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สำหรับกระบวนการอัลคิลเลชัน

นางสาวเสาวนีย์ เดชเจริญสี

PLANTWIDE CONTROL STRUCTURES DESIGN

FOR ALKYLATION PROCESS

Miss Saowani Detjareansri

A Thesis Submitted in Partial Fulfillment of the Requirements

for the Degree of Master of Engineering Program in Chemical Engineering

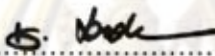Department of Chemical Engineering

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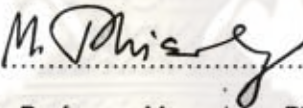Chulalongkorn University

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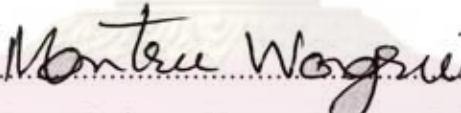
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เสาวนีย์ เดชเจริญสี : การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการอัลคิเลชัน. (PLANTWIDE CONTROL STRUCTURES DESIGN FOR ALKYLATION PROCESS) อ. ที่ปรึกษาวิทยานิพนธ์หลัก : ผศ. ดร. มนตรี วงศ์ศรี, 212 หน้า.

การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์มีความซับซ้อน เนื่องมาจากในกระบวนการมีการนำสารตั้งต้นและพลังงานกลับมาใช้ใหม่ เพื่อให้เกิดความคุ้มค่าในการดำเนินการผลิตมากที่สุด ดังนั้นในงานวิจัยนี้จึงได้ศึกษาการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์ตามขั้นตอนของวงศ์ศรี (2009) มาพัฒนาโครงสร้างการควบคุมสำหรับกระบวนการอัลคิเลชัน เพื่อให้เป็นไปตามวัตถุประสงค์ของการควบคุม โดยกระบวนการอัลคิเลชันเป็นกระบวนการที่มีความซับซ้อนประกอบไปด้วยหน่วยปฏิบัติการหลายหน่วย มีปฏิกิริยาหลักที่เกิดขึ้นคือปฏิกิริยาระหว่างไอโซบิวเทนและบิวทีน ทำให้ได้ผลิตภัณฑ์คือไอโซออกเทน ซึ่งไอโซออกเทนนั้นมีความสำคัญในการนำไปผลิตเป็นแก๊สโซลีนที่ใช้กันอย่างกว้างขวางในปัจจุบัน โดยงานวิจัยนี้ใช้โปรแกรมไฮซีสเพื่อจำลองกระบวนการอัลคิเลชันทั้งสภาวะคงตัวและสภาวะพลวัต จากนั้นทำการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการอัลคิเลชันทั้งหมด 8 โครงสร้าง โดยใช้ขั้นตอนการออกแบบของวงศ์ศรี (2009) และประเมินสมรรถนะของโครงสร้างการควบคุมที่ออกแบบเปรียบเทียบกับโครงสร้างการควบคุมพื้นฐานของ ของลูเบน (2002) โดยใช้ตัวรบกวนกระบวนการ 2 ชนิด ได้แก่ การรบกวนอัตราการไหลของสาร และการรบกวนทางความร้อน พบว่าโครงสร้างที่ออกแบบมีสมรรถนะที่ดี เนื่องจากสามารถกำจัดสิ่งรบกวนที่เข้าสู่ระบบ และสามารถรักษาคุณภาพของผลิตภัณฑ์ได้ดี โดยเปรียบเทียบจากปริพันธ์ของค่าคลาดเคลื่อนสัมบูรณ์ และพลังงานที่ใช้ทั้งหมดน้อยกว่า

ดังนั้นจึงสรุปได้ว่าการออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์โดยใช้ขั้นตอนของวงศ์ศรี (2009) ซึ่งรวมวิธีฮิวริสติกส์ วิธีการวิเคราะห์ และการจำลองแบบพลวัต เป็นวิธีที่นำไปสู่โครงสร้างการควบคุมที่มีสมรรถนะที่ดี

ภาควิชา...วิศวกรรมเคมี.................. ลายมือชื่อนิสิต เสาวนีย์ เดชเจริญสี

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

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The plantwide control structure design is complex because of material and energy recycles designed for economic. This research used plantwide control procedure of Wongsri (2009) to develop the control structures for alkylation process. The alkylation process is a complex multi-unit process, which features several unit operations. The main reaction is the combination of isobutane and butene to form iso-octane. Iso-octane is valuable for making gasoline. In this research using software HYSYS to simulate alkylation process at steady state and dynamic. Then we design eight plantwide control structures (CS1 to CS8) for alkylation process using new design procedure of Wongsri (2009) and evaluate the dynamic performance of the designed control structures compare with base case control structure (Luyben, 2002) by two types of disturbances: material and thermal disturbances. The designed control structure has a good performance because it can handle disturbances entering the process and can maintain product quality as compared by integral absolute error (IAE) and total energy use low.

Therefore this research establishes that the Wongsri's procedure, which combines heuristics, analytical method and dynamic simulation, a useful design procedure that leads to a good-performance plantwide control system.

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

Field of Study : Chemical Engineering...... Advisor's Signature ....

Academic Year : 2009...............................

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CHAPTER I

INTRODUCTION

1.1 Importance and Reasons for Research

Nowadays, industries are very competitive both in quality and cost of production. Therefore, production process should have high quality and high efficiency. The process should always operate under the design condition, use little energy, low waste production and meet the required specification of the products. In the real situation, the process will not operate smoothly. All factors do not meet the designed conditions. The process always changes due to disturbance from the external factors and the internal factors. However, not matter what factors cause the change, in case of having deviation or disturbance come in to the process, the effect should be eliminated from the process as soon as possible so that the process will have the least deviation from the designed condition. Moreover, due to the restriction of the environment, safety and operating condition, it is very necessary to have the control system to control the condition and compensate for any deviation occurred.

Previously, the control system of the process is designed to control only each individual unit. In general, industries have an integrated process (material recycle and energy integrations). Individual control systems for each unit operation are not available because they increase the interactions among the process units. The design process control using plantwide can resolve the previous problem. Because plantwide process control design a control system from the viewpoint of the entire plant.

An important problem in process control is to develop effective control structures for complex multi-unit process. In this research, it will focus on new plantwide control design procedure of Montree Wongsri (2009) applied to alkylation process. The alkylation process of isobutane with butene to from iso-octane is a widely used method for producing a high-octane blending component for gasoline. Iso-octane has an octane number of 100, so it is valuable for making gasoline for high-compressor engines that require high octane number (suppress pre-ignition). The alkylation process was commercialized just before World War II, and it provided high-octane gasoline for the

airplanes in that historic conflict. It is still widely used around the world in many refineries as a way to upgrade light component and to produce a high-valve, non-aromatic gasoline blending material.

1.2 Research Objectives

The objective of this research is:

To design plantwide control structures for alkylation process using new design procedure of Wongsri (2009).

1.3 Scopes of Research

The Scopes of this work are listed below:

1. Simulation of the alkylation process is performed by using a commercial process simulator – HYSYS (version 3.1).

2. Description and data of alkylation process is obtained from William L. Luyben, 2002.

3. Plantwide control structures for alkylation process are designed using new design procedure of Wongsri (2009).

4. New two plantwide control structures for alkylation process are designed.

1.4 Contributions of Research

The contributions of this work are as follows:

1. Steady state and dynamic HYSYS model for Alkylation process.

2. The new plantwide control structures of alkylation process are designed and compared with the work given by William L. Luyben, 2002.

3. Evaluation of the new plantwide control structures design procedure.

1.5 Research Procedures

The procedures of this research are as follows:

1. Study of plantwide process control theory.

2. Study of alkylation process and concerned information.

3. Simulations of alkylation process at steady state and dynamic.

4. Study the new design procedure.

5. Design new plantwide control structures follow the new design procedure.

6. Simulations of alkylation process at dynamic.

7. Evaluate the dynamic performance of the designed control structures.

8. Analyze the design and simulation results.

9. Conclude of the thesis.

1.6 Research Contents

This thesis is devided into six chapters:

Chapter I is an introduction to this research. This chapter consists of importance and reasons for research, research objectives, scopes of research, contributions of research and research procedures.

Chapter II reviews the work carried out on plantwide control design and related to plantwide control structure design procedure and the method of selection set of controlled variables.

Chapter III cover some background information of Luyben's theory concerning with plantwide control and new plantwide control structure design procedure of Wongsri (2009).

     Chapter IV describes the process description and process simulation via HYSYS

     Chapter V describes the design of plantwide control structures, dynamic simulation results and compares control structure of Luyben with new control structures.

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

     This is follow by:

References

Appendix A: Tuning of Control Structures

Appendix B: Parameter Tuning

Appendix C: Fixture Point Theorem Data

CHAPTER II

LITERATURE REVIEW

This chapter is a literature review of this research. It presents a review of the previous works on plantwide control design.

Luyben (1994) presented a mathematical analysis of the problem for several typical kinetic systems. In the simple binary first-order case of A→B, an analytical solution can be found for the recycle flow rate as a function of the fresh feed flow rate and fresh feed composition. Two different control structures are explored. It is shown analytically why the control structure proposed by Luyben prevents snowballing and why the conventional structure results in severe snowballing. Two other kinetic systems are studied numerically: consecutive first-order reactions A→B→C and a second-order reaction A+B→C. Results confirm the snowball problems can be prevented by using a control structure that fixes the flow rate of one stream somewhere in a liquid recycle loop. In processes with one recycle, the flow rate of the reactor effluent can be set. In processes with two or more recycle streams, the flow rate of each recycle can be fixed.

Luyben (1996) presented one of the central problems in developing a steady-state process flowsheet is finding the number of variables that must be specified to completely define the process. This number is called the design degrees of freedom. Once this number has been found, the number of design optimization variables can be calculated by subtracting all variables that are set by specifications on production rate, product qualities, safety constraints, and environmental limitations. In principle, the design degrees of freedom are easily calculated by simply subtracting the number of equations from the number of variables. However, for typically complex industrial processes, there are many hundreds of variables and equations, and it is not a trivial job to make sure that the correct variables and equations have been defined. In addition, this conventional variables-minus equations approach requires that a detailed model of the process be available. Once the plant has been specified, the design of a control structure requires that the control degrees of freedom be known. This is the number of

variables that can be controlled. It is very easy to calculate this number, even for quite complex processes, because it is equal to the number of manipulated variables (the number of control valves in the process). These variables are different than the design optimization variables. He illustrated that the number of design degrees of freedom is equal to the number of control degrees of freedom for an important class of processes. For a much broader class of processes a slight modification of this equality must be used. Several progressively more complex recycle process case studies are used to demonstrate these results. The practical significance of this approach is that we do not need a model and we can avoid the tedious and error-prone procedure of accounting for all variables and equations.

Luyben, Tyreus and Luyben (1997) presented a general heuristic design procedure. Their procedure generated an effective plantwide control structure for an entire complex process flowsheet and not simply individual units. The nine step of the proposed procedure center around the fundamental principles of plantwide control: energy management, production rate, product quality, operational, environmental and safety constraints, liquid level and gas-pressure inventories, makeup of reactants, component balances and economic or process optimization. Application of the procedure was illustrated with three industrial examples: the vinyl acetate monomer process, Eastman process and HDA process. The procedure produced a workable plantwide control strategy for a given process design.

Mahajanam, Zheng, and Douglas (2001) proposed a shortcut method to eliminate poor choices and to generate and rank attractive alternatives without solving optimization problems. The method is based on scaling of all of the candidate controlled variables, so that they have similar effects on the steady-state profit. The procedure is illustrated on a simplified butane alkylation process.

Larsson, Govatsmark, Skogestad and Yu (2003) presented control structure selection for a simple plant with a liquid-phase reactor, a distillation column, and recycle of unreacted reactants. The starting point is a clear definition of the operational objectives, constraints, and degrees of freedom. Active constraints should

be controller to optimize the economic performance. This implies for this case study that the reactor level should be kept at its maximum, that being economically attractive. Maximizing the reactor holdup also minimizes the "snowball effect". The main focus is no the selection of a suitable controlled variable for the remaining unconstrained degree of freedom, that use the concept of self-optimizing control to search for a constant setpoint strategy with an acceptable economic loss. Both for the case with a given feed rate where the energy costs should be minimized and for the case where the production rate should be maximized, they find that a good controlled variable is the reflux ratio L/F. This applies to single-loop control as well as multivariable model predictive control.

Sumitra Kasemchainun (2003) presented plantwide control strategies for designing control structures of a vinyl acetate monomer plant to achieve the objective of this process and can operate within constraint of safety, environment, and operation. Three alternative plantwide control structures are designed, tested and compared the performance with Luyben's structure. All of control structures can operate within constraint, achieve the objectives and a good control structure was quickly response to disturbance and adjust itself to steady state while minimizing the deviation of the product quality.

Skogestad (2004) interested in control structure design deals with the structural decisions of the control system, including what to control and how to pair the variables to form control loops. He presented a systematic procedure for control structure design for complete chemical plants (plantwide control). It started with carefully defining the operational and economic objectives, and the degrees of freedom available to fulfill them. Other issues, discussed in the paper, include inventory and production rate control, decentralized versus multivariable control, loss in performance by bottom-up design, and a definition of a the "complexity number" for the control system.

Konda, Rangaiah and Krishnaswamy (2005) presented the novel plantwide control (PDC) methodologies are becoming increasingly important as chemical processes are becoming more and more integrated with recycle for reasons of

safety, environmental considerations, and economics. Hence, in the present work, an integrated framework of simulation and heuristics is proposed. The main emphasis here is on vertical integration of simulation and heuristics which exploits the inherent interlink between them. By adopting third framework, simulators can be more efficiently utilized and they also offer invaluable support to the decisions taken by heuristics. The proposed framework is then successfully applied to an industrially relevant case study: the hydrodealkylation of toluene (HDA) process. An analysis of results shows that the proposed framework builds synergies between the powers of both the simulation and the heuristics, thereby resulting in a practical PWC methodology that leads to a viable control system.

Chotirat Kiatpiriya (2007) presented a plantwide control design procedure base on basic idea of self-optimizing control to select controlled variables which when kept constant lead to minimum economic loss. The maximum scaled gain is used to selecting and paring controlled variables with manipulated variables. In her study, three control structures were designed and compared. In order to illustrate the dynamic behaviors of the control structures in reaction section of HDA plant when economic disturbance load occurred. All of control structures can operate to achieve the objective and within process constraints. The performance of designed control structures were presented in IAE value and compared with reference structure. The designed structures are faster response than reference structure and can handle disturbance of large heat exchanger which small furnace duty.

Suchada Suntisrikomol (2008) presented the "Fixture Point Theorem" for Hydrodealkylation process (HDA) to select appropriate the set of controlled variables from a large number of candidate output. The fixture point control theorem states that the most disturbed points must be satisfactorily controlled by giving them consideration before other controlled variables. The maximum (scaled) gain is used to selecting and paring controlled variables with manipulated variables. In her study, the set of first rank of controlled variables are same as Luyben (1998). She selected three set of controlled variables (second and third rank from fixture point) and five control structures were designed and compared. In order to illustrate the dynamic behaviors of the control

structures when economic disturbance load occur (such as change in methane composition in fresh feed gas and quencher outlet temperature), the performance of designed control structures were presented in IAE value and compared with reference structure. The designed structures are fast response and the most effective on compared with Araujo et al, 2006 and Luyben, 1998.

CHAPTER III

THEORY

3.1 The main function of control system

In general, the control system installed in process has main function.

3.1.1 To reject disturbance

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

3.1.2 To maintain stability

The stability is necessary for every process. As a result the control system is set to improve the process stability for the guarantee of quality of product, safety to equipment of process and plant.

3.1.3 To keep the process performing highest efficiency

Besides rejecting disturbance and maintaining stability, the control system can achieve the great profit because it losses less energy and raw materials during the operating. Moreover the product will meet the required specification and have high production rate.

3.2 Integrated Process

Three basic features of integrated chemical process lie at the root of the need to consider the entire plant's control system.

(1) The effect of material recycle

(2) The effect of energy integration

(3) The need to account for chemical component inventories.

If these issues did not have to worry about, then a complex plantwide control problem was not had to deal with. However, there are fundamental reasons why each there exists in virtually all real processes.

3.2.1 Material Recycles

Material is recycled for six basic and important reasons.

1. Increase conversion. For chemical processes involving reversible reactions, conversion of reactants to products is limited by thermodynamic equilibrium constraints. Therefore the reactor effluent by necessity contains both reactants and products. Separation and recycle of reactants are essential if the process is to be economically viable.

2. Improve economics. In most systems it is simply cheaper to build a reactor with incomplete conversion and recycle reactants than it is to reach the necessary conversion level in one reactor or several in series. A reactor followed by a stripping column with recycle is cheaper than one large reactor or three reactors in series.

3. Improve yields. In reaction system such as A $\longrightarrow$ B $\longrightarrow$ C, where B is the desired product, the per-pass conversion of A must be kept low to avoid producing too much of the undesirable product C. Therefore the concentration of B is kept fairly low in the reactor and a large recycle of A is required.

4. Provide thermal sink. In adiabatic reactors and in reactors where cooling is difficult and exothermic heat effects are large, it is often necessary to feed excess material to the reactor (an excess of one reactant or a product) so that the reactor temperature increase will not be too large. High temperature can potentially create several unpleasant events: it can lead to thermal runaways, it can deactivate catalysts, it can cause undesirable side reactions, it can cause mechanical failure of

equipment, etc. So the heat of reaction is absorbed by the sensible heat required to rise the temperature of the excess material in the stream flowing through the reactor.

5. Prevent side reactions. A large excess of one of the reactants is often used so that the concentration of the other reactant is kept low. If this limiting reactant is not kept in low concentration, it could react to produce undesirable products. Therefore the reactant that is in excess must be separated from the product components in the reactor effluent stream and recycled back to the reactor.

6. Control properties. In many polymerization reactors, conversion of monomer is limited to achieve the desired polymer properties. These include average molecular weight, molecular weight distribution, degree of branching, particle size, etc. Another reason for limiting conversion to polymer is to control the increase in viscosity that is typical of polymer solutions. This facilitates reactor agitation and heat removal and allows the material to be further processed.

3.2.2 Energy Integration

The fundamental reason for the use of energy integration is to improve the thermodynamics efficiency of the process. This translates into a reduction in utility cost. For energy-intensive processes, the savings can be quite significant.

3.2.3 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 Plantwide Control Problem

3.3.1 Units in series

If process units are arranged in a purely series configuration, where the products of each unit feed downstream units and there is no recycle of material or energy, the plantwide control problem is greatly simplified. We do not have to worry about the issues discussed in the previous section and it can be simply configure the control scheme on each individual unit operation to handle load disturbances.

If production rate is set at the front end of process, each unit will only see load disturbances coming from its upstream neighbor. If the plant is set up for "on-demand" production, changes in throughput will propagate back through the process. So any individual unit will see load disturbances coming from both its downstream neighbor (flowrate changes to achieve different throughputs) and its upstream neighbor (composition changes as the upstream units adjust to the load changes they see).

Figure 3.1 compares these two possible configurations for a simple plant. A fresh feed stream containing a mixture of chemical components A, B and C is fed into a two-column distillation train. The relation volatility are $\propto_A >\propto_B > \propto_C$ and the "direct" (or "light-out-first") separation sequence is selected: A is taken out the top of the first column and B out the top of the second column.

Figure 3.1 (a) shows the situation where the fresh feed stream is flow-controlled into the process. The inventory loops (liquid levels) in each unit are controlled

by manipulating flows leaving that unit. All disturbances propagate from unit to unit down the series configuration. The only disturbances that each unit sees are changes in its feed conditions.

Figure 3.1 (b) shows the on-demand situation where the flowrate of product C leaving the bottom of the second column is set by the requirements of a downstream unit. Now some of the inventory loops (the base of both columns) are controlled by manipulating the feed into each column.

When the units are arranged in series with no recycles, the plantwide control problem can be electively broken up into the control of each individual unit operation. There is no recycle effect, no coupling, and no feedback of material from downstream to upstream units. The plant's dynamic behavior is governed by the individual unit operations and the only path for disturbance propagation is linear along the process.

(a)
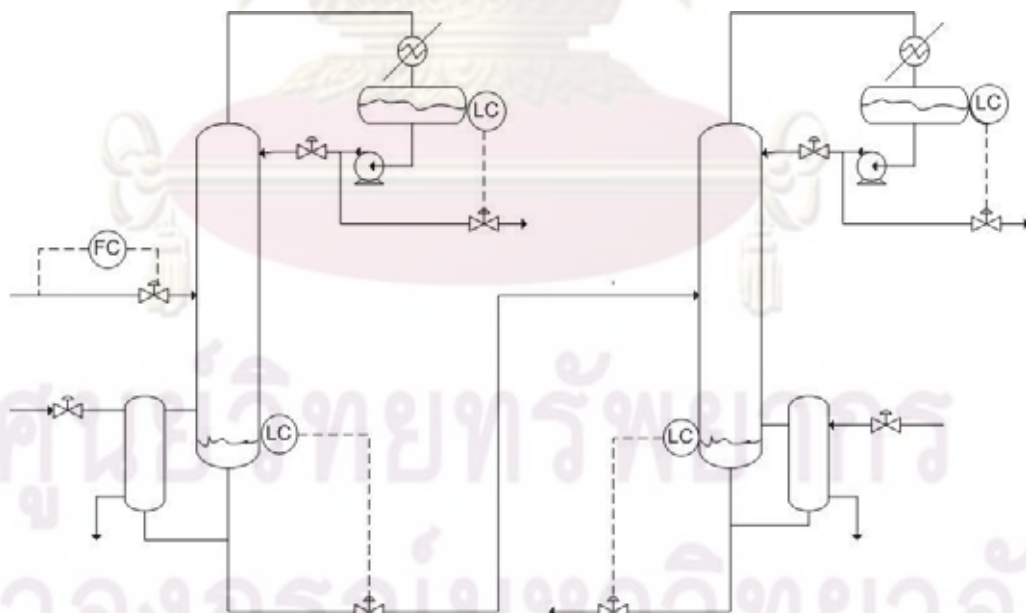
(b)
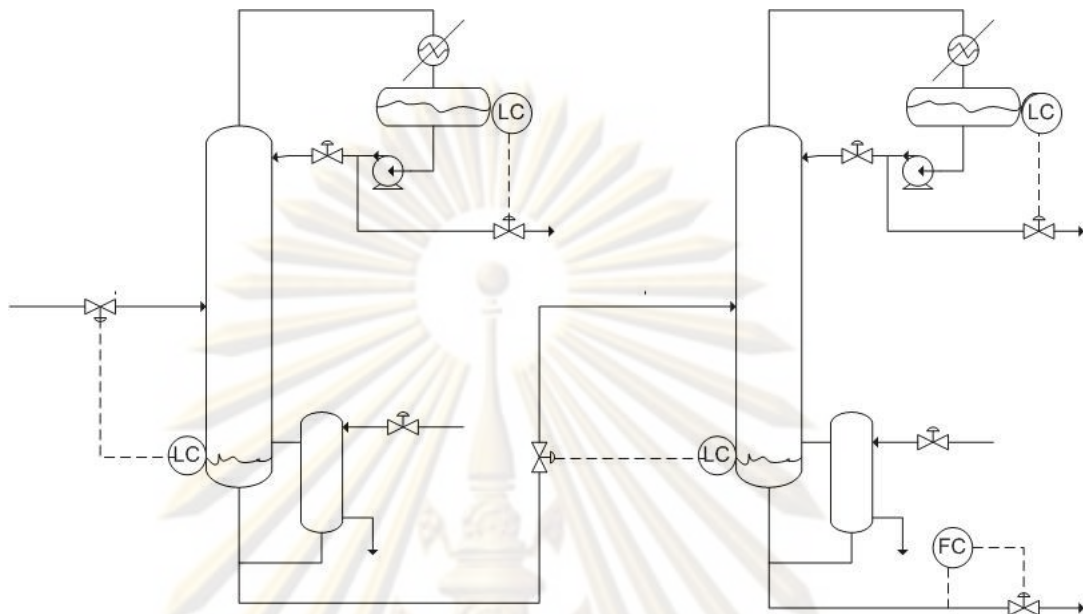
Figure 3.1 Unit in series. (a) Level control in direction of flow; (b) level control in direction opposite flow.

3.3.2 Effects of recycle

Most real processes contain recycle streams. In this case the plantwide control problem becomes much more complex and its solution is not intuitively obvious. The presence of recycle streams profoundly alters the plant's dynamic and steady-state behavior. To gain an understanding of these effects, some very simple recycle systems are looked. The insight they are obtained from these idealized, simplistic systems can be extended to the complex flowsheets of typical chemical processes. First the groundwork must be laid and had some feel for the complexities and phenomena that recycle stream produce in a plant.

Two basic effects of recycle:

(1) Recycle has an impact on the dynamics of the process. The overall time constant can be much different than the sum of the time constants of the individual units.

(2) Recycle leads to the "snowball" effect. This has two manifestations, one steady state and one dynamic. A small change in throughput or feed composition can lead to a large change in steady-state recycle stream flowrates. These disturbances can lead to even larger dynamic changes in flows, which propagate around the recycle loop. Both effects have implications for the inventory control of components.

3.3.3 Snowball effects

Another interesting observation that has been made about recycle systems is their tendency to exhibit large variations in the magnitude of the recycle flows. Plant operators report extended periods of operation when very small recycle flows occur. It is often difficult to turn the equipment down to such low flowrates. Then, during other periods when feed conditions are not very different, recycle flowrates increase drastically, usually over a considerable period of time. Often the equipment cannot handle such a large load.

This high sensitivity of the recycle flowrate to small disturbances is called the snowball effect. It is important to note that this is not a dynamic effect; it is a steady-state phenomenon. But it dose have dynamic implications for disturbance propagation and for inventory control. It has nothing to do with closed-loop stability. However, this does not imply that it is independent of the plant's control structure. On the contrary, the extent of the snowball effect is very strongly dependent upon the control structure used.

The large swings in recycle flowrates are undesirable in a plant because they can overload the capacity of the separation section or move the separation section into a flow region below its minimum turndown. Therefore it is important to select a plantwide control structure that avoids this effect. As the example below illustrates and as more complex processes also show, a very plant wide control heuristic "A stream somewhere in each liquid recycle loop should be flow controlled".

Let us consider one of the simplest recycle processes imaginable: a continuous stirred tank reactor (CSTR) and a distillation column. As shown in Figure 3.2,

a fresh reactant stream is fed into the reactor. Insider the reactor, a first-order isothermal irreversible reaction of component A to produce component B occurs A $\longrightarrow$ B. The specific reaction rate is $k(h^{-1})$ and the reactor holdup is V_R(moles). The fresh feed flowrate is F_0(moles/h) and its composition is z_0(mole fraction component A). the system is binary with only two components: reactant A and product B. the composition in the reactor is Z (mole fraction A). the reactor effluent, with flowrate F (moles/h) is fed into a distillation column that separates unreacted A from product B.
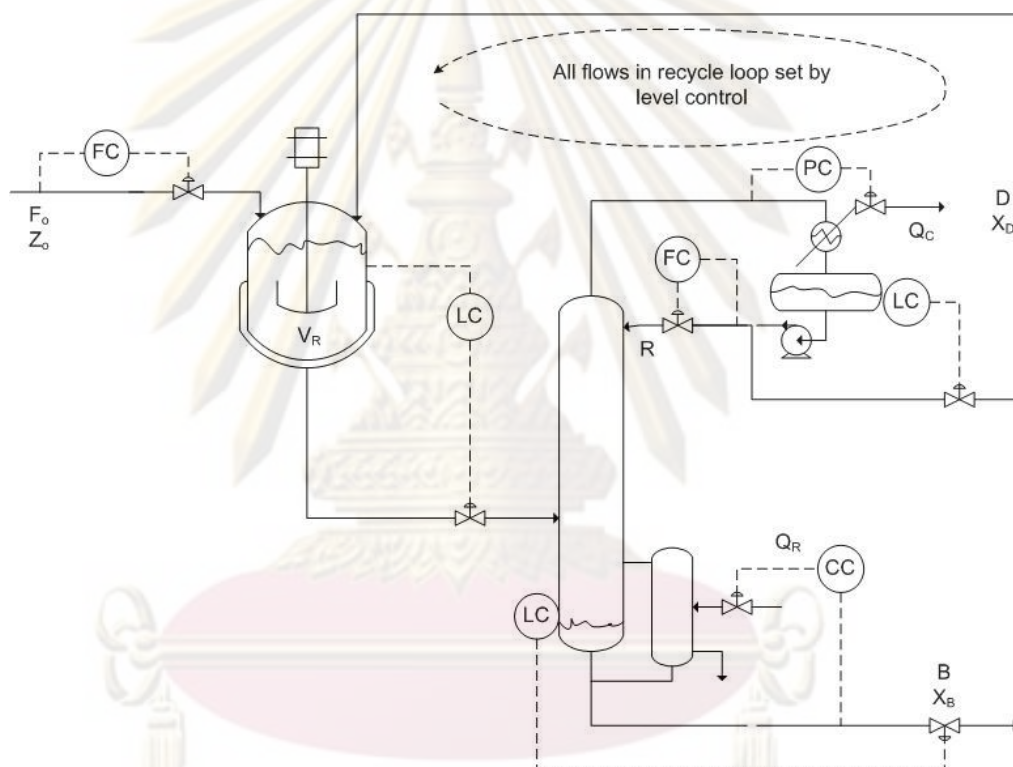
Figure 3.2 Conventional control structure with fixed reactor holdup

The relative volatilities are such that A is more volatile than B, so the bottom from the column is the product stream. Its flowrate is B (moles/h) and its composition is x_B (mole fraction A). The amount of A impurity in this product stream is an important control objective and must be maintained at some specified level to satisfy the product quality requirements of the customer.

The overhead distillate stream from the column contains almost all of component A that leaves the reactor because of the purity specification on the bottoms stream. It is recycled back to the reactor at a flowrate D and with a composition x_D (mole fraction A). The column has N_T trays and the feed tray is N_F (counting from the bottom). The reflux flowrate is R and the vapor boilup is V (moles/h).

Conventional control structure As shown in Figure 3.2, the following control loops are chosen:

1. Fresh feed flow is controlled.

2. Reactor level is controlled by manipulating reactor effluent flow.

3. Bottoms product purity is controlled by manipulating heat input to the reboiler.

4. Distillate purity is controlled by manipulating reflux flow. Note that dual composition control (controlling both distillate and bottoms purities) have been chosen to use in the distillation column, but there is no a priori reason for holding the composition of the recycle stream constant since it does not leave the process. It may be useful to control the composition of this recycle stream for reactor yield purposes or for improved dynamic response. The "best" recycle purity levels in both the design and operation of the plant are been often free to find.

5. Reflux drum level is held by distillate flow (recycle).

6. Base level is held by bottoms flow.

7. Column pressure is controlled by manipulating coolant flowrate to the condenser.

This control scheme is probably what most engineers would devise if given the problem of designing a control structure for this simple plant. Our tendency is to start with setting the flow of the fresh reactant feed stream as the means to regulate plant production rate, and then work downstream from there as if looking at a steady-state flowsheet and simply connect the recycle stream back to the reactor based upon a standard control strategy for the column.

However, this strategy is no flow controller anywhere in the recycle loop. The flows around the loop are set based upon level control in the reactor and reflux drum. This control structure is expected to find that exhibiting the snowball effect. By writing the various overall steady-state mass and component balances around the whole process and around the reactor and column, the flow of the recycle stream can be calculated at steady state for any given fresh reactant feed flow and composition.

With the control structure in Figure 3.2 and the base-case fresh feed flow and composition, the recycle flowrate is normally 260.5 moles/h. However, the recycle flow must decrease to 205 moles/h when the fresh feed composition is 0.80 mole fraction A. it must increase to 330 moles/h when the fresh feed composition changes to pure A. Thus a 25 percent change in the disturbance (fresh feed composition) results in a 60 percent change in recycle flow. With this same control structure and the base-case fresh reactant feed composition, the recycle flow drops to 187 moles/h if the fresh feed flow changes to 215 moles/h. It must increase to 362 moles/h when the fresh feed flowrate is changes to 265 moles/h. Thus a 23 percent change in fresh feed flowrate results in a 94 percent change in recycle flowrate. These snowball effects are typical for many recycle systems when control structure such as that shown in Figure 3.2 are used and there is no flow controller somewhere in the recycle loop.

Variable reactant holdup structure An alternative control structure is shown in Figure 3.3. This strategy differs from the previous one in two simple but important ways.

1. Reactor effluent flow is controlled.

2. Reactor holdup is controlled by manipulating the fresh reactant feed flowrate.

All other control loops are the same. The production rate cannot change directly by manipulating the fresh feed flow, because it is used to control reactor level. However, the plant throughput can achieved indirectly in this scheme by changing the setpoint of the reactor level controller. Using the same numerical case considered previously, the recycle flowrate does not change at all when the fresh feed composition changes. To alter production rate from 215 moles/h to 265 moles/h (a 23 percent change), the reactor holdup must be change from 1030 moles/h to 1520 moles/h (a 48 percent change), Recycle flow also changes, but only from 285 to 235 moles/h. This is an 18 percent change in recycle flow compared with 94 percent in the alternative strategy.
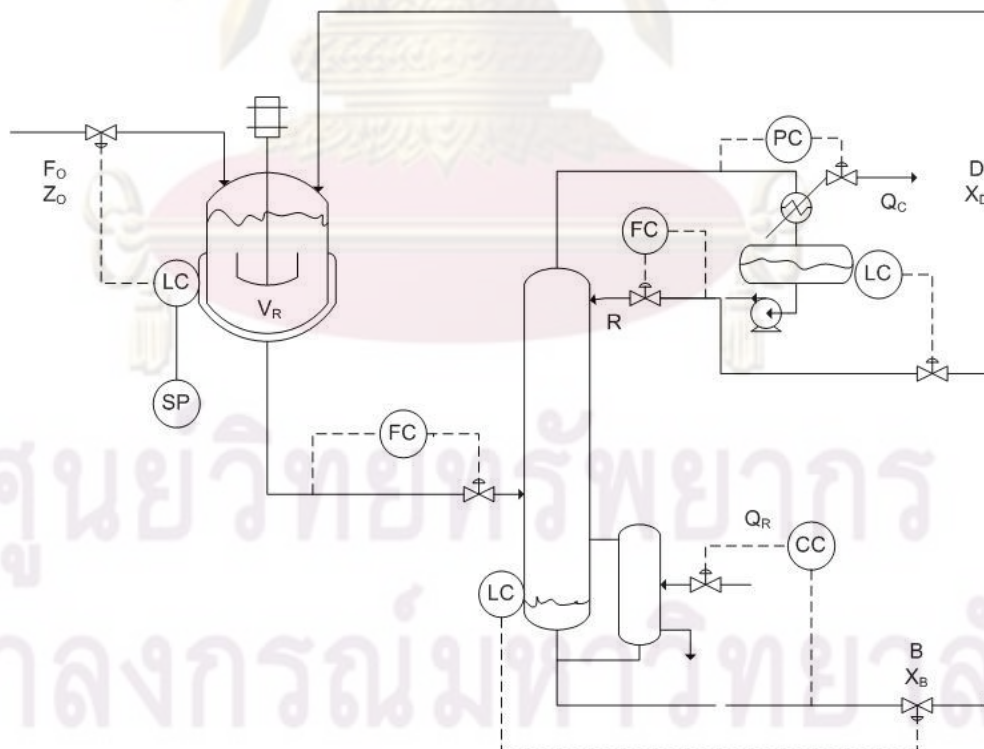
Figure 3.3 Control structure with variable reactor hold

3.3.4 Reaction/Separation Section Interaction

For the process considered in the previous section where the reaction is, the over reaction rate depends upon reactor holdup, temperature (rate constant), and reactant composition (mole fraction A) R= $V_R$kz . The two control structures considered above produce fundamentally different behavior in handling disturbances. In the first, the separation section must absorb almost all of the changes. For example, to increase production rate of component B by 20 percent, the overall reaction rate must increase by 20 percent since both reactor temperature and reactor holdup VR are held constant, reactor composition z must increase 20 percent. This translates into a very significant change in the composition of the feed stream to the separation section. This means the load on the separation section changes significantly, producing large variations in recycle flowrates.

In the second structure, both reactor holdup and reactor composition z can change, so the separation section sees a smaller load disturbance. This reduces the magnitude of the resulting change in recycle flow because the effects of the disturbance can be distributed between the reaction and separation sections.

If the tuning of the reactor level controller in the conventional structure (Figure 3.2) is modified from normal PI to P only, then changes in production rate also produce changes in reactor holdup. This tends to compensate somewhat for the required changes in overall reaction rate and lessens the impact on the separation section. So both control system structure and the algorithm used in the inventory controller of the reactor affect the amount of this snowball phenomenon.

This example has a liquid-phase reactor, where volume can potentially be varied. If the reactor were vapor phase, reactor volume would be fixed. However, an additional degree of freedom are had and could vary reactor pressure to affect reaction rate.

A very useful general conclusion from this simple binary system can be depicted that is applicable to more complex processes: changes in production rate can be achieved only by changing conditions in the reactor. This means something that affects reaction rate in the reactor must vary: holdup in liquid-phase reactor, pressure in gas-phase reactors, temperature, concentrations of reactants (and products in reversible reactions), and catalyst activity or initiator addition rate. Some of these variables affect the conditions in the reactor more than others. Variable with a large effect are called dominant. By controlling the dominant variables in a process, partial control is achieved. The term partial control arises because it typically have fewer available manipulators than variables that would like to control. The setpoints of the partial control loops are then manipulated to hold the important economic objectives in the desired ranges.

The plantwide control implication of this idea is that production rate changes should preferentially be achieved by modifying the setpoint of a partial control loop in the reaction section. This means that the separation section will not be significantly disturbed. Using the control structure in Figure 3.2, changes in production rate require large changes in reactor composition, which disturb the column. Using the control structure shown in Figure 3.3, changes in production rate are achieved by altering the setpoint of a controlled dominant variable, reactor holdup, with only small changes in reactor composition. This means that the column is not disturbed as much as with the alternative control scheme.

3.4 Plantwide process control

Control analysis and control system design for chemical and petroleum processes have traditionally followed the "unit operations approach". First, all of the control loops were established individually for each unit or piece of equipment in the plant. Then the pieces were combined together into an entire plant. This meant that any conflicts among the control loops somehow had to be reconciled. The implicit assumption of this approach was that the sum of the individual parts could effectively comprise the whole of the plant's control system. Over the last few decades, process

control researchers and practitioners have developed effective control schemes for many of the traditional chemical unit operations. And for processes where these unit operations are arranged in series, each downstream unit simply sees disturbances from its upstream neighbor.

Most industrial processes contain a complex flowsheet with several recycle streams, energy integration, and many different unit operation. Essentially, the plantwide control problem is how to development the control loops needed to operate an entire process and achieve its design objective. Recycle streams and energy integration introduce a feedback of material and energy among units upstream and downstream. They also interconnect separate unit operations and create a pate for disturbance propagation. The presence of recycle streams profoundly alters that is not localized to an isolated part of the process.

Despite this process complexity, the unit operations approach to control system design has worked reasonably well. In the past, plants with recycle streams contained many surge tanks to buffer disturbance, to minimize interaction, and to isolate units in the sequence of material flow. This allowed each unit to be controlled individually. Prior to the 1970s, low energy costs meant little economic incentive for energy integration. However, there is growing pressure to reduce capital environmental concerns. This has prompted design engineers to start eliminating many surge tanks, increasing recycle streams, and introducing heat integration for both exiting and new plants. Often this is done without a complete understanding of their effects on plant operability.

So economic force within the chemical industry are compelling improved capital productivity. Requirements for on-aim product quality control grow increasingly tighter. More energy integration occurs. Improved product yields, which reduce raw material costs, are achieved via lower reactant per-pass conversion and higher material recycle rates through the process. Better product quality, energy integration, and higher yields are all economically attractive in the steady-state flowsheet by they present

significant challenges to smooth dynamic plant operation. Hence an effective control system regulating the entire plants operation and a process designed with good dynamic performance play critical parts in achieving the business objectives of reducing operating and capital costs.

Buckley (1964) proposed a control design procedure for the plantwide control problem that consisted of two stages. The first stage determined the material balance control structure to handle vessel inventories for low-frequency disturbances. The second established the product quality control structure to regulate high-frequency disturbances. This procedure has been widely and effectively utilized. It has served as the conceptual framework in many subsequent ideas for developing control systems for complete plants. However, the two-stage Buckley procedure provides little guidance concerning three important aspects of a plantwide control strategy. First, it does not explicitly discuss energy management. Second, it does not address the specific issues of recycle systems. Third, it does not deal with component balance in the context of inventory control. By placing the priority on material balance over product quality control, the procedure can significantly limit the flexibility in choosing the latter.

The goals for an effective plantwide process control system include.

1. Safe and smooth process operation.

2. Tight control of product quality in the face of disturbances.

3. Avoidance of unsafe process conditions.

4. A control system runs in automatic, not manual, requiring minimal operator attention.

5. Rapid rate and product quality transitions.

6. Zero unexpected environmental releases.

3.5 Basic Concepts of Plantwide Control

Buckley Basic

Page Buckley (1964) was the first to suggest the idea of separating the plantwide control problem into two parts:

(1) Material balance control.

(2) Production quality control.

He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loop is established, using the flowrates of liquid and gas process streams. He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated as small as possible. The most level controllers should be proportional-only (P) to achieve flow smoothing.

Douglas doctrines

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

(1) Minimize losses of reactants and products.

(2) Maximize flowrates through gas recycle systems.

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

The control structure implication is that we do not attempt to regulate the gas recycle flow and we do not worry about what we control with its manipulation. We

simply maximize its flow. This removes one control degree of freedom and simplifies the control problem.

Downs drill

Jim Downs (1992) pointed out the importance of looking at the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively. We must ensure that all components (reactants, product, and inerts) have a way to leave or be consumed within the process. Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. Because we usually want to minimize raw material costs and maintain high-purity products, most of the reactants fed into the process must be chewed up in the reactions. And the stoichiometry must be satisfied down to the last molecule. Chemical plants often act as pure integrators in terms of reactants will result in the process gradually filling up with the reactant component that is in excess. There must be a way to adjust the fresh feed flowrates so that exactly the right amounts of the two reactants are fed in.

Luyben laws

Three laws have been developed as a result of a number of case studies of many types of system:

(1) All recycle loops should be flow controlled. This is to prevent the snowball effect.

(2) A fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants.

(3) If the final product from a process comes out the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor (Cantrell et al., 1995). Even if steady-state economics favor a liquid feed stream, the profitability of an operating plant with a

product leaving the bottom of a column may be much better if the feed to column is vaporized.

Richardson rule

Bob Richardson suggested the heuristic that the largest stream should be selected to control the liquid level in a vessel. (The bigger the handle you have to affect a process, the better you can control it).

Shinskey schemes

Greg Shinskey (1988) has produced a number of "advanced control" structures that permit improvements in dynamic performance.

Tyreus tuning

Use of P-only controllers for liquid levels, turning of P controller is usually trivial: set the controller gain equal to 1.67. This will have the valve wide open when the level is at 80 percent and the valve shut when the level is at 20 percent.

For other control loops, suggest the use of PI controllers. The relay-feedback test is a simple and fast way to obtain the ultimate gain (K_u) and ultimate period (P_u). Then either the Ziegler-Nichols setting or the Tyreus-Luyben (1992) settings can be used:

$$K_{ZN} \;=\; K_u/2.2 \quad \tau_{ZN} \;=\; P_u/1.2$$

$$K_{TL} \;=\; K_u/2.2 \quad \tau_{TL} \;=\; P_u/1.2$$

The use of PID controllers, the controlled variable should have a very large signal-to-noise ratio and tight dynamic control is really essential from a feedback control stability perspective.

3.6 Step of Plantwide Process Control Design Procedure

Luyben et al. (1998) presented nine steps of the design procedure center around the fundamental principles of plantwide control: energy management; production rate; product quality; operational, environmental, and safety component balance; and economic or process optimization.

Step 1: Establish control objectives

Assess the steady-state design and dynamic control objects for the process. This is probably the most important aspect of the problem because different control objectives lead to different control structures. The "best" control structure for a plant depends upon the design and control criteria established.

These objectives include reactor and separation yields, product quality specification, product grades and demand determination, environmental restrictions, and the range of safe operating conditions.

Step 2: Determine control degrees of freedom

Count the number of control values available.

This is the number of degrees of freedom for control, i.e., the number of variables that can be controlled to set point. The placement of these control valves can sometimes be made to improve dynamic performance, but often there is no choice in their location.

Most of these valves will be used to achieve basic regulatory control of the process: set production rate, maintain gas and liquid inventories, control product qualities, and avoid safety and environmental constraints. Any valves that remain after these vital tasks have been accomplished can be utilized to enhance steady-state economic objectives or dynamic controllability (e.g. minimizes energy consumption, maximize yield, or reject disturbances).

Step 3: Establish energy management system

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

We use the term energy management to describe two functions.

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

2. If heat integration does occur between process streams, then the second function of energy management is to provide a control system that prevents the propagation of thermal disturbances and ensure the exothermic reactor heat is dissipated and not recycled. Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom for control.

Heat removal in exothermic reactors is crucial because of the potential for thermal runaways. In endothermic reactions, failure to add enough heat simply results in the reaction slowing up. If the exothermic reactor is running adiabatically, the control system must prevent excessive temperature rise through the reactor.

Heat integration of a distillation column with other columns or with reactors is widely used in chemical plants to reduce energy consumption. While these designs look great in terms of steady-state economics, they can lead to complex dynamic behavior and poor performance due to recycling of disturbances. If not already included in the design, trim heater/cooler or heat exchanger bypass line must be added to prevent this. Energy disturbances should be transferred to the plant utility system whenever possible to remove this source of variability from the process units.

Step 4: Set production rate

Establish the variable that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate.

To obtain higher production rate, we must increase overall reaction rates. This can be accomplished by raising temperature, increasing reactant concentrations, increasing reactor holdup, or increasing reactor pressure. The variable we select must be dominant for the reactor.

We often want to select a variable that has the least effect on the separation section but also has a rapid and direct effect on reaction rate in the reactor without hitting an operational constraint.

Step 5: Control product quality and handle safety, operational, and environmental constraints

Select the best value to control each of the product-quality, safety and environmental variables.

We should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and dead times and large steady-state gains.

It should be note that, since product quality considerations have become more important, so it should be establish the product-quality loops first, before the material balance control structure.

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

Fix a flow in every recycle loop and then select the manipulated variable to control inventories

In most process 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 level. We have to determine what valve should be used to control each inventory variable. Inventories include all liquid levels (except for surge volume in certain liquid recycle streams) and gas pressures. An inventory variable should be controlled with the manipulate variable that has the largest effect on it within that unit (Richardson rule).

Gas recycle loops are normally set at maximum circulation rate, as limited by compressor capacity, to achieve maximum yields (Douglas doctrine) Proportional-only control should be used in non-reactive level loops for cascade units in series. Even in reactor level control, proportional control should be considered to help filter flowrate disturbances to the downstream separation system.

Step 7: Check component balances

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

Component balances are particularly important in process with recycle streams because of their integrating effect. We must identify the specific mechanism or control loop to guarantee that there will be no uncontrollable buildup of any chemical component within the process (Downs drill).

In process, we don't want reactant components to leave in the product streams because of the yield loss and the desired product purity specification. Hence we are limited to the use of two methods: consuming the reactants by reaction or adjusting their fresh feed flow. 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.

A tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust the fuel flowrate to a furnace supplying energy to the reactor.

Step 9: Optimize economics or improve dynamic controllability

Establish the best way to use the remaining control degrees of freedom.

After satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and setpoints in some controllers that can be adjusted. These can be used either to optimize steady-state economic process performance (e.g. minimize energy, maximize selectivity) or improve dynamic response.

3.7 New Plantwide Control Structure Design Procedure

The book by Luyben et al. (1998) outlines the control structure design procedure based on heuristic and their process engineering insight. Several case studies are given in the book. Luyben's procedure is widely studied and used the plantwide process control community. However, the structural design procedure is not given explicitly. Their case study designs followed the outline and collected heuristic law but need the designer's process engineering insight to pair CVs and MVs. Skogestad (2004) presented the new design procedure mainly based on the mathematical analysis. First, the dynamic and steady state degree of freedom are identified. Then the set of primary controlled variables are determined. They basically are the active constraints and the variables that must be maintained to achieve minimal economic loss when disturbances occur. Then the control variable determining the production rate is selected based on the optimization resulted in the previous step. The pairings of the

selected sets of MVs and CVs are done hierarchically: regulatory control, supervisory control (loop enhancement for SISO or constraint handling for MPC), and optimization layers. Several analysis tools are utilized in these steps, e.g. pole vector analysis, RGA, CLDG, linear and nonlinear optimization. However, he did not discuss which controlled variables should take precedence? Normally the plant would have a large number of variables; the precedence of the control variables must be established to assure the optimality of the designs and to avoid ambiguity in the design decision. Second, how to handle the disturbances is not discussed.

New design procedure of Wongsri (2009) presented plantwide control structure design procedure based on heuristics and mathematical analysis. In this procedure, the precedence of control variables is established. The major disturbances are directed or managed explicitly to achieve the minimal interaction between loops by using the extended (thermal) disturbance propagation method (Wongsri, 1990) to cover the material disturbances. The purposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward.

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

New design procedure of Wongsri (2009) are:

Step 1 Establishment of control objectives.

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

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

Step 4 Energy management via heat exchanger networks.

Step 5 Selection of control configuration using various tools available.

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

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

Step 8 Validation via rigorous dynamic simulation.

Fixture point theorem analysis:

1. The process is considered at dynamic mode (we run the process until the process responses are at steady state).

2. Controlled variable (CV) can be arranged to follow the most sensibility of the process variable by step change of the MV in open loop control (change only one MV, the other should be fixed then alternate to other until complete).

3. Study the magnitude of integral absolute error (IAE) of all process variables that deviates from steady state. This research considers six process variables including temperature, pressure, flow rate, level, Tray temperature and Tray iC4 composition for de-isobutanizer (DIB) column.

4. Select CV by considering CV that gave the most deviation from steady state (high value score).

CHAPTER IV

ALKYLATION PROCESS

4.1 Introduction

In alkylation process we explore another complex multi-unit process, which features several unit operations, two recycle streams and many control loops. There is a reaction section consisting of three CSTRs in series with one of the reactant feeds split among the reactors. The other reactant is kept in excess by a large recycle stream. The reactors operate at low temperature, so refrigeration is required to remove the exothermic heat of reaction. Reactor cooling is achieved by autorefrigeration (evaporative cooling) in the HYSYS simulation. The two fresh feeds contain inert components, so the separation section must separate these from the reactants and the products. The separation section contains three distillation columns.

4.2 Process Studied

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

The main reaction is the combination of isobutane and butene to form iso-octane.

$$iC_4 + C_4^= \rightarrow iC_8$$
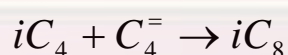
However, there is an undesirable consecutive reaction of butene with iso-octane to form dodecane.

$$iC_8 + C_4^= \rightarrow C_{12}$$
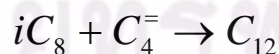
The kinetic data is taken from a case study given by Mahajanam et al. (2001). These exothermic reactions are irreversible and occur in the liquid phase.

The kinetic expressions assumed to be valid for the system are

$$R_1 = 9.6 \times 10^{13} e^{\frac{-28,000}{RT}} \left(C_{iC_4}\right)\left(C_{C_4^=}\right)$$

$$R_2 = 2.4 \times 10^{17} e^{\frac{-35,000}{RT}} \left(C_{C_8}\right)\left(C_{C_4^=}\right)$$

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 $^{\circ}$F).

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

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

4.3 HYSYS Simulation

Figure 4.1 gives the HYSYS simulation flowsheet. There are three CSTRs, each with a volume of 100 ft^3 and each is 80% full of liquid. They operate at 50 $^{\circ}$F and pressures of about 30 psia. The large recycle stream (630 lb-mol/hr) is fed to the first reactor along with 25 lb-mol/hr of BB fresh feed. The composition of the BB fresh feed is 5 mol% propane, 20 mol% isobutane, 60 mol% 1-butene and 15 mol% normal butane. The recycle stream concentrations are 3.5 mol% propane, 94.7 mol% isobutane, 0.2 mol% butene and 1.6 mol% normal butane.
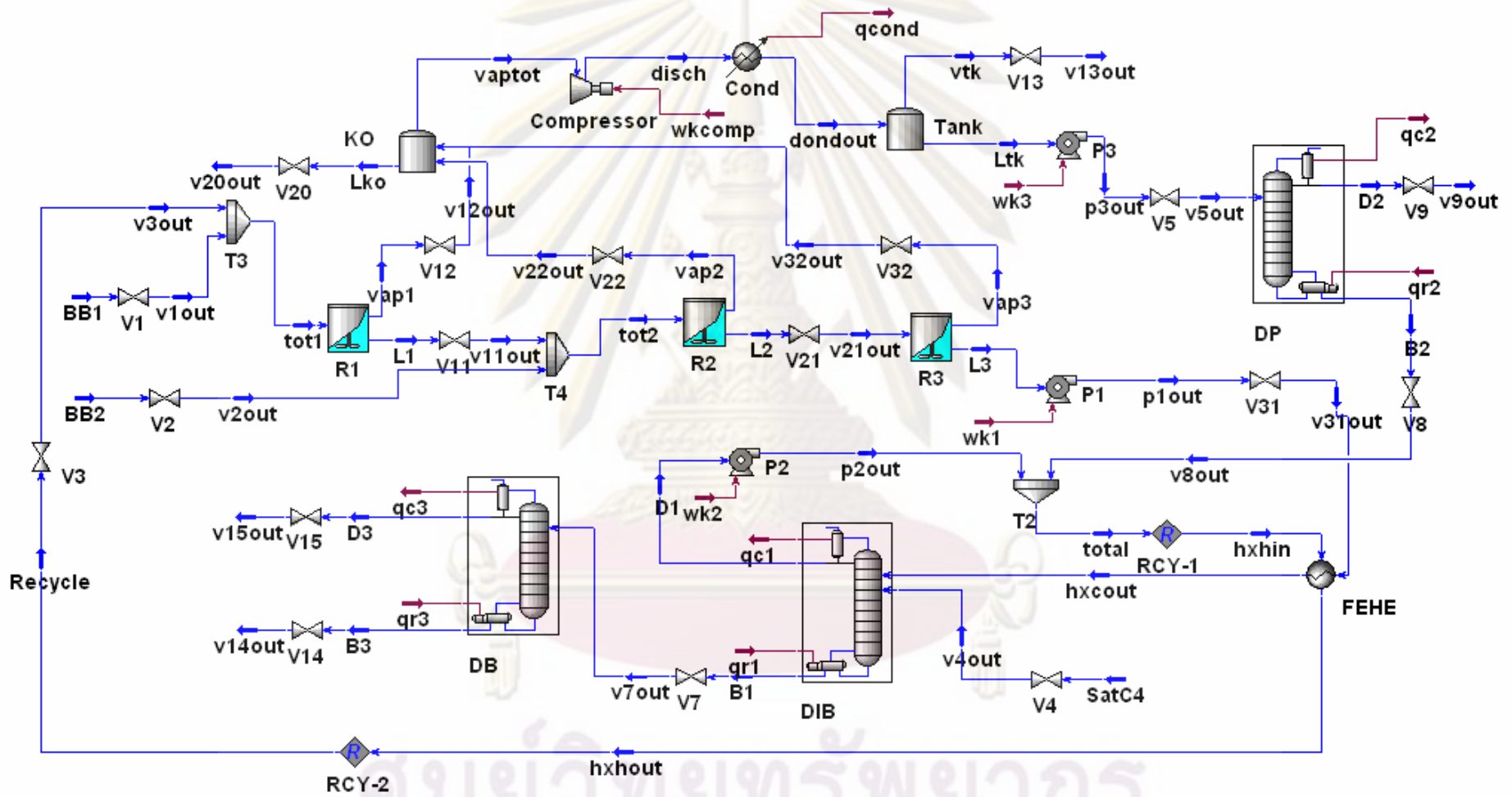
Figure 4.1 HYSYS flowsheet of alkylation process (Luyben, 2002)

The reactors have no external heat transfer. They operate adiabatically, and the required low temperature is attained by autorefrigeration (evaporative cooling). The boiling liquid in the reactor generates a vapor stream, and the latent heat of vaporization removes the exothermic heat of reaction and cools the reactor to the desired temperature. The vapor and liquid phases are in phase equilibrium.

The liquid effluent from the first reactor goes to the second CSTR where a second fresh BB stream (25 lb-mol/hr) is added. The reactor is also cooled by removing vapor stream. The liquid effluent from the second reactor goes to the final reactor, which is cooled in the same way.

The liquid effluent from the reaction section is heated in heat exchanger, which warms the reactor effluent and cools the recycle stream. This reduces both refrigeration requirements in the reaction section and stream consumption in the downstream distillation column. The de-isobutanizer (DIB) column produces a recycle stream of fairly pure isobutane out the top, which is recycled back to the reaction section. The column has 50 trays, operates at 80 psia (reflux drum temperature 108 ^{o}F) and has a reflux ratio of 2. The reactor effluent is fed on Tray 35. There is a second feed stream, the Sat.C$_4$ fresh feed, which is fed on Tray 26. Thus one of the fresh feed is introduced in to the process in the separation section, and the other is introduced into the process in the separation section.

The bottoms from the DIB is fed into a debutanizer (DB) column, which takes the normal butane out the top and the product stream of iso-octane (with some impurity of dodecane) out the bottom. The overall yield of iso-octane is 88%. This column operates at 60 psia.

The final column in the separation section is used to remove propane from the unit. The highest concentration of propane in the system occurs in the vapor streams leaving the reactors (6.9 mol% propane). This vapor is compressed to a high enough pressure (100 psia) so that it can be condensed using cooling water. The liquid is pumped to a higher pressure and fed into a depropanizer distillation column, which

operates at 200 psia. This column has 30 trays and a reflux ratio of 5. The small distillation stream (4.4 lb-mol/hr) with a purity of 95 mol% removes the propane from the system. The bottoms from the column is combined with the distillate from the DIB, and the total is cooled in a feed-effluent heat exchanger (FEHE) before recycling back to the reaction section.

The heat exchanger accomplishes two things: it lessens the cooling load of the refrigeration system in the reaction section, and it lessens the heating load in the DIB reboiler. The heat exchanger has 500 ft^2 of area with an overall heat-transfer coefficient of 100 Btu/hr-ft^2-^{o}F. Its heat duty is 1.06×10^6 Btu/hr.

Thus the flowsheet has two recycle streams, three reactors, three distillation columns and one process/process heat exchanger. Table 4.1 gives stream information.
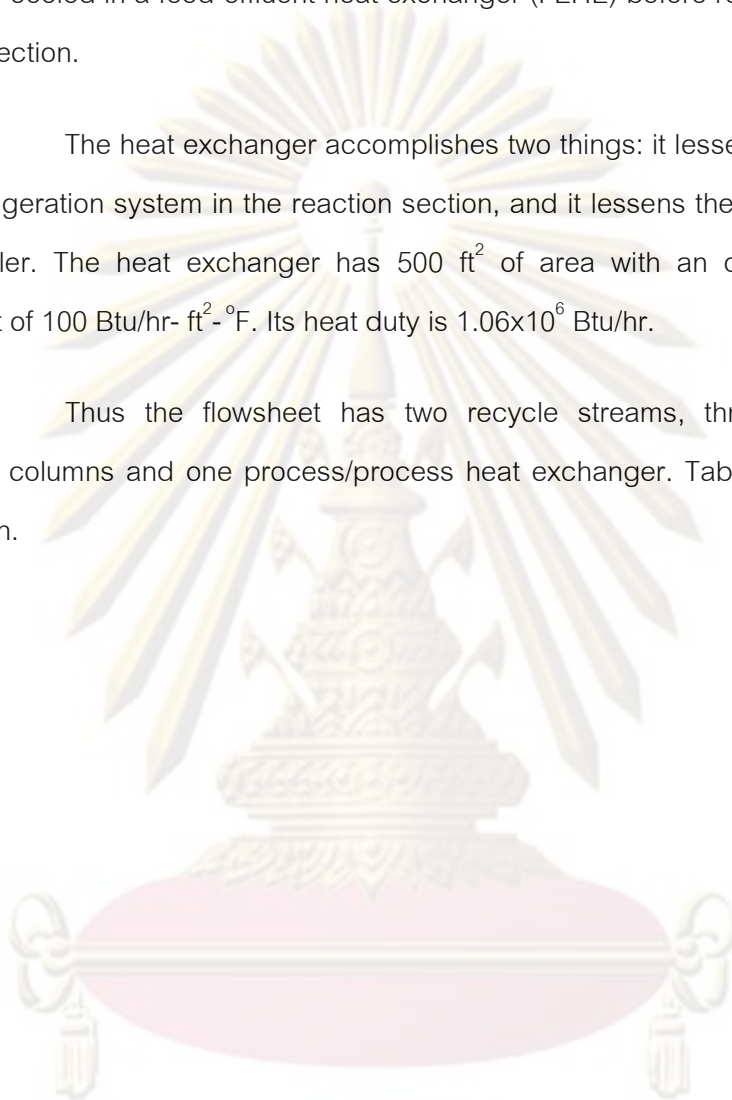
Table 4.1 Data stream of alkylation process

Name	recycle	v3out	BB1	v1out	tot1	vap1	L1	v12out	v11out	BB2
Vapor Fraction	0.0000	0.1411	0.0000	0.1243	0.1404	1.0000	0.0000	1.0000	0.0071	0.0000
Temperature (F)	83.66	47.62	90.00	56.26	47.94	49.38	49.38	45.56	47.55	90.00
Pressure (psia)	95.00	31.00	100.00	31.00	31.00	31.00	31.00	20.00	30.00	100.00
Molar Flow (lbmole/hr)	633.05	633.05	25.00	25.00	658.05	139.62	505.35	139.62	505.35	25.00
Mass Flow (lb/hr)	36484.14	36484.14	1405.32	1405.32	37889.46	7977.56	29911.90	7977.56	29911.90	1405.32
Comp Mole Frac (Propane)	0.0352	0.0352	0.0500	0.0500	0.0357	0.0704	0.0271	0.0704	0.0271	0.0500
Comp Mole Frac (i-Butane)	0.9477	0.9477	0.2000	0.2000	0.9193	0.9099	0.9213	0.9099	0.9213	0.2000
Comp Mole Frac (1-Butene)	0.0019	0.0019	0.6000	0.6000	0.0246	0.0040	0.0051	0.0040	0.0051	0.6000
Comp Mole Frac (n-Butane)	0.0152	0.0152	0.1500	0.1500	0.0203	0.0154	0.0222	0.0154	0.0222	0.1500
Comp Mole Frac (3-Mheptane)	0.0001	0.0001	0.0000	0.0000	0.0001	0.0002	0.0230	0.0002	0.0230	0.0000
Comp Mole Frac (n-C12)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0015	0.0000	0.0015	0.0000

Table 4.1 (Continued) Data stream of alkylation process

Name	v2out	tot2	vap2	L2	v22out	v21out	vap3	L3	v32out	p1out
Vapor Fraction	0.1304	0.0131	1.0000	0.0000	1.0000	0.0075	1.0000	0.0000	1.0000	0.0000
Temperature (F)	54.45	47.85	49.81	49.81	46.34	47.92	48.26	48.26	45.12	48.84
Pressure (psia)	30.00	30.00	30.00	30.00	20.00	29.00	29.00	29.00	20.00	100.00
Molar Flow (lbmole/hr)	25.00	530.35	59.82	455.80	59.82	455.80	11.09	442.56	11.09	442.56
Mass Flow (lb/hr)	1405.32	31317.23	3422.83	27894.40	3422.83	27894.40	635.19	27259.20	635.19	27259.20
Comp Mole Frac (Propane)	0.0500	0.0281	0.0655	0.0242	0.0655	0.0242	0.0639	0.0233	0.0639	0.0233
Comp Mole Frac (i-Butane)	0.2000	0.8873	0.9077	0.8849	0.9077	0.8849	0.9129	0.8843	0.9129	0.8843
Comp Mole Frac (1-Butene)	0.6000	0.0331	0.0046	0.0056	0.0046	0.0056	0.0007	0.0009	0.0007	0.0009
Comp Mole Frac (n-Butane)	0.1500	0.0282	0.0217	0.0299	0.0217	0.0299	0.0220	0.0303	0.0220	0.0303
Comp Mole Frac (3-Mheptane)	0.0000	0.0219	0.0005	0.0498	0.0005	0.0498	0.0005	0.0549	0.0005	0.0549
Comp Mole Frac (n-C12)	0.0000	0.0014	0.0000	0.0056	0.0000	0.0056	0.0000	0.0064	0.0000	0.0064

Table 4.1 (Continued) Data stream of alkylation process

Name	v31out	vaptot	Lko	v20out	disch	dondout	vtk	Ltk	v13out	p3out
Vapor Fraction	0.0000	1.0000	0.0000	1.0000	1.0000	0.0000	1.0000	0.0000	1.0000	0.0000
Temperature (F)	48.89	45.76	45.76	1421.55	151.53	116.10	116.10	116.10	1439.71	117.55
Pressure (psia)	87.00	20.00	20.00	10.00	100.00	95.00	95.00	95.00	45.00	225.00
Molar Flow (lbmole/hr)	442.56	210.53	0.00	0.00	210.53	210.53	0.00	210.53	0.00	210.53
Mass Flow (lb/hr)	27259.20	12035.58	0.00	0.00	12035.58	12035.58	0.00	12035.58	0.00	12035.58
Comp Mole Frac (Propane)	0.0233	0.0687	0.0254	0.0254	0.0687	0.0687	0.1387	0.0687	0.1387	0.0687
Comp Mole Frac (i-Butane)	0.8843	0.9094	0.9059	0.9059	0.9094	0.9094	0.8456	0.9094	0.8456	0.9094
Comp Mole Frac (1-Butene)	0.0009	0.0040	0.0050	0.0050	0.0040	0.0040	0.0032	0.0040	0.0032	0.0040
Comp Mole Frac (n-Butane)	0.0303	0.0175	0.0249	0.0249	0.0175	0.0175	0.0124	0.0175	0.0124	0.0175
Comp Mole Frac (3-Mheptane)	0.0549	0.0003	0.0353	0.0353	0.0003	0.0003	0.0000	0.0003	0.0000	0.0003
Comp Mole Frac (n-C12)	0.0064	0.0000	0.0034	0.0034	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Table 4.1 (Continued) Data stream of alkylation process

Name	v5out	hxcout	hxhin	hxhout	SatC4	v4out	B1	D1	p2out	v7out
Vapor Fraction	0.0000	0.0041	0.0405	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0945
Temperature (F)	117.57	115.24	123.19	83.64	90.00	90.10	211.11	107.66	107.88	197.88
Pressure (psia)	205.00	82.00	100.00	95.00	125.00	85.00	85.00	80.00	100.00	65.00
Molar Flow (lbmole/hr)	210.53	442.56	633.05	633.05	35.00	35.00	47.19	430.37	430.37	47.19
Mass Flow (lb/hr)	12035.58	27259.20	36484.14	36484.14	2009.79	2009.79	4423.84	24845.16	24845.16	4423.84
Comp Mole Frac (Propane)	0.0687	0.0233	0.0352	0.0352	0.0500	0.0500	0.0000	0.0280	0.0280	0.0000
Comp Mole Frac (i-Butane)	0.9094	0.8843	0.9477	0.9477	0.6000	0.6000	0.0050	0.9575	0.9575	0.0050
Comp Mole Frac (1-Butene)	0.0040	0.0009	0.0019	0.0019	0.0000	0.0000	0.0006	0.0008	0.0008	0.0006
Comp Mole Frac (n-Butane)	0.0175	0.0303	0.0152	0.0152	0.3500	0.3500	0.4195	0.0136	0.0136	0.4195
Comp Mole Frac (3-Mheptane)	0.0003	0.0549	0.0001	0.0001	0.0000	0.0000	0.5148	0.0000	0.0000	0.5148
Comp Mole Frac (n-C12)	0.0000	0.0064	0.0000	0.0000	0.0000	0.0000	0.0601	0.0000	0.0000	0.0601

Table 4.1 (Continued) Data stream of alkylation process

Name	D2	B2	v9out	v8out	total	B3	D3	v15out	v14out
Vapor Fraction	0.0000	0.0000	0.1038	0.2802	0.0401	0.0000	0.0000	0.1654	0.2050
Temperature (F)	110.52	177.58	89.11	122.85	123.19	365.25	112.24	68.89	333.26
Pressure (psia)	200.00	202.20	150.00	100.00	100.00	60.00	60.00	30.00	40.00
Molar Flow (lbmole/hr)	4.37	206.15	4.37	206.15	636.52	27.16	20.03	20.03	27.16
Mass Flow (lb/hr)	195.94	11839.64	195.94	11839.64	36684.80	3259.61	1164.23	1164.23	3259.61
Comp Mole Frac (Propane)	0.9500	0.0500	0.9500	0.0500	0.0351	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (i-Butane)	0.0500	0.9277	0.0500	0.9277	0.9479	0.0000	0.0118	0.0118	0.0000
Comp Mole Frac (1-Butene)	0.0000	0.0041	0.0000	0.0041	0.0019	0.0000	0.0014	0.0014	0.0000
Comp Mole Frac (n-Butane)	0.0000	0.0179	0.0000	0.0179	0.0150	0.0010	0.9868	0.9868	0.0010
Comp Mole Frac (3-Mheptane)	0.0000	0.0003	0.0000	0.0003	0.0001	0.8945	0.0000	0.0000	0.8945
Comp Mole Frac (n-C12)	0.0000	0.0000	0.0000	0.0000	0.0000	0.1045	0.0000	0.0000	0.1045

The three columns, the tank after the compressor, the knock-out drum before the compressor and the heat exchanger must be sized for dynamic simulation. The usual sizing methods are used, and result are summarized in Table 4.2

Table 4.2 Equipment data specifications of alkylation process (Luyben, 2002)

Unit		
DIB	Column ID	6 ft
	Reflux Drum	360 ft3
	Base	300 ft3
DP	Column ID	1.2 ft
	Reflux Drum	20 ft3
	Base	100 ft3
DB	Column ID	1 ft
	Reflux Drum	9 ft3
	Base	40 ft3
KO Drum	Volumn	100 ft3
Drum	Volumn	62 ft3
FEHE	Shell c	11 ft3
	Tube Volume	11 ft3
Reactor (3)	Total Volumn (80% liquid full)	100 ft3

CHAPTER V

CONTROL STRUCTURE DESIGN AND DYNAMIC SIMULATION

Maintaining the plant energy and mass balances are the essential task of plantwide for a complex plant consists of recycle streams and energy integration when the disturbance load come through the process. The control system is needed to reject loads and regulate an entire process into a design condition to achieve its objectives therefore our purpose of this chapter is to present the new control structures of alkylation process. Moreover, the eight designed control structures are also compared between base case of alkylation process based on rigorous dynamic simulation by using the commercial software HYSYS version 3.1.

5.1 New Plantwide Control Strategies

The proposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward.

In this research the plantwide control structures in the typical of alkylation process are designed based on the new design procedure given by Wongsri (2009) for all designed control structures and discussed below.

Step 1: Establishment of control objectives

In this research, the objectives we decomposed into two levels: Plantwide level and Unit level.

Plantwide Level: For this process, the control objectives maintain product purity of iso-octane composition at 88 mol%

Unit Level: Stabilitation and smooth operation.

Process constraints during operation: The three reactors temperature should be around 50 $^{\circ}$F and pressure about 30 psia. This is an optimization decision to have better reaction rate. Providing a large recycle stream to maintain the desired yield

at 632.9 lbmol/hr with 94 mol% iso-butane purity to suppress the second undesirable reaction.

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

Plantwide Level: Consider material recycle loop because it causes a system to be born "snowball effect". Alkylation process has a large recycle stream. Therefore to avoid snowball effect is control the molar flow rate of recycle stream in the process.

Unit Level: Use the Fixture Point theorem to select appropriate controlled variables from a candidate output to maintain product quality, to satisfy safety operational, environmental constrains and to setting the production rate. The most disturbed points must be satisfactorily controlled by giving them consideration before other variables. Screening output variables for identification controlled variables by using input variables change (change five percent of manipulated variables).

Table 5.1 shows the IAE summation result of reaction section from Fixture Point method to select the appropriate controlled variables to satisfy safety operational and environmental constrains from a candidate output deviation. We select reactor 1 (R1) level, reactor 2 (R2) level, reactor 3 (R3) level because they are more sensitive than the others (molar flow rate variables), and reactor 1 (R1) temperature, reactor 2 (R2) temperature and reactor 3 (R3) temperature are selected as the controlled variables to setting the production rate because they direct to the point to control. Table 5.2 shows the IAE summation result of two fresh feeds from Fixture Point method to select the appropriate controlled variables to setting the production rate from a candidate output deviation. From this table, fresh feed BB1 molar flow rate and Fresh feed BB2 molar flow rate are selected as the controlled variables because they are more sensitive than the others (temperature and pressure). Table 5.3 shows the IAE summation result of Tank and KO from Fixture Point method to select the appropriate controlled variables of Tank and KO to satisfy safety operational and environmental

constrains from a candidate output deviation. From this table, Tank level, Tank temperature and KO pressure are selected as the controlled variables for Tank and KO section because they direct to the point to control.

Table 5.1 IAE summation result of reaction section

Rank	Variables	SUM IAE
1	Reactor 1 level	0.1653
2	Reactor 2 level	0.1459
3	Reactor 3 level	0.0870
4	Reactor 2 outlet flow rate	0.0364
5	Reactor 3 inlet flow rate	0.0364
6	Reactor 1 outlet flow rate	0.0100
7	Reactor 2 inlet flow rate	0.0099
8	Reactor 3 outlet flow rate	0.0053
9	Reactor 3 temperature	0.0053
10	Reactor 3 outlet temperature	0.0053
11	Reactor 3 inlet temperature	0.0052
12	Reactor 2 temperature	0.0052
13	Reactor 2 outlet temperature	0.0052
14	Reactor 2 inlet temperature	0.0050
15	Reactor 1 temperature	0.0048
16	Reactor 1 outlet temperature	0.0048
17	Reactor 1 inlet temperature	0.0047
18	Reactor 1 inlet flow rate	0.0042
19	Reactor 3 pressure	0.0024
20	Reactor 2 pressure	0.0021
21	Reactor 1 pressure	0.0021

Table 5.2 IAE summation result of two fresh feeds

Rank	Variables	SUM IAE
1	Fresh feed BB1 molar flow rate	0.0001
2	Fresh feed BB2 molar flow rate	0.0001
3	Fresh feed BB1 temperature	0.0000
4	Fresh feed BB2 temperature	0.0000
5	Fresh feed BB1 pressure	0.0000
6	Fresh feed BB2 pressure	0.0000

Table 5.3 IAE summation result of KO and Tank

Rank	Variables	SUM IAE
1	Tank level	0.2603
2	Tank inlet temperature	0.0249
3	Tank temperature	0.0248
4	Tank outlet temperature	0.0248
5	Tank outlet molar flow	0.0066
6	Tank inlet molar flow	0.0062
7	KO pressure	0.0046
8	KO inlet pressure	0.0046
9	KO inlet pressure	0.0046
10	KO inlet pressure	0.0046
11	KO outlet pressure	0.0046

The control scheme for the depropanizer (DP) column and debutanizer (DB) column are standard stand-alone schemes. The controlled variables are Tray temperature, pressure condenser, condenser level and reboiler level.

The control scheme for de-isobutanizer (DIB) column is not standard. It is different from traditional schemes. A composition controller is used instead of a temperature controller in the DIB column because there is very little difference in the

boiling points between the two key components in this column (iso-butane and n-butane). So we control the iso-butane composition of DIB column.

Table 5.4 and Figure 5.1 show the IAE summation result of Tray iso-butane composition for de-isobutanizer (DIB) column from Fixture Point method to select the appropriate controlled variables of Tray iso-butane composition for DIB column from a candidate output deviation. The iso-butane composition on Tray 19 for DIB column is the appropriate controlled variables for all designed control structures (CS1 to CS8) because it is the most sensitive.

Figure 5.2 shows the composition gradient for DIB column (this value is the slope value of Tray iso-butane composition for de-isobutanizer (DIB) column from the steady state value). The appropriate controlled variables of Tray iso-butane composition for DIB column from Fixture Point method is similar when compare with slope value of Tray iso-butane composition for DIB column from the steady state value. Figure 5.1 and Figure 5.2 show the similar appropriate controlled variables of Tray iso-butane composition for DIB column.

Table 5.4 IAE summation result of Tray iso-butane composition deviation for DIB column

Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE
1	0.0011	11	0.0083	21	0.0189	31	0.0043	41	0.0011
2	0.0013	12	0.0099	22	0.0178	32	0.0034	42	0.0010
3	0.0016	13	0.0116	23	0.0163	33	0.0027	43	0.0009
4	0.0019	14	0.0134	24	0.0146	34	0.0021	44	0.0009
5	0.0024	15	0.0151	25	0.0128	35	0.0016	45	0.0008
6	0.0030	16	0.0168	26	0.0109	36	0.0015	46	0.0007
7	0.0037	17	0.0182	27	0.0094	37	0.0014	47	0.0006
8	0.0046	18	0.0191	28	0.0079	38	0.0013	48	0.0005
9	0.0056	19	0.0196	29	0.0065	39	0.0012	49	0.0004
10	0.0069	20	0.0195	30	0.0053	40	0.0011	50	0.0003
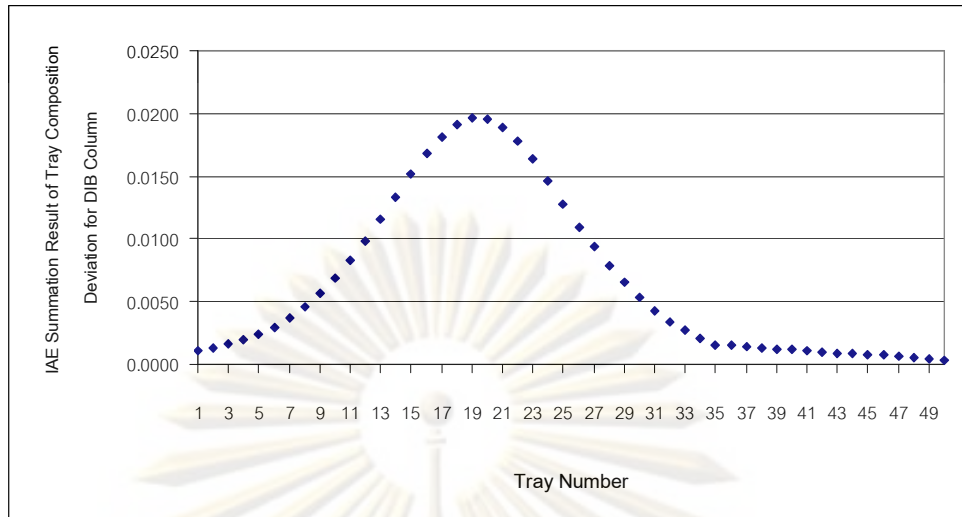
Figure 5.1 IAE summation result of Tray iso-butane composition deviation for DIB column
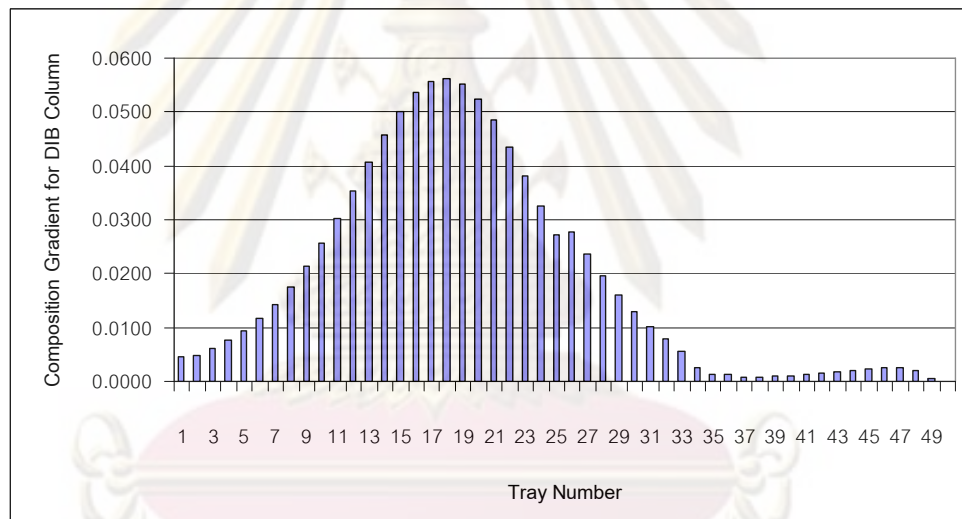
Figure 5.2 Composition Gradient for DIB column

Table 5.5 and Figure 5.3 show the IAE summation result of Tray temperature for depropanizer (DP) column from Fixture Point method to select the appropriate controlled variables of Tray temperature for DP column from a candidate output deviation. The temperature on Tray 25 for DP column is the appropriate controlled variables for all designed control structures (CS1 to CS8) because it is the most sensitive.

Figure 5.4 shows the temperature gradient for DP column (this value is the slope value of Tray temperature for depropanizer (DP) column from the steady state value). The appropriate controlled variables of Tray temperature for DP column from Fixture Point method is similar when compare with slope value of Tray temperature for DP column from the steady state value. Figure 5.3 and Figure 5.4 show the similar appropriate controlled variables of Tray temperature for DP column.

Table 5.5 IAE summation result of Tray temperature deviation for DP column

Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE
1	0.0169	7	0.0174	13	0.0177	19	0.0184	25	0.0262
2	0.0170	8	0.0174	14	0.0177	20	0.0191	26	0.0260
3	0.0172	9	0.0174	15	0.0177	21	0.0200	27	0.0245
4	0.0173	10	0.0174	16	0.0178	22	0.0215	28	0.0222
5	0.0173	11	0.0177	17	0.0179	23	0.0234	29	0.0200
6	0.0174	12	0.0177	18	0.0181	24	0.0251	30	0.0182
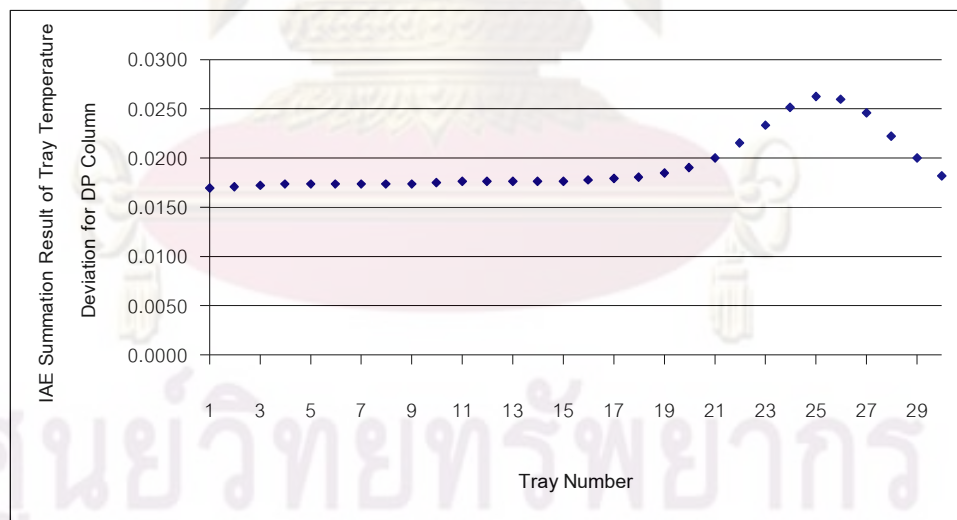
Figure 5.3 IAE summation result of Tray temperature deviation for DP column
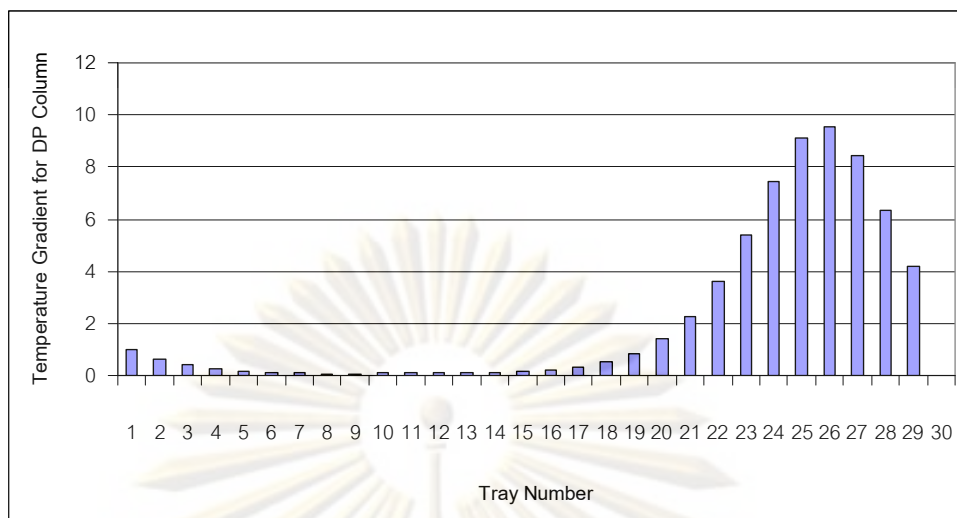
Figure 5.4 Temperature Gradient for DP column

Table 5.6 and Figure 5.5 show the IAE summation result of Tray temperature for debutanizer (DB) column from Fixture Point method to select the appropriate controlled variables of Tray temperature for DB column from a candidate output deviation. The temperature on Tray 3 for DB column is the appropriate controlled variables for all designed control structures (CS1 to CS8) because it is the most sensitive.

Figure 5.6 shows the temperature gradient for DB column (this value is the slope value of Tray temperature for debutanizer (DB) column from the steady state value). The appropriate controlled variables of Tray temperature for DB column from Fixture Point method is similar when compare with slope value of Tray temperature for DB column from the steady state value. Figure 5.5 and Figure 5.6 show the similar appropriate controlled variables of Tray temperature for DB column.

Also the appropriate controlled variable of Tray temperature for debutanizer (DB) column is 3rd Tray temperature because it is the most sensitive but the temperature on Tray 10 for DB column is sensitive too because it has high IAE summation result of Tray temperature deviation from Fixture Point method and high slope value of Tray temperature for DB column from the steady state value. So some designed control structure (CS3, CS5, CS7 and CS8), we use the two point control to control the Tray temperature of DB column on Tray 3 and Tray 10.

Table 5.6 IAE summation result of Tray temperature deviation for DB column

Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE	Tray	SUM IAE
1	0.0968	4	0.3191	7	0.1552	10	0.1211	13	0.1168
2	0.2568	5	0.2335	8	0.1189	11	0.1274	14	0.1047
3	0.3749	6	0.1873	9	0.0803	12	0.1246	15	0.0814
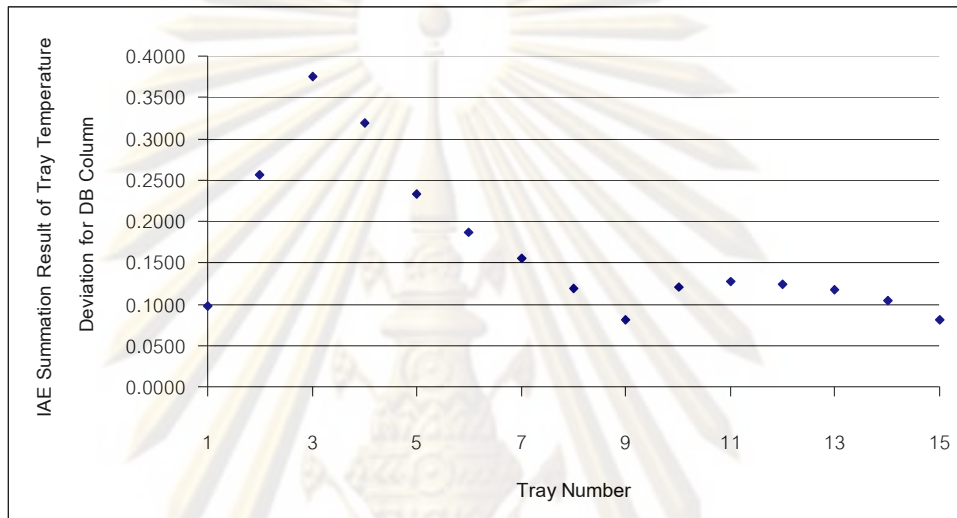
Figure 5.5 IAE summation result of Tray temperature deviation for DB column
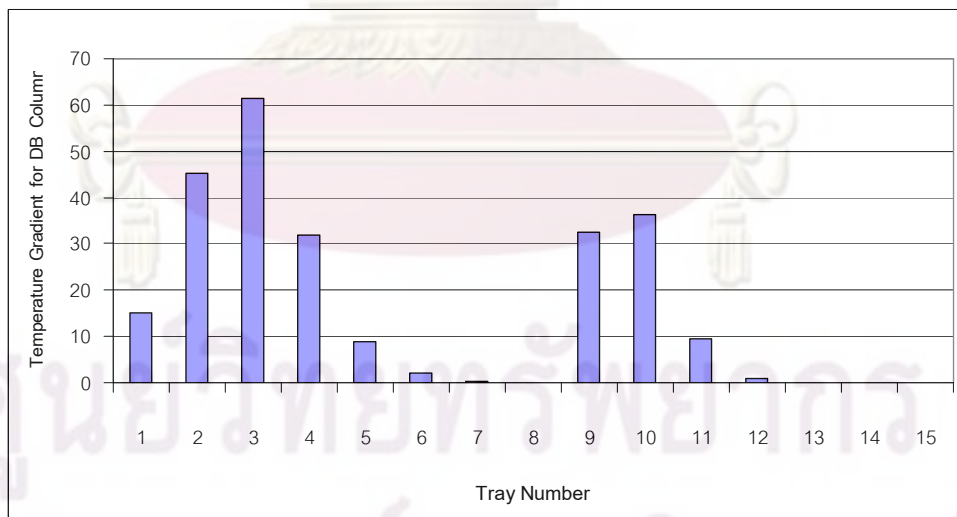
Figure 5.6 Temperature Gradient for DB column

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

The base case control structure (CS0) for alkylation process has 24 control degrees of freedom; 7 utility streams, 1 compressor power and 16 control valves.

For designed control structure I (CS1), designed control structure II (CS2), there are 24 control degrees of freedom same as the base case.

For designed control structure III (CS3) and designed control structure VIII (CS8), there are 25 control degrees of freedom; 7 utility streams, 1 compressor power, 1 reflux and 16 control valves.

For designed control structure IV (CS4), there are 24 control degrees of freedom; 7 utility streams, 1 compressor power, 1 reflux and 15 control valves.

For designed control structure V (CS5) and designed control structure VII (CS7), there are 25 control degrees of freedom; 7 utility streams, 1 compressor power, 2 reflux and 15 control valves.

For designed control structure VI (CS6), there are 24 control degrees of freedom; 7 utility streams, 1 compressor power, 1 reflux and 15 control valves.

The control degrees of freedom for each control structures are shown in Table 5.7.

Table 5.7 The control degrees of freedom for each control structures

Position	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
BB1 fresh feed valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
BB2 fresh feed valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
recycle stream valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 1 (R1) vapor stream valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 1 (R1) liquid effluent valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 2 (R2) vapor stream valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 2 (R2) liquid effluent valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 3 (R3) vapor stream valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reactor 3 (R3) liquid effluent valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
Compressor power	✓	✓	✓	✓	✓	✓	✓	✓	✓
Cooling duty	✓	✓	✓	✓	✓	✓	✓	✓	✓
Tank effluent liquid valve	✓	✓	✓	✓	✓	✓	✓	✓	✓
SatC4 fresh feed valve	✓	✓	✓	✓	✓	✓	✓	✓	✓

Table 5.7 (Continued) The control degrees of freedom for each control structures

Position	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
Condenser duty of DIB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reboiler duty of DIB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Bottom valve of DIB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Condenser duty of DP column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reboiler duty of DP column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Distillate valve of DP column	✓	✓	✓	✓			✓		✓
Bottom valve of DP column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reflux of DP column					✓	✓	✓	✓	
Condenser duty of DB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reboiler duty of DB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Distillate valve of DB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Bottom valve of DB column	✓	✓	✓	✓	✓	✓	✓	✓	✓
Reflux of DB column				✓	✓	✓		✓	✓
Degree of Freedom	24	24	24	25	24	25	25	25	25

Step 4: Energy management via heat exchanger networks

In this research, we no design the new heat exchanger network, since alkylation process is low temperature operation.

The heat exchanger in alkylation process accomplishes two things: it lessens the cooling load of the refrigeration system in the reaction section, and it lessens the heating load in DIB reboiler.

Step 5: Selection of control configuration using various tools available

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

For reactor section: There are three reactors (CSTR) exothermic reactions. The reactors have no external heat transfer. They operate adiabatically, and the required low temperature. The boiling liquid in the reactor generates a vapor stream, and the latent heat of vaporization removes the exothermic heat of reaction and cools the reactor to the desired temperature. The reactor is cooled by removing a vapor stream so we select the vapor stream valve to manipulate the reactor temperature. And we select the liquid effluent stream valve to manipulate the reactor level because the liquid effluent from the reactor is out of this stream.

For KO unit: KO has to prevent the flow of fluid liquid into the compressor. KO pressure is controlled by manipulating compressor power because if the KO pressure down can adjust at the compressor power immediately.

For Tank unit: The cooling water for the condenser would be minimized compressor discharge pressure. We control Tank temperature because if the temperature increase will become a liquid substance that makes pump (P3) break and

manipulate at cooling duty in the condenser because if the Tank temperature increase can adjust at the cooling duty in the condenser immediately.

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

In some designed control structure (CS2, CS3, CS6 and CS7), for de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column reboiler duty (qr1) because it is direct to manipulate the iso-butane composition.

In some designed control structure (CS4, CS5, CS6 and CS7), for debutanizer (DP) column, The DP column condenser level is controlled by manipulating the DP column reflux because there are higher molar flow rate than distillate stream.

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

Alkylation process has six components to be accounted for: There are propane (C_3), isobutane (iC_4), butene (C_4), n-butane (nC_4), iso-octane (iC_8) and dodecane (C_{12}). The compositions of two fresh feeds are butene and isobutene with two impurities (propane and n-butane). The product (iC_8) and by-product (C_{12}) leave in product stream of DB column. N-butane leaves in distillate stream of DB column. Propane leaves in distillate stream of DP column. Unreact isobutane from product stream of DP column is combined with the distillate stream of de-isobutanizer (DIB) column and recycling back to the reaction section.

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

In this research, all controller type is single-input-single-output loop. There are temperature controller, composition controller, pressure controller, flow controller and level controller. Temperature controllers and composition controller are PIDs which are tuned using relay feedback. Pressure controllers and flow controllers are

PIs and their parameters are heuristics values. Proportional-only level controllers are used and their parameters are heuristics values. All control valves are half-open at nominal operating condition.

Step 8: Validation via rigorous dynamic simulation

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

5.2 Design of Plantwide Control Structure

In this research, we apply the new design procedure by Wongsri (2009) to all design control structures for alkylation process. The objectives were decomposed into two levels: Plantwide level and Unit level. In all of these control structures (CS0, CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8) the same loops are used as follows:

Plantwide level

- Valve V3 is manipulated to control the recycle flow rate

Unit level

Reactor 1 (R1) unit

- Valve V12 is manipulated to control the reactor temperature.

- Valve V11 is manipulated to control the reactor level.

- Valve V1 is manipulated to control fresh feed (BB1) flow rate.

Reactor 2 (R2) unit

- Valve V22 is manipulated to control the reactor temperature.

- Valve V21 is manipulated to control the reactor level.

- Valve V2 is manipulated to control fresh feed (BB1) flow rate.

Reactor 3 (R3) unit

- Valve V32 is manipulated to control the reactor temperature.

- Valve V31 is manipulated to control the reactor level.

KO vessel unit

- Compressor power is manipulated to control the KO vessel pressure.

Tank vessel unit

- Condenser duty is manipulated to control the Tank vessel temperature.

- Valve V5 is manipulated to control the Tank vessel level.

De-isobutanizer (DIB) column unit

- Heat duty of condenser (qc1) is manipulated to control the de-isobutanizer (DIB) column pressure.

- Valve V4 is manipulated to control the de-isobutanizer (DIB) column level condenser.

Depropanizer (DP) column unit

- Heat duty of condenser (qc1) is manipulated to control the depropanizer (DP) column pressure.

- Heat duty of reboiler (qr2) is manipulated to control the depropanizer (DP) column Tray temperature.

- Valve V8 is manipulated to control the depropanizer (DP) column level reboiler.

Debutanizer (DB) column unit

- Heat duty of condenser (qc3) is manipulated to control the debutanizer (DB) column pressure.

- Heat duty of reboiler (qr3) is manipulated to control the debutanizer (DB) column Tray temperature.

- Valve V15 is manipulated to control the debutanizer (DB) column level condenser.

- Valve V14 is manipulated to control the debutanizer (DB) column level reboiler.

In all of these control structures (CS0, CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8) the difference loops are used as follows

5.2.1 Base case control structure (CS0)

For de-isobutanizer (DIB) column, the 10th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column bottom product valve (V7). The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1). As show in Figure 5.7

5.2.2 Design of control structure I (CS1)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column bottom product valve (V7). The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1). As show in Figure 5.8

5.2.3 Design of control structure II (CS2)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column reboiler

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duty (qr1). The DIB column condenser level is controlled by manipulating the DIB column bottom product valve (V7). As show in Figure 5.9

5.2.4 Design of control structure III (CS3)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column reboiler duty (qr1). The DIB column condenser level is controlled by manipulating the DIB column bottom product valve (V7). For depropanizer (DP) column, 10th Tray temperature of depropanizer (DP) column is controlled by manipulating the DP column reflux. As show in Figure 5.10

5.2.5 Design of control structure IV (CS4)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column bottom product valve (V7). The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1). For debutanizer (DP) column, The DP column condenser level is controlled by manipulating the DP column reflux. As show in Figure 5.11

5.2.6 Design of control structure V (CS5)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column bottom product valve (V7). The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1). For debutanizer (DP) column, The DP column condenser level is controlled by manipulating the DP column reflux. For depropanizer (DP) column, 10th Tray temperature of depropanizer (DP) column is controlled by manipulating the DP column reflux. As show in Figure 5.12

5.2.7 Design of control structure VI (CS6)

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

column bottom product valve (V7). For debutanizer (DP) column, The DP column condenser level is controlled by manipulating the DP column reflux. As show in Figure 5.13

5.2.8 Design of control structure VII (CS7)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column reboiler duty (qr1). The DIB column condenser level is controlled by manipulating the DIB column bottom product valve (V7). For debutanizer (DP) column, The DP column condenser level is controlled by manipulating the DP column reflux. For depropanizer (DP) column, 10th Tray temperature of depropanizer (DP) column is controlled by manipulating the DP column reflux. As show in Figure 5.14

5.2.9 Design of control structure VIII (CS8)

For de-isobutanizer (DIB) column, the 19th Tray iso-butane composition of de-isobutanizer (DIB) column is controlled by manipulating the DIB column bottom product valve (V7). The DIB column condenser level is controlled by manipulating the DIB column reboiler duty (qr1). For depropanizer (DP) column, 10th Tray temperature of depropanizer (DP) column is controlled by manipulating the DP column reflux. As show in Figure 5.15
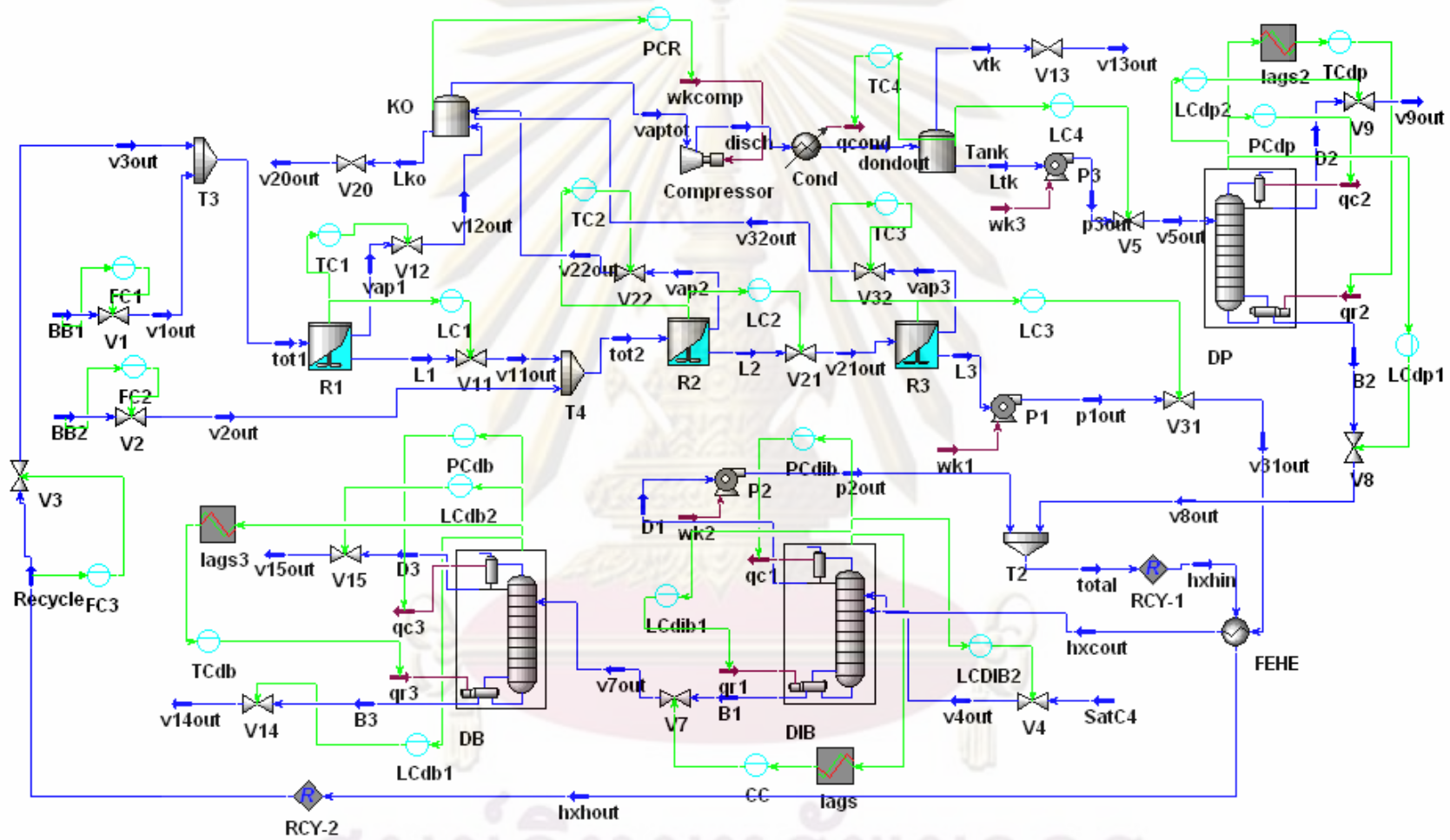
Figure 5.7 Base case control structure (CS0) for alkylation Process, Luyben (2002)
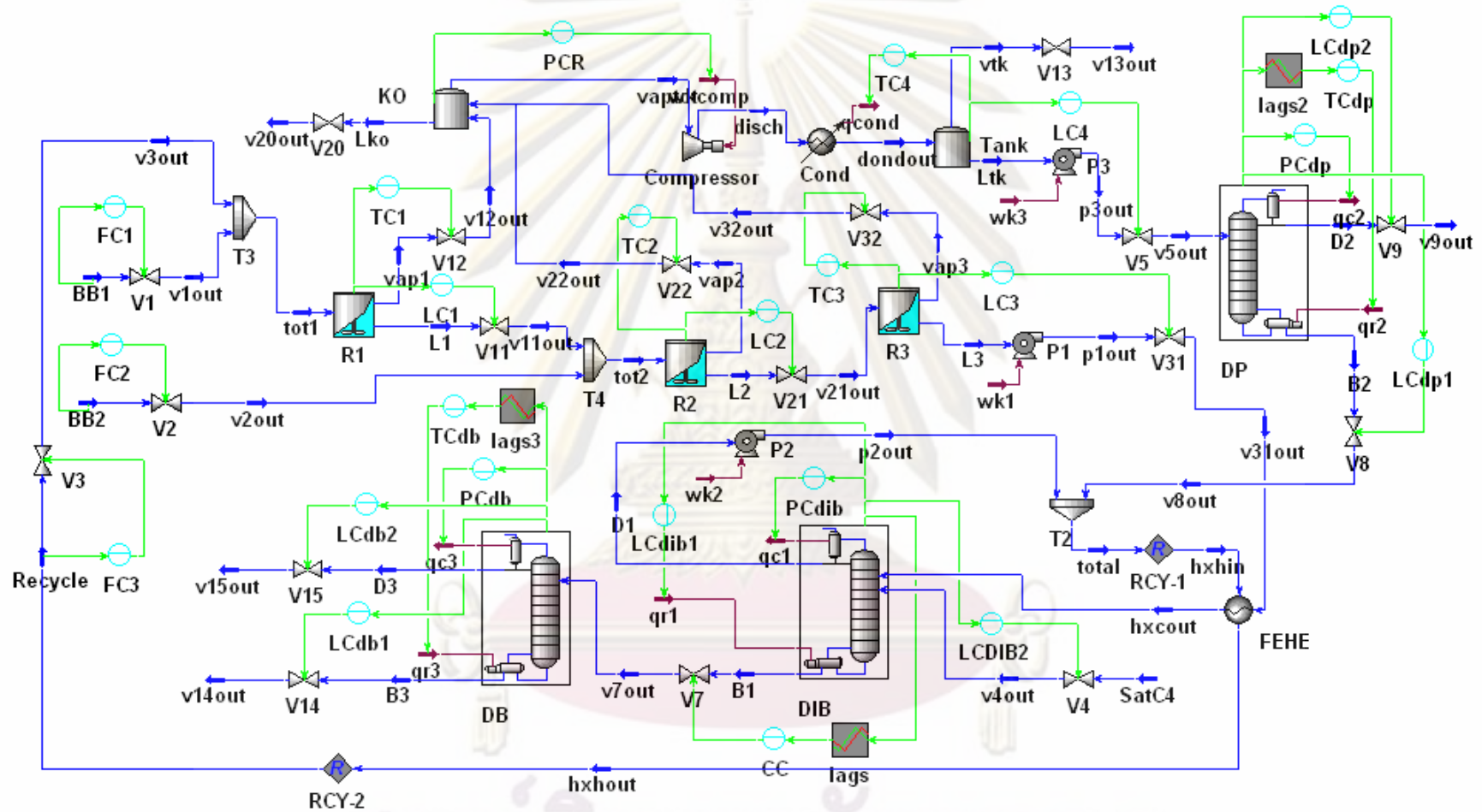
Figure 5.8 Designed control structure I (CS1) for alkylation Process
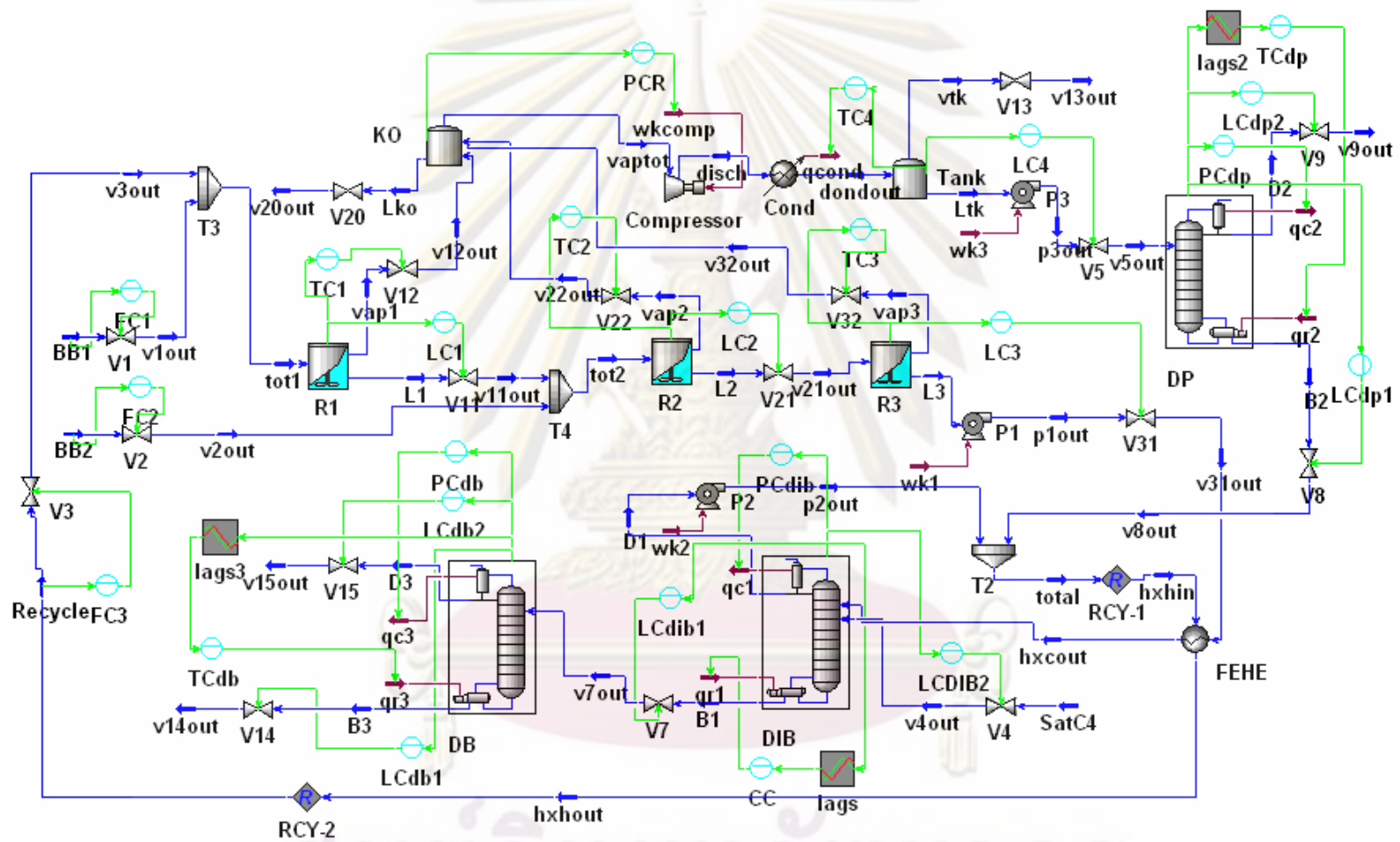
Figure 5.9 Designed control structure II (CS2) for alkylation Process
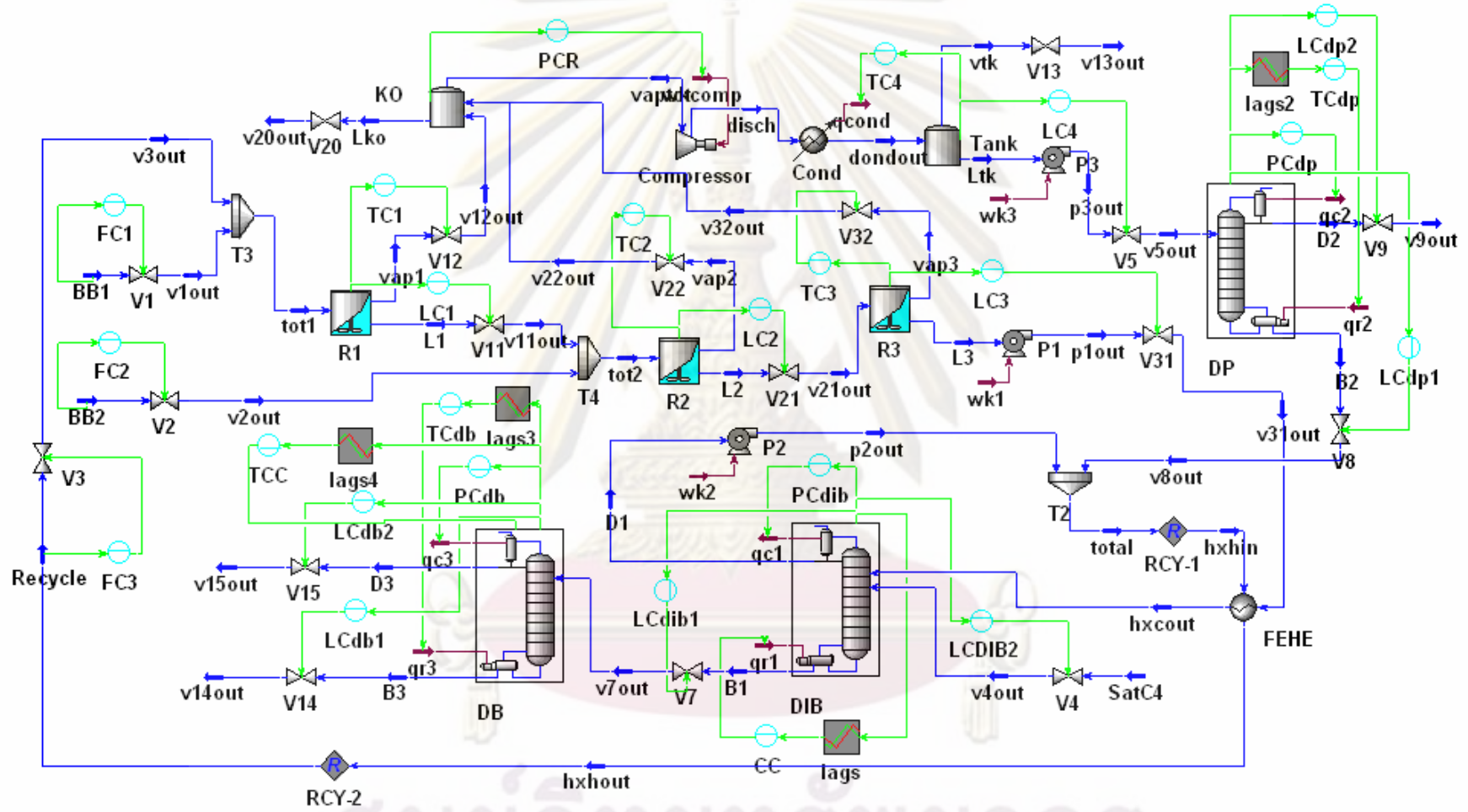
Figure 5.10 Designed control structure III (CS3) for alkylation Process
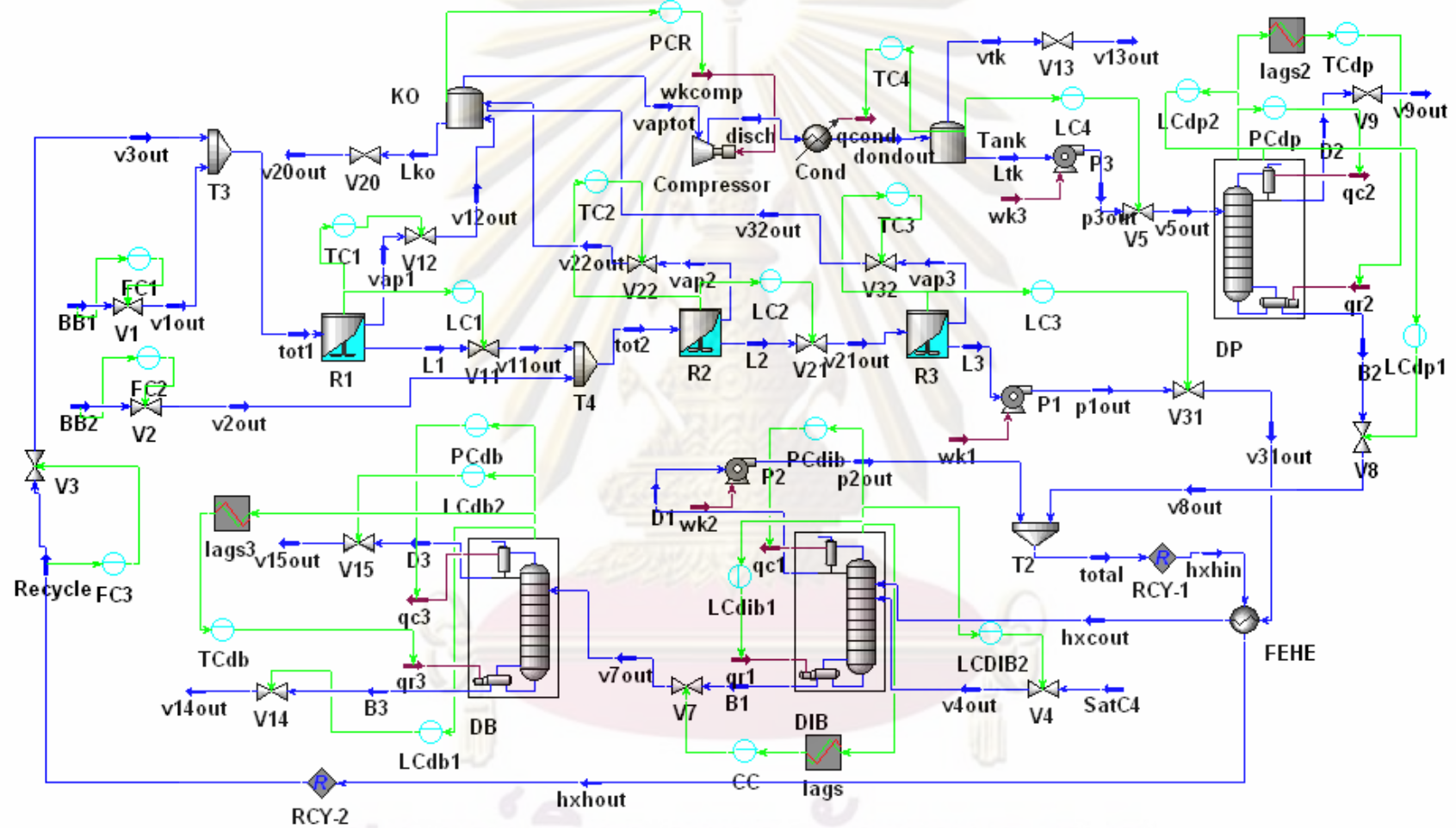
Figure 5.11 Designed control structure IV (CS4) for alkylation Process
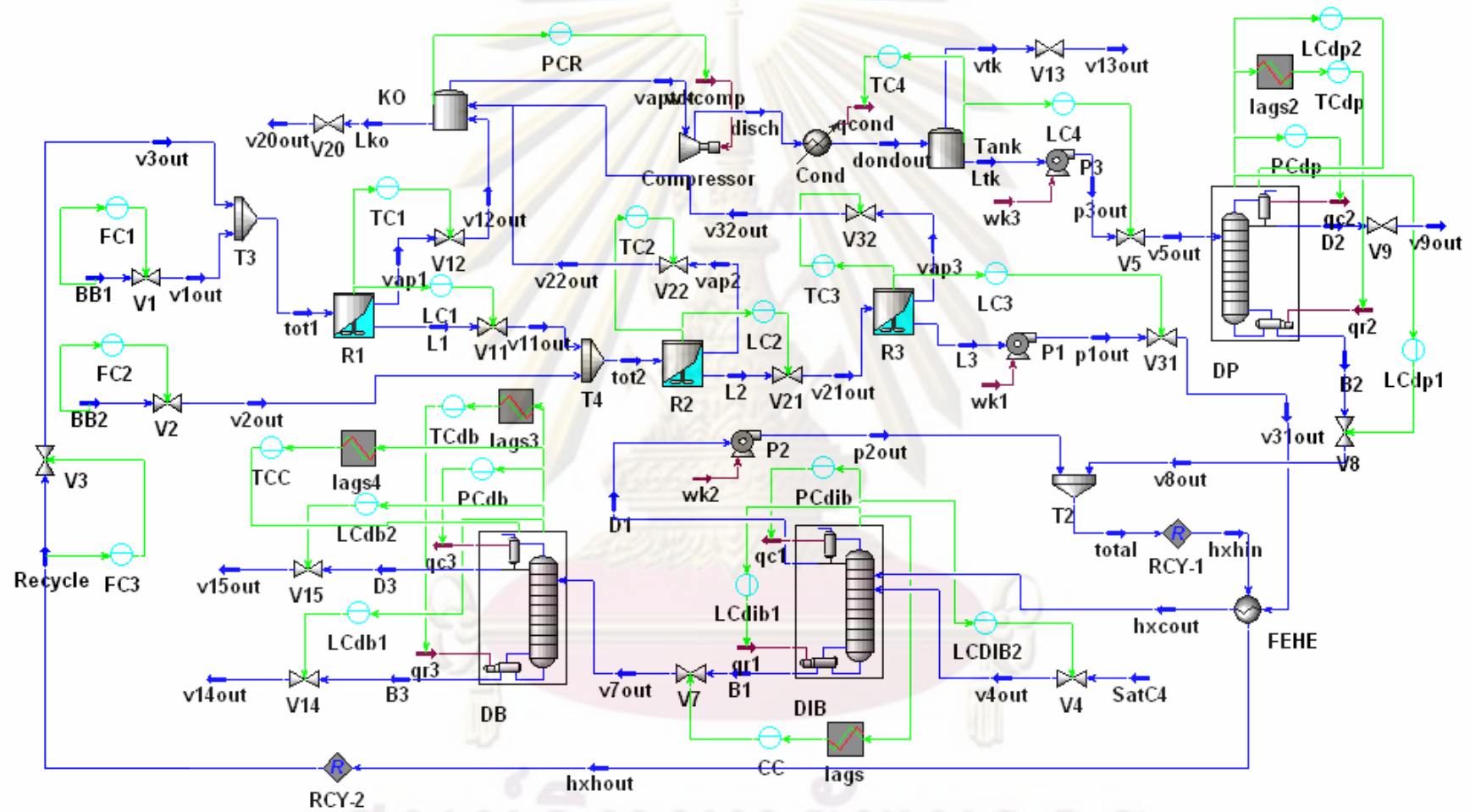
Figure 5.12 Designed control structure V (CS5) for alkylation Process
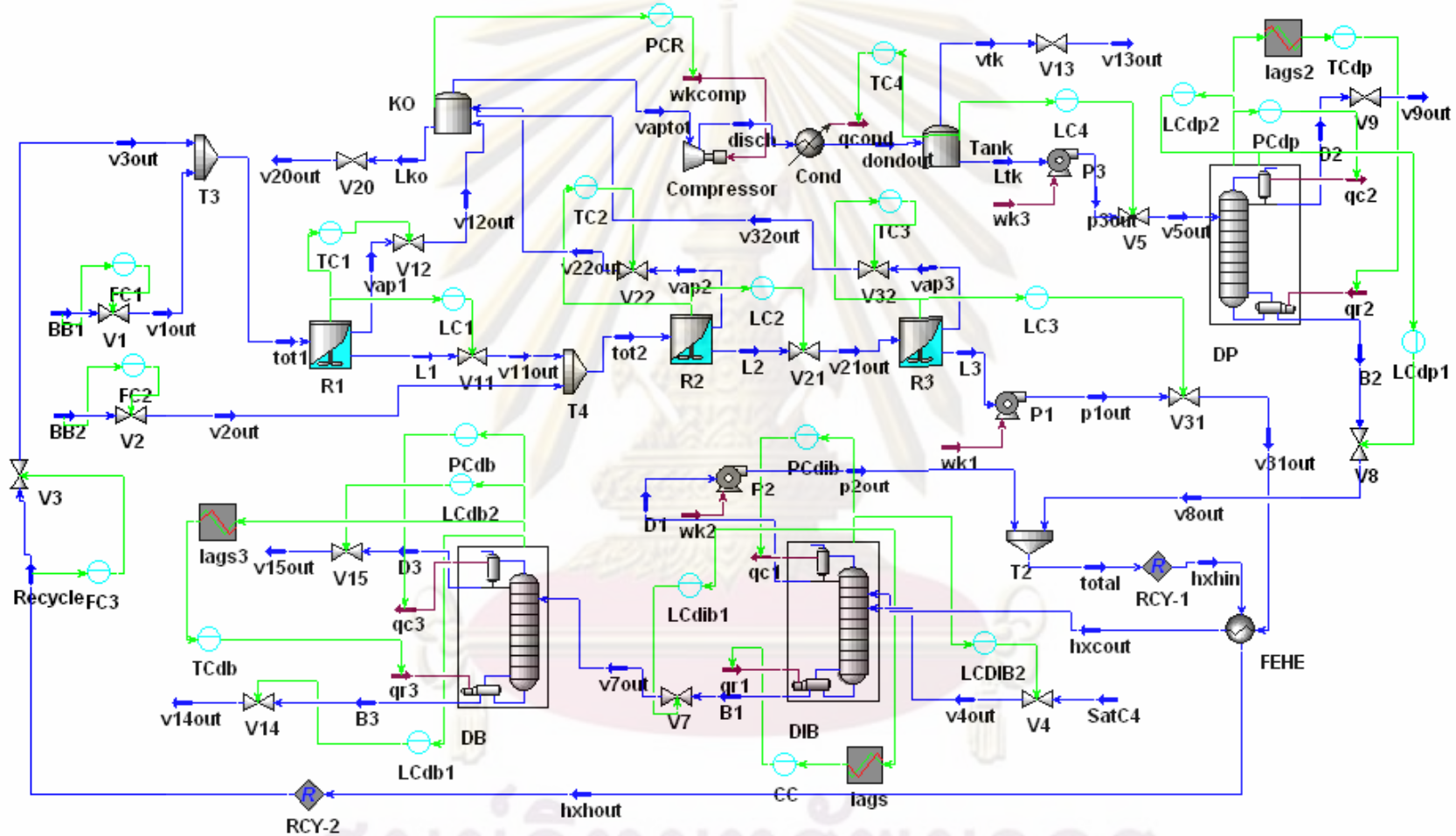
Figure 5.13 Designed control structure VI (CS6) for alkylation Process
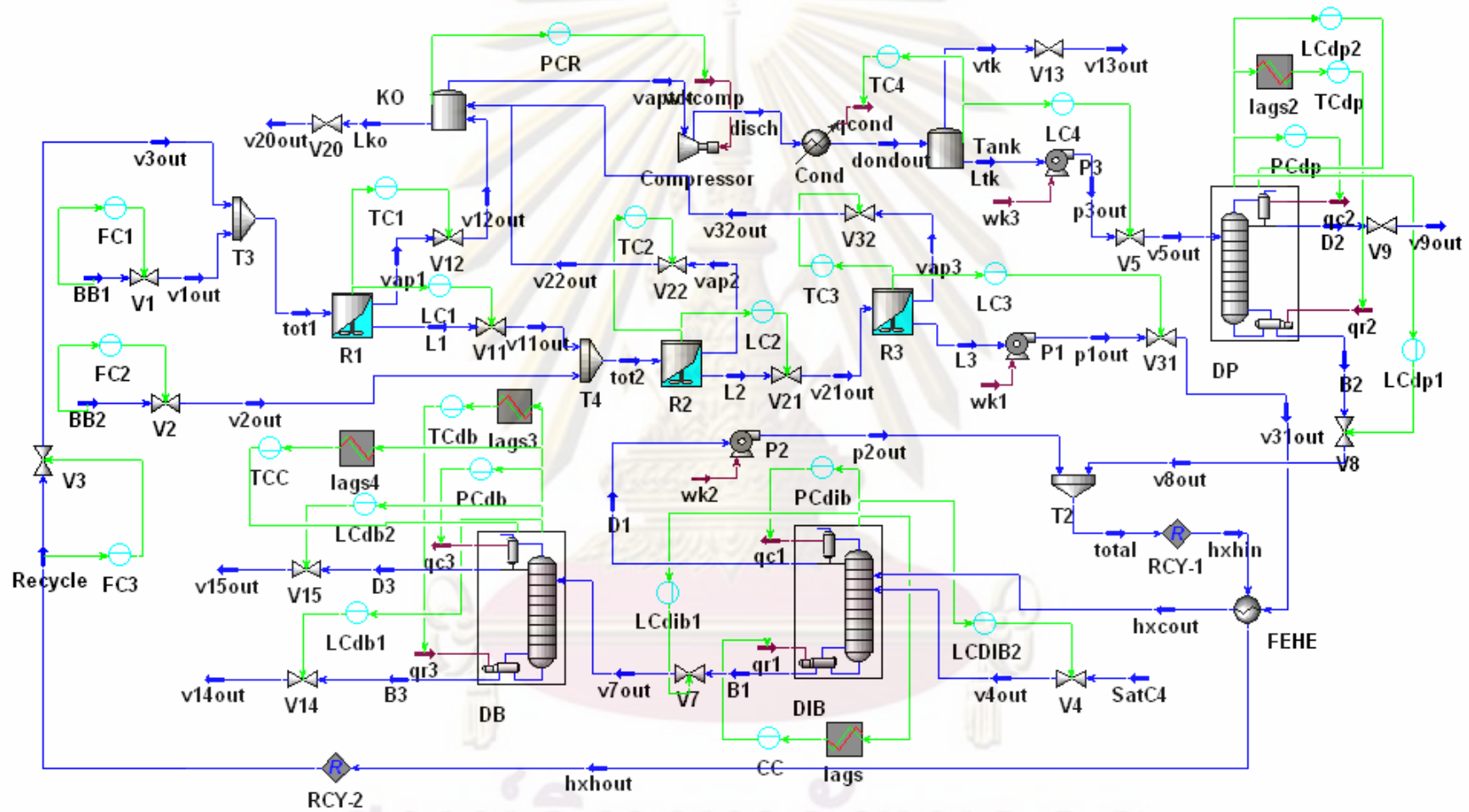
Figure 5.14 Designed control structure VII (CS7) for alkylation Process
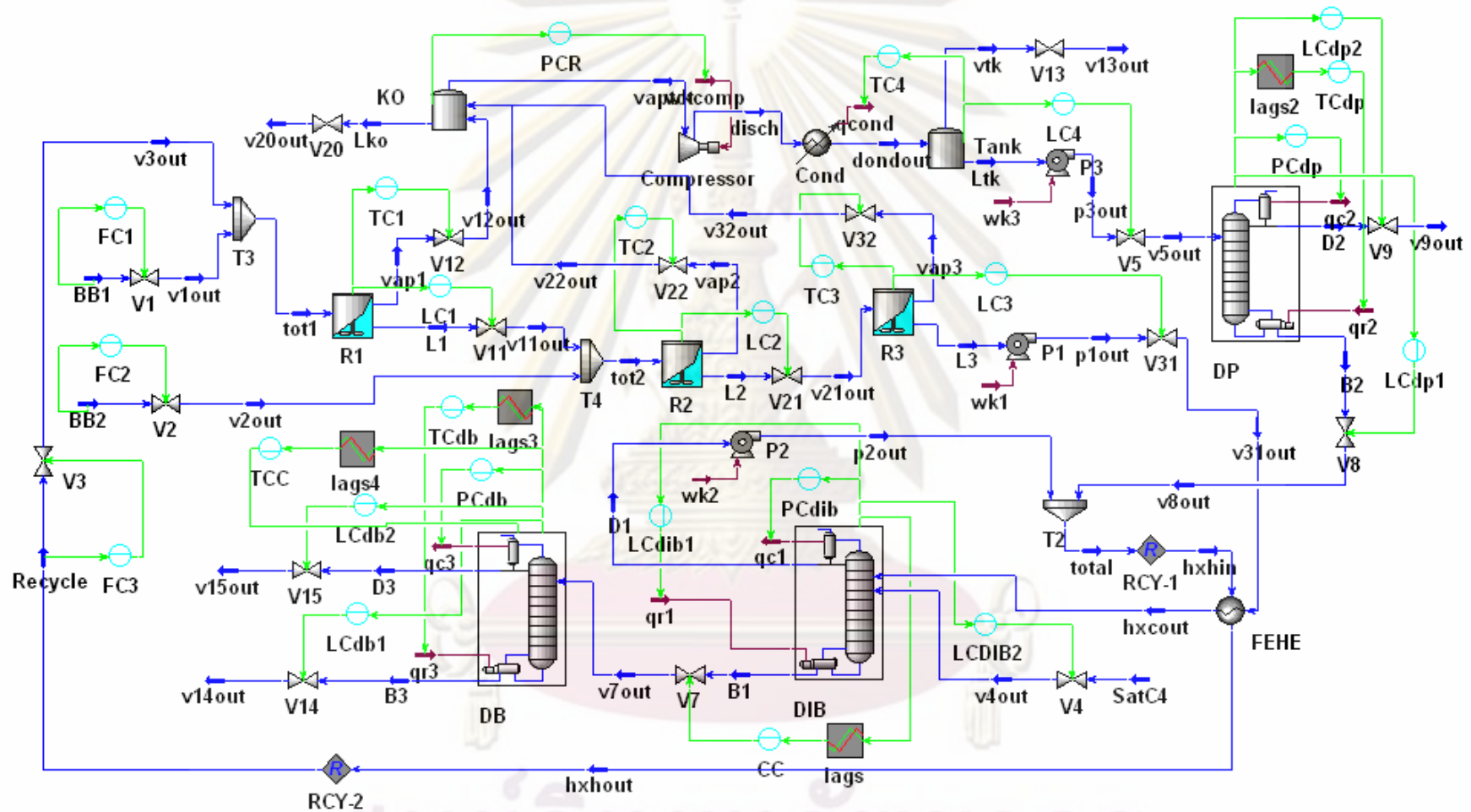
Figure 5.15 Designed control structure VIII (CS8) for alkylation Process

5.3 Dynamic Simulation Results

In order to illustrate the dynamic behaviors of base case control structure by Luyben (2002) and all designed control structures using new design procedure of Wongsri (2009), two types of disturbances: material and thermal disturbances are used in evaluation of the base case control structure (CS0), the designed control structure I (CS1) to the designed control structure VIII (CS8) respectively for alkylation process.

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

(1) Change in two fresh feeds molar flow rate by step change of molar flow rate ±10% of two fresh feeds (BB1 and BB2) from 25 lbmol/hr to 27.5 lbmol/hr at 50 min to 250 min and from 27.5 lbmol/hr to 22.5 lbmol/hr at 250 min to 450 min.

(2) Change in two fresh feeds temperature by step change of temperature ±10% of two fresh feeds (BB1 and BB2) from 90°F to 99 $^{\circ}$F at 50 min to 250 min and from 99°F to 81 $^{\circ}$F at 250 min to 450 min.

Temperature controllers are PIDs which are tuned using relay feedback. Three temperature measurement lags of 0.5 minute are included in the three temperature loops (25th Tray temperature of depropanizer (DP) column, 3rd and 10th Tray temperature of debutanizer (DB) column). Iso-butane composition is measured and controlled using PID controller. Two iso-butane composition measurement lags of 0.5 minute are the 10th and 19th de-isobutanizer (DIB) column. Flow and pressure controller are PIs and their parameters are heuristics values. Proportional-only level controllers are used and their parameters are heuristics values. All control valves are half-open at nominal operating condition.

5.3.1 Change in material disturbances of two fresh feeds flow rate for all control structures (base case control structure (CS0), designed control structure I (CS1) to designed control structure VIII (CS8))

Dynamic responses for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2) increase from 25 lbmol/hr to 27.5 lbmol/hr at 50 min to 250 min and decrease from 27.5 lbmol/hr to 22.5 lbmol/hr at 250 min to 450 min. As shown in Figure 5.16. The graph of dynamic responses of base case control structure (CS0), designed control structure I (CS1) to designed control structure VIII (CS8) show in Figure 5.17 to Figure 5.25.
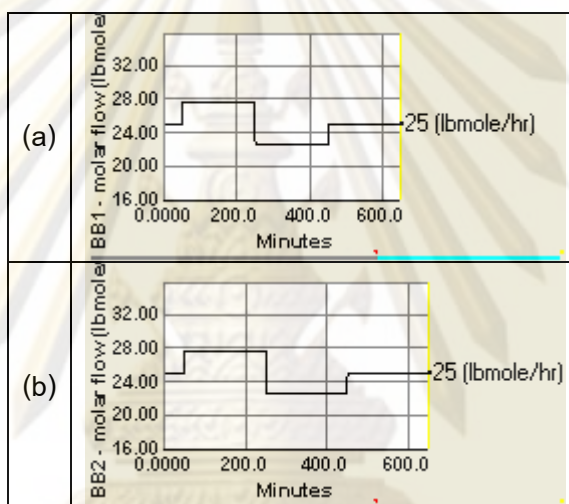
Figure 5.16 Dynamic responses for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2) where (a) is a molar flow rate of BB1 and (b) is a molar flow rate of BB2

Figure 5.17 Dynamic responses of base case control structure (CS0) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.17 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 10th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.17 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column

Figure 5.17 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column

Figure 5.17 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.18 Dynamic responses of designed control structure I (CS1) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.18 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.18 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column

Figure 5.18 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column

Figure 5.18 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.19 Dynamic responses of designed control structure II (CS2) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.19 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.19 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column

Figure 5.19 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is a condenser duty of DB column and (t) is a reboiler duty of DB column

Figure 5.19 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.20 Dynamic responses of designed control structure III (CS3) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.20 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.20 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column

Figure 5.20 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column

Figure 5.20 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (v) and (w) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (x) is a composition of iso-octane in product stream of DIB column, (y) is a composition of propane in distillate stream of DP column, (z) is a composition of iso-butane in product stream of DP column, (aa) is a composition of n-butane in distillate stream of DB column and (ab) and (ac) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.21 Dynamic responses of designed control structure IV (CS4) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.21 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.21 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column

Figure 5.21 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column
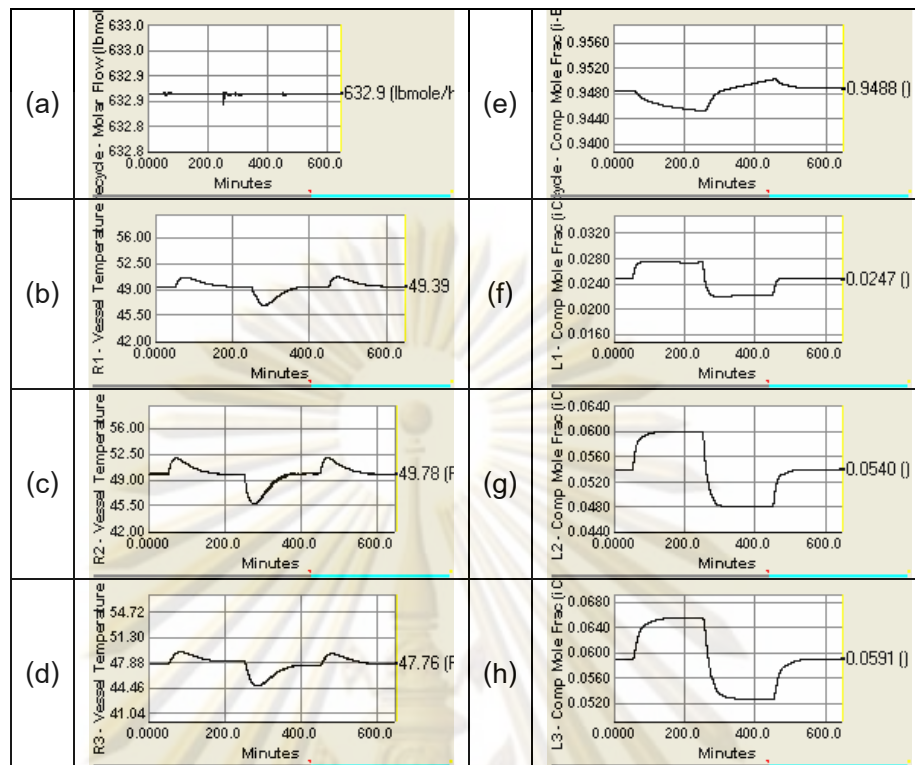
Figure 5.21 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.22 Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
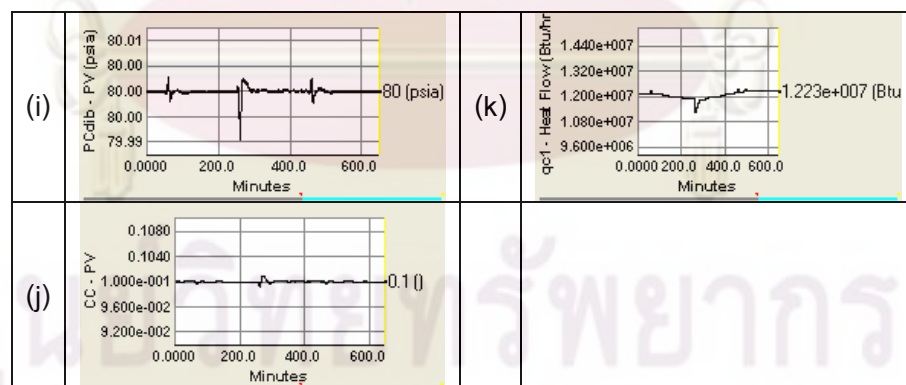

Figure 5.22 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column
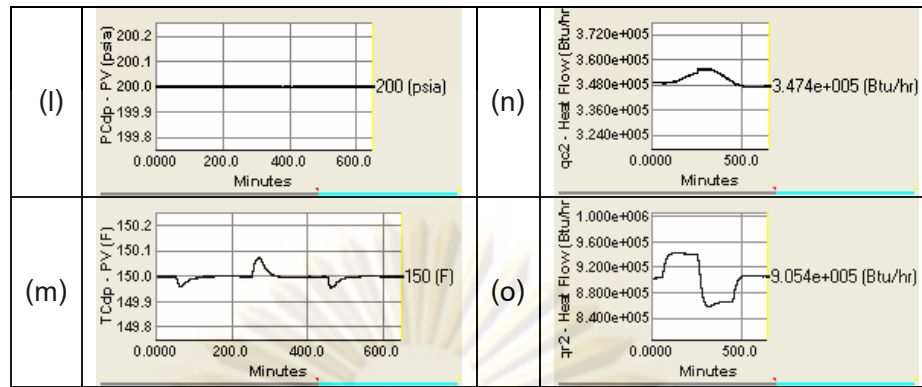
Figure 5.22 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column
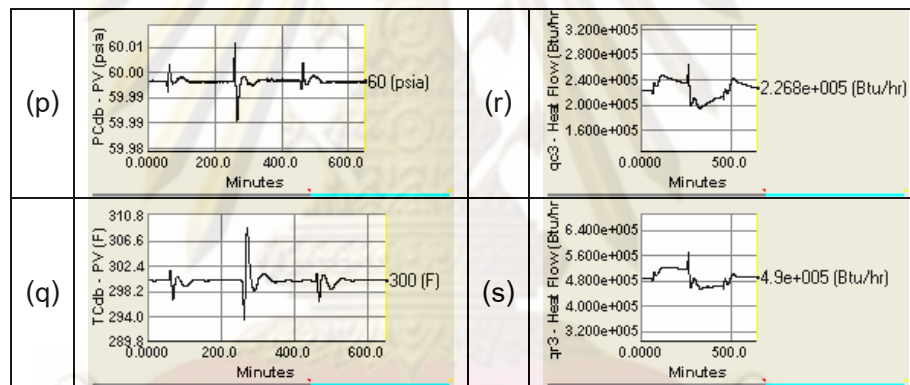

Figure 5.22 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column
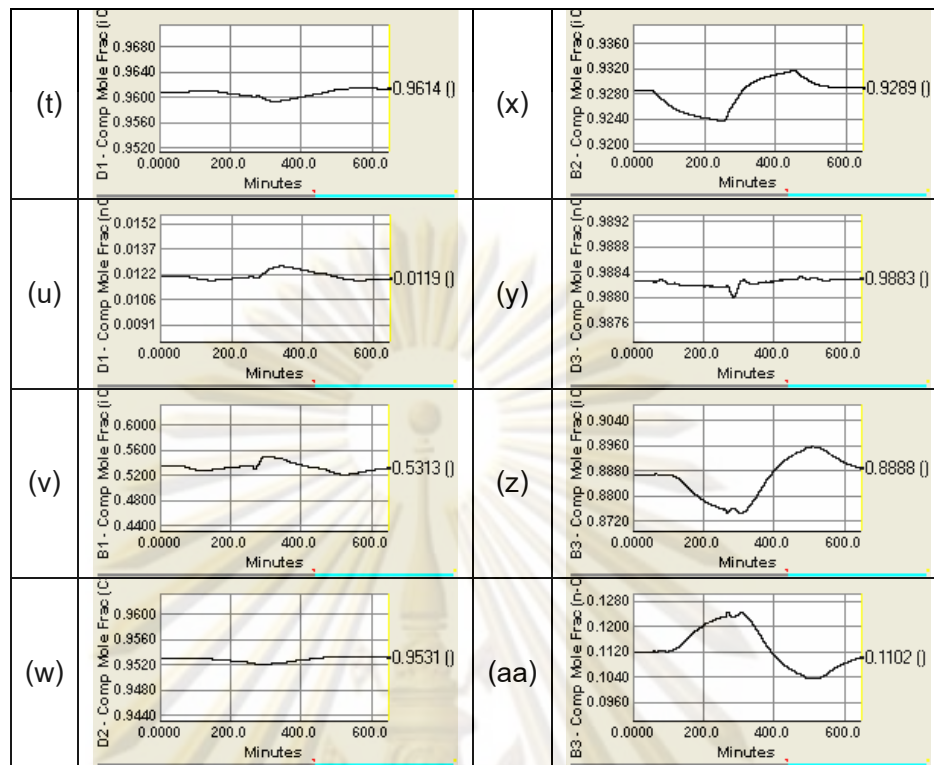
Figure 5.22 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively
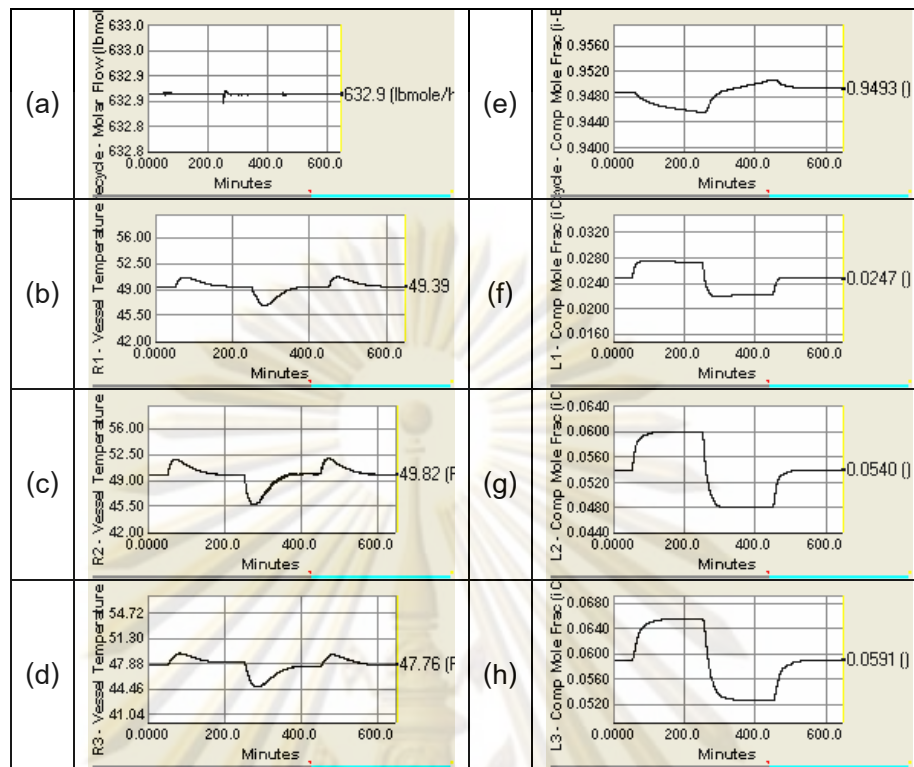
Figure 5.23 Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
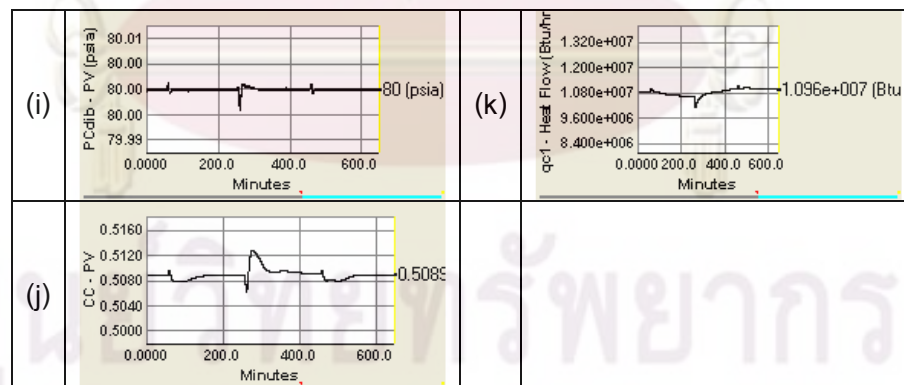

Figure 5.23 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column
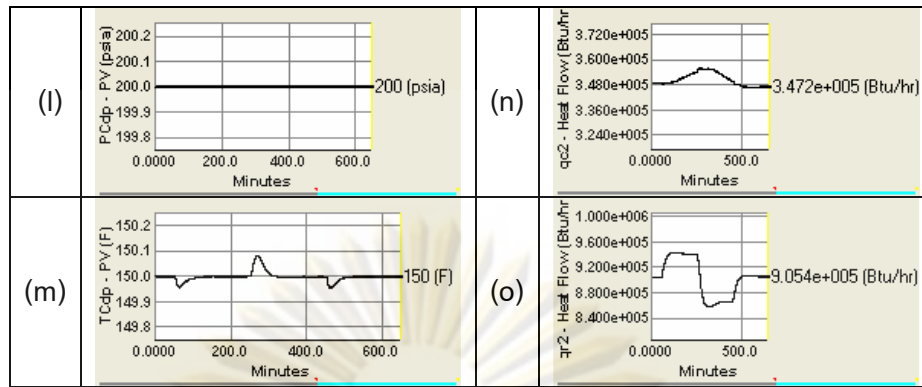
Figure 5.23 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column

Figure 5.23 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is a condenser duty of DB column and (t) is a reboiler duty of DB column
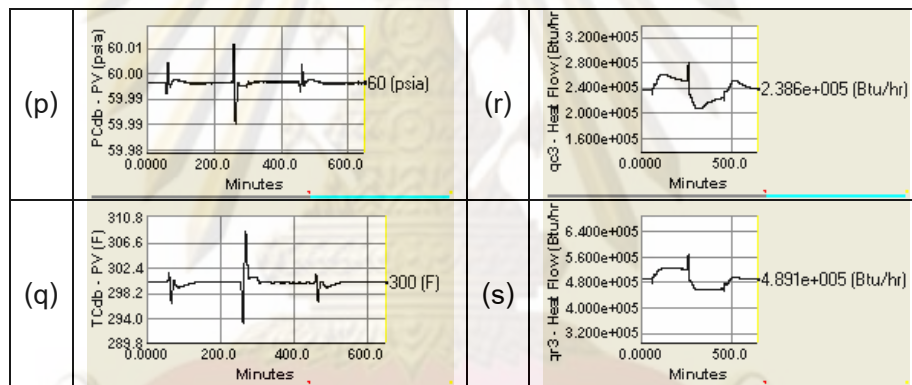
Figure 5.23 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.24 Dynamic responses of designed control structure VII (CS7) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

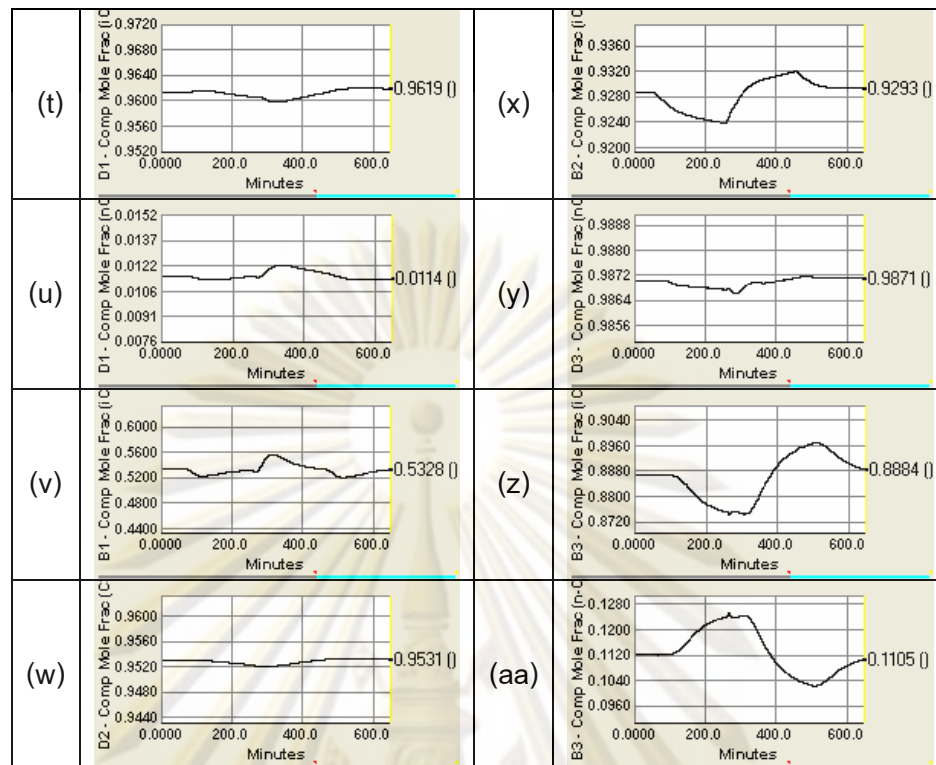**Figure 5.24 (Continued)** Dynamic responses of designed control structure VII (CS7) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.24 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column
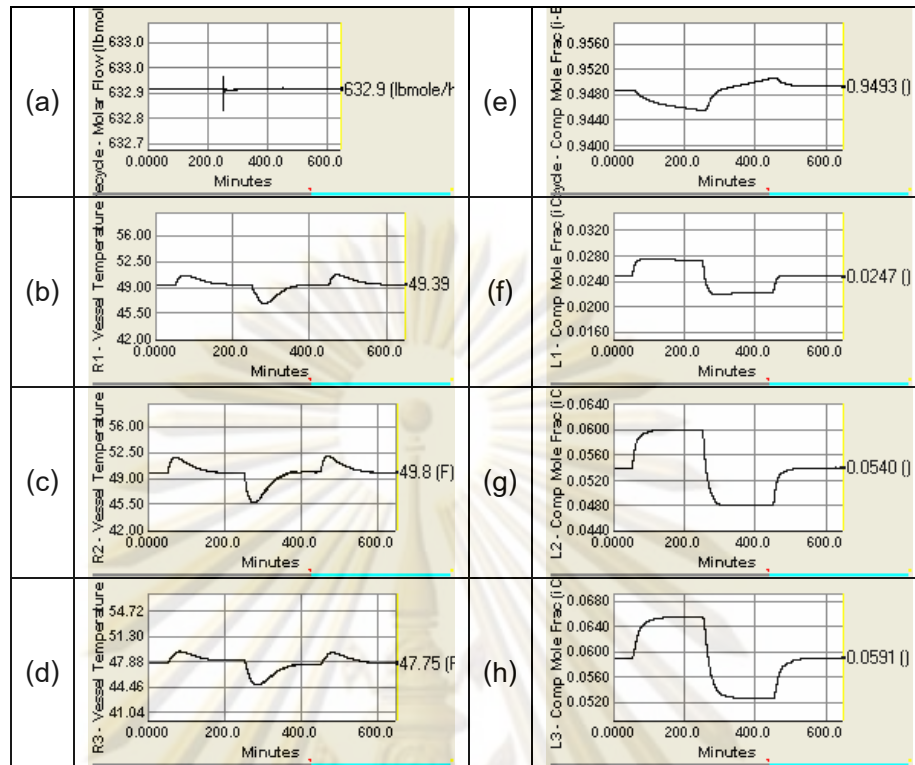

Figure 5.24 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column

Figure 5.24 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (v) and (w) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (x) is a composition of iso-octane in product stream of DIB column, (y) is a composition of propane in distillate stream of DP column, (z) is a composition of iso-butane in product stream of DP column, (aa) is a composition of n-butane in distillate stream of DB column and (ab) and (ac) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.25 Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
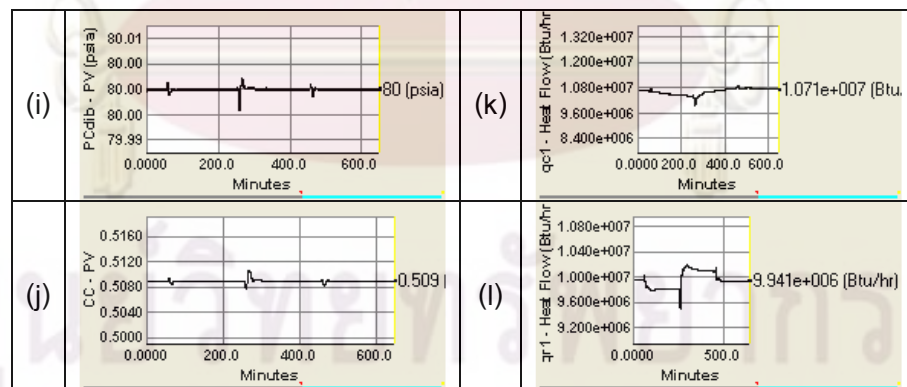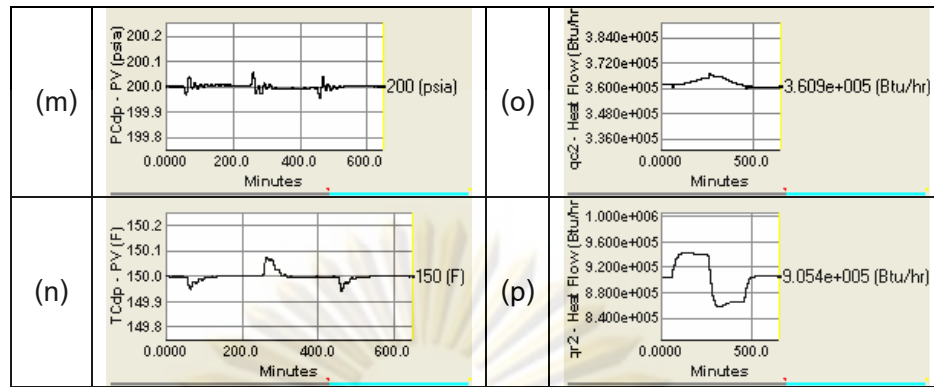

Figure 5.25 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.25 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column
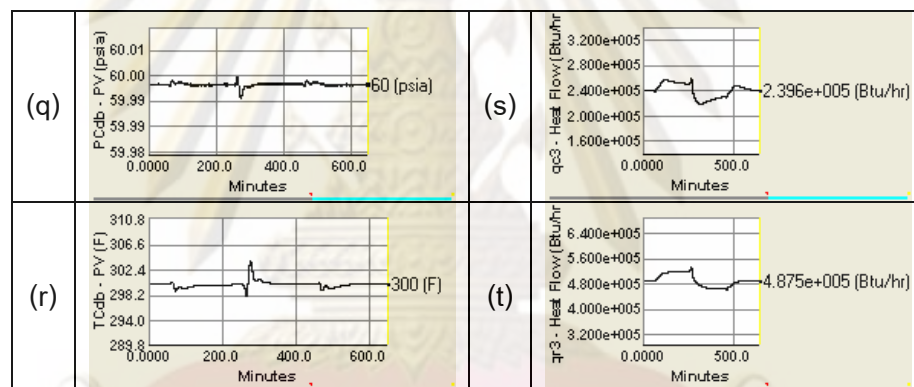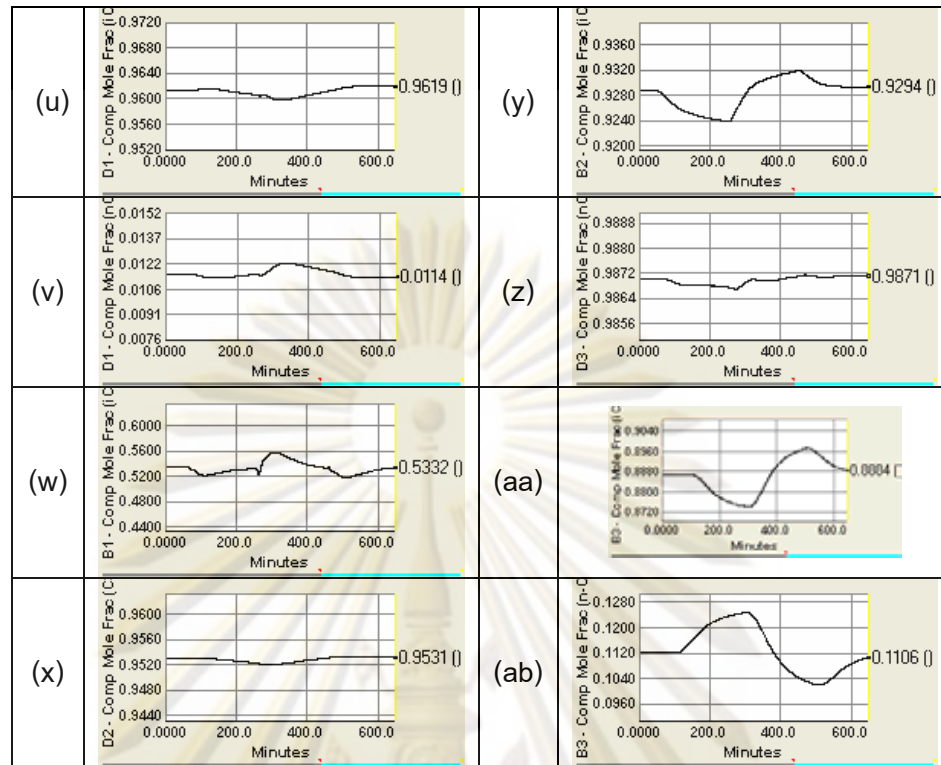

Figure 5.25 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column

Figure 5.25 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar flow rate ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively
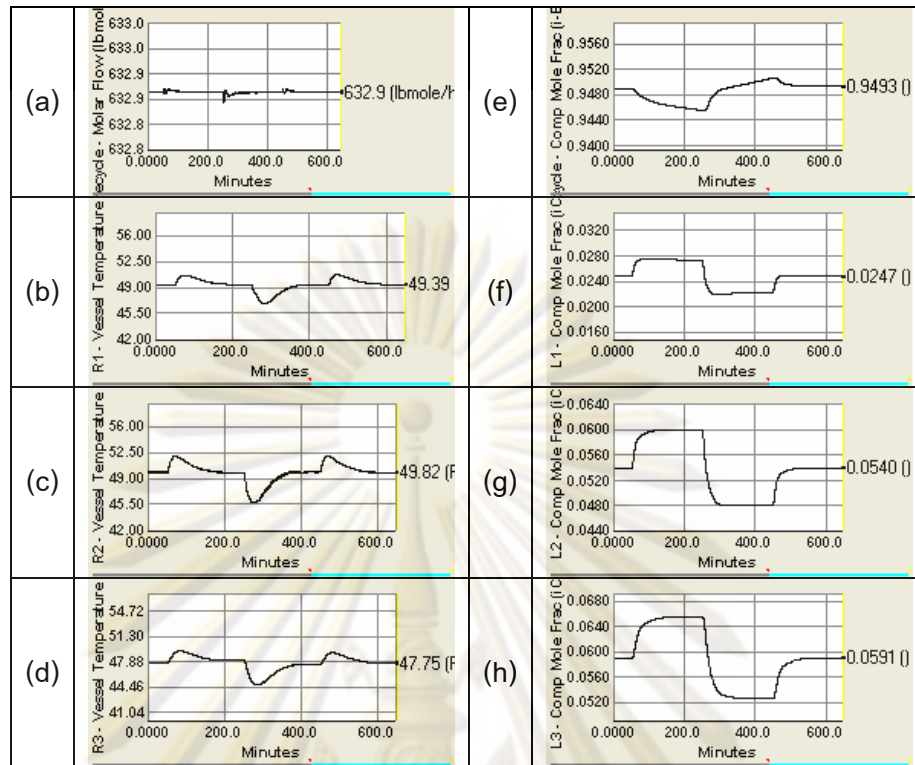
5.3.2 Change in thermal disturbances of two fresh feeds temperature for all control structures (base case control structure (CS0), designed control structure I (CS1) to designed control structure VIII (CS8))

Dynamic responses for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2) increase from 90°F to 99 $^{\circ}$F at 50 min to 250 min and decrease from 99°F to 81 $^{\circ}$F at 250 min to 450 min. As shown in Figure 5.26. The graph of dynamic responses of base case control structure (CS0), designed control structure I (CS1) to designed control structure VIII (CS8) show in Figure 5.27 to Figure 5.35.
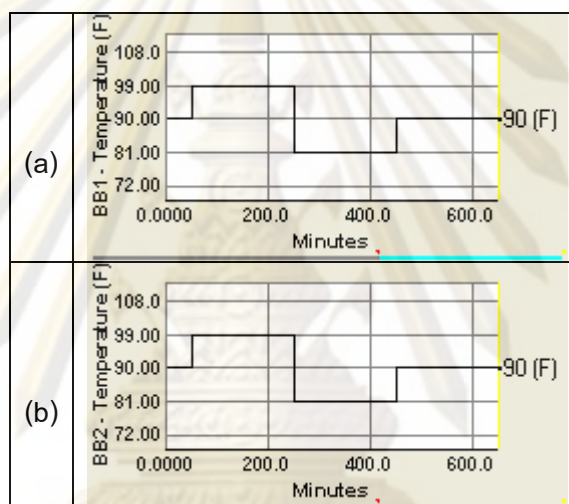
Figure 5.26 Dynamic responses for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2) where (a) is a temperature of BB1 and (b) is a temperature of BB2

Figure 5.27 Dynamic responses of base case control structure (CS0) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
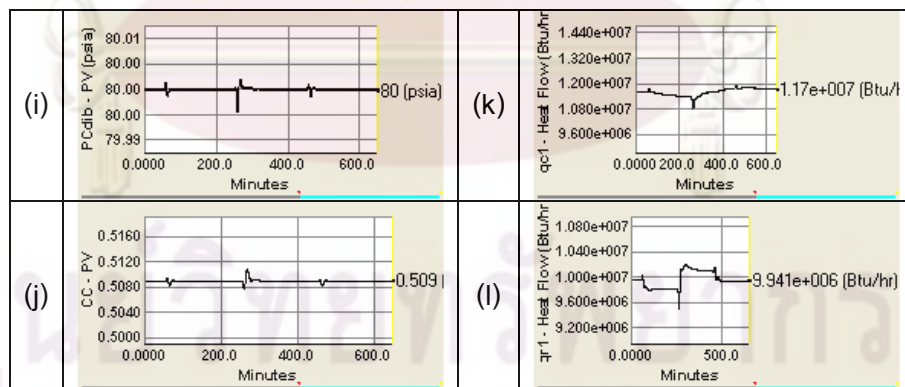

Figure 5.27 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 10th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.27 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column
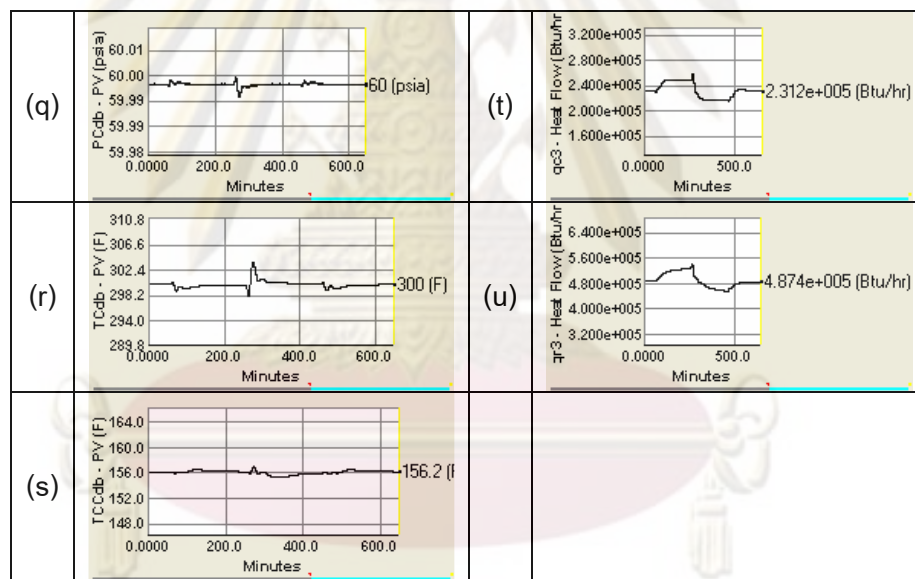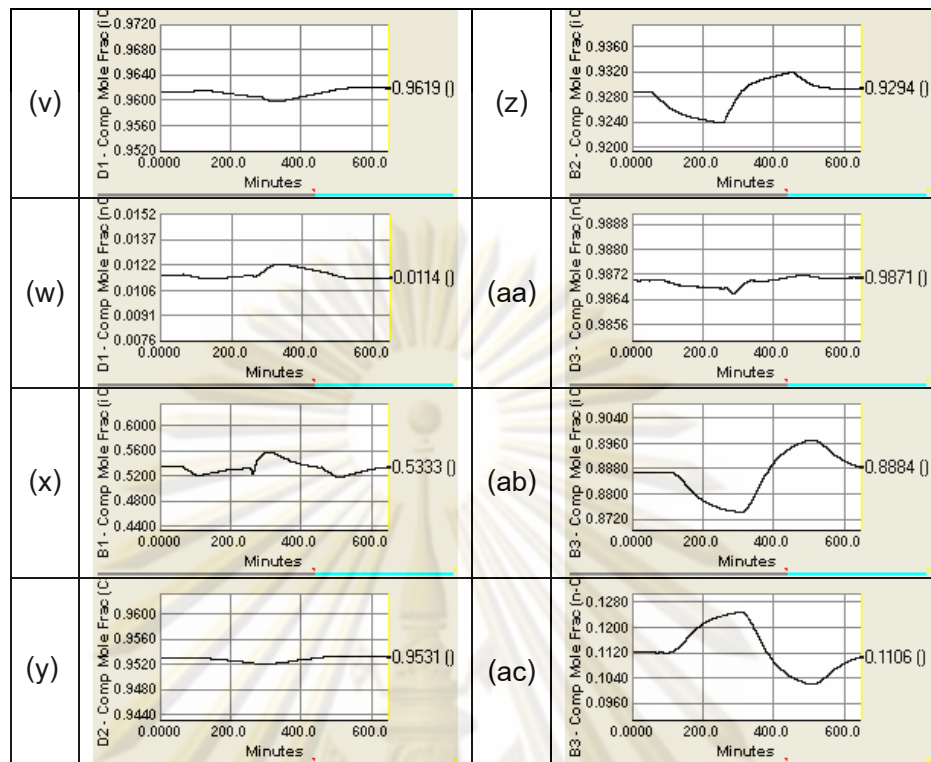

Figure 5.27 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column

Figure 5.27 (Continued) Dynamic responses of base case control structure (CS0) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively
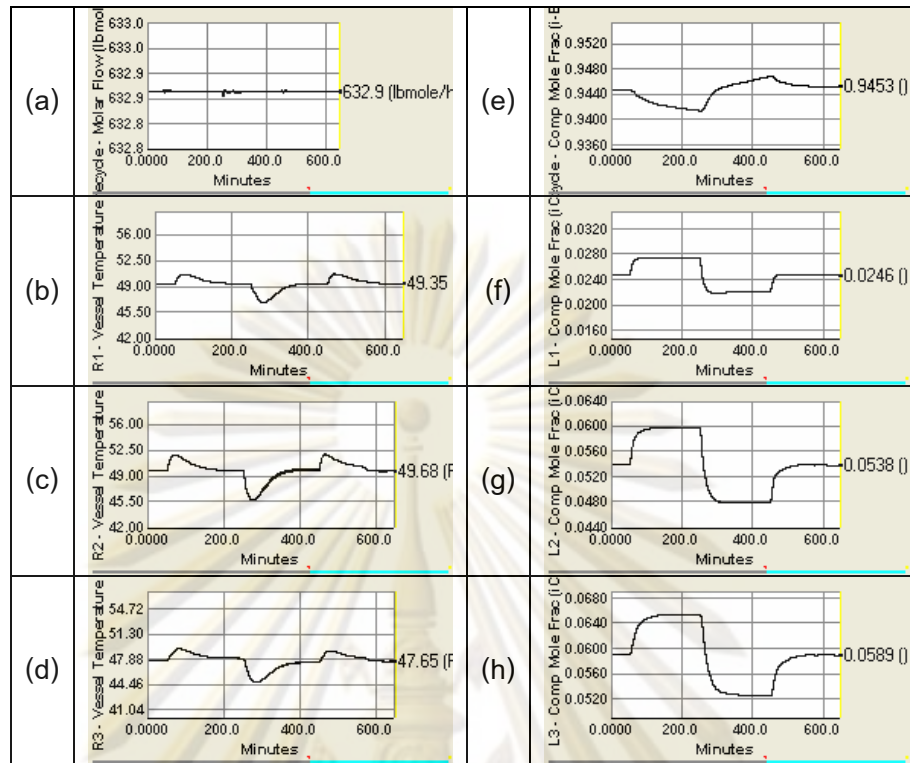
Figure 5.28 Dynamic responses of designed control structure I (CS1) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
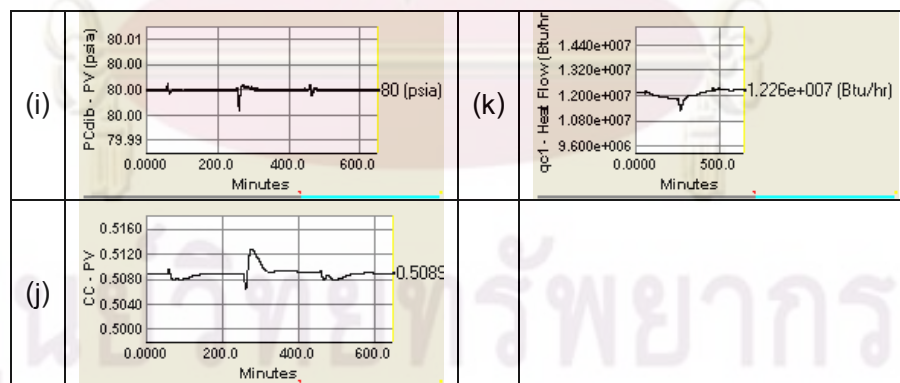

Figure 5.28 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column
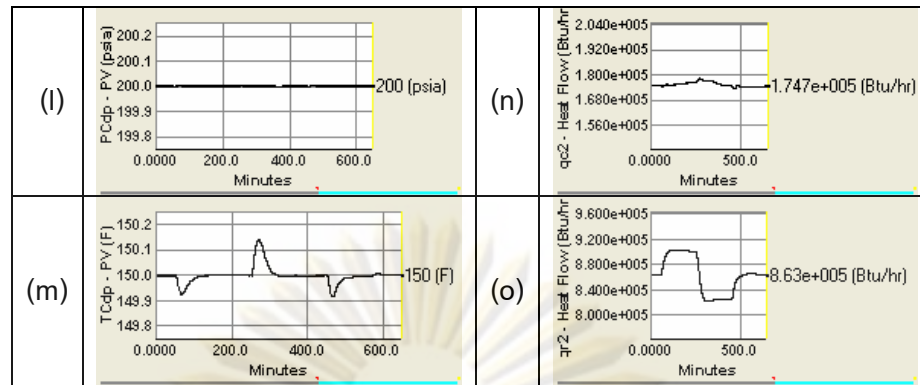
Figure 5.28 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column
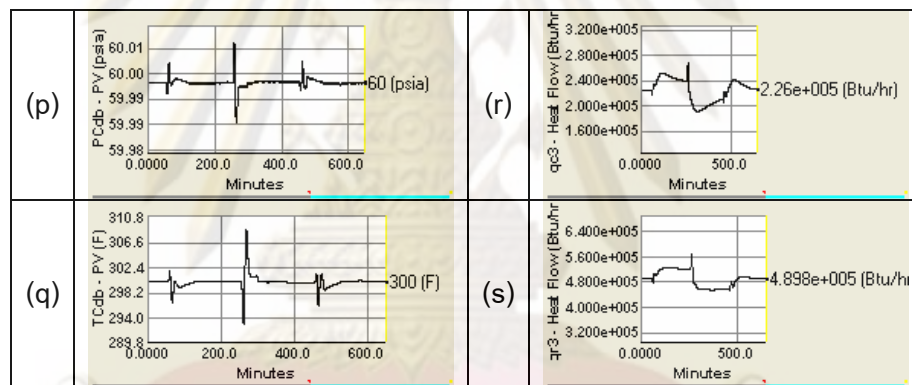

Figure 5.28 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column
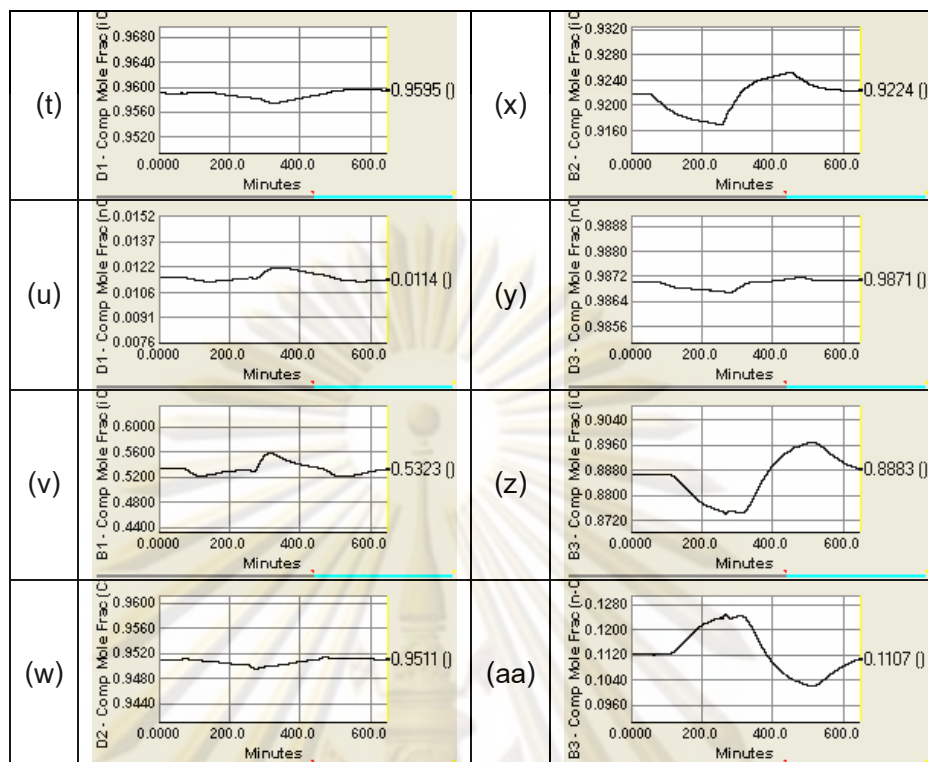
Figure 5.28 (Continued) Dynamic responses of designed control structure I (CS1) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively
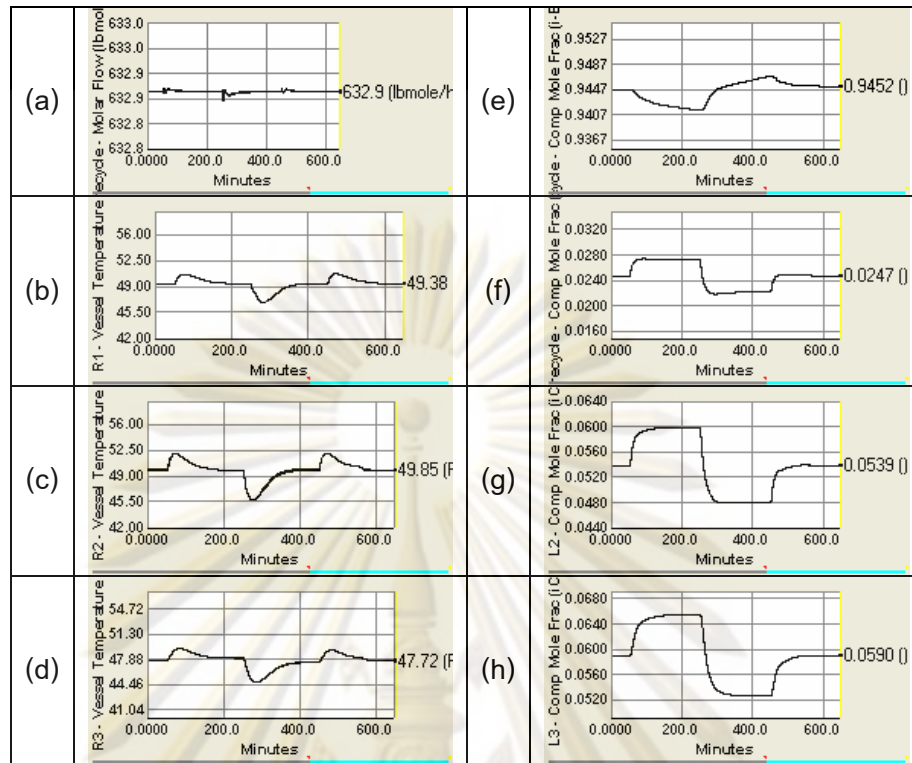
Figure 5.29 Dynamic responses of designed control structure II (CS2) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.29 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.29 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column
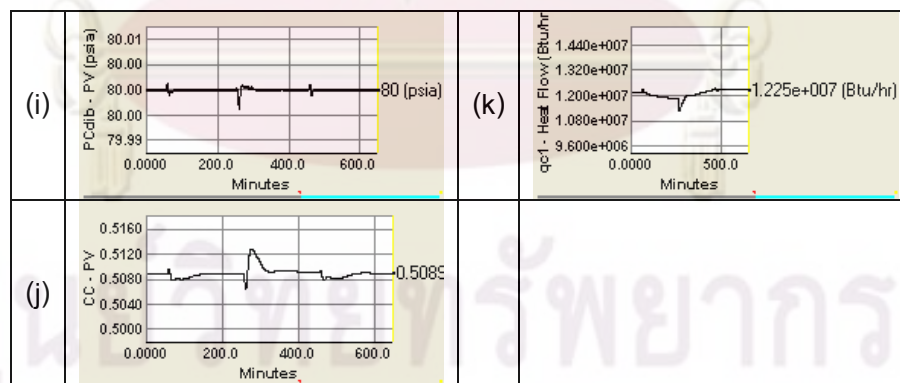

Figure 5.29 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is a condenser duty of DB column and (t) is a reboiler duty of DB column
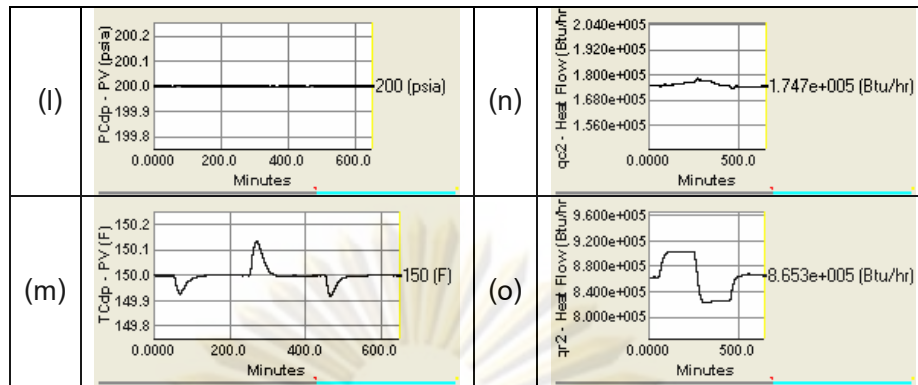
Figure 5.29 (Continued) Dynamic responses of designed control structure II (CS2) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.30 Dynamic responses of designed control structure III (CS3) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.30 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.30 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column
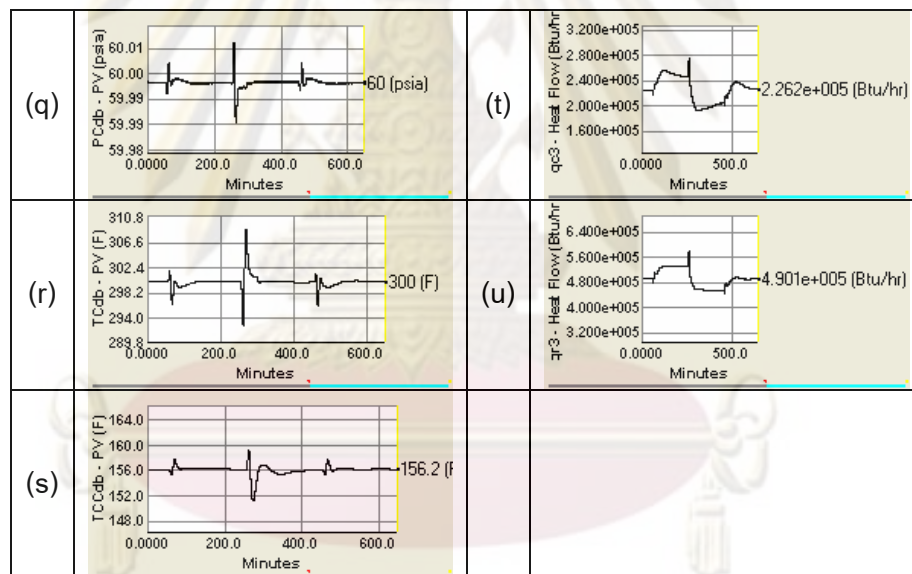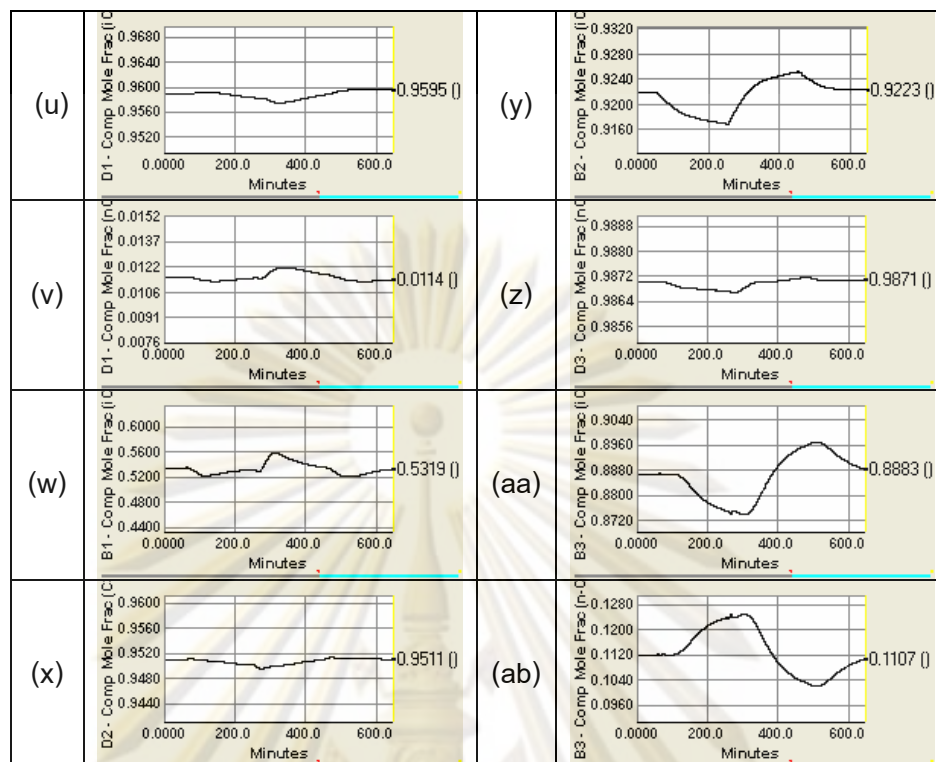

Figure 5.30 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column
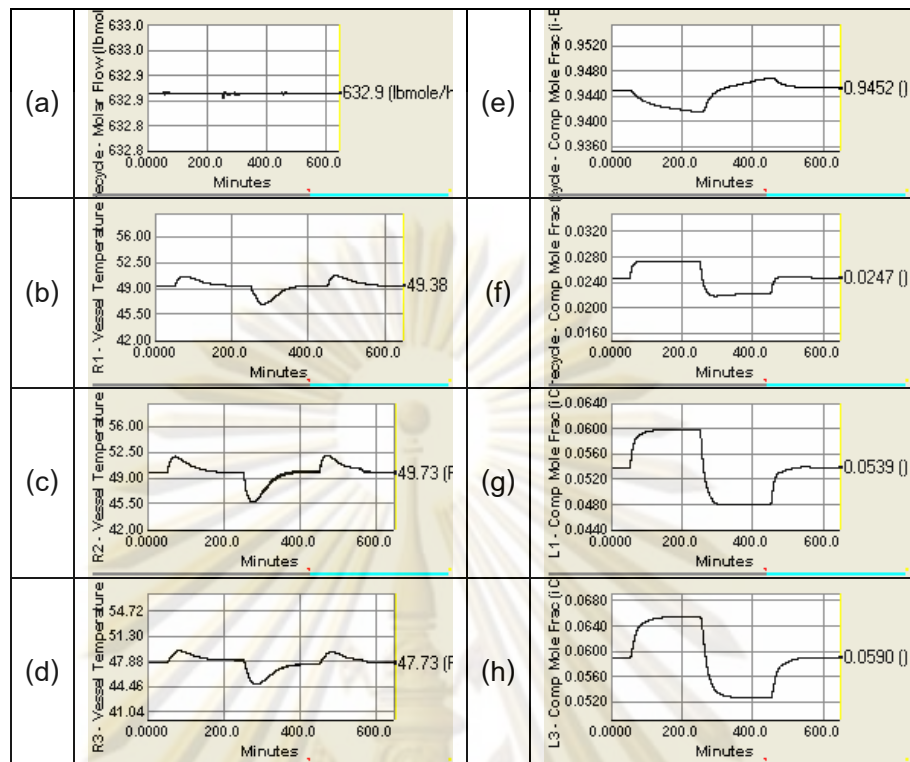
Figure 5.30 (Continued) Dynamic responses of designed control structure III (CS3) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (v) and (w) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (x) is a composition of iso-octane in product stream of DIB column, (y) is a composition of propane in distillate stream of DP column, (z) is a composition of iso-butane in product stream of DP column, (aa) is a composition of n-butane in distillate stream of DB column and (ab) and (ac) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.31 Dynamic responses of designed control structure IV (CS4) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
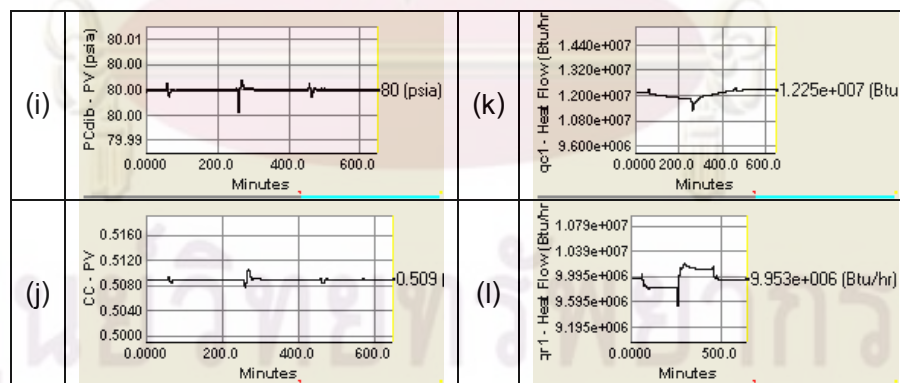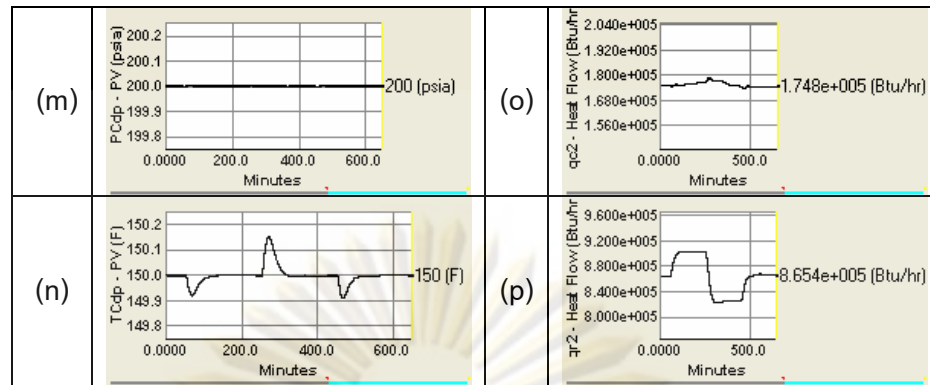

Figure 5.31 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

Figure 5.31 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column
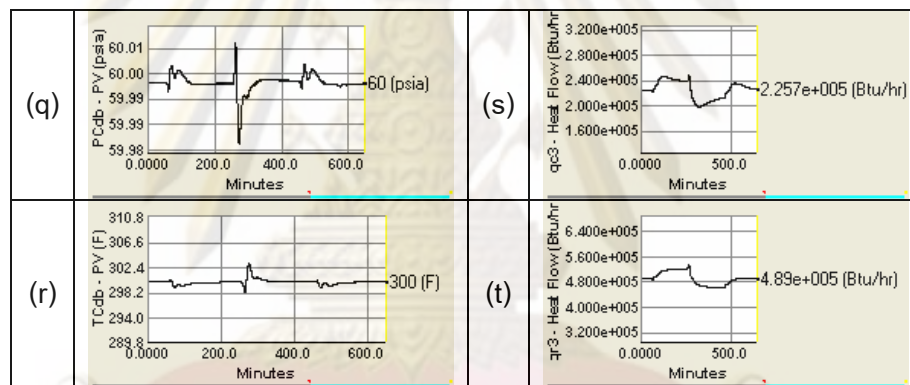

Figure 5.31 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (p) is a pressure condenser of debutanizer (DB) column, (q) is 3rd Tray temperature of DB column, (r) is a condenser duty of DB column and (s) is a reboiler duty of DB column
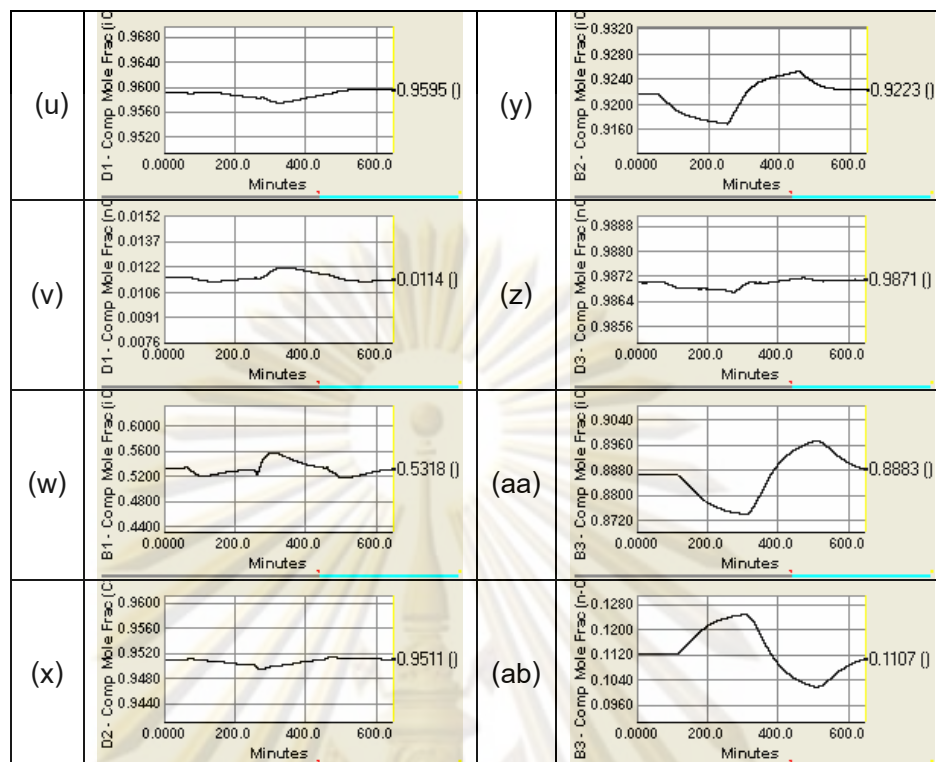
Figure 5.31 (Continued) Dynamic responses of designed control structure IV (CS4) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (t) and (u) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (v) is a composition of iso-octane in product stream of DIB column, (w) is a composition of propane in distillate stream of DP column, (x) is a composition of iso-butane in product stream of DP column, (y) is a composition of n-butane in distillate stream of DB column and (z) and (aa) is a composition of iso-octane and dodecane in product stream of DB column respectively
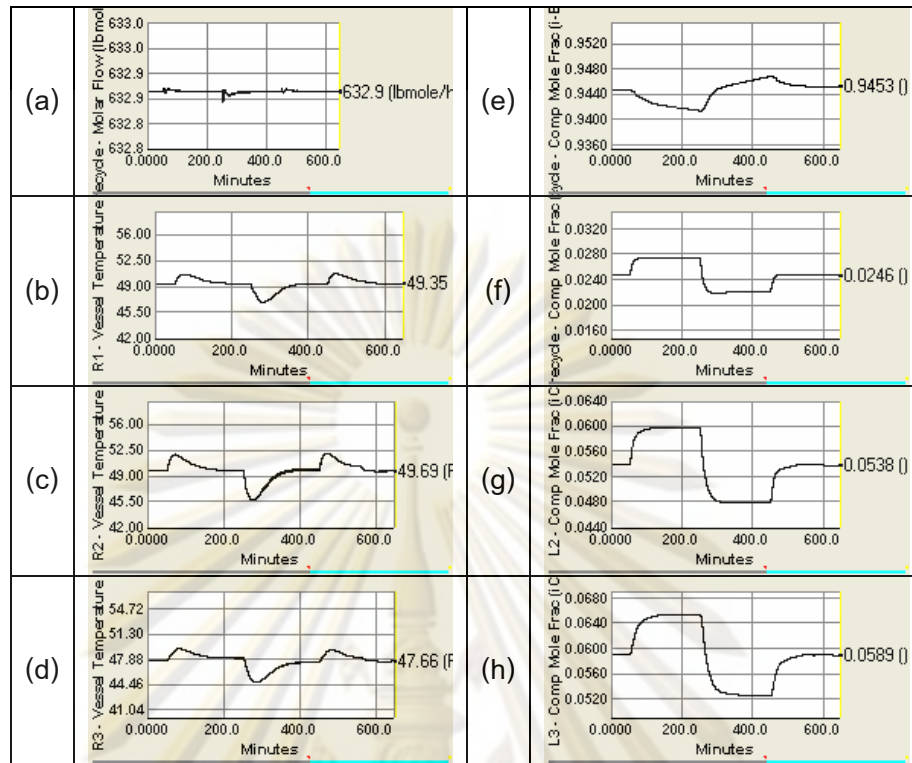
Figure 5.32 Dynamic responses of designed control structure V (CS5) for alkylation process by step change molar temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.32 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column
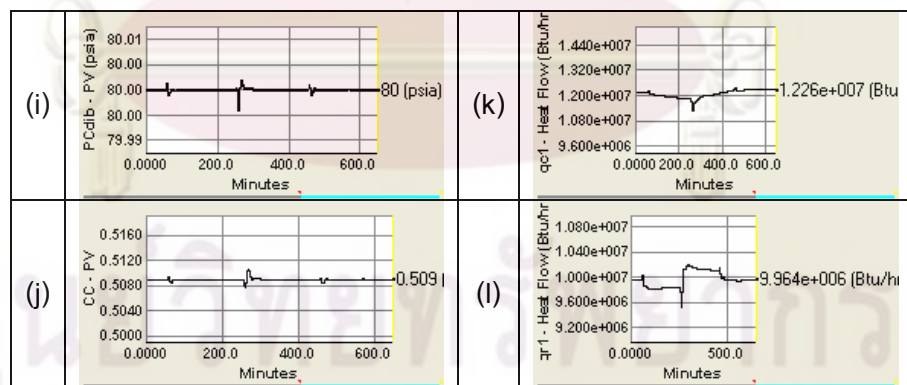
Figure 5.32 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column

Figure 5.32 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column
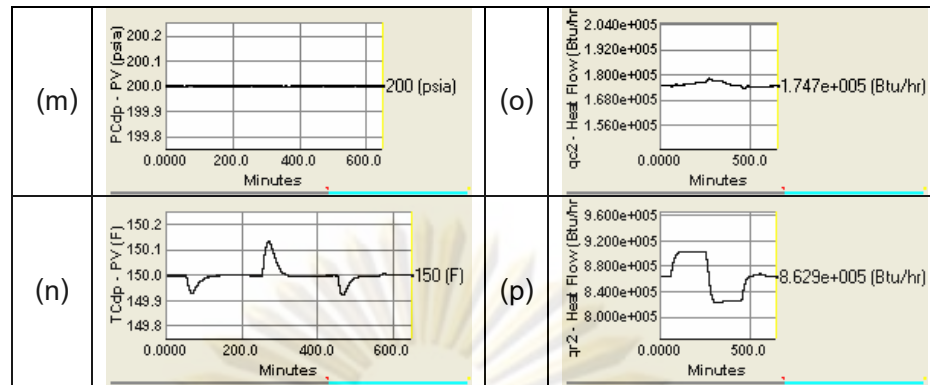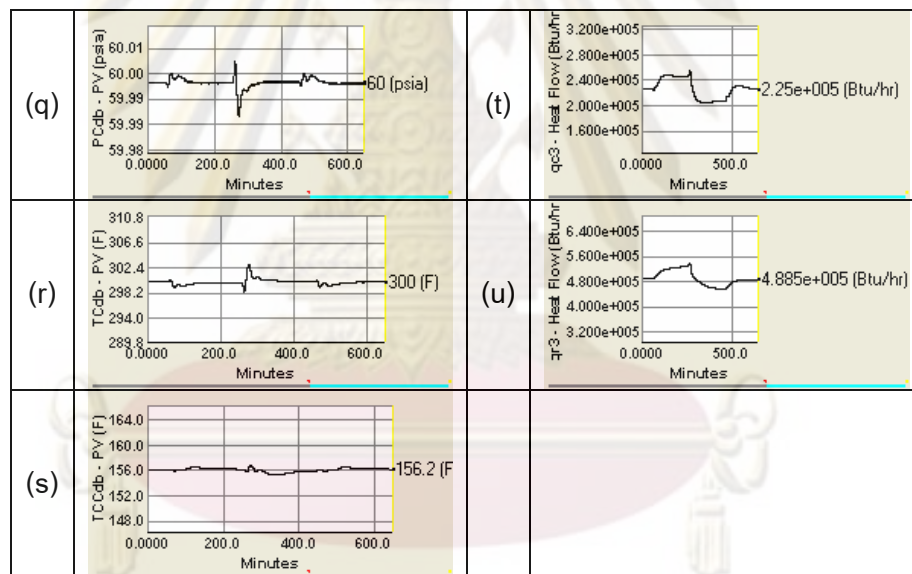
Figure 5.32 (Continued) Dynamic responses of designed control structure V (CS5) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.33 Dynamic responses of designed control structure VI (CS6) for alkylation process by step change molar temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

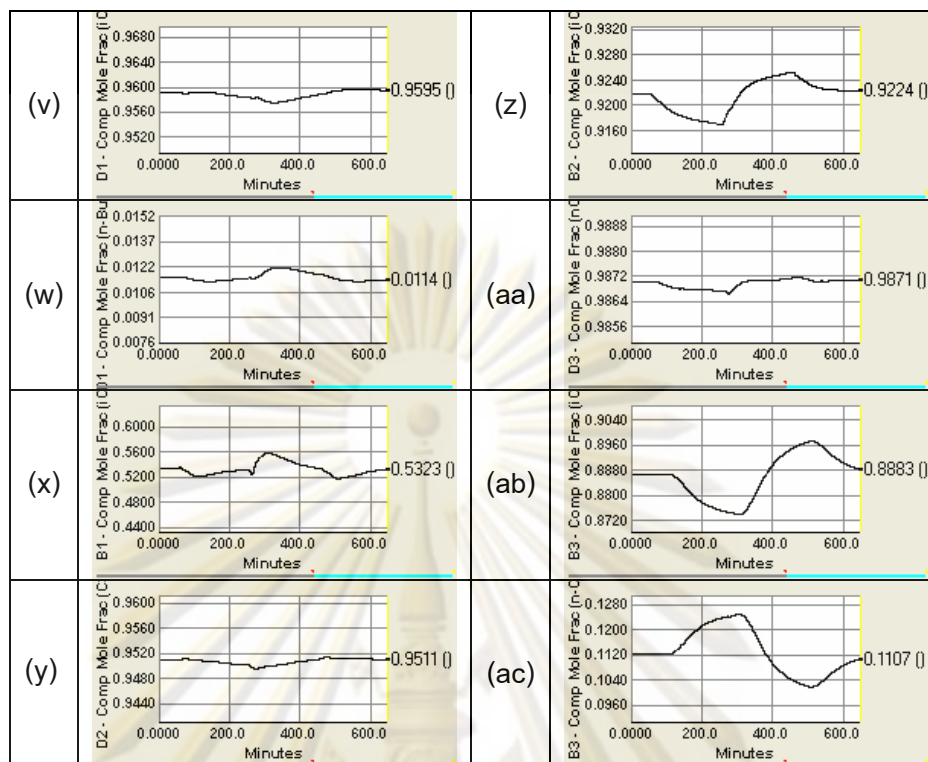**Figure 5.33 (Continued)** Dynamic responses of designed control structure VI (CS6) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column

Figure 5.33 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column

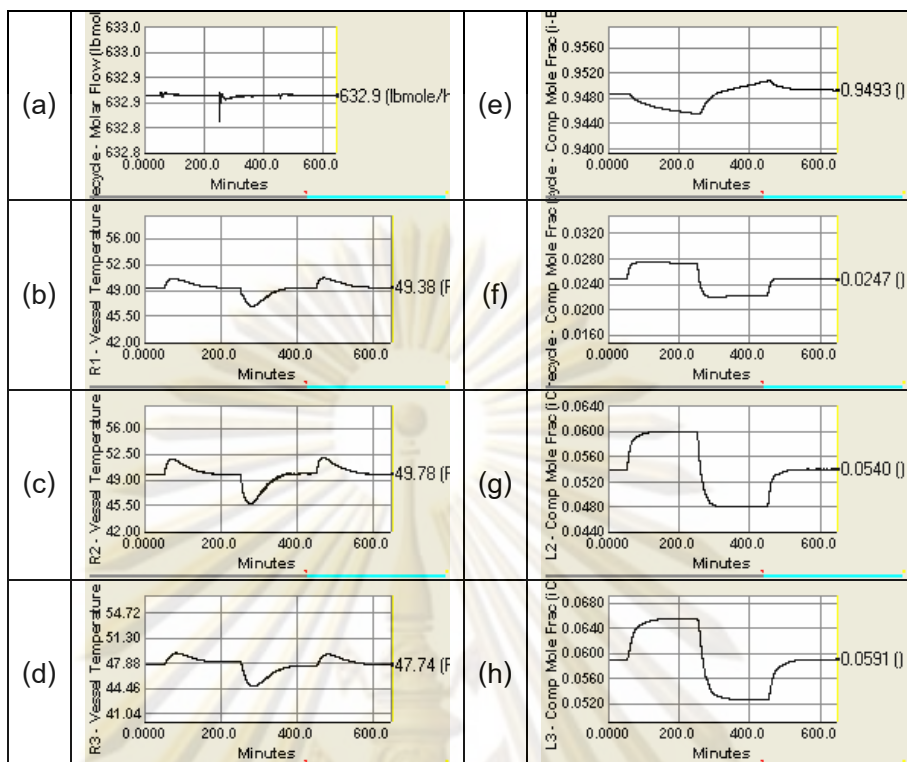Figure 5.33 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is a condenser duty of DB column and (t) is a reboiler duty of DB column

Figure 5.33 (Continued) Dynamic responses of designed control structure VI (CS6) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.34 Dynamic responses of designed control structure VII (CS7) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively
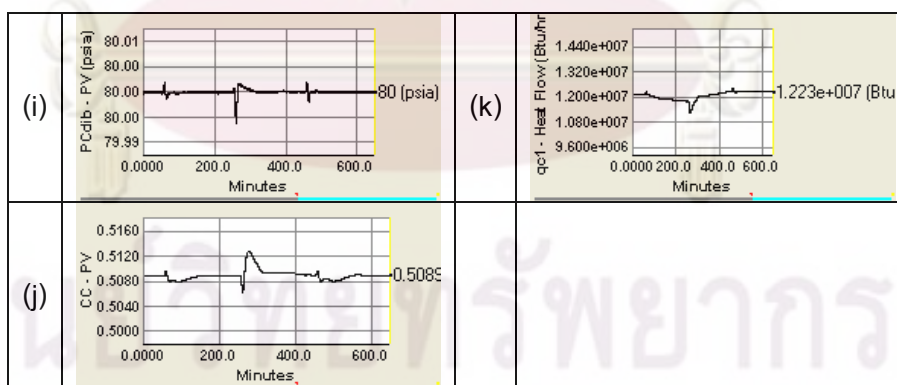

Figure 5.34 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column, (k) is a condenser duty of DIB column and (l) is a reboiler duty of DIB column
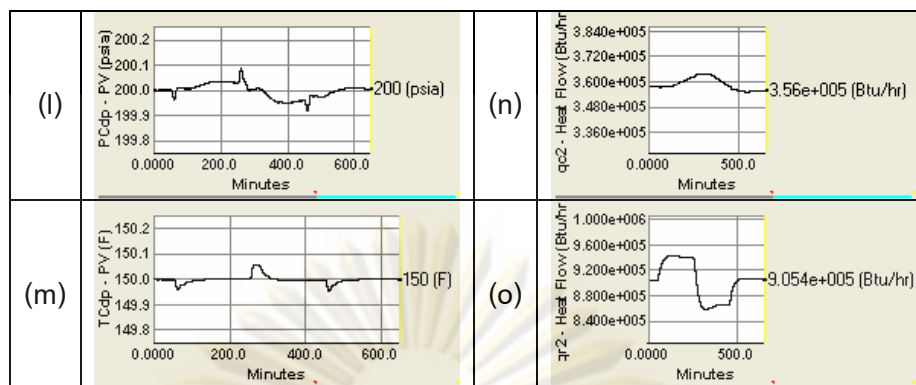
Figure 5.34 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (m) is a pressure condenser of depropanizer (DP) column, (n) is 25th Tray temperature of DP column, (o) is a condenser duty of DP column and (p) is a reboiler duty of DP column
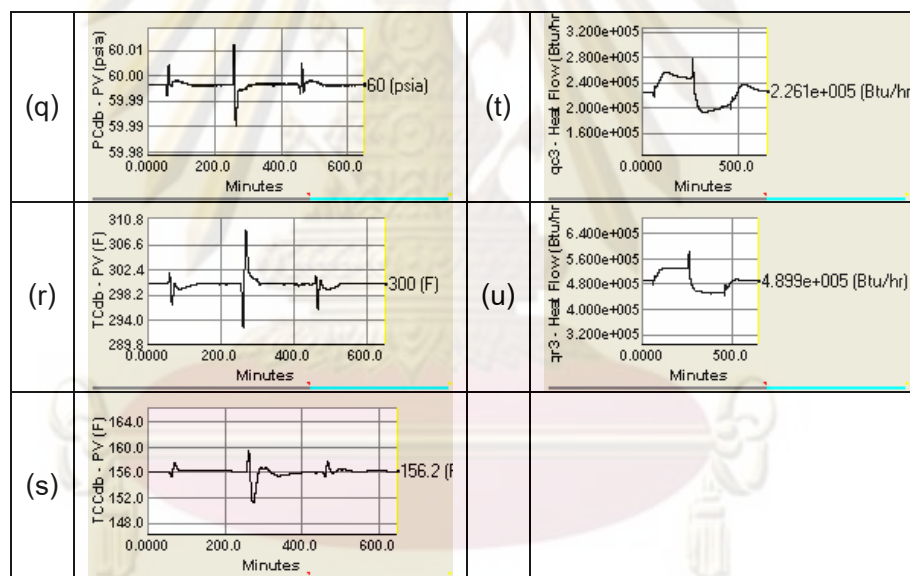

Figure 5.34 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column
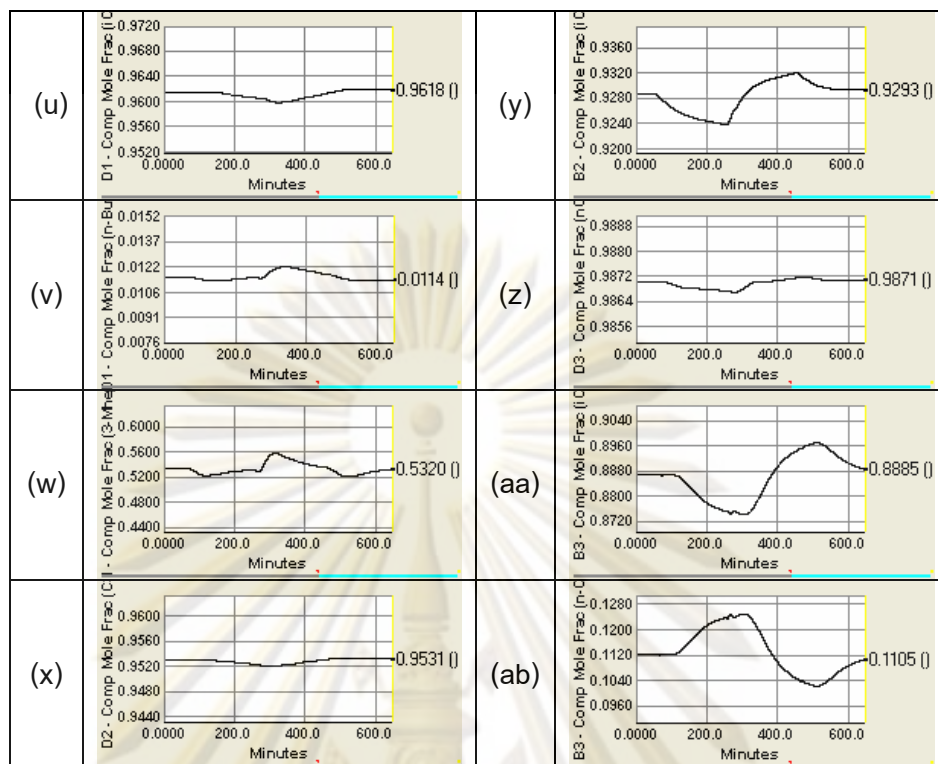
Figure 5.34 (Continued) Dynamic responses of designed control structure VII (CS7) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (v) and (w) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (x) is a composition of iso-octane in product stream of DIB column, (y) is a composition of propane in distillate stream of DP column, (z) is a composition of iso-butane in product stream of DP column, (aa) is a composition of n-butane in distillate stream of DB column and (ab) and (ac) is a composition of iso-octane and dodecane in product stream of DB column respectively

Figure 5.35 Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change molar temperature ±10% in two fresh feeds (BB1 and BB2), where (a) is a molar flow rate of recycle stream, (b), (c), (d) is a temperature of reactor 1, 2 and 3 respectively, (e) is a iso-butane composition of recycle stream and (f), (g) and (h) is a iso-octane composition in liquid effluent reactor 1, 2 and 3 respectively

Figure 5.35 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (i) is a pressure condenser of de-isobutanizer (DIB) column, (j) is 19th Tray iso-butane composition of DIB column and (k) is a condenser duty of DIB column

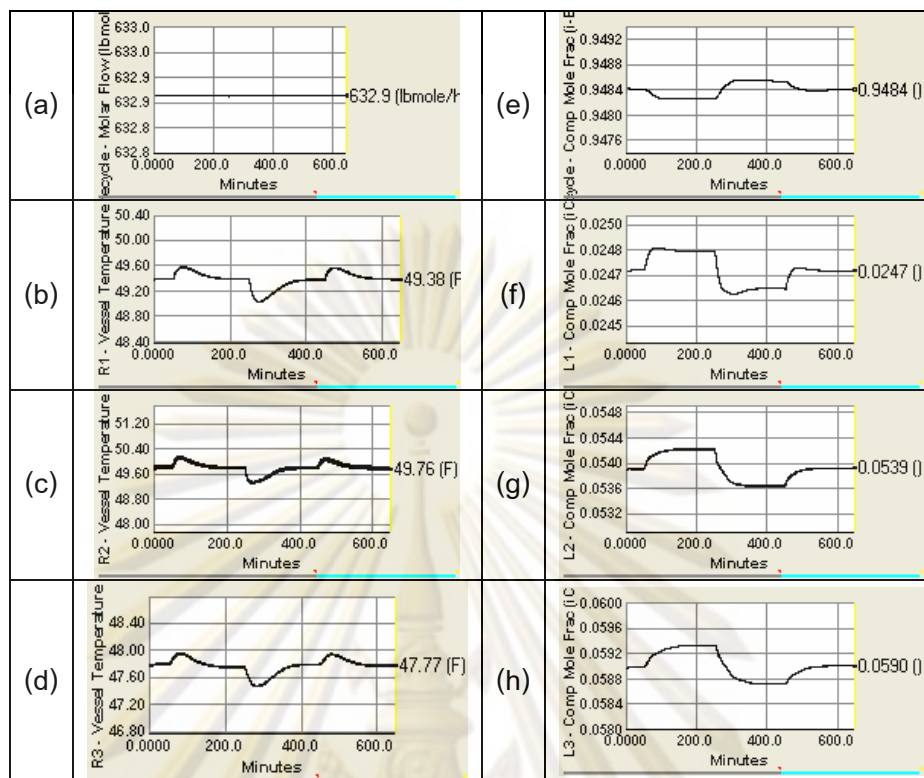**Figure 5.35 (Continued)** Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (l) is a pressure condenser of depropanizer (DP) column, (m) is 25th Tray temperature of DP column, (n) is a condenser duty of DP column and (o) is a reboiler duty of DP column

Figure 5.35 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (q) is a pressure condenser of debutanizer (DB) column, (r) is 3rd Tray temperature of DB column, (s) is 10th Tray temperature of DB column, (t) is a condenser duty of DB column and (u) is a reboiler duty of DB column
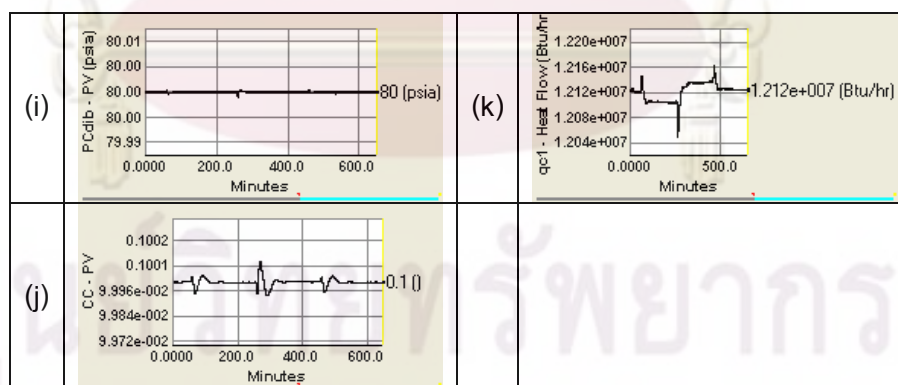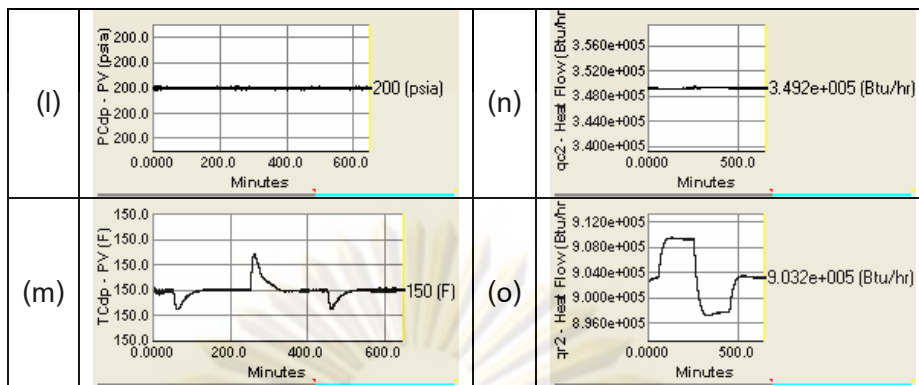
Figure 5.35 (Continued) Dynamic responses of designed control structure VIII (CS8) for alkylation process by step change temperature ±10% in two fresh feeds (BB1 and BB2), where (u) and (v) is a composition of iso-butane and n-butane in distillate stream of DIB column respectively, (w) is a composition of iso-octane in product stream of DIB column, (x) is a composition of propane in distillate stream of DP column, (y) is a composition of iso-butane in product stream of DP column, (z) is a composition of n-butane in distillate stream of DB column and (aa) and (ab) is a composition of iso-octane and dodecane in product stream of DB column respectively

For changing in material disturbances and thermal disturbances of two fresh feeds for base case control structure (CS0) and designed control structure I (CS1) to designed control structure VIII (CS8).

For de-isobutanizer (DIB) column, the pressure condenser of all designed control structures (CS1 to CS8) have a smaller deviation than base case control structure (CS0) because all designed control structures (CS1 to CS8) control the appropriate controlled variable of Tray iso-butane composition of de-isobutanizer (DIB) column is 19th Tray iso-butane composition. The 19th Tray iso-butane composition of DIB column for the designed control structure II (CS2), the designed control structure III (CS3), the designed control structure VI (CS6) and the designed control structure VII (CS7) have a smaller deviation than base case control structure (CS0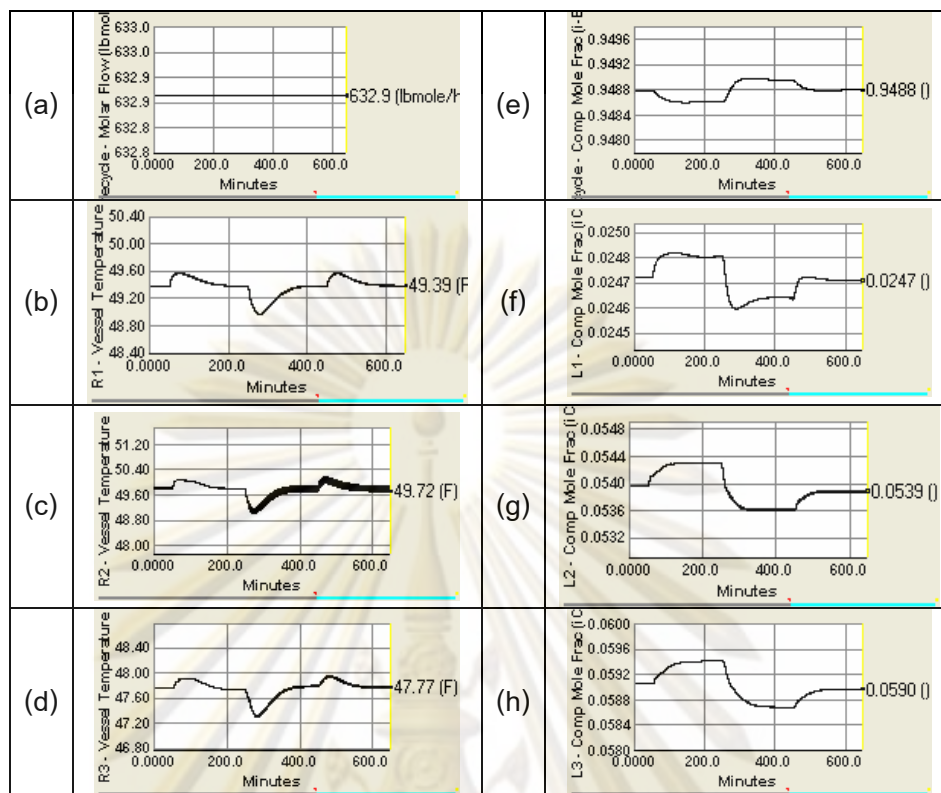) and the other designed control structures (CS1, CS4, CS5 and CS8) because they control the 19th Tray iso-butane composition of de-isobutanizer (DIB) column by manipulating the DIB column reboiler duty (qr1).

For depropanizer (DP) column, the pressure condenser of designed control structure IV (CS4), the designed control structure V (CS5), the designed control structure VI (CS6) and the designed control structure VII (CS7) have a smaller deviation than the other control structures because the DP column condenser level is controlled by manipulating the DP column reflux. Since the distillate stream has a small molar flow rate so control the condenser level of DP column by manipulating the DP column reflux is better.

For debutanizer (DB) column, the pressure condenser of designed control structure II (CS2) and designed control structure III (CS3) have a smaller deviation than the other control structures and The 3rd Tray temperature of DB column for the designed control structure II (CS2), the designed control structure III (CS3), the designed control structure VI (CS6) and the designed control structure VII (CS7) have a smaller deviation than the other control structures because the good effect from the DIB column.

5.4 Evaluation of the Dynamic Performance

The dynamic performance index is focused on time related characteristics of the controller's response to setpoint changes or deterministic disturbances. There exist several candidate performance measures such as settling time and integral absolute error (IAE). Integral absolute error is well known and widely used. For the formulation of a dynamic performance as written below:

$$\text{IAE} = \int |\mathcal{E}(t)|\, dt$$

Note that $\mathcal{E}(t) = y_{sp}(t) - y(t)$ is the deviation (error) of the response from the desired setpoint.

In this research, IAE method is used to evaluate the dynamic performance of the base case control structure and designed control structures. In the process have many types of variables (temperature, pressure, molar flow rate and level) so to compare it we must divide by span (the largest expected change in disturbance) of each variable. The IAE results consider in handle disturbances and maintain product quality. For energy use, the summation value of all energy use is used to evaluate the dynamic performance of the base case control structure and designed control structures.

For changing in material disturbances of two fresh feeds for base case control structure (CS0) and designed control structure I (CS1) to designed control structure VIII (CS8). The IAE results for handle disturbances and maintain product quality and the summation value of all energy use are shown in Table 5.8, Table 5.11 and Table 5.14 respectively.

For changing in thermal disturbances of two fresh feeds for base case control structure (CS0) and designed control structure I (CS1) to designed control structure VIII (CS8). The IAE results for handle disturbances and maintain product quality and the summation value of all energy use are shown in Table 5.9, Table 5.12 and Table 5.15 respectively.

Table 5.8 The IAE Result for handle disturbances to the change in the two fresh feeds flow rate

Controller	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
TC1	0.1009	0.1003	0.1003	0.1004	0.0997	0.1007	0.0996	0.1001	0.0998
TC2	0.1514	0.1507	0.1507	0.1515	0.1540	0.1625	0.1540	0.1560	0.1497
TC3	0.1415	0.1389	0.1388	0.1408	0.1417	0.1443	0.1417	0.1430	0.1390
TC4	0.0010	0.0005	0.0005	0.0007	0.0007	0.0006	0.0007	0.0006	0.0007
PCR	0.0004	0.0002	0.0002	0.0003	0.0002	0.0003	0.0003	0.0002	0.0002
PCdib	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PCdp	0.0000	0.0001	0.0011	0.0010	0.0000	0.0000	0.0000	0.0000	0.0035
PCdb	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0003	0.0002	0.0001
CCdib - Tray 10	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
CCdib - Tray 19	0.0002	0.0001	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0001
TCdp - Tray 25	0.0018	0.0019	0.0018	0.0018	0.0034	0.0033	0.0037	0.0033	0.0015
TCdb - Tray 3	0.0885	0.0678	0.0481	0.0585	0.0721	0.0774	0.0420	0.0483	0.0785
SUM	0.4860	0.4608	0.4416	0.4552	0.4722	0.4895	0.4425	0.4517	0.4733

Table 5.9 The IAE Result for handle disturbances to the change in the two fresh feeds temperature

Controller	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
TC1	0.0158	0.0166	0.0166	0.0151	0.0123	0.0160	0.0122	0.0122	0.0163
TC2	0.0283	0.0283	0.0283	0.0255	0.0103	0.0346	0.0102	0.0103	0.0278
TC3	0.0153	0.0171	0.0170	0.0140	0.0082	0.0158	0.0080	0.0081	0.0170
TC4	0.0006	0.0003	0.0003	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004
PCR	0.0002	0.0001	0.0001	0.0002	0.0000	0.0002	0.0000	0.0000	0.0001
PCdib	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PCdp	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0002
PCdb	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CCdib - Tray 10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CCdib - Tray 19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
TCdp - Tray 25	0.0003	0.0003	0.0003	0.0002	0.0003	0.0005	0.0003	0.0003	0.0002
TCdb - Tray 3	0.0109	0.0091	0.0044	0.0046	0.0075	0.0121	0.0027	0.0028	0.0114
SUM	0.0714	0.0718	0.0672	0.0601	0.0391	0.0798	0.0340	0.0342	0.0735

For all disturbances testing, Figure 5.36 shows the results of Integral Absolute Error (IAE) of all control structure for handle disturbances. Control structure VI (CS6) can handle disturbances the best. Control structure VII (CS7) is the second and control structure II (CS2) is the third. From this result, if the designed control structure controls the 19th Tray iso-butane composition of de-isobutanizer (DIB) column by manipulating the DIB column reboiler duty (qr1) it can handle disturbance well.
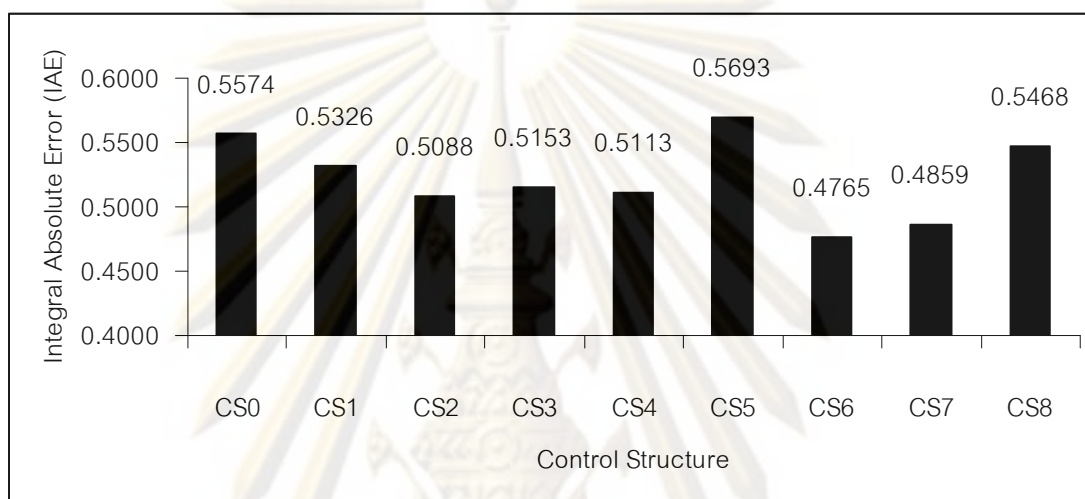
Figure 5.36 The IAE summation for handle disturbances to the change in all disturbances testing

Table 5.10 The IAE Result of product variation to the change in the two fresh feeds flow rate

Product	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
Molar Flow of iC8	0.2185	0.2183	0.2161	0.2169	0.2178	0.2207	0.2139	0.2150	0.2208
Comp Mole Frac of iC8	0.0011	0.0012	0.0012	0.0012	0.0012	0.0012	0.0012	0.0012	0.0012
SUM	0.2195	0.2194	0.2173	0.2180	0.2190	0.2218	0.2150	0.2162	0.2220

Table 5.11 The IAE Result of product variation to the change in the two fresh feeds temperature

Product	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
Molar Flow of iC8	0.0094	0.0129	0.0066	0.0058	0.0078	0.0128	0.0039	0.0039	0.0125
Comp Mole Frac of iC8	0.0000	0.0001	0.0001	0.0001	0.0000	0.0001	0.0000	0.0000	0.0001
SUM	0.0095	0.0130	0.0066	0.0059	0.0078	0.0129	0.0039	0.0039	0.0126

For all disturbances testing, Figure 5.37 shows the results of Integral Absolute Error (IAE) of all control structure for product variation. Control structure VI (CS6) can maintain product quality the best. Control structure VII (CS7) is the second. Control structure II (CS2) is the third. From this result, if the designed control structure controls the 19th Tray iso-butane composition of de-isobutanizer (DIB) column by manipulating the DIB column reboiler duty (qr1) it can maintain product quality well.
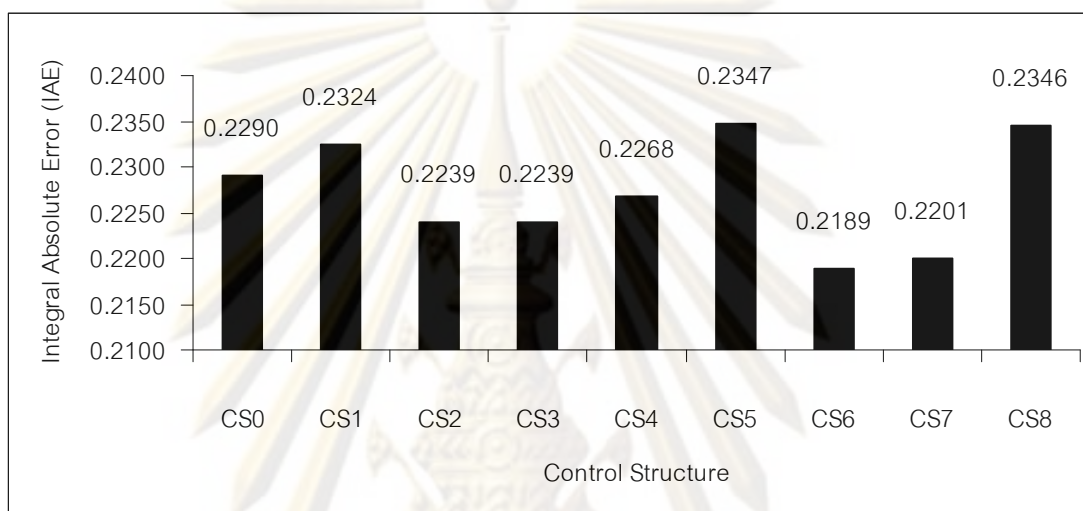
Figure 5.37 The IAE summation of product variation to the change in all disturbances testing

Table 5.12 The summation value of all energy use to the change in the two fresh feeds flow rate

Energy	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
wkcomp - Heat Flow	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05
qcond - Heat Flow	4.17E+05	4.17E+05	4.17E+05	4.17E+05	4.16E+05	4.16E+05	4.16E+05	4.16E+05	4.17E+05
qc1 - Heat Flow	2.42E+06	2.17E+06	2.12E+06	2.32E+06	2.42E+06	2.42E+06	2.42E+06	2.42E+06	2.42E+06
qc2 - Heat Flow	7.01E+04	7.01E+04	7.25E+04	7.11E+04	3.51E+04	3.51E+04	3.51E+04	3.51E+04	7.18E+04
qc3 - Heat Flow	4.51E+04	4.78E+04	4.82E+04	4.67E+04	4.52E+04	4.53E+04	4.52E+04	4.53E+04	4.53E+04
qr1 - Heat Flow	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06
qr2 - Heat Flow	1.81E+05	1.81E+05	1.81E+05	1.81E+05	1.73E+05	1.73E+05	1.73E+05	1.73E+05	1.81E+05
qr3 - Heat Flow	9.82E+04	9.82E+04	9.82E+04	9.84E+04	9.82E+04	9.84E+04	9.82E+04	9.84E+04	9.83E+04
SUM	5.32E+06	5.08E+06	5.04E+06	5.23E+06	5.28E+06	5.28E+06	5.28E+06	5.28E+06	5.32E+06

Table 5.13 The summation value of all energy use to the change in the two fresh feeds temperature

Energy	Integral Absolute Error (IAE)								
	CS0	CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
wkcomp - Heat Flow	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05	1.06E+05
qcond - Heat Flow	4.16E+05	4.17E+05	4.17E+05	4.16E+05	4.15E+05	4.15E+05	4.15E+05	4.15E+05	4.17E+05
qc1 - Heat Flow	2.42E+06	2.17E+06	2.13E+06	2.32E+06	2.43E+06	2.43E+06	2.43E+06	2.43E+06	2.42E+06
qc2 - Heat Flow	6.99E+04	6.99E+04	7.24E+04	7.10E+04	3.50E+04	3.50E+04	3.50E+04	3.50E+04	7.16E+04
qc3 - Heat Flow	4.48E+04	4.75E+04	4.79E+04	4.65E+04	4.49E+04	4.50E+04	4.49E+04	4.51E+04	4.50E+04
qr1 - Heat Flow	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06	1.99E+06
qr2 - Heat Flow	1.81E+05	1.81E+05	1.81E+05	1.81E+05	1.73E+05	1.73E+05	1.73E+05	1.73E+05	1.81E+05
qr3 - Heat Flow	9.79E+04	9.80E+04	9.80E+04	9.82E+04	9.80E+04	9.82E+04	9.80E+04	9.82E+04	9.81E+04
SUM	5.33E+06	5.08E+06	5.04E+06	5.23E+06	5.29E+06	5.29E+06	5.29E+06	5.29E+06	5.33E+06

For all disturbances testing, Figure 5.38 shows the summation value for all energy use of all control structure. Control structure II (CS2) is the most minimize energy use. Control structure I (CS1) is the second. Control structure VI (CS6) is the third.
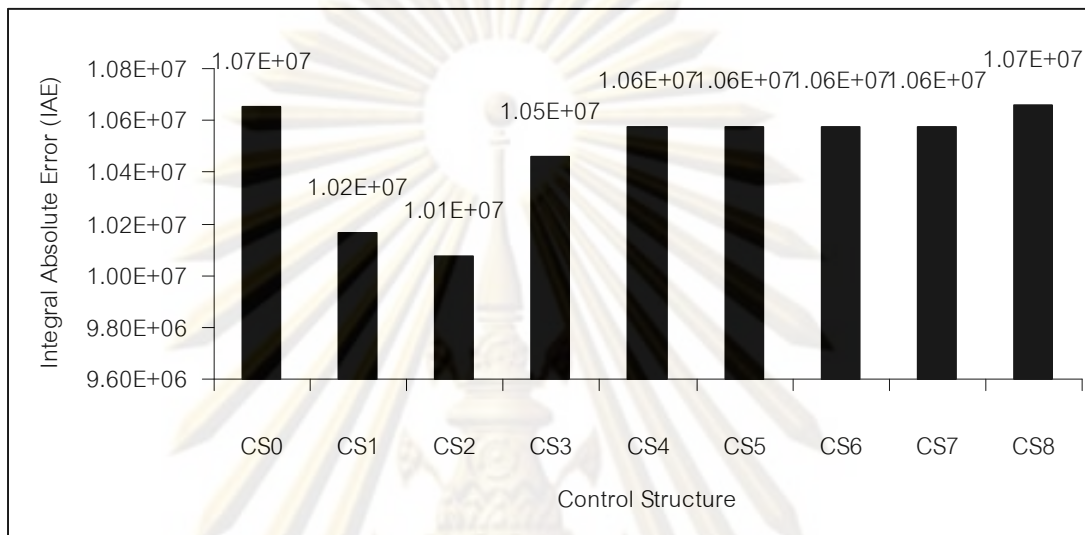
Figure 5.38 The summation value of all energy use to the change in all disturbances testing

5.5 Calculation of utility costs

The calculation of utility costs can be quite complicated, and the "true" cost of such streams is often difficult to estimate in a large facility. For estimating operation costs associated with supplying utilities to different processes, the approach taken here is to assume that the capital investment required to build a facility to supply the utility, for example, a cooling tower, a steam boiler, and so forth, has been already made. This would be the case when a grass-roots cost has been used for the fixed capital investment. The costs associated with supplying a given utility are then obtained by calculating the operating costs to generate the utility. These are the costs that have been presented in Table 5.14. The result of utility costs for each control structure to the change in all disturbances testing is shown in Figure 5.39.

Table 5.14 Utility costs (Analysis, Synthesis, and Design of Chemical Processes, 2007)

Utility	Description	Cost ($/GJ)
Steam from boilers	Process steam: latent heat only	
	a. Low pressure (5 barg, 160 $^{\circ}$C) from HP steam	6.08
	b. Medium pressure (10 barg, 184 $^{\circ}$C) from HP steam	6.87
	c. High pressure (41 barg, 254 $^{\circ}$C) from HP steam	9.83
Electrical substation	Electric Distribution	16.8
	a. 110 V	
	b. 220 V	
	c. 440 V	
Refrigeration	a. Moderately low temperature Refrigerated water	
	in at T = 5 $^{\circ}$C and returned at 15 $^{\circ}$C	4.43
	b. Low temperature Refrigerated water	
	in at T = -20 $^{\circ}$C	7.89
	c. Very low temperature Refrigerated water	
	in at T = -50 $^{\circ}$C	13.11
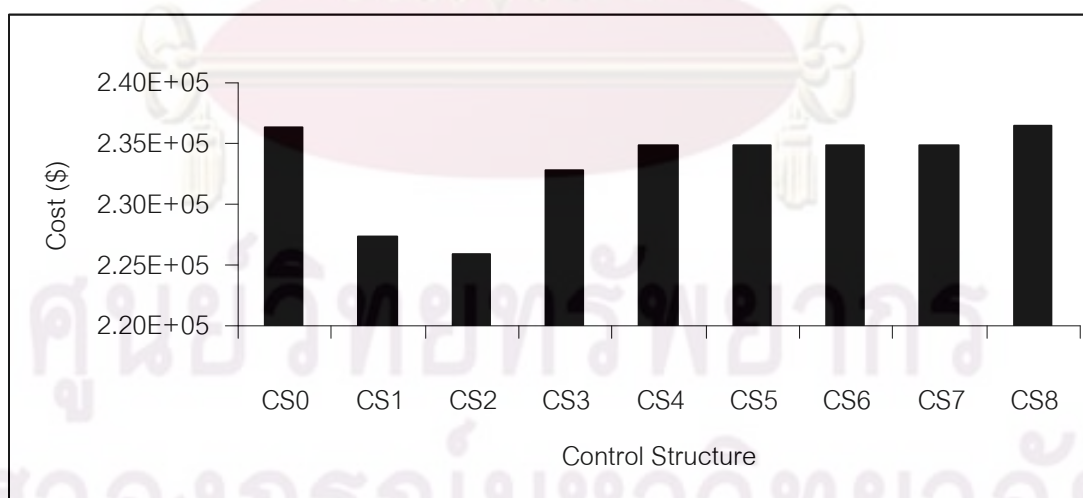
Figure 5.39 Utility costs for each control structure to the change in all disturbances testing

CHAPTER VI

CONCLUTIONS AND RECOMMENDATIONS

6.1 Conclusion

In this research has discussed control structure design for alkylation process, using new design procedure of Wongsri (2009). This procedure based on heuristics and mathematical analysis. The precedence of control variables is established. The purposed plantwide control structure design procedure for selection the best set of control structure is intuitive, simple and straightforward.

The best control structure should handle disturbances entering the process, maintain product quality and minimize energy use. The major disturbances are directed or managed explicitly to achieve the minimal interaction between loops by using the material disturbances and the thermal disturbances.

For the material disturbances, designed control structure II (CS2) can handle disturbances the best, designed control structure VI (CS6) can maintain product quality the best and designed control structure II (CS2) is the most minimize energy use.

For the thermal disturbances, designed control structure VI (CS6) can handle disturbances the best, designed control structure VI (CS6) can maintain product quality the best and designed control structure II (CS2) is the most minimize energy use.

For all disturbances testing, designed control structure VI (CS6) can handle disturbances the best, designed control structure VI (CS6) can maintain product quality the best and designed control structure II (CS2) is the most minimize energy use.

New design procedure of Wongsri (2009) can find the appropriate set of controlled variables to achieve form fixture point theorem. The best control configurations depend on the direction of controlled variable with manipulated variable. Therefore this research establishes that the Wongsri's procedure, which combines heuristics, analytical method and dynamic simulation, a useful design procedure that leads to a good-performance plantwide control system.

6.2 Recommendations

Study and design the control structure of the other process in plantwide control via new design procedure of Wongsri (2009).

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APPENDICES

APPENDIX A

TUNING OF CONTROL STRUCTURES

A.1 Tuning Controllers

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

1. Three tuning constants must be specified.

2. Signal noise is amplified.

3. Several types of PID control algorithms are used, so important to careful that the right algorithm is used with its matching tuning method.

4. The simulation is an approximation of the real plant. If high performance controllers are required to get good dynamics from the simulation, the real plant may not work well.

A.2 Tuning Flow, Level, Pressure and Temperature Loops

Flow Controllers

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

Level Controllers

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady state offset (the level will not be returned to its setpoint value).

However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is $K_C = 2$.

Pressure Controllers

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

Temperature Controllers

Temperature dynamic responses are generally slow, so PID control is used. Typically, the controller gain, Kc, should be set between 2 to 10, the integral time,

A.3 Relay-Feedback Testing

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

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

$$K_U = \frac{4h}{a\pi} \qquad (1)$$

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

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

1. Only one parameter has to be specified (relay height).

2. The time it takes to run the test is short, particularly compared to the extended periods required for methods like PRBS.

3. The test is closedloop, so the process is not driven away from the setpoint.

4. The information obtained is very accurate in the frequency range that is important for the design of a feedback controller.

5. The impact of load changes that occur during the test can be detected by a change to asymmetric pulses in the manipulated variable.

These entire features make relay-feedback testing a useful identification tool.

Knowing the ultimate gain, K_U and the ultimate period, P_U permits us to calculate controller settings. There are several methods that require only these two parameters. The Ziegler-Nichols tuning equations for a PI controller are:

$$K_C = K_U / 2.2 \qquad (2)$$

$$\tau_I = P_U / 1.2 \qquad (3)$$

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

$$K_C = K_U / 3.2 \qquad (4)$$

$$\tau_I = 2.2 P_U \qquad (5)$$

A.4 Inclusion of Lags

Any real physical system has many lags. Measurement and actuator lags always exist. In simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed. Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint. Table A.1 summarizes some recommended lags to include in several different types of control loops.
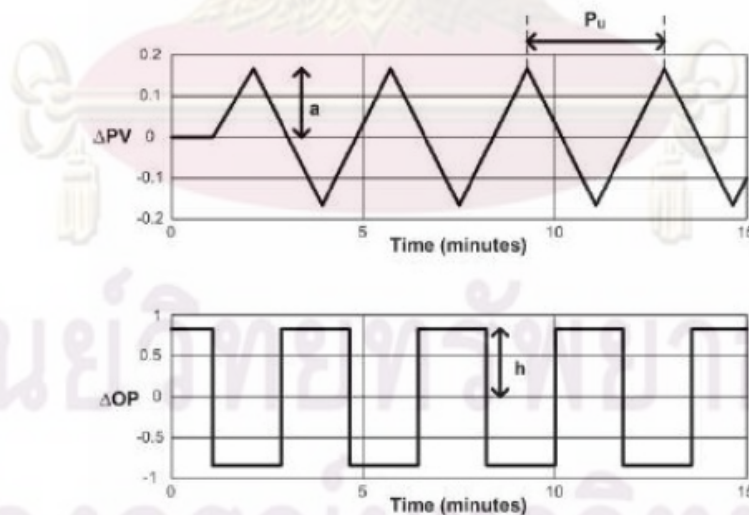
Figure A.1 Input and Output from Relay-Feedback Test

Table A.1 Typical measurement lags

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

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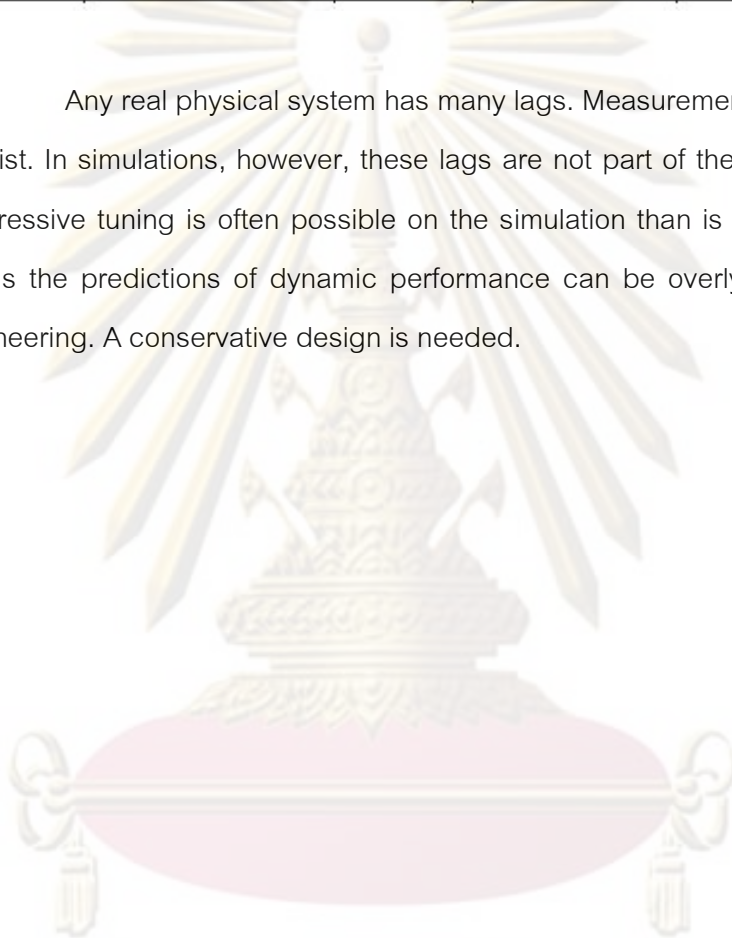
APPENDIX B

PARAMETER TUNING

Table B.1 Tuning parameters for the base case control structure (CS0)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.171	8.39e-3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.693	7.02e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.23	0.274	6.09e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %

Table B.1 (Continued) Tuning parameters for the base case control structure (CS0)

Controller	Controlled Variables	Manipulated variables	K_c	τ_I	τ_D	Action Controller	PV Range
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	28	0.285	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	15.7	0.626	-	Direct	150-250 psia
PCdb	Pressure condenser of DB column	qc3	21.5	0.185	-	Direct	0-100 psia
CC	Tray 10 iso-butane composition of DIB column	V7	10.7	8.06	1.79	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	5	2.93	0.651	Reverse	100-200 $^\circ$F
TCdb	Tray 3 temperature of DB column	qr3	0.93	4.01	0.891	Reverse	200-400 $^\circ$F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	qr1	5	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	V9	2	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	2	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

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

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.171	8.39e-3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.806	5.59e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.89	0.206	4.57e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	28.2	0.165	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	15.9	0.588	-	Direct	150-250 psia

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

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	22.4	0.181	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	V7	10.7	20.8	4.63	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	4.84	2.97	0.659	Reverse	100-200 ^{o}F
TCdb	Tray 3 temperature of DB column	qr3	0.921	4.02	0.893	Reverse	200-400 ^{o}F
LCdib2	Level condenser of DIB column	V4	2.5	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	qr1	5	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	V9	2	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	3	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	3	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

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

Controller	Controlled Variables	Manipulated variables	K_c	τ_I	τ_D	Action Controller	PV Range
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.171	8.39e-3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.805	5.60e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.51	0.193	4.30e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	27.5	0.163	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	19.2	0.536	-	Direct	150-250 psia

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

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	23	0.179	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	qr1	6.74	3.76	0.836	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	5.06	2.90	0.645	Reverse	100-200 $^\circ$F
TCdb	Tray 3 temperature of DB column	qr3	0.796	4.41	0.981	Reverse	200-400 $^\circ$F
LCdib2	Level condenser of DIB column	V4	3	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	V7	2	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	V9	5	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	3	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	5	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

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

Controller	Controlled Variables	Manipulated variables	Tuning Parameter K_c	Tuning Parameter τ_I	Tuning Parameter τ_D	Action Controller	PV Range
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.149	1.54e-2	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.788	5.37e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.66	0.210	4.68e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	29.4	0.159	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	2	3	-	Direct	150-250 psia

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

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	22.3	0.182	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	qr1	6.63	3.80	0.844	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	5.20	2.87	0.637	Reverse	100-200 $^{\circ}$F
TCdb	Tray 3 temperature of DB column	qr3	0.775	4.49	0.999	Reverse	200-400 $^{\circ}$F
TCC	Tray 10 temperature of DB column	reflux	1.29	5.47	1.22	Direct	56.1-256.1 $^{\circ}$F
LCdib2	Level condenser of DIB column	V4	3	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	V7	2	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	V9	5	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	3	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	5	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

Table B.5 Tuning parameters for the designed control structure IV (CS4)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter K_c	τ_I	τ_D	Action Controller	PV Range
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.172	8.39e-3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.787	5.37e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.47	0.281	6.25e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	30.7	0.154	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	9.85	0.538	-	Direct	150-250 psia

Table B.5 (Continued) Tuning parameters for the designed control structure IV (CS4)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	21.4	0.187	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	V7	11.4	19.1	4.25	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	4.84	2.97	0.659	Reverse	100-200 $^\circ$F
TCdb	Tray 3 temperature of DB column	qr3	0.888	4.14	0.920	Reverse	200-400 $^\circ$F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	qr1	5	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	reflux	0.1	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	2	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

Table B.6 Tuning parameters for the designed control structure V (CS5)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter K_c	τ_I	τ_D	Action Controller	PV Range
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.149	1.54e-2	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.788	5.37e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 $^{\circ}$F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 $^{\circ}$F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 $^{\circ}$F
TC4	Tank vessel temperature	qcond	3.60	0.165	3.67e-2	Direct	50-150 $^{\circ}$F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	30.5	0.150	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	9.85	0.538	-	Direct	150-250 psia

Table B.6 (Continued) Tuning parameters for the designed control structure V (CS5)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	21.5	0.187	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	V7	11.4	19.1	4.25	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	3.59	3.81	0.846	Reverse	100-200 $^{\circ}$F
TCdb	Tray 3 temperature of DB column	qr3	0.886	4.14	0.920	Reverse	200-400 $^{\circ}$F
TCC	Tray 10 temperature of DB column	reflux	1.31	5.48	1.22	Direct	56.1-256.1 $^{\circ}$F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	qr1	5	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	reflux	0.1	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	2	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

Table B.7 Tuning parameters for the designed control structure VI (CS6)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.172	8.39e-3	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.643	7.05e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	3.42	0.283	6.28 e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	30.4	0.167	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	9.8	0.54	-	Direct	150-250 psia

Table B.7 (Continued) Tuning parameters for the designed control structure VI (CS6)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	13.0	0.482	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	qr1	6.60	3.81	0.848	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	3.34	4.04	0.899	Reverse	100-200 ^{o}F
TCdb	Tray 3 temperature of DB column	qr3	0.884	4.14	0.921	Reverse	200-400 ^{o}F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	V7	2	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	reflux	0.1	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	2	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

Table B.8 Tuning parameters for the designed control structure VII (CS7)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.149	1.54e-2	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.792	5.37e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 $^{\circ}$F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 $^{\circ}$F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 $^{\circ}$F
TC4	Tank vessel temperature	qcond	3.65	0.211	4.68e-2	Direct	50-150 $^{\circ}$F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	30.5	0.155	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	9.78	0.54	-	Direct	150-250 psia

Table B.8 (Continued) Tuning parameters for the designed control structure VII (CS7)

Controller	Controlled Variables	Manipulated variables	K_c	τ_I	τ_D	Action Controller	PV Range
PCdb	Pressure condenser of DB column	qc3	16.1	0.335	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	qr1	6.60	3.81	0.847	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	3.59	3.81	0.847	Reverse	100-200 ^{o}F
TCdb	Tray 3 temperature of DB column	qr3	0.884	4.14	0.921	Reverse	200-400 ^{o}F
TCC	Tray 10 temperature of DB column	reflux	1.3	5.5	1.22	Direct	56.1-256.1 ^{o}F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	V7	2	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	reflux	0.1	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	2	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

Table B.9 Tuning parameters for the designed control structure VIII (CS8)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
FC1	Fresh feed flowrate	V1	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC2	Fresh feed flowrate	V2	0.171	1.76e-2	-	Reverse	0-50 lbmole/hr
FC3	Recycle flowrate	V3	0.15	1.54e-2	-	Reverse	0-1000 lbmole/hr
PCR	KO vessel pressure	wkcomp	0.789	5.37e-2	-	Direct	0-50 psia
TC1	Reactor 1 vessel temperature	V12	0.5	10	-	Direct	0-100 ^{o}F
TC2	Reactor 2 vessel temperature	V22	0.5	10	-	Direct	0-100 ^{o}F
TC3	Reactor 3 vessel temperature	V32	0.5	-	-	Direct	0-100 ^{o}F
TC4	Tank vessel temperature	qcond	2.91	0.198	4.41e-2	Direct	50-150 ^{o}F
LC1	Reactor 1 liquid percent level	V11	4	-	-	Direct	0-100 %
LC2	Reactor 2 liquid percent level	V21	4	-	-	Direct	0-100 %
LC3	Reactor 3 liquid percent level	V31	4	-	-	Direct	0-100 %
LC4	Tank liquid percent level	V5	2	-	-	Direct	0-100 %
PCdib	Pressure condenser of DIB column	qc1	29.3	0.23	-	Direct	50-150 psia
PCdp	Pressure condenser of DP column	qc2	2	10	-	Direct	150-250 psia

Table B.9 (Continued) Tuning parameters for the designed control structure VIII (CS8)

Controller	Controlled Variables	Manipulated variables	Tuning Parameter			Action Controller	PV Range
			K_c	τ_I	τ_D		
PCdb	Pressure condenser of DB column	qc3	21.4	0.187	-	Direct	0-100 psia
CC	Tray 19 iso-butane composition of DIB column	V7	11.4	19.1	4.25	Reverse	0-0.2
TCdp	Tray 25 temperature of DP column	qr2	5.42	2.65	0.59	Reverse	100-200 ^{o}F
TCdb	Tray 3 temperature of DB column	qr3	0.889	4.14	0.920	Reverse	200-400 ^{o}F
TCC	Tray 10 temperature of DB column	reflux	1.31	5.49	1.22	Direct	56.1-256.1 ^{o}F
LCdib2	Level condenser of DIB column	V4	2	-	-	Reverse	0-100 %
LCdib1	Level reboiler of DIB column	qr1	5	-	-	Direct	0-100 %
LCdp2	Level condenser of DP column	V9	2.5	-	-	Direct	0-100 %
LCdp1	Level reboiler of DP column	V8	1.5	-	-	Direct	0-100 %
LCdb2	Level condenser of DB column	V15	2	-	-	Direct	0-100 %
LCdb1	Level reboiler of DB column	V14	2	-	-	Direct	0-100 %

APPENDIX C

FIXTURE POINT THEOREM DATA

Table C.1 List of Manipulated Variables for Alkylation Process

Manipulated Variable	Description
V1	Fresh feed valve
V2	Fresh feed valve
V3	Iso-butane recycle valve
V12	Reactor1effluence vapor valve
V11	Reactor1effluence liquid valve
V22	Reactor2 effluence vapor valve
V21	Reactor2 effluence liquid valve
V32	Reactor3 effluence vapor valve
V31	Reactor3 effluence liquid valve
wkcomp	Compressor power
qcond	Condenser heat duty
V5	Tank effluence liquid valve
V4	SatC4 fresh feed valve
qc1	DIB column condenser duty
qr1	DIB column reboiler duty
V7	DIB column bottom valve
qc2	DP column condenser duty
qr2	DP column reboiler duty
V9	DP column distillate valve
V8	DP column bottom valve
qc3	DB column condenser duty
qr3	DB column reboiler duty
V15	DB column distillate valve
V14	DB column bottom valve

Table C.2 IAE Results of Flow Rate Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
B1 - Molar Flow	6.3064	0.1229	0.3505	1.0254	0.0536	0.0677
B2 - Molar Flow	26.7745	5.4030	4.4079	7.6075	0.6413	0.1916
B3 - Molar Flow	2.9086	0.4974	0.3204	0.3696	0.0386	0.1361
BB1 - Molar Flow	0.0001	4.4280	0.0000	0.0000	0.0000	0.0000
BB2 - Molar Flow	0.0002	0.0000	4.4280	0.0000	0.0000	0.0000
D1 - Molar Flow	57.5294	5.6016	4.0529	7.0921	0.7764	0.2988
D2 - Molar Flow	3.6012	0.1888	0.1037	0.3200	0.0264	0.0066
D3 - Molar Flow	3.6513	0.2668	0.1465	0.1862	0.0186	0.0732
disch - Molar Flow	47.7052	9.6793	16.7679	15.4728	1.1040	4.7771
dondout - Molar Flow	48.0077	9.6691	16.9086	16.0703	1.1333	5.0421
hxcout - Molar Flow	25.0517	0.7320	3.0295	1.8648	0.3850	1.6432
hxhin - Molar Flow	71.4927	0.7586	0.7106	0.8767	0.1375	0.1970
hxhout - Molar Flow	71.3806	0.7470	0.7134	0.8674	0.1344	0.1996
L1 - Molar Flow	122.4688	4.9019	23.5570	29.8565	5.0996	14.0739
L2 - Molar Flow	254.9521	7.9646	16.4226	18.9064	3.9535	5.0614
L3 - Molar Flow	25.1467	0.7172	3.0330	1.8600	0.3881	1.6405
Lko - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ltk - Molar Flow	51.7359	6.2009	13.4202	13.7975	0.5334	1.7317
p1out - Molar Flow	25.1467	0.7172	3.0330	1.8600	0.3881	1.6405
p2out - Molar Flow	57.5294	5.6016	4.0529	7.0921	0.7764	0.2988
p3out - Molar Flow	51.7359	6.2009	13.4202	13.7975	0.5334	1.7317
SatC4 - Molar Flow	3.7592	0.2757	0.1600	0.4057	0.0176	0.0273
tot1 - Molar Flow	71.3807	4.2468	0.7134	0.8674	0.1344	0.1996
tot2 - Molar Flow	122.4689	4.9019	19.1750	29.8565	5.0996	14.0739
total - Molar Flow	71.4927	0.7586	0.7106	0.8767	0.1375	0.1970

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
v11out - Molar Flow	122.4688	4.9019	23.5570	29.8565	5.0996	14.0739
v12out - Molar Flow	34.1797	7.5722	15.9679	23.3849	1.5021	4.4240
v13out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Molar Flow	2.9086	0.4974	0.3204	0.3696	0.0386	0.1361
v15out - Molar Flow	3.6513	0.2668	0.1465	0.1862	0.0186	0.0732
v1out - Molar Flow	0.0001	4.4280	0.0000	0.0000	0.0000	0.0000
v20out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Molar Flow	254.9521	7.9646	16.4226	18.9064	3.9535	5.0614
v22out - Molar Flow	15.6845	1.9675	2.6355	3.7793	0.3170	9.3969
v2out - Molar Flow	0.0002	0.0000	4.4280	0.0000	0.0000	0.0000
v31out - Molar Flow	25.1467	0.7172	3.0330	1.8600	0.3881	1.6405
v32out - Molar Flow	108.9299	0.1778	0.5466	0.6423	0.0754	0.4158
v3out - Molar Flow	71.3806	0.7470	0.7134	0.8674	0.1344	0.1996
v4out - Molar Flow	3.7592	0.2757	0.1600	0.4057	0.0176	0.0273
v5out - Molar Flow	51.7359	6.2009	13.4202	13.7975	0.5334	1.7317
v7out - Molar Flow	6.3064	0.1229	0.3505	1.0254	0.0536	0.0677
v8out - Molar Flow	26.7745	5.4030	4.4079	7.6075	0.6413	0.1916
v9out - Molar Flow	3.6012	0.1888	0.1037	0.3200	0.0264	0.0066
vap1 - Molar Flow	34.1797	7.5722	15.9679	23.3849	1.5021	4.4240
vap2 - Molar Flow	15.6845	1.9675	2.6355	3.7793	0.3170	9.3969
vap3 - Molar Flow	108.9299	0.1778	0.5466	0.6423	0.0754	0.4158
vaptot - Molar Flow	47.7052	9.6793	16.7679	15.4728	1.1040	4.7771
vtk - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
recycle - Molar Flow	71.3806	0.7470	0.7134	0.8674	0.1344	0.1996

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
B1 - Molar Flow	0.0539	0.0288	4.2986	0.5145	2.1581	0.0645
B2 - Molar Flow	0.1180	0.4753	2.2971	11.6573	23.7012	0.8662
B3 - Molar Flow	0.1464	0.0215	5.5886	1.1612	0.7462	0.0998
BB1 - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Molar Flow	0.2674	0.4536	4.7995	10.7021	21.2337	0.6898
D2 - Molar Flow	0.0052	0.0182	0.0629	0.3711	4.7634	0.0357
D3 - Molar Flow	0.0719	0.0097	1.8782	0.5849	0.5723	0.0528
disch - Molar Flow	0.0978	1.1159	4.5922	8.7515	3.9163	0.5282
dondout - Molar Flow	0.1302	1.0854	4.5967	8.8151	4.3871	0.5681
hxcout - Molar Flow	0.8015	0.4412	26.8977	6.6875	11.9546	0.5263
hxhin - Molar Flow	0.1541	0.0542	2.5024	1.9838	6.0775	0.4239
hxhout - Molar Flow	0.1525	0.0539	2.4865	1.9769	6.0031	0.4217
L1 - Molar Flow	2.4746	2.8475	10.4509	9.6885	13.4993	1.4172
L2 - Molar Flow	6.4141	7.3783	13.0608	245.3811	7.6075	1.7248
L3 - Molar Flow	0.8028	0.4403	26.9145	6.7075	11.9963	0.5282
Lko - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ltk - Molar Flow	0.2522	0.5060	2.2320	11.7580	20.3802	4.6610
p1out - Molar Flow	0.8028	0.4403	26.9145	6.7075	11.9963	0.5282
p2out - Molar Flow	0.2674	0.4536	4.7995	10.7021	21.2337	0.6898
p3out - Molar Flow	0.2522	0.5060	2.2320	11.7580	20.3802	4.6610
SatC4 - Molar Flow	0.0222	0.0204	0.0699	0.6626	1.4658	0.0784
tot1 - Molar Flow	0.1525	0.0539	2.4865	1.9768	6.0031	0.4217
tot2 - Molar Flow	2.4746	2.8475	10.4509	9.6886	13.4993	1.4172
total - Molar Flow	0.1541	0.0542	2.5024	1.9838	6.0775	0.4239

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
v11out - Molar Flow	2.4746	2.8475	10.4509	9.6885	13.4993	1.4172
v12out - Molar Flow	0.1972	0.4572	3.4087	5.4561	13.3758	0.4369
v13out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Molar Flow	0.1464	0.0215	5.5886	1.1612	0.7462	0.0998
v15out - Molar Flow	0.0719	0.0097	1.8782	0.5849	0.5723	0.0528
v1out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v20out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Molar Flow	6.4141	7.3783	13.0608	245.3811	7.6075	1.7248
v22out - Molar Flow	0.0872	0.2618	1.0348	2.0596	2.2701	0.0992
v2out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v31out - Molar Flow	0.8028	0.4403	26.9145	6.7075	11.9963	0.5282
v32out - Molar Flow	0.1837	1.8340	0.1675	1.3681	0.3002	0.0100
v3out - Molar Flow	0.1525	0.0539	2.4865	1.9769	6.0031	0.4217
v4out - Molar Flow	0.0222	0.0204	0.0699	0.6626	1.4658	0.0784
v5out - Molar Flow	0.2522	0.5060	2.2320	11.7580	20.3802	4.6610
v7out - Molar Flow	0.0539	0.0288	4.2986	0.5145	2.1581	0.0645
v8out - Molar Flow	0.1180	0.4753	2.2971	11.6573	23.7012	0.8662
v9out - Molar Flow	0.0052	0.0182	0.0629	0.3711	4.7634	0.0357
vap1 - Molar Flow	0.1972	0.4572	3.4087	5.4561	13.3758	0.4369
vap2 - Molar Flow	0.0872	0.2618	1.0348	2.0596	2.2701	0.0992
vap3 - Molar Flow	0.1837	1.8340	0.1675	1.3681	0.3002	0.0100
vaptot - Molar Flow	0.0978	1.1159	4.5922	8.7515	3.9163	0.5282
vtk - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
recycle - Molar Flow	0.1525	0.0539	2.4865	1.9769	6.0031	0.4217

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
B1 - Molar Flow	0.1891	0.5363	28.7745	7.0480	0.0665	0.6724
B2 - Molar Flow	0.3655	1.4340	43.7524	0.0330	1.1975	15.2081
B3 - Molar Flow	1.3867	0.4276	17.6678	2.5776	0.1122	1.9598
BB1 - Molar Flow	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000
BB2 - Molar Flow	0.0000	0.0000	0.0002	0.0000	0.0000	0.0001
D1 - Molar Flow	0.3650	1.5843	33.9721	0.0921	0.9443	13.3702
D2 - Molar Flow	0.0126	0.0493	3.4760	0.0013	0.0624	0.4766
D3 - Molar Flow	0.6789	0.2317	8.2206	0.9809	0.0592	0.9975
disch - Molar Flow	0.6937	2.4247	101.8350	0.0869	0.8008	8.2932
dondout - Molar Flow	0.6675	2.4471	102.0083	0.0882	0.8270	8.2774
hxcout - Molar Flow	0.6288	3.5132	54.0995	0.1499	0.5121	7.2947
hxhin - Molar Flow	0.7273	3.0033	53.9329	0.1140	0.3497	3.7407
hxhout - Molar Flow	0.7272	3.0114	53.9503	0.1157	0.3500	3.7281
L1 - Molar Flow	0.8719	3.7345	87.8493	1.0939	1.8806	11.2214
L2 - Molar Flow	1.2663	1.9650	149.4929	0.6264	1.1112	337.1695
L3 - Molar Flow	0.6283	3.5201	54.1219	0.1514	0.5134	7.2995
Lko - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ltk - Molar Flow	0.2288	1.4044	105.2072	0.1042	2.1101	11.7169
p1out - Molar Flow	0.6283	3.5201	54.1219	0.1514	0.5134	7.2995
p2out - Molar Flow	0.3650	1.5843	33.9721	0.0921	0.9443	13.3702
p3out - Molar Flow	0.2288	1.4044	105.2072	0.1042	2.1101	11.7169
SatC4 - Molar Flow	6.3471	0.4559	6.0429	0.0154	0.0841	1.0595
tot1 - Molar Flow	0.7272	3.0114	53.9503	0.1157	0.3500	3.7281
tot2 - Molar Flow	0.8719	3.7345	87.8493	1.0939	1.8806	11.2215
total - Molar Flow	0.7273	3.0033	53.9329	0.1140	0.3497	3.7407

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
v11out - Molar Flow	0.8719	3.7345	87.8493	1.0939	1.8806	11.2214
v12out - Molar Flow	0.5288	1.8685	4.2755	0.0626	0.6261	6.5676
v13out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Molar Flow	1.3867	0.4276	17.6678	2.5776	0.1122	1.9598
v15out - Molar Flow	0.6789	0.2317	8.2206	0.9809	0.0592	0.9975
v1out - Molar Flow	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000
v20out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Molar Flow	1.2663	1.9650	149.4929	0.6264	1.1112	337.1695
v22out - Molar Flow	0.1495	0.5059	3.5137	0.0241	0.1616	1.0085
v2out - Molar Flow	0.0000	0.0000	0.0002	0.0000	0.0000	0.0001
v31out - Molar Flow	0.6283	3.5201	54.1219	0.1514	0.5134	7.2995
v32out - Molar Flow	0.0160	0.0602	129.6801	0.0040	0.0185	2.4162
v3out - Molar Flow	0.7272	3.0114	53.9503	0.1157	0.3500	3.7281
v4out - Molar Flow	6.3471	0.4559	6.0429	0.0154	0.0841	1.0595
v5out - Molar Flow	0.2288	1.4044	105.2072	0.1042	2.1101	11.7169
v7out - Molar Flow	0.1891	0.5363	28.7745	7.0480	0.0665	0.6724
v8out - Molar Flow	0.3655	1.4340	43.7524	0.0330	1.1975	15.2081
v9out - Molar Flow	0.0126	0.0493	3.4760	0.0013	0.0624	0.4766
vap1 - Molar Flow	0.5288	1.8685	4.2755	0.0626	0.6261	6.5676
vap2 - Molar Flow	0.1495	0.5059	3.5137	0.0241	0.1616	1.0085
vap3 - Molar Flow	0.0160	0.0602	129.6801	0.0040	0.0185	2.4162
vaptot - Molar Flow	0.6937	2.4247	101.8350	0.0869	0.8008	8.2932
vtk - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
recycle - Molar Flow	0.7272	3.0114	53.9503	0.1157	0.3500	3.7281

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
B1 - Molar Flow	0.0641	1.1253	0.2024	6.8921	1.2319	0.0586
B2 - Molar Flow	1.0216	21.5057	0.0401	0.0242	0.0427	0.0392
B3 - Molar Flow	0.0722	2.1395	0.2168	5.3740	2.0107	3.0945
BB1 - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Molar Flow	0.0000	0.0013	0.0000	0.0000	0.0000	0.0000
D1 - Molar Flow	0.9608	15.7256	0.0304	0.0948	0.0220	0.0329
D2 - Molar Flow	0.7127	0.0947	0.0016	0.0010	0.0017	0.0015
D3 - Molar Flow	0.0376	1.1287	0.1208	1.2565	2.3890	0.0352
disch - Molar Flow	0.6234	8.2564	0.0235	0.0879	0.0301	0.0214
dondout - Molar Flow	0.6470	8.3297	0.0551	0.0869	0.0616	0.0528
hxcout - Molar Flow	0.2493	13.3018	0.0169	0.0915	0.0281	0.0147
hxhin - Molar Flow	0.1219	11.4602	0.0101	0.0841	0.0211	0.0078
hxhout - Molar Flow	0.1221	11.4438	0.0117	0.0828	0.0227	0.0085
L1 - Molar Flow	1.9250	37.3617	0.5825	1.0461	0.5852	0.5860
L2 - Molar Flow	1.3724	332.8224	0.7546	0.9346	0.7400	0.7594
L3 - Molar Flow	0.2515	13.2806	0.0174	0.0902	0.0290	0.0149
Lko - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ltk - Molar Flow	1.8821	8.1930	0.0340	0.0874	0.0326	0.0330
p1out - Molar Flow	0.2515	13.2806	0.0174	0.0902	0.0290	0.0149
p2out - Molar Flow	0.9608	15.7256	0.0304	0.0948	0.0220	0.0329
p3out - Molar Flow	1.8821	8.1930	0.0340	0.0874	0.0326	0.0330
SatC4 - Molar Flow	0.0507	1.1841	0.0033	0.0111	0.0049	0.0028
tot1 - Molar Flow	0.1221	11.4438	0.0117	0.0828	0.0227	0.0085
tot2 - Molar Flow	1.9250	37.3618	0.5825	1.0461	0.5852	0.5860
total - Molar Flow	0.1219	11.4602	0.0101	0.0841	0.0211	0.0078

Table C.2 (Continued) IAE Results of Flow Rate Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
v11out - Molar Flow	1.9250	37.3617	0.5825	1.0461	0.5852	0.5860
v12out - Molar Flow	0.4911	5.9559	0.0222	0.0607	0.0273	0.0205
v13out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Molar Flow	0.0722	2.1395	0.2168	5.3740	2.0107	3.0945
v15out - Molar Flow	0.0376	1.1287	0.1208	1.2565	2.3890	0.0352
v1out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v20out - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Molar Flow	1.3724	332.8224	0.7546	0.9346	0.7400	0.7594
v22out - Molar Flow	0.1252	18.0247	0.0057	0.0261	0.0066	0.0057
v2out - Molar Flow	0.0000	0.0013	0.0000	0.0000	0.0000	0.0000
v31out - Molar Flow	0.2515	13.2806	0.0174	0.0902	0.0290	0.0149
v32out - Molar Flow	0.0135	20.0291	0.0014	0.0033	0.0015	0.0013
v3out - Molar Flow	0.1221	11.4438	0.0117	0.0828	0.0227	0.0085
v4out - Molar Flow	0.0507	1.1841	0.0033	0.0111	0.0049	0.0028
v5out - Molar Flow	1.8821	8.1930	0.0340	0.0874	0.0326	0.0330
v7out - Molar Flow	0.0641	1.1253	0.2024	6.8921	1.2319	0.0586
v8out - Molar Flow	1.0216	21.5057	0.0401	0.0242	0.0427	0.0392
v9out - Molar Flow	0.7127	0.0947	0.0016	0.0010	0.0017	0.0015
vap1 - Molar Flow	0.4911	5.9559	0.0222	0.0607	0.0273	0.0205
vap2 - Molar Flow	0.1252	18.0247	0.0057	0.0261	0.0066	0.0057
vap3 - Molar Flow	0.0135	20.0291	0.0014	0.0033	0.0015	0.0013
vaptot - Molar Flow	0.6234	8.2564	0.0235	0.0879	0.0301	0.0214
vtk - Molar Flow	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
recycle - Molar Flow	0.1221	11.4438	0.0117	0.0828	0.0227	0.0085

Table C.3 IAE Results of Temperature Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
B1 - Temperature	165.5618	9.2092	8.4723	23.7314	1.0797	5.9352
B2 - Temperature	144.2621	21.1159	12.3240	30.7229	2.6937	1.0703
B3 - Temperature	52.8530	5.0227	2.1109	5.7902	0.2598	0.6795
BB1 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Temperature	47.1241	3.1640	1.7656	4.4311	0.2222	0.1409
D2 - Temperature	149.2492	20.8965	15.2122	34.3889	2.5975	0.6807
D3 - Temperature	62.7894	9.7775	5.4024	7.0158	0.6830	2.6882
disch - Temperature	276.8774	11.9707	103.6471	143.2876	0.8168	23.3394
dondout - Temperature	275.1308	34.7005	18.5132	51.7825	4.2610	3.3369
hxcout - Temperature	72.3891	12.0945	3.2215	12.9044	0.8196	1.3761
hxhin - Temperature	51.9750	3.7705	1.4027	3.7375	0.3540	0.1369
hxhout - Temperature	81.6096	16.5774	4.1554	17.3405	1.0086	2.3243
L1 - Temperature	57.5807	17.4624	10.6635	16.2672	2.7983	6.4501
L2 - Temperature	67.6833	18.3214	13.1509	15.7873	0.5073	6.3614
L3 - Temperature	74.2688	18.5560	14.3788	15.4162	0.3263	6.7248
Lko - Temperature	277.1558	20.6882	101.3120	145.0368	2.1598	20.0909
Ltk - Temperature	266.6388	34.7682	17.0897	49.1258	4.2807	3.0291
p1out - Temperature	74.4964	18.5538	14.3543	15.4115	0.3329	6.7045
p2out - Temperature	47.2621	3.1743	1.7671	4.4407	0.2242	0.1444
p3out - Temperature	267.4342	34.8094	17.4549	49.6901	4.2963	3.0434
SatC4 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Temperature	53.9838	16.4105	10.1813	16.3604	2.4460	5.5220
tot2 - Temperature	63.0473	17.6852	12.0804	15.4973	0.4543	7.1226

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
total - Temperature	51.9750	3.7705	1.4027	3.7375	0.3540	0.1369
v11out - Temperature	62.8308	17.7481	11.7927	15.5799	0.4613	7.1155
v12out - Temperature	65.5950	20.4308	110.6548	150.3355	2.9550	5.2390
v13out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Temperature	10.3521	5.7181	3.2768	3.4788	0.4720	0.9127
v15out - Temperature	83.7417	0.0484	0.0223	0.1566	0.0034	0.0184
v1out - Temperature	54.4219	15.7149	9.6946	16.4176	2.3701	5.7000
v20out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Temperature	78.2110	18.4519	14.0657	15.3999	0.3759	6.5335
v22out - Temperature	86.5417	21.3041	12.2669	16.0803	0.4590	46.2254
v2out - Temperature	59.7066	16.0095	10.8169	15.7375	0.6148	6.3645
v31out - Temperature	74.3568	18.5535	14.3607	15.4079	0.3324	6.7060
v32out - Temperature	234.7060	21.6952	13.4639	15.8563	0.3791	5.2473
v3out - Temperature	53.8721	16.1630	10.1961	16.3460	2.4487	5.5148
v4out - Temperature	0.1343	0.0094	0.0054	0.0139	0.0005	0.0008
v5out - Temperature	267.0615	34.7461	17.4155	49.5927	4.2890	3.0323
v7out - Temperature	189.8202	6.3601	8.0924	25.2966	0.9497	5.0977
v8out - Temperature	41.5201	4.2133	1.4560	3.8747	0.4096	0.4205
v9out - Temperature	4.2987	2.0808	0.7974	2.1130	0.2536	0.1294
vap1 - Temperature	57.5780	17.4602	10.6628	16.2630	2.7980	6.4494
vap2 - Temperature	67.6756	18.3228	13.1508	15.7870	0.5074	6.3601
vap3 - Temperature	74.2133	18.5560	14.3788	15.4162	0.3264	6.7246
vaptot - Temperature	277.0800	20.6882	101.3109	145.0333	2.1598	20.0910
vtk - Temperature	264.6895	34.7687	17.0885	49.0570	4.2807	2.9702
recycle - Temperature	81.6096	16.5774	4.1554	17.3405	1.0086	2.3243

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
B1 - Temperature	3.7763	0.5435	113.5445	13.5451	69.5378	2.5028
B2 - Temperature	0.5503	1.8813	10.6473	45.9338	185.2885	3.8764
B3 - Temperature	0.8911	0.0562	24.9358	13.9932	15.6540	0.4015
BB1 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Temperature	0.2218	0.2738	0.5196	7.5955	18.2401	0.9834
D2 - Temperature	0.5379	1.8267	9.5512	45.8742	182.4623	3.6540
D3 - Temperature	2.6410	0.3572	67.9494	21.3793	13.0988	1.9460
disch - Temperature	0.2925	1.2519	2.0867	70.6070	261.4891	1.3693
dondout - Temperature	0.6348	2.6127	17.5192	66.7837	300.8152	5.0940
hxcout - Temperature	0.7254	0.5701	11.4756	18.6087	55.6754	1.9262
hxhin - Temperature	0.1217	0.3179	1.1229	8.4879	25.0605	1.0349
hxhout - Temperature	1.1273	0.6749	13.8930	23.1561	66.5348	2.3008
L1 - Temperature	0.4929	0.5088	6.4767	7.9740	13.8584	0.9571
L2 - Temperature	0.7314	0.8133	7.1631	8.4830	13.7058	0.9795
L3 - Temperature	3.2375	0.8926	8.0043	4.3623	13.7605	1.0282
Lko - Temperature	0.4320	0.2898	7.8898	14.0798	257.8578	1.2968
Ltk - Temperature	0.6873	2.7176	17.3607	68.0196	291.6473	5.1843
p1out - Temperature	3.2478	0.8764	8.1645	3.9654	13.8260	1.0193
p2out - Temperature	0.2156	0.2705	0.5186	7.6213	18.2721	0.9899
p3out - Temperature	0.6861	2.7415	17.1866	68.2487	289.0760	2.8376
SatC4 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Temperature	0.3349	0.3609	6.4839	8.3614	13.6608	1.0079
tot2 - Temperature	0.5780	0.6163	6.7800	7.9626	13.5098	0.9299

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
total - Temperature	0.1217	0.3179	1.1229	8.4879	25.0605	1.0349
v11out - Temperature	0.5858	0.6195	6.8184	7.9614	13.6390	0.9465
v12out - Temperature	0.5162	0.2419	7.6730	7.5529	278.6349	1.2927
v13out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Temperature	1.7328	0.2628	42.4247	15.9487	5.3650	0.7396
v15out - Temperature	0.0109	0.0013	0.3344	0.0808	4.6661	0.0099
v1out - Temperature	0.3286	0.3952	5.5842	7.9203	12.1533	0.8557
v20out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Temperature	2.9483	1.1734	7.8707	4.1103	13.7458	0.9993
v22out - Temperature	0.7361	0.4129	8.3244	8.3564	32.3402	1.3072
v2out - Temperature	0.4223	0.5071	5.9277	8.0178	11.9487	0.8430
v31out - Temperature	3.2474	0.8776	8.2726	3.9587	13.8030	1.0187
v32out - Temperature	2.7132	0.5034	9.0043	42.7237	32.5636	1.2784
v3out - Temperature	0.3361	0.3607	6.5013	8.3637	13.6579	1.0132
v4out - Temperature	0.0008	0.0007	0.0025	0.0225	0.0510	0.0026
v5out - Temperature	0.6851	2.7366	17.1583	68.1195	288.6695	2.8432
v7out - Temperature	2.6889	0.5974	76.8378	11.7317	79.6542	2.6730
v8out - Temperature	0.1234	0.3495	1.3422	9.3722	20.1754	1.1876
v9out - Temperature	0.0629	0.1869	0.9409	4.9005	7.8681	0.4480
vap1 - Temperature	0.4929	0.5088	6.4754	7.9729	13.8544	0.9570
vap2 - Temperature	0.7314	0.8133	7.1628	8.5378	13.7053	0.9795
vap3 - Temperature	3.2387	0.8916	8.0042	4.3514	13.7605	1.0280
vaptot - Temperature	0.4320	0.2898	7.8898	14.0798	257.8480	1.2968
vtk - Temperature	0.6873	2.7176	17.3607	68.0198	291.5963	5.1844
recycle - Temperature	1.1273	0.6749	13.8930	23.1561	66.5348	2.3008

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
B1 - Temperature	32.7830	27.3300	1154.3888	0.4850	2.0752	22.3314
B2 - Temperature	0.9422	5.6858	107.3838	0.1350	4.5391	63.0076
B3 - Temperature	21.3325	5.5329	542.1098	110.8793	0.4159	21.7842
BB1 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Temperature	0.8009	5.3847	65.9352	0.1820	0.9510	12.2799
D2 - Temperature	1.1697	5.5275	118.1368	0.1297	1.4178	61.2131
D3 - Temperature	24.8272	8.5220	1250.2807	35.9240	2.1806	36.3377
disch - Temperature	0.2220	1.0194	131.8125	0.0609	0.5030	10.9780
dondout - Temperature	1.9744	8.8192	157.2440	0.3072	3.1863	15.7295
hxcout - Temperature	1.0161	5.7952	27.6787	0.1184	2.1167	29.3440
hxhin - Temperature	0.5542	4.0979	47.3158	0.1357	0.9531	13.4173
hxhout - Temperature	1.1868	6.2844	15.8798	0.0928	2.6208	37.0316
L1 - Temperature	0.6896	3.1279	9.7547	0.1247	1.0284	22.8417
L2 - Temperature	0.7473	3.1335	7.6527	0.1463	1.0433	23.6966
L3 - Temperature	0.7381	3.1601	13.8957	0.1770	1.0825	12.6406
Lko - Temperature	0.8852	3.7615	120.5385	0.1512	1.2707	11.1306
Ltk - Temperature	1.9467	8.7424	151.2652	0.3195	3.2645	15.4367
p1out - Temperature	0.7509	3.1730	13.7300	0.1741	1.0731	16.6547
p2out - Temperature	0.7951	5.3877	65.9249	0.1864	0.9584	12.3148
p3out - Temperature	1.9576	8.7862	153.4129	0.3170	3.3218	15.3742
SatC4 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Temperature	0.6620	3.2114	11.4193	0.1089	1.0620	23.0491
tot2 - Temperature	0.7289	3.0544	8.0763	0.1469	0.9961	23.0785

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
total - Temperature	0.5542	4.0979	47.3158	0.1357	0.9531	13.4173
v11out - Temperature	0.7107	3.0754	8.1723	0.1442	1.0072	23.1256
v12out - Temperature	0.8673	3.7569	9.9720	0.1484	1.2663	28.2203
v13out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Temperature	21.0600	2.0497	452.8032	73.3305	0.8366	26.0388
v15out - Temperature	0.0851	0.0640	1105.4556	0.0625	0.0119	0.1358
v1out - Temperature	0.8538	3.0790	7.1283	0.1097	1.0601	21.6768
v20out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Temperature	0.7351	3.1403	8.2272	0.1582	1.0586	17.0133
v22out - Temperature	0.9282	3.7780	18.3844	0.1616	1.2839	29.2552
v2out - Temperature	0.8778	3.0210	7.0242	0.1272	1.0304	22.1383
v31out - Temperature	0.7490	3.1663	13.6266	0.1744	1.0725	16.6400
v32out - Temperature	0.9977	3.7402	117.0058	0.2019	1.2450	28.1481
v3out - Temperature	0.6557	3.2197	11.6815	0.1087	1.0618	23.0765
v4out - Temperature	0.0033	0.0155	0.1925	0.0005	0.0028	0.0358
v5out - Temperature	1.9535	8.7707	152.8992	0.3165	3.3160	15.3636
v7out - Temperature	20.6305	28.2765	1183.8187	23.0250	2.0475	19.0696
v8out - Temperature	0.6117	4.1554	56.0817	0.1377	1.0126	15.0994
v9out - Temperature	0.0605	0.4689	4.3765	0.0115	2.7297	6.4655
vap1 - Temperature	0.6894	3.1274	9.7376	0.1246	1.0282	22.8391
vap2 - Temperature	0.7473	3.1333	7.6322	0.1459	1.0432	23.7785
vap3 - Temperature	0.7382	3.1601	13.8594	0.1769	1.0824	12.6229
vaptot - Temperature	0.8852	3.7615	120.5357	0.1512	1.2707	11.1310
vtk - Temperature	1.9467	8.7424	151.2653	0.3195	3.2645	15.4367
recycle - Temperature	1.1868	6.2844	15.8798	0.0928	2.6208	37.0316

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
B1 - Temperature	1.7302	53.4890	0.1207	0.2783	0.1344	0.1370
B2 - Temperature	3.8204	12.6378	0.1599	0.0977	0.1704	0.1540
B3 - Temperature	0.2849	29.6601	3.7318	96.2467	40.7424	3.3133
BB1 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Temperature	0.5629	13.1137	0.0352	0.1432	0.0534	0.0278
D2 - Temperature	1.0249	12.2835	0.1604	0.0950	0.1706	0.1562
D3 - Temperature	1.3846	41.0664	4.4773	88.3860	37.2037	1.3062
disch - Temperature	0.4032	57.6979	0.0757	0.0602	0.0774	0.0743
dondout - Temperature	2.4640	24.9419	0.1891	0.2679	0.2108	0.1798
hxcout - Temperature	1.5992	13.4378	0.0771	0.1206	0.0934	0.0698
hxhin - Temperature	0.6303	11.8220	0.0405	0.0940	0.0535	0.0343
hxhout - Temperature	2.0352	14.8864	0.0905	0.1318	0.1073	0.0831
L1 - Temperature	0.8020	8.7760	0.0437	0.1156	0.0525	0.0395
L2 - Temperature	0.8006	10.6702	0.0434	0.1460	0.0512	0.0403
L3 - Temperature	0.8298	10.9292	0.0914	0.1504	0.0990	0.0878
Lko - Temperature	0.9879	60.0093	0.0581	0.1391	0.0681	0.0535
Ltk - Temperature	2.5550	22.5071	0.2649	0.2800	0.2868	0.2561
p1out - Temperature	0.8184	7.7770	0.0779	0.1448	0.0855	0.0742
p2out - Temperature	0.5698	13.1409	0.0395	0.1371	0.0575	0.0317
p3out - Temperature	2.6034	22.5062	0.2593	0.2779	0.2810	0.2503
SatC4 - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Temperature	0.8173	8.6446	0.0454	0.1112	0.0543	0.0410
tot2 - Temperature	0.7693	9.9652	0.0364	0.1464	0.0442	0.0337

Table C.3 (Continued) IAE Results of Temperature Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
total - Temperature	0.6303	11.8220	0.0405	0.0940	0.0535	0.0343
v11out - Temperature	0.7793	9.8017	0.0415	0.1396	0.0497	0.0381
v12out - Temperature	0.9891	10.1670	0.0611	0.1340	0.0713	0.0564
v13out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Temperature	0.5609	27.0321	0.4976	21.6752	0.7235	0.6974
v15out - Temperature	0.0080	0.1523	0.0062	27.5077	0.6472	0.0023
v1out - Temperature	0.8301	9.0676	0.0490	0.1028	0.0572	0.0462
v20out - Temperature	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Temperature	0.8007	7.7155	0.0575	0.1370	0.0651	0.0538
v22out - Temperature	0.9891	23.1994	0.0533	0.1572	0.0630	0.0492
v2out - Temperature	0.7987	10.3393	0.0380	0.1280	0.0459	0.0356
v31out - Temperature	0.8183	7.7876	0.0785	0.1448	0.0861	0.0748
v32out - Temperature	0.9462	90.5709	0.0476	0.1903	0.0540	0.0466
v3out - Temperature	0.8163	8.6366	0.0458	0.1116	0.0547	0.0413
v4out - Temperature	0.0017	0.0400	0.0000	0.0004	0.0000	0.0000
v5out - Temperature	2.5988	22.4681	0.2589	0.2774	0.2807	0.2500
v7out - Temperature	1.4853	50.7231	2.6647	39.9557	23.5416	0.6460
v8out - Temperature	0.6685	11.7557	0.0468	0.0904	0.0601	0.0407
v9out - Temperature	2.4980	0.9167	0.0121	0.0062	0.0134	0.0118
vap1 - Temperature	0.8018	8.7755	0.0437	0.1155	0.0525	0.0395
vap2 - Temperature	0.8005	10.7261	0.0432	0.1459	0.0511	0.0401
vap3 - Temperature	0.8296	10.9405	0.0912	0.1502	0.0988	0.0876
vaptot - Temperature	0.9879	59.9962	0.0581	0.1391	0.0681	0.0535
vtk - Temperature	2.5550	22.5073	0.2650	0.2799	0.2868	0.2561
recycle - Temperature	2.0352	14.8864	0.0905	0.1318	0.1073	0.0831

Table C.4 IAE Results of Pressure Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
B1 - Pressure	44.1917	3.0822	1.7992	4.5838	0.1981	0.3085
B2 - Pressure	83.0725	41.8549	28.0944	46.9199	5.1004	1.3001
B3 - Pressure	30.0874	7.3365	3.9327	5.0275	0.5010	1.9528
BB1 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Pressure	44.1765	3.0748	1.8009	4.5793	0.1982	0.3064
D2 - Pressure	259.7187	41.7309	30.8969	68.0333	5.0846	1.2963
D3 - Pressure	30.6448	7.4121	4.0109	5.1687	0.5108	2.0118
disch - Pressure	196.1463	59.8642	16.8050	48.6769	5.4051	4.5000
dondout - Pressure	209.2819	61.0618	17.8330	51.7522	5.4960	4.7141
hxcout - Pressure	44.1664	3.0745	1.8013	4.5801	0.1981	0.3062
hxhin - Pressure	52.3041	4.1839	1.7688	4.7483	0.3213	0.2942
hxhout - Pressure	46.3789	1.6633	0.6866	1.6505	0.1241	0.3346
L1 - Pressure	23.5481	7.4908	4.5908	7.5671	1.1054	2.6542
L2 - Pressure	24.8575	7.4155	4.9775	7.0466	0.2812	2.8789
L3 - Pressure	29.1034	7.2291	5.2148	6.6192	0.2015	2.8614
Lko - Pressure	40.1397	13.9939	3.5774	8.0689	1.3506	1.0421
Ltk - Pressure	209.2819	61.0618	17.8330	51.7522	5.4960	4.7141
p1out - Pressure	37.5635	4.6088	1.7334	5.1497	0.1634	1.2162
p2out - Pressure	52.3041	4.1839	1.7688	4.7483	0.3213	0.2942
p3out - Pressure	230.0393	47.7024	21.7579	59.9466	5.1183	2.2715
SatC4 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Pressure	23.5481	7.4908	4.5908	7.5671	1.1054	2.6542
tot2 - Pressure	24.8575	7.4155	4.9775	7.0466	0.2812	2.8789

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V3	V1	V2	V12	V11	V22
total - Pressure	52.3041	4.1839	1.7688	4.7483	0.3213	0.2942
v11out - Pressure	24.8575	7.4155	4.9775	7.0466	0.2812	2.8789
v12out - Pressure	40.1397	13.9939	3.5774	8.0689	1.3506	1.0421
v13out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v15out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	23.5481	7.4908	4.5908	7.5671	1.1054	2.6542
v20out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Pressure	29.1034	7.2291	5.2148	6.6192	0.2015	2.8614
v22out - Pressure	40.1397	13.9939	3.5774	8.0689	1.3506	1.0421
v2out - Pressure	24.8575	7.4155	4.9775	7.0466	0.2812	2.8789
v31out - Pressure	43.0156	3.3366	1.6055	4.6773	0.1910	0.4609
v32out - Pressure	40.1397	13.9939	3.5774	8.0689	1.3506	1.0421
v3out - Pressure	23.5481	7.4908	4.5908	7.5671	1.1054	2.6542
v4out - Pressure	44.1605	3.0754	1.8003	4.5797	0.1980	0.3066
v5out - Pressure	260.4668	41.8182	30.9616	68.1725	5.0964	1.3002
v7out - Pressure	30.5591	7.4116	4.0103	5.1695	0.5107	2.0115
v8out - Pressure	52.3041	4.1839	1.7688	4.7483	0.3213	0.2942
v9out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
vap1 - Pressure	23.5481	7.4908	4.5908	7.5671	1.1054	2.6542
vap2 - Pressure	24.8575	7.4155	4.9775	7.0466	0.2812	2.8789
vap3 - Pressure	29.1034	7.2291	5.2148	6.6192	0.2015	2.8614
vaptot - Pressure	40.1397	13.9939	3.5774	8.0689	1.3506	1.0421
vtk - Pressure	209.2819	61.0618	17.8330	51.7522	5.4960	4.7141
recycle - Pressure	46.3789	1.6633	0.6866	1.6505	0.1241	0.3346

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
B1 - Pressure	0.2490	0.2284	0.8150	7.3501	16.7702	0.8800
B2 - Pressure	1.0238	3.6084	18.6639	92.6037	136.1149	6.9725
B3 - Pressure	1.9508	0.2640	54.2651	16.3623	10.3338	1.4203
BB1 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Pressure	0.2468	0.2282	0.7707	7.3491	16.7818	0.8809
D2 - Pressure	1.0197	3.5998	18.6069	92.3340	323.4194	6.9557
D3 - Pressure	1.9826	0.2670	54.3968	16.4566	10.6652	1.4548
disch - Pressure	0.8193	3.5430	20.5072	112.8263	287.7420	6.3918
dondout - Pressure	0.8464	3.5525	20.7840	115.8095	307.1253	6.6184
hxcout - Pressure	0.2474	0.2283	0.7729	7.3489	16.7837	0.8806
hxhin - Pressure	0.1864	0.3109	0.5083	9.4401	17.9743	0.9877
hxhout - Pressure	0.2682	0.1295	0.9796	4.0377	1.6970	0.6094
L1 - Pressure	0.1510	0.1826	2.5999	3.7541	5.6209	0.3976
L2 - Pressure	0.1906	0.2291	2.6807	3.6923	5.3675	0.3825
L3 - Pressure	1.4106	0.3831	2.9039	1.9310	5.2155	0.3842
Lko - Pressure	0.1787	0.9666	5.2300	5.8227	49.7293	1.1700
Ltk - Pressure	0.8464	3.5525	20.7840	115.8095	307.1253	6.6184
p1out - Pressure	0.6527	0.1766	20.3402	4.5315	12.0314	0.6722
p2out - Pressure	0.1864	0.3109	0.5083	9.4401	17.9743	0.9877
p3out - Pressure	0.9399	3.5206	19.7844	99.1819	327.6183	37.0554
SatC4 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Pressure	0.1510	0.1826	2.5999	3.7541	5.6209	0.3976
tot2 - Pressure	0.1906	0.2291	2.6807	3.6923	5.3675	0.3825

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V21	V32	V31	wkcomp	qcond	V5
total - Pressure	0.1864	0.3109	0.5083	9.4401	17.9743	0.9877
v11out - Pressure	0.1906	0.2291	2.6807	3.6923	5.3675	0.3825
v12out - Pressure	0.1787	0.9666	5.2300	5.8227	49.7293	1.1700
v13out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v15out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.1510	0.1826	2.5999	3.7541	5.6209	0.3976
v20out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Pressure	1.4106	0.3831	2.9039	1.9310	5.2155	0.3842
v22out - Pressure	0.1787	0.9666	5.2300	5.8227	49.7293	1.1700
v2out - Pressure	0.1906	0.2291	2.6807	3.6923	5.3675	0.3825
v31out - Pressure	0.3162	0.2106	2.7685	6.8687	15.9781	0.8452
v32out - Pressure	0.1787	0.9666	5.2300	5.8227	49.7293	1.1700
v3out - Pressure	0.1510	0.1826	2.5999	3.7541	5.6209	0.3976
v4out - Pressure	0.2477	0.2283	0.7828	7.3468	16.7760	0.8802
v5out - Pressure	1.0225	3.6066	18.6483	92.5262	324.0352	6.9703
v7out - Pressure	1.9823	0.2670	54.3958	16.4560	10.6393	1.4546
v8out - Pressure	0.1864	0.3109	0.5083	9.4401	17.9743	0.9877
v9out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
vap1 - Pressure	0.1510	0.1826	2.5999	3.7541	5.6209	0.3976
vap2 - Pressure	0.1906	0.2291	2.6807	3.6923	5.3675	0.3825
vap3 - Pressure	1.4106	0.3831	2.9039	1.9310	5.2155	0.3842
vaptot - Pressure	0.1787	0.9666	5.2300	5.8227	49.7293	1.1700
vtk - Pressure	0.8464	3.5525	20.7840	115.8095	307.1253	6.6184
recycle - Pressure	0.2682	0.1295	0.9796	4.0377	1.6970	0.6094

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
B1 - Pressure	1.0709	5.1357	62.9282	0.1736	0.9444	11.6989
B2 - Pressure	2.4423	10.9978	74.2394	0.2596	9.0382	124.6451
B3 - Pressure	18.9892	6.0196	102.2080	27.5074	1.5913	28.2668
BB1 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Pressure	1.0604	5.1361	62.8130	0.1741	0.9450	11.6933
D2 - Pressure	2.4368	10.9665	212.8858	0.2581	9.0307	124.2002
D3 - Pressure	19.0989	6.3411	102.0757	27.6280	1.6306	28.3781
disch - Pressure	3.1308	11.2083	61.0313	0.4133	4.3067	131.5269
dondout - Pressure	3.1759	11.3805	64.6468	0.4220	4.3961	135.8313
hxcout - Pressure	1.0614	5.1359	62.8520	0.1738	0.9447	11.6941
hxhin - Pressure	0.9713	4.6588	58.4546	0.1494	1.1191	14.2314
hxhout - Pressure	0.8581	4.1275	69.5960	0.1436	0.6060	7.1718
L1 - Pressure	0.4033	1.4359	3.3464	0.0507	0.4932	10.4217
L2 - Pressure	0.4020	1.3697	3.1055	0.0580	0.4670	10.3487
L3 - Pressure	0.3783	1.3332	4.4749	0.0632	0.4635	5.2340
Lko - Pressure	0.7970	2.8167	10.7359	0.1021	1.0187	22.6936
Ltk - Pressure	3.1759	11.3805	64.6468	0.4220	4.3961	135.8313
p1out - Pressure	0.7826	3.5956	39.8251	0.0934	0.7406	8.9948
p2out - Pressure	0.9713	4.6588	58.4546	0.1494	1.1191	14.2314
p3out - Pressure	2.6339	10.8892	119.5078	0.2707	7.3316	128.4805
SatC4 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Pressure	0.4033	1.4359	3.3464	0.0507	0.4932	10.4217
tot2 - Pressure	0.4020	1.3697	3.1055	0.0580	0.4670	10.3487

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V4	qc1	qr1	V7	qc2	qr2
total - Pressure	0.9713	4.6588	58.4546	0.1494	1.1191	14.2314
v11out - Pressure	0.4020	1.3697	3.1055	0.0580	0.4670	10.3487
v12out - Pressure	0.7970	2.8167	10.7359	0.1021	1.0187	22.6936
v13out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v15out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.4033	1.4359	3.3464	0.0507	0.4932	10.4217
v20out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Pressure	0.3783	1.3332	4.4749	0.0632	0.4635	5.2340
v22out - Pressure	0.7970	2.8167	10.7359	0.1021	1.0187	22.6936
v2out - Pressure	0.4020	1.3697	3.1055	0.0580	0.4670	10.3487
v31out - Pressure	1.0141	4.8745	58.9423	0.1601	0.9101	11.2359
v32out - Pressure	0.7970	2.8167	10.7359	0.1021	1.0187	22.6936
v3out - Pressure	0.4033	1.4359	3.3464	0.0507	0.4932	10.4217
v4out - Pressure	1.0619	5.1351	62.8695	0.1737	0.9443	11.6919
v5out - Pressure	2.4413	10.9901	213.5132	0.2591	9.0329	124.4569
v7out - Pressure	19.0978	6.3372	102.2296	27.6289	1.6305	28.3775
v8out - Pressure	0.9713	4.6588	58.4546	0.1494	1.1191	14.2314
v9out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
vap1 - Pressure	0.4033	1.4359	3.3464	0.0507	0.4932	10.4217
vap2 - Pressure	0.4020	1.3697	3.1055	0.0580	0.4670	10.3487
vap3 - Pressure	0.3783	1.3332	4.4749	0.0632	0.4635	5.2340
vaptot - Pressure	0.7970	2.8167	10.7359	0.1021	1.0187	22.6936
vtk - Pressure	3.1759	11.3805	64.6468	0.4220	4.3961	135.8313
recycle - Pressure	0.8581	4.1275	69.5960	0.1436	0.6060	7.1718

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
B1 - Pressure	0.5703	13.0954	0.0377	0.1240	0.0550	0.0316
B2 - Pressure	7.6719	24.3626	0.3121	0.1896	0.3320	0.3037
B3 - Pressure	1.0136	32.1474	3.2679	49.7021	26.8196	0.9890
BB1 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
BB2 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
D1 - Pressure	0.5709	13.0956	0.0385	0.1231	0.0557	0.0324
D2 - Pressure	7.6700	24.2750	0.3098	0.1881	0.3297	0.3015
D3 - Pressure	1.0363	32.2672	3.3112	49.0372	27.3674	0.9704
disch - Pressure	3.3957	29.5525	0.3195	0.3488	0.3458	0.3096
dondout - Pressure	3.4688	30.3219	0.3368	0.3553	0.3634	0.3269
hxcout - Pressure	0.5705	13.0965	0.0381	0.1234	0.0554	0.0320
hxhin - Pressure	0.7587	14.7689	0.0438	0.1012	0.0589	0.0385
hxhout - Pressure	0.2971	14.0286	0.0250	0.1082	0.0393	0.0200
L1 - Pressure	0.3858	4.2548	0.0199	0.0484	0.0239	0.0185
L2 - Pressure	0.3619	4.6356	0.0163	0.0592	0.0198	0.0152
L3 - Pressure	0.3553	4.3170	0.0272	0.0535	0.0305	0.0260
Lko - Pressure	0.7991	7.4964	0.0564	0.0904	0.0635	0.0539
Ltk - Pressure	3.4688	30.3219	0.3368	0.3553	0.3634	0.3269
p1out - Pressure	0.4742	6.9194	0.0325	0.0919	0.0442	0.0284
p2out - Pressure	0.7587	14.7689	0.0438	0.1012	0.0589	0.0385
p3out - Pressure	6.1359	24.1943	0.3108	0.2253	0.3325	0.3022
SatC4 - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
tot1 - Pressure	0.3858	4.2548	0.0199	0.0484	0.0239	0.0185
tot2 - Pressure	0.3619	4.6356	0.0163	0.0592	0.0198	0.0152

Table C.4 (Continued) IAE Results of Pressure Deviation for the Process Stream

Stream	V9	V8	qc3	qr3	V15	V14
total - Pressure	0.7587	14.7689	0.0438	0.1012	0.0589	0.0385
v11out - Pressure	0.3619	4.6356	0.0163	0.0592	0.0198	0.0152
v12out - Pressure	0.7991	7.4964	0.0564	0.0904	0.0635	0.0539
v13out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v14out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v15out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v1out - Pressure	0.3858	4.2548	0.0199	0.0484	0.0239	0.0185
v20out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
v21out - Pressure	0.3553	4.3170	0.0272	0.0535	0.0305	0.0260
v22out - Pressure	0.7991	7.4964	0.0564	0.0904	0.0635	0.0539
v2out - Pressure	0.3619	4.6356	0.0163	0.0592	0.0198	0.0152
v31out - Pressure	0.5541	12.0468	0.0372	0.1181	0.0535	0.0314
v32out - Pressure	0.7991	7.4964	0.0564	0.0904	0.0635	0.0539
v3out - Pressure	0.3858	4.2548	0.0199	0.0484	0.0239	0.0185
v4out - Pressure	0.5702	13.0945	0.0380	0.1236	0.0552	0.0319
v5out - Pressure	7.6677	24.3391	0.3113	0.1891	0.3312	0.3029
v7out - Pressure	1.0362	32.2663	3.3085	49.0970	27.3184	0.9703
v8out - Pressure	0.7587	14.7689	0.0438	0.1012	0.0589	0.0385
v9out - Pressure	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
vap1 - Pressure	0.3858	4.2548	0.0199	0.0484	0.0239	0.0185
vap2 - Pressure	0.3619	4.6356	0.0163	0.0592	0.0198	0.0152
vap3 - Pressure	0.3553	4.3170	0.0272	0.0535	0.0305	0.0260
vaptot - Pressure	0.7991	7.4964	0.0564	0.0904	0.0635	0.0539
vtk - Pressure	3.4688	30.3219	0.3368	0.3553	0.3634	0.3269
recycle - Pressure	0.2971	14.0286	0.0250	0.1082	0.0393	0.0200

Table C.5 IAE Result of Level Deviation for the Process Unit

Unit	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
R1 - Level	747.721	356.275	384.558	394.944	224.175	506.747	134.229	122.265	118.475	429.809	350.385	62.120
R2 - Level	750.707	201.474	402.112	204.695	83.453	402.396	156.956	164.897	108.945	26.433	75.242	6.787
R3 - Level	479.013	330.404	614.412	875.024	164.089	63.776	290.401	283.404	769.298	302.033	603.651	52.559
Tank - Level	856.284	594.552	950.738	1044.030	95.672	735.878	23.047	133.916	374.886	678.743	930.513	927.141
DIB-Level-Condenser	36.311	3.460	1.847	3.398	0.390	0.082	0.032	0.284	1.866	7.178	9.876	0.581
DIB-Level -Reboiler	330.116	63.014	46.901	34.371	3.065	18.487	5.894	2.397	336.324	43.666	78.943	3.892
DP-Level -Condenser	41.826	8.700	8.221	16.592	1.058	0.469	0.179	0.722	4.200	18.889	37.055	1.155
DP-Level -Reboiler	615.565	62.350	420.769	514.730	16.174	5.628	6.673	16.905	51.517	291.370	663.625	300.841
DB-Level -Condenser	332.160	2.547	1.320	3.290	0.281	0.659	0.726	0.118	20.688	5.414	50.010	0.672
DB-Level -Reboiler	59.865	55.846	12.232	28.491	7.362	17.300	18.617	3.045	590.867	96.515	5.891	10.737

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Table C.5 (Continued) IAE Result of Level Deviation for the Process Unit

Unit	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
R1 - Level	8.146	102.335	753.388	14.017	56.936	501.187	36.849	379.884	1.447	3.502	0.002	1.313
R2 - Level	12.109	26.808	502.020	10.542	3.981	21.859	1.534	223.819	0.159	4.513	0.000	0.120
R3 - Level	26.734	244.456	470.388	8.334	43.529	383.440	28.453	404.724	0.713	5.507	0.002	0.534
Tank - Level	28.898	222.848	806.254	3.255	353.659	1023.233	206.912	154.205	2.750	1.555	0.005	2.627
DIB-Level-Condenser	0.180	69.051	14.043	0.025	0.716	9.828	0.600	9.036	0.019	0.014	0.023	0.019
DIB-Level -Reboiler	66.554	31.509	814.113	131.101	5.612	78.397	4.293	89.302	6.327	126.644	21.004	1.367
DP-Level -Condenser	0.600	2.389	25.199	0.068	106.596	25.476	12.111	7.238	0.067	0.051	0.069	0.064
DP-Level -Reboiler	3.877	38.791	586.956	2.052	100.302	704.940	51.673	894.399	1.089	1.418	1.098	1.080
DB-Level -Condenser	6.563	3.946	186.610	11.738	0.687	10.660	0.301	16.844	124.412	198.837	81.438	0.403
DB-Level -Reboiler	186.134	65.800	615.424	604.151	12.278	210.019	5.425	480.330	69.370	716.173	213.630	692.150

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Table C.6 IAE Result of Temperature Deviation for the Process Unit

Unit	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
R1 - Temperature	57.5807	17.4623	10.6635	16.2672	2.7983	6.4501	0.4929	0.5088	6.4767	7.9740	13.8584	0.9571
R2 - Temperature	67.6832	18.3214	13.1509	15.7873	0.5073	6.3614	0.7314	0.8133	7.1631	8.4827	13.7058	0.9795
R3 - Temperature	74.2688	18.5560	14.3788	15.4162	0.3263	6.7248	3.2375	0.8926	8.0043	4.3623	13.7605	1.0282
Tank - Temperature	266.6373	34.7682	17.0897	49.1256	4.2807	3.0290	0.6873	2.7176	17.3607	68.0196	291.6473	5.1844
KO - Temperature	277.1584	20.6882	101.3121	145.0362	2.1598	20.0903	0.4320	0.2898	7.8898	14.0797	257.8633	1.2968

Table C.6 (Continued) IAE Result of Temperature Deviation for the Process Unit

Unit	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
R1 - Temperature	0.6896	3.1279	9.7549	0.1247	1.0284	22.8417	0.8020	8.7760	0.0437	0.1156	0.0525	0.0395
R2 - Temperature	0.7473	3.1335	7.6527	0.1463	1.0433	23.6963	0.8006	10.6700	0.0434	0.1460	0.0512	0.0403
R3 - Temperature	0.7381	3.1601	13.8957	0.1770	1.0825	12.6406	0.8298	10.9292	0.0914	0.1504	0.0990	0.0878
Tank - Temperature	1.9467	8.7424	151.2652	0.3195	3.2645	15.4367	2.5550	22.5071	0.2649	0.2800	0.2868	0.2561
KO - Temperature	0.8852	3.7615	120.5408	0.1512	1.2707	11.1304	0.9879	60.0121	0.0581	0.1391	0.0681	0.0535

Table C.7 IAE Result of Pressure Deviation for the Process Unit

Unit	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
R1 - Pressure	23.5498	7.4909	4.5914	7.5658	1.1047	2.6534	0.1503	0.1820	2.5991	3.7544	5.6197	0.3969
R2 - Pressure	24.8941	7.4143	4.9768	7.0467	0.2807	2.8793	0.1913	0.2297	2.6811	3.7220	5.3677	0.3830
R3 - Pressure	28.3882	7.2261	5.2123	6.6215	0.1998	2.8640	1.4080	0.3859	2.9065	5.8651	5.2178	0.3867
Tank - Pressure	209.1803	61.0661	17.7956	51.1784	5.4957	4.7883	0.8463	3.5522	20.7824	115.8036	307.1484	6.6181
KO - Pressure	40.1318	13.9933	3.5762	8.0616	1.3498	1.0414	0.1780	0.9672	5.2289	5.8214	49.7261	1.1692
DIB-Pressure-Condenser	44.1730	3.0745	1.8007	4.5789	0.1982	0.3063	0.2468	0.2282	0.7707	7.3482	16.7805	0.8808
DIB-Pressure-Reboiler	44.1882	3.0818	1.7990	4.5834	0.1981	0.3085	0.2489	0.2283	0.8150	7.3493	16.7689	0.8800
DP-Pressure-Condenser	259.6840	41.7277	30.8928	68.0255	5.0841	1.2959	1.0196	3.5994	18.6053	92.3283	323.3950	6.9555
DP-Pressure-Reboiler	83.0275	41.8516	28.0904	46.9160	5.0999	1.2997	1.0238	3.6080	18.6622	92.5979	136.1019	6.9723
DB-Pressure-Condenser	30.6413	7.4114	4.0104	5.1686	0.5108	2.0116	1.9824	0.2670	54.3932	16.4551	10.6644	1.4547
DB-Pressure-Reboiler	30.0839	7.3358	3.9322	5.0274	0.5010	1.9526	1.9506	0.2640	54.2615	16.3608	10.3330	1.4203

Table C.7 (Continued) IAE Result of Pressure Deviation for the Process Unit

Unit	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
R1 - Pressure	0.4038	1.4351	3.3419	0.0506	0.4925	10.4217	0.3851	4.2555	0.0192	0.0484	0.0233	0.0178
R2 - Pressure	0.4013	1.3703	3.1838	0.0579	0.4676	10.3916	0.3625	4.7336	0.0167	0.0589	0.0202	0.0156
R3 - Pressure	0.3766	1.3359	4.8007	0.0636	0.4663	13.0947	0.3581	8.0703	0.0300	0.0542	0.0333	0.0288
Tank - Pressure	3.1757	11.3798	64.6421	0.4220	4.3958	135.8245	3.4685	30.3160	0.3368	0.3552	0.3633	0.3268
KO - Pressure	0.7974	2.8158	10.7329	0.1019	1.0179	22.6990	0.7983	7.4969	0.0557	0.0903	0.0628	0.0532
DIB-Pressure-Condenser	1.0603	5.1358	62.8093	0.1741	0.9450	11.6921	0.5709	13.0951	0.0385	0.1231	0.0557	0.0324
DIB-Pressure-Reboiler	1.0708	5.1355	62.9245	0.1736	0.9443	11.6978	0.5702	13.0948	0.0377	0.1240	0.0550	0.0316
DP-Pressure-Condenser	2.4365	10.9655	212.8616	0.2580	9.0302	124.1930	7.6693	24.2690	0.3098	0.1881	0.3296	0.3015
DP-Pressure-Reboiler	2.4420	10.9969	74.2079	0.2596	9.0377	124.6380	7.6712	24.3565	0.3120	0.1895	0.3319	0.3037
DB-Pressure-Condenser	19.0972	6.3410	102.0695	27.6265	1.6306	28.3763	1.0362	32.2665	3.3112	49.0363	27.3660	0.9703
DB-Pressure-Reboiler	18.9876	6.0194	102.2018	27.5058	1.5912	28.2650	1.0135	32.1468	3.2679	49.7012	26.8181	0.9890

Table C.8 IAE Result of Tray Composition Deviation for DIB Column

Tray	V3	V1	V2	V12	V11	V22	V21	V32
1	7.5620	0.7362	0.2624	1.1736	0.0376	0.2582	0.1625	0.0125
2	9.0461	0.9074	0.2893	1.4047	0.0435	0.3080	0.1941	0.0147
3	11.3386	1.1378	0.3610	1.7666	0.0549	0.3886	0.2448	0.0178
4	14.2171	1.4220	0.4534	2.2211	0.0697	0.4896	0.3090	0.0216
5	17.7749	1.7690	0.5690	2.7838	0.0880	0.6146	0.3881	0.0261
6	22.1405	2.1897	0.7126	3.4743	0.1109	0.7676	0.4857	0.0315
7	27.4534	2.6937	0.8901	4.3152	0.1387	0.9540	0.6039	0.0379
8	33.8501	3.2909	1.1073	5.3264	0.1727	1.1775	0.7465	0.0454
9	41.4511	3.9863	1.3704	6.5263	0.2134	1.4422	0.9156	0.0541
10	50.3337	4.7801	1.6850	7.9244	0.2615	1.7494	1.1131	0.0640
11	60.4979	5.6630	2.0550	9.5173	0.3172	2.0973	1.3389	0.0749
12	71.8274	6.6088	2.4842	11.2859	0.3787	2.4841	1.5863	0.0872
13	84.0284	7.5831	2.9686	13.1746	0.4465	2.8919	1.8531	0.0998
14	96.6082	8.5249	3.5017	15.1052	0.5167	3.3052	2.1240	0.1126
15	108.8432	9.3629	4.0635	16.9598	0.5866	3.6946	2.3848	0.1243
16	119.8227	10.0105	4.6306	18.6019	0.6503	4.0320	2.6122	0.1351
17	128.5276	10.3904	5.1694	19.8774	0.7043	4.2792	2.7886	0.1427
18	134.0054	10.4364	5.6511	20.6578	0.7427	4.4113	2.8917	0.1475
19	135.5468	10.1216	6.0462	20.8516	0.7634	4.4064	2.9124	0.1481
20	132.8698	9.4578	6.3373	20.4378	0.7651	4.2624	2.8472	0.1446
21	126.2034	8.4969	6.5209	19.4700	0.7481	4.0032	2.7019	0.1382
22	116.2434	7.3255	6.6018	18.0591	0.7171	3.6610	2.4954	0.1290
23	104.0093	6.0391	6.5983	16.3567	0.6760	3.2648	2.2475	0.1189
24	90.6267	4.7303	6.5336	14.5200	0.6298	2.8475	1.9804	0.1090
25	77.1336	3.4755	6.4307	12.6863	0.5840	2.4368	1.7147	0.1002

Table C.8 (Continued) IAE Result of Tray Composition Deviation for DIB Column

Tray	V3	V1	V2	V12	V11	V22	V21	V32
26	64.3616	2.3929	6.3134	10.9639	0.5416	2.0556	1.4641	0.0935
27	53.3232	1.6062	6.3165	9.5429	0.5168	1.7446	1.2582	0.0915
28	43.3458	1.1014	6.2607	8.2487	0.4973	1.4616	1.0689	0.0895
29	34.6204	0.8912	6.2122	7.1223	0.4840	1.2174	0.8974	0.0882
30	27.2109	1.1101	6.1709	6.1634	0.4789	1.0081	0.7554	0.0878
31	21.0846	1.6846	6.1548	5.3787	0.4702	0.8368	0.6303	0.0874
32	16.1131	2.1680	6.1598	4.7458	0.4631	0.6960	0.5258	0.0875
33	12.1269	2.5697	6.1806	4.2342	0.4601	0.5781	0.4423	0.0889
34	8.9631	2.9097	6.2213	3.8194	0.4481	0.4786	0.3645	0.0897
35	6.7330	3.1841	6.2630	3.4450	0.4268	0.4020	0.2931	0.0971
36	9.8927	1.0478	3.6187	3.1989	0.2883	0.2004	0.2596	0.0809
37	9.7146	0.8790	3.3115	3.0104	0.2674	0.1831	0.2428	0.0823
38	9.3961	0.7675	3.0717	2.8254	0.2483	0.1725	0.2245	0.0816
39	9.0714	0.6619	2.8420	2.6442	0.2302	0.1654	0.2073	0.0815
40	8.7450	0.5592	2.6194	2.4664	0.2146	0.1628	0.1928	0.0839
41	8.4187	0.4634	2.4073	2.2948	0.1973	0.1620	0.1770	0.0844
42	8.0872	0.3748	2.2045	2.1286	0.1785	0.1620	0.1600	0.0826
43	7.7373	0.2903	2.0066	1.9632	0.1623	0.1668	0.1455	0.0832
44	7.3599	0.2161	1.8162	1.7998	0.1448	0.1697	0.1299	0.0817
45	6.9338	0.1520	1.6297	1.6345	0.1283	0.1725	0.1152	0.0805
46	6.4310	0.1008	1.4448	1.4638	0.1123	0.1746	0.1007	0.0785
47	5.8034	0.0653	1.2572	1.2810	0.0969	0.1781	0.0864	0.0763
48	4.9701	0.0484	1.0634	1.0761	0.0792	0.1828	0.0695	0.0702
49	3.7658	0.0423	0.8457	0.8194	0.0611	0.2065	0.0514	0.0679
50	1.9600	0.0661	0.5813	0.4851	0.0363	0.2821	0.0327	0.0714

Table C.8 (Continued) IAE Result of Tray Composition Deviation for DIB Column

Tray	V31	wkcomp	qcond	V5	V4	qc1	qr1	V7
1	5.9504	1.2143	2.0142	0.1229	1.2837	0.7709	17.7813	0.0294
2	7.2640	1.5021	2.4100	0.1557	1.5067	0.9538	20.7488	0.0408
3	9.2099	1.8865	3.0244	0.1951	1.8924	1.1974	22.9290	0.0527
4	11.6426	2.3646	3.7958	0.2435	2.3743	1.4992	24.9739	0.0677
5	14.6398	2.9534	4.7516	0.3032	2.9627	1.8715	30.3426	0.0859
6	18.3034	3.6732	5.9262	0.3757	3.6757	2.3269	38.0993	0.1087
7	22.7370	4.5434	7.3575	0.4638	4.5295	2.8796	47.7509	0.1358
8	28.0367	5.5833	9.0812	0.5685	5.5399	3.5423	59.5715	0.1690
9	34.2707	6.8062	11.1285	0.6917	6.7150	4.3260	73.9040	0.2083
10	41.4588	8.2170	13.5177	0.8333	8.0538	5.2359	91.0699	0.2546
11	49.5389	9.8053	16.2443	0.9916	9.5389	6.2687	111.3146	0.3084
12	58.3285	11.5356	19.2735	1.1676	11.1263	7.4107	134.7377	0.3654
13	67.5065	13.3523	22.5162	1.3492	12.7519	8.6237	161.1841	0.4292
14	76.5800	15.1603	25.8348	1.5315	14.3123	9.8551	190.1585	0.4934
15	84.9152	16.8404	29.0281	1.6997	15.6817	11.0271	220.7136	0.5565
16	91.7859	18.2484	31.8538	1.8438	16.7126	12.0496	251.4146	0.6101
17	96.4925	19.2472	34.0449	1.9450	17.2700	12.8228	280.3602	0.6529
18	98.4821	19.7183	35.3669	1.9960	17.2430	13.2642	305.3553	0.6763
19	97.4875	19.6010	35.6553	1.9867	16.5859	13.3167	324.1736	0.6800
20	93.5884	18.9001	34.8641	1.9182	15.3221	12.9697	334.9285	0.6624
21	87.2010	17.6883	33.0774	1.8004	13.5450	12.2611	336.4227	0.6233
22	78.9904	16.0955	30.4879	1.6431	11.4040	11.2657	328.3845	0.5690
23	69.7218	14.2737	27.3597	1.4632	9.0640	10.0840	311.5312	0.5035
24	60.1284	12.3733	23.9701	1.2754	6.6787	8.8190	287.4118	0.4323
25	50.8118	10.5214	20.5655	1.0917	4.4342	7.5603	258.1014	0.3615

Table C.8 (Continued) IAE Result of Tray Composition Deviation for DIB Column

Tray	V31	wkcomp	qcond	V5	V4	qc1	qr1	V7
26	42.1967	8.8099	17.3378	0.9222	2.7018	6.3777	225.8302	0.2938
27	35.1261	7.3232	14.6264	0.7751	1.9898	5.3840	197.6993	0.2408
28	28.7007	5.9747	12.1119	0.6422	1.4683	4.4753	169.2819	0.1926
29	23.0279	4.7891	9.8609	0.5307	1.0712	3.6782	141.8025	0.1456
30	18.1470	3.7802	7.8895	0.4306	0.7766	2.9934	116.2829	0.1104
31	14.0081	2.9282	6.2006	0.3530	0.5620	2.4257	93.3459	0.0751
32	10.5361	2.2233	4.7692	0.2919	0.4079	1.9658	73.2254	0.0452
33	7.6410	1.6526	3.5651	0.2429	0.2968	1.6037	55.8593	0.0283
34	5.2065	1.1975	2.5869	0.2173	0.2211	1.3483	40.9680	0.0188
35	3.1547	0.8788	1.9602	0.2160	0.1825	1.2005	28.1942	0.0199
36	2.8943	0.9162	2.3297	0.1961	0.2197	1.0128	28.9581	0.0143
37	2.5767	0.9408	2.4060	0.1961	0.2277	1.0116	26.2744	0.0136
38	2.2659	0.9547	2.4316	0.1960	0.2361	1.0046	23.8280	0.0116
39	1.9759	0.9667	2.4385	0.1944	0.2424	0.9923	21.5875	0.0106
40	1.7074	0.9785	2.4279	0.1903	0.2452	0.9734	19.5353	0.0124
41	1.4555	0.9852	2.4036	0.1870	0.2489	0.9516	17.6612	0.0125
42	1.2195	0.9854	2.3655	0.1846	0.2537	0.9266	15.9469	0.0108
43	1.0029	0.9825	2.3088	0.1792	0.2558	0.8936	14.3697	0.0118
44	0.8039	0.9698	2.2354	0.1740	0.2591	0.8557	12.9096	0.0115
45	0.6257	0.9471	2.1407	0.1672	0.2616	0.8094	11.5384	0.0120
46	0.4677	0.9094	2.0214	0.1586	0.2645	0.7535	10.2213	0.0124
47	0.3319	0.8503	1.8721	0.1481	0.2691	0.6850	8.9035	0.0130
48	0.2306	0.7539	1.6925	0.1405	0.2868	0.6042	7.4791	0.0101
49	0.3141	0.6095	1.4850	0.1335	0.3330	0.4994	5.6515	0.0086
50	0.7926	0.4170	1.3800	0.1491	0.4814	0.3917	3.0141	0.0024

Table C.8 (Continued) IAE Result of Tray Composition Deviation for DIB Column

Tray	qc2	qr2	V9	V8	qc3	qr3	V15	V14
1	0.1535	2.0653	0.1119	2.2565	0.0068	0.0336	0.0039	0.0082
2	0.1919	2.5684	0.1356	2.8111	0.0079	0.0467	0.0037	0.0099
3	0.2410	3.2247	0.1707	3.5285	0.0099	0.0604	0.0045	0.0125
4	0.3015	4.0388	0.2145	4.4160	0.0122	0.0769	0.0052	0.0155
5	0.3764	5.0396	0.2689	5.5068	0.0153	0.0977	0.0064	0.0195
6	0.4675	6.2608	0.3354	6.8379	0.0187	0.1228	0.0075	0.0240
7	0.5784	7.7349	0.4168	8.4452	0.0235	0.1540	0.0094	0.0301
8	0.7106	9.4928	0.5142	10.3636	0.0288	0.1912	0.0112	0.0370
9	0.8665	11.5560	0.6295	12.6177	0.0352	0.2357	0.0136	0.0454
10	1.0464	13.9305	0.7632	15.2158	0.0423	0.2874	0.0160	0.0548
11	1.2485	16.5967	0.9140	18.1385	0.0494	0.3457	0.0180	0.0646
12	1.4731	19.4943	1.0831	21.3233	0.0605	0.4138	0.0230	0.0786
13	1.7069	22.5245	1.2595	24.6632	0.0692	0.4839	0.0257	0.0904
14	1.9425	25.5274	1.4391	27.9860	0.0793	0.5571	0.0296	0.1037
15	2.1618	28.3005	1.6074	31.0705	0.0871	0.6260	0.0318	0.1146
16	2.3509	30.6050	1.7548	33.6553	0.0959	0.6892	0.0356	0.1262
17	2.4867	32.2119	1.8630	35.4851	0.1005	0.7363	0.0369	0.1330
18	2.5582	32.9303	1.9239	36.3469	0.1045	0.7661	0.0389	0.1384
19	2.5529	32.6649	1.9270	36.1278	0.1039	0.7718	0.0389	0.1382
20	2.4711	31.4311	1.8719	34.8377	0.0999	0.7530	0.0373	0.1334
21	2.3241	29.3580	1.7670	32.6117	0.0946	0.7138	0.0360	0.1266
22	2.1248	26.6650	1.6210	29.6848	0.0860	0.6556	0.0330	0.1155
23	1.8947	23.6074	1.4502	26.3356	0.0765	0.5858	0.0298	0.1031
24	1.6530	20.4348	1.2692	22.8386	0.0669	0.5104	0.0266	0.0901
25	1.4155	17.3555	1.0899	19.4244	0.0566	0.4338	0.0227	0.0765

Table C.8 (Continued) IAE Result of Tray Composition Deviation for DIB Column

Tray	qc2	qr2	V9	V8	qc3	qr3	V15	V14
26	1.1960	14.5201	0.9231	16.2609	0.0473	0.3612	0.0194	0.0640
27	1.0061	12.0603	0.7800	13.5584	0.0385	0.3034	0.0157	0.0526
28	0.8337	9.8391	0.6486	11.1081	0.0300	0.2499	0.0122	0.0419
29	0.6873	7.8965	0.5363	8.9554	0.0268	0.2060	0.0122	0.0365
30	0.5566	6.2482	0.4331	7.1152	0.0174	0.1614	0.0073	0.0254
31	0.4526	4.8672	0.3510	5.5893	0.0147	0.1284	0.0067	0.0211
32	0.3673	3.7339	0.2827	4.4313	0.0127	0.1009	0.0067	0.0176
33	0.2930	2.8274	0.2214	3.5500	0.0069	0.0738	0.0032	0.0109
34	0.2365	2.1289	0.1741	2.8534	0.0068	0.0557	0.0051	0.0098
35	0.1877	1.6809	0.1313	2.2875	0.0064	0.0388	0.0060	0.0082
36	0.1578	1.7476	0.1084	1.6199	0.0021	0.0296	0.0010	0.0042
37	0.1549	1.7524	0.1032	1.6334	0.0018	0.0255	0.0009	0.0040
38	0.1525	1.7422	0.0994	1.6484	0.0019	0.0231	0.0016	0.0045
39	0.1490	1.7304	0.0954	1.6641	0.0016	0.0207	0.0017	0.0043
40	0.1436	1.7187	0.0899	1.6765	0.0017	0.0169	0.0010	0.0025
41	0.1392	1.7017	0.0861	1.6758	0.0022	0.0148	0.0014	0.0021
42	0.1359	1.6775	0.0836	1.6603	0.0013	0.0140	0.0009	0.0027
43	0.1300	1.6491	0.0791	1.6327	0.0025	0.0121	0.0017	0.0016
44	0.1246	1.6091	0.0756	1.5855	0.0025	0.0111	0.0016	0.0015
45	0.1173	1.5560	0.0709	1.5171	0.0031	0.0100	0.0020	0.0015
46	0.1079	1.4836	0.0650	1.4204	0.0035	0.0091	0.0022	0.0016
47	0.0950	1.3850	0.0564	1.2850	0.0049	0.0081	0.0036	0.0022
48	0.0811	1.2463	0.0473	1.0840	0.0026	0.0069	0.0019	0.0009
49	0.0566	1.0702	0.0295	0.7768	0.0025	0.0046	0.0020	0.0010
50	0.0333	0.9056	0.0096	0.3594	0.0008	0.0012	0.0008	0.0009
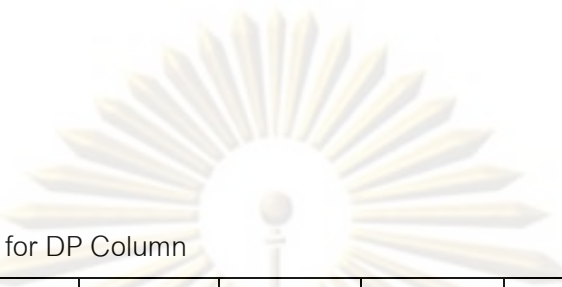
Table C.9 IAE Result of Tray Temperature Deviation for DP Column

Tray	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
1	143.1370	21.2817	12.9250	31.3756	2.7280	1.0796	0.5578	1.8983	10.9723	46.1664	185.0196	3.8871
2	144.3188	21.4205	13.6811	32.5840	2.7540	1.1082	0.5680	1.9001	11.1576	46.3790	187.9321	3.8711
3	146.9762	21.5222	14.0046	33.1598	2.7696	1.1212	0.5739	1.9008	11.2623	46.5252	190.1771	3.8489
4	148.4353	21.5948	14.1461	33.4423	2.7772	1.1298	0.5752	1.9032	11.3238	46.6272	191.4171	3.8291
5	149.3245	21.6469	14.2096	33.5838	2.7801	1.1371	0.5735	1.9073	11.3637	46.7013	192.1317	3.8135
6	149.8894	21.6848	14.2419	33.6594	2.7818	1.1438	0.5713	1.9116	11.3941	46.7576	192.5728	3.8033
7	150.2741	21.7059	14.2669	33.7105	2.7920	1.1427	0.5783	1.9077	11.4293	46.7951	192.8800	3.8061
8	150.5491	21.7317	14.2822	33.7386	2.7899	1.1512	0.5734	1.9143	11.4506	46.8369	193.1013	3.7994
9	150.7703	21.7464	14.3012	33.7665	2.7970	1.1525	0.5776	1.9130	11.4796	46.8672	193.2929	3.8033
10	150.9578	21.7635	14.3182	33.7874	2.7990	1.1579	0.5770	1.9160	11.5037	46.8999	193.4595	3.8031
11	154.4760	21.6988	14.4006	33.9141	2.7912	0.9858	0.5751	1.9120	11.4260	46.6802	193.5363	3.7984
12	154.5318	21.6490	14.3984	33.9195	2.7903	0.9588	0.5793	1.9025	11.3582	46.6188	193.4927	3.7982
13	154.4858	21.6078	14.3902	33.9131	2.7852	0.9421	0.5794	1.8972	11.2880	46.5694	193.5261	3.7940
14	154.4953	21.5716	14.3837	33.9065	2.7809	0.9279	0.5801	1.8917	11.2228	46.5239	193.7664	3.7911
15	154.6254	21.5472	14.3765	33.9005	2.7743	0.9184	0.5777	1.8904	11.1607	46.4924	194.4153	3.7874

Table C.9 (Continued) IAE Result of Tray Temperature Deviation for DP Column

Tray	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
16	154.9925	21.5294	14.3764	33.9083	2.7792	0.9068	0.5845	1.8825	11.1198	46.4763	195.8414	3.7962
17	155.7941	21.5456	14.3717	33.9237	2.7772	0.9062	0.5819	1.8867	11.0877	46.5151	198.6385	3.8020
18	157.4154	21.6034	14.3734	33.9738	2.7858	0.9102	0.5847	1.8924	11.0910	46.6358	203.8098	3.8236
19	160.5142	21.7323	14.3911	34.0947	2.8155	0.9199	0.5986	1.8990	11.1570	46.9033	212.7657	3.8757
20	166.0825	22.0043	14.4316	34.3382	2.8599	0.9480	0.6121	1.9251	11.3064	47.4523	226.9868	3.9602
21	175.2763	22.5113	14.5295	34.8174	2.9368	1.0003	0.6308	1.9782	11.5963	48.4756	247.0480	4.1045
22	188.5728	23.3831	14.7471	35.7098	3.0680	1.0825	0.6637	2.0676	12.0981	50.2514	271.1240	4.3461
23	204.3035	24.7492	15.1736	37.2244	3.2772	1.1939	0.7217	2.1996	12.8711	53.0736	294.2508	4.7202
24	217.8798	26.6249	15.8720	39.4249	3.5484	1.3269	0.7923	2.3813	13.8589	57.0172	309.3259	5.2065
25	223.4470	28.6332	16.7881	41.9208	3.8327	1.4302	0.8744	2.5603	14.7999	61.3998	309.8849	5.7066
26	217.5801	29.9395	17.5871	43.6227	3.9650	1.4536	0.8996	2.6811	15.1545	64.5076	293.8253	5.9584
27	202.3172	29.6306	17.8559	43.4639	3.8588	1.3456	0.8658	2.6412	14.5656	64.3416	266.0249	5.7922
28	183.8476	27.6857	17.4547	41.4585	3.5426	1.1449	0.7829	2.4469	13.2130	60.5871	235.7882	5.2587
29	167.8032	25.0360	16.6599	38.6889	3.1570	0.9388	0.6822	2.1985	11.6763	55.0371	210.6893	4.6033
30	156.4164	22.6399	15.8477	36.1974	2.8279	0.7809	0.5979	1.9813	10.4146	49.7982	193.2756	4.0447

Table C.9 (Continued) IAE Result of Tray Temperature Deviation for DP Column

Tray	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
1	0.8738	5.7315	105.8405	0.1341	4.5107	62.8079	3.7905	12.8048	0.1571	0.0971	0.1679	0.1510
2	0.8392	5.7653	105.5395	0.1378	4.4941	62.3631	3.7780	12.7505	0.1629	0.1014	0.1735	0.1563
3	0.8235	5.7865	106.8735	0.1405	4.4822	61.9116	3.7708	12.6670	0.1668	0.1044	0.1775	0.1602
4	0.8166	5.7991	107.3467	0.1408	4.4728	61.5641	3.7651	12.5958	0.1673	0.1053	0.1781	0.1606
5	0.8137	5.8066	107.5459	0.1395	4.4653	61.3301	3.7599	12.5441	0.1653	0.1046	0.1762	0.1586
6	0.8118	5.8126	107.6279	0.1379	4.4606	61.1865	3.7565	12.5096	0.1631	0.1035	0.1740	0.1563
7	0.8070	5.8273	107.6630	0.1422	4.4664	61.0986	3.7630	12.4929	0.1698	0.1076	0.1807	0.1629
8	0.8066	5.8293	107.6699	0.1387	4.4617	61.0635	3.7582	12.4760	0.1645	0.1049	0.1755	0.1577
9	0.8028	5.8407	107.6697	0.1410	4.4663	61.0479	3.7628	12.4697	0.1683	0.1073	0.1793	0.1615
10	0.8009	5.8473	107.6612	0.1402	4.4668	61.0528	3.7628	12.4629	0.1673	0.1067	0.1783	0.1604
11	0.8121	5.8309	113.8600	0.1390	4.4495	60.7763	3.7442	12.7127	0.1648	0.1049	0.1758	0.1581
12	0.8223	5.8195	114.3432	0.1413	4.4344	60.7254	3.7297	12.7651	0.1677	0.1060	0.1786	0.1610
13	0.8345	5.8031	114.6014	0.1406	4.4077	60.6891	3.7054	12.7976	0.1663	0.1048	0.1772	0.1598
14	0.8466	5.7883	114.8728	0.1405	4.3677	60.6549	3.6694	12.8294	0.1659	0.1042	0.1767	0.1593
15	0.8597	5.7717	115.2125	0.1383	4.2992	60.6339	3.6070	12.8632	0.1623	0.1017	0.1731	0.1558

Table C.9 (Continued) IAE Result of Tray Temperature Deviation for DP Column

Tray	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
16	0.8686	5.7675	115.7014	0.1419	4.1954	60.6295	3.5121	12.9145	0.1680	0.1048	0.1788	0.1615
17	0.8809	5.7583	116.4635	0.1390	4.0055	60.6843	3.3357	12.9810	0.1635	0.1017	0.1743	0.1572
18	0.8914	5.7635	117.7472	0.1381	3.6906	60.8327	3.0413	13.0874	0.1628	0.1008	0.1735	0.1564
19	0.8969	5.7963	119.9913	0.1430	3.1712	61.1536	2.5534	13.2620	0.1709	0.1053	0.1817	0.1644
20	0.9051	5.8571	123.9209	0.1442	2.3086	61.8089	1.7305	13.5358	0.1735	0.1062	0.1845	0.1671
21	0.9171	5.9727	130.5917	0.1436	1.0369	63.0404	0.4831	13.9710	0.1743	0.1057	0.1854	0.1677
22	0.9357	6.1818	141.0362	0.1446	1.4430	65.2043	1.6826	14.6443	0.1780	0.1063	0.1896	0.1710
23	0.9653	6.5249	155.0849	0.1524	4.4303	68.6949	4.6583	15.6126	0.1921	0.1130	0.2043	0.1846
24	1.0197	6.9927	169.6396	0.1619	8.3498	73.6589	8.4082	16.8058	0.2079	0.1196	0.2207	0.1998
25	1.0942	7.4990	178.6906	0.1824	12.1648	79.3240	12.0871	17.9082	0.2369	0.1356	0.2505	0.2284
26	1.1888	7.7853	177.1539	0.1823	14.2490	83.5861	14.1383	18.3341	0.2379	0.1336	0.2522	0.2295
27	1.2564	7.6635	165.5927	0.1760	13.2198	83.8683	13.1997	17.6886	0.2294	0.1285	0.2435	0.2216
28	1.2718	7.1588	149.7070	0.1654	9.4891	79.5744	9.6192	16.2063	0.2141	0.1213	0.2275	0.2075
29	1.2473	6.5073	135.2624	0.1512	4.9401	72.8079	5.2019	14.5455	0.1933	0.1112	0.2055	0.1877
30	1.2073	5.9361	124.7934	0.1390	1.2611	66.2321	1.4934	13.1968	0.1747	0.1024	0.1860	0.1701

Table C.10 IAE Result of Tray Temperature Deviation for DB Column

Tray	V3	V1	V2	V12	V11	V22	V21	V32	V31	wkcomp	qcond	V5
1	48.4416	1554.0847	10.0361	14.9562	4.2188	10.3927	16.4674	0.9602	328.2463	142.9029	14.0297	7.9244
2	82.0000	5202.2466	61.2315	89.2094	19.1204	56.4356	70.6329	5.0496	712.6402	373.1440	91.4279	41.7484
3	386.6306	7129.0079	180.9171	307.3663	42.8950	171.2050	137.7911	12.3018	691.4464	415.0516	343.6438	117.4492
4	792.5965	3926.8491	250.8339	563.7864	33.3091	217.9827	94.2529	10.2313	365.5278	209.1736	654.3078	122.8602
5	987.4736	1254.2576	183.6241	630.1062	11.5756	115.2146	31.6762	3.5251	164.4321	64.5964	754.7704	50.0121
6	1019.2582	417.6943	83.6950	566.9648	3.3482	37.5147	9.8930	0.9800	105.7086	21.4441	697.8078	12.8322
7	987.6347	224.7134	30.1140	442.2126	1.4646	13.2848	4.7375	0.6644	92.9085	14.6360	576.2137	3.8705
8	869.6411	184.1746	14.0255	256.5040	1.1833	7.6884	3.6392	0.6629	90.8264	13.7928	388.1299	3.0472
9	567.7727	176.7709	10.3418	98.6805	1.1485	6.5342	3.4205	0.6774	91.0619	13.8424	180.8583	3.0807
10	800.5394	407.0781	22.2231	204.4446	2.4813	14.4127	7.4819	1.2606	107.1387	25.4759	318.3601	6.1151
11	1004.0442	88.3795	4.9917	237.6499	0.5695	3.5005	1.0745	0.4768	33.1899	12.7262	443.6817	1.8885
12	1012.7862	192.3825	4.4412	174.8593	0.5920	2.1066	2.3373	0.3332	64.5164	20.4227	423.8749	1.8126
13	981.4595	207.9003	5.3333	107.6701	0.6807	2.6436	2.6232	0.3553	67.6943	21.3114	367.9204	1.9334
14	907.9156	209.0145	5.3980	41.1311	0.6836	2.6861	2.6408	0.3568	67.9585	21.3783	267.4848	1.9457
15	628.8034	209.3015	5.4067	10.7104	0.6882	2.6879	2.6479	0.3574	68.0156	21.3994	100.6023	1.9447

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Table C.10 (Continued) IAE Result of Tray Temperature Deviation for DB Column

Tray	V4	qc1	qr1	V7	qc2	qr2	V9	V8	qc3	qr3	V15	V14
1	185.0294	19.3298	385.5067	457.1093	8.8031	215.5708	5.3508	200.4328	5.6772	110.4467	17.6646	10.4092
2	466.2653	121.0972	407.6824	804.2454	46.5859	494.6528	28.2662	456.5908	42.6380	215.7565	87.7536	36.1685
3	508.8683	425.0230	706.3533	691.3761	132.4715	500.9272	78.2828	497.6666	126.6498	537.6673	402.4730	74.5432
4	261.1580	797.7762	1133.9585	292.2529	141.6068	243.0294	80.0574	303.8647	139.1408	949.2483	805.6307	55.0350
5	90.9590	883.0033	1348.8576	59.0199	59.0859	74.6101	32.9477	151.5156	58.6742	1111.7043	963.5635	18.3298
6	39.7730	711.6278	1394.0974	12.0226	15.8122	28.0843	9.5951	95.9156	13.6100	1087.5635	913.0603	4.8246
7	28.1914	415.5654	1385.9249	24.2590	4.3431	22.5863	3.5225	71.5482	1.3903	987.1110	714.9843	1.5774
8	25.8986	160.1043	1352.9089	27.6041	2.5193	22.1093	2.1963	61.3784	1.5305	775.4102	370.1419	0.8721
9	25.6076	52.8946	1284.1441	28.3633	2.4360	22.3644	1.9328	59.0565	2.0847	403.6206	95.5047	0.7176
10	46.8040	145.4662	1493.2889	59.1204	5.1971	38.5617	4.3315	108.9219	8.5078	598.3038	297.0085	1.3900
11	10.1745	100.7885	1691.5259	51.6389	1.5295	24.8620	0.5814	73.3087	3.8703	765.5822	411.8304	1.5073
12	23.0928	10.6843	1689.2759	38.0450	1.9959	35.1852	1.2067	45.0732	4.1810	763.2840	344.9723	1.3358
13	24.6980	7.1985	1655.9632	36.1000	2.1642	36.2565	1.3682	40.8201	4.4453	724.3656	250.4868	1.3025
14	24.8208	8.4360	1616.1053	35.9308	2.1800	36.3393	1.3831	41.0610	4.4708	645.8477	137.1243	1.3033
15	24.8481	8.5275	1521.9084	35.9425	2.1792	36.3689	1.3822	41.1011	4.4727	401.3050	42.1847	1.3014

VITA

Miss Saowani Detjareansri was born in Bangkok, Thailand on August 16, 1986. She graduated Bachelor Degree in department of Chemical Engineering from Mahidol University in 2007. After that she studied for master degree in Chemical Engineering and joined Control and Systems Engineering research group at Chulalongkorn University in 2008.