	0.0571	0.0732	0.0872	0.0653	0.0653
v1out	0.0874	0.0123	0.0874	0.0474	0.0476	0.0908	0.0395	0.0488	0.0579	0.0494	0.0494
v2out	0.1154	0.0957	0.1154	0.0616	0.0592	0.0618	0.0485	0.0000	0.0722	0.0553	0.0553
v3out	0.1196	0.0968	0.1196	0.0686	0.0659	0.0647	0.0537	0.0000	0.0811	0.0612	0.0612

Table C.1 (Continue) I AE result of temperature of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
v4out	0.1206	0.0993	0.0969	0.0644	0.0617	0.0644	0.0507	0.0000	0.0000	0.0576	0.0576
v5out	0.0000	0.0000	0.0000	0.0002	0.0002	0.0002	0.0001	0.0000	0.0057	0.0002	0.0002
v7out	0.1293	0.1094	0.0948	0.0851	0.0944	0.3270	0.1107	0.0000	0.0644	0.1321	0.1321
v8out	0.0000	0.0000	0.0000	0.2089	0.1742	0.1160	0.0897	0.0000	0.5305	0.1046	0.1046
v11out	0.0136	0.0139	0.0144	0.0469	0.0475	0.0753	0.0396	0.0000	0.0561	0.0498	0.0498
v12out	0.1292	0.1050	0.1060	0.0731	0.0703	0.0682	0.0570	0.0000	0.0869	0.0652	0.0652
v21out	0.1790	0.1632	0.0695	0.1280	0.1274	0.1633	0.1158	0.0000	0.1466	0.1332	0.1332
v22out	0.1448	0.0886	0.1560	0.1269	0.1234	0.0953	0.0931	0.0000	0.1618	0.1106	0.1106
v31out	0.1340	0.1066	0.0884	0.1126	0.1161	0.1535	0.1148	0.0000	0.0996	0.1306	0.1306
v32out	0.1376	0.1103	0.1011	0.1391	0.1311	0.0884	0.0998	0.0000	0.1729	0.1179	0.1179
vap1	0.1527	0.1145	0.0000	0.1106	0.1079	0.1011	0.0805	0.0000	0.1420	0.0975	0.0975
vap2	0.1277	0.1053	0.0713	0.0000	0.0635	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
vap3	0.1296	0.1010	0.0959	0.0780	0.0750	0.0713	0.0603	0.0000	0.0936	0.0692	0.0692
vaptot	0.1516	0.1009	0.1625	0.1250	0.1231	0.0959	0.0925	0.0000	0.1616	0.1100	0.1100
vtk	0.1488	0.1316	0.3402	0.1312	0.1323	0.1625	0.1205	0.0000	0.1515	0.1394	0.1394

Table C.1 (Continue) I AE result of temperature of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
BB1	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
B1	0.3732	0.3087	0.1260	0.1109	0.0874	0.0645	0.1322	0.3272	0.0657	0.0728	0.1246
B2	0.0698	0.0711	0.1105	0.0579	0.0617	0.0630	0.0608	0.1077	0.0595	0.0627	0.0592
B3	0.0766	0.0643	0.1342	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1	0.0782	0.0745	0.0679	0.0507	0.0642	0.0760	0.0576	0.0643	0.0701	0.0683	0.0579
D2	0.0829	0.0828	0.1190	0.0571	0.0732	0.0872	0.0653	0.0681	0.0798	0.0781	0.0655
D3	0.1735	0.1203	0.0655	0.0395	0.0489	0.0579	0.0494	0.0908	0.0514	0.0494	0.0496
disch	0.0719	0.0701	0.0594	0.0486	0.0000	0.0722	0.0553	0.0618	0.0668	0.0652	0.0556
dondout	0.0849	0.0836	0.0698	0.0537	0.0000	0.0811	0.0612	0.0647	0.0744	0.0730	0.0617
hxcount	0.1230	0.1423	0.1071	0.0507	0.0000	0.0000	0.0576	0.0644	0.0701	0.0683	0.0578
hxhin	0.1431	0.1313	0.1501	0.0001	0.0000	0.0057	0.0002	0.0002	0.0015	0.0007	0.0002
hxhout	0.0693	0.0779	0.1289	0.1107	0.0000	0.0644	0.1321	0.3270	0.0656	0.0727	0.1245
L1	0.0407	0.0404	0.0391	0.0897	0.0000	0.5305	0.1046	0.1160	0.3715	0.3037	0.1110
L2	0.0694	0.0765	0.1318	0.0396	0.0000	0.0561	0.0498	0.0753	0.0501	0.0486	0.0499

Table C.1 (Continue) I AE result of temperature of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
Lko	0.0514	0.0493	0.0496	0.1158	0.0000	0.1466	0.1332	0.1633	0.1385	0.1066	0.1444
Ltk	0.0595	0.0627	0.0592	0.0931	0.0000	0.1618	0.1106	0.0953	0.1279	0.1406	0.1084
tot1	0.1488	0.1482	0.1173	0.1148	0.0000	0.0996	0.1306	0.1535	0.1023	0.1032	0.1282
tot2	0.0781	0.0744	0.0678	0.0998	0.0000	0.1729	0.1179	0.0884	0.1486	0.1480	0.1171
total	0.0656	0.0728	0.1246	0.0805	0.0000	0.1420	0.0975	0.1011	0.1117	0.1233	0.0948
SatC4	0.0595	0.0627	0.0592	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0594
p1out	0.0000	0.0000	0.0000	0.0603	0.0000	0.0936	0.0692	0.0713	0.0851	0.0835	0.0698
p2out	0.0701	0.0683	0.0579	0.0925	0.0000	0.1616	0.1100	0.0959	0.1255	0.1405	0.1073
p3out	0.0798	0.0781	0.0655	0.1206	0.0000	0.1515	0.1394	0.1625	0.1431	0.1314	0.1502
v1out	0.0514	0.0493	0.0496	0.1154	0.0000	0.0684	0.1372	0.3402	0.0693	0.0779	0.1289
v2out	0.0668	0.0652	0.0556	0.0002	0.0003	0.0003	0.0004	0.0000	0.0007	0.0010	0.0007
v3out	0.0744	0.0730	0.0617	0.0916	0.0531	0.0678	0.0919	0.0696	0.0764	0.0723	0.0635
v4out	0.0701	0.0683	0.0578	0.0907	0.0528	0.0673	0.0909	0.0691	0.0757	0.0717	0.0633
v5out	0.0015	0.0007	0.0002	0.0899	0.0524	0.0668	0.0898	0.0685	0.0750	0.0711	0.0630
v7out	0.0656	0.0727	0.1245	0.0892	0.0521	0.0663	0.0887	0.0679	0.0742	0.0705	0.0628

Table C.1 (Continue) I AE result of temperature of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
v11out	0.0501	0.0485	0.0499	0.0877	0.0513	0.0652	0.0864	0.0666	0.0726	0.0692	0.0624
v12out	0.0796	0.0779	0.0654	0.0871	0.0509	0.0647	0.0852	0.0660	0.0717	0.0685	0.0622
v21out	0.1385	0.1066	0.1444	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v22out	0.1279	0.1406	0.1084	0.0857	0.0501	0.0635	0.0829	0.0647	0.0701	0.0671	0.0618
v31out	0.1023	0.1032	0.1282	0.0850	0.0497	0.0630	0.0817	0.0640	0.0692	0.0664	0.0616
v32out	0.1485	0.1480	0.1171	0.0843	0.0493	0.0624	0.0805	0.0633	0.0684	0.0657	0.0614
vap1	0.1117	0.1233	0.0948	0.0837	0.0489	0.0618	0.0793	0.0626	0.0675	0.0650	0.0612
vap2	0.0000	0.0000	0.0594	0.0830	0.0485	0.0612	0.0780	0.0619	0.0666	0.0643	0.0610
vap3	0.0851	0.0835	0.0697	0.0824	0.0480	0.0607	0.0768	0.0612	0.0657	0.0636	0.0608
vaptot	0.1255	0.1405	0.1073	0.0818	0.0477	0.0601	0.0756	0.0605	0.0648	0.0629	0.0606
vtk	0.1431	0.1314	0.1502	0.0839	0.0491	0.0616	0.0764	0.0618	0.0659	0.0643	0.0604
L3	0.1771	0.1561	0.1307	0.0570	0.0000	0.0869	0.0652	0.0682	0.0796	0.0779	0.0654
v8out	0.3715	0.3037	0.1110	0.0885	0.0517	0.0657	0.0875	0.0673	0.0734	0.0698	0.0626

Table C.2 IAE result of Molar flow of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
BB1	0.0584	0.0295	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2	0.0450	0.0354	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
B1	0.0155	0.0160	0.0441	0.0262	0.0218	0.0562	0.0260	0.0281	0.0288	0.0538	0.0288
B2	0.0584	0.0561	0.0661	0.0618	0.0656	0.1431	0.0696	0.0816	0.0621	0.0548	0.0621
B3	0.0134	0.0138	0.0338	0.0256	0.0240	0.0460	0.0245	0.0274	0.0265	0.0388	0.0265
D1	0.0792	0.0677	0.1159	0.0740	0.0667	0.0195	0.0288	0.0421	0.0776	0.1344	0.0776
D2	0.0007	0.0008	0.0098	0.0078	0.0091	0.0144	0.0089	0.0131	0.0079	0.0060	0.0079
D3	0.0020	0.0021	0.0173	0.0093	0.0081	0.0103	0.0045	0.0062	0.0099	0.0218	0.0099
disch	0.2341	0.2954	0.6025	0.4552	0.4346	0.1815	0.2934	0.3541	0.4613	0.5823	0.4613
dondout	0.1337	0.3063	1.0000	0.8082	0.7361	0.4246	0.5973	0.6611	0.7423	0.9408	0.7423
hxcount	0.1153	0.0585	0.7356	0.6096	0.5855	0.3497	0.4484	0.5211	0.6176	0.7556	0.6176
hxhin	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
hxhout	0.0531	0.0345	0.1536	0.1077	0.1016	0.1466	0.0759	0.0925	0.1090	0.1604	0.1090

Table C.2 (Continue) IAE result of Molar flow of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
L1	0.1153	0.0574	0.7123	0.5885	0.5657	0.3483	0.4406	0.5101	0.5955	0.7326	0.5955
L2	0.0550	0.0373	0.2084	0.1377	0.1164	0.1456	0.0769	0.0943	0.1345	0.2332	0.1345
L3	0.0552	0.0376	0.2049	0.1359	0.1152	0.1490	0.0748	0.0879	0.1322	0.2278	0.1322
Lko	0.0210	0.0000	0.1810	0.1330	0.1286	0.1374	0.0937	0.1175	0.1349	0.1840	0.1349
Ltk	0.0219	0.0115	0.1773	0.1319	0.1272	0.1366	0.0924	0.1165	0.1360	0.1833	0.1360
tot1	0.1153	0.0585	0.7356	0.6096	0.5855	0.3497	0.4484	0.5211	0.6176	0.7556	0.6176
tot2	0.0792	0.0677	0.1159	0.0740	0.0667	0.0195	0.0288	0.0421	0.0776	0.1344	0.0776
total	0.1983	0.0345	0.1536	0.1077	0.1016	0.1466	0.0759	0.0925	0.1090	0.1604	0.1090
SatC4	0.0219	0.0115	0.1773	0.1319	0.1272	0.1366	0.0924	0.1165	0.1360	0.1833	0.1360
p1out	0.0035	0.0036	0.0145	0.0143	0.0140	0.0173	0.0113	0.0138	0.0146	0.0152	0.0146
p2out	0.0232	0.0185	0.1773	0.1319	0.1272	0.1366	0.0924	0.1165	0.1360	0.1833	0.1360
p3out	0.2531	0.3181	0.6025	0.4552	0.4346	0.1815	0.2934	0.3541	0.4613	0.5823	0.4613
v1out	0.0219	0.0118	0.1810	0.1330	0.1286	0.1374	0.0937	0.1175	0.1349	0.1840	0.1349
v2out	0.0384	0.0295	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v3out	0.0450	0.0354	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v4out	0.0219	0.0115	0.1773	0.1319	0.1272	0.1366	0.0924	0.1165	0.1360	0.0000	0.1360

Table C.2 (Continue) IAE result of Molar flow of Isobutane

PV	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
v5out	0.0035	0.0036	0.0145	0.0143	0.0140	0.0173	0.0113	0.0138	0.0146	0.0152	0.0146
v7out	0.0531	0.0345	0.1536	0.1077	0.1016	0.1466	0.0759	0.0925	0.1090	0.1604	0.1090
v8out	0.0277	0.0423	0.0441	0.0262	0.0218	0.0562	0.0260	0.0281	0.0288	0.0538	0.0288
v11out	0.0584	0.0561	0.0661	0.0618	0.0656	0.1431	0.0696	0.0816	0.0621	0.0548	0.0621
v12out	0.2341	0.2954	0.6025	0.4552	0.4346	0.1815	0.2934	0.3541	0.4613	0.5823	0.4613
v21out	0.1242	0.1923	0.3353	0.2694	0.2535	0.1514	0.1959	0.2302	0.2668	0.3487	0.2668
v22out	0.1337	0.3063	1.0000	0.8082	0.7361	0.4246	0.5973	0.6611	0.7423	0.9408	0.7423
v31out	0.0404	0.0572	0.1230	0.1166	0.1117	0.0680	0.1015	0.1014	0.1159	0.1288	0.1159
v32out	0.1153	0.0585	0.4613	0.6096	0.5855	0.3497	0.4484	0.5211	0.4613	0.7556	0.4613
vap1	0.0036	0.0261	0.0021	0.0018	0.0016	0.0008	0.0013	0.0014	0.0017	0.0022	0.0017
vap2	0.1242	0.1923	0.3353	0.0474	0.0026	0.0468	0.0372	0.2302	0.0473	0.0558	0.0473
vap3	0.0404	0.0572	0.1230	0.1166	0.1117	0.0680	0.1015	0.1014	0.1159	0.1288	0.1159
vaptot	0.0036	0.0261	0.0021	0.0018	0.0016	0.0008	0.0013	0.0014	0.0017	0.0022	0.0017
vtk	0.0552	0.0376	0.2049	0.1359	0.1152	0.1490	0.0748	0.0879	0.1322	0.2278	0.1322

Table C.2 (Continue) IAE result of Molar flow of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
BB1	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
B1	0.0277	0.0423	0.0363	0.0281	0.0260	0.0281	0.0288	0.0538	0.0000	0.0277	0.0423
B2	0.0822	0.0560	0.0558	0.0816	0.0696	0.0816	0.0621	0.0548	0.0000	0.0822	0.0560
B3	0.0270	0.0323	0.0297	0.0274	0.0245	0.0274	0.0265	0.0388	0.0000	0.0270	0.0323
D1	0.0417	0.0999	0.0934	0.0421	0.0288	0.0421	0.0776	0.1344	0.0000	0.0417	0.0999
D2	0.0130	0.0058	0.0062	0.0131	0.0089	0.0131	0.0079	0.0060	0.0000	0.0130	0.0058
D3	0.0059	0.0156	0.0131	0.0062	0.0045	0.0062	0.0099	0.0218	0.0000	0.0059	0.0156
disch	0.3799	0.5134	0.4711	0.3541	0.2934	0.3541	0.4613	0.5823	0.0000	0.3799	0.5134
dondout	0.6680	0.8137	0.8296	0.6611	0.5973	0.6611	0.7423	0.9408	0.0000	0.6680	0.8137
hxcount	0.5096	0.6846	0.6543	0.5211	0.4484	0.5211	0.6176	0.7556	0.0000	0.5096	0.6846
hxhin	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
hxhout	0.0933	0.1288	0.1206	0.0925	0.0759	0.0925	0.1090	0.1604	0.0000	0.0933	0.1288
L1	0.4990	0.6622	0.6321	0.5101	0.4406	0.5101	0.5955	0.7326	0.0000	0.4990	0.6622
L2	0.0939	0.1908	0.1663	0.0943	0.0769	0.0943	0.1345	0.2332	0.0000	0.0939	0.1908
L3	0.0919	0.1877	0.1644	0.0879	0.0748	0.0879	0.1322	0.2278	0.0000	0.0919	0.1877

Table C.2 (Continue) IAE result of Molar flow of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
Lko	0.1152	0.1547	0.1480	0.1175	0.0937	0.1175	0.1349	0.1840	0.0000	0.1152	0.1547
Ltk	0.1143	0.1558	0.1470	0.1165	0.0924	0.1165	0.1360	0.1833	0.0000	0.1143	0.1558
tot1	0.5096	0.6846	0.6543	0.5211	0.4484	0.5211	0.6176	0.7556	0.0000	0.5096	0.6846
tot2	0.0417	0.0999	0.0934	0.0421	0.0288	0.0421	0.0776	0.1344	0.0000	0.0417	0.0999
total	0.0933	0.1288	0.1206	0.0925	0.0759	0.0925	0.1090	0.1604	0.0000	0.0933	0.1288
SatC4	0.1143	0.1558	0.1470	0.1165	0.0924	0.1165	0.1360	0.1833	0.0000	0.1143	0.1558
p1out	0.0137	0.0152	0.0149	0.0138	0.0113	0.0138	0.0146	0.0152	0.0000	0.0137	0.0152
p2out	0.1143	0.1558	0.1470	0.1165	0.0924	0.1165	0.1360	0.1833	0.0000	0.1143	0.1558
p3out	0.3799	0.5134	0.4711	0.3541	0.2934	0.3541	0.4613	0.5823	0.0000	0.3799	0.5134
v1out	0.1152	0.1547	0.1480	0.1175	0.0937	0.1175	0.1349	0.1840	0.0000	0.1152	0.1547
v2out	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v3out	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v4out	0.1143	0.1558	0.1470	0.1165	0.0924	0.1165	0.1360	0.0000	0.0000	0.1143	0.1558
v5out	0.0137	0.0152	0.0149	0.0138	0.0113	0.0138	0.0146	0.0152	0.0000	0.0137	0.0152
v7out	0.0933	0.1288	0.1206	0.0925	0.0759	0.0925	0.1090	0.1604	0.0000	0.0933	0.1288
v8out	0.0277	0.0423	0.0363	0.0281	0.0260	0.0281	0.0288	0.0538	0.0000	0.0277	0.0423

Table C.2 (Continue) IAE result of Molar flow of Isobutane

PV	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
v11out	0.0822	0.0560	0.0558	0.0816	0.0696	0.0816	0.0621	0.0548	0.0000	0.0822	0.0560
v12out	0.3799	0.5134	0.4711	0.3541	0.2934	0.3541	0.4613	0.5823	0.0000	0.3799	0.5134
v21out	0.2173	0.3115	0.0131	0.2302	0.1959	0.2302	0.2668	0.3487	0.0000	0.2173	0.3115
v22out	0.6680	0.8137	0.8296	0.6611	0.5973	0.6611	0.7423	0.9408	0.0000	0.6680	0.8137
v31out	0.1033	0.1268	0.1216	0.1014	0.1015	0.1014	0.1159	0.1288	0.0000	0.1033	0.1268
v32out	0.5096	0.6846	0.6543	0.5211	0.4484	0.5211	0.4613	0.7556	0.0000	0.5096	0.6846
vap1	0.0014	0.0019	0.0019	0.0014	0.0013	0.0014	0.0017	0.0022	0.0000	0.0014	0.0019
vap2	0.0425	0.0515	0.0502	0.2302	0.0372	0.2302	0.0473	0.0558	0.0000	0.0425	0.0515
vap3	0.1033	0.1268	0.1216	0.1014	0.1015	0.1014	0.1159	0.1288	0.0000	0.1033	0.1268
vaptot	0.0014	0.0019	0.0019	0.0014	0.0013	0.0014	0.0017	0.0022	0.0000	0.0014	0.0019
vtk	0.0919	0.1877	0.1644	0.0879	0.0748	0.0879	0.1322	0.2278	0.0000	0.0919	0.1877

Table C.3 IAE result of DIB Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(50__Main TS)	0.0334	0.0121	0.0922	0.0687	0.0679	0.0691	0.0907	0.0528	0.0673	0.0909	0.0909
(49__Main TS)	0.0334	0.0121	0.0922	0.0687	0.0679	0.0691	0.0907	0.0528	0.0673	0.0909	0.0909
(48__Main TS)	0.0337	0.0120	0.0912	0.0681	0.0673	0.0685	0.0899	0.0524	0.0668	0.0898	0.0898
(47__Main TS)	0.0338	0.0120	0.0901	0.0675	0.0668	0.0679	0.0892	0.0520	0.0663	0.0887	0.0887
(46__Main TS)	0.0339	0.0120	0.0891	0.0668	0.0662	0.0673	0.0885	0.0517	0.0657	0.0875	0.0875
(45__Main TS)	0.0340	0.0119	0.0880	0.0662	0.0656	0.0666	0.0877	0.0513	0.0652	0.0864	0.0864
(44__Main TS)	0.0341	0.0119	0.0870	0.0656	0.0650	0.0660	0.0870	0.0509	0.0647	0.0852	0.0852
(43__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
(42__Main TS)	0.0343	0.0118	0.0848	0.0642	0.0637	0.0647	0.0857	0.0501	0.0635	0.0829	0.0829
(41__Main TS)	0.0344	0.0117	0.0837	0.0636	0.0631	0.0640	0.0850	0.0497	0.0630	0.0817	0.0817
(40__Main TS)	0.0345	0.0117	0.0826	0.0629	0.0625	0.0633	0.0843	0.0493	0.0624	0.0805	0.0805
(39__Main TS)	0.0346	0.0116	0.0815	0.0622	0.0618	0.0626	0.0837	0.0489	0.0618	0.0793	0.0793
(38__Main TS)	0.0347	0.0116	0.0804	0.0615	0.0611	0.0619	0.0830	0.0485	0.0612	0.0780	0.0780
(37__Main TS)	0.0347	0.0115	0.0793	0.0608	0.0605	0.0612	0.0824	0.0480	0.0606	0.0768	0.0768
(36__Main TS)	0.0347	0.0115	0.0782	0.0601	0.0599	0.0605	0.0818	0.0476	0.0601	0.0756	0.0756
(35__Main TS)	0.0306	0.0116	0.0792	0.0614	0.0612	0.0618	0.0839	0.0491	0.0616	0.0764	0.0764

Table C.3 (Continue) IAE result of DIB Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(34__Main TS)	0.0314	0.0117	0.0766	0.0601	0.0602	0.0605	0.0835	0.0487	0.0609	0.0731	0.0731
(33__Main TS)	0.0322	0.0117	0.0735	0.0585	0.0588	0.0589	0.0833	0.0483	0.0601	0.0692	0.0692
(32__Main TS)	0.0330	0.0119	0.0698	0.0566	0.0572	0.0571	0.0831	0.0477	0.0591	0.0649	0.0649
(31__Main TS)	0.0337	0.0120	0.0658	0.0544	0.0554	0.0549	0.0830	0.0471	0.0580	0.0601	0.0601
(30__Main TS)	0.0346	0.0121	0.0615	0.0520	0.0533	0.0525	0.0831	0.0464	0.0567	0.0551	0.0551
(29__Main TS)	0.0354	0.0123	0.0570	0.0494	0.0511	0.0499	0.0833	0.0457	0.0554	0.0500	0.0500
(28__Main TS)	0.0363	0.0125	0.0525	0.0467	0.0488	0.0473	0.0837	0.0450	0.0540	0.0451	0.0451
(27__Main TS)	0.0371	0.0128	0.0483	0.0441	0.0466	0.0447	0.0843	0.0443	0.0527	0.0405	0.0405
(26__Main TS)	0.0380	0.0131	0.0445	0.0416	0.0445	0.0423	0.0851	0.0438	0.0514	0.0366	0.0366
(25__Main TS)	0.0390	0.0134	0.0400	0.0389	0.0422	0.0396	0.0865	0.0433	0.0501	0.0319	0.0319
(24__Main TS)	0.0399	0.0138	0.0363	0.0366	0.0402	0.0373	0.0884	0.0430	0.0491	0.0282	0.0282
(23__Main TS)	0.0408	0.0141	0.0334	0.0347	0.0386	0.0355	0.0907	0.0429	0.0484	0.0254	0.0254
(22__Main TS)	0.0416	0.0143	0.0315	0.0336	0.0376	0.0343	0.0934	0.0432	0.0481	0.0239	0.0239
(21__Main TS)	0.0422	0.0145	0.0307	0.0331	0.0374	0.0340	0.0964	0.0438	0.0483	0.0243	0.0243
(20__Main TS)	0.0426	0.0146	0.0311	0.0336	0.0378	0.0344	0.0997	0.0448	0.0491	0.0257	0.0257
(19__Main TS)	0.0428	0.0145	0.0326	0.0347	0.0390	0.0356	0.1031	0.0461	0.0503	0.0276	0.0276

Table C.3 (Continue) IAE result of DIB Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(18__Main TS)	0.0428	0.0143	0.0352	0.0366	0.0407	0.0375	0.1064	0.0476	0.0520	0.0302	0.0302
(17__Main TS)	0.0426	0.0141	0.0386	0.0390	0.0430	0.0399	0.1093	0.0492	0.0539	0.0336	0.0336
(16__Main TS)	0.0423	0.0138	0.0426	0.0418	0.0454	0.0426	0.1115	0.0506	0.0558	0.0382	0.0382
(15__Main TS)	0.0419	0.0135	0.0470	0.0446	0.0479	0.0454	0.1128	0.0518	0.0576	0.0431	0.0431
(14__Main TS)	0.0414	0.0132	0.0515	0.0473	0.0503	0.0481	0.1131	0.0527	0.0591	0.0481	0.0481
(13__Main TS)	0.0409	0.0128	0.0557	0.0498	0.0523	0.0505	0.1123	0.0531	0.0602	0.0529	0.0529
(12__Main TS)	0.0405	0.0126	0.0597	0.0520	0.0541	0.0526	0.1104	0.0531	0.0609	0.0573	0.0573
(11__Main TS)	0.0401	0.0123	0.0632	0.0538	0.0554	0.0544	0.1076	0.0528	0.0612	0.0612	0.0612
(10__Main TS)	0.0398	0.0121	0.0662	0.0552	0.0564	0.0557	0.1041	0.0522	0.0612	0.0646	0.0646
(9__Main TS)	0.0396	0.0119	0.0688	0.0563	0.0572	0.0568	0.1002	0.0514	0.0609	0.0675	0.0675
(8__Main TS)	0.0394	0.0117	0.0709	0.0571	0.0576	0.0576	0.0961	0.0506	0.0605	0.0699	0.0699
(7__Main TS)	0.0392	0.0116	0.0726	0.0578	0.0580	0.0582	0.0920	0.0497	0.0600	0.0719	0.0719
(6__Main TS)	0.0391	0.0115	0.0739	0.0582	0.0581	0.0586	0.0881	0.0489	0.0594	0.0735	0.0735
(5__Main TS)	0.0391	0.0114	0.0750	0.0585	0.0582	0.0589	0.0845	0.0481	0.0588	0.0747	0.0747
(4__Main TS)	0.0391	0.0113	0.0758	0.0587	0.0582	0.0590	0.0813	0.0474	0.0583	0.0756	0.0756
(3__Main TS)	0.0398	0.0112	0.0761	0.0587	0.0581	0.0591	0.0783	0.0467	0.0577	0.0723	0.0723
(2__Main TS)	0.0516	0.0115	0.0693	0.0569	0.0565	0.0573	0.0697	0.0443	0.0551	0.0820	0.0820
(1__Main TS)	0.1534	0.0179	0.0851	0.0367	0.0373	0.0369	0.1111	0.0547	0.0597	0.2385	0.2385

Table C.3 (Continue) IAE result of DIB Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(50__Main TS)	0.0757	0.0717	0.0633	0.0775	0.0739	0.0745	0.0961	0.0582	0.0727	0.0963	0.0963
(49__Main TS)	0.0757	0.0717	0.0633	0.0775	0.0739	0.0745	0.0961	0.0582	0.0727	0.0963	0.0963
(48__Main TS)	0.0749	0.0711	0.0630	0.0767	0.0732	0.0739	0.0953	0.0578	0.0722	0.0952	0.0952
(47__Main TS)	0.0742	0.0705	0.0628	0.0758	0.0724	0.0733	0.0946	0.0575	0.0717	0.0941	0.0941
(46__Main TS)	0.0734	0.0698	0.0626	0.0750	0.0717	0.0727	0.0939	0.0571	0.0712	0.0930	0.0930
(45__Main TS)	0.0726	0.0691	0.0624	0.0742	0.0709	0.0721	0.0932	0.0567	0.0706	0.0918	0.0918
(44__Main TS)	0.0717	0.0685	0.0622	0.0733	0.0702	0.0714	0.0925	0.0564	0.0701	0.0906	0.0906
(43__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0054	0.0054	0.0054	0.0054	0.0054	0.0054
(42__Main TS)	0.0701	0.0671	0.0617	0.0716	0.0686	0.0701	0.0911	0.0556	0.0690	0.0883	0.0883
(41__Main TS)	0.0692	0.0664	0.0615	0.0707	0.0678	0.0694	0.0904	0.0551	0.0684	0.0871	0.0871
(40__Main TS)	0.0684	0.0657	0.0614	0.0698	0.0670	0.0688	0.0898	0.0547	0.0678	0.0859	0.0859
(39__Main TS)	0.0675	0.0650	0.0612	0.0689	0.0662	0.0681	0.0891	0.0543	0.0673	0.0847	0.0847
(38__Main TS)	0.0666	0.0643	0.0610	0.0679	0.0653	0.0674	0.0884	0.0539	0.0667	0.0835	0.0835
(37__Main TS)	0.0657	0.0636	0.0608	0.0670	0.0644	0.0666	0.0878	0.0534	0.0661	0.0822	0.0822
(36__Main TS)	0.0648	0.0629	0.0606	0.0661	0.0637	0.0660	0.0873	0.0531	0.0656	0.0811	0.0811
(35__Main TS)	0.0659	0.0643	0.0604	0.0672	0.0648	0.0672	0.0893	0.0546	0.0671	0.0818	0.0818

Table C.3 (Continue) IAE result of DIB Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(34__Main TS)	0.0638	0.0630	0.0603	0.0649	0.0630	0.0660	0.0890	0.0542	0.0664	0.0785	0.0785
(33__Main TS)	0.0613	0.0614	0.0602	0.0623	0.0607	0.0644	0.0887	0.0537	0.0655	0.0747	0.0747
(32__Main TS)	0.0584	0.0595	0.0600	0.0592	0.0581	0.0625	0.0885	0.0531	0.0645	0.0703	0.0703
(31__Main TS)	0.0551	0.0574	0.0599	0.0557	0.0551	0.0603	0.0884	0.0525	0.0634	0.0655	0.0655
(30__Main TS)	0.0515	0.0550	0.0598	0.0519	0.0518	0.0579	0.0885	0.0518	0.0621	0.0605	0.0605
(29__Main TS)	0.0478	0.0524	0.0597	0.0481	0.0483	0.0553	0.0887	0.0511	0.0608	0.0554	0.0554
(28__Main TS)	0.0440	0.0498	0.0596	0.0442	0.0448	0.0527	0.0891	0.0504	0.0594	0.0505	0.0505
(27__Main TS)	0.0404	0.0472	0.0595	0.0405	0.0415	0.0501	0.0897	0.0498	0.0581	0.0459	0.0459
(26__Main TS)	0.0372	0.0448	0.0594	0.0373	0.0384	0.0477	0.0905	0.0492	0.0568	0.0420	0.0420
(25__Main TS)	0.0335	0.0421	0.0593	0.0335	0.0349	0.0450	0.0920	0.0487	0.0556	0.0374	0.0374
(24__Main TS)	0.0303	0.0398	0.0592	0.0304	0.0319	0.0427	0.0938	0.0484	0.0545	0.0336	0.0336
(23__Main TS)	0.0279	0.0380	0.0591	0.0280	0.0297	0.0409	0.0961	0.0483	0.0538	0.0309	0.0309
(22__Main TS)	0.0264	0.0368	0.0590	0.0266	0.0282	0.0398	0.0988	0.0486	0.0535	0.0293	0.0293
(21__Main TS)	0.0259	0.0364	0.0589	0.0262	0.0277	0.0394	0.1018	0.0493	0.0538	0.0297	0.0297
(20__Main TS)	0.0264	0.0368	0.0588	0.0269	0.0281	0.0398	0.1052	0.0502	0.0545	0.0311	0.0311
(19__Main TS)	0.0279	0.0379	0.0587	0.0286	0.0295	0.0410	0.1086	0.0515	0.0558	0.0331	0.0331

Table C.3 (Continue) IAE result of DIB Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(18__Main TS)	0.0303	0.0398	0.0586	0.0311	0.0318	0.0429	0.1119	0.0530	0.0574	0.0357	0.0357
(17__Main TS)	0.0335	0.0422	0.0586	0.0344	0.0347	0.0453	0.1147	0.0546	0.0593	0.0390	0.0390
(16__Main TS)	0.0370	0.0449	0.0585	0.0380	0.0381	0.0481	0.1170	0.0561	0.0612	0.0436	0.0436
(15__Main TS)	0.0408	0.0477	0.0584	0.0419	0.0416	0.0508	0.1183	0.0573	0.0630	0.0486	0.0486
(14__Main TS)	0.0446	0.0503	0.0583	0.0458	0.0451	0.0535	0.1185	0.0581	0.0645	0.0535	0.0535
(13__Main TS)	0.0482	0.0527	0.0582	0.0494	0.0484	0.0560	0.1177	0.0585	0.0656	0.0583	0.0583
(12__Main TS)	0.0515	0.0548	0.0582	0.0527	0.0514	0.0581	0.1158	0.0585	0.0663	0.0627	0.0627
(11__Main TS)	0.0543	0.0565	0.0581	0.0556	0.0540	0.0598	0.1130	0.0582	0.0666	0.0666	0.0666
(10__Main TS)	0.0567	0.0579	0.0580	0.0581	0.0562	0.0612	0.1095	0.0576	0.0666	0.0701	0.0701
(9__Main TS)	0.0587	0.0589	0.0579	0.0601	0.0580	0.0622	0.1056	0.0569	0.0663	0.0729	0.0729
(8__Main TS)	0.0604	0.0597	0.0579	0.0618	0.0595	0.0630	0.1015	0.0560	0.0659	0.0754	0.0754
(7__Main TS)	0.0617	0.0603	0.0578	0.0631	0.0606	0.0636	0.0974	0.0552	0.0654	0.0773	0.0773
(6__Main TS)	0.0627	0.0606	0.0577	0.0642	0.0615	0.0640	0.0935	0.0543	0.0648	0.0789	0.0789
(5__Main TS)	0.0635	0.0609	0.0577	0.0650	0.0622	0.0643	0.0900	0.0535	0.0642	0.0802	0.0802
(4__Main TS)	0.0642	0.0610	0.0576	0.0656	0.0627	0.0645	0.0867	0.0528	0.0637	0.0810	0.0810
(3__Main TS)	0.0644	0.0611	0.0576	0.0658	0.0629	0.0645	0.0837	0.0521	0.0631	0.0778	0.0778
(2__Main TS)	0.0586	0.0594	0.0575	0.0577	0.0590	0.0627	0.0751	0.0497	0.0605	0.0875	0.0875
(1__Main TS)	0.0759	0.0376	0.0575	0.0937	0.0587	0.0423	0.1165	0.0601	0.0652	0.2439	0.2439

Table C.4 IAE result of DP Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(30__Main TS)	0.0399	0.0461	0.1062	0.1026	0.1081	0.1037	0.2049	0.1139	0.1308	0.0891	0.0891
(29__Main TS)	0.0396	0.0538	0.1249	0.1221	0.1277	0.1233	0.2244	0.1343	0.1513	0.1076	0.1076
(28__Main TS)	0.0397	0.0608	0.1563	0.1541	0.1598	0.1553	0.2567	0.1668	0.1848	0.1380	0.1380
(27__Main TS)	0.0405	0.0641	0.2037	0.2008	0.2068	0.2021	0.3041	0.2118	0.2333	0.1823	0.1823
(26__Main TS)	0.0418	0.0621	0.2649	0.2580	0.2647	0.2595	0.3630	0.2632	0.2924	0.2358	0.2358
(25__Main TS)	0.0429	0.0565	0.3289	0.3133	0.3213	0.3151	0.4217	0.3082	0.3496	0.2849	0.2849
(24__Main TS)	0.0431	0.0503	0.3800	0.3521	0.3624	0.3543	0.4665	0.3349	0.3915	0.3126	0.3126
(23__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
(22__Main TS)	0.0422	0.0418	0.4088	0.3550	0.3748	0.3592	0.4960	0.3252	0.4131	0.2773	0.2773
(21__Main TS)	0.0418	0.0395	0.3884	0.3245	0.3516	0.3301	0.4883	0.2978	0.4001	0.2296	0.2296
(20__Main TS)	0.0416	0.0382	0.3512	0.2807	0.3157	0.2877	0.4736	0.2628	0.3780	0.1803	0.1803
(19__Main TS)	0.0415	0.0374	0.3034	0.2315	0.2716	0.2391	0.4557	0.2245	0.3496	0.1392	0.1392
(18__Main TS)	0.0415	0.0369	0.2514	0.1851	0.2249	0.1922	0.4368	0.1876	0.3162	0.1096	0.1096
(17__Main TS)	0.0416	0.0363	0.1610	0.1211	0.1475	0.1254	0.3980	0.1330	0.2379	0.0784	0.0784
(16__Main TS)	0.0415	0.0365	0.2022	0.1476	0.1819	0.1535	0.4177	0.1563	0.2783	0.0903	0.0903
(15__Main TS)	0.0417	0.0362	0.1304	0.1040	0.1233	0.1071	0.3760	0.1175	0.1993	0.0715	0.0715

Table C.4 (Continue) IAE result of DP Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(14__Main TS)	0.0418	0.0360	0.1098	0.0936	0.1077	0.0959	0.3493	0.1080	0.1671	0.0675	0.0675
(13__Main TS)	0.0419	0.0359	0.0969	0.0875	0.0983	0.0892	0.3158	0.1023	0.1438	0.0651	0.0651
(12__Main TS)	0.0421	0.0358	0.0893	0.0840	0.0928	0.0854	0.2776	0.0991	0.1287	0.0637	0.0637
(11__Main TS)	0.0418	0.0357	0.0848	0.0819	0.0895	0.0831	0.2420	0.0972	0.1196	0.0628	0.0628
(10__Main TS)	0.0361	0.0356	0.0807	0.0800	0.0873	0.0812	0.2154	0.0960	0.1142	0.0604	0.0604
(9__Main TS)	0.0360	0.0355	0.0806	0.0799	0.0871	0.0810	0.2141	0.0957	0.1138	0.0604	0.0604
(8__Main TS)	0.0359	0.0355	0.0805	0.0798	0.0870	0.0809	0.2127	0.0954	0.1135	0.0603	0.0603
(7__Main TS)	0.0357	0.0355	0.0805	0.0797	0.0869	0.0808	0.2111	0.0951	0.1132	0.0603	0.0603
(6__Main TS)	0.0355	0.0355	0.0805	0.0796	0.0868	0.0808	0.2094	0.0948	0.1129	0.0604	0.0604
(5__Main TS)	0.0350	0.0354	0.0807	0.0796	0.0867	0.0808	0.2075	0.0945	0.1125	0.0608	0.0608
(4__Main TS)	0.0343	0.0354	0.0809	0.0797	0.0867	0.0808	0.2052	0.0941	0.1122	0.0612	0.0612
(3__Main TS)	0.0330	0.0353	0.0814	0.0798	0.0867	0.0810	0.2025	0.0937	0.1118	0.0618	0.0618
(2__Main TS)	0.0300	0.0352	0.0820	0.0800	0.0868	0.0812	0.1993	0.0933	0.1114	0.0626	0.0626
(1__Main TS)	0.0330	0.0353	0.0814	0.0798	0.0867	0.0810	0.2025	0.0937	0.1118	0.0618	0.0618

Table C.4 (Continue) IAE result of DP Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(30__Main TS)	0.0911	0.1088	0.1311	0.0945	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0911
(29__Main TS)	0.1101	0.1282	0.1511	0.1144	0.1221	0.1277	0.1233	0.2245	0.1344	0.1514	0.1101
(28__Main TS)	0.1412	0.1603	0.1839	0.1466	0.1541	0.1599	0.1554	0.2567	0.1668	0.1849	0.1412
(27__Main TS)	0.1862	0.2075	0.2317	0.1919	0.2008	0.2069	0.2022	0.3042	0.2118	0.2334	0.1862
(26__Main TS)	0.2402	0.2663	0.2901	0.2446	0.2580	0.2648	0.2596	0.3631	0.2632	0.2924	0.2402
(25__Main TS)	0.2899	0.3247	0.3474	0.2901	0.3134	0.3214	0.3152	0.4217	0.3082	0.3496	0.2899
(24__Main TS)	0.3194	0.3683	0.3900	0.3132	0.3521	0.3624	0.3543	0.4665	0.3350	0.3916	0.3194
(23__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0000
(22__Main TS)	0.2922	0.3866	0.4137	0.2735	0.3551	0.3748	0.3592	0.4960	0.3252	0.4131	0.2922
(21__Main TS)	0.2469	0.3659	0.4015	0.2263	0.3245	0.3517	0.3301	0.4884	0.2979	0.4002	0.2469
(20__Main TS)	0.1967	0.3317	0.3799	0.1785	0.2808	0.3157	0.2877	0.4736	0.2629	0.3780	0.1967
(19__Main TS)	0.1526	0.2884	0.3519	0.1392	0.2315	0.2717	0.2392	0.4557	0.2246	0.3497	0.1526
(18__Main TS)	0.1198	0.2409	0.3192	0.1114	0.1851	0.2250	0.1923	0.4369	0.1876	0.3163	0.1198
(17__Main TS)	0.0845	0.1575	0.2431	0.0828	0.1211	0.1475	0.1255	0.3980	0.1330	0.2380	0.0845
(16__Main TS)	0.0980	0.1953	0.2825	0.0936	0.1477	0.1819	0.1535	0.4178	0.1563	0.2784	0.0980
(15__Main TS)	0.0766	0.1300	0.2046	0.0765	0.1040	0.1233	0.1071	0.3761	0.1176	0.1993	0.0766

Table C.4 (Continue) IAE result of DP Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(14__Main TS)	0.0720	0.1119	0.1718	0.0728	0.0937	0.1078	0.0959	0.3493	0.1080	0.1671	0.0720
(13__Main TS)	0.0693	0.1008	0.1476	0.0707	0.0876	0.0983	0.0893	0.3158	0.1024	0.1439	0.0693
(12__Main TS)	0.0677	0.0943	0.1315	0.0694	0.0840	0.0928	0.0854	0.2777	0.0991	0.1288	0.0677
(11__Main TS)	0.0667	0.0905	0.1218	0.0686	0.0820	0.0896	0.0832	0.2421	0.0972	0.1196	0.0667
(10__Main TS)	0.0649	0.0879	0.1160	0.0667	0.0801	0.0873	0.0812	0.2154	0.0960	0.1142	0.0649
(9__Main TS)	0.0648	0.0877	0.1157	0.0666	0.0799	0.0872	0.0811	0.2141	0.0957	0.1139	0.0648
(8__Main TS)	0.0647	0.0876	0.1153	0.0665	0.0798	0.0871	0.0810	0.2127	0.0955	0.1136	0.0647
(7__Main TS)	0.0647	0.0875	0.1149	0.0665	0.0797	0.0869	0.0809	0.2112	0.0952	0.1133	0.0647
(6__Main TS)	0.0648	0.0874	0.1145	0.0666	0.0797	0.0868	0.0808	0.2095	0.0949	0.1129	0.0648
(5__Main TS)	0.0650	0.0873	0.1141	0.0667	0.0797	0.0867	0.0808	0.2075	0.0945	0.1126	0.0650
(4__Main TS)	0.0652	0.0873	0.1137	0.0670	0.0797	0.0867	0.0809	0.2052	0.0942	0.1122	0.0652
(3__Main TS)	0.0656	0.0873	0.1132	0.0674	0.0799	0.0867	0.0810	0.2025	0.0938	0.1119	0.0656
(2__Main TS)	0.0662	0.0874	0.1126	0.0680	0.0801	0.0868	0.0812	0.1993	0.0934	0.1115	0.0662
(1__Main TS)	0.0656	0.0873	0.1132	0.0674	0.0799	0.0867	0.0810	0.2025	0.0938	0.1119	0.0656

Table C.5 IAE result of DB Column

Tray	V1	V2	V3	V4	V5	V7	V8	V9	V11	V12	V14
(15__Main TS)	0.2366	0.1413	0.3459	0.0672	0.3176	0.3302	0.4665	0.3176	0.1413	0.3938	0.3938
(14__Main TS)	0.2025	0.2241	0.8124	0.0672	0.6299	0.6616	0.2308	0.3811	0.2241	0.8429	0.8429
(13__Main TS)	0.1669	0.1501	0.9701	0.0672	0.6799	0.7209	0.0707	0.3347	0.1501	1.0000	1.0000
(12__Main TS)	0.1574	0.0828	0.9496	0.0672	0.6315	0.6730	0.0953	0.2855	0.0828	0.9800	0.9800
(11__Main TS)	0.1560	0.0618	0.8779	0.0672	0.5543	0.6015	0.1007	0.2390	0.0618	0.9301	0.9301
(10__Main TS)	0.1564	0.0576	0.8204	0.0672	0.5003	0.5526	0.1014	0.2073	0.0576	0.8935	0.8935
(9__Main TS)	0.1572	0.0573	0.7776	0.0672	0.4615	0.5168	0.1015	0.1840	0.0573	0.8651	0.8651
(8__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
(7__Main TS)	0.1585	0.0584	0.6306	0.0664	0.3269	0.3849	0.1015	0.1286	0.0584	0.7489	0.7489
(6__Main TS)	0.1067	0.1408	0.7386	0.0658	0.3778	0.4488	0.1142	0.1266	0.1408	0.8860	0.8860
(5__Main TS)	0.0433	0.2221	0.7552	0.0650	0.3743	0.4483	0.1078	0.0981	0.2221	0.9017	0.9017
(4__Main TS)	0.0318	0.0956	0.7509	0.0642	0.3685	0.4436	0.1061	0.0927	0.0956	0.9003	0.9003
(3__Main TS)	0.0320	0.0129	0.7387	0.0632	0.3568	0.4322	0.1061	0.0849	0.0129	0.8895	0.8895
(2__Main TS)	0.0321	0.0103	0.7097	0.0621	0.3313	0.4057	0.1065	0.0748	0.0103	0.8635	0.8635
(1__Main TS)	0.0321	0.0116	0.5743	0.0607	0.2267	0.2905	0.1070	0.0569	0.0116	0.7376	0.7376

Table C.5 (Continue) IAE result of DB Column

Tray	V22	V31	V32	QC1	QC2	QC3	QR1	QR2	QR3	WKCOMP	QCOND
(15__Main TS)	0.3594	0.3507	0.3573	0.3785	0.1260	0.1109	0.0874	0.0645	0.1322	0.3272	0.3594
(14__Main TS)	0.7347	0.6677	0.4821	0.7570	0.1105	0.0579	0.0617	0.0630	0.0608	0.1077	0.7347
(13__Main TS)	0.8414	0.7126	0.4341	0.8571	0.1342	0.0000	0.0000	0.0000	0.0000	0.0000	0.8414
(12__Main TS)	0.8020	0.6639	0.3940	0.8217	0.0679	0.0507	0.0642	0.0760	0.0576	0.0643	0.8020
(11__Main TS)	0.7478	0.5946	0.3390	0.7692	0.1190	0.0571	0.0732	0.0872	0.0653	0.0681	0.7478
(10__Main TS)	0.7109	0.5470	0.2975	0.7329	0.0655	0.0395	0.0489	0.0579	0.0494	0.0908	0.7109
(9__Main TS)	0.6823	0.5119	0.2655	0.7048	0.0594	0.0486	0.0000	0.0722	0.0553	0.0618	0.6823
(8__Main TS)	0.0000	0.0000	0.0000	0.0000	0.0698	0.0537	0.0000	0.0811	0.0612	0.0647	0.0000
(7__Main TS)	0.5608	0.3799	0.1734	0.5861	0.1071	0.0507	0.0000	0.0000	0.0576	0.0644	0.5608
(6__Main TS)	0.6617	0.4426	0.1833	0.6939	0.1501	0.0001	0.0000	0.0057	0.0002	0.0002	0.6617
(5__Main TS)	0.6683	0.4435	0.1694	0.7018	0.1289	0.1107	0.0000	0.0644	0.1321	0.3270	0.6683
(4__Main TS)	0.6661	0.4389	0.1629	0.6997	0.0391	0.0897	0.0000	0.5305	0.1046	0.1160	0.6661
(3__Main TS)	0.6555	0.4277	0.1531	0.6891	0.1318	0.0396	0.0000	0.0561	0.0498	0.0753	0.6555
(2__Main TS)	0.6288	0.4014	0.1372	0.6629	0.1307	0.0570	0.0000	0.0869	0.0652	0.0682	0.6288
(1__Main TS)	0.5020	0.2872	0.0913	0.5372	0.0496	0.1158	0.0000	0.1466	0.1332	0.1633	0.5020

VITA

Ms. Panisara Khamanarm graduated Bachelor Degree in department of Chemical from Chandrakasem University in 2004. After that she studied for master degree in Chemical Engineering and joined Control and Systems Engineering research group at Chulalongkorn University in 2011.