การประเมินสมรรถนะของโครงสร้างการเบ็ดเสร็จพลังงานและโครงสร้างการควบคุม ของโรงงานบิวเทนไอโซเมอไรเซชัน

นางสาวสุพาพร วิไลลักษณ์

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PERFORMANCE EVALUATION OF HEAT INTEGRATED AND CONTROL STRUCTURES OF BUTANE ISOMERIZATION PLANT

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A Thesis Submitted in Partial Fulfillment of the Requirements
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Field of Study	Chemical Engineering
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สุพาพร วิไลลักษณ์ : การประเมินสมรรถนะของโครงสร้างการเบ็ดเสร็จพลังงานและโครงสร้างการควบคุมของโรงงานบิวเทนไอโซเมอไรเซชัน. (PERFORMANCE EVALUATION OF HEAT INTEGRATED AND CONTROL STRUCTURES OF BUTANE ISOMERIZATION PLANT) อ. ที่ปรึกษา : ผศ. ดร. มนตรี วงศ์ศรี, 229 หน้า.

ในงานวิจัยนี้ได้ออกแบบข่ายงานเครื่องแลกเปลี่ยนความร้อนใหม่จำนวน 3 ทางเลือก (RHEN5-7) เพื่อลดการใช้พลังงานสำหรับโรงงานบิวเทนไอโซเมอไรเซชัน โดยใช้วิธีการออกแบบข่ายงานเครื่องแลกเปลี่ยนความร้อนแบบยืดหยุ่นของ Wongsri (1990) ข่ายงานที่ถูกออกแบบนี้ใช้การแลกเปลี่ยนพลังงานด้วยเครื่องแลกเปลี่ยนความร้อน 1 ตัว แทนการใช้เครื่องแลกเปลี่ยนความร้อน 2 ตัว ที่เป็นการออกแบบข่ายงานเครื่องแลกเปลี่ยนความร้อนแบบยืดหยุ่น 4 แบบ ก่อนหน้านี้ (Kunajitpimol, 2006) ข่ายงานเครื่องแลกเปลี่ยนความร้อนแบบยืดหยุ่นทั้ง 7 แบบ และ กรณีฐาน ถูกนำมาทดสอบทางพลวัตด้วยการใช้โปรแกรมไฮซีส การประหยัดพลังงานอยู่ในช่วงร้อยละ 16.36-25.24 เมื่อเทียบกับกรณีฐาน โครงสร้างการควบคุมใหม่ (CS5-CS8) รวมกับของ Luyben (CS1-CS2) และของ Kunajitpimol (CS3-CS4) ถูกนำมาติดตั้งกับกระบวนการที่มีการเบ็ดเสร็จพลังงานทั้ง 8 แบบ ในการประเมินสมรรถนะโครงสร้างการควบคุมได้ใช้ตัวรบกวนกระบวนการ 2 ชนิด ได้แก่ การรบกวนทางความร้อน (การเปลี่ยนแปลงองค์ประกอบของสารที่ออกจากเครื่องปฏิกรณ์) และการรบกวนอัตราการไหลของสาร เราพบว่าโครงสร้างการควบคุม CS3 และ CS7 เป็นโครงสร้างที่เหมาะสมเมื่อมีการรบกวนทางความร้อนกับทุกโครงสร้างการเบ็ดเสร็จพลังงาน CS1 และ CS5 เป็นโครงสร้างที่เหมาะสมเมื่อเกิดการรบกวนทางอัตราการไหลกับ HIP3 HIP4 HIP6 และ HIP7 ดังนั้นเราสามารถสรุปได้ว่า การประเมินสมรรถนะที่ถูกต้องมีความจำเป็นในการเลือกโครงสร้างที่มีการเบ็ดเสร็จพลังงานและโครงสร้างการควบคุมที่เหมาะสม ซึ่งสามารถทำได้ด้วยการจำลองกระบวนการที่สภาวะพลวัตด้วยโปรแกรมไฮซีส

ภาควิชา........วิศวกรรมเคมี........ ลายมือชื่อนิสิต........สุพาพร วิไลลักษณ์........
สาขาวิชา......วิศวกรรมเคมี........ ลายมือชื่ออาจารย์ที่ปรึกษา........
ปีการศึกษา.........2550…………

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SUPAPORN WILAILAK : PERFORMANCE EVALUATION OF HEAT INTEGRATED AND CONTROL STRUCTURES OF BUTANE ISOMERIZATION PLANT. THESIS ADVISOR: ASST. PROF. MONTREE WONGSRI, D.Sc., 229 pp.

In this work, three new heat exchanger networks (RHEN5-7) are designed to reduce energy consumption for Butane Isomerization plant. The resilient heat exchanger networks design follows Wongsri's resilient HEN synthesis method (1990). These structures has 1 feed effluent heat exchanger (FEHE) instead of 2 FEHEs as in the previous four RHEN designs (Kunajitpimol, 2006). The seven RHENs and the base case (Luyben, 1998) are evaluate dynamically using HYSYS. The energy saving ranges from 16.36-25.24% compared to the base case. The new control structures (CS5-CS8) together with Luyben's (CS1-CS2) and Kunajitpimol's (CS3-CS4) are implemented on the eight heat integrated process structures (HIPs). Two kinds of disturbances are used to evaluate control structure performances: thermal disturbance (transformed into composition disturbance after leaving the reactor) and the material flow disturbance. We found that the control structure CS3 and CS7 are the best control performance with all HIPs for thermal disturbance. The control structures CS1 and CS5 with HIP3, HIP4, HIP6 and HIP7 are the best control performance for material recycle flowrate disturbance. We can conclude that the closed-to-realistic performance evaluation is necessary to choose the suitable heat integrated structure and proper control structure. This can be done in real time domain using rigorous process dynamic simulator like HYSYS.

Department……Chemical Engineering…
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Advisor's signature……………………

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NOMENCLATURES

Alt1	Alternative 1
Alt2	Alternative 2
Alt3	Alternative 3
C_3	Propane
C_i	Cold stream
CS1	Reference control structure I
CS2	Reference control structure II
CS3	Reference control structure III
CS4	Reference control structure IV
CS5	Design of control structure 1
CS6	Design of control structure 2
CS7	Design of control structure 3
CS8	Design of control structure 4
D	Disturbance
DMER	Dynamic maximum energy recovery
e	error
H_i	Hot stream
HEN	Heat exchanger network
HIP	Heat integrated process

HPH	Heat pathway heuristics
IAE	Integral absolute error
iC_4	iso-Butane
k_F	The kinetic expression for the isomerization reaction
MER	Maximum energy recovery
nC_4	normal-Butane
P	Pressure, psia
r	Reaction rate of isomerization reaction
RHEN	Resilient heat exchanger network
T	Temperature, ^oF
V_R	Reactor volume
W	The heat capacity flowrate units of Btu/hr-^oF

CHAPTER I

INTRODUCTION

This chapter is an introduction of this research. It consists of importance and reasons for research, research objectives, scope of research, procedure and method, contribution of research, and the research contents.

1.1 Importance and reasons for research

Today petrochemical industry is challenged by many circumstantial variations. Oil price has risen radically during past three years and stays at high level. Especially energy saving is the most important issue in the petrochemical industry associated with cost, regulations, and social relationships. Heat exchanger network (HEN) is now received more and more attention and is widely used for heat recovery purpose in various kind of industries because it determines to a large extent the net energy consumption of the process. Tremendous efforts have been expanded to establish a series of systematic approaches toward conserving energy and also minimizing losses in the process industries. Moreover, 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 design conditions. The process always changes due to disturbance from the external factors and the internal factor. It is necessary to have the control system to control the condition and compensate for any deviation occurred.

In general, most industrial processes contain a complex flowsheet with several recycle streams, energy integration, and many different unit operations. The economic can be improved by introducing recycle streams and energy integration into the process. However, the 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 path for disturbance propagation. Therefore, strategies for plantwide control are required to operate an entire plant safely and achieve its design objectives. Essentially, the plantwide control problem is how to develop the control loops needed to operate an entire process and achieves its design objectives. The problem is extremely complex and is very much opened. There are a combinatorial number of possible choices and alternative strategies to control and manage the disturbance load entering the process.

This study design new heat exchanger network to save energy in the butane isomerization plant and control structure will be design using disturbance load propagation method (Wongsri, 1990) and Luyben heuristic design method (1983), respectively. The main objective is to use plantwide control strategies to develop the new control structures for the butane isomerization process with energy integration schemes that are designed to achieve the control objective and reduce the cost of production. In this work, the performances of the heat exchanger networks are designed and their control structures are evaluated via simulation using HYSYS.

1.2 Research objective

1. To design heat exchanger networks of the butane isomerization plant by using disturbance load propagation method (Wongsri, 1990).

2. To design control structures for heat exchanger network in butane isomerization plant.

3. To assess performance of the designed control structures for heat exchanger network in butane isomerization plant.

4. To assess performance of the designed control structures for heat exchanger network in butane isomerization plant.

1.3 Scope of research

1. The target for design heat exchanger networks of the butane isomerization plant is to achieve possible maximum energy recovery or the minimum utility requirement (Wongsri, 1990).

2 The heat exchanger network with control structures of the butane isomerization plant are programmed using HYSYS for control structure performance tests.

3 Description and data of the heat exchanger network in butane isomerization plant are obtained from William L. Luyben, Bjorn D. Tyreus, and Michael L. Luyben (1998), Poothanakul P. (2002) and Kunajitpimol B. (2006).

4 The design control structures for energy-integrated butane isomerization plant are design using Luyben's heuristics method.

5 The number of design heat exchanger network in the butane isomerization plant are 3 alternative (not include Lyben and Kunajitpimol B.).

6 The number of control structure design is 4 alternatives which are designed and compared with 4 alternatives from earlier work (2 alternative by Luyben and 2 alternatives by Kunajitpimol B.)

1.4 Contribution of Research

1. The new control structures of the butane isomerization plant with heat integration are designed and compared with the earlier work given by L. Luyben with no energy recovery.

2. New energy integrated designs of the butane isomerization plant.

1.5 Procedure Plan

1. Study of plantwide process control theory, the butane isomerization plant and concerned in formation.

2. Study and Design heat exchanger networks of the butane isomerization plant by using HEN heristics.

3. Steady state modeling and simulation of heat exchanger networks of the butane isomerization plant.

4. Study of dynamic modeling and simulation of the heat exchanger network in butane isomerization plant with no energy integation.

5. Design of control structures for heat exchanger network in butane isomerization plant.

6. Dynamic Simulation for the energy-integrated butane isomerization plant with control structures design.

7. Assessment of the dynamic performance of the control structure.

8. Analysis of the design and simulation results.

9. Conclusion of the thesis.

1.6 Research Contents

This thesis is divided into six chapters.

Chapter I is an introduction to this research. This chapter consists of research objectives, scope of research, contribution of research, and procedure plan.

Chapter II reviews the work carried out on heat exchanger networks design, heat integrated processes and plantwide control design.

Chapter III cover some background information of heat exchanger network design, disturbance transfer technique plantwide (Wongsri, 1990) and theory concerning with plantwide control.

Chapter IV describes the process description and the design of heat exchanger networks for the butane isomerization plant.

Chapter V describes the design of plantwide control structures and dynamic simulation results and compare with control structures of Luyben and Kunajitpimol B.

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

This is follow by:

References

Appendix A: Butane Isomerization Process Stream and Equipment Data

Appendix B: Parameter Tuning of Control Structures

Appendix C: Dynamic Responses

CHAPTER II

LITERATURE REVIEW

2.1 Conceptual Design

A synthesis/analysis procedure for developing first flowsheets and base case designs has been established by Douglas (1985). The procedure is described in terms of a hierarchy of decision levels, as follows:

1. Batch versus continuous
2. Input-output structure of the flowsheet
3. Recycle structure of the flowsheet
4. Separation system specification, including vapor and liquid recovery system
5. Heat exchanger network (HEN)

Douglas (1985) considered a continuous process for producing benzene by

hydrodealkylation of toluene (HDA process) to illustrate the procedure. The complete process is always considered at each decision level, but additional fine structure is added to the flowsheet as he proceeds to the later decision level. Each decision level terminates in an economic analysis. Experience indicates that less than one percent of the ideals for new designs are ever commercialized, and therefore it is highly desirable to discard poor projects quickly. Similarly, the later level decisions are guided by the economic analysis of the early level decisions.

In a series of papers, Fisher et al. (1988a,b,c) presented a study of the interface between design and control including process controllability, process operability and selecting a set of controlled variables. At the preliminary stages of a process design, most plants are uncontrollable. That is normally there are not enough manipulative variables in the flowsheet to be able to satisfy all of the process constraints and to optimize all of the operating variables as disturbances enter the plant. In order to develop a systematic procedure for controllability analysis, Fisher et al. (1988a) used the design decision hierarchy described by Douglas (1985) as the decomposition procedure and considered HDA process as a case study. Where at some levels, that are level 1, 2 and 3, the process is uncontrollable, but controllable at level 4 and level 5. If the available manipulated variables are compared with the constraints and operating variables introduced at each level, the preliminary controllability criterion can often be satisfied.

Beside controllability analysis, Fisher et al. (1988b) also focused on operability analysis. The goal of operability analysis is to ensure that there is an adequate amount of equipment over design so that they could satisfy the process constraints and minimize a combination of the operating costs and over design costs over the entire range of anticipated process disturbances. They also followed the same hierarchical procedure to develop operability analysis. For HDA process, the operability decisions were encountered at each level. Fisher et al. (1988c) proposed steady state control structure for HDA process using an optimum steady state control analysis. They found the values of manipulated variables (that minimize the total operating costs for various values of the disturbances) and used it to define the controlled variables.

D. L. Terrill and J. M. Douglas (1988) have studied HDA process from a steady state point of view and determined that the process can be held very close to its optimum for a variety of expected load disturbances by using the following strategy: (1) Fix the flow of recycle gas through the compressor at its maximum value, (2) Hold a constant heat input flowrate in the stabilizer, (3) Eliminate the reflux entirely in the recycle column, (4) Maintain a constant hydrogen-to-aromatic ratio in the reactor inlet by adjusting hydrogen fresh feed, (5) Hold the recycle toluene flowrate constant by adjusting fuel to the furnace, (6) Hold the temperature of the cooling water leaving the partial condenser constant.

Downs and Vogel (1993) described a model of an industrial chemical process for the purpose of developing, studying and evaluating process control technology. It consisted of a reactor/separator/recycle arrangement involving two simultaneous gas-liquid exothermic reactions. This process was well suited for a wide variety of studies including both plantwide control and multivariable control problems.

Tyreus and W. L. Luyben (1993) considered second order kinetics with two fresh feed makeup streams. Two cases are considered: (1) instantaneous and complete one pass conversion of one of the two components in the reactor so there is an excess of only one component that must be recycled and (2) incomplete conversion per pass so there are two recycle streams. It is shown that the generic liquid-recycle rule proposed by Luyben applies in both of these cases: “snowballing” is prevented by fixed the flowrate somewhere in the recycle system. An additional generic rule is proposed fresh feed makeup of any component cannot be fixed unless the component undergoes complete single-pass conversion. In the complete on-pass conversion case, throughput can be set by to fix the flowrate of the limiting reactant. The makeup of the other reactant should be set by level control in the reflux drum of the distillation column.

Yi and Luyben (1995) presented a method that was aimed at helping to solve this problem by providing a preliminary screening of candidate plantwide control structures in order to eliminate some poor structures. Only steady state information was required. Equation-based algebraic equation solvers were used to find the steady state changes that occur in all manipulated variables for a candidate control structure when load changes occur. Each control structure fixed certain variables: flows, compositions, temperatures, etc. The number of fixed variables was equal to the number of degrees of freedom of the closed-loop system. If the candidate control structure required large changes in manipulated variables, the control structure was a poor one because valve saturation and/or equipment overloading will occur. The effectiveness of the remaining structures was demonstrated by dynamic simulation. Some control structures were found to have multiple steady states and produce closed-loop instability.

2.2 Heat Exchanger Networks (HENs)

Linhoff, B. and Hindmarsh, E. (1983) presented a novel method for the design of HEN. The method is the first to combine sufficient simplicity to be used by hand with near certainty to identify “best” designs, even for large problems. Best design features the highest degree of energy recovery possible with a given number of capital items. Moreover, they feature network patterns required for good controllability, plant layout, intrinsic safety, etc. Typically, 20-30 percent energy savings, coupled with capital saving, can be realized in state of the art flowsheets by improved HEN design. The task involves the placement of process and utility heat exchangers to heat and cool process streams from specified supply to specified target temperatures.

Generally, minimum cost networks feature the correct degree of energy recovery and the correct number of units. This is achieved in two stages. First, the method aims for a minimum energy solution, corresponding to a specified ΔT_{min} with no more units than is compatible with minimum energy. This task is achieved through understanding of the pinch phenomenon, hence the method is called the pinch design method. Second, the method involves a controlled reduction in number of units. This may require “backing-off” from minimum utility usage.

The pinch design method also identifies situations where stream splitting is inevitable for a minimum utility design. The pinch design method incorporates five important stages. These are:

1. The HEN problem is divided at the pinch into separate problems.

2. The design for this separate problem is started at the pinch and developed moving away from the pinch. At the pinch essential matches, match options and stream splitting requirements are identified by applying the feasibility criteria.

3. When options exist at the pinch, the engineer is free to base his selection to suit the process requirements.

4. The heat loads of exchangers at the pinch are determined using the stream tick-off heuristic. In case of difficulty, a different exchanger topology at the pinch can be chosen or the load on the offending match can be reduced.

5. Away from the pinch there is generally a free choice of matches. The procedure does not insist on particular matches but allows the designers to discriminate between matches based on his judgment and process knowledge.

Linhoff, B., Dunford, H., and Smith, R., (1983) studied heat integration of distillation columns into overall process. This study reveals that good integration between distillation and the overall process can result in column operating at effectively zero utility cost. Generally, the good integration is when the integration as column not crossing heat recovery pinches of the process and either the reboiler or the condenser being integrated with the process. If these criteria can be met, energy cost for distillation can effectively be zero.

Saboo and Morari (1983) classified flexible HENs into two classes according to the kind and magnitude of disturbances that effect the pinch location. For the temperature variation, they show that if the MER can be expressed explicitly as a function of stream supply and target conditions the problem belongs to Class I, i.e. the case that small variations in inlet temperatures do not affect the pinch temperature location. If an explicit function for the minimum utility requirement valid over the whole disturbance range does not exist, the problem is of Class II, i.e. the case that large changes in inlet temperature of flowrate variations cause the discrete changes in pinch temperature locations.

Calandranis and Stephanopoulos (1988) proposed a new approach to address the following problems: design the configuration of control loops in a network of heat exchangers and sequence the control action of the loops, to accommodate set point changes and reject load disturbances. The approach proposed exploits the structure characteristics of a HEN by identifying routes through the HEN structure that can allocate load (disturbances, or set point changes) to available sinks (external coolers or heaters). They also discussed several design issues such as the placement of bypass lines and the restrictions imposed by the existence of a process pinch. An online, real-time planning of control actions is the essence of implementation strategies generated by an expert controller, which selects path through the HEN is to be used for each entering disturbance or set point change, and what loops should be activated (and in what sequence) to carry the associated load (disturbance or set point change) to a utility unit.

In a series papers, studies of the sensitivity of the total processing cost to heat exchanger network alternatives and steady state operability evaluation were undertaken by Terrill and Douglas (1987a,b,c). They considered the temperature-enthalpy (T-H) diagram and developed six HEN alternatives for a base case design for HDA process which energy savings ranging between 29 and 43%. The simplest of these designs is alternative 1, recovers an additional 29 percent of the base case heat consumption by making the reactor preheater larger and the furnace smaller. The most complicated of the design is alternative 6, recovers 43 percent of the base case net energy consumption.

Several terms have been used in the literature to describe the additional attributes of HENs that have a capability to tolerate change in input or operational parameters while achieving the targets. Operability has been used to describe the ability of the system to perform satisfactorily under normal and abnormal conditions different design condition. Normal refers to the steady state operation while abnormal refers to the transient operation during failure, start up or shut down periods. Flexibility has been used to describe the ability of process systems to readily adjust to meet the requirement of changes, i.e. different feed stocks, product specifications or process conditions. Resiliency refers to the ability of HEN to tolerate and recover from undesirable parameter variations, and the term static resiliency or simply resiliency has been used in the same sense as flexibility. Dynamic resiliency refers to ability to handle the unsteady state operation.

Colberg (1989) suggested that flexibility should deal with planed, desirable changed that often have a discrete set of values. Whereas resilience deals with unplanned, undesirable changes which are naturally continuous values. Thus a flexibility problem is a ‘multiple period’ type pf problem. A resilience problem should be a problem with a continuous range of operating conditions in the neighborhood of nominal operating points.

Wongsri, M., (1990) studied a resilient HEN design. He presented a simple but effective systematic synthesis procedure for the design of resilient HEN. His heuristic design procedure is used to design or synthesize HENs with pre-specified resiliency. It used physical and heuristic knowledge in finding resilient HEN structures. The design must not only feature minimum cost, but must also be able cope with fluctuation or changers in operating conditions. The ability of a HEN to tolerate unwanted changes is called resiliency. It should be noted that the ability of a HEN to tolerate wanted changes is called flexibility. A resilient HEN synthesis procedure was developed based on the match pattern design and a physical understanding of the disturbances propagation concept. The disturbance load propagation technique was developed from the shift approach and was used in a systematic synthesis method. The design condition was selected to be the minimum heat load condition for easy accounting and interpretation. This is a condition where all process streams are at their minimum heat loads, e.g. the input temperatures of hot streams are at the lowest and those of cold streams are at the highest.

Generating designs at a base case and some extreme conditions and combining those designs to a base design requires that the designs should be similar. However, the networks designed at extreme conditions can be very different from each other. This poses difficulties in the combination. These methods involve repetitive effort in funding a resilient structure because the resiliency objective has not been included in their models. Also, the problem of selecting extreme conditions is far from trivial. Grossmann and Morari (1984) show that the extreme conditions that seem logical can load to a poor design. Most of us would select the maximum and minimum operating conditions as design conditions. However, in their example the extreme condition is located at the intermediate value. Extra units in a combined design are then eliminated by either inspection or using optimization methods to obtain a minimum unit solution. A minimum unit solution is tested for resiliency using mathematical programming or inspection techniques.

Marselle et al. (1982) addressed the problem of synthesizing heat recovery networks, where the inlet temperatures vary within given ranges and presented the design procedure for a flexible HEN by finding the optimal network structures for four selected extreme operating conditions separately. The specified worst cases of operating conditions are the maximum heating, the maximum cooling, the maximum total exchange and the minimum total exchange. The network configurations of each worst condition are generated and combined by a designer to obtain the final design. The strategy is to derive similar design in order to have as many common units as possible in order to minimize number of units.

Linnhoff and Kotjabasakis (1984) developed a design procedure for operable HENs by inspection and using the concept of downstream paths, i.e. the paths that connect the disturbed variables downstream to the controlled variables. They generated HEN design alternatives by the pinch method for the nominal operating condition. Then, the alternative designs are inspected for the effects of disturbances on the controlled variables and they are removed by breaking the troublesome downstream paths. Path breaking can be done by relocating and/or removing exchangers. If this procedure is not feasible, control action is inserted into the structure.

Saboo and Morari (1984) proposed the corner point theorem which states that for temperature variation only, if a network allows MER without violating ΔT_{min} at M corner points, then the network is structurally resilient or flexible. This is the case where the constraint is convex, so examining the vertices of the polyhedron is sufficient. This procedure again can only apply to restricted classes of HEN problem. Their design procedure is similar to Marselle et al. (1982), but using two extreme cases to develop the network structure. The strategy for both procedures is finding similar optional network structures for the extreme cases and the base case design in order that they may be easily merged and not have too many units. Two extreme cases are:

1. When all streams enter at their maximum inlet temperatures and the heat capacity flowrates of hot streams are maximal and those of cold streams minimal. This is the case of maximum cooling.

2. When all streams enter at their minimum inlet temperatures and the heat capacity flowrates of hot streams are minimal and those of cold streams maximal. This is an opposite case the above one and in this case maximum heating is required.

The ‘base’ design is then generated by using an optimization technique and the final design is obtained by combining these designs. A test for resiliency (calculating, RI) is required. If the design is not feasible a modification is done by attempting to reduce ΔT_{min} and if not successful, a new heat exchanger will added or some heat exchangers are located. If the modified network is still not resilient, synthesize network structures at all corner points where the current design is not feasible. The new structures should be as similar to the current design as possible. The new design is obtained by superimposing the current structure and the new structures. The unneeded heat exchangers are inspected and removed.

Floudas and Grossmann (1987) presented a synthesis procedure for resilient HENs. Their multiperiod operation transshipment model is used to find a match structure for selected design points. The design obtained for feasibility at the match level. If it is not feasible, the critical point is added as an additional operating point and the problem is reformulated and solved. If the match network is feasible then the multiperiod superstructure is derived and formulated as an NLP problem to find a minimum unit solution.

Ploypaisansang A., (2003) presented to redesign six alternatives for HDA process to be the resiliency networks for maintain the target temperature and also achieve maximum energy recovery (MER). The best resilient network is selected by to trade-off between cost and resiliency. The auxiliary unit should be added in the network for cope safely with the variations and easy to design control structure to the network.

2.3 Design and Control of Energy-Integrated Process

In the last few decades, Douglas, Orcutt, and Berthiaume (1962) studied design and control of feed-effluent heat exchanger – reactor systems. They obtained a simultaneous solution of the steady state heat and material balances for a first order reaction occurring in the system and used it to calculate the values of exchanger and reactor lengths that minimized the equipment cost of the system. A dynamic study indicated that the desired steady state conditions were met sable. However, proportional controller could be used to stabilize the process.

Silverstein and Shinnar (1982) discussed the linear and nonlinear stability analysis of a fixed bed catalytic reactor with heat exchanger between the feed and product streams, with special emphasis on case which are open loop unstable. They used classical frequency response techniques, contains the implicit assumption that the designer should evaluate the effect of the overall design on stability. Tyreus, B.D. and Luyben, W. L., (1993) presented a mathematical analysis of the unusual dynamic in coupled reactor/preheater process. The outlet temperature of the reactor exhibits inverse response for a change in the inlet reactor temperature and a large dead time.

Handogo, R. and Luyben, W. L., (1987) studied the dynamics and control of a heat-integrated reactor/column system. An exothermic reactor was the heat source, and a distillation column reboiler was the heat sink. Two types of heat-integrated system were examined: indirect and direct heat integration. Both indirect and direct heat-integration systems are found in industry. In the indirect heat-integration system, steam generation was used to cool the reactor, and the steam was used as the heating medium for the reboiler. The direct heat-integration system used the reactor fluid to directly heat the column reboiler. The indirect heat-integration system was found to have several advantages over the direct heat-integration system in terms of its dynamic performance. Both systems were operable for both large and small temperature differences between the reactor and column base. Somewhat unexpectedly, the heat-integration system with a small temperature difference was found to be more controllable than a system with a larger temperature difference. However, the cost of the heat exchanger increased rapidly as the temperature difference decreased. An important thing in this study is how to solve some of control difficulties in the process associated with heat integration schemes. They suggested adding auxiliary utility coolers and reboilers to the process.

M.L. Luyben and W.L. Luyben (1995) examined the plantwide design and control of a complex process. The plant contains two reactions steps, three distillation columns, two recycle streams, and six chemical components. Two methods, a heuristic design procedure and a nonlinear optimization, have been used to determine an approximate economically optimal steady state design. The designs differ substantially in terms of the purities and flowrates of the recycle streams. The total annual cost of the nonlinear optimization design i2 about 20 percent less than the cost of the heuristic design. An analysis has also been done to examine the sensitivity to design parameters and specifications. Two effect control strategies have been developed using guidelines from previous plantwide control studies; both require reactor composition control as well as flow control of a stream somewhere in each recycle loop. Several alternative control strategies that might initially have seemed obvious do not work.

Jones, W.E., and Wilson, J.A., (1997) considered the range ability of flows in the bypass line of heat exchanger through interesting heat exchanger problems. Difficulty is immediately encountered when considering heat exchanger between two process streams; changing the flowrate of one will certainly affect the exit temperature of the other. Unfortunately, interfering with a process stream flowrate immediately upsets the plant mass balance, which is undesirable. The difficulty is overcome by using a bypass that does not affect the total flowrate but changes the proportion actually passing through the heat exchanger and hence the heat transfer. Good engineering practice would maintain a minimum flowrate of 5-10% through the bypass. This bypass is expected to be able to handle disturbances.

Luyben, M.L., Tyreus, B.D. and Luyben, W.L., (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. The control system was tested on a dynamic model built with TMODS, Dupont’s in-house simulator.

From the W.L. Luyben (2000) studied the process had the exothermic, irreversible, gas – phase reaction $A + B \longrightarrow C$ occurring in an adiabatic tubular reactor. A gas recycle returns unconverted reactants from the separation section. Four alternative plantwide control structures for achieving reactor exit temperature control were explored. The reactor exit temperature controller changed different manipulated variables in three of the four control schemes: (1) CS1, the setpoint of the reactor inlet temperature controller was changed; (2) CS2, the recycle flowrate was changed; and (3) CS3, the flowrate of one of the reactant fresh feeds was changed. The fourth control scheme, CS4, uses an “on – demand” structure. Looking that the dynamics of the reactor in isolation would lead one to select CS2 because CS1 had a very large deadtime and CS3 had a very small gain. Dynamic simulations demonstrated that in the plantwide environment, with the reactor and separation operating together, the CS3 structure gave effective control and offered an attractive alternative in those cases where manipulation of recycle flowrate was undesirable because of compressor limitations. The on – demand CS4 structure was the best for handling feed composition disturbances.

Kunlawaniteewat, J., (2001) proposed the rules and procedure for design control structure of heat exchanger network using heuristic approach for to achieve outlet temperature targets and maintain maximum energy recovery (MER). The rules are categorized as following: generals, match pattern, loop placement, bypass placement, and split fraction rules.

Wongsri and Kietawarin (2002) presented a comparison among four control structures designed for withstanding disturbances that cause production rate change of HDA process. The changes had been introduced to the amount of toluene and feed temperature before entering the reactor. Compared with the reference control structure using a level control to control toluene quantity in the system, the first control scheme measured toluene flowrate in the process and adjusted the fresh toluene feed rate. This structure resulted in faster dynamic response than the reference structure. The second control scheme was modified from the first scheme by adding a cooling unit to control the outlet temperature from the reactor, instead of using internal process flow. The result was to reduce material and separation ratio fluctuations within the process. The product quality was also quite steadily. In the third control scheme, a ratio control was introduced to the second control scheme for controlling the ratio of hydrogen and toluene within the process. This scheme showed that it could withstand large disturbances. Dynamic study showed that the control structure had significant effect on process behavior. A good system control should quickly response to disturbances and adjusts itself to steady state while minimizing the deviation of the product quality.

Chen, T.H., and Yu, C.C. (2003) proposed systematic approach to complex FEHE schemes. Because a loss of controllability come the positive feedback loop, several design parameters were studied, and the design heuristic were proposed to give more controllable heat integration schemes. They used two examples, a simple two-FEHE example and an HDA process example to illustrate the assessment of controllability based on process flowsheet. The results showed that, contrary to expectations, some complex heat-integrated reactor design alternatives (e.g., alternative 6 of HDA example) were indeed more controllable than some of the simpler heat-integration schemes (e.g., alternative 1). The increased number of FEHEs allows for a greater number of candidate manipulated inputs and thus provides opportunities for multivariable control.

Wongsri and Thaicharoen (2004) presented the new control structures for HDA process with energy integration schemes alternative 3. Five control structures have been designed, tested and compared the performance with Luyben’s structure (CS1). The result showed that the HDA process with heat integration can reduce energy cost. Furthermore, this process can be operated well by using plantwide methodology to design the control structure. The dynamics responses of the designed control structures and the reference structure are similar. The CS2 has been limited in bypass. So, it is able to handle in small disturbances. The CS3 has been designed to improve CS2 in order to handle more disturbances by using auxiliary heater instead of bypass valve to control temperature of stabilizer column. The recycle column temperature control response of the CS4 is faster than that of the previous control structures, because reboiler duty of column can control the column temperature more effective than bottom flow. The CS5, on – demand structure has an advantage when downstream customer desires immediate responses in the availability of the product stream from this process. The energy used in CS6 control structure is less than CS1 and CS4.

Wongsri and Hermawan Y.D., (2004) studied the control strategies for energy integrated HDA plant (i.e. alternative 1, 4 and 6) based on the heat pathway heuristics, i.e. selecting an appropriate heat pathway to carry associated load to a utility unit, so that the dynamic maximum energy recovery (DMER) can be achieved with some trade-off. In addition, a selective controller with low selector switch (LSS) is employed to select an appropriate heat pathway through the network. The new control structure with the LSS has been applied in the HDA plant.

Wongsri and Kunajitpimol B., (2006) presented four alternatives of heat exchanger network (HEN) designs and two alternatives of control structure of Butane Isomerization plant. The control difficulties associated with heat integration were solved by adding auxiliary utilities which was kept minimal. Heat exchanger network designs used the heat from the reactor effluent stream to provide the heat for the column reboiler. The energy saving was 24.88% from the design without heat integration, but the capital cost raised about 0.67% due to adding of a process to process exchanger and an auxiliary utility exchanger to the process. The plantwide control structures were designed following Luyben’s heuristic method. The result showed that the control structures can reject disturbances better than base case which were designed by Luyben. The designed control structure was evaluated based on the rigorous dynamic simulation using the commercial software HYSYS.

Wongsri and Sae-leaw B., (2006) designed the control structures of energy integrated HAD plant with minimum auxiliary reboilers. They outlined the plantwide control design approach that would be taken for a complex heat-integrated scheme like Alternative 6 of the HAD process. It started with specifying the disturbances and their magnitudes, and then designing the resilient heat exchanger network was designed at the minimum heat supply and maximum heat demand condition. They can solve the control difficulties associated with alternative 6 by adding an auxiliary reboiler to the process instead of three as suggested by Luyben (1999). The three new control structures were proposed and their performances were evaluated. CS2 is the best control structure for handle disturbances due to it gives better control performances. In these control structure, the recycle column feed flowrate is flow-controlled so that fluctuations in the process are not propagated to the next downstream unit operations.

CHAPTER III

THEORY

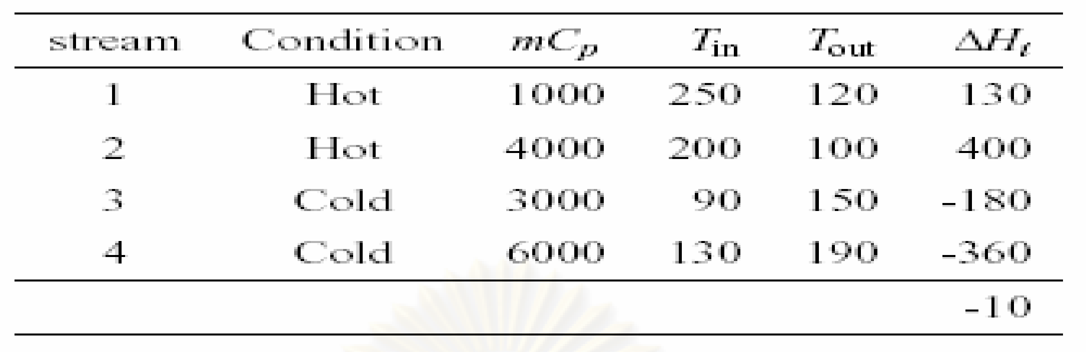
The starting point for an energy integration analysis is the calculation of the minimum heating and cooling requirement for a heat exchanger network. These calculations can be performed without having to specify any heat-exchanger network. Similarly, we can calculate the minimum number of exchangers required to obtain the minimum energy requirements without having to specify a network. Then the minimum energy requirements and minimum number of exchanger provide targets for the subsequent design of a heat-exchanger network.

3.1 Minimum Heating and Cooling Requirements

3.1.1 First law Analysis

Suppose we consider a very simple problem where we have two streams that need to be heated and two streams that need to be cooled (see the data in Table 3.1). If we simply calculate the heat available in the hot streams and heat required for the cold streams, the difference between these two values is the net amount of heat that we would have to remove or supply to satisfy the first law. The results are also shown in the Table 3.1, and the first two entries are determined as follows:

Table 3.1 First-law calculation



stream	Condition	mC_p	T_{in}	T_{out}	ΔH_t
1	Hot	1000	250	120	130
2	Hot	4000	200	100	400
3	Cold	3000	90	150	-180
4	Cold	6000	130	190	-360
					-10

NOTE mCp = FCp is the heat capacity flowrate units of Btu/hr-°F

$Q = \Delta H_t = FCp(T_{in}-T_{out})$ is the heat available for each stream 10^3 Btu/hr

$$Q_1 = F_1Cp_1\Delta T_1 = [1000 \text{ Btu } / (\text{hr.}^{o}\text{F})]\ (250-120)$$

$$= 130 \times 10^3 \text{ Btu/hr}$$

$$Q_2 = F_2Cp_2\Delta T_2 = [4000 \text{ Btu } / (\text{hr.}^{o}\text{F})]\ (200-100)$$

$$= 400 \times 10^3 \text{ Btu/hr.}$$

Thus, 10 x 10^3 Btu/hr must be supplied from utilities if there are no restrictions on temperature - driving forces.

3.1.2 Minimum Utility Loads

The net result of the operation is that the minimum utility requirements have been predicted, i.e., 70 x 10^3 Btu/hr of hot utilities and 60 x 10^3 Btu/hr cold utilities.

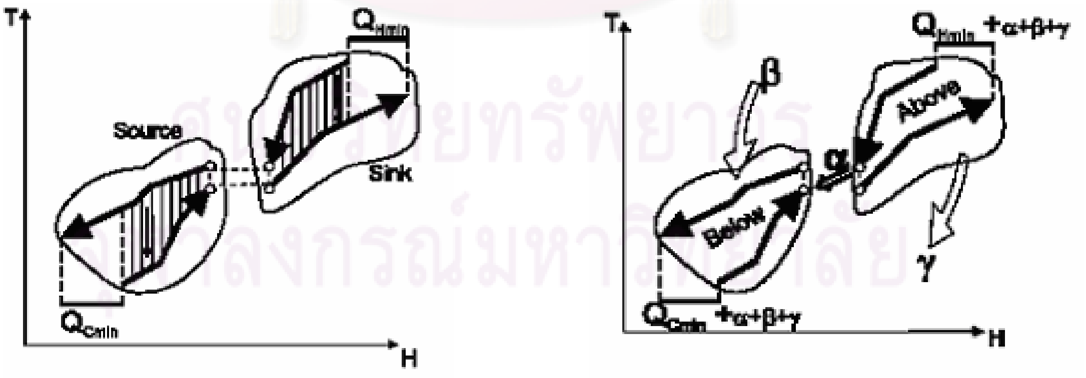
3.1.3 Pinch Temperature

At the third and fourth temperature intervals, there is no energy transfer. We call this the *pinch temperature* (130 °C for the cold streams or 140 °C for the hot streams) Thus, the pinch temperature provides a decomposition of the design problem. That is, above the pinch temperature we only supply heat, whereas below the pinch temperature we only reject heat to a cold utility.

3.1.4 The Pinch Principle

The point where ΔTmin is observed is known as the “Pinch” and recognizing its implications allows energy targets to be realized in practice. Once the pinch has been identified, it is possible to consider the process as two separate systems: one above and one below the pinch, as shown in Figure 3.3(a). The system above the pinch requires a heat input and is therefore a net heat sink. Below the pinch, the system rejects heat and so is a net heat source.

In Figure 3.1 (b) α amount of heat is transferred from above the pinch to below the pinch. The system above the pinch, which was before in heat balance with Q_{Hmin}, now loses α units of heat to the system below the pinch. To restore the heat balance, the hot utility must be increased by the same amount, that is α units. Below the pinch, α units of heat are added to the system that had an excess of heat, therefore the cold utility requirement also increases by α units. In conclusion, the consequence of a *cross-pinch heat transfer* (α) is that both the hot and cold utility will increase by the cross-pinch duty (α).



(a) **(b)**

Figure 3.1 The Pinch Principle

Figure 3.3 (b) also shows γ amount of external cooling above the pinch and β amount of external heating below the pinch. The external cooling above the pinch of γ amount increases the hot utility demand by the same amount. Therefore on an overall basis both the hot and cold utilities are increased by γ amount. Similarly external heating below the pinch of β amount increases the overall hot and cold utility requirement by the same amount (i.e. β).

To summarize, the understanding of the pinch gives three rules that must be obeyed in order to achieve the minimum energy targets for a process:

Heat must not be transferred across the pinch

* *There must be no external cooling above the pinch*

* *There must be no external heating below the pinch*

Violating any of these rules will lead to cross-pinch heat transfer resulting in an increase in the energy requirement beyond the target. The rules form the basis for the network design procedure which is described in "Heat Exchanger Network Design". The design procedure for heat exchanger networks ensures that there is no cross pinch heat transfer. For retrofit applications the design procedure “corrects” the exchangers that are passing the heat across the pinch.

3.1.5 Design of minimum-Energy Heat-Exchanger Networks

There is design heuristic for feasible matches at the pinch condition.

Above the pinch: $(FCp)_{\text{hot streams}} \le (FCp)_{\text{cold streams}}$

Below the pinch: $(FCp)_{\text{hot streams}} \ge (FCp)_{\text{cold streams}}$

Violating any of these rules will lead to violate the minimum approach temperature.

3.1.6 Problem table

For more complication problems, it would be a lengthy procedure to evaluate possible designs for each subnetwork. Fortunately, a for more rapid procedure can adopted (Linnhoff and Flower, 1978).

Table 3.2 Data for test case

stream	condition	mCp	T(supply)	T(target)	Heat Load
1	cold	3	60	180	-360
2	hot	2	180	40	280
3	cold	2.6	30	105	-195
4	hot	4	150	40	140
					Toal =165

NOTE mCp = heat capacity flowrate units of kW/C.

Heat load = mCp(Ts-Tt) unit.

Table 3.3 Problem table for test case Table 3.2

	COLUMNS						1	2	3	4	5
SN	Stream and Temperature Hot			Cold			Deficit	Accumulate		Maximun permissible	
	(2)	(4)		180	(1)	(3)		input	output	input	output
SN(1)	↓		180	170	↑		+30	0	-30	**+60**	+30
SN(2)	↓		150	140	↑		+30	-30	-60	+30	0
SN(3)	↓	↓	115	105	↑		-105	-60	+45	0	+105
SN(4)	↓	↓	70	60		↑	-18	+45	+63	+105	+123
SN(5)		↓	40	30		↑	-102	+63	**+165**	+123	**+225**

In Table 3.3, the search for the upper limits to the loads of heaters and coolers in the subnetworks is carried out in a systematic way. The data used refer to test case Table 3.2, and Table 3.3 represents what will be referred to as a problem table. In column1, the values are given of net heat requirement for each subnetwork. This deficit D_k is difference between the heat input I_k, which correspond to the heat supplied by the heater(s), and the heat output O_k, that is, the heat removed by the cooler(s). For the K^{th} subnetwork, the D_k may be calculated by means of Equation (3.1)

$$D_k = I_k - O_k = (T_k - T_{k+1}) \left(\sum Cp_{cold} - \sum Cp_{hot} \right) \tag{3.1}$$

The summations only include the streams present in SN (K).Since Equation (1) is just an enthalpy balance, the results will be independent of any subnetwork design subsequently adopted. D_k will be positive or negative, depending on whether the heat capacity flow rates of the hot streams are less or greater than those of the cold streams. If D_k is positive, more heating than cooling is required.

Consider, now, the output form SN (K) is passed to SN (K+1) to satisfy any requirements for heat in SN(K+1). If there is no separate connection to a process utility heat source in (K+1), Equation (3.2) can be used to calculate the maximum amount of heat made available to SN(K+1):

$$I_{k+1} = O_k \tag{3.2}$$

Form SN(K+1):

$$O_{k+1} = O_k - D_{k+1} \tag{3.3}$$

Thus, assuming no heat supply to SN(1), the Table 3.3 for the inputs and outputs for each subnetwork are found in columns 2 and 3.

The physical significance of this Table is as follows. If no process utility heat is supplied to any of the subnetworks, and all surplus heat form the matches between the streams in one subnetwork is passed to the next, the heat inputs to each subnetwork would be given as the values in column 2 and the heat outputs in column 3. If any of the values in column 3 are negative, as is the case here for SN(1) and SN(2), process utility heat must be introduced to these subnetworks to increase these outputs to zero. It follows that if one must use process utility heat anywhere in the system, it may as well be introduced at its highest available temperature, that is, into SN(1) and then passed through the sequence of subnetworks. In this way, the amount of heat available in the intervening subnetwork designs. Accordingly, columns 4 and 5 in Table 3.3 have been drawn up. They are base on exactly the same sequence of calculations as column 2 and 3, with the single difference that the minimum heat requirement for the whole network (the most negative Table in column 3) is introduced as the input to SN(1) from process utility heat sources. As result, the Table 3.3 in columns 4 and 5 represent the heat flows into and out of the subnetworks for the case where the necessary minimum process utility heat is received at highest possible temperature.

The transfer of this heat form one subnetwork to the next creates the maximum degree of choice for subnetwork design without any adverse effect on consumption of resources. Any further increase of a subnetwork's heat input must be provided by additional heat form process utilities. Thus, the Table 3.3 in column4 and 5 represent the upper limits for the heater and cooler loads in the subnetworks which must not be exceed if subnetworks are to be designed which do not prevent maximum energy recovery. In these sense, they are maximum permissible values (see Table 3.3). Comparing the limits obtained in Table 3.3 for SN(3), SN(4), and SN(5) with the alternative designs for these subnetworks, it could be adopted for SN(4) and SN(5), but not for SN(3).

Owing to the logic on which the problem table is based , three values in columns 3, 4 and 5 will have a significance not just for the subnetwork to which they belong , but also for whole problem.

In Table 3.3, These figures are shown boxed:

1. The bottom Table3.3 in column 3 denotes the net cooling requirement for the whole problem as found by an overall enthalpy balance (see Table 3.2).

2. The top Table 3.3 in column 4 is the minimum process utility heat requirement for the whole problem.

3. The bottom Table 3.3 in column 5 is the corresponding cooling requirement for the whole problem.

In summary, the following procedure is used:

1. The temperature T_1, T_2….. T_{n+1} are identified.

2. An enthalpy balance, that is Equation (3.1), must be solved for each subnetwork, giving Table 3.3 for net heat requirements, column 1.

3. Columns 2and 3 are calculated by means of Equation (3) and Equation (4), assuming $I_1 = 0$.

4. Columns 4 and 5 are produced by adding the value of the most negative entry in columns 3 to each entry in columns 2 and 3. If there is no negative entry in column 3.

The problem table will then show:

1. Values for the total process heat and cooling loads which will be required if maximum energy recovery is achieved.

2. Maximum permissible Table 3.3 for the heater and cooler loads of each subnetwork which must not be exceeded if the final network is to be optimum form an energy recovery point of view.

3.2 Rule –Based Heat Exchanger Network synthesis

3.2.1 Problem Definition

The problem of heat exchanger network synthesis can be described as follows:

A set of a cold streams (i =1, n_c) initially at supply temperature T_i^s and at heat capacity flowrate W_i is to be heated to target T_i^t. Concurrently, a set of hot streams (j =1, n_h) initially at supply temperature T_i^s and at heat capacity flow rate W_j is to be cooled to target temperature T_j^t. Variations in these temperatures and heat capacity flow rates may arise due to real world situations. Hot and cold utilities are available for use. The enthalpy versus temperature relationship is known for all streams. The appropriate physical properties for determining heat transfer characteristics are also given. The objective is to design the optimal network of heat exchangers, coolers and heaters to accomplish the desired temperature changes. Optimal usually means most economic for the capital and utility costs available.

Three major properties of a HEN may be:

1. The maximum energy recovery (MER) or minimum utility usage.

2. The number of heat exchanger units (N_{min}).

3. The minimum approach temperature difference between hot process and cold process streams which is a bottleneck in a design.

3.2.2 Preanalysis

This step determines targets for a network to be designed. The design targets of a network are the maximum energy (MER) and the minimum number of matches. Other targets proposed in replacing the minimum number of units are (1) heat transfer area, (2) capital cost, and (3) number of shells.

The maximum Energy Recovery (MER). MER can be determined by using the temperature-enthalpy diagram (Hohmann, 1971) or by the problem table (Hohmann, 1971; Linnhoff et al., 1982) or by mathematical programming techniques, i.e. the northwest corner algorithm, (Cerda and Westerberg, 1983a). The idea of the first two merges all hot streams into a single composite hot stream and all cold into a single composite cold stream. By shifting the position of the composite cold stream curve along the enthalpy axis to produce a separation between these two curves equal to the specified minimum approach temperature in the temperature-enthalpy diagram, the pinch temperature and the MER can be the obtained.

The Pinch Temperature. The pinch temperature arises in heat exchanger network which require both heating and cooling utilities. It is the point of closest approach, on the temperature scale, of the composite heating and cooling curves as dictated by the network ΔT_{min} or the minimum approach temperature between hot process streams and cold process streams. The pinch temperature divides the network into subnetworks with the requirement that no heat is allowed to transmit through that point in order to achieve MER. In each subnetwork, only one utility (heating or cooling) is required.

The Minimum Number of Matches. The probable minimum number of matches (heat exchangers, heaters and coolers) can be predicted by the following equation (Hohmann, 1971),

$$N_{min} = N_h + N_c + N_{hu} + N_{cu} - 1$$

Where N_h and N_{hu} are the numbers of hot process and utility streams; N_c and N_{hu} are the numbers of process and utility streams. For problems with a pinch this equation should be applied separately to the subnetworks above and below the pinch (Linnhoff et al., 1982).

3.2.3 Heat Exchanger Network Synthesis

Usually, heat exchanger network synthesis is divided into 2 steps:

1. Network targeting.

In the targeting step, the following important properties are determined before the actual network is designed. They are used as the targets for a design.

(1) The maximum energy (or the minimum utilities). The minimum utilities can be calculated by constructing the problem table (Linnhoff and Flower, 1978a). The values depend on the minimum approach temperature, ΔT_{min} on the generation of network configurations is such that, at its higher value, some configuration will be prohibited from appearing.

(2) The minimum number of matches (or units). The minimum number of matches is calculated from

$$N_{match,min} = N_{HotStream} + N_{ColdStream} - 1$$

If a problem is separated by pinch, this equation must be applied separately to each separated problem. If a matching procedure follows the ‘tick-off ’ heuristics, a solution obtained will feature the minimum matches predicted by the above equation. In general, the cost of a minimum matches network solution is close to the minimum capital cost network (Nishida et al.(1997)

2. Network Synthesis

The heuristic approaches find a HEN solution in a sequence of steps. This can be viewed a math operator to map one design state to another. There can be many match operators or in contrast just one operator. Mehta and Fan (1987) use the following condition in testing a math:

For a math at hot end position,

$$\Delta T_{he} \geq \Delta T_{\min}$$

where $\Delta T_{he} = T_h^{supply} - T_c^{target}$, h = hot stream, c = cold stream

For a math at cold end position,

$$\Delta T_{ce} \geq \Delta T_{\min}$$

where $\Delta T_{ce} = T_h^{target} - T_c^{supply}$

Using one operator does not use any heuristic knowledge at all. To make use of heuristics we must discriminate among matches according to criteria or preferences. We can classify matches into several categories and give them different priorities. In this way, heuristics are used.

The HEN heuristics have appeared in the literature over two decades. The following is a summary of published HENS heuristics. The best known one is to make use of the pinch temperature which is called *pinch* heuristics (see Linnhoff and Hindmarsh, 1983). However, it alone cannot solve difficult problems.

3.2.4 Heuristics for HEN Synthesis

Several HEN matching rules with minimum energy and investment costs have been presented (Masso and Rudd, 1969, Ponton and Donalson, 1974 Rathore and Powers, 1975 Linnhoff and Hindmarsh, 1983, Jezowski and Hahne1986, Huang, Metha and fan, 1988, etc.),

The following are heuristics from the literature classified according to the design criteria

The heuristics to minimize the capital cost (the number of heat exchangers):

Heuristic C1. To generate a network featuring the minimum number of heat exchanger units, let each match eliminate at least one of the two streams; a tick-off rule (Hohmann, 1971).

Heuristic C2. Prefer the matches that will leave a residual stream at its cold end for a heating problem, or its hot end for a cooling problem. A match of this type will feature the maximum temperature difference.

Heuristic C3. Prefer matching large heat load streams together. The significance of this rule is that the control problem (a capital cost) of a mach of this type (whether it is implemented by one or many heat exchangers) should be less than that of heating or cooling a large stream with many small streams.

The heuristics to minimize the energy cost (the minimum utility requirement):

Heuristic E1. Divide the problem at the pinch into subproblems, one a heat sink (heating subproblem or hot end problem) and the other a heat source (cooling subproblem or cold end problem), and solve them separately (Linnhoff and Hindmarsh, 1983).

Heuristic E2. Do not transfer heat across the pinch.

Heuristic E3. Do not cool above the pinch.

Heuristic E4. Do not heat below the pinch.

The laws of thermodynamics:

Heuristic T1. In a heating problem, if a supply temperature of a cold stream is less than a target temperature of a hot stream by ΔT_{min} or more and the heat capacity flowrate of a hot stream is less than or equal to the heat capacity flowrate of a cold stream, the match between these two streams is feasible.(Immediately above the pinch tem perature, the heat capacity flowrate of a cold stream must be greater than or equal to that of a hot stream.)

Heuristic T2. In a cooling problem, if a supply temperature of a hot stream is greater than a target temperature of a cold stream by ΔT_{min} or more and the heat capacity flowrate of a hot stream is greater than equal to the heat capacity flowrate of a cold stream, the mach between these two streams is feasible. (Immediately below the pinch temperature, the heat capacity flowrate of a hot stream must be greater than or equal to that of a cold stream.)

Heuristic T3. For a situation different from the above rule, a match fesibility must be determined by checking whether the minimum temperature difference of a match violates the minimum approach , ΔT_{min} , specific by the design.

3.2.5 Math Classification

In order to make use of the heuristics we must classify matches. The following criteria are considered important in this research:

1. Position of a Match

One heuristic prefers a match at the cold end and another prefers a match at the hot end. Pinch heuristics prefers a match at the cold end in a heating subproblem and a match at the hot end in a cooling subproblem. However, there are other possibilities. By using the tick-off heuristic, there are four ways that two streams can match. This leads to the basic four match patterns, (Wongsri, 1990).

2. Heat capacity flowrate (between hot and cold stream).

See Heuristic T.1 and T.2.

3. Heat Load (between hot and cold streams).

The heuristic that concerns heat load state that one must match large heat load hot and cold streams first. This leads to two additional heuristic:

Heuristic N1. For a heating subproblem , a match where the heat load of a cold stream is greater than that of a hot stream should be given higher priority than the other .The reason is that the net heat load heating subproblem is in deficit. The sum of heat loads of cold streams is greater than of hot streams. The purposed match will likely be part of a solution, (Wongsri, 1990).

Heuristic N2. Conversely, we prefer a mach where the heat load of a hot stream is greater than that of a cold in a cooling subproblem, (Wongsri, 1990).

4. Residual Heat Load.

No heuristics for this quantity have thus far appeared in the literature. Two new heuristics are introduced.

For a match in a heating subproblem that satisfies the heat load preference heuristics N.1;

Heuristic N3. We prefer a match where the residual heat load is less than or equal to the minimum heating requirement (Wongsri, 1990).

For a match in a cooling subproblem that satisfies the heat load preference or heuristics N.2:

Heuristic N4. We prefer a match where the residual heat load is less than or equal to the minimum cooling requirement, (Wongsri, 1990).

The reasoning behind the above two heuristics N3 and N4 is that the residual may be matched to a utility stream. One has the possibility of eliminating two streams at once.

3.2.6 Match Patterns

HEN synthesis is usually considered as a combinatorial matching problem. For a HEN in which a design property is regarded as a network property, or an structural property, we need to look beyond the match level to a higher level where such a property exists, e.g. to a match structure or match pattern. Match patterns are the descriptions of the match configuration of two, and possibly more, process streams and their properties that are thermally connected with heat exchangers. Not only the match description, e.g. heat duty of an exchanger and inlet and outlet temperatures is required but also the position of a match, e.g. upstream or downstream, the magnitude of the residual heat load and the heat capacity flow rates between a pair of matched streams.

By using the 'tick off rule' there are four match patterns for a pair of hot and cold streams according to the match position and the length (heat load) of streams. The four patterns are considered to the basic match pattern classes. The members of these classes are the patterns where other configurations and properties are specified. The four match pattern classes are simply called A, B, C and D and are shown in Figure 3.2, 3.3, 3.4 and 3.5 respectively. Any eligible match must belong to one of the four match pattern classes.

Definition 3.1 Class A Match Pattern: The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the cold end of the cold stream. The residual heat load is on the hot portion of the cold stream. (See Figure 3.2)

A match of this class is a first type match at cold end position and the heat load of the cold stream is greater than that of the hot stream. This is a upstream match. For a heating subproblem, a Class A match is favored, because it leaves a cold process stream at the hot end (Heuristic N1) and follows the pinch heuristics. (See Table 3.4)

Definition 3.2 Class B Match Pattern: The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the hot end of the hot stream. The residual heat load is on the cold portion of the hot stream. (See Figure 3.3)

A match of this class is a second type match; a hot end match and the heat load of the hot stream is greater than that of the cold stream. This is an upstream match. For a cooling subproblem, a Class B match is favored, because it leaves a hot process stream at the cold end (Heuristic N2) and also follows the pinch heuristics. (See Table 3.4)

Definition 3.3 Class C Match Pattern: The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the cold end of the hot stream. The residual heat load is on the hot portion of the hot stream. (See Figure 3.4)

A match of this class is a first type match; a cold end match and the heat load of the hot stream is greater than that of the cold stream. This is a downstream match. (See Table 3.5)

Definition 3.4 Class D Match Pattern: The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the hot end of the cold stream. The residual heat load is on the cold portion of the cold stream. (See Figure 3.5)

A match of this class is a second type match; a hot end match and the heat load of the cold stream is greater than that of the hot stream. This is a downstream match. (See Table 3.5)

When the residual heat load in a match pattern is matched to a utility stream, it is closed or completed pattern. Otherwise, it is an open or incomplete pattern. It can be seen that if the heat load of the residual stream is less than the minimum heating or cooling requirement then the chances that the match pattern will be matched to a utility stream is high. So we give a match pattern which its residual less than the minimum heating or cooling requirement a high priority in match pattern.

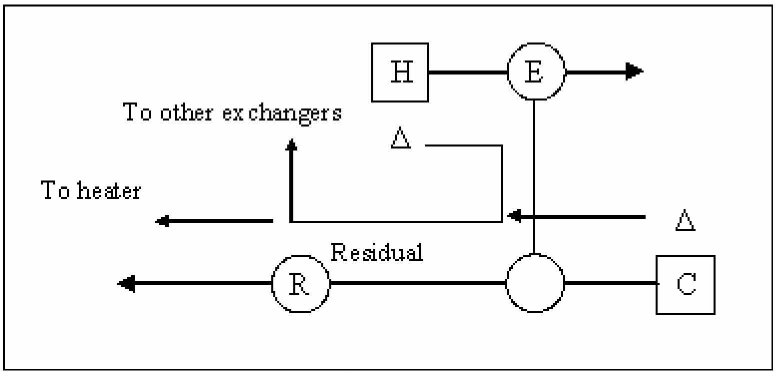


Figure 3.2 Class A Match Patter

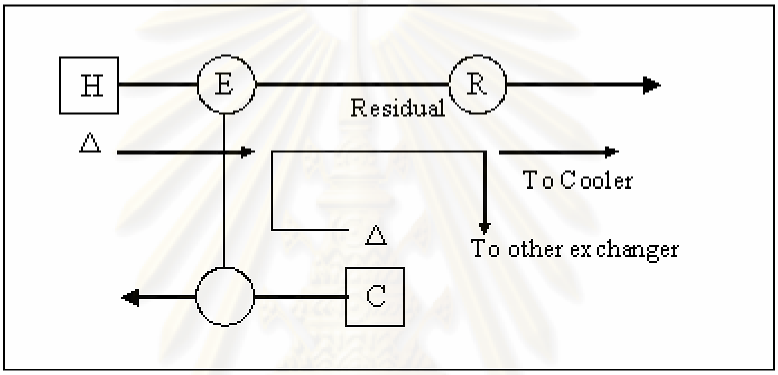


Figure 3.3 Class B Match Pattern

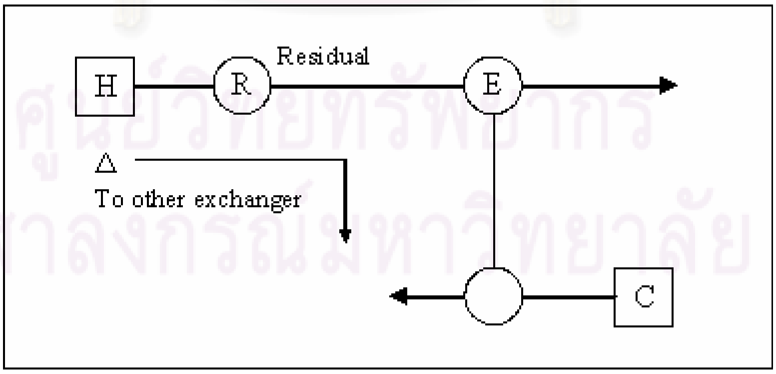


Figure 3.4 Class C Match Pattern

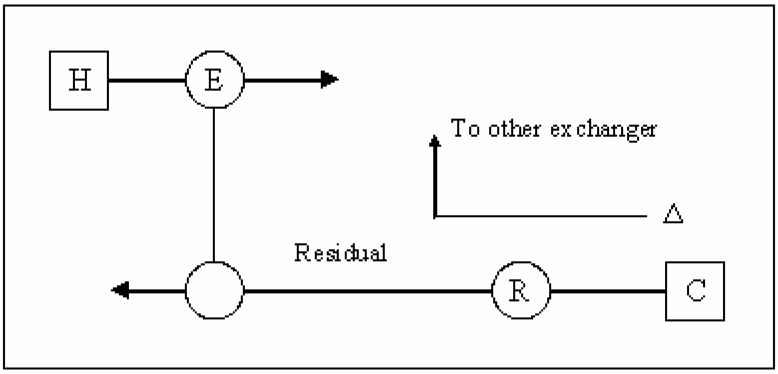
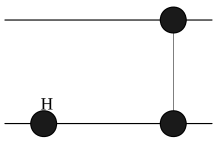
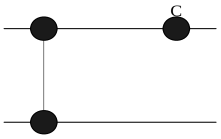
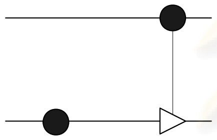
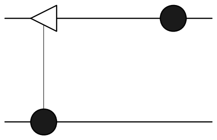
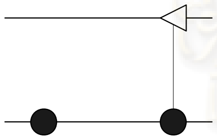
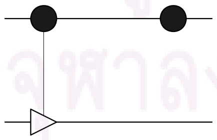


Figure 3.5 Class D Match Pattern

A match of Class A or Class C will leave a residual at the hot end, while a match of ClassB or D will leave a residual at the cold end. Heuristics N.3 and N.4 will be use heuristics to further subclassify matches of Class A and B into matches of high priority.

Table 3.4 Match Pattern Operators of Class A and B

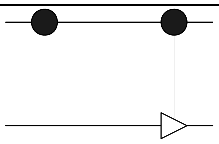
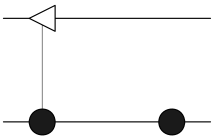
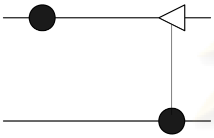
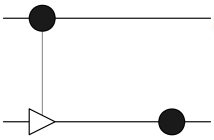
Match Operators	Conditions	Actions
 Pattern AH	$T_H^s{}^{*} \geq T_C^t{}^{**}$ $L_H \leq L_C$ $T_H^s \geq T_C^s + L_H\ W_C^{-1}$ $L_C - L_H \leq Q_{\min}^{heating}$	Match H and C Status of H $\Leftarrow$ Matched*** $T_C^s \Leftarrow T_C^s + L_H\ W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern BK	$T_H^s \geq T_C^t$ $L_C \leq L_H$ $T_C^s \leq T_H^s - L_C\ W_H^{-1}$ $L_H - L_C \leq Q_{\min}^{cooling}$	Match H and C Status of C $\Leftarrow$ Matched $T_H^s \Leftarrow T_H^s - L_C\ W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern A[H]	$T_H^t \geq T_C^s$ $L_H \leq L_C$ $W_C \geq W_H$	Match H and C Status of H $\Leftarrow$ Matched $T_C^s \Leftarrow T_C^s + L_H\ W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern B[C]	$T_H^s \geq T_C^t$ $L_C \leq L_H$ $W_C \leq W_H$	Match H and C Status of C $\Leftarrow$ Matched $T_H^s \Leftarrow T_H^s - L_C\ W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern A[C]	$T_H^t \geq T_C^s$ $L_H \leq L_C$ $W_C < W_H$ $T_H^s \geq T_C^s + L_H\ W_C^{-1}$	Match H and C Status of H $\Leftarrow$ Matched $T_C^s \Leftarrow T_C^s + L_H\ W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern B[H]	$T_H^s \geq T_C^t$ $L_C \leq L_H$ $W_H < W_C$ $T_C^s \leq T_H^s - L_C\ W_H^{-1}$	Match H and C Status of C $\Leftarrow$ Matched $T_H^s \Leftarrow T_H^s - L_C\ W_H^{-1}$ $L_H \Leftarrow L_H - L_C$

* T^t=target temp, T^s=supply temp, W=heat capacity flowrate, L, Q=heat load.

** Cold stream temperatures are shifted up by ΔT_{min}.

*** There are two status of process streams, ‘active’ and ‘matched’. This will exclude this stream from a set of process streams to be selected next.

Table 3.5 Match Pattern Operators of Class C and D

Match Operators	Conditions	Actions
 Pattern C[H]	$T_H^t \ge T_C^s$ $L_H > L_C$ $W_H \le W_C$	Match H and C Status of C ⇐ Matched $T_H^t \Leftarrow T_H^t - L_C\, W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern D[C]	$T_H^s \ge T_C^t$ $L_H < L_C$ $W_H \ge W_C$	Match H and C Status of H ⇐ Matched $T_C^t \Leftarrow T_C^t + L_H\, W_C^{-1}$ $L_C \Leftarrow L_C - L_H$
 Pattern C[C]	$T_H^t \ge T_C^s$ $L_H > L_C$ $W_C < W_H$ $T_C^t \le T_H^t + L_C\, W_H^{-1}$	Match H and C Status of C ⇐ Matched $T_H^t \Leftarrow T_H^t - L_C\, W_H^{-1}$ $L_H \Leftarrow L_H - L_C$
 Pattern D[H]	$T_H^s \ge T_C^t$ $L_H \le L_C$ $W_H < W_C$ $T_H^t \ge T_C^t - L_H\, W_C^{-1}$	Match H and C Status of H ⇐ Matched $T_C^t \Leftarrow T_C^t + L_H\, W_C^{-1}$ $L_C \Leftarrow L_C - L_H$

3.2.7 Disturbance Propagation Design Method

In order for a stream to be resilient with a specified disturbance load, the disturbance load must be transferred to heat sinks or heat sources within the network. With the use of the heuristic: To generate a heat exchanger network featuring the minimum number of heat transfer units, let each match eliminate at lease one of the two streams.

We can see that in a match of two heat load variable streams, the variation in heat load of the smaller stream S1 will cause a variation to the residual of the larger stream S2 by the same degree: in effect the disturbance load of S1 is shifted to the residual of S2. If the residual stream S2 is matched to S3 which has larger heat load, the same situation will happen. The combined disturbance load of S1 and S2 will cause the variation in the heat load to the residual S3. Hence, it is easy to see that the disturbance load in residual S3 is the combination of its own disturbance load and those obtained from S1 and S2. Or, if S2 is matched to a smaller heat load stream S4, the new disturbance load of residual S2 will be the sum of the disturbance loads of S1 and S4. Form this observation, in order to be resilient, a smaller process stream with specified disturbance load must be matched to a larger stream that can tolerate its disturbance. In other words, the propagated disturbance will not overshoot the target temperature of the larger process stream.

However, the amount of disturbance load that can be shifted from one stream to another depends upon the type of match patterns and the residual heat load. Hence, in design we must choose a pattern that yields the maximum resiliency. We can state that the resiliency requirement for a match pattern selection is that the entire disturbance load from a smaller heat load stream must be tolerated by a residual stream. Otherwise, the target temperature of the smaller stream will fluctuate by the unshifted disturbance. Of course, the propagated disturbance will be finally handled by utility exchangers. In short, the minimum heat load value of a larger stream must be less than a maximum heat load value of a smaller stream.

By choosing the minimum heat load condition for the design, the new input temperature of a residual stream to its design condition according to the propagated disturbance. The propagated disturbance will proportionally cause more temperature variation in the residual stream and the range of temperature variation of the residual stream will be larger than its original range.

Definition 3.5 Propagated Disturbance. The propagated disturbance of a stream is the disturbance caused by a variation in heat load of ‘up-path’ streams to which such a stream is matched. Only a residual stream will have a propagated disturbance. The new disturbance load of a residual stream will be the sum of its own disturbance (if any) and the propagated disturbance. See Figure 3.6 and 3.7.

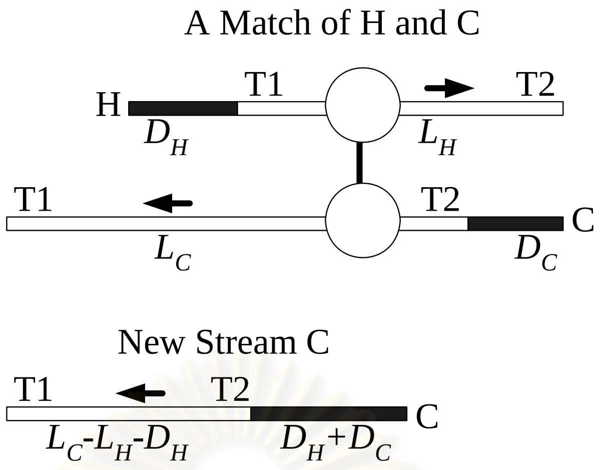


Figure 3.6 A Concept of Propagated Disturbance

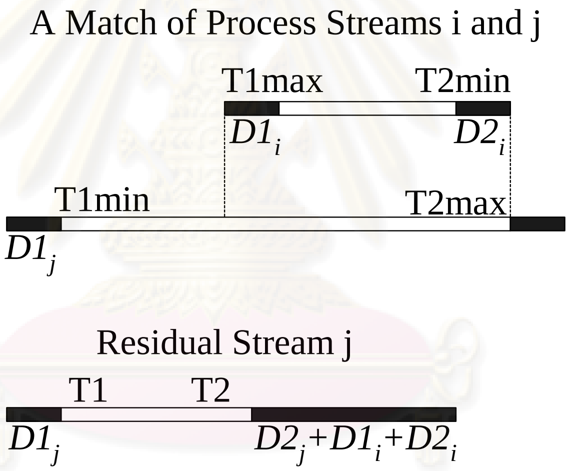


Figure 3.7 A General Concept of Propagated Disturbance

Hence, a stream with no original variation in heat load will be subjected to variation in heat load if it is matched to a stream with disturbance. Another design consideration is that the disturbance load travel path should be as short as possible, i.e. the lease number of streams involved. Otherwise, the accumulated disturbance will be at high level. From the control point of view, it is difficult to achieve good control if the order of the process and the transportation lag are high. From the design viewpoint, are may not find heat sinks or sources that can handle the large amount of propagated disturbance. (Wongsri, 1990).

3.2.8 Synthesis Procedure

A procedure of HEN synthesis by using math operators and a notion of a design state can be carried in step as follow:

1. Push the match operators to a stack in proper order. This is a beginning of a new state.

2. While there is an operator on a stack.

   (a) Pop a match operator form a stack to operate on process streams.

   (b) If a match is found, exclude matched streams from a set of process stream. Change the condition of residual streams. Include the residual streams in to a set of process streams. Go to a new design state (the first step)

3. If there are only hot or cold process streams left in the set of stream, a solution is found. If there are other solutions, they can be found by backtracking to the previous states to try the unused operators in those states

4. If no matches is found in a current design state, back track to a previous stare to try an available operator on the stack of that state. (Go to Step 2 in the previous loop.) It is a recursive procedure here. If a math still could not be found, backtrack again to the more previous

The above sequences represent a loop of one design state. A total generation procedure a loop composing of these sequences.

3.3 Plantwide Control

A typical chemical plant flowsheet has a mixture of multiple units connected both in series and parallel that consist of reaction sections, separation sections and heat exchanger network. So Plantwide Process Control involves the system and strategies required to control entire plant consisting of many interconnected unit operations.

3.3.1 Integrated Process

Figure 3.8 shows integrated process flowsheet. Three basic features of integrated chemical process lie at the root of our need to consider the entire plant’s control system: the effect of material recycle, the effect of energy integration, and the need to account for chemical component inventories.

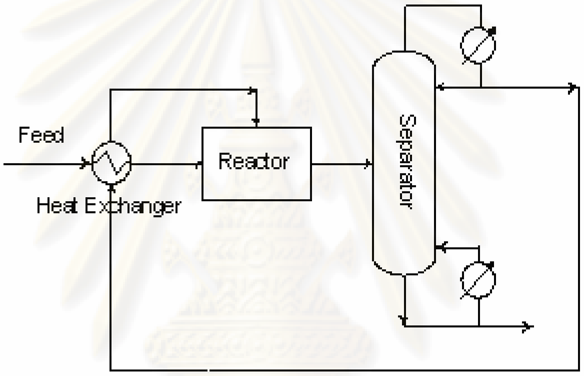


Figure 3.8 Integrated Process flowsheet

3.3.1.1 Material recycle

Material is recycled for six basic and important reasons.

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.

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.

Improve yields.

In reaction system such as $A \rightarrow B \rightarrow C$, 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.

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.

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.

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.

3.3.1.2 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.

3.3.1.3 Chemical component inventories

In chemical processes can characterize a plant’s chemical species into three types: reactants, products, and inerts. 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 units, chemical component balancing is not a problem because exit streams from the unit automatically adjust their flows and composition. However, when 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.

3.3.2 Effects of Recycle

Most real processes contain recycle streams. In this case the plantwide control problem becomes much more complex. Two basic effect of recycle is: 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 time constants of the individual units. Recycle leads to the “snowball” effect. A small change in throughput or feed composition can lead to a large change in steady-state recycle stream flowrates.

Snowball effect

Snowball effect is high sensitivity of the recycle flowrates to small disturbances. 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. It is a steady-state phenomenon but it does have dynamic implications for disturbance propagation and for inventory control.

The large swings in recycle flowrates are undesirable in plant because they can overload the capacity of separation section or move the separation section into a flow region below its minimum turndown. Therefore it is important to select a plantwide control structure that avoids this effect.

3.3.3 Plantwide Control Design Procedures

In plantwide control design procedure satisfies the two fundamental chemical engineering principles, namely the overall conservation of energy and mass. Additionally, the procedure accounts for nonconserved entities within a plant such as chemical components (produced and consumed) and entropy (produced).

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 run in automatic, not manual, requiring minimal operator attention

5. Rapid rate and product quality transitions.

6. Zero unexpected environmental releases.

3.3.3.1 Basic Concepts of Plantwide Control

1. *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.

2. *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.

3. *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.

4. *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.

(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.

5. *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).

6. *Shinskey schemes*

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

7. *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:

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

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

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.

3.3.3.2 Step of Plantwide Process Control Design Procedure

Step1: 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 setpoint. 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. minimize 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 deadtimes 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 variables 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 nonreactive 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. Crytallizers require manipulation of refrigeration load to control temperature. Oxygen concentration in strack gas form a furnace is controlled to prevent excess fuel usage. Liquid sovent feed flow to an absorber is controlled as some ratio to the gas feed.

Step 9: Optimize economics or improve dynamic controllability

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

After satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and 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.

3.4 Control of process-to-process exchanger

Process-to-process (P/P) exchangers are used for heat recover within a process. We can control the two exit temperatures provided we can independently manipulate the two inlet flowrates. However, these flowrates are normally unavailable for us to manipulate and we therefore give up two degrees of freedom fairly easily. It is possible to oversize the P/P exchanger and provides a controlled bypass around it as in Fig 3.9.a. It is possible to combine the P/P exchanger with a utility exchanger as in Fig 3.9.b.

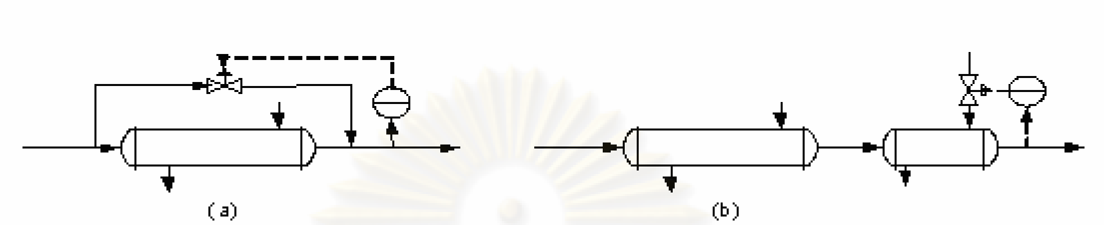


Figure 3.9 Control of P/P heat exchangers: (a) use of bypass; (b) use of auxiliary utility exchanger.

3.4.1 Use of Bypass Control

When the bypass method is used for unit operation control, we have several choices about the bypass location and the control point. Figure 3.3 shows the most common alternatives. For choosing the best option, it depends on how we define the best. Design consideration might suggest, we measure and bypass on the cold side since it is typically less expensive to install a measurement device and a control valve for cold service than it is for high-temperature service. Cost consideration would also suggest a small bypass flow to minimize the exchanger and control valve sizes.

From a control standpoint, we should measure the most important stream, regardless of temperature, and bypass on the same side as well we control (see Figure 3.10 a and c). This minimizes the effects of exchanger dynamics in the loop. We should also want to bypass a large fraction of the controlled stream since it improves the control range. This requires a large heat exchanger. There are several general heuristic guidelines for heat exchanger bypass streams. We typically want to bypass the flow of the stream whose temperature we want to control. The bypass should be about 5 to 10 percent of the flow to be able to handle disturbances. Finally, we must carefully consider the fluid mechanics of the bypass design for the pressure drops through the control valves and heat exchanger.

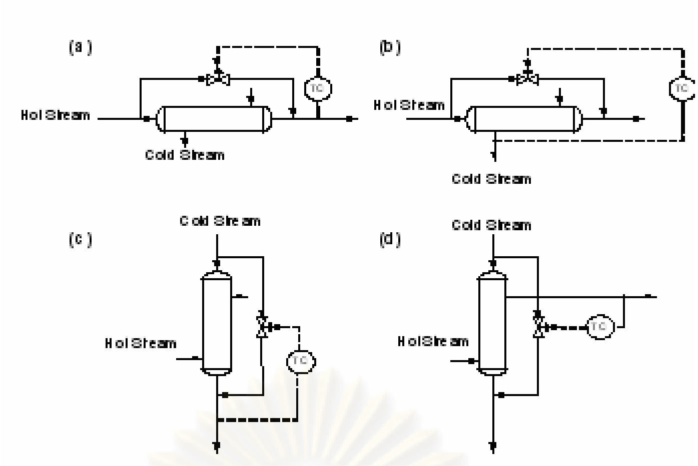


Figure 3.10 Bypass control of process-to process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing hot stream.

3.4.2 Use of Auxiliary Exchangers

When the P/P exchanger is combined with a utility exchanger, we also have a few design decisions to make. We must first establish the relative sizes between the recovery and the utility exchanger large and the utility exchanger small. This gives us the most heat recovery, and it is also the least expensive alternative from an investment standpoint. However, a narrow control range and the inability to reject disturbance make this choice this choice the least desirable from a control standpoint.

Next, we must decide how to combine the utility exchanger with the P/P exchanger. This could be done either in a series or parallel arrangement. Physical implementation issues may dictate this choice but it could affect controllability. Finally, we have to design hoe to control the utility exchanger for best overall control performance. Consider a distillation column that uses a large amount of high-pressure stream in its thermo siphon reboiler. To reduce operating costs we would like to heat-integrate this column with the reactor. A practical way of suggested. We can then use some or all of this stream to help reboil the column by condensing the stream in the tubes of a stab-in reboiler. However, the total heat from the reactor may not be enough to reboil the column, so the remaining heat must come from the thermo siphon reboiler that noe serves as an auxiliary reboiler. The column tray temperature controller would manipulate the stream to the thermo siphon reboiler.

3.5 Cascade Control

One of the most useful concepts in advanced control is cascade control. A cascade control structure has two feedback controllers with the output of the primary (or master) controller changing the setpoint of the secondary (or slave) controller. The output of the secondary goes to the slave.

There are two purposes for cascade control: (1) to eliminate the effects of some disturbances, and (2) to improve the dynamic performance of the control loop.

To illustrate the disturbance rejection effect, consider the distillation column reboiler. Suppose the steam supply pressure increases. The pressure drop over the control valve will be larger, so the steam flow rate will increase. With the single-loop temperature controller, no correction will be made until the higher steam flow rate increases the vapor boilup and the higher vapor rate begins to rise the temperature on tray. Thus the whole system is disturbed by a supply-steam pressure change.

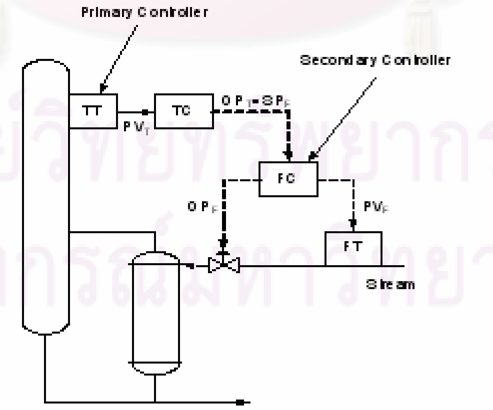


Figure 3.11 Cascade control in distillation-column-reboiler

With the cascade control system, the steam flow controller will immediately see the increase in steam flow and will pinch back on the steam valve to return the steam flow rate to its setpoint. Thus the reboiler and the column are only slightly affected by the steam supply-pressure disturbance.

Figure 3.11 shows another common system where cascade control is used. The reactor temperature controller is the primary controller; the jacket temperature controller is the secondary controller. The reactor temperature control is isolated by the cascade system from disturbance in cooling-water inlet temperature and supply pressure.

CHAPTER IV

ISOMERIZATION PROCESS

4.1 Process Description

The isomerization process is quite important in the petroleum industry because isobutane is usually more valuable as a chemical feedstock than normal butane. The typical amount of iC4 contained in crude oil and produced in refinery operations such as catalytic cracking is sometime not enough to satisfy the demand. On the other hand the supply of nC4 sometimes exceeds the demand, particularly in the summer when less nC4 can be blended into gasoline because of vapor pressure limitations.

Some of the many uses of isobutane include the production of high-octane gasoline blending components by reacting it with various olefins in alkylation processes and the production of propylene oxide and tertiary butane alcohol.

The process consists of a reactor, two distillation columns, and liquid recycle stream. There are four components to consider.

Figure 4.1 shows the flowsheet of the isomerization process to convert normal butane (nC4) into isobutane (iC4).

$$nC4 \longrightarrow iC4$$

The reaction of nC4 to iC4 occurs in the vapor phase and is run at elevated temperatures (400^{o}F) and pressures (600 psia). The reaction of is exothermic (heat of reaction -3600 Btu/lb.mol), so there is a temperature rise as the process stream flows through the adiabatic tubular reactor. Following heat exchanger with the reactor inlet stream and condensation with cooling water, the reactor effluent is introduced a large distillation column that separates the C4's. The iso/normal separation is difficult because of the similar relative volatilities, so many tray (50) and high reflux ratio (7) are required. For the design case considered, this column ends up being 16 feet in dimeter. This column is called a deisobutanizer (DIB)

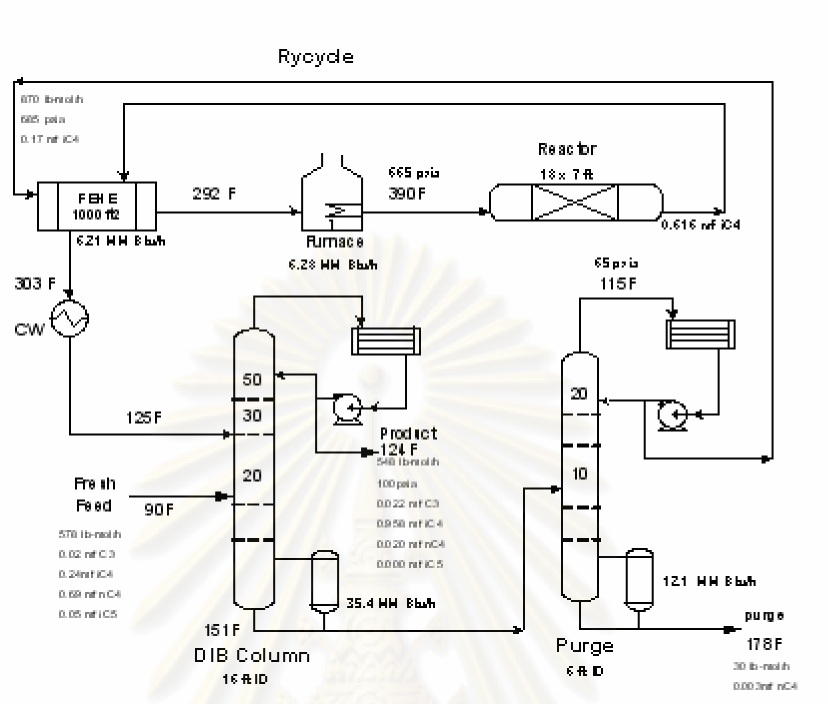


Figure 4.1 The Butane Isoomerization Process (Plantwide Process Control, 1998)

The fresh feed stream is a mixture of nC4 and iC4 (with some propane and isopentane impurities). It is also introduced into the column, not directly into the reactor. It is fed at a lower tray in the column than the reactor effluent stream because the concentration of iC4 and all of the C3 in the fresh feed before sending the nC4 to the reactor from the recycle stream. The ratio of the recycle flow to the fresh feed flow is about 1:2. The DIB column operates at 100 psia so that cooling water can be used in the condenser (reflux drum temperature is 124 ^oF). The base temperature is 150 ^oF, so low-pressure stream can be used.

The distillate product from the DIB is the isobutane product. It has a specification of 2 mol% nC4. Since the fresh feed contains some propane, there is also some propane in the distillate product. All of the propane in the feed leaves the process in the distillate stream.

The bottoms from the DIB contain most of the nC4, along with some iC4 impurity and all of the heavy isopentane impurity. Since this heavy component will build up in the process unless it is removed, a second distillation column is used to purge out a small stream that contains the isopentane. Some nC4 is lost in this purge stream. The purge column has 20 trays and is 6 ft in diameter. The distillate product from the second column is the recycle stream to the reactor, which is pumped up to the required pressure and sent through a feed-effluent heat exchanger and a furnace before entering the reactor in the vapor phase.

The numerical case studied is derived from a flowsheet given in Stanford Research Institute Report 91, “Isomerization of Paraffins for Gasoline”. Since no kinetic information is given in this in this report, only reactor inlet and exit conditions. We consider that the reactor is irreversible. The activation energy of 30,000 Btu/lb.mol is used, and the pre-exponential factor is adjusted to give the same conversation reported in the SRI report. Table A1 gives stream data for the butane isomerization process. Table A2 and Table A3 list the process parameter values.

4.2 Design of heat exchanger networks

At this point, the heat exchanger network design method provide by Wongsri (1990) is used to design the heat exchanger networks for Butane Isomerization process. The design procedures and definitions from previous chapters will be methods to design and compare with the preliminary stage of a process design without energy integration. The Problem Table Method is applied to find pinch temperature and reach maximum energy recovery (MER). The cost estimated will be consequence to compare and choose the best network that more optimal for the Butane Isomerization process. The information for design is shown in the following Table 4.1

Table 4.1 The information of Butane Isomerization Process

Stream Name	Tin(^{o}F)	Tout(^{o}F)	W	Duty (BTU/hr)
H1: Reactor Product Stream (RPS)	424.10	125.00	4.46E+04	1.34E+07
H2: DIB Column Condenser (q)	122.40	121.20	2.47E+07	2.97E+07
H3: Purge Column Condenser (qq)	110.38	108.92	9.09E+06	1.33E+07
C1: Reactor Feed Stream (RFS)	117.67	390.10	4.57E+04	1.25E+07
C2: DIB Column Reboiler (q)	142.08	143.87	1.73E+07	3.10E+07
C3: Purge Column Reboiler (qq)	171.50	173.76	5.39E+06	1.22E+07

4.2.1 HEN Base Case

According to table 4.1, it can be simply translated to a heat exchanger network for Butane Isomerization Process (Base Case) in Figure 4.2.

There are two streams in the network. We do not find Pinch temperature using Problem table method.

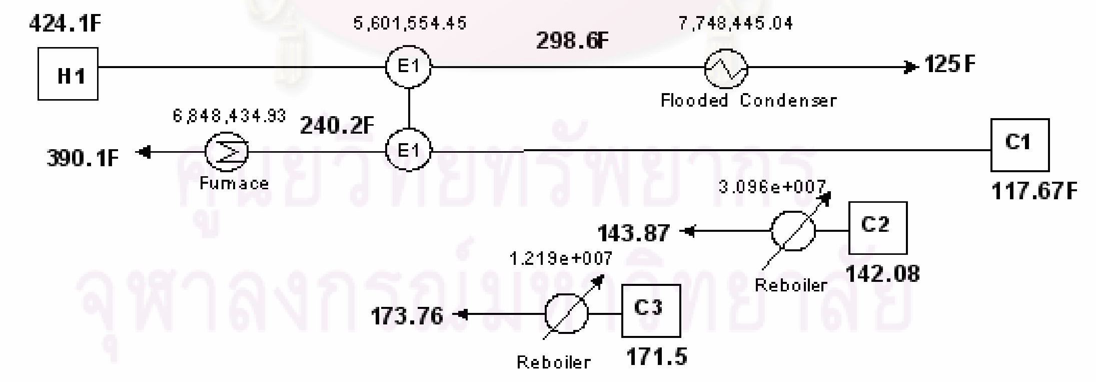


Figure 4.2 The heat exchanger network, Base Case

4.2.2 RHEN Alternative 1

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.3. At the minimum heat load condition, the pinch temperature occurs at 137.67/117.67 ^{o}F. The minimum utility requirements have been predicted 30.626 x 10^{6} Btu/hr of hot utilities and 5.655 x 10^{5} Btu/hr of cold utilities.

However, RHEN1 is modified and to meet the ΔT_{min} constraint, so the utilities requirements are 43.901×10^{6} Btu/hr of hot utilities and 16.514×10^{5} Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.2. Figure 4.3 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 1. In our case as shown in Fig. 4.3, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

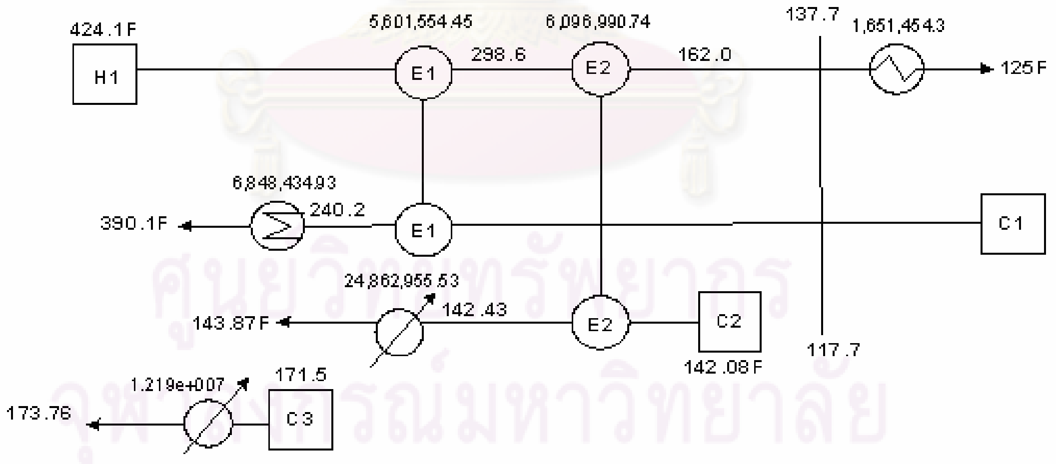


Figure 4.3 The resilient heat exchanger network alternative 1, RHEN-1

Table 4.2 Process stream data for alternative 1

Stream	W (Btu/hr-°F)	T_{supply} (°F)			T_{target} (°F)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.10	444.10	404.10	125.00
C1	45,699.80	117.67	137.67	117.67	390.10
C2	17,296,089.39	142.08	143.08	141.08	143.87

Synthesis Table for Cold End of RHEN-1							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-1							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.10	137.67	892,678	892,678	Matched with C1
C1	11,536,000.51	45,699.80	137.67	390.1	913,996	0	Matched with H1
C2	13,663,910.62	17,296,089.39	143.08	143.87	17,296,089.39	0	
b) State 2	Selected match pattern AH *						
H1	5,601,554.45	44,633.90	404.10	278.60	892,678	0	Matched with C2
C1	5,245,744.99	45,699.80	275..31	390.10	0	21,318	To Heater
C2	13,663,910.62	17,296,089.39	143.08	143.87	17,296,089.39	0	Matched with H1
c) State 3							
H1							
C2	7,373,655.09	17,296,089.39	143.51	143.87	18,188,767.39	0	To Heater

Table 4.3 Problem table for alternative 1

w			T hot	T cold	DT	Sum W	Require	Interval (H)	Casecade	Sum Interval
h1	c1	c2								
0.00	0.00	0.00	424.10	404.10			Qh			
44,633.90	0.00	0.00	410.10	390.10	14.00	44,633.90	30,625,508.55	624,874.60	31,250,383.15	624,874.60
44,633.90	45,699.80	0.00	163.87	143.87	246.23	-1,065.90	31,250,383.15	-262,456.56	30,987,926.59	362,418.04
44,633.90	45,699.80	17,296,089.39	162.08	142.08	1.79	-17,297,155.29	30,987,926.59	-30,961,907.97	26,018.62	-30,599,489.93
44,633.90	45,699.80	0.00	137.67	117.67	24.41	-1,065.90	26,018.62	-26,018.62	0.00	-30,625,508.55
44,633.90	0.00	0.00	125.00	105.00	12.67	44,633.90	0.00	565,511.51	565,511.51	-30,059,997.03
									Qc	

4.2.3 RHEN Alternative 2

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.5. At the minimum heat load condition, the pinch temperature occurs at 137.67/117.67 ^{o}F. The minimum utility requirements have been predicted 30.626 x 10^{6} Btu/hr of hot utilities and 5.655 x 10^{5} Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.4. Figure 4.4 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 2. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

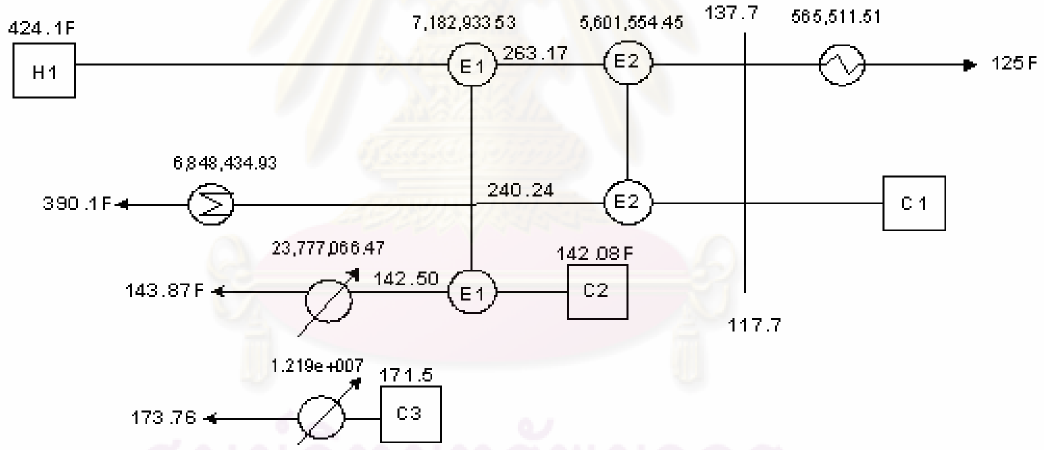


Figure 4.4 The resilient heat exchanger network alternative 2, RHEN-2

Table 4.4 Process stream data for alternative 2

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1
C2	17,296,089.39	142.08	143.08	141.08	143.87

Synthesis Table for Cold End of RHEN-2

Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-2

Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.10	137.67	892,678	892,678	Matched with C1
C1	11,536,000.51	45,699.80	137.67	390.1	913,996	0	Matched with H1
C2	13,663,910.62	17,296,089.39	143.08	143.87	17,296,089.39	0	
b) State 2	Selected match pattern AH*						
H1	6,290,255.53	44,633.90	404.10	263.17	892,678	0	Matched with C2
C1	5,934,446.06	45,699.80	260.24	390.10	0	21,318	To Heater
C2	13,663,910.62	17,296,089.39	143.08	143.87	17,296,089.39	0	Matched with H1
c) State 3							
H1							
C2	7,373,655.09	17,296,089.39	143.51	143.87	18,188,767.39	0	To Heater

Table 4.5 Problem table for alternative 2

w			T hot	T cold	DT	Sum W	Require	Interval (H)	Casecade	Sum Interval
h1	c1	c2								
0.00	0.00	0.00	424.10	404.10			Qh			
44,633.90	0.00	0.00	410.10	390.10	14.00	44,633.90	30,625,508.55	624,874.60	31,250,383.15	624,874.60
44,633.90	45,699.80	0.00	163.87	143.87	246.23	-1,065.90	31,250,383.15	-262,456.56	30,987,926.59	362,418.04
44,633.90	45,699.80	17,296,089.39	162.08	142.08	1.79	-17,297,155.29	30,987,926.59	-30,961,907.97	26,018.62	-30,599,489.93
44,633.90	45,699.80	0.00	137.67	117.67	24.41	-1,065.90	26,018.62	-26,018.62	0.00	-30,625,508.55
44,633.90	0.00	0.00	125.00	105.00	12.67	44,633.90	0.00	565,511.51	565,511.51	-30,059,997.03
									Qc	

4.2.4 RHEN Alternative 3

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.7. At the minimum heat load condition, the pinch temperature occurs at 137.67/117.67 ^{o}F. The minimum utility requirements have been predicted 11.856 x 10^{6} Btu/hr of hot utilities and 5.655 x 10^{5} Btu/hr of cold utilities.

However, RHEN3 is modified and to meet the ΔT_{min} constraint, so the utilities requirements are 45.218×10^{6} Btu/hr of hot utilities and 29.682×10^{5} Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.6. Figure 4.5 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 3. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

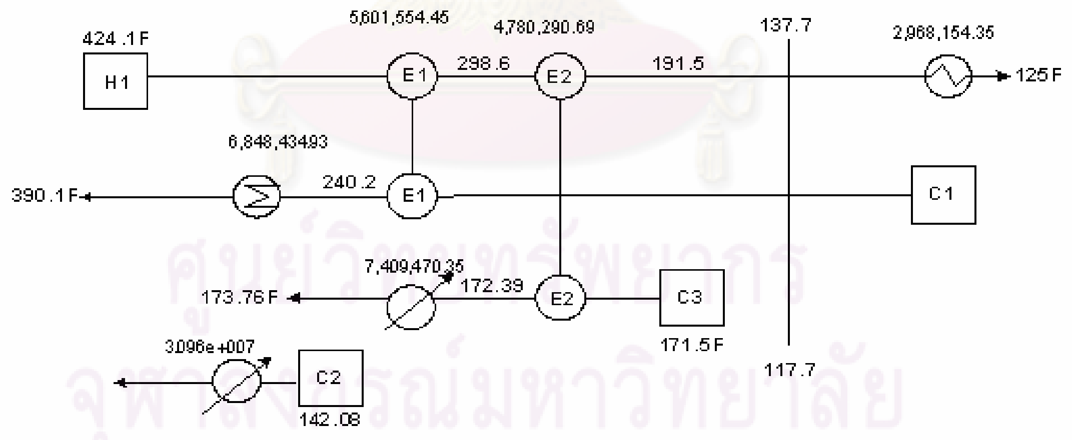


Figure 4.5 The resilient heat exchanger network alternative 3, RHEN-3

Table 4.6 Process stream data for alternative 3

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1
C3	5,393,805.31	171.5	172.5	170.5	173.76

Synthesis Table for Cold End of RHEN-3							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-3							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.10	137.67	892,678	892,678	Matched with C1
C1	11,536,000.51	45,699.80	137.67	390.1	913,996	0	Matched with H1
C3	6,796,194.69	5,393,805.31	172.50	173.76	5,393,805.31	0	
b) State2	Selected match pattern AH *						
H1	5,601,554.45	44,633.90	404.10	278.60	892,678	0	Matched with C3
C1	5,245,744.99	45,699.80	275.31	390.10	0	21,318	To Heater
C3	6,796,194.69	5,393,805.31	172.50	173.76	5,393,805.31	0	Matched with H1
d) State 4							
H1							
C3	505,939.16	5,393,805.31	172.59	173.76	6,286,483.31	0	To Heater

Table 4.7 Problem table for alternative 3

w			T hot	T cold	DT	Sum W	Require	Interval (H)	Casecade	Sum Interval
h1	c1	c3								
0.00	0.00	0.00	424.10	404.10			Qh			
44,633.90	0.00	0.00	410.10	390.10	14.00	44,633.90	11,855,508.54	624,874.60	12,480,383.14	624,874.60
44,633.90	45,699.80	0.00	193.76	173.76	216.34	-1,065.90	12,480,383.14	-230,596.81	12,249,786.33	394,277.79
44,633.90	45,699.80	5,393,805.31	191.50	171.50	2.26	-5,394,871.21	12,249,786.33	-12,192,408.93	57,377.40	-11,798,131.14
44,633.90	45,699.80	0.00	137.67	117.67	53.83	-1,065.90	57,377.40	-57,377.40	0.00	-11,855,508.54
44,633.90	0.00	0.00	125.00	105.00	12.67	44,633.90	0.00	565,511.51	565,511.51	-11,289,997.02
									Qc	

4.2.5 RHEN Alternative 4

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.9. At the minimum heat load condition, the pinch temperature occurs at 137.67/117.67 ^{o}F. The minimum utility requirements have been predicted 11.856 x 10^{6} Btu/hr of hot utilities and 5.655 x 10^{5} Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.8. Figure 4.6 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 3. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

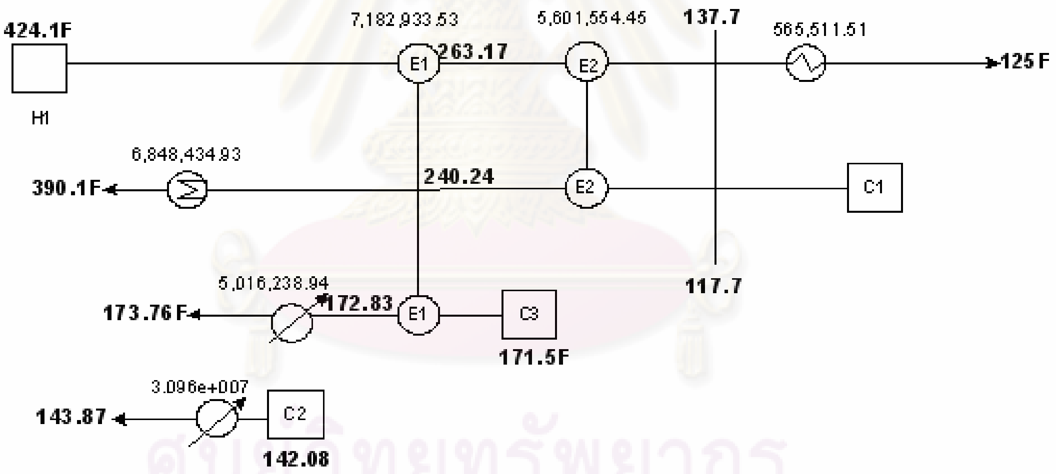


Figure 4.6 The resilient heat exchanger network alternative 4, RHEN-4

Table 4.8 Process stream data for alternative 4

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1
C3	5,393,805.31	171.5	172.5	170.5	173.76

Synthesis Table for Cold End of RHEN-4							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-4							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.10	137.67	892,678	892,678	Matched with C1
C1	11,536,000.51	45,699.80	137.67	390.1	913,996	0	Matched with H1
C3	6,796,194.69	5,393,805.31	172.50	173.76	5,393,805.31	0	
b) State 2	Selected match pattern AH *						
H1	6,290,255.53	44,633.90	404.10	263.17	892,678	0	Matched with C3
C1	5,934,446.06	45,699.80	260.24	390.10	0	21,318	To Heater
C3	6,796,194.69	5,393,805.31	172.50	173.76	5,393,805.31	0	Matched with H1
c) State 3							
H1							
C3	505,939.16	5,393,805.31	172.59	173.76	6,286,483.31	0	To Heater

Table 4.9 Problem table for alternative 4

w										
h1	c1	c3	T hot	T cold	DT	Sum W	Require	Interval (H)	Casecade	Sum Interval
0.00	0.00	0.00	424.10	404.10			Qh			
44,633.90	0.00	0.00	410.10	390.10	14.00	44,633.90	11,855,508.54	624,874.60	12,480,383.14	624,874.60
44,633.90	45,699.80	0.00	193.76	173.76	216.34	-1,065.90	12,480,383.14	-230,596.81	12,249,786.33	394,277.79
44,633.90	45,699.80	5,393,805.31	191.50	171.50	2.26	-5,394,871.21	12,249,786.33	-12,192,408.93	57,377.40	-11,798,131.14
44,633.90	45,699.80	0.00	137.67	117.67	53.83	-1,065.90	57,377.40	-57,377.40	0.00	-11,855,508.54
44,633.90	0.00	0.00	125.00	105.00	12.67	44,633.90	0.00	565,511.51	565,511.51	-11,289,997.02
									Qc	

4.2.6 RHEN Alternative 5

There are two streams in the network. We can find Pinch temperature using Problem table method as shown in Table 4.11. At the minimum heat load condition, the pinch temperature occurs at 162.08/142.08^{o}F. The minimum utility requirements have been predicted 19.265 x 10^{6} Btu/hr of hot utilities and 1.655 x 10^{6} Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.10. Figure 4.7 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 5. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

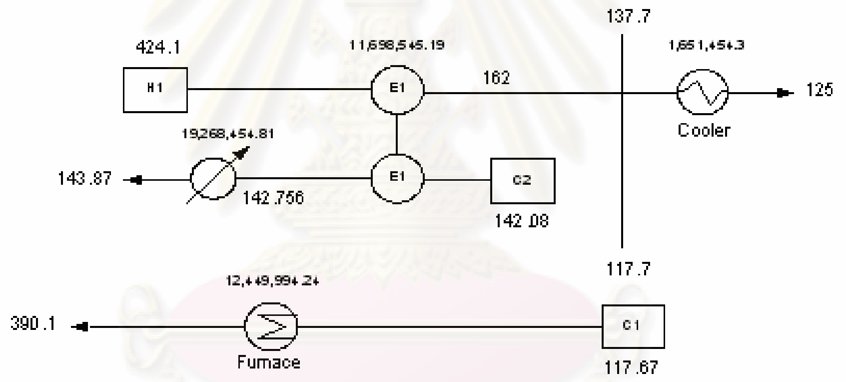


Figure 4.7 The resilient heat exchanger network alternative 5, RHEN-5

Table 4.10 Process stream data for alternative 5

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1
C2	17,296,089.39	142.08	143.08	141.08	143.87

Synthesis Table for Cold End of RHEN-5							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-5							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.1	137.67	892,678.80	892,678.80	Selected
C2	13,663,910.62	17,296,089.39	143.87	143.08	0	34,592,178.78	Selected
b) State 2	Selected match pattern AH *						
H1	1,169,408.18	44,633.90	163.87	137.67		892,678	Selected
C2	2,048,830.82	17,296,089.39	143.87	143.75		35,484,856.78	To heater

Table 4.11 Problem table for alternative 5

w		T hot	T cold	DT	Sum W	Require	Interval (H)	Casecade	Sum Interval
h1	c2								
0.00	0.00	424.10	404.10			Qh			
44,633.90	0.00	163.87	143.87	260.23	44,633.90	19,265,025.53	11,615,079.80	30,880,105.33	11,615,079.80
44,633.90	17,296,089.39	162.08	142.08	1.79	-17,251,455.49	30,880,105.33	-30,880,105.33	0.00	-19,265,025.53
44,633.90	0.00	137.67	117.67	24.41	44,633.90	0.00	1,089,513.50	1,089,513.50	-18,175,512.03
44,633.90	0.00	125.00	105.00	12.67	44,633.90	1,089,513.50	565,511.51	1,655,025.01	-17,610,000.52
								Qc	

4.2.7 RHEN Alternative 6

There are two streams in the network. We don’t need to find Pinch temperature using Problem table method. From resilient heat exchanger network design, the heat exchanger is great ability so the furnace is not used in this case. For this reason, the hot utility requirement is zero. The minimum utility requirements have been predicted is 1.814x 10^6 Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.12. Figure 4.8 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 5. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^{o}F.

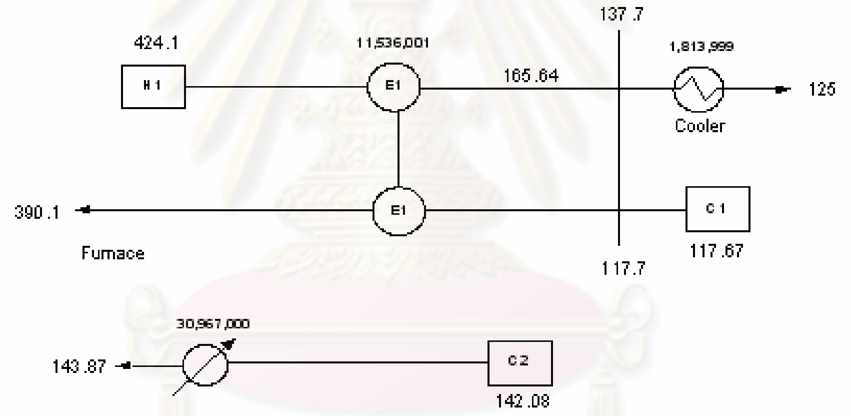


Figure 4.8 The resilient heat exchanger network alternative 6, RHEN-6

Table 4.12 Process stream data for alternative 6

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1

Synthesis Table for Cold End of RHEN-6							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-6							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.1	137.67	892,678	892,678	Selected
C1	11,536,000.51	45,699.80	390.1	137.67	0	913,996	Selected
b) State 2	Selected match pattern AH *						
H1	-558,186.53	44,633.90	125.16	137.67	1,806,674	892,678	To cooler*
C1							

4.2.8 RHEN Alternative 7

There are two streams in the network. We don’t need to find Pinch temperature using Problem table method. From resilient heat exchanger network design, the heat exchanger is great ability so the Purge reboiler is not used in this case. The minimum utility requirements have been predicted is 6.554 x 10^6 Btu/hr of cold utilities and the hot utility requirement is zero.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.13. Figure 4.9 shows a design of resilient heat exchanger network for Butane Isomerization process alternative 7. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 20 ^oF.

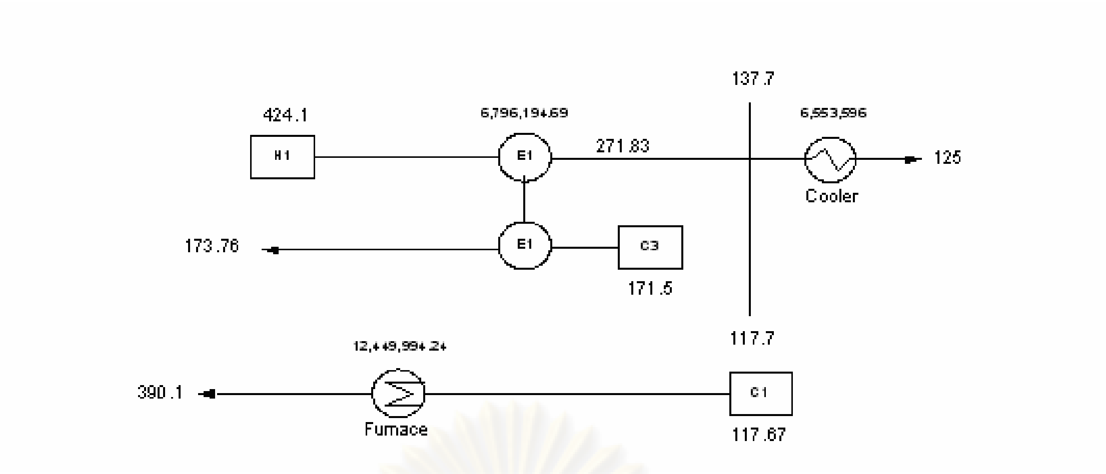


Figure 4.9 The resilient heat exchanger network alternative 7, RHEN-7

Table 4.13 Process stream data for alternative 7

Stream	W (Btu/hr-^oF)	T_{supply} (^oF)			T_{target} (^oF)
		Nominal	Max	Min	Nominal
H1	44,633.90	424.1	444.1	404.1	125
C1	45,699.80	117.67	137.67	117.67	390.1
C3	5,393,805.31	171.5	172.5	170.5	173.76

Synthesis Table for Cold End of RHEN-7							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern B[H]*						
H1	565,511.51	44,633.90	137.67	125	892,678	0	Matched with C1
C1	0	45,699.80	117.67	117.67	0	913,996	Matched with H1
b) State 2							
H1	565,511.51	44,633.90	137.67	125	21,318	0	To Cooler
C1							

Synthesis Table for Hot End of RHEN-7							
Stream	Load	W	T1	T2	D1	D2	Match Action
a) State 1	Selected match pattern C[H]*						
H1	11,891,809.98	44,633.90	404.1	137.67	892,678	892,678	Selected
C3	6,796,194.69	5,393,805.31	173.76	172.5	0	10,787,610.62	Selected
b) State 2	Selected match pattern AH *						
H1	-5,691,995.33	44,633.90	10.14	137.67	11,680,289	892,678	To cooler*
C3							

The various alternatives of heat exchanger network are designed for the Butane Isomerization process, the energy saved from the Base case as in Table 4.1

Table 4.14 Energy integration for Butane Isomerization process

	Base case	RHEN-1	RHEN-2	RHEN-3	RHEN-4	RHEN-5	RHEN-6	RHEN-7
Hot utilities usage, (x10^7Btu/hr)	4.9998	4.3901	4.2816	4.5218	4.2825	4.3908	4.3150	4.3410
Furnace	0.6848	0.6848	0.6848	0.6848	0.6848	1.2450	-	1.2450
DIB column reboiler	3.0960	-	-	3.0960	3.0960		3.0960	3.0960
Purge column reboiler	1.2190	1.2190	1.2190	-	-	1.2190	1.2190	-
DIB auxiliary reboiler	-	2.4863	2.3777	-	-	1.9268	-	-
Purge auxiliary reboiler	-	-	-	0.7409	0.5016	-	-	-
Cold utilities usage, (x10^7Btu/hr)	0.7748	0.1651	0.0566	0.2968	0.0566	0.1651	0.1814	0.6554
Energy saving, %	-	21.12	24.88	16.56	24.86	21.10	22.14	13.48

4.3 The Butane Isomerization Alternatives

Seven alternatives of heat exchanger networks (HEN) designs of the Butane Isomerization plant are proposed to save energy from the Base Case and use to evaluate performance of control structures are designed both simply energy-integrated plant and complex energy-integrated plant.

In Figure 4.10 show the Base Case of Butane Isomerization process with simply energy integration, we used a feed-effluent heat exchanger (FEHE) to reduce the amount of fuel burned in the furnace. The heat of reaction and the heat added in the furnace are therefore removed in the flooded condenser.

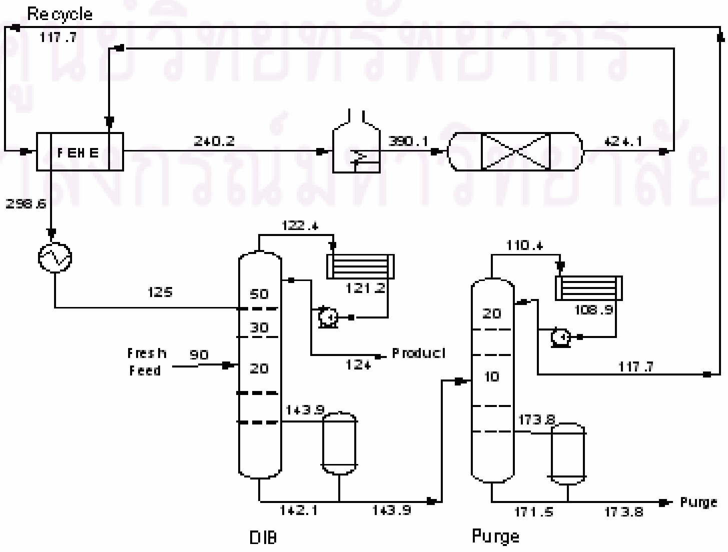


Figure 4.10 Butane Isomerization process-Base Case

In alternative 1 have two heat exchangers for preheat feed and the reboiler in the DIB column is driven by the reactor effluent stream. The first heat exchanger is used to reduce the amount of fuel burned in the furnace and the second heat exchanger is used to reboil for DIB column. In this work, the DIB column is added an auxiliary reboiler to supply the heat because the total heat from the reactor is not enough to reboil the column. The heat of reaction and the heat added in the furnace are removed in the flooded condenser as in Figure 4.11.

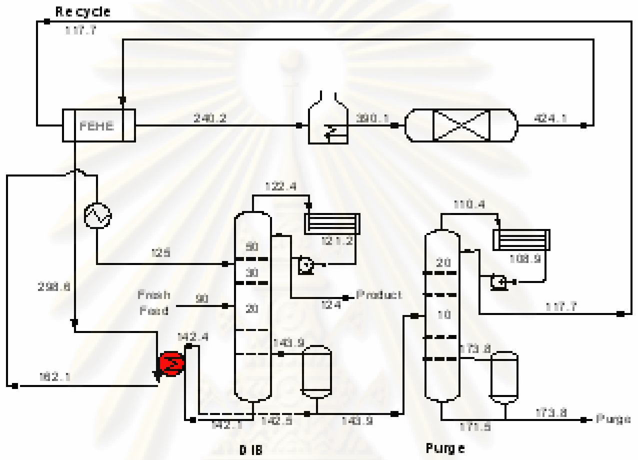


Figure 4.11 Butane Isomerization process alternative 1

In alternative 2 have two heat exchangers for preheat feed and the reboiler in the DIB column is driven by the reactor effluent stream. The first heat exchanger is used to reboil for DIB column and the second heat exchanger is used to reduce the amount of fuel burned in the furnace. In this work, the DIB column is added an auxiliary reboiler to supply the heat because the total heat from the reactor is not enough to reboil the column. The heat of reaction and the heat added in the furnace are removed in the flooded condenser as in Figure 4.12.

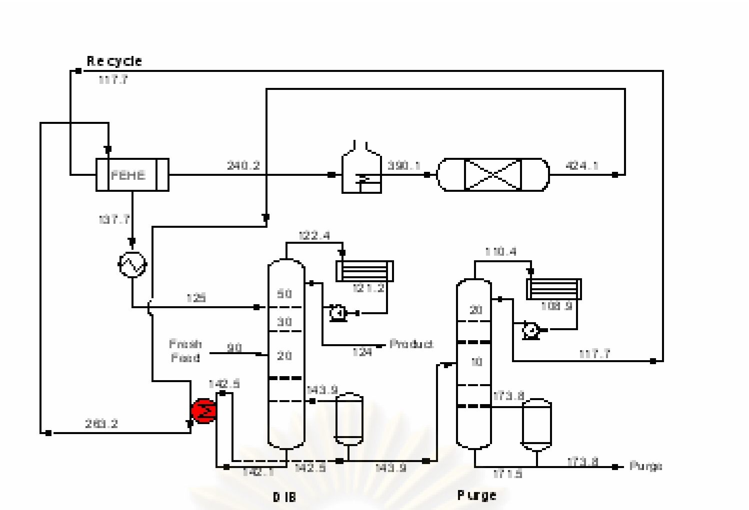


Figure 4.12 Butane Isomerization process alternative 2

In alternative 3 have two heat exchangers for preheat feed and the reboiler in the Purge column is driven by the reactor effluent stream. The first heat exchanger is used to reduce the amount of fuel burned in the furnace and the second heat exchanger is used to reboil for Purge column. In this work, the Purge column is added an auxiliary reboiler to supply the heat because the total heat from the reactor is not enough to reboil the column. The heat of reaction and the heat added in the furnace are removed in the flooded condenser as in Figure 4.13.

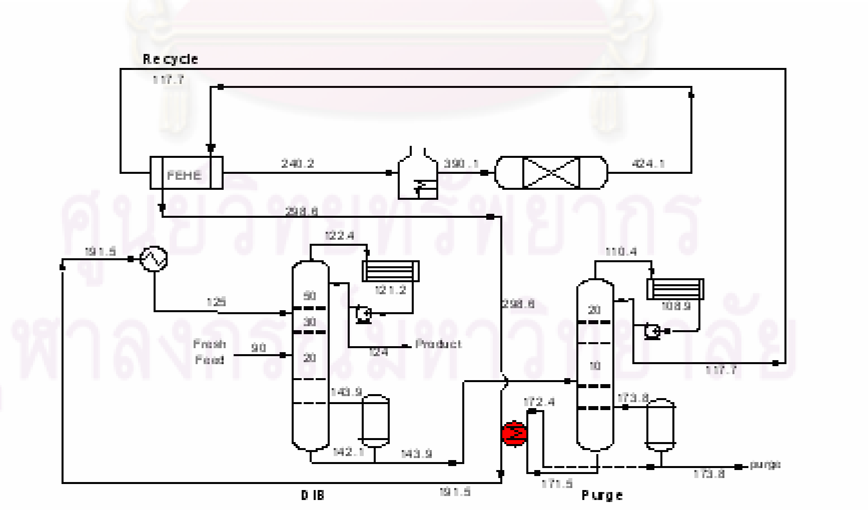


Figure 4.13 Butane Isomerization process alternative 3

In alternative 4 have two heat exchangers for preheat feed and the reboiler in the Purge column is driven by the reactor effluent stream. The first heat exchanger is used to reboil for Purge column and the second heat exchanger is used to reduce the amount of fuel burned in the furnace. In this work, the Purge column is added an auxiliary reboiler to supply the heat because the total heat from the reactor is not enough to reboil the column. The heat of reaction and the heat added in the furnace are removed in the flooded condenser as in Figure 4.14.

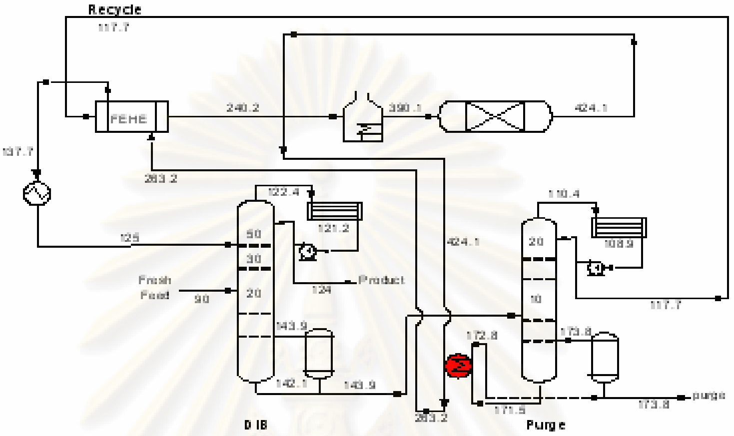


Figure 4.14 Butane Isomerization process alternative 4

In alternative 5 has a heat exchanger for preheat the reboiler in the DIB column is driven by the reactor effluent stream. The DIB column is added an auxiliary reboiler to supply the heat because the total heat from the reactor is not enough to reboil the column. The heat of reaction and the heat added in the furnace are removed in the flooded condenser as in Figure 4.15.

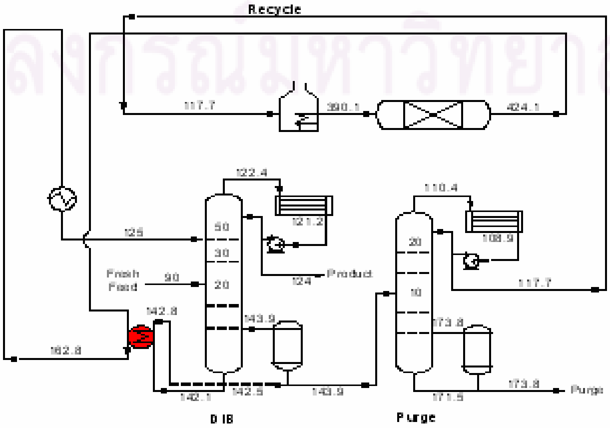


Figure 4.15 Butane Isomerization process alternative 5

In alternative 6 has a heat exchanger for preheat the reactor feed stream. We use a large size of heat exchanger. Because the heat is enough to reaction then the furnace is not used. The heat from the reaction is removed in the flooded condenser as in Figure 4.16.

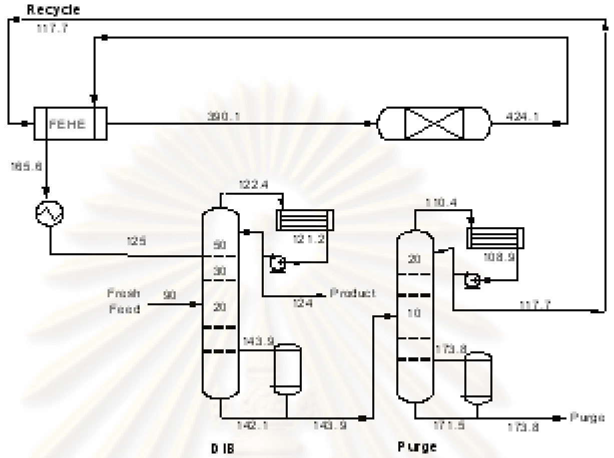


Figure 4.16 Butane Isomerization process alternative 6

In alternative 7 has a heat exchanger for preheat reboiler in the Purge column. Because the heat are enough to reboil the column then the reboiler of Purge column is not used. The reactor feed stream used the heat from the furnace and the heat from the reaction are removed in the flooded condenser as in Figure 4.17

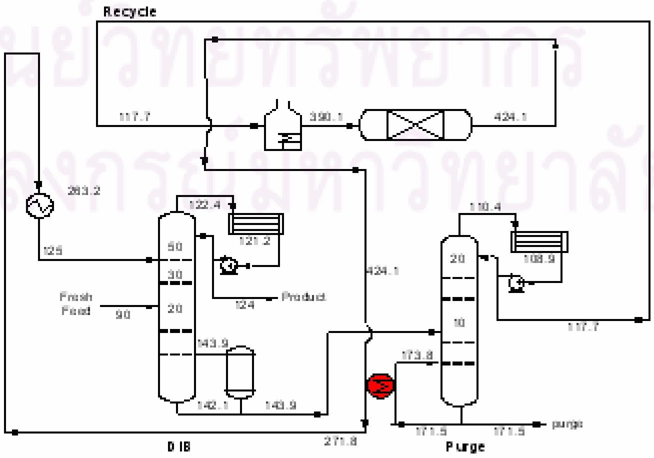


Figure 4.17 Butane Isomerization process alternative 7

4.4 Steady-State Modeling

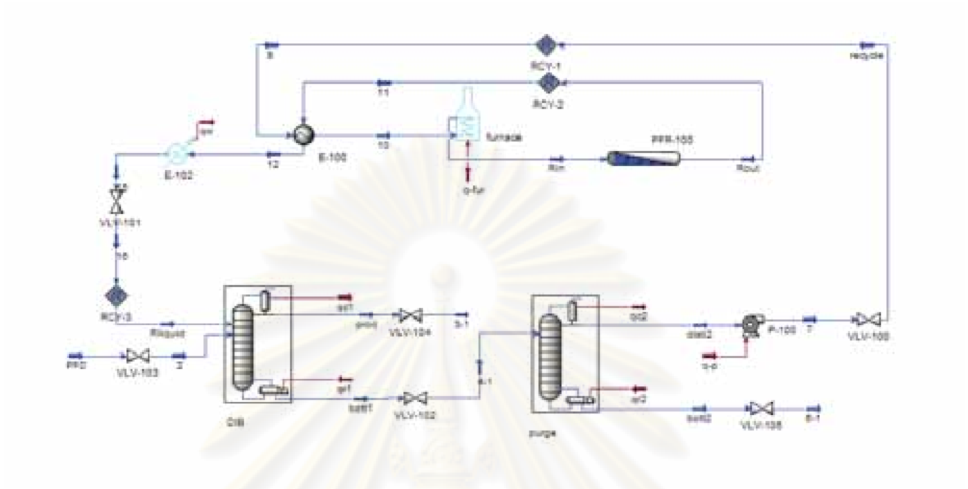
First, a steady-state model is built in HYSYS.PLANT, using the flowsheet and equipment design information, mainly taken from Luyben et al. (1998) and Poothanakul P. (2002). Appendix A presents the data and specifications for the different equipment. For the simulation, the Peng-Robinson model is selected for physical property calculations because of its reliability in predicting the properties of most hydrocarbon-based fluids over a wide range of operating conditions. The reaction kinetics of both reactions are modeled with standard Arrhenius kinetic expressions available in HYSYS.PLANT, and the kinetic data are taken from Luyben et al. (1998).

4.4.1 Steady State Simulation of Butane Isomerization Process (Base Case)

Figure 4.18 shows the HYSYS flowsheet of Butane Isomerization process (Base Case). The steady state simulation results are summarized in Table A.1 and Figure 4.19. For the comparison, the steady state simulation results give by Poothanakul P. (2002) are also listed in Table A.2.

Since there are three material recycle streams in Butane Isomerization process, three recycle modules are inserted in the streams: hot stream to FEHE, the recycle stream to the reactor and DIB-Feed. Proper initial values should be chosen for these streams; otherwise the iterative calculations might converge to another steady state due to the non-linearity and unstable characteristics of the process.

All of the two columns are simulated using the “distillation column” module. When columns are modeled in steady state, besides the specification of inlet streams, pressure profiles, the number of trays and the feed tray, two additional variables should be additionally specified for columns with condenser or reboiler. These could be the duties, reflux rate, draw stream rates, composition fraction, etc. We chose to specify a priori overhead and bottom component mole fraction for all columns. These mole fractions are specified to meet the required purity of product given by Luyben et al. (1998) and Poothanakul P. (2002). The tray sections of the columns are calculated using the tray sizing utility in HYSYS, which calculates tray diameters based on sieve trays. The column specifications of Butane Isomerization process (Base Case) are given in Table A.5. Although the tray diameter and spacing, weir length and high are not required for steady state modeling, they are required for dynamic simulation.



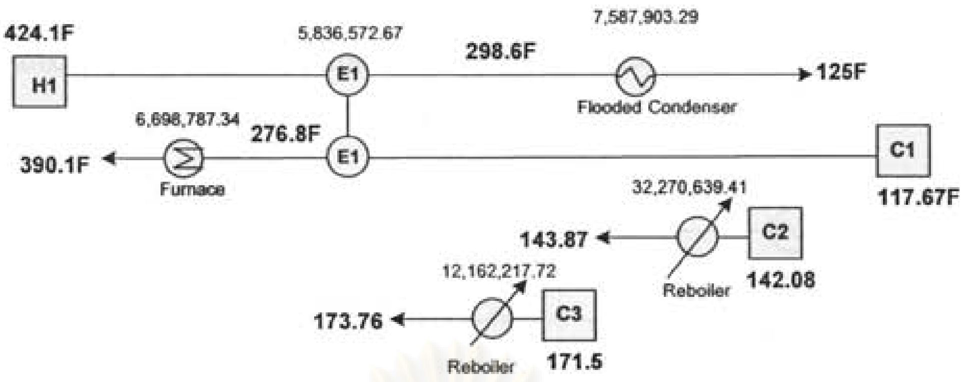


Figure 4.19 The steady state simulation results of heat exchanger network, Base Case

4.4.2 Steady State Simulation of Butane Isomerization Process Alternative 1 and 2

In alternative 1 and 2, there are two heat exchangers and additionally the reboiler in the DIB column is driven by the reactor effluent stream. The first heat exchanger for alternative 1 is used to preheat the recycle stream before entering furnace and reboil the DIB column for alternative 2. The second heat exchanger is used to reboil the DIB column for alternative 1 and preheat the recycle stream for alternative 2. All process-to-process heat exchangers are simulated using a heat exchanger with a hot stream on the shell side and a cold stream on the tube side. Note that, a minimum ΔT of about 20 °F is assumed for all process-to-process heat exchangers.

The DIB column is simulated using a "refluxed absorber" that it does not include a reboiler. The Purge column is simulated using the "distillation column" module. Since a "reflux absorber" module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a "refluxed absorber" module.

In alternatives 1 and 2, a tank is needed to accommodate liquid from the bottom of DIB column.

Figure 4.22 and 4.23 show the HYSYS flowsheets of the Butane Isomerization process with energy integration schemes for alternative 1 and 2. The selected process streams data for these alternatives are not included in this chapter but listed in Appendix A. The steady state simulation results are shown in Figure 20 and 21.

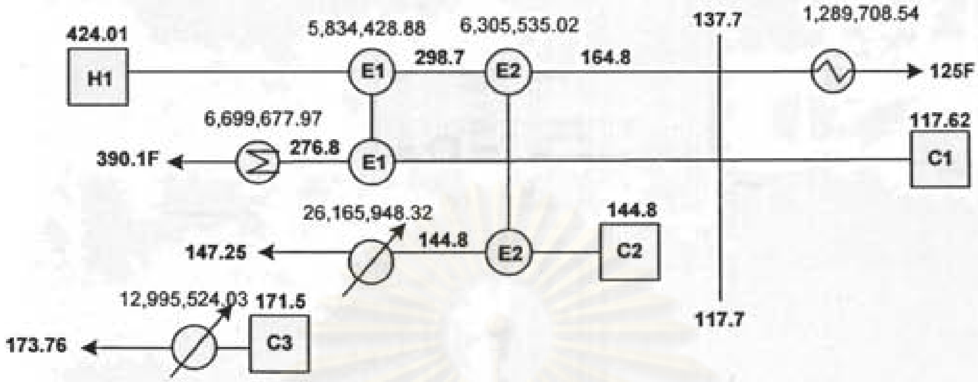


Figure 4.20 The steady state simulation results of resilient heat exchanger network, alternative 1

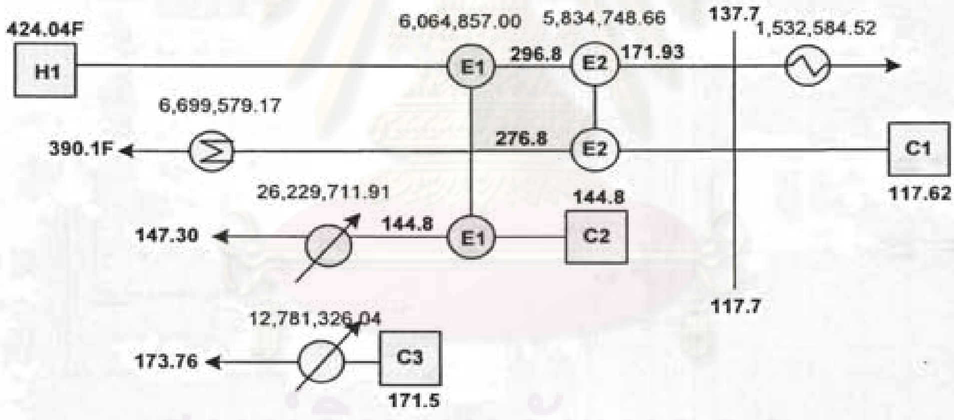


Figure 4.21 The steady state simulation results of resilient heat exchanger network, alternative 2

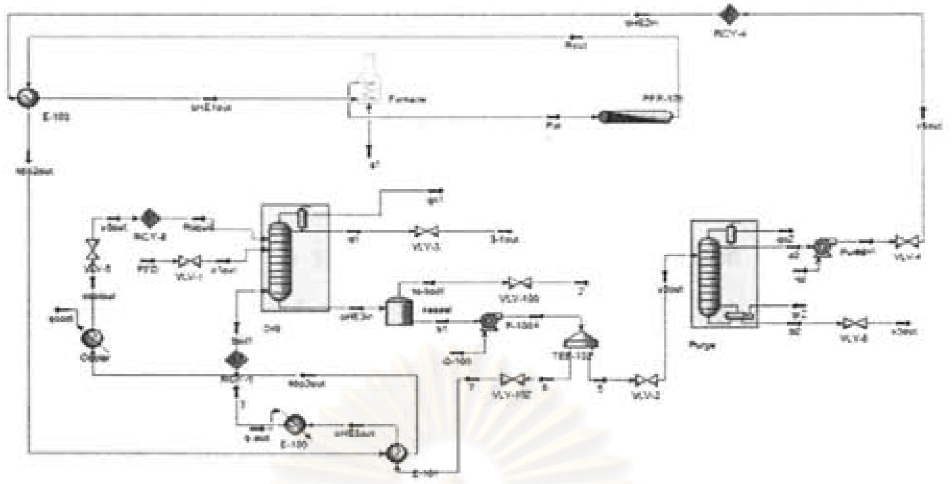


Figure 4.22 The simulated Butane Isomerization process (alt.1) at steady-state by HYSYS

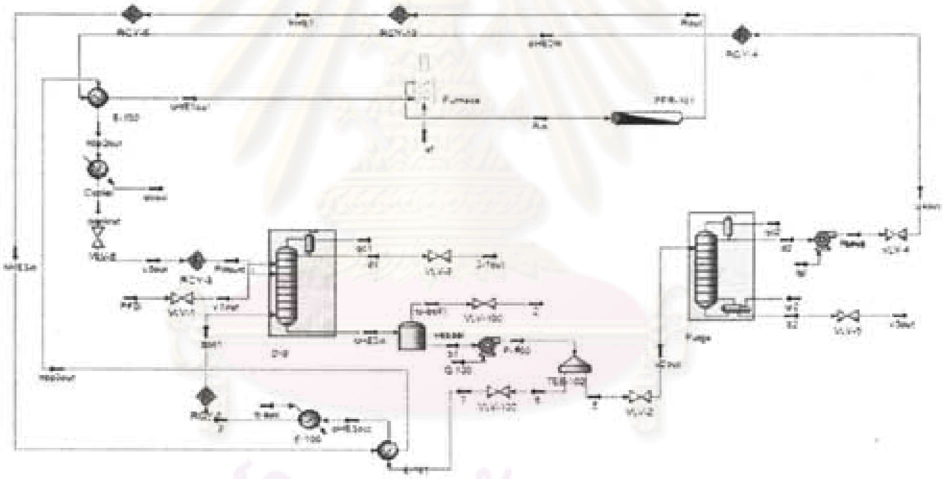


Figure 4.23 The simulated Butane Isomerization process (alt.2) at steady-state by HYSYS

4.4.3 Steady State Simulation of Butane Isomerization Process Alternative 3 and 4

In alternative 3 and 4, there are two heat exchangers and additionally the reboiler in the Purge column is driven by the reactor effluent stream. The first heat exchanger for alternative 3 is used to preheat the recycle stream before entering furnace and reboil the Purge column for alternative 2. The second heat exchanger is used to reboil the Purge column for alternative 3 and preheat the recycle stream for alternative 4. All process-to-process heat exchangers are simulated using a heat exchanger with a hot stream on the shell side and a cold stream on the tube side. Note that, a minimum ΔT of about 20 ^{o}F is assumed for all process-to-process heat exchangers.

The DIB column is simulated using the “distillation column” module. The Purge column is simulated using “refluxed absorber” that it does not include a reboiler. Since a “reflux absorber” module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a “refluxed absorber” module.

In alternatives 3 and 4, a tank is needed to accommodate liquid from the bottom of Purge column.

Figure 4.26 and 4.27 show the HYSYS flowsheets of the Butane Isomerization process with energy integration schemes for alternative 3 and 4. The selected process streams data for these alternatives are not included in this chapter but listed in Appendix A. The steady state simulation results are shown in Figure 4.24 and 4.25.

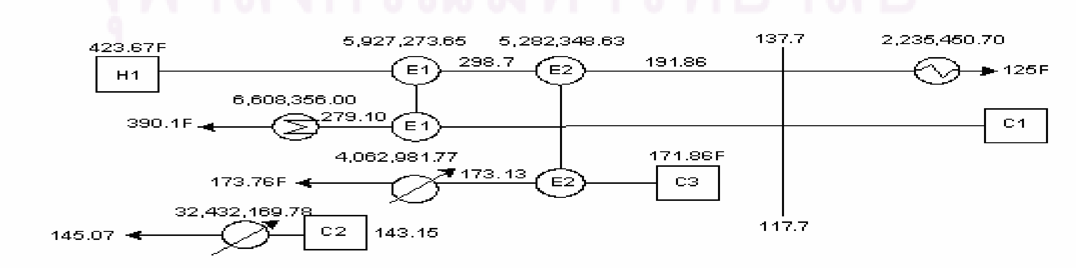


Figure 4.24 The steady state simulation results of resilient heat exchanger network, alternative 3

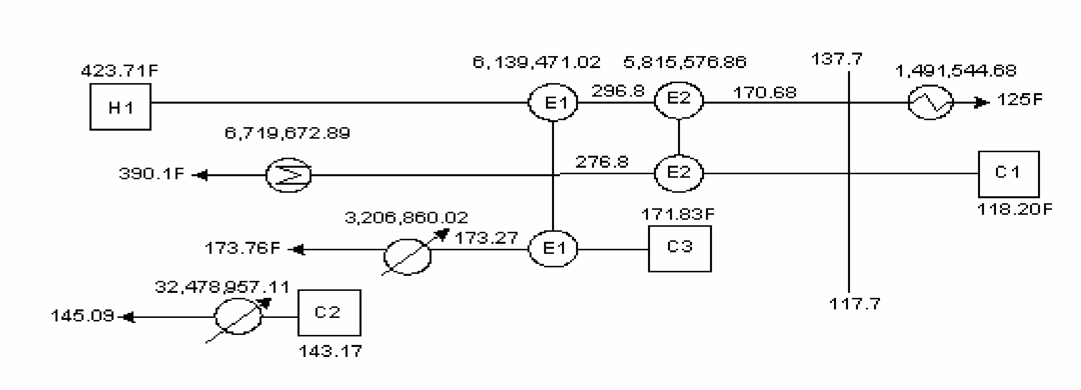


Figure 4.25 The steady state simulation results of resilient heat exchanger network, alternative 4

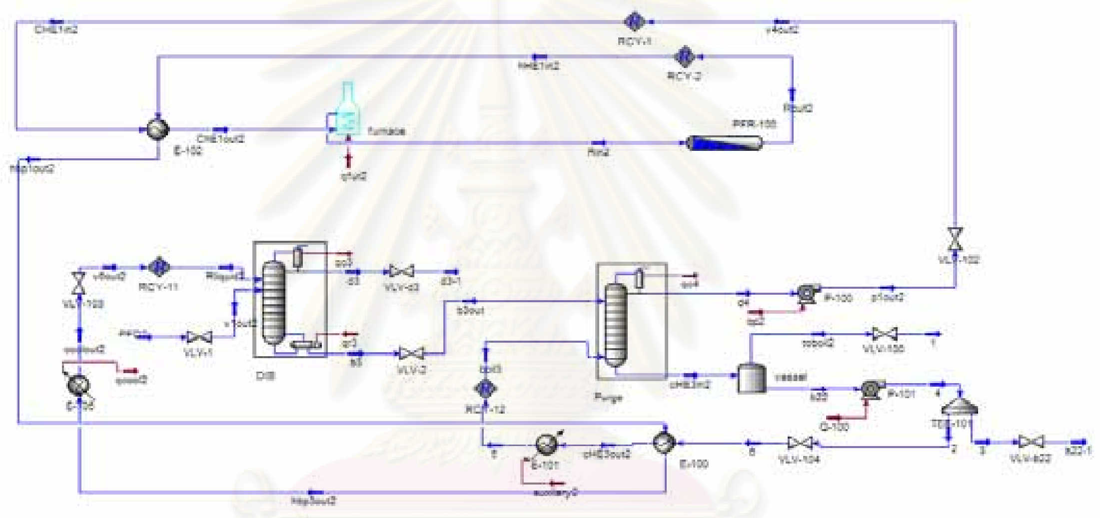


Figure 4.26 The simulated Butane Isomerization process (alt.3) at steady-state by HYSYS

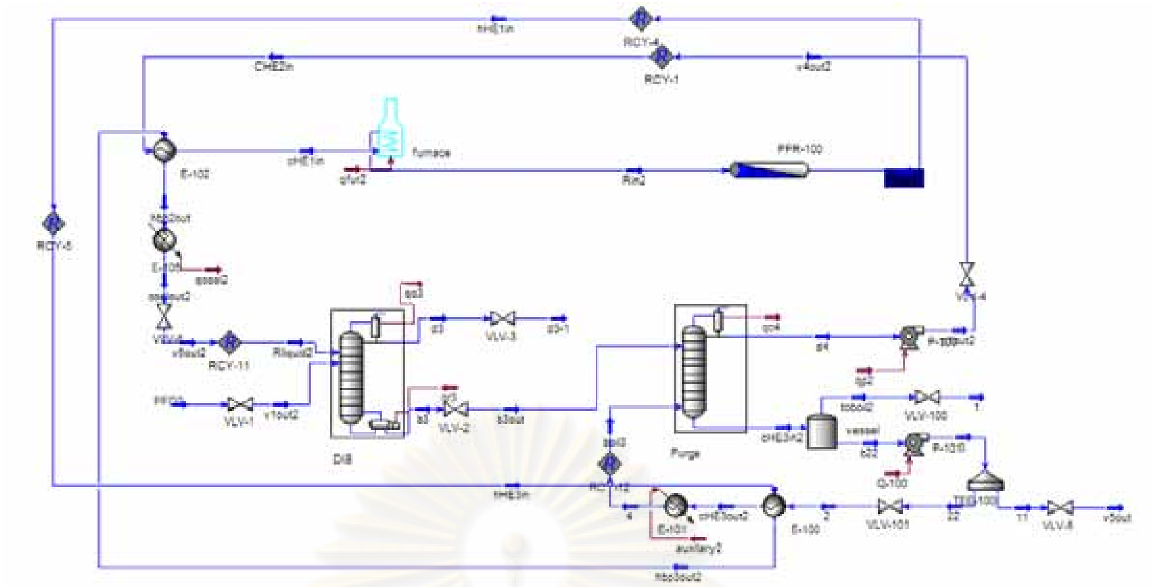


Figure 4.27 The simulated Butane Isomerization process (alt.4) at steady-state by HYSYS

4.4.4 Steady State Simulation of Butane Isomerization Process Alternative 5

In alternative 5 there is a heat exchanger and additionally the reboiler in the DIB column is driven by the reactor effluent stream. The heat exchanger is used to reboil the DIB column. The process-to-process heat exchanger is simulated using a heat exchanger with a hot stream on the shell side and a cold stream on the tube side. Note that, a minimum ΔT of about 20 ^{o}F is assumed for all process-to-process heat exchangers.

The DIB column is simulated using a “refluxed absorber” that it does not include a reboiler. The Purge column is simulated using the “distillation column” module. Since a “reflux absorber” module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a “refluxed absorber” module.

In alternatives 5, a tank is needed to accommodate liquid from the bottom of DIB column.

Figure 4.29 shows the HYSYS flowsheets of the Butane Isomerization process with energy integration schemes for alternative 5. The selected process streams data for this alternative is not included in this chapter but listed in Appendix A. The steady state simulation result is shown in Figure 4.28.

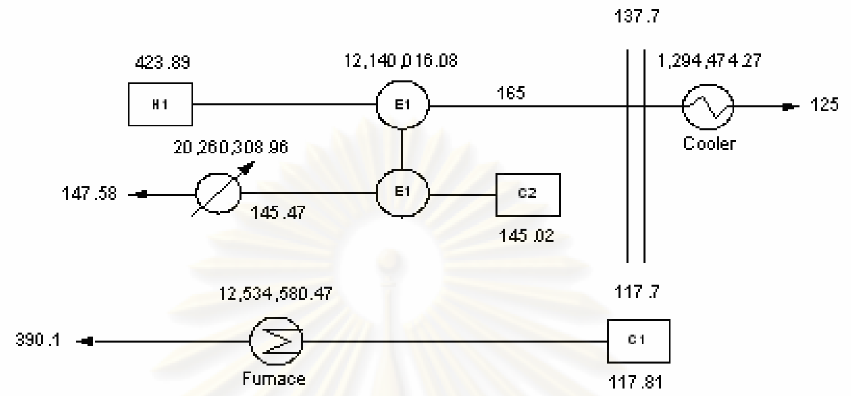


Figure 4.28 The steady state simulation results of resilient heat exchanger network, alternative 5

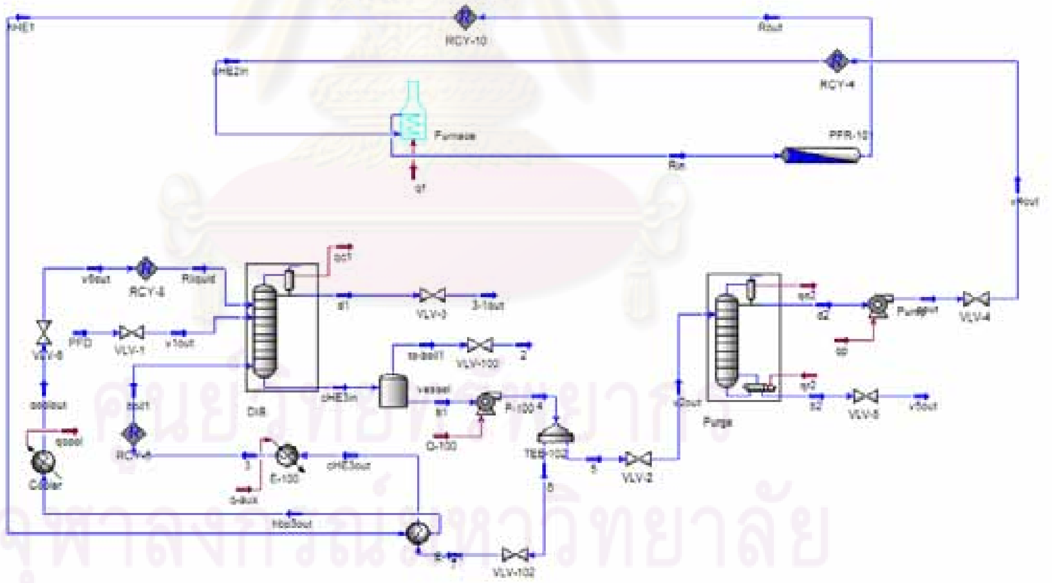


Figure 4.29 The simulated Butane Isomerization process (alt.5) at steady-state by HYSYS

4.4.5 Steady State Simulation of Butane Isomerization Process Alternative 6

In alternative 6, there is a heat exchanger and is driven by the reactor effluent stream. The heat exchanger for alternative 6 is used to preheat the reactor feed stream. Because we used a large heat exchanger which the heat is enough to increase reactor feed temperature, then the furnace is not used. The process-to-process heat exchanger is simulated using a heat exchanger with a hot stream on the shell side and a cold stream on the tube side. Note that, a minimum ΔT of about 20 ^{o}F is assumed for all process-to-process heat exchangers.

The DIB column is simulated using a "refluxed absorber" that it does not include a reboiler. The Purge column is simulated using the "distillation column" module. Since a "reflux absorber" module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a "refluxed absorber" module.

In alternative 6, a tank is needed to accommodate liquid from the bottom of DIB column.

Figure 4.31 shows the HYSYS flowsheets of the Butane Isomerization process with energy integration schemes for alternative 6. The selected process streams data for this alternative is not included in this chapter but listed in Appendix A. The steady state simulation result is shown in Figure 4.30.

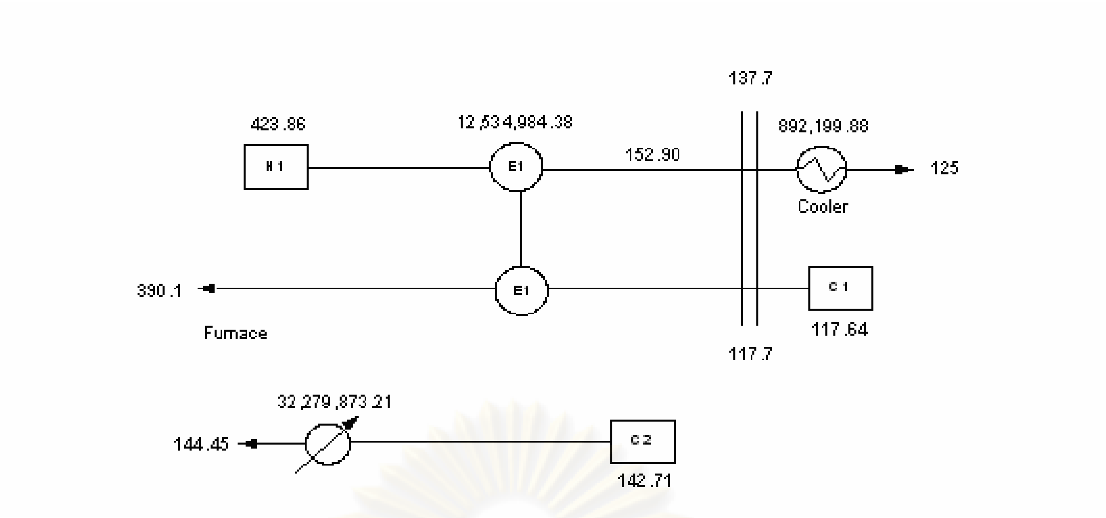


Figure 4.30 The steady state simulation results of resilient heat exchanger network, alternative 6

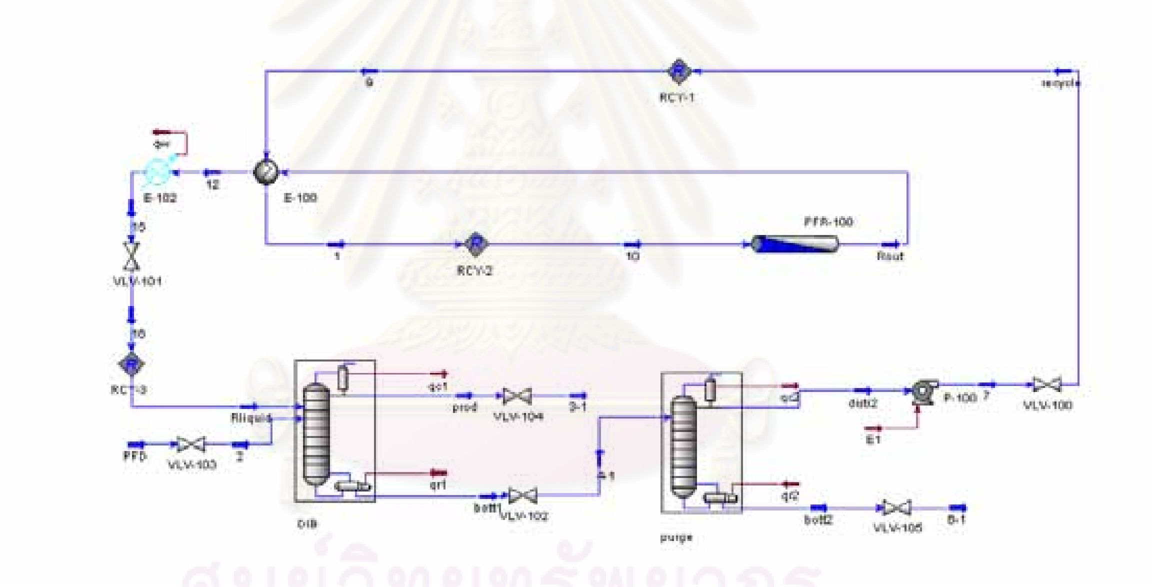


Figure 4.31 The simulated Butane Isomerization process (alt.6) at steady-state by HYSYS

4.4.6 Steady State Simulation of Butane Isomerization Process Alternative 7

In alternative 7, there is a heat exchanger and additionally the reboiler in the Purge column is driven by the reactor effluent stream. The heat exchanger for alternative 7 is used to reboil the Purge column. Because we used a large heat exchanger which the heat is enough to reboil the column, then the reboiler is not used.

The process-to-process heat exchanger is simulated using a heat exchanger with a hot stream on the shell side and a cold stream on the tube side. Note that, a minimum ΔT of about 20 ^{o}F is assumed for all process-to-process heat exchangers.

The DIB column is simulated using the “distillation column” module. The Purge column is simulated using “refluxed absorber” that it does not include a reboiler. Since a “reflux absorber” module is used, only one variable need to be specified for the columns with condenser. The overhead mole fraction is chosen to be specified for a “refluxed absorber” module.

In alternative 7, a tank is needed to accommodate liquid from the bottom of Purge column.

Figure shows 4.33 the HYSYS flowsheets of the Butane Isomerization process with energy integration schemes for alternative 7. The selected process streams data for this alternative is not included in this chapter but listed in Appendix A. The steady state simulation result is shown in Figure 4.32.

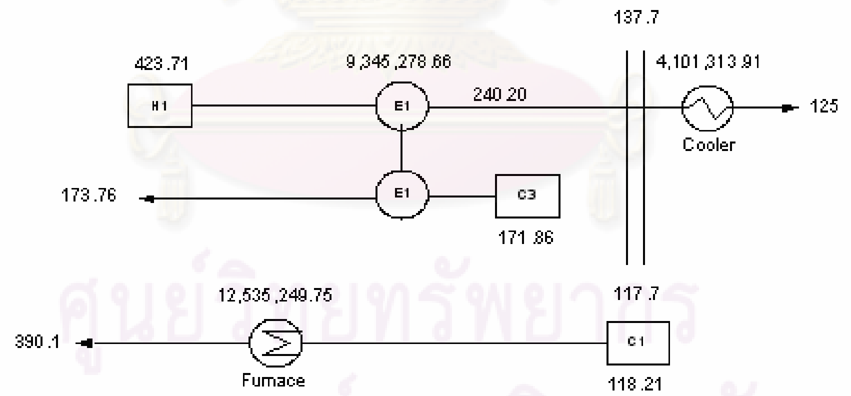


Figure 4.32 The steady state simulation results of resilient heat exchanger network, alternative 7

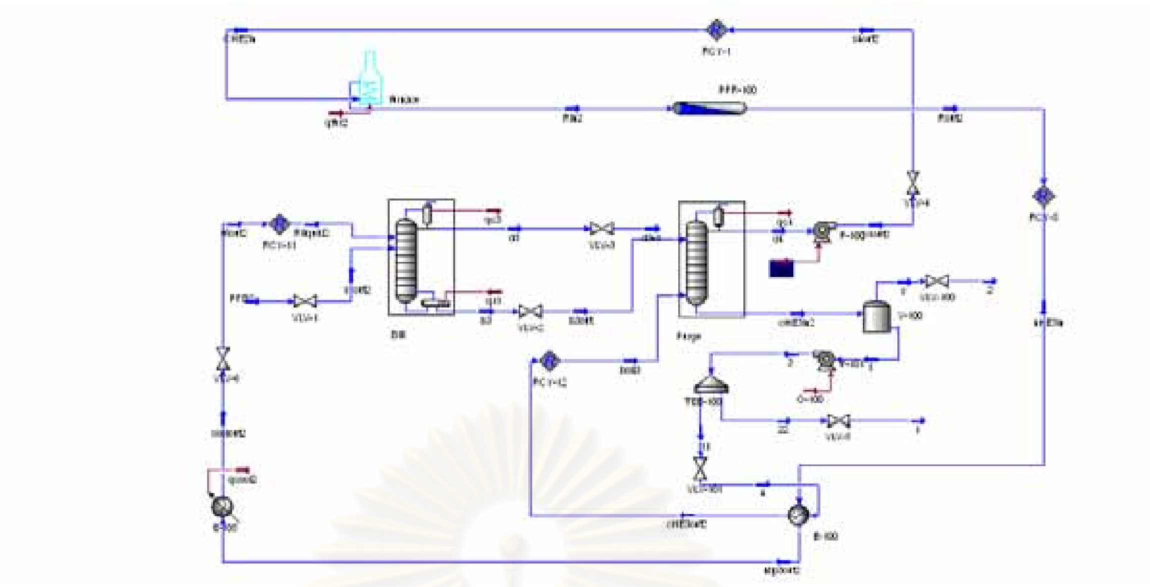


Figure 4.33 The simulated Butane Isomerization process (alt.7) at steady-state by HYSYS

4.4.7 Energy Integration from Steady State Simulation of Butane Isomerization Process

From steady state simulation results by HYSYS, the energy saved from the base case heat consumption as shown in Table 4.15.

Table 4.15 Energy integration for Butane Isomerization process (steady state simulation)

	Alternatives							
	Base case	**RHEN-1**	**RHEN-2**	**RHEN-3**	**RHEN-4**	**RHEN-5**	**RHEN-6**	**RHEN-7**
1. Utilities usage (Btu/hr) x 10^7								
Furnace	6.6988	6.6997	6.6996	6.6084	6.7197	12.5346	-	12.5352
DIB reboiler	32.2706	-	-	32.4322	32.4790	0.0000	32.2799	32.4790
Purge reboiler	12.1624	12.9955	12.7813	-	-	11.2306	12.2602	-
DIB auxiliary reboiler	-	26.1659	26.2297	-	-	20.2603	-	-
Purge auxiliary reboiler	-	-	-	4.0630	3.2069	-	-	-
Hot utilities usage	51.1318	45.8612	45.7106	43.1035	42.4055	44.0255	44.5400	45.0142
Cold utilities usage	7.5879	1.2897	1.5326	2.2355	1.4915	1.2945	0.8922	4.1013
2. Energy saving, %	-	19.70	19.54	22.79	25.24	22.82	22.63	16.36

CHAPTER V

CONTROL STRUCTURE DESIGN

5.1 Plantwide control design procedure

Step 1: Establish control objectives

In this process we want to achieve the desired production rate and control the impurity of normal butane in the isobutene product at 2 mol%. Reactor pressure cannot exceed the design operating pressure of 700 psia. We assume that we are free to choose the production rate handle. Neither the fresh feed nor product flowrates are fixed other plant considerations. In the pentane purge column, we do not want to lose too much nC4.

Step 2: Determine control degrees of freedom

The Butane Isomerization process for Base case has 14 control degrees of freedom. They include fresh feed valve; DIB column stream, cooling water, reflux, distillate, and bottoms valve; Purge column stream, cooling water, reflux, distillate, and bottom valve; furnace fuel valve; flooded condenser cooling water valve; and DIB column feed valve. For the resilient heat exchanger networks, each has a bypass valve of heat exchanger. So there are 16 degrees of freedom for 2 heat exchangers (RHEN-1, RHEN-2, RHEN-3 and RHEN-4) and 15 degrees of freedom for a heat exchanger (RHEN-5). Because the furnace and the purge column reboiler are not used for RHEN-6 and RHEN-7, respectively, so there are 14 degrees of freedom for a heat exchanger.

Step 3: Establish energy management system

The exothermic heat of reaction must be removed, and the reactor feed must be heated to a high enough temperature to initiate the reaction. Since the heat of reaction is not large and complete one pass conversion is not achieved, the reactor exit temperature is only 32 ^{o}F higher than the reactor inlet temperature. Since heat transfer coefficients in gas-to-gas systems are typically quite low, this small temperature differential would require a very large heat exchanger if only the reactor effluent is used to heat the reactor feed and no furnace is used. Therefore, a furnace is required to bring the reactor inlet up to the desired level.

The use of a feed-effluent heat exchanger (FEHE) reduces the amount of fuel burned in the furnace. So form a steady-state viewpoint, the economic trade-off between utility and capital costs would produces a fairly large heat exchanger and a small furnace. However, the exothermic heat of reaction and the heat of vaporization supplied in the furnace must be dissipated to utilities at the flooded condenser. If the FEHE is too large, reactor heat will be recycled. Also the large heat exchanger, the smaller heat input in the furnace. This could potentially be solved using a bypass around the FEHE on the cold side. This should enable us to prevent reactor runaway to high temperature and would guarantee that the furnace4 is in operation at all times. However, unless the furnace is large enough, there is no guarantee that the system will never quench to low temperature when a large disturbance occurs to drop the reactor inlet temperature.

A second difficulty that can occur with a large FEHE is a hydraulic problem. The recycle stream entering the heat exchanger is subcooled liquid (115^{o}F) at the pressure in the reactor section (685 psia). As this stream is heated to the required reactor inlet temperature of 390 ^{o}F, it begins to vaporize. It is superheated vapor when fed into the reactor (390 ^{o}F at 665 psia). When a small FEHE is used, the exit is 292^{o}F for the recycle stream, which means the stream is still all liquid. All the vaporization occurs in the furnace. If a large heat exchanger were used, vaporization would begin to occur in the heat exchanger. This would make the hydraulic design of this FEHE much more difficult. The dynamic response could also be adversely affected as changes in flowrates and temperature make the stream go in and out of the two-phase region.

The same problem occurs on the hot side of FEHE. As the hot reactor effluent is cooled, it starts to condense at some temperature. And this dewpoint temperature could occur in the heat exchanger and not in the flooded condenser if a large area is used.

Because of both the heat dissipation and hydraulic concerns, we use a relatively small FEHE: 1000 ft^2 compared to the Stanford Research Report's listed area of 3100 ft^2. So the energy management system consists of controlling reactor inlet temperature by furnace firing and controlling the rate of removal in the flooded condenser by cooling water flowrate. The heat of reaction and the heat added in the furnace are therefore removed in the flooded condenser. Because of this design we do not need a bypass around the FEHE.

In this work, we designed the new heat exchanger network to be chosen for each proper control structure. For RHEN-1, RHEN-2, RHEN-3 and RHEN-4 are complex heat-integrated process that they use two FEHEs to reduce utilities consumption at furnace and column reboiler. In RHEN-5, RHEN-6, and RHEN-7 are simply heat-integrated process, they use a large FEHE to save energy from Base Case.

Step4: Set production rate

We are not constrained either by reactant supply or product demand to set production rate at a certain point in the process. We need to examine which variables affect reactor productivity.

The kinetic expression for the isomerization reaction is relatively simple. For the irreversible case, reaction rate depends upon the forward rate constant, reactor volume, and normal butane concentration.

$$R = k_F V_R C_{n\mathrm{C}_4}$$

Form this expression we see that only three variables could possibly be dominant: temperature, pressure, and mole fraction of nC_4 in the reactor feed.

Pressure affects productivity through its influence on the reactant concentration. Since the normal operating reactor pressure is close to the design limit, we are constrained in how much we can move pressure to achieve the desired production rate change. The nC_4 mole fraction in the reactant feed is about 0.81.

Therefore large absolute changes in the reactant feed mole fraction would have to be made to be achieve a significant relative change in throughput.

Finally, we are then left with temperature. The relative change in reaction rate depends upon the temperature through the activation energy. For a 10 ^{o}F change in temperature, the reaction rate increases by 20 percent. Clearly temperature is a dominant variable for reactor productivity.

In this work, we use the isomerization reaction in the irreversible case and we express the isomerization reaction in the reversible case too.

For the reversible case, reaction rate depends upon the forward and reverse rate constants, reactor volume, and nC_4 and iC_4 concentrations:

$$R = k_F V_R C_{nC4} - k_R V_R C_{iC4}$$

The activation energy of the reverse reaction is always greater than the activation energy of the forward reaction since the reaction is exothermic. Therefore the reverse reaction will increase more quickly with an increase in the temperature than will the forward reaction. Temperature may still dominate for reactor productivity, but in the opposite direction compared with the irreversible case, since conversion increases with lower temperature. However, when the temperature becomes too low, both reaction rates slow down such that we cannot achieve the desired production rate with this variable alone. Instead, the concentrations of nC_4 and iC_4 dominate the rate through the relationship imposed by the equilibrium constant $K_{eq} = C_{iC4} / C_{nC4}$.

Therefore we choose the reactor inlet temperature setpoint as the production rate handle for the irreversible case. However, for the reversible case we need to look for variables that affect the ratio of nC4 to iC4 in the recycle stream. For this case we will not have unit control for the reactor since these concentrations depend upon operation in other parts of the process.

Note that setting the production rate with variables at the reactor or within the process specifies the amount of fresh reactant feed flow required at steady state. The choices for the control system made in Steps 6 and 7 must recognize this relationship between production rate and fresh reactant feed flowrate.

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

The final isobutane product is the distillate from the DIB column, and we want to keep the composition of the nC_4 impurity at 2 mol%. Nothing can be done about the propane impurity. Whatever propane is in the fresh feed must leave in the product stream. Because the separation involves two isomers, the temperature profile is flat in the DIB column. Use of an overhead composition analyzer is necessary.

The choice of manipulated variables that can be used to control nC_4 composition in the DIB distillate include reflux flowrate, distillate flowrate and reboiler heat input. If the reflux ratio is high, control of reflux drum level using distillate flow may be ineffective, particularly if the distillate were going directly to a downstream process. If we use reflux flow to control reflux drum level, we must control distillate composite by manipulating the distillate flowrate. The reason is that distillate flow must math production rate, which is dependently set in the reactor. However, in this case we assume that the distillate is going to a storage tank or cavern, so large changes in distillate flowrate are not important. Distillate can then be used for reflux drum level control, allowing us to consider other variables for composition control.

Most distillate column responds more quickly to vapor rate changes than to changes in liquid rates. Therefore, we can select reboiler heat input to control nC_4 impurity in the distillate but this choice is poor because we are controlling something at the top of the column by changing a variable near the base. However, vapor changes affect all trays in the column quite quickly, so tight control of distillate composition should be possible by manipulating vapor boilup. A viable alternative is to control distillate composition with distillate flowrate and control reflux drum level with reflux flowrate.

To avoid the high-pressure safety constraint, we must control reactor pressure. We can use the distillate valve from the purge column, the flooded condenser cooling water valve, or the DIB column feed valve. The most logical variable to use for control of the flooded condenser (reactor) pressure is the DIB column feed valve. Base upon the discussion in step 3, we would then use the flooded condenser cooling water valve to keep the liquid leveling a good control range.

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

We have only two choices, DIB column base valve or purge column distillate valve, for fixing a flow in the recycle loop. Either of these would work. The rationale for picking one is based upon avoiding disturbances to the unit downstream of the fixed flow location. Since the purge column is not critical from the viewpoint of product quality, we elect to fix the flow upstream of reactor (purge column distillate flow) so that we minimize disturbance in reactor temperature and pressure.

We must control the two column pressure. This is best done by manipulating the condenser cooling water flowrates.

There are four liquid levels to be controlled i.e. DIB column base level, DIB column reflux drum, Purge column base level and Purge column reflux drum. The choices of manipulated variables that can be used to control DIB column reflux drum level include distillate product flowrate and reflux flowrate. We must also control the level in the DIB column base and in the purge column reflux drum and base.

Having made the choice to fix the purge column distillate flow, we are faced with the problem of how to control purge column reflux drum level. We have two primary choices: reflux flow or heat input. We choose the latter because the flowrate of the purge column reflux is small relative to the vapor coming overhead from the top of the column. Remember the Richardson rule, which says we select the largest stream. So we choose the heat input manipulating to control the Purge column reflux drum.

The flowrate of the purge stream from the base of the purge column is quite small, so it would not do a good job in controlling base level. This is especially true when the large stream flow has been selected to control the reflux drum level. Base level in the purge column can be controlled by manipulating the bottoms flowrate from the DIB column.

We are then left with controlling base level in the DIB column. The only remaining valve is the fresh nC_4 feed flowrate into the column. The feed is liquid and there only 20 trays between the lower feed point and the column base, so base level control using feed should be possible. This base level is also an indication of the nC_4 inventory within the process.

The material balance control structure works opposite to the direction of flow. Purge column distillate is fixed; purge column reflux drum level is controlled by vapor boilup; purge column base level is controlled by feed to the purge column; and DIB column base level is controlled by the fresh feed to the DIB column.

Had we started to assign the DIB column base level control first, we would have ended up with the same inventory control structure. The reason is as follow. Assume we had chosen the DIB column base valve to control base level. After resolving the purge column inventory loops, we would have found that we need to control the purge column base or reflux drum level with the fresh feed flow to the DIB column. The dynamic lags associated with these loops would have forced us back to the control strategy as described above.

An obvious question at this point is “Why don’t we just flow control the fresh feed into the process?”. If we did this, we could not fix the flowrate in the recycle loop. For example, suppose we select the following control structure: fix fresh feed flowrate, control DIB column base level with DIB bottoms, control purge column base level with heat input, and control purge column reflux drum level with distillate. This structure is intuitively attractive and permits us to fix the production rate directly by setting the fresh feed flowrate. However, only level controllers set the flows around the recycle loop, so we would expect problems with snowballing. Flow disturbances can propagate around the liquid recycle loop. In this work, we test the recycle flow disturbance so this structure is not work.

Step 7: Check component balances

Four components need to be accounted for; they are C_3, iC_4, nC_4 and iC_4. The light inert propane leaves in the product stream. The heavy inert component isopentane (iC_5) leaves in the purge stream. Any the iC_4 coming into the process in the fresh feed and the iC_4 produced by the reaction can leave in the product stream.

The only component that is trapped inside the stream and must be consumed by the reaction is the nC_4. The composition controller on the DIB distillate stream permits only a small amount (2 mol %) of the nC_4 to leave in the product stream. The purge stream from the bottom of the purge column permits only a small amount of the nC_4 to escape. This purge stream can be simple flow controlled if we don't mind losing a small amount of the nC_4 with the iC_5 purge. If the amount of the iC_5 in the fresh feed is small, this may be the simplest strategy and may have little economic penalty. Alternatively we could control the amount of the nC_4 in the purge column bottoms by manipulating bottom flowrate. Since there is a fairly large temperature change in the purge column, controlling the temperature on a suitable tray (tray 2) may be more practical since it eliminates the need for an on-line analyzer. In the simulations given later in this chapter, we adopt the simple strategy of flow-controlling the purge stream.

The amount of reactant the nC4 fed into the system must somehow be exactly balance by the amount of the nC4 converted to product the iC4. The process acts almost like a purge integration in term of the moles of the nC4. The way this balancing of the nC4 is accomplished in the control structure shown in Figure 5.1 is by using the level in the base of the DIB column to indicate if the nC4 is building up in the system or is being depleted. The material in the DIB base is mostly the nC4. There is a little the iC4 (16 percent) and a little the iC5 (5 percent), and the remained is the nC4. So DIB base level changes reflect changes in the nC4 inventory in the process. If the level is decreasing, fresh feed should be increased because we are consuming more reactant than we are feeding. Table 5.1 summarizes the component balance control strategy.

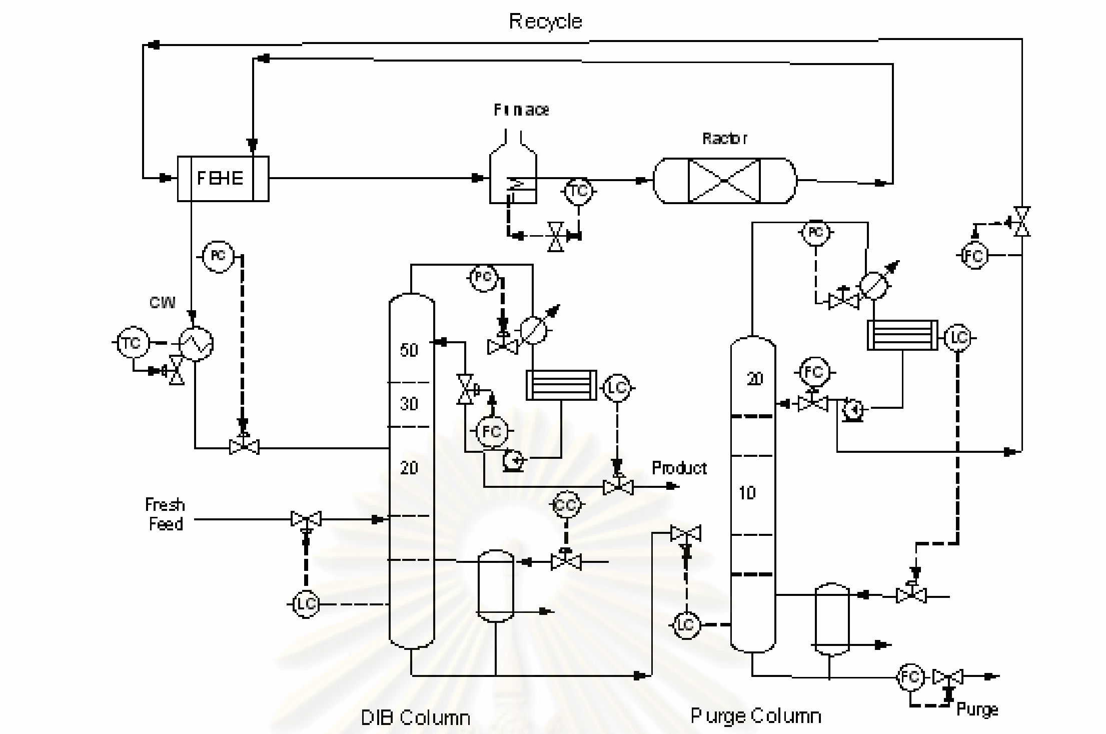


Figure 5.1 Control structure for Isomerization process (Plantwide Process Control, 1998)

Table 5.1 Component Material Balance for irreversible case

Component	Input	+Generation	-Output	-Consumption	=Accumulation Inventory Controlled by
C3	Fresh Feed	0	Product stream	0	Self-regulating by product quality controller
iC4	Fresh Feed	$k_F V_R Cn_{C4}$	Product stream	0	Controlled by product quality controller
nC4	Fresh Feed	0	0	$k_F V_R Cn_{C4}$	Indicated by DIB column base level
iC5	Fresh Feed	0	Purge stream	0	Self-regulating by composition change in purge

Where V_R = reaction volume

Step 8: Control individual unit operations

The previous steps have left us at this point with two unassigned control valves, which are the reflux flows to each column (case of the DIB reflux flow is not used). We do not need dual composition control for irreversible case because only one end of both columns is a product stream leaving the process. These two reflux flowrates are available in the step 9 to use as optimizing variables or to improve dynamic response.

Step 9: Optimize economics or improve dynamic controllability

When we use reactor inlet temperature for production rate control (irreversible case), the only remaining degrees of freedom for optimization are the reflux flows for the two columns and the setpoint of the distillate flowrate from the purge column (recycle flow).

5.2 Design of plantwide control structures

In this current work, we apply the first control structure of Luyben (1998) namely control structure 1 (CS1), the second control structure of Luyben (1998) namely control structure 2 (CS2), the third control structure of Kunajitpimol (2006) namely control structure 3 (CS3) and the forth control structure of Kunajitpimol (2006) namely control structure 4 (CS4) to the Butane Isomerization process with energy integration schemes for RHEN-1, RHEN-2, RHEN-3, RHEN-4, RHEN-5, RHEN-6, RHEN-7 and Base Case.. The new plantwide control structures CS5, CS6, CS7 and CS8 are designed for all processes. In all of these control structures, the same loops are used as follows:

- The DIB column reboiler level is controlled by manipulating the fresh feed valve.

- The Purge column reboiler level is controlled by manipulating the Purge column feed valve.

- The flowrate of recycle stream is controlled by manipulating the Purge column distillate valve.

- The Purge column bottom frowrate is controlled by manipulating the Purge column bottom valve.

- The flooded condenser pressure is controlled by manipulating the DIB column feed recycle valve.

- The flooded condenser outlet temperature is controlled by manipulating the cooler duty.

- The reactor inlet temperature is controlled by manipulating the furnace duty.

- The DIB column pressure is controlled by manipulating the DIB column condenser duty.

- The Purge column pressure is controlled by manipulating the Purge column condenser duty.

- The Purge column condenser level is controlled by manipulating the Purge column reboiler duty.

5.2.1 Reference control structure I (CS1)

For DIB column of this control structure, the impurity of nC4 in the product stream is controlled by manipulating the DIB column reboiler duty. The DIB column reflux drum level is controlled by manipulating the DIB column distillate valve. The reflux flow is fixed with reflux valve. This control structure is designed to reduce the effects of disturbance in order to achieved impurity of normal butane in product and desired production rate. For the Purge column, the condenser level is controlled by manipulating the Purge column reboiler duty.

5.2.2 Reference control structure II (CS2)

This control structure develops from CS1. A ratio controller is used to increase the reflux flowrate in the DIB column. The fractionating capability of the column increases as the load is increased, so the bottom does not contain less nC4 reactant. Product rate increases. The other control structures are the same as CS1.

5.2.3 Reference control structure III (CS3)

This control structure uses the DIB column distillate valve to control impurity liquid of nC4 in the top of DIB column (tray 50), DIB column reflux flow is manipulated to control reflux drum level, and DIB column reboiler duty is manipulated to control the temperature on tray 1 of DIB column. Alternatively we could control the amount of the nC4 in the purge column bottoms by manipulating purge column reboiler duty. Since there is a fairly large temperature change in the purge column, controlling the temperature on some suitable trays (tray 1 to 7) are controlled at the average temperature. We choose the level of purge column condenser is controlled by cascade to the tray temperature controller is relative to the vapor coming overhead from the top of the column.

5.2.4 Reference control structure IV (CS4)

This control structure develops from CS3. Due to both the DIB column distillate valve which is manipulated to control impurity liquid of nC4 in the top of DIB column (tray50) and DIB column reflux flow which is manipulated to control reflux drum level obtain poor performance. Because the distillate flowrate has large changes, this control structure switches those two variables. The other control structures are the same as CS3.

5.2.5 Design of control structure V (CS5)

This control structure develops from CS1 which we control the impurity of nC4 in the product stream by manipulating the DIB column reboiler duty. The DIB column reflux drum level is controlled by manipulating the DIB column distillate valve. The reflux flow is fixed with reflux valve. For the Purge column, we want to control the amount of nC4 in the purge column bottoms by manipulating purge column reboiler duty. Since there is a fairly large temperature change in the purge column, controlling the temperature on some suitable trays (tray 1 to 7) are controlled at the average temperature. The purge column reflux drum level is controlled by cascade to the tray temperature controller is relative to the vapor coming overhead from the top of the column.

5.2.6 Design of control structure VI (CS6)

This control structure develops from CS2, i.e., the control structures of the DIB column are the same as CS2. For the control structures of the purge column differ from CS2, which the purge column is controlled the average temperature on tray 1 to 7 by manipulating the purge column reboiler duty to achieve the desired nC4 in the column bottoms. So the purge column reflux drum level is controlled by cascade to the tray temperature controller.

5.2.7 Design of control structure VII (CS7)

This control structure develops from CS3. Because we want to evaluate the importance of temperature in the purge column controlling of CS3 so we control the tray temperature in DIB column while the Purge column does not control. The Purge column reflux drum level is controlled by manipulating the Purge column reboiler duty.

5.2.8 Design of control structure VIII (CS8)

This control structure develops from CS4. The control structures of DIB column are the same as CS4 but the control structure of Purge column is different. The temperature in the Purge column is not controlled because we want to evaluate the importance of the tray temperature controlling.

For all of the control structures, we apply them to the 8 alternatives of the heat exchanger networks (Base case and RHEN-1 to 7). Because the RHEN-7 has not the Purge column reboiler so we can not to use the purge column reboiler duty to be manipulated variable. At this constraint, there are 4 control structures to apply to RHEN-7 (CS1-4). Furthermore, the control structure of the Purge column in CS1 and CS2 are different from the other case that we use the bypass valve to control the purge column reflux drum level by cascade to the temperature controller.

So, there are 60 alternatives of the heat exchanger networks with control structures for Butane Isomerization plant include Base case with 8 control structures, RHEN-1 to 6 with 8 control structures and RHEN-7 with 4 control structures as show in Figures 5.12 to 5.71

5.3 Energy management of heat integrated Butane Isomerization Plant

As the operating conditions change, the designed control system must regulate the entire process to meet the desired condition. On the other hand, changes in the heat load disturbance of the cold or hot stream affect energy consumption of its unity units. Therefore, for a complex energy-integrated plant, it is important to study the heat pathway control in order to manage the heat load disturbance in such a way that the maximum energy recovery (MER) can always be achieved.

We now look at the plantwide control issues around energy management. The control configurations of RHEN are determined using the Heat Pathway Heuristics (HPH) (Wongsri and Hermawan, 2005). The objective of HPH design is to find proper heat pathways to achieve the dynamic HEN operation objective which is desired target variables and maximum energy recovery. As the operating conditions change or heat load disturbances enter, the designed control system must regulate the heat flow within the network to meet the desired goal.

HPH is used in design and operation of RHEN. HPH is about how to properly direct heat load disturbance throughout the network to heat sinks or heat sources in order to achieve MER at all time. First two kinds of disturbances is needed to be introduced: Positive disturbance load, D+, an entering disturbance resulting in increasing heat load of a stream; Negative disturbance load, D--, an entering disturbance resulting in decreasing heat load of a stream. D+ of a hot stream and D-- of a cold stream must be directed to heaters and vice versa for D-- of a hot stream and D+ of a cold stream. The heat pathway should be short to minimize the input and propagated disturbances, simply a path with minimized upsets.

5.3.1 Heat Pathways and HEN control configuration designs for RHEN-1 and RHEN-3

The design of the heat pathways for RHEN-1 and RHEN-3 shown in Figure 5.2 shifts the positive and negative disturbance loads of C1 to furnace. Thus, the negative disturbance load of a cold stream will result in decrease of the furnace duty which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C2 are shifted to the auxiliary reboiler of DIB column as in RHEN-1 or Purge column for RHEN-3. As a result, the auxiliary reboiler duty of the engaged column is decreased or increased subject to the type of the disturbances. The negative or positive disturbance load of H1 is directed to the furnace; the furnace duty of corresponding column is increased or decreased accordingly.

Note that RHEN-1 is similar to RHEN-3, except C2 (reboiler feed of DIB column) is replaced by C3 (reboiler feed of Purge column).

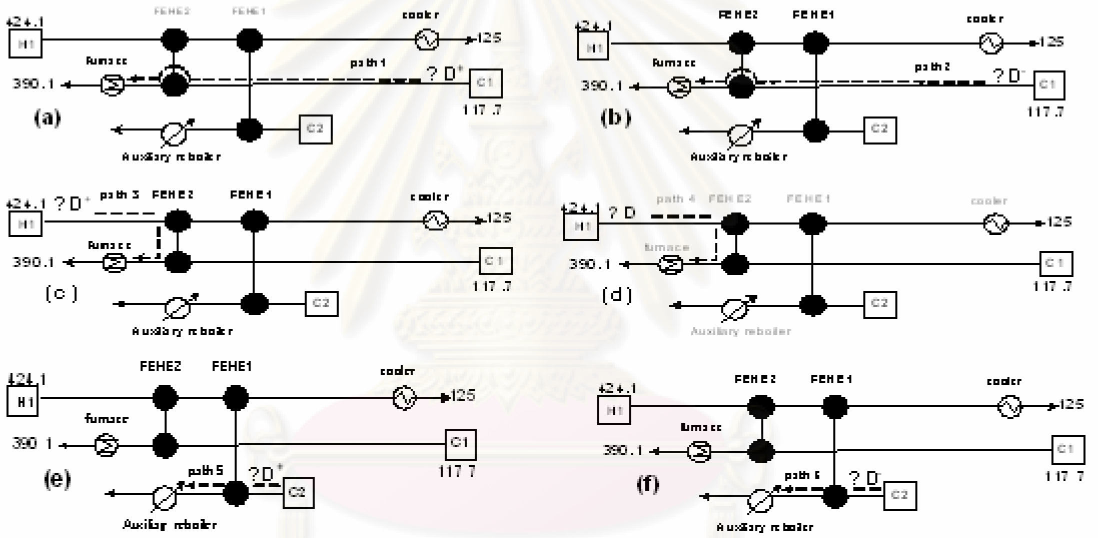


Figure 5.2 Heat pathways through RHEN-1 and RHEN-3, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to the furnace, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the furnace , (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to the furnace, (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to the furnace, (e) path 5 is used to shift the positive disturbance load of the cold stream C_2 to the auxiliary reboiler and (f) path 6 is used to shift the negative disturbance load of the cold stream C_2 to the auxiliary reboiler

From designed the heat pathways for RHEN-1 and RHEN-3, we can design the control configurations as show in Figure 5.3. These control systems involve one manipulated variable and one controlled variable and work as follows: the hot outlet temperature of FEHE2 is controlled at its nominal set point by manipulating the valve on the bypass line (VBP2). At the same time, the hot outlet temperature of FEHE2 should not be allowed to drop below a lower value, which is necessary to keep the furnace duty at a good level. Whenever the hot outlet temperature of FEHE2 drops below the allowable limit due to, for example, a negative disturbance load entering the hot stream H1, the control action to the hot temperature control (TC-hHE2out) open the valve VBP2. As a result, the hot outlet temperature of FEHE2 will rise to its normal temperature and the cold outlet temperature of FEHE2 will be further decreased, so the furnace duty will also be increased but, the auxiliary reboiler duty will be kept in the same as duty in the condition without-disturbance. Whenever, the hot outlet temperature of FEHE2 increases above a lower limit, i.e., a desired-condition during operation, due to the positive disturbance load entering the hot stream H1, the control action to TC-hHE2out close the valve VBP2. Consequently, the hot outlet temperature of FEHE2 will drop to its normal temperature and the cold outlet temperature of FEHE2 will be increased, so the furnace duty will also be deceased. For another control system, the hot outlet temperature of FEHE1 is controlled at its normal set point by manipulating the valve on the bypass line (VBP1). At the same time, the hot outlet temperature of FEHE1 should not be allowed to drop below a lower limit value, which is necessary to keep the auxiliary reboiler duty at a good level.

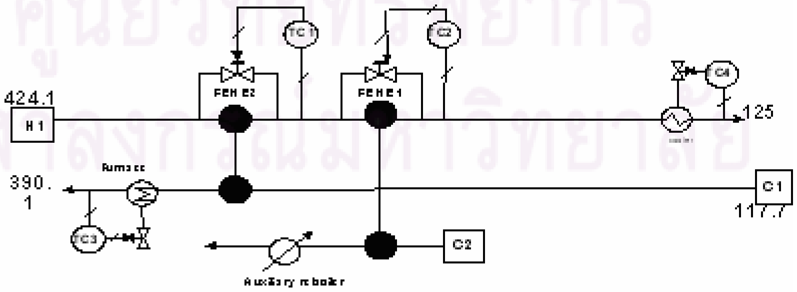


Figure 5.3 Control configurations of RHEN-1 and RHEN-3

5.3.2 Heat Pathways and HEN control configuration designs for RHEN-2 and RHEN-4

The design of the heat pathways for RHEN-2 and RHEN-4 shown in Figure 5.4 shifts the positive and negative disturbance loads of C1 to cooler. Thus, the positive disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load will result in increase of the cooler duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C2 are shifted to the auxiliary reboiler of DIB column as in RHEN-2 or Purge column for RHEN-4. As a result, the auxiliary reboiler duty of the engaged column is decreased or increased subject to the type of the disturbances. The negative or positive disturbance load of H1 is directed to the auxiliary reboiler; the auxiliary reboiler duty of corresponding column is increased or decreased accordingly.

Note that RHEN-2 is similar to RHEN-4, except C2 (reboiler feed of DIB column) is replaced by C3 (reboiler feed of Purge column).

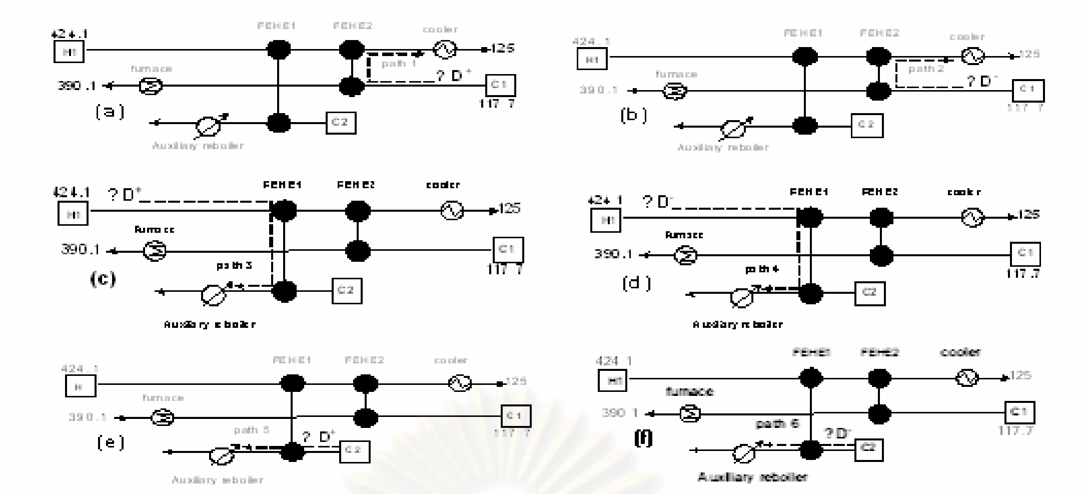


Figure 5.4 Heat pathways through RHEN-2 and RHEN-4 to achieve the highest possible dynamic MER, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to the auxiliary reboiler, (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to the auxiliary reboiler, (e) path 5 is used to shift the positive disturbance load of the cold stream C_2 to the auxiliary reboiler and (f) path 6 is used to shift the negative disturbance load of the cold stream C_2 to the auxiliary reboiler

These control systems involve one manipulated variable and one controlled variable and work as follows: Figure 5.5 shows the hot outlet temperature of FEHE1 is controlled at its nominal set point by manipulating the valve on the bypass line (VBP1). At the same time, the hot outlet temperature of FEHE1 should not be allowed to drop below a lower limit value, which is necessary to keep the auxiliary reboiler duty at a good level. Whenever the hot outlet temperature of FEHE1 drops below the allowable limit due to, for example, a negative disturbance load entering the hot stream H1, the control action to the hot temperature control (TC-hHE1out) open the valve VBP1. As a result, the hot outlet temperature of FEHE1 will rise to its normal temperature and the cold outlet temperature of FEHE1 will be further decreased, so the auxiliary reboiler duty will also be increased but, the furnace duty will be kept in the same as duty in the condition without-disturbance. If the hot outlet temperature of FEHE1 increases above a lower limit, i.e., a desired-condition during operation, due to the positive disturbance load entering the hot stream H1, the control action to TC-cHE1out close the valve VBP1. Consequently, the hot outlet temperature of FEHE1 will drop to its normal temperature and the cold outlet temperature of FEHE1 will be increased, so the auxiliary reboiler duty will also be deceased. Another control system, the cold outlet temperature of FEHE2 is controlled at its normal set point by manipulating the valve on the bypass line (VBP2). At the same time, the cold outlet temperature of FEHE2 should not be allowed to rise above an upper limit valve, which is necessary to keep the cooler duty at a good level.

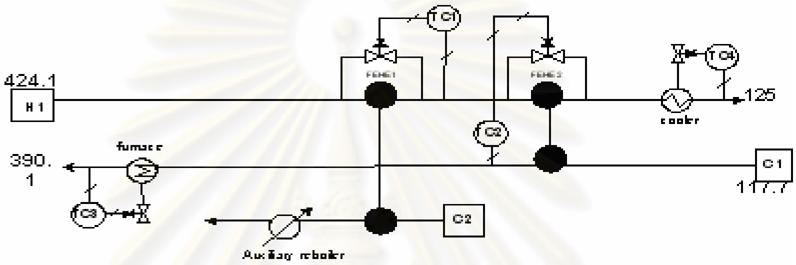


Figure 5.5 Control configurations of RHEN-2 and RHEN-4

5.3.3 Heat Pathways and HEN control configuration designs for RHEN-5

The design of the heat pathways for RHEN-5 shown in Figures 5.6 shifts the positive and negative disturbance loads of C2 to auxiliary reboiler of DIB column. Thus, the negative disturbance load of a cold stream will result in decrease of the auxiliary reboiler duty which is good. The positive disturbance load will result in increase of the auxiliary reboiler duty which is ruled by T_{min} constraint. The negative or positive disturbance load of H1 is directed to the auxiliary reboiler; the auxiliary reboiler duty of corresponding column is increased or decreased accordingly.

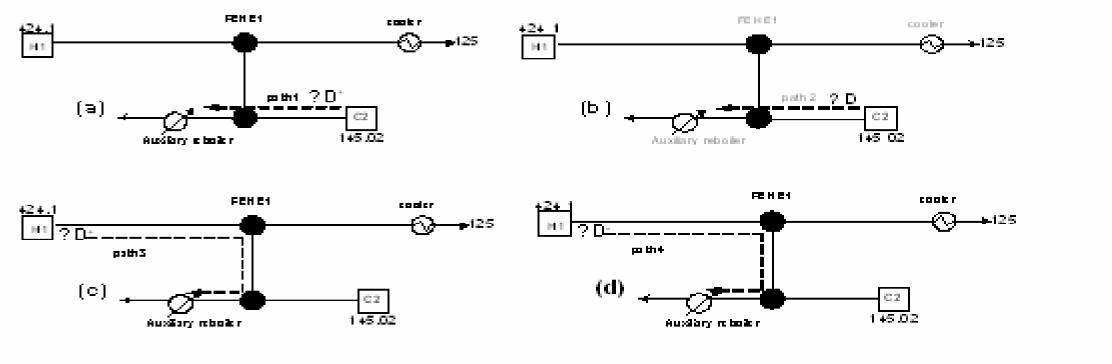


Figure 5.6 Heat pathways through RHEN-5, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_2 to the auxiliary reboiler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_2 to the auxiliary reboiler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to the auxiliary reboiler and (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to the auxiliary reboiler.

The control systems for RHEN-5 work as follows: Figure 5.7 shows the hot outlet temperature of FEHE1 is controlled at its nominal set point by manipulating the valve on the bypass line (VBP1). At the same time, the hot outlet temperature of FEHE1 should not be allowed to drop below a lower limit value, which is necessary to keep the auxiliary reboiler duty at a good level. Whenever the hot outlet temperature of FEHE1 drops below the allowable limit due to, for example, a negative disturbance load entering the hot stream H1, the control action to the hot temperature control (TC-hHE1out) open the valve VBP1. As a result, the hot outlet temperature of FEHE1 will rise to its normal temperature and the cold outlet temperature of FEHE1 will be further decreased, so the auxiliary reboiler duty will also be increased. If the hot outlet temperature of FEHE1 increases above a lower limit, i.e., a desired-condition during operation, due to the positive disturbance load entering the hot stream H1, the control action to TC-hHE1out close the valve VBP1. Consequently, the hot outlet temperature of FEHE1 will drop to its normal temperature and the cold outlet temperature of FEHE1 will be increased, so the auxiliary reboiler duty will also be decreased. At the same time, the hot outlet temperature of FEHE1 should not be allowed to drop below a lower limit value, which is necessary to keep the cooler duty at a good level.

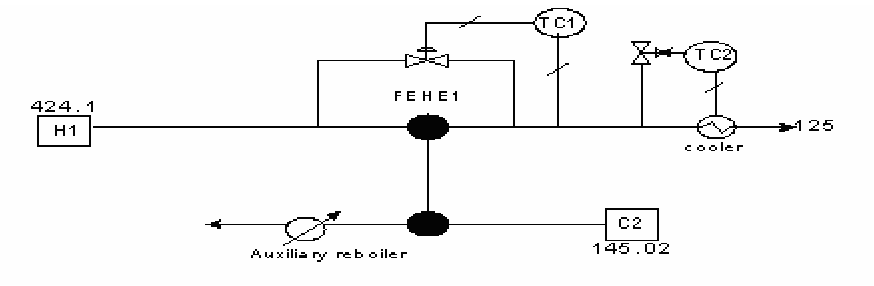


Figure 5.7 Control configuration of RHEN-5

5.3.4 Heat Pathways and HEN control configuration designs for RHEN-6

The design of the heat pathways for RHEN-6 shown in Figures 5.8 shifts the positive and negative disturbance loads of C1 to cooler. Thus, the positive disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load will result in increase of the cooler duty due to this case has not a furnace to support the heat to the reactor. Because the pinch temperature for this heat exchanger network is necessary including the furnace is not used, the negative or positive disturbance loads of H1 are shifted to the cooler.

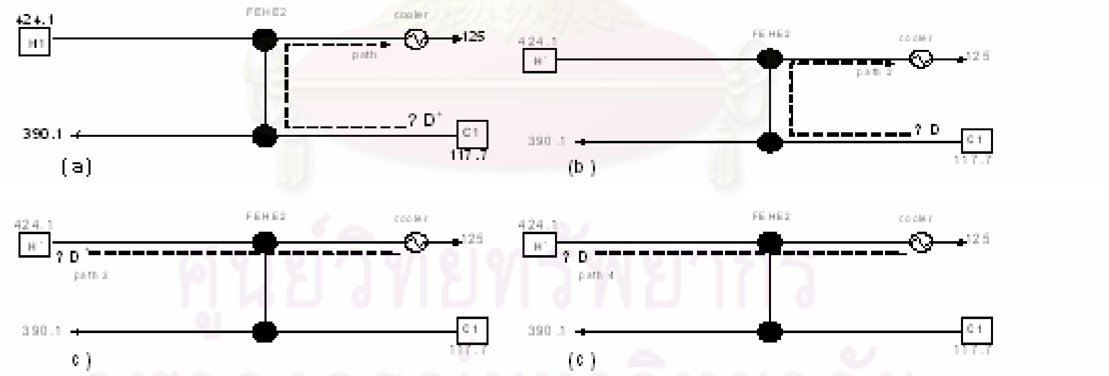


Figure 5.8 Heat pathways through RHEN-6, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to the cooler and (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to the cooler.

The control systems for RHEN-6 work as follows: Figure 5.9 shows the cold outlet temperature of FEHE1 is controlled at its nominal set point by manipulating the valve on the bypass line (VBP1). Whenever the cold outlet temperature of FEHE1 drops below the allowable limit due to, for example, a negative disturbance load entering the hot stream H1, the control action to the cold temperature control (TC-cHE1out) close the valve VBP1. As a result, the cold outlet temperature of FEHE1 will rise to its normal temperature and the hot outlet temperature of FEHE1 will be further decreased, so the cooler duty will also be decreased. If the cold outlet temperature of FEHE1 increases above a lower limit, i.e., a desired-condition during operation, due to the positive disturbance load entering the hot stream H1, the control action to TC-cHE1out open the valve VBP1. Consequently, the cold outlet temperature of FEHE1 will drop to its normal temperature and the hot outlet temperature of FEHE1 will be increased, so the cooler duty will also be increased. At the same time, the cold outlet temperature of FEHE1 should not be allowed to rise above an upper limit value, which is necessary to keep the cooler duty at a good level.

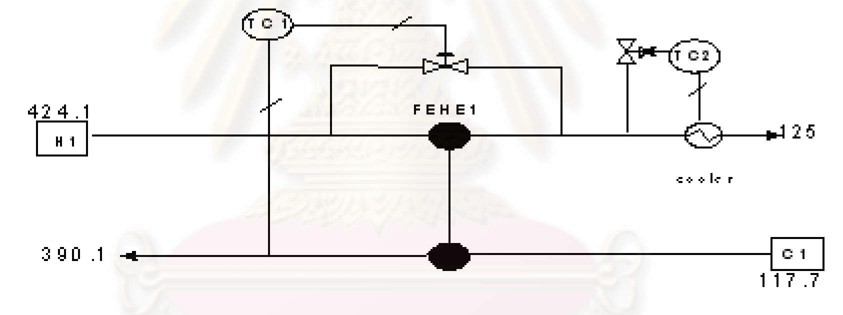


Figure 5.9 Control configuration of RHEN-6

5.3.5 Heat Pathways and HEN control configuration designs for RHEN-7

The design of the heat pathways for RHEN-7 shown in Figures 5.10 shifts the positive and negative disturbance loads of C3 to cooler. Thus, the positive disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load will result in increase of the cooler duty due to this case has not the auxiliary reboiler for Purge column to support the heat to the reactor. Because the pinch temperature for this heat exchanger network is necessary including reboiler for Purge column is not used, the negative or positive disturbance loads of H1 are shifted to the cooler.

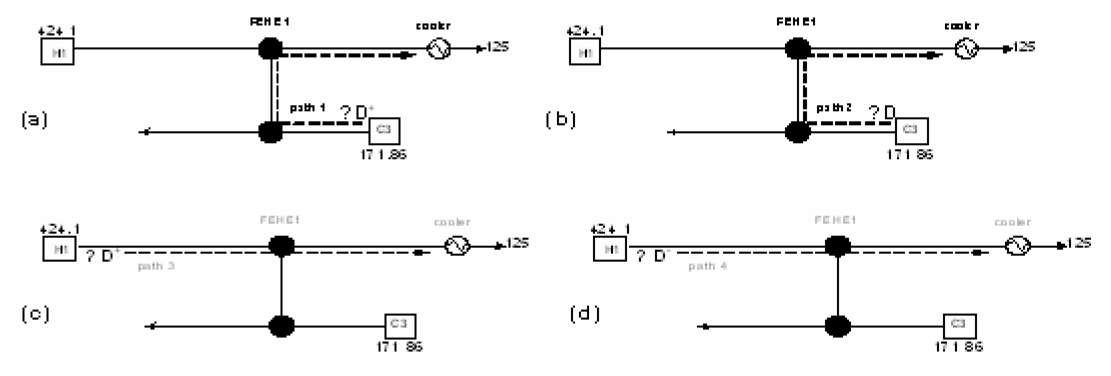


Figure 5.10 Heat pathways through RHEN-7, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_3 to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_3 to the cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to the cooler and (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to the cooler.

The control systems for RHEN-7 work as follows: Figure 5.11 shows the cold outlet temperature of FEHE1 is controlled at its nominal set point by manipulating the valve on the bypass line (VBP1). Whenever the cold outlet temperature of FEHE1 drops below the allowable limit due to, for example, a negative disturbance load entering the hot stream H1, the control action to the cold temperature control (TC-cHE1out) close the valve VBP1. As a result, the cold outlet temperature of FEHE1 will rise to its normal temperature and the hot outlet temperature of FEHE1 will be further decreased, so the cooler duty will also be decreased. If the cold outlet temperature of FEHE1 increases above a lower limit, i.e., a desired-condition during operation, due to the positive disturbance load entering the hot stream H1, the control action to TC-cHE1out open the valve VBP1. Consequently, the cold outlet temperature of FEHE1 will drop to its normal temperature and the hot outlet temperature of FEHE1 will be increased, so the cooler duty will also be increased. At the same time, the cold outlet temperature of FEHE1 should not be allowed to rise above an upper limit value, which is necessary to keep the cooler duty at a good level.

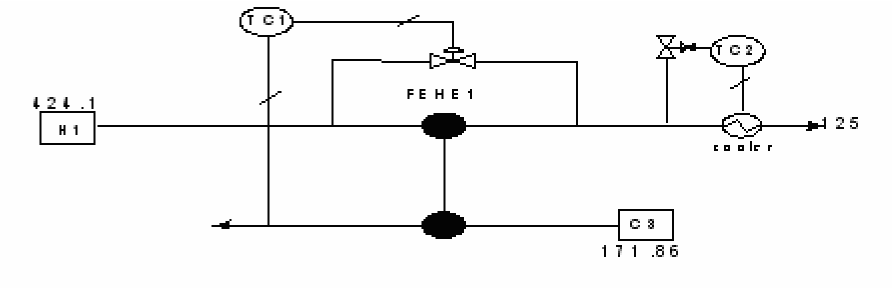


Figure 5.11 Control configuration of RHEN-7

For all the heat integration units, the bypass streams are designed to control the outlet temperatures of FEHEs. The bypass stream should be about 5 to 10 percent of the total flow to be able to handle disturbances (Jones and Wilson, 1997). In normal operation, a control valve should operate with an opening between 20 to 80 percent (Jones and Wilson, 1997). In our study, the bypass valves in the process-to-process-heat-exchangers are designed with the valve opening of 50%, i.e. this translates into the bypass flow rate of about 5% of the total flow. In practice we have to overdesign the process-to-process-heat-exchanger, in order to be able to handle the disturbances. In this work, it is not our intention to study the best overdesign policy. The oversize of the heat exchanger is related to the estimated maximum size of disturbance loads of both the cold and hot streams. The size of disturbance in this study is about 5 to 10% according to Luyben’s recommendations.

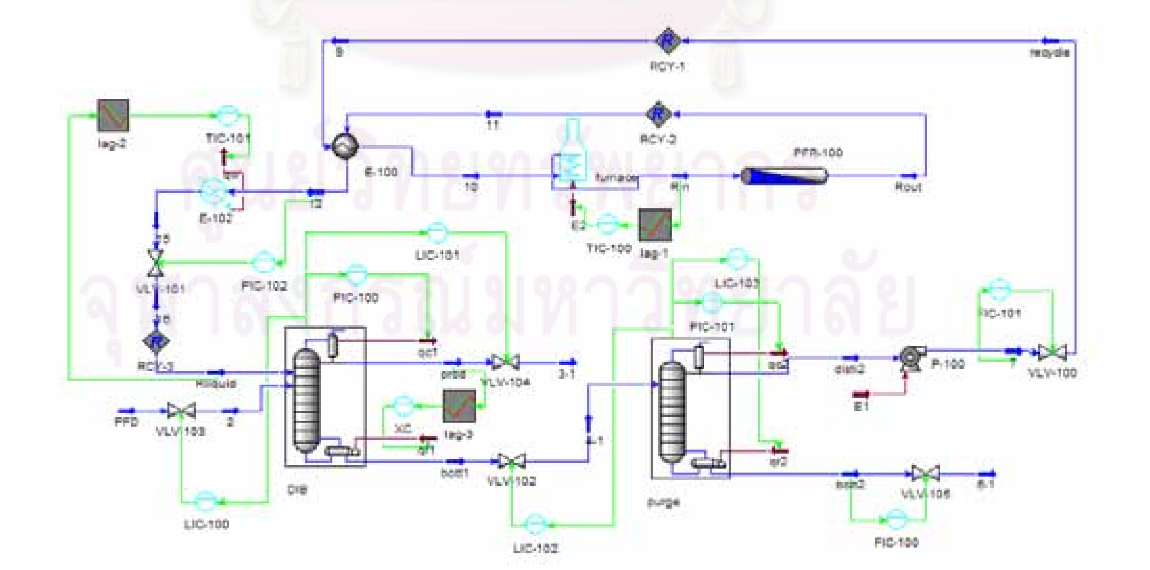


Figure 5.12 Application of reference control structure 1 (CS1) to the Butane Isomerization plant (Base Case)

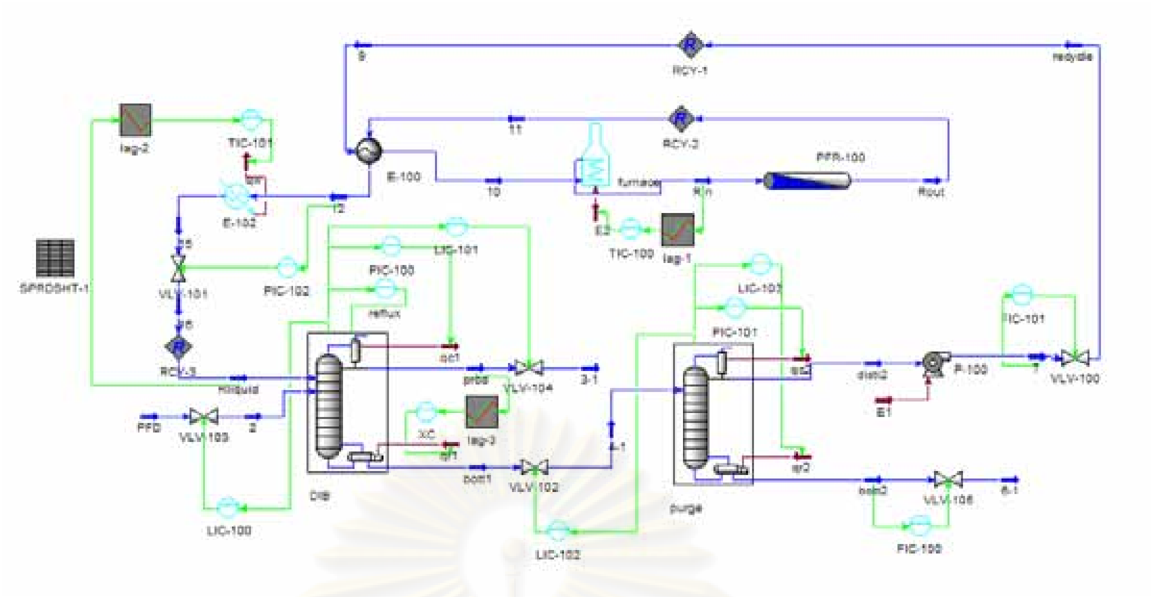


Figure 5.13 Application of reference control structure 2 (CS2) to the Butane Isomerization plant (Base Case)

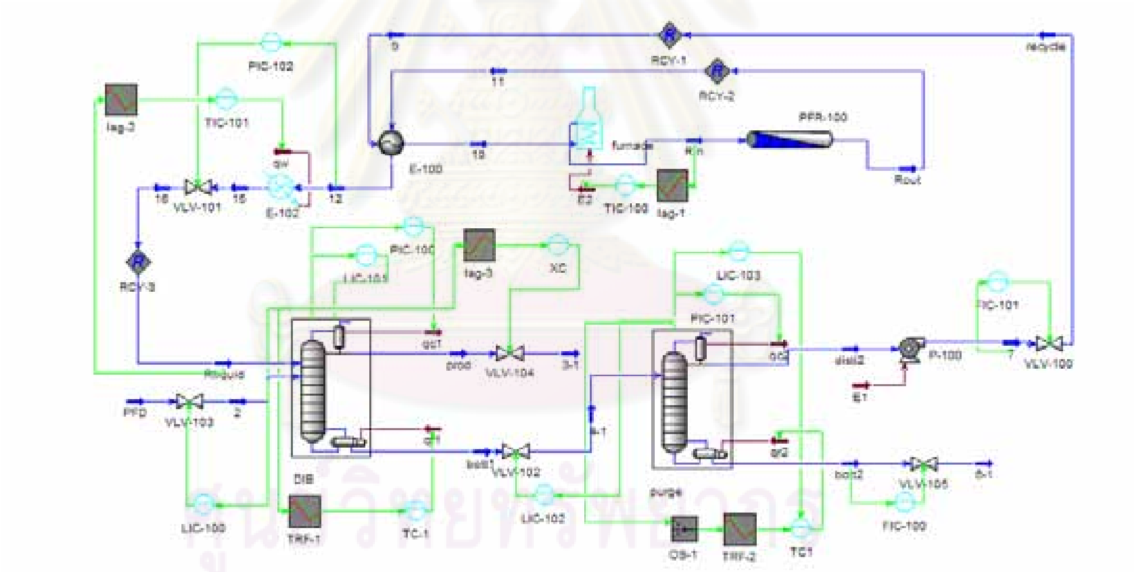


Figure 5.14 Application of reference control structure 3 (CS3) to the Butane Isomerization plant (Base Case)

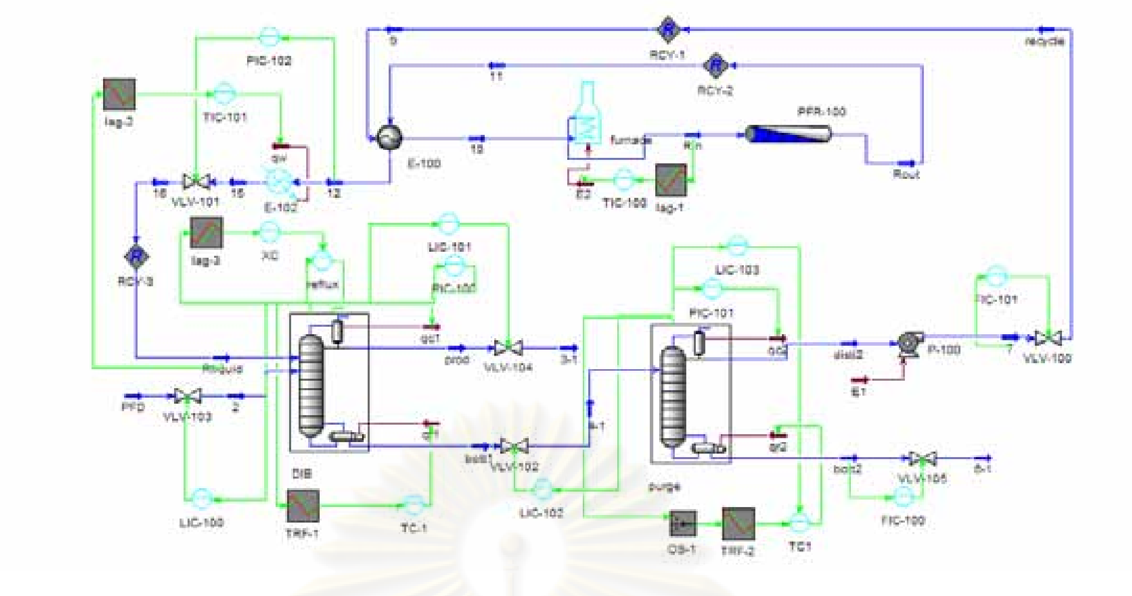


Figure 5.15 Application of reference control structure 4 (CS4) to the Butane Isomerization plant (Base Case)

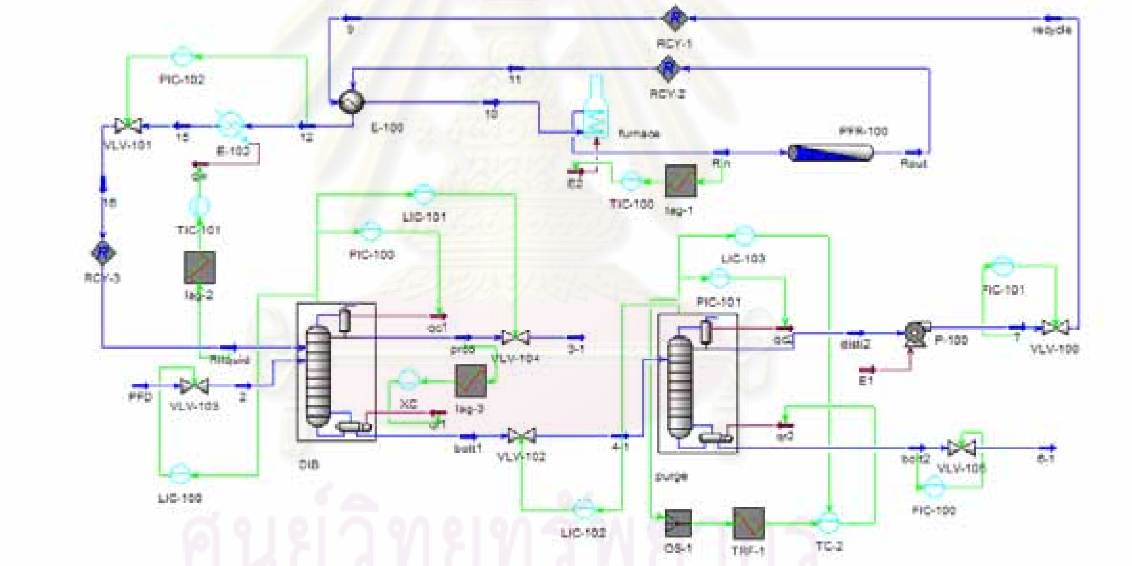


Figure 5.16 Application of reference control structure 5 (CS5) to the Butane Isomerization plant (Base Case)

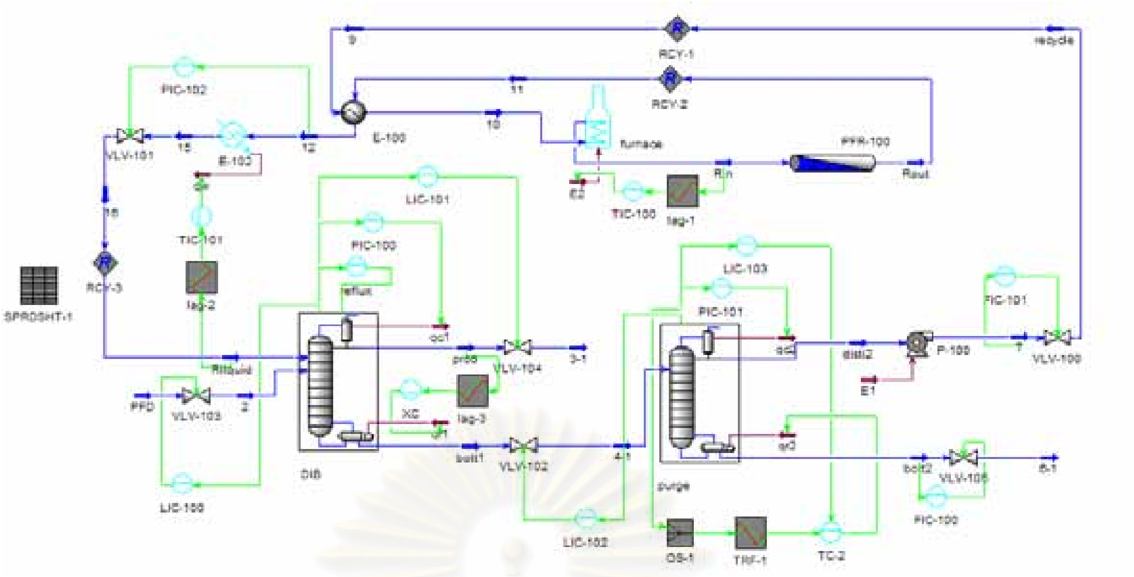


Figure 5.17 Application of reference control structure 6 (CS6) to the Butane Isomerization plant (Base Case)

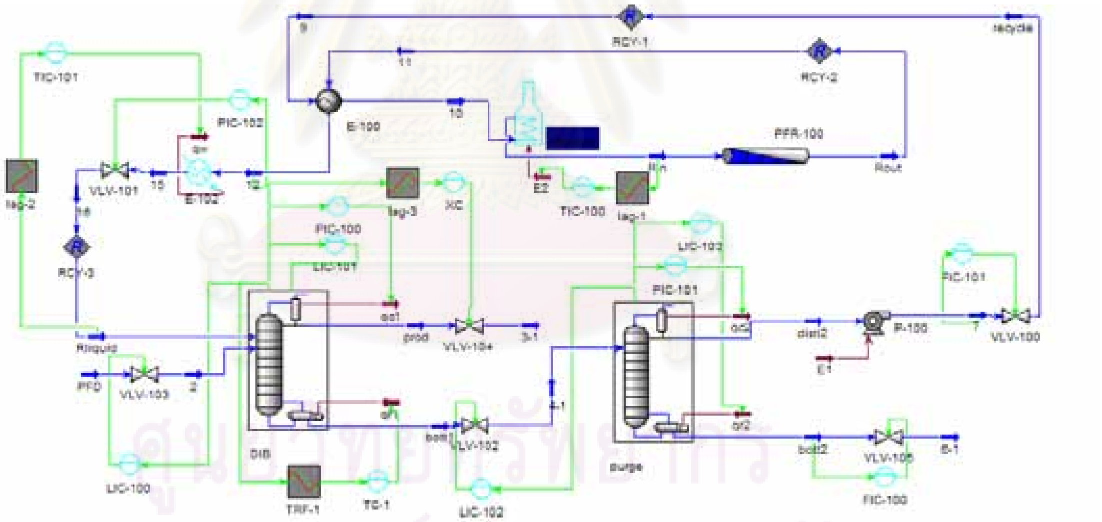


Figure 5.18 Application of reference control structure 7 (CS7) to the Butane Isomerization plant (Base Case)

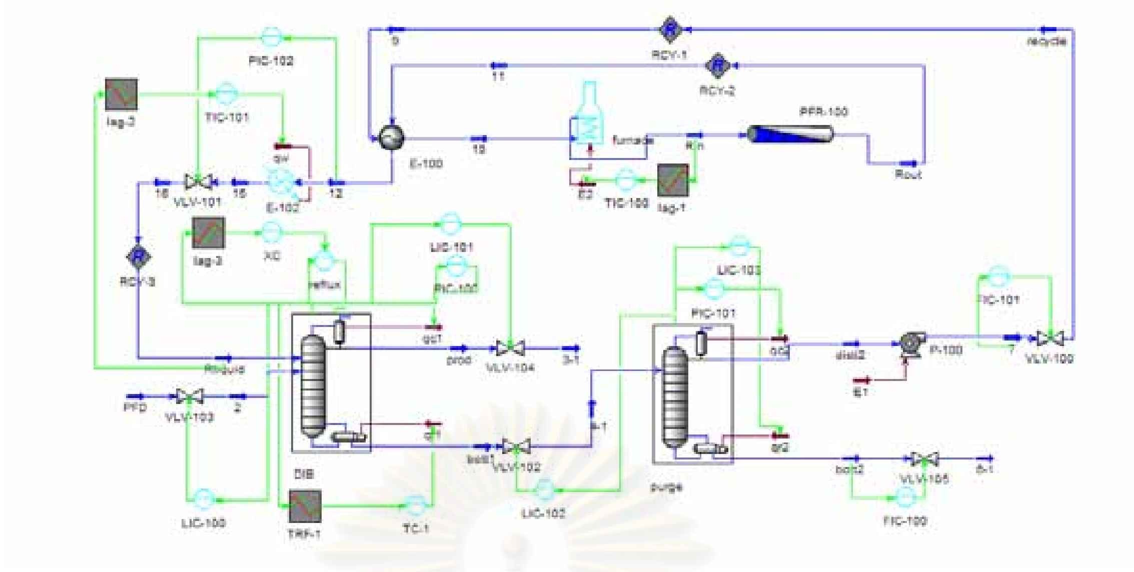


Figure 5.19 Application of reference control structure 8 (CS8) to the Butane Isomerization plant (Base Case)

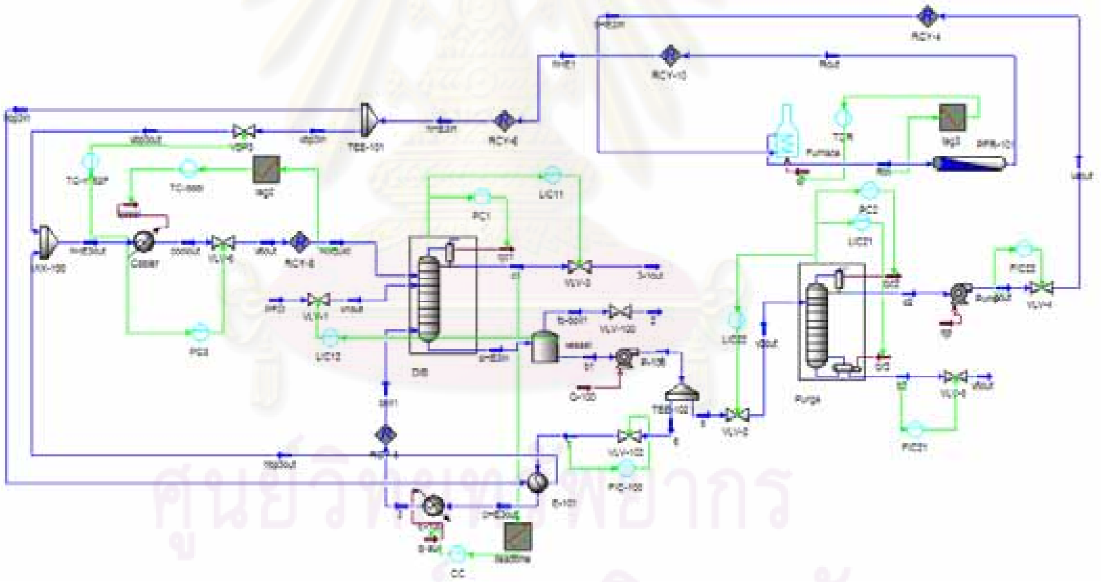


Figure 5.20 Application of reference control structure 1 (CS1) to the Butane Isomerization plant alternative 5 (RHEN-5)

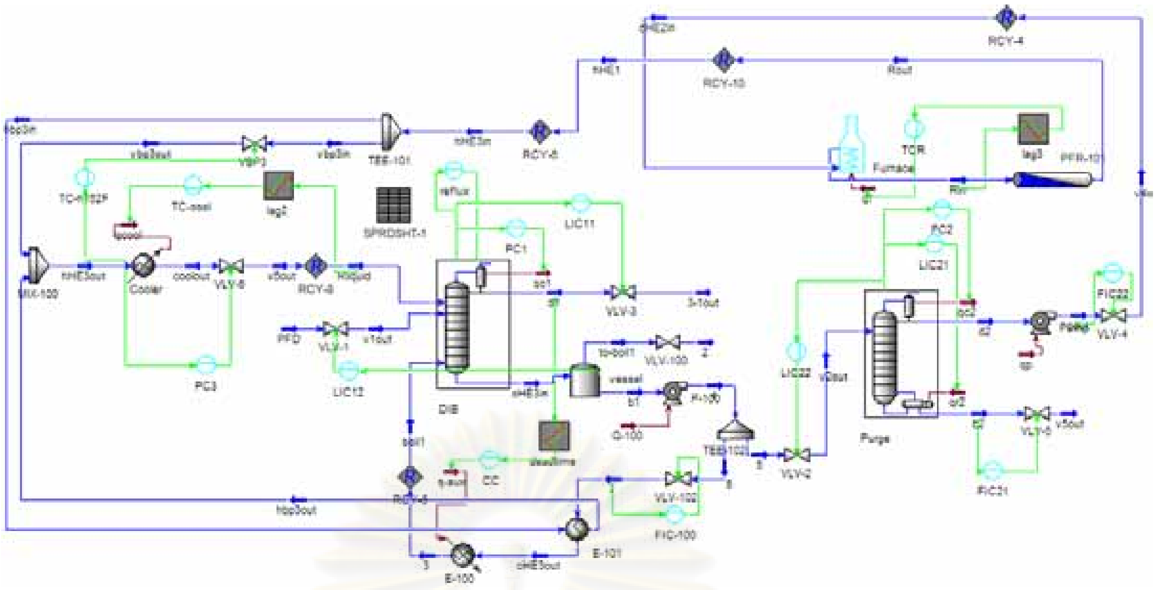


Figure 5.21 Application of reference control structure 2 (CS2) to the Butane Isomerization plant alternative 5 (RHEN-5)

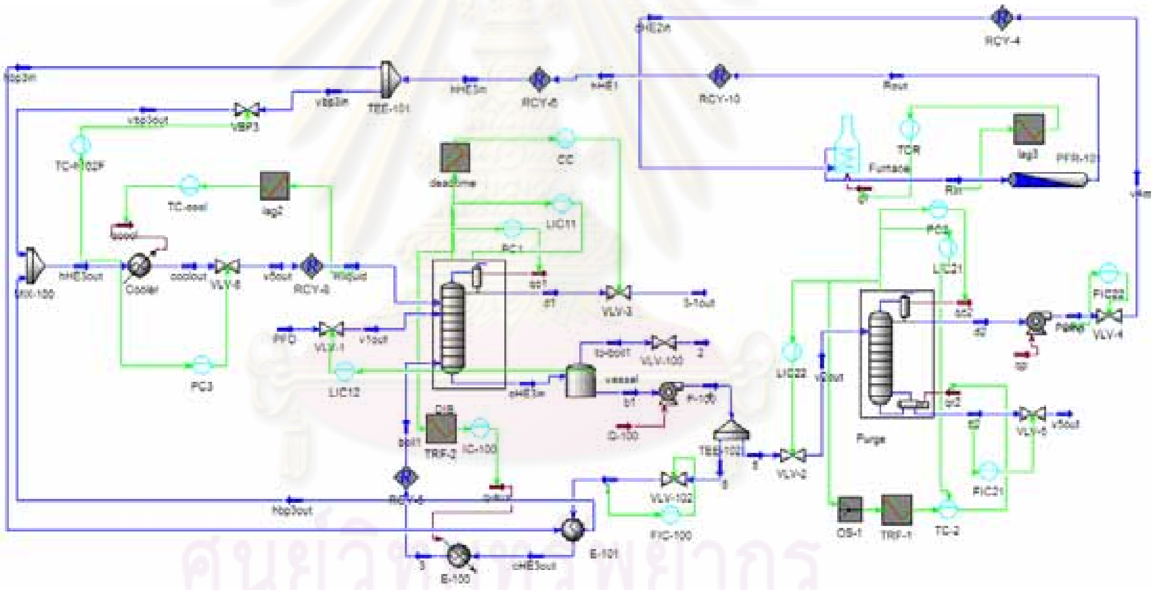


Figure 5.22 Application of reference control structure 3 (CS3) to the Butane Isomerization plant alternative 5 (RHEN-5)

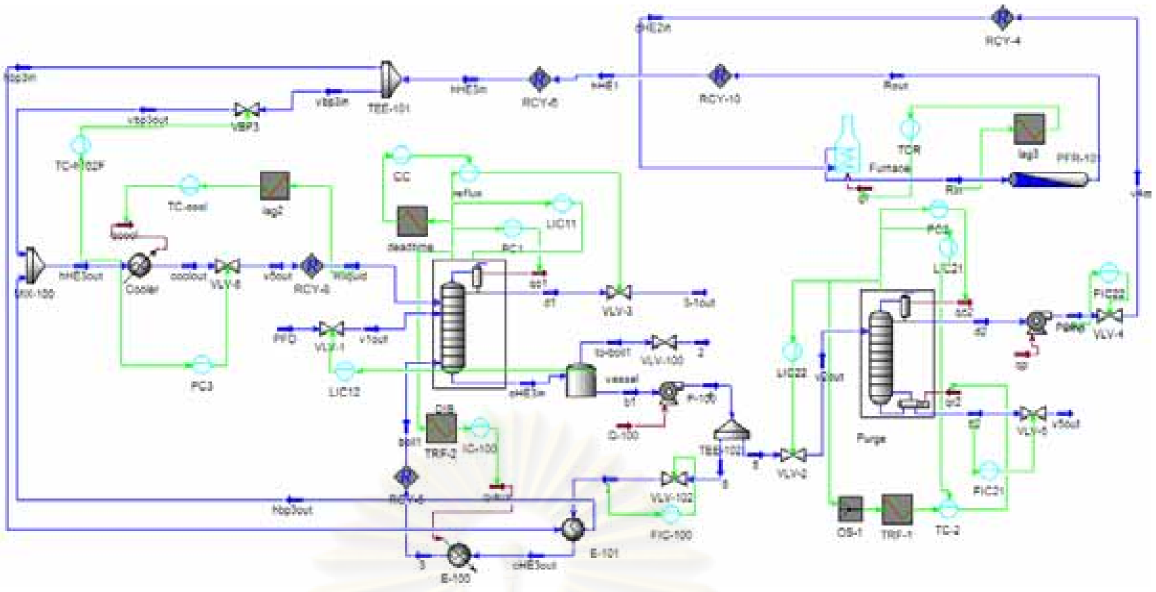


Figure 5.23 Application of reference control structure 4 (CS4) to the Butane Isomerization plant alternative 5 (RHEN-5)

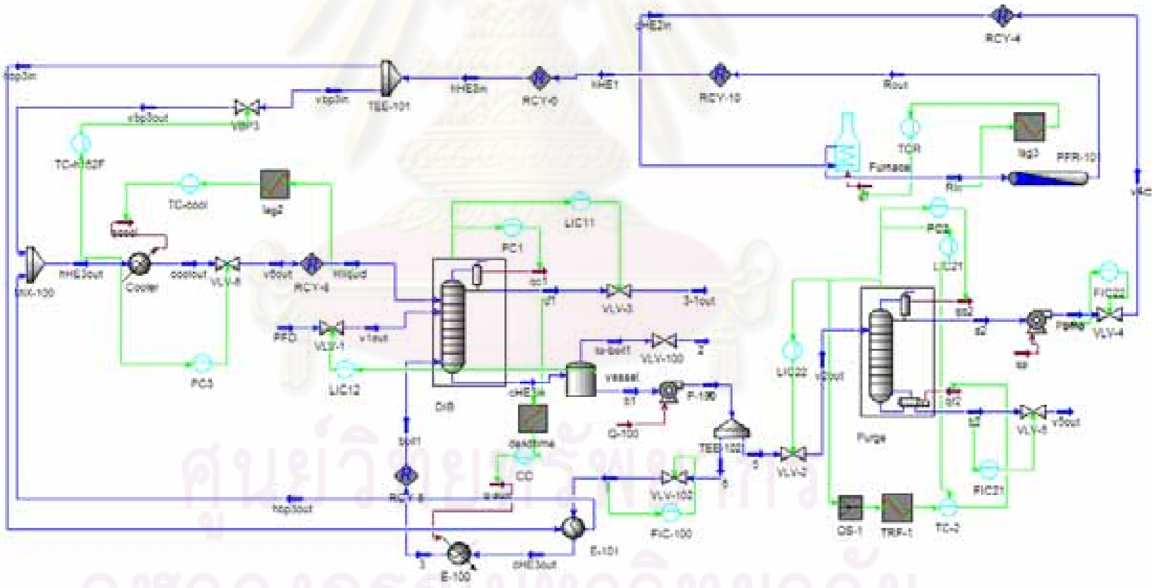


Figure 5.24 Application of reference control structure 5 (CS5) to the Butane Isomerization plant alternative 5 (RHEN-5)

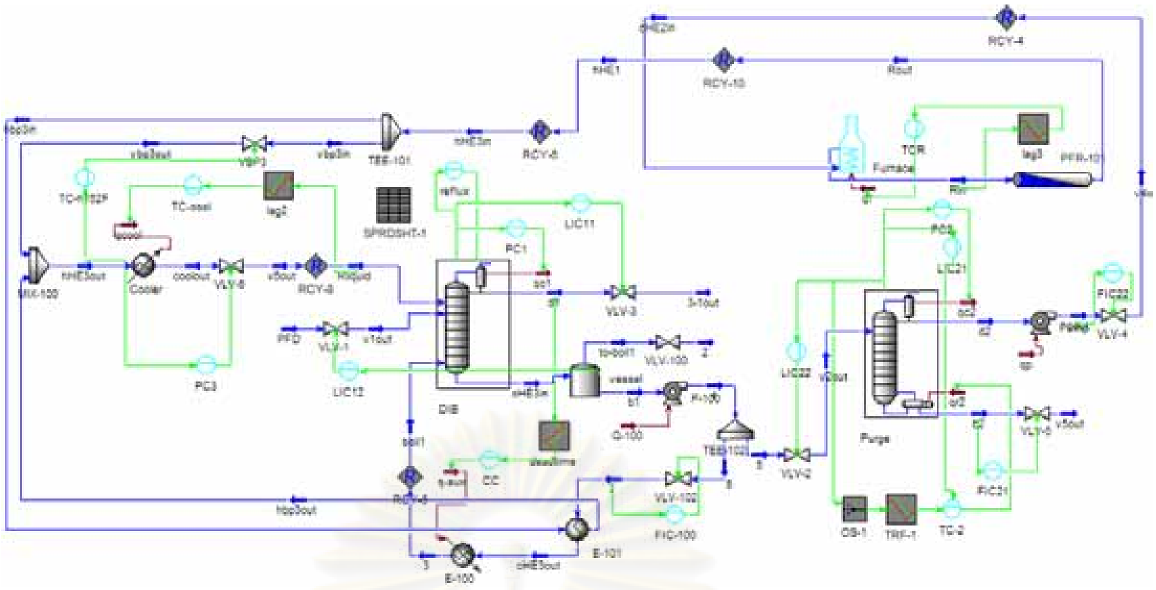


Figure 5.25 Application of reference control structure 6 (CS6) to the Butane Isomerization plant alternative 5 (RHEN-5)

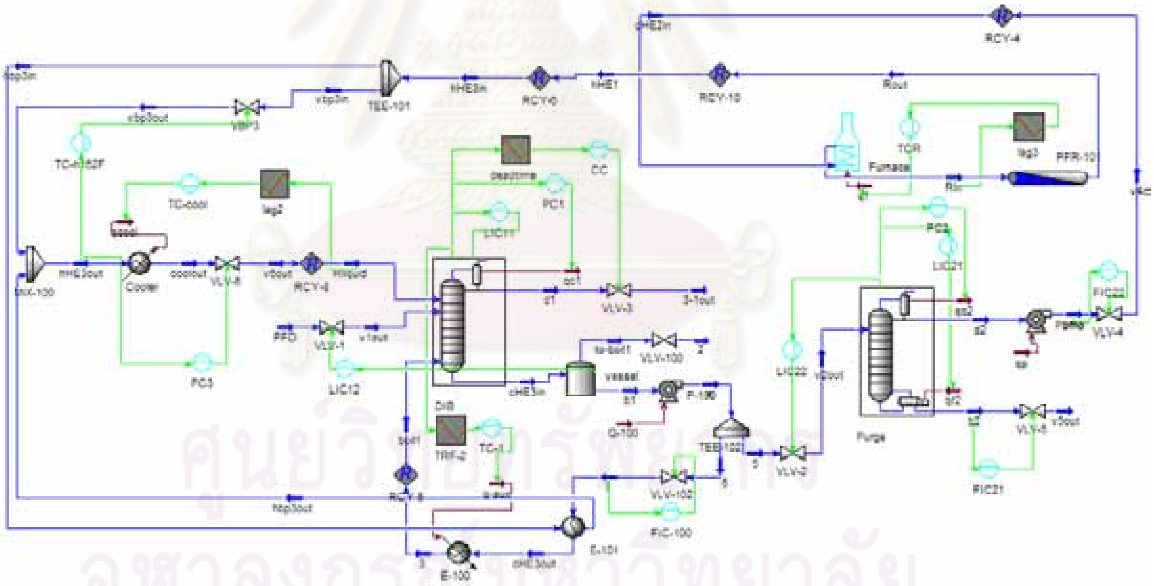


Figure 5.26 Application of reference control structure 7 (CS7) to the Butane Isomerization plant alternative 5 (RHEN-5)

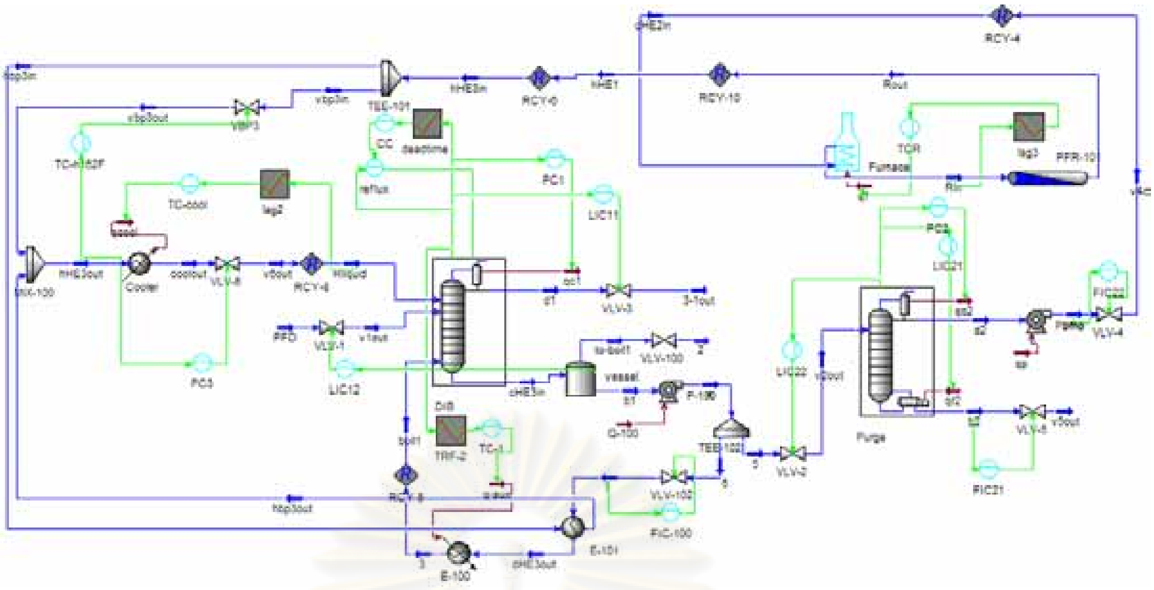


Figure 5.27 Application of reference control structure 8 (CS8) to the Butane Isomerization plant alternative 5 (RHEN-5)

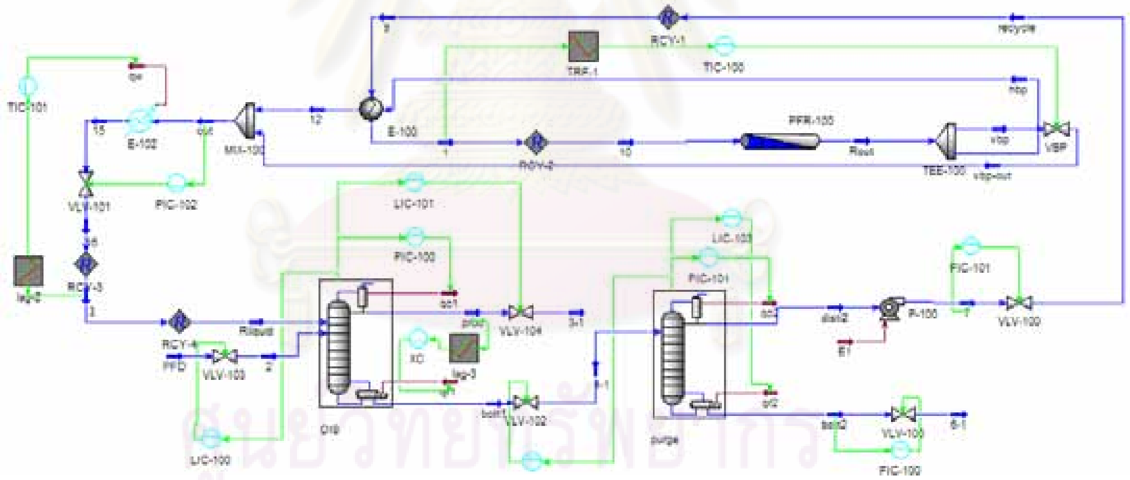


Figure 5.28 Application of reference control structure 1 (CS1) to the Butane Isomerization plant alternative 6 (RHEN-6)

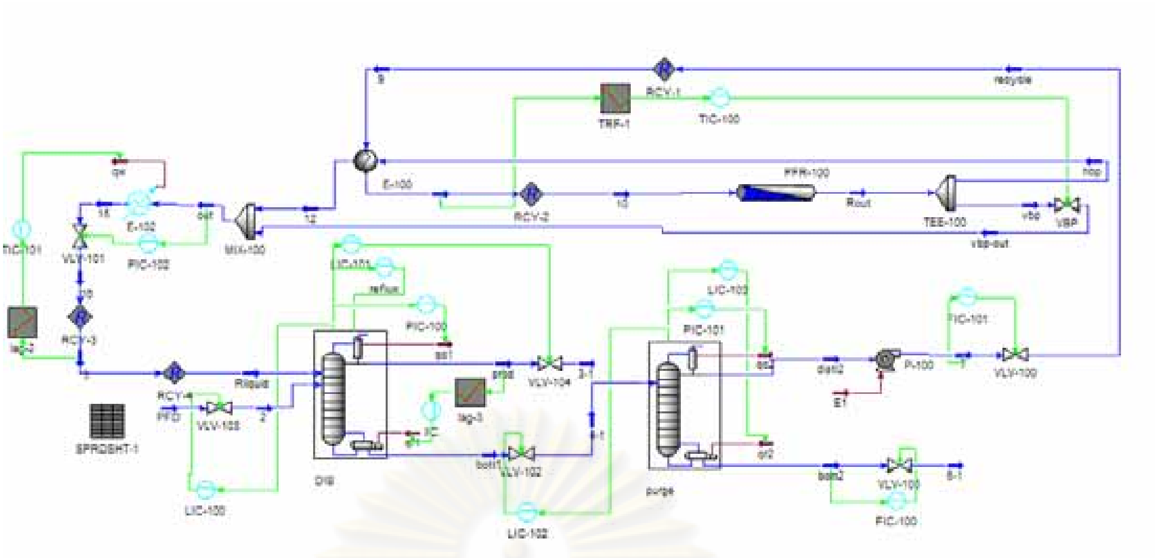


Figure 5.29 Application of reference control structure 2 (CS2) to the Butane Isomerization plant alternative 6 (RHEN-6)

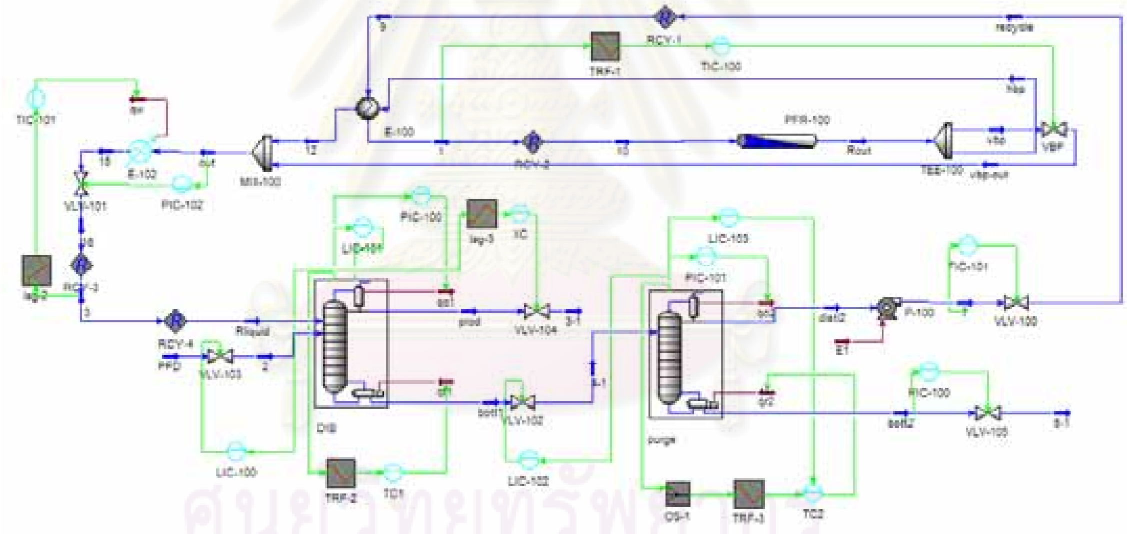


Figure 5.30 Application of reference control structure 3 (CS3) to the Butane Isomerization plant alternative 6 (RHEN-6)

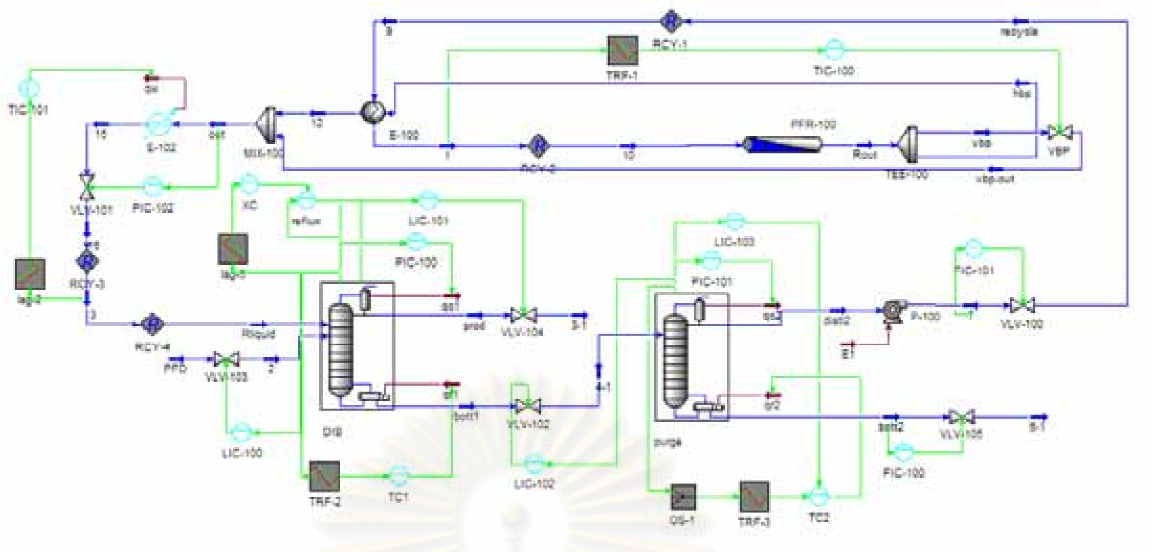


Figure 5.31 Application of reference control structure 4 (CS4) to the Butane Isomerization plant alternative 6 (RHEN-6)

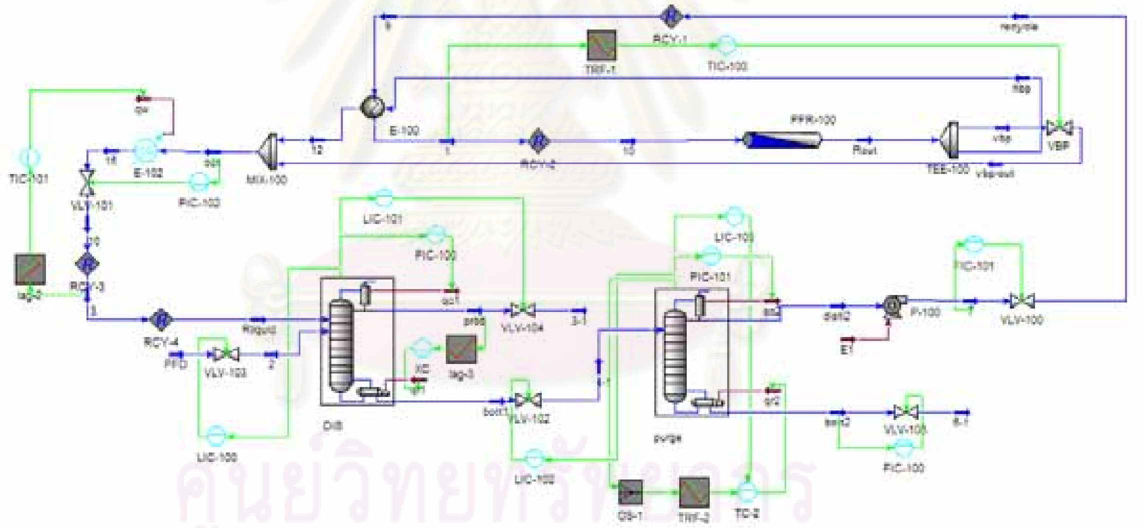


Figure 5.32 Application of reference control structure 5 (CS5) to the Butane Isomerization plant alternative 6 (RHEN-6)

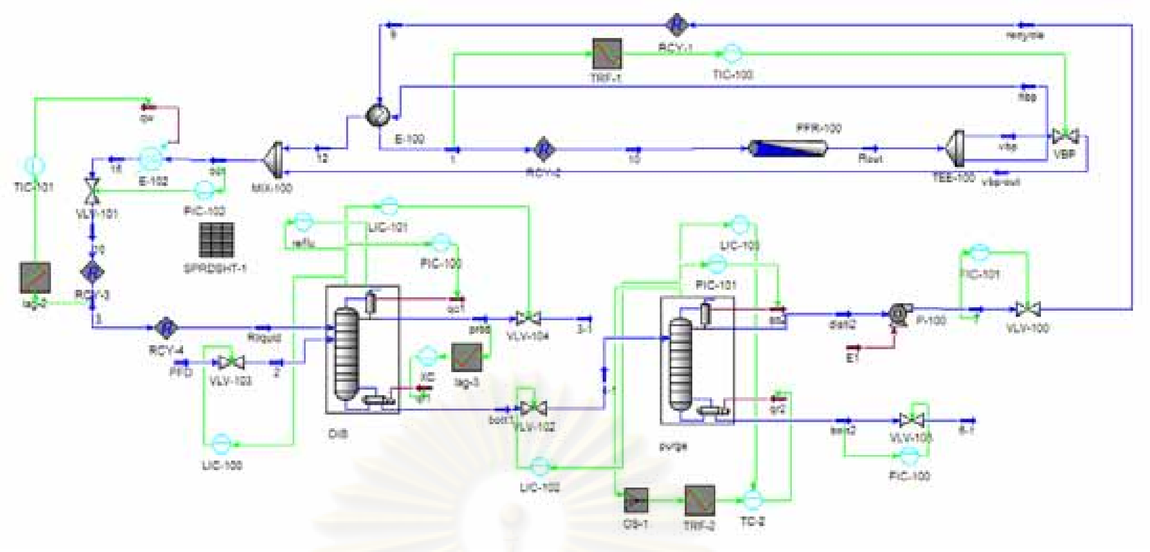


Figure 5.33 Application of reference control structure 6 (CS6) to the Butane Isomerization plant alternative 6 (RHEN-6)

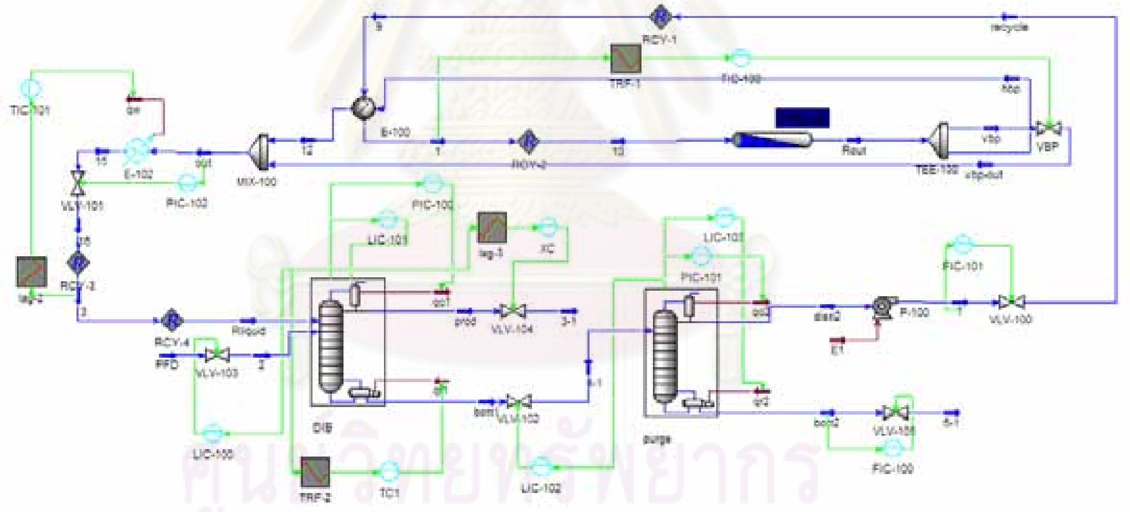


Figure 5.34 Application of reference control structure 7 (CS7) to the Butane Isomerization plant alternative 6 (RHEN-6)

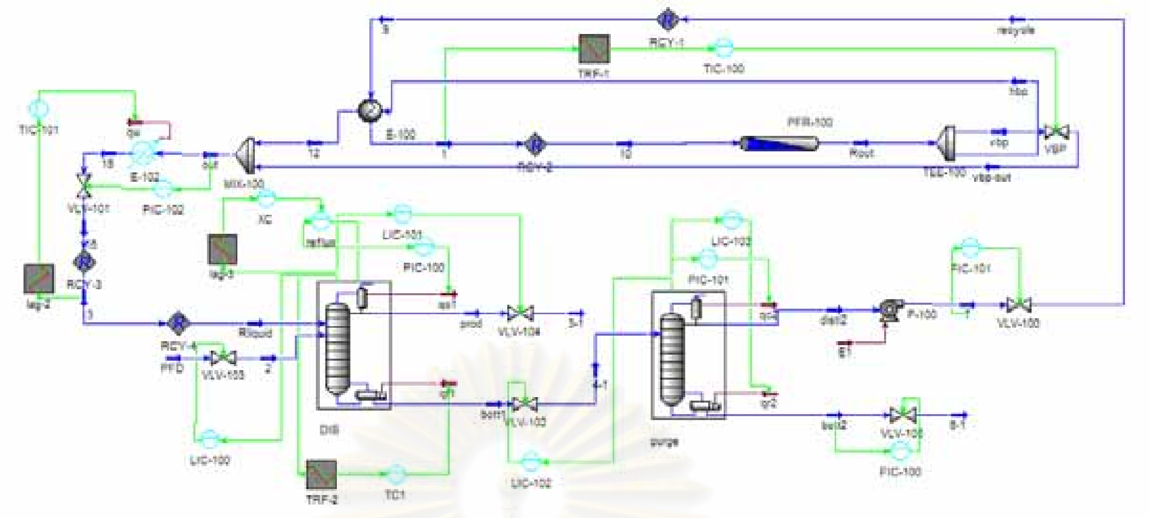


Figure 5.35 Application of reference control structure 8 (CS8) to the Butane Isomerization plant alternative 6 (RHEN-6)

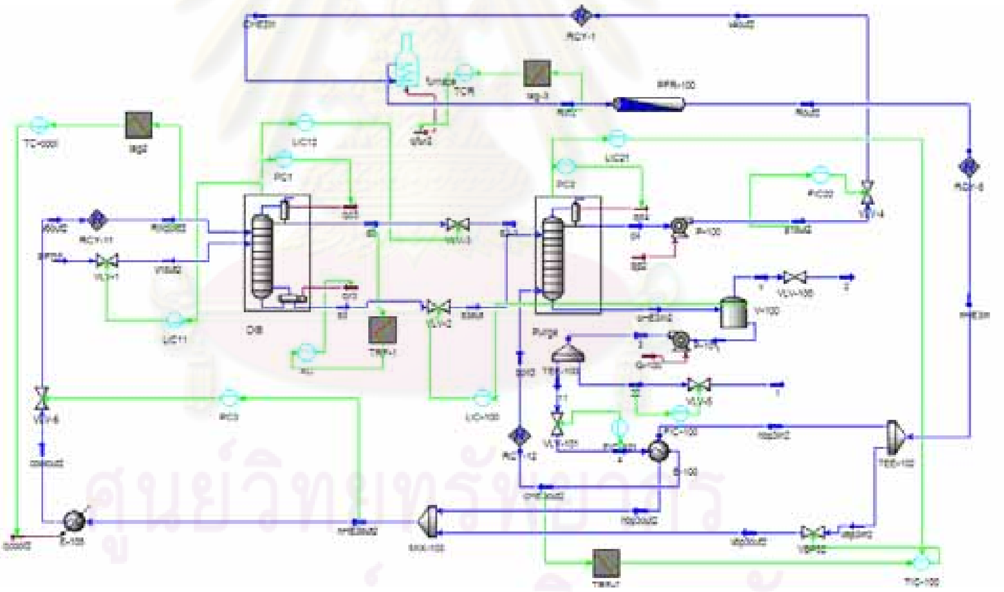


Figure 5.36 Application of reference control structure 1 (CS1) to the Butane Isomerization plant alternative 7 (RHEN-7)

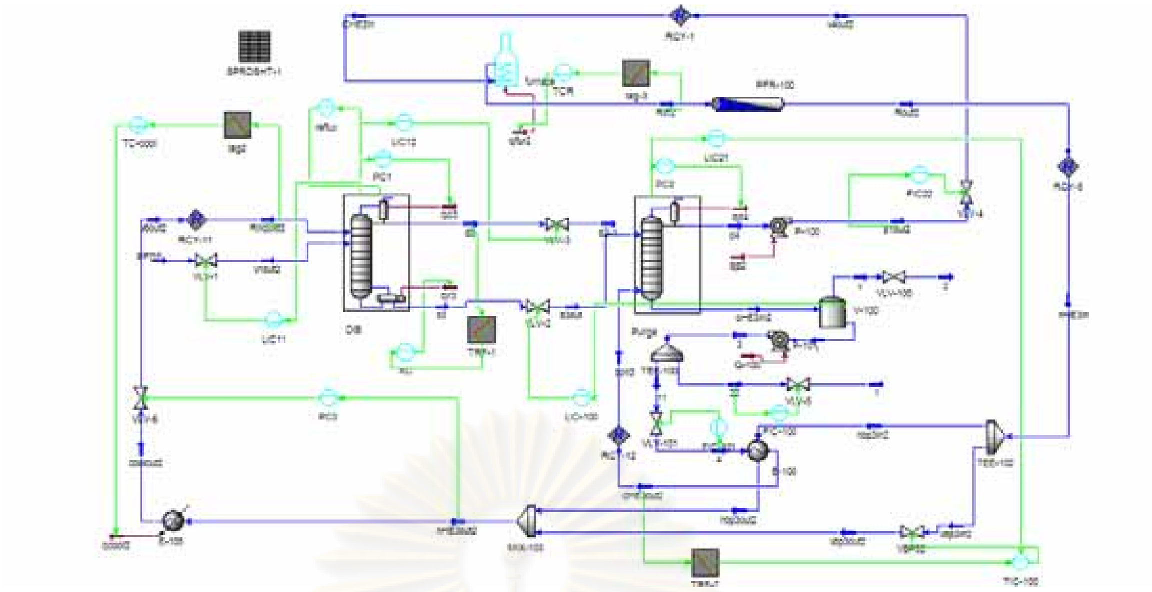


Figure 5.37 Application of reference control structure 2 (CS2) to the Butane Isomerization plant alternative 7 (RHEN-7)

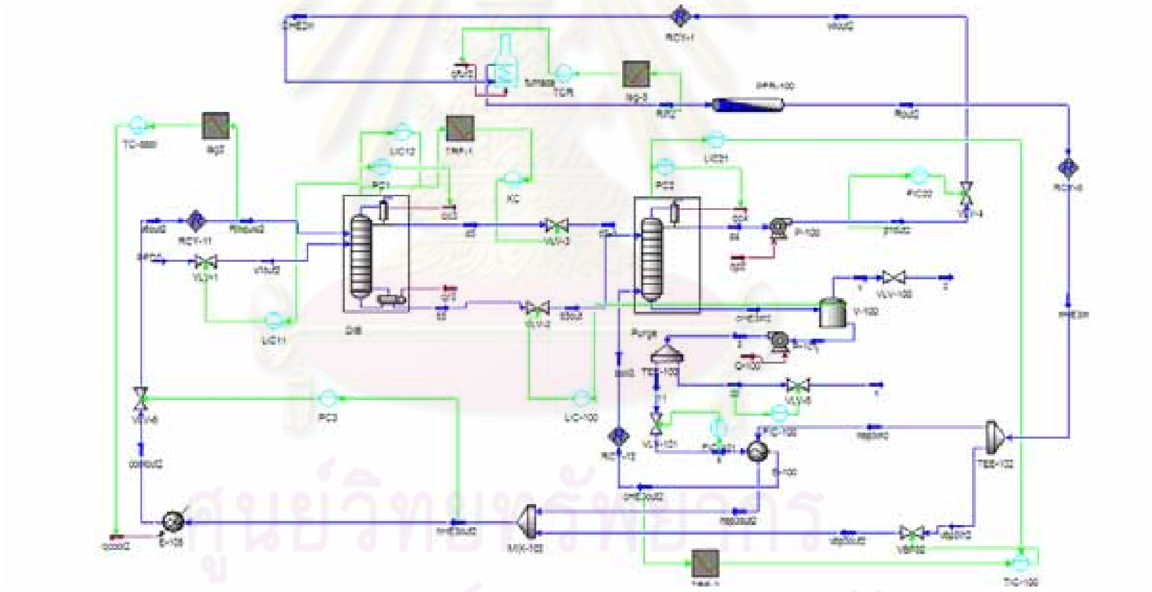


Figure 5.38 Application of reference control structure 3 (CS3) to the Butane Isomerization plant alternative 7 (RHEN-7)

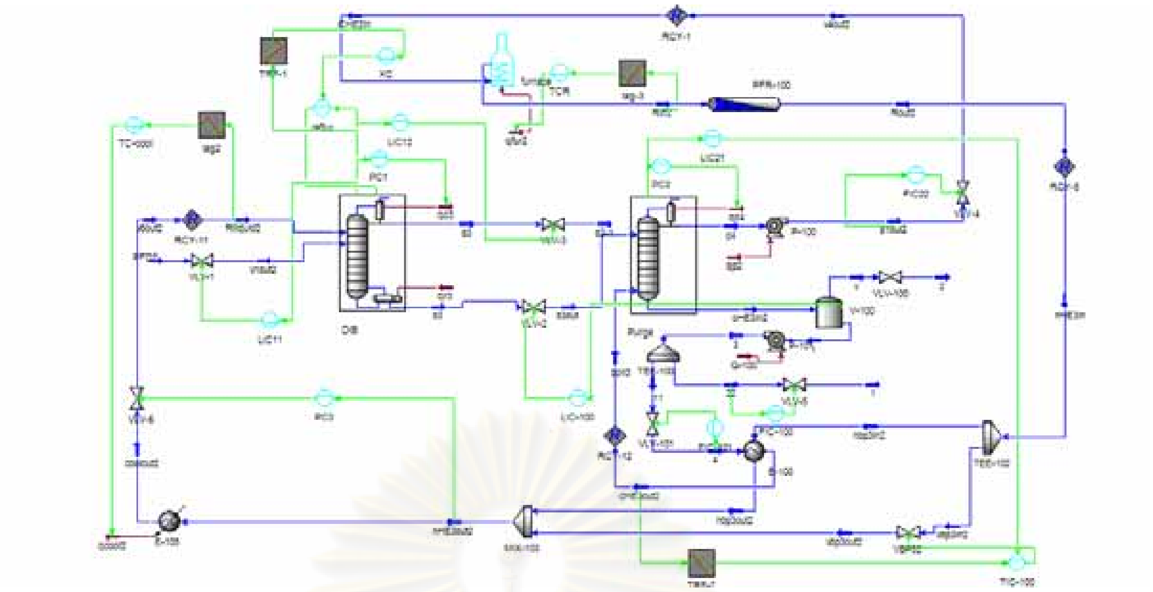


Figure 5.39 Application of reference control structure 4 (CS4) to the Butane Isomerization plant alternative 7 (RHEN-7)

5.4 Dynamic simulation results

In order to illustrate the dynamic behaviors of our control structures and the previous control structures (Luyben et al., 1999 and Kunajitpimol, 2006), two types of disturbance are used to test response of the system: inlet reactor temperature step increase 10 °F, and recycle flowrate increases from 870.1 lb mol/hr to 940.1 lb mol/hr. Temperature controllers are PIDs which are tuned using relay feedback. Two temperature measurement lags of 0.1 minute are included in the two temperature loops (reactor inlet temperature and DIB feed temperature). A 3-minute deadtime is assumed in the product composition measurement (distillate from the DIB). Flow and pressure controller are PIs and their parameters are heuristics values. Filters of 6 seconds are added in the flow loop to smooth out the measurement. Proportional-only level controllers are used and their parameters are heuristics values. Butane composition is measured and controlled using PID controller. All control valves are half-open at nominal operating condition.

Eight control structures CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8 are implemented on 8 integrated plants which are Base case, HIP1, HIP2, HIP3, HIP4, HIP5, HIP6 and HIP7. The dynamic responses of these plants are discussed next. If CS1 is paired with HIP1, we call this setout ‘HIP1- CS1’.

5.4.1 Change in the heat load disturbance of hot stream for CS1 to CS8 in Butane Isomerization plant for Base Case (Luyben et al., 1999)

Figure 5.72 show results when the reactor inlet temperature is changed by step increases 10^{o}F (from 390.1^{o}F to 400.1^{o}F), occurring at time equals 10 minutes.

The normal butane (nC4) for all of control structures decrease at first and rises to its set point afterwards as show in Figure. For CS1 and CS2, changes for impurity of nC4 in product are larger than CS3 and CS4 because there is the base temperature controller of DIB column for CS3 and CS4 to fix the composition. As for CS5 and CS6 are the same as CS1 and CS2 while CS7 and CS8 are the same as CS3 and CS4 respectively, because the control structures of DIB column are similar.

The product flowrates increase due to reaction rate increases as temperature increases and fresh feed flowrates are fed to the column increasingly. As a result, CS1, CS2, CS5 and CS6 increase more CS3, CS4, CS7 and CS8 as show in Figure 5.72a.

The IAEs impurity, *n*C4, composition loops at the top of DIB column is shown in Table 5.2. Control structures CS3, CS4, CS7 and CS8 handle the change better than CS1, CS2, CS5 and CS6 since their composition and temperature loops have smaller lags and delays.

Table 5.2 IAEs of *n*C4 composition loops at the top of DIB column

Integral Absolute Error							
CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
1.5993	1.5688	0.2129	0.2180	1.6031	1.5679	0.2114	0.2196

It is interesting to note that the heat inputs of the reboiler of DIB column of control structure CS3, CS4, CS7 and CS8 are lower than CS1, CS2, CS5 and CS6, see Table 5.3. Since CS3, CS4, CS7 and CS8 reject the thermal disturbance by keeping the base temperature constant, its resulted DIB reboiler duty is lower and resulted furnace duty is higher than that of CS1, CS2, CS5 and CS6. The product stream (v3out) is increased; in the case of CS3, CS4, CS7 and CS8, it is lower than in the case of CS1, CS2, CS5 and CS6, and so is the fresh feed (PFD). Figure 5.72d shows dynamic responses of the temperature on tray 1 of DIB column.

Control structure CS3, CS4, CS7 and CS8 control the impurity in the product by manipulating distillate (for CS3 and CS7) and reflux (for CS4 and CS8). The dynamic responses are faster than the others resulting in lower in IAEs. However, large change in the product flow is not important since it is assumed that the distillate is going to a storage tank or cavern.

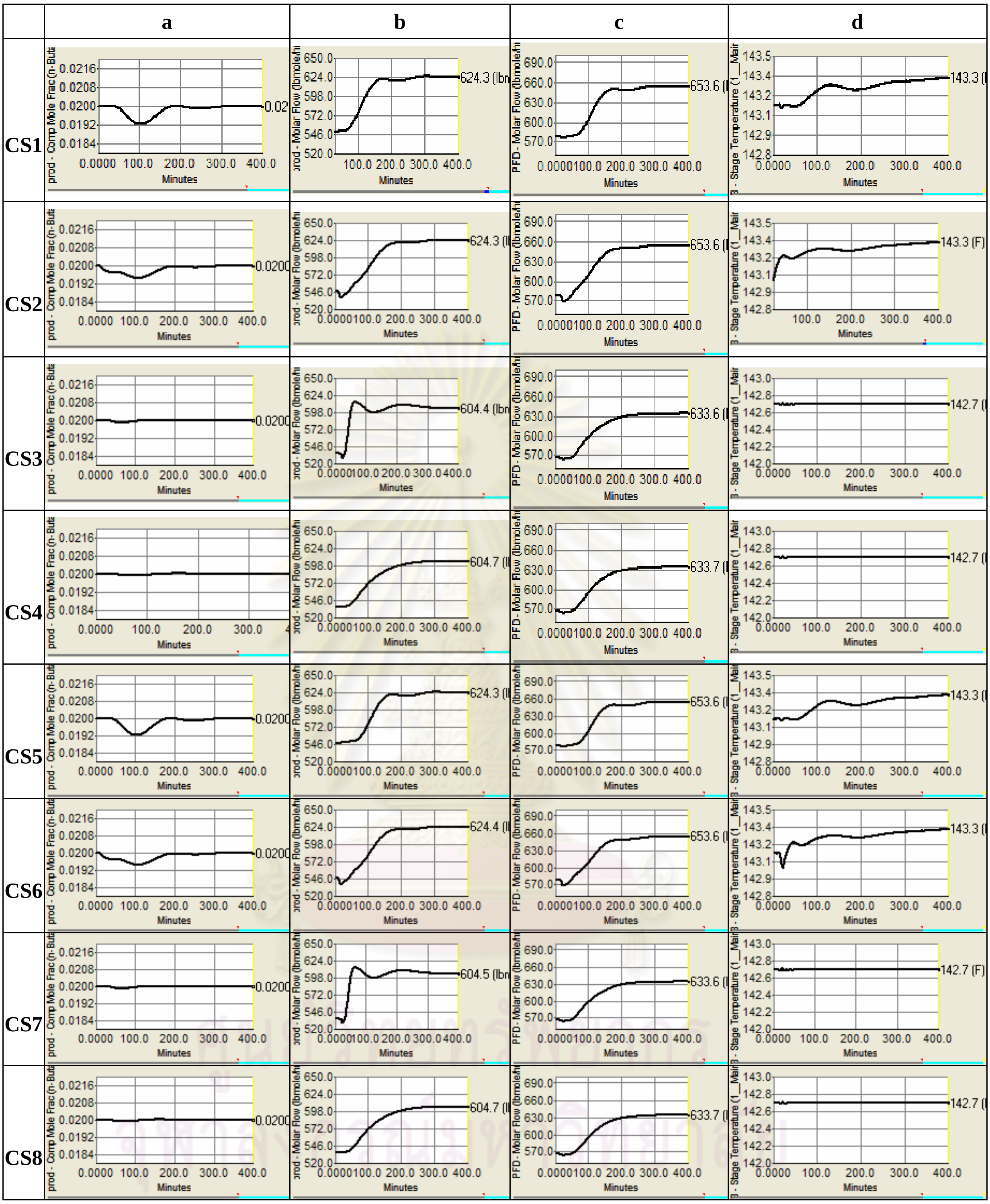


Figure 5.40 Dynamic responses to 10 ^{o}F increase in reactor inlet temperature of Base case, where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column

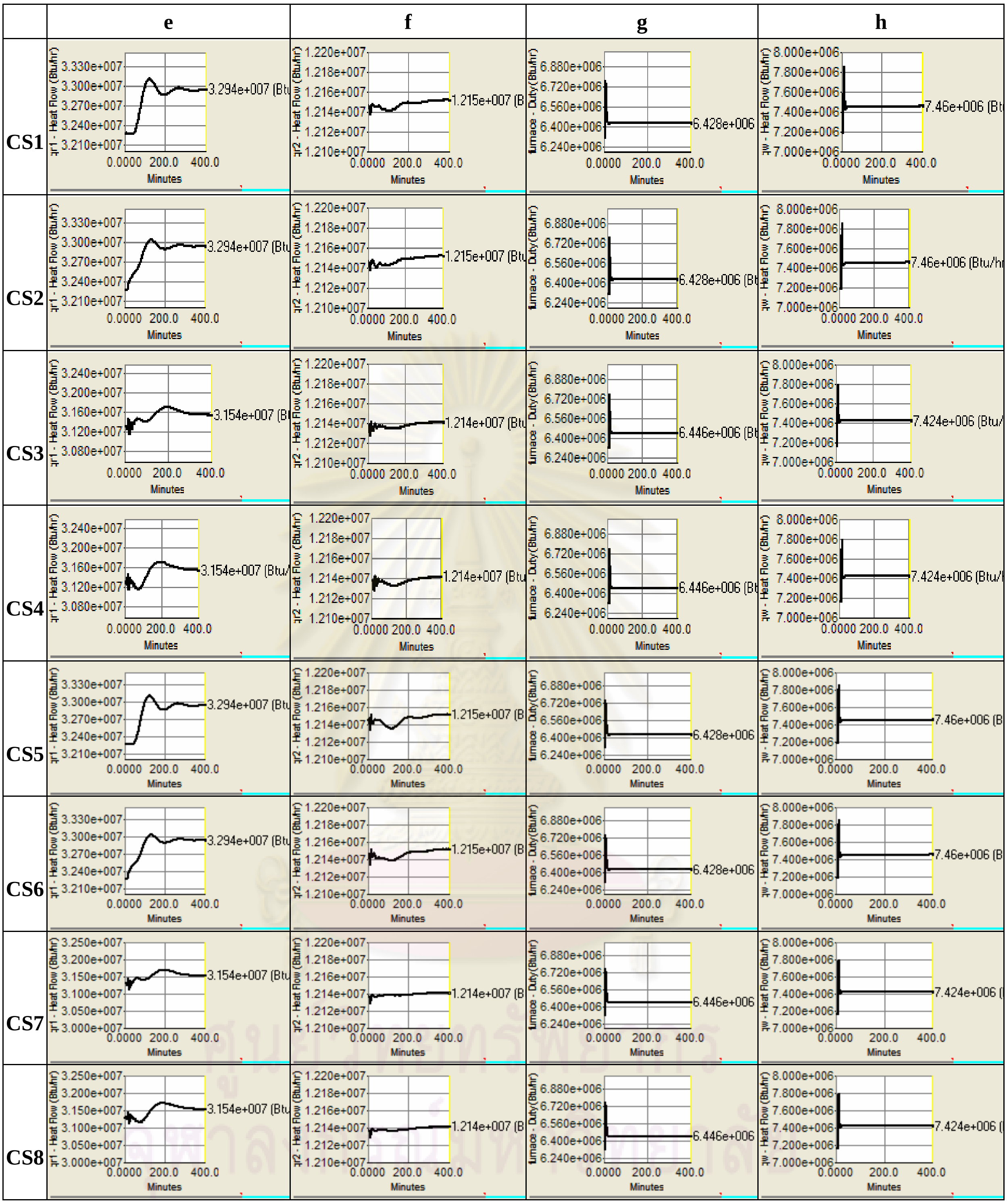


Figure 5.40 Continued Dynamic responses to 10 °F increase in reactor inlet temperature of Base Case, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

Table 5.3 Energy increases as the temperature of reactor feed increase 10 °F for Base Case

Energy Increases (10^5,Btu/ hr)						
	DIB column reboiler	Purge column reboiler	Total reboiler	Furnace	Total hot utilities	Total cold utilities
CS1	67.2126	0.6094	67.8220	12.0636	79.8856	27.3552
CS2	67.0629	0.6125	67.6754	12.0580	79.7334	27.3603
CS3	23.4715	0.3331	23.8046	12.6350	36.4396	26.1154
CS4	22.9736	0.3567	23.3303	12.6363	35.9667	26.1144
CS5	67.3701	0.6156	67.9857	12.0641	80.0497	27.3473
CS6	67.0320	0.6433	67.6754	12.0550	79.7304	27.3637
CS7	23.4039	0.3559	23.7597	12.6373	36.3971	26.1124
CS8	23.0927	0.3477	23.4403	12.6366	36.0769	26.1120

5.4.2 Change in the heat load disturbance of hot stream for CS1 to CS8 in Butane Isomerization plant for HIP1, HIP2, HIP3 and HIP4 (Kunajitpimol, 2006)

Figure 5.73-5.76 show dynamic responses of the Butane Isomerization plant HIP1 to HIP4 when the reactor inlet temperature is changed by step increases 10°F (from 390.1°F to 400.1°F), occurring at time equals 10 minutes.

The heat integration structure HIP1, normal butane (nC4) of all control structures decrease at first and rises to its set point after 350 minutes for CS1 and CS5, 300 minutes for CS2 and CS6, 100 minutes for CS3 and CS7 and 50 minutes for CS4 and CS8 as show in Figure 5.73a. For CS1 and CS2, changes for impurity of nC4 in product are larger than CS3 and CS4 because there is the base temperature controller of DIB column for CS3 and CS4 to fix the bottom composition. As for CS5 and CS6 are the same as CS1 and CS2 while CS7 and CS8 are the same as CS3 and CS4 respectively, because the control structures of DIB column are similar. The product flowrates increase due to reaction rate increases as temperature increases and fresh feed flowrates are fed to the column increasingly. As a result, CS1, CS2, CS5 and CS6 increase more CS3, CS4, CS7 and CS8 as show in Figure 5.73b-c.

Similar conclusion can be drawn for heat integrated structure HIP2, HIP3 and HIP4 with control structures CS1-CS8. (see Figure 5.74-5.76).

The thermal disturbance for HIP1 and HIP3 are shifted to the furnace, their changes in furnace duties are lower than HIP2 and HIP4 because increased temperature from thermal disturbance assist furnace duties consumption decreasing. HIP2 and HIP4 shift the thermal disturbance to the DIB and Purge auxiliary reboiler, so the least energy increases of DIB reboiler and Purge reboiler are HIP2 and HIP4 respectively. (see Table 5.5)

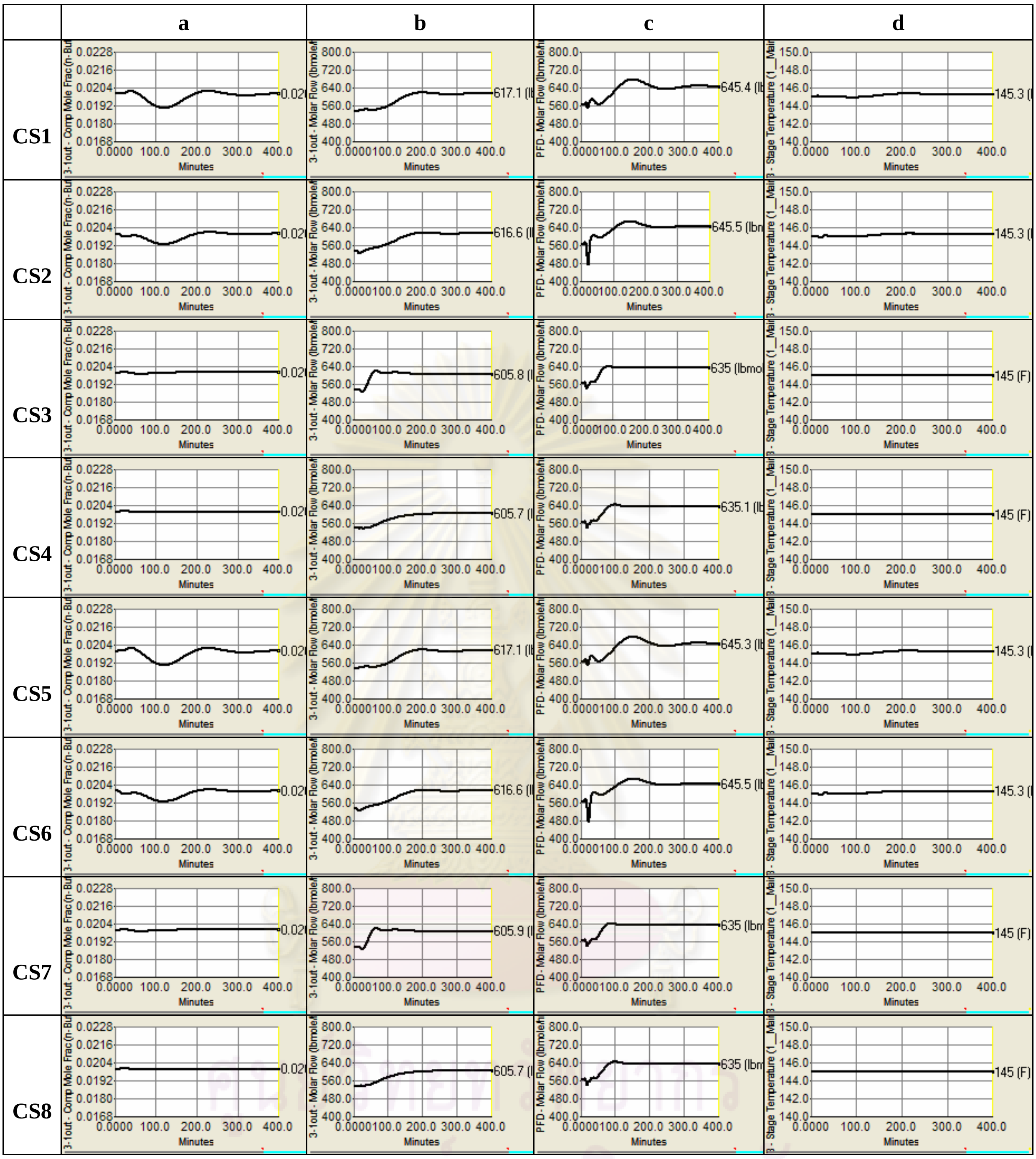


Figure 5.41 Dynamic responses of the Butane Isomerization plant HIP1 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

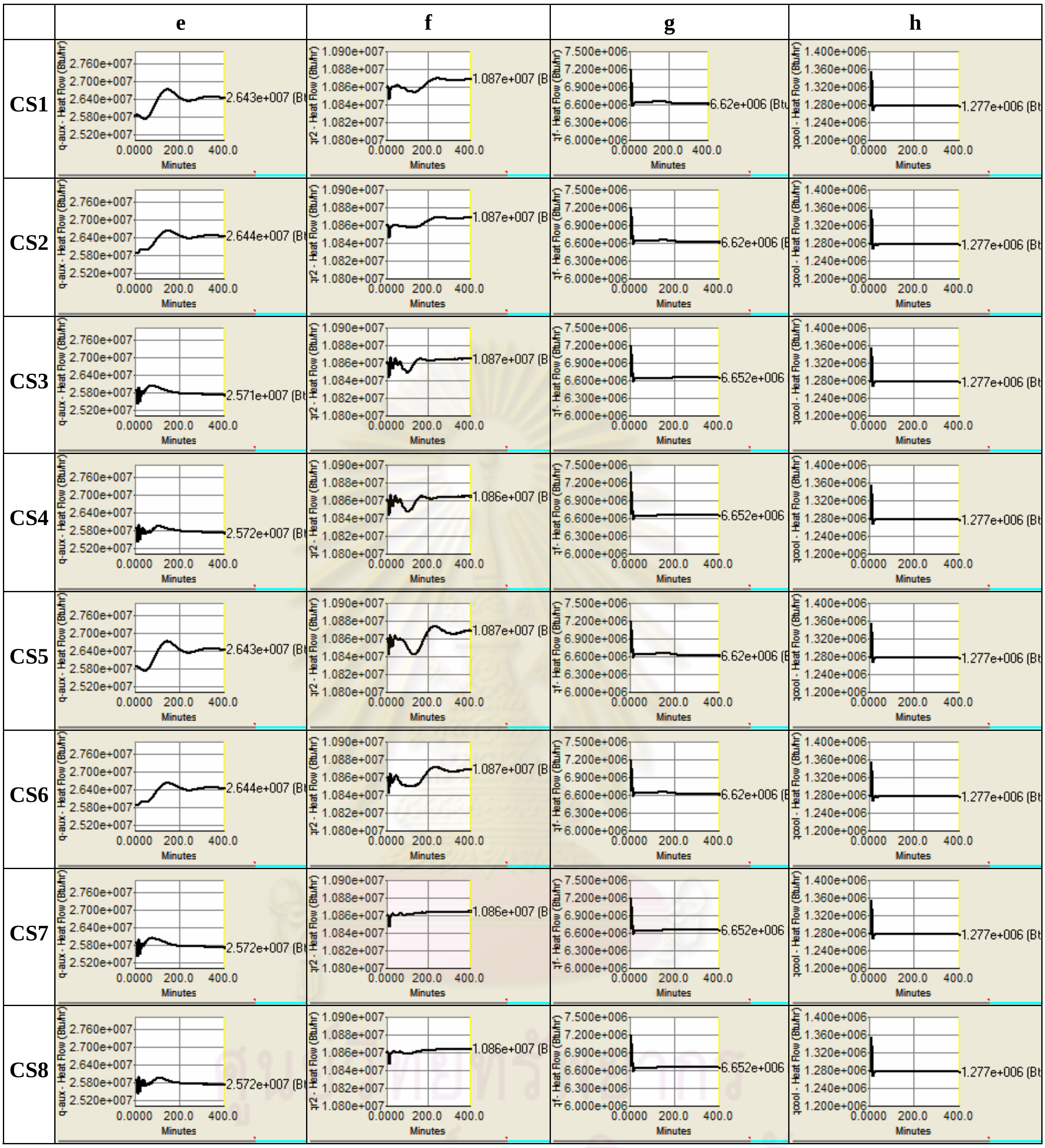


Figure 5.41 Continued Dynamic responses of the Butane Isomerization plant HIP1 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

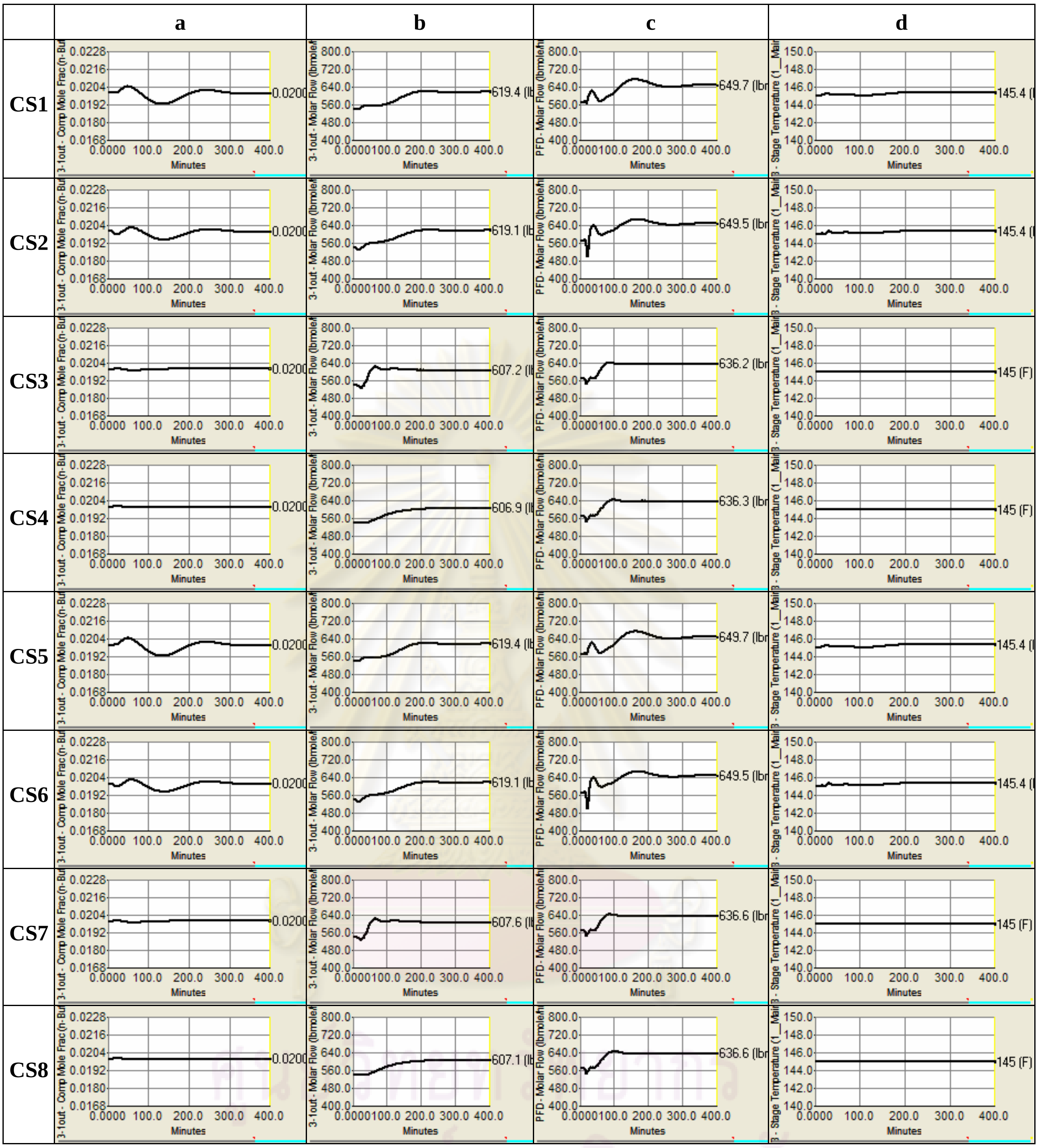


Figure 5.42 Dynamic responses of the Butane Isomerization plant HIP2 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

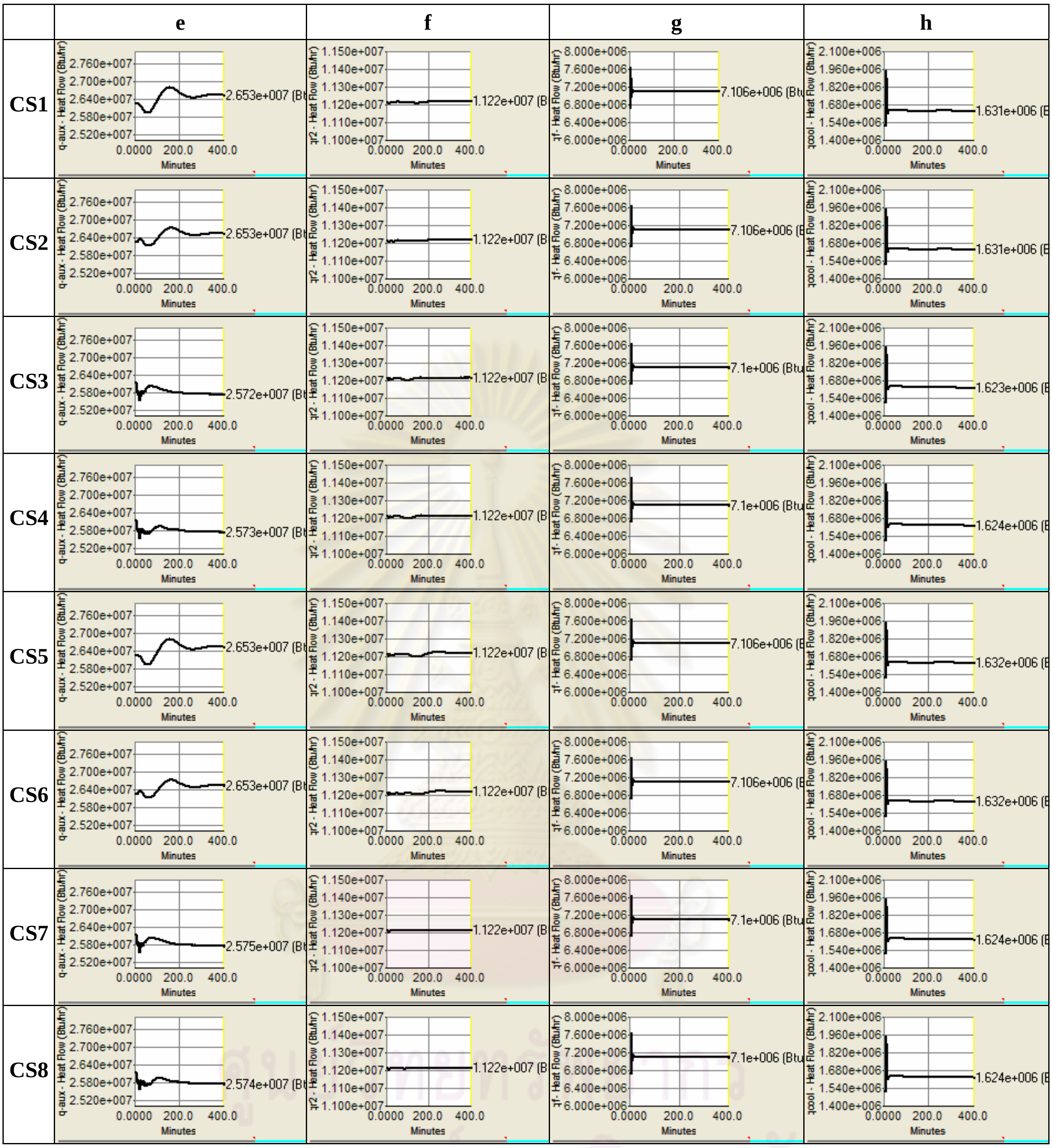


Figure 5.42 Continued Dynamic responses of the Butane Isomerization plant HIP2 to10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

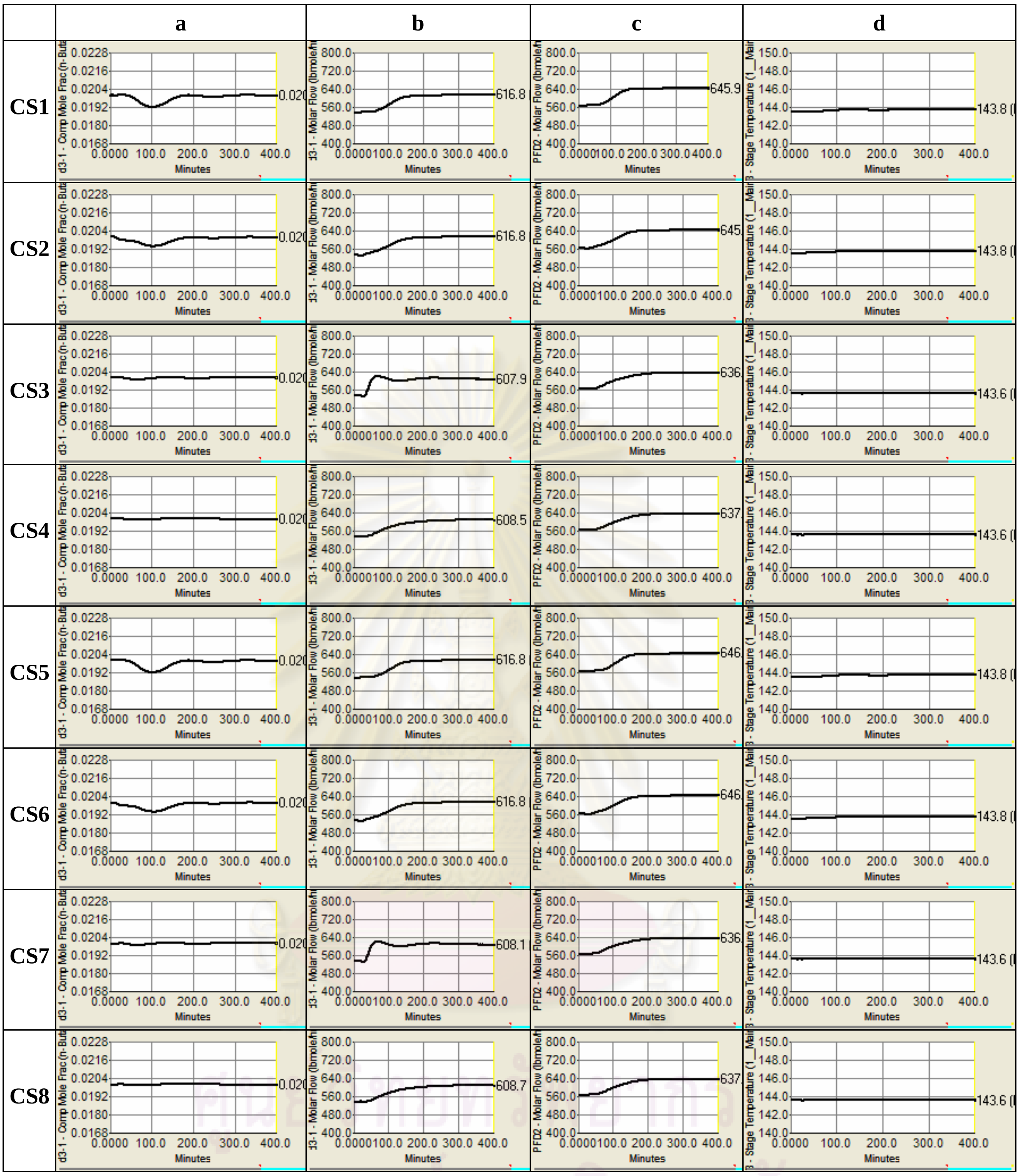


Figure 5.43 Dynamic responses of the Butane Isomerization plant HIP3 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

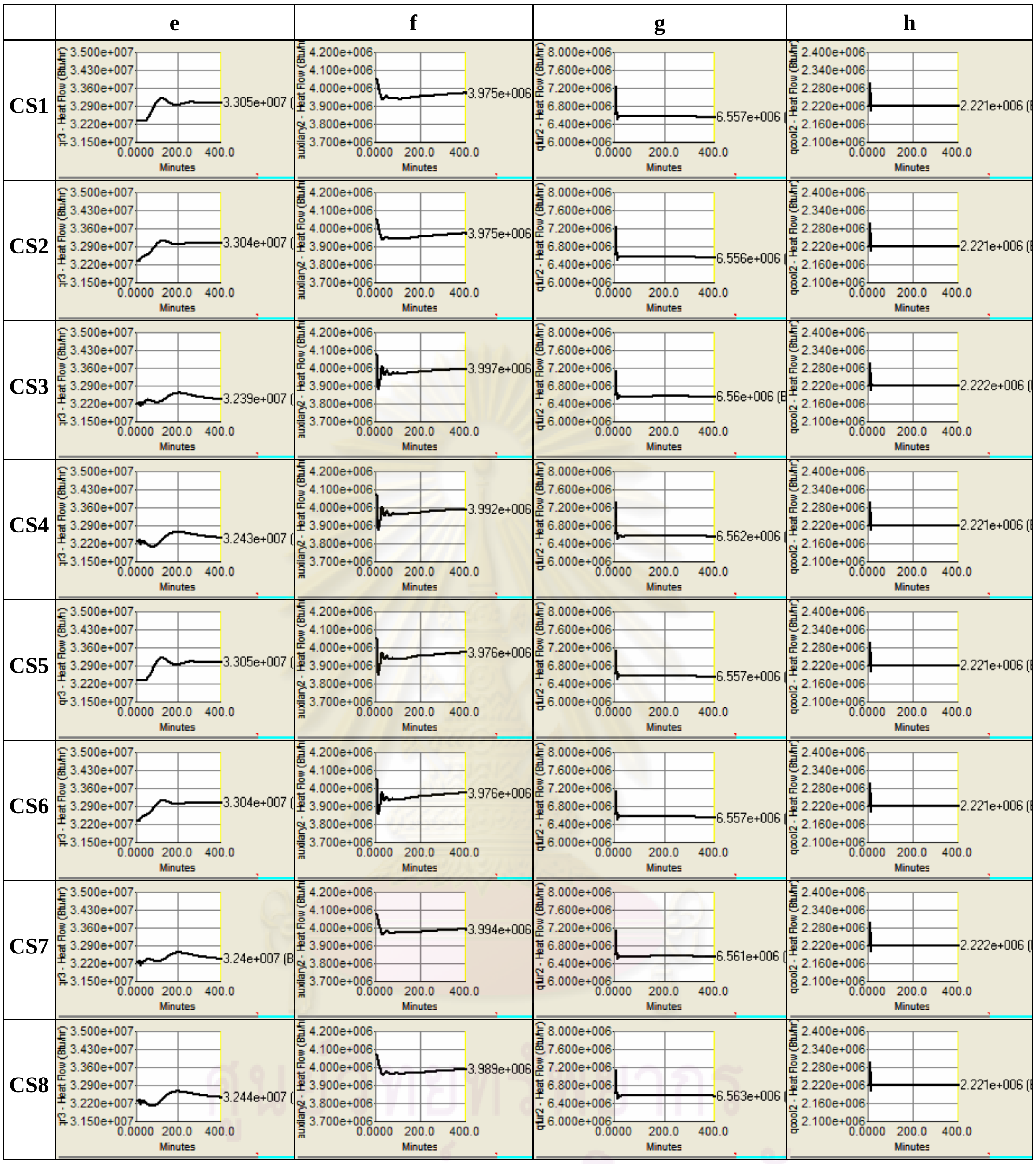


Figure 5.43 Continued Dynamic responses of the Butane Isomerization plant HIP3 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

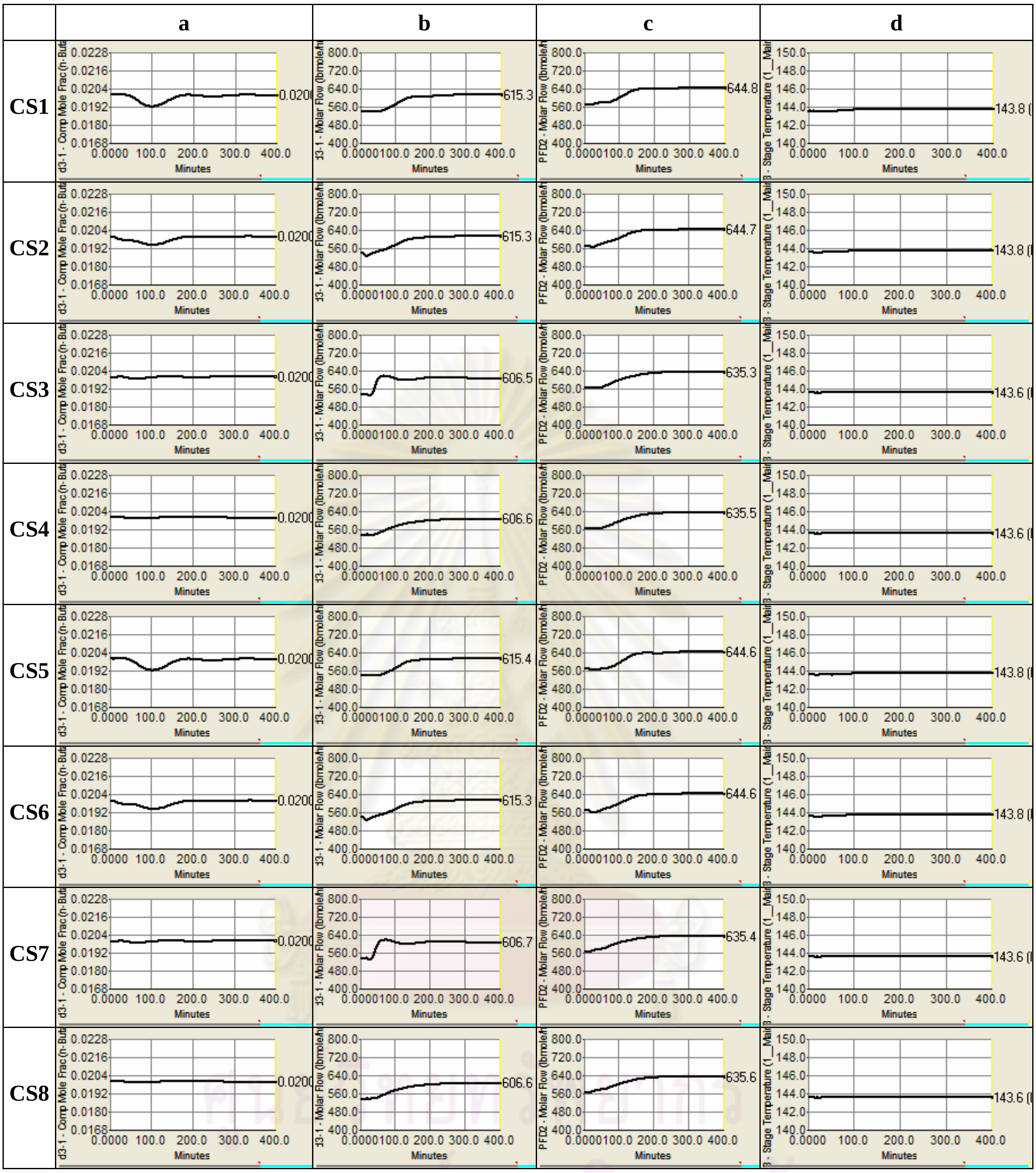


Figure 5.44 Dynamic responses of the Butane Isomerization plant HIP4 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

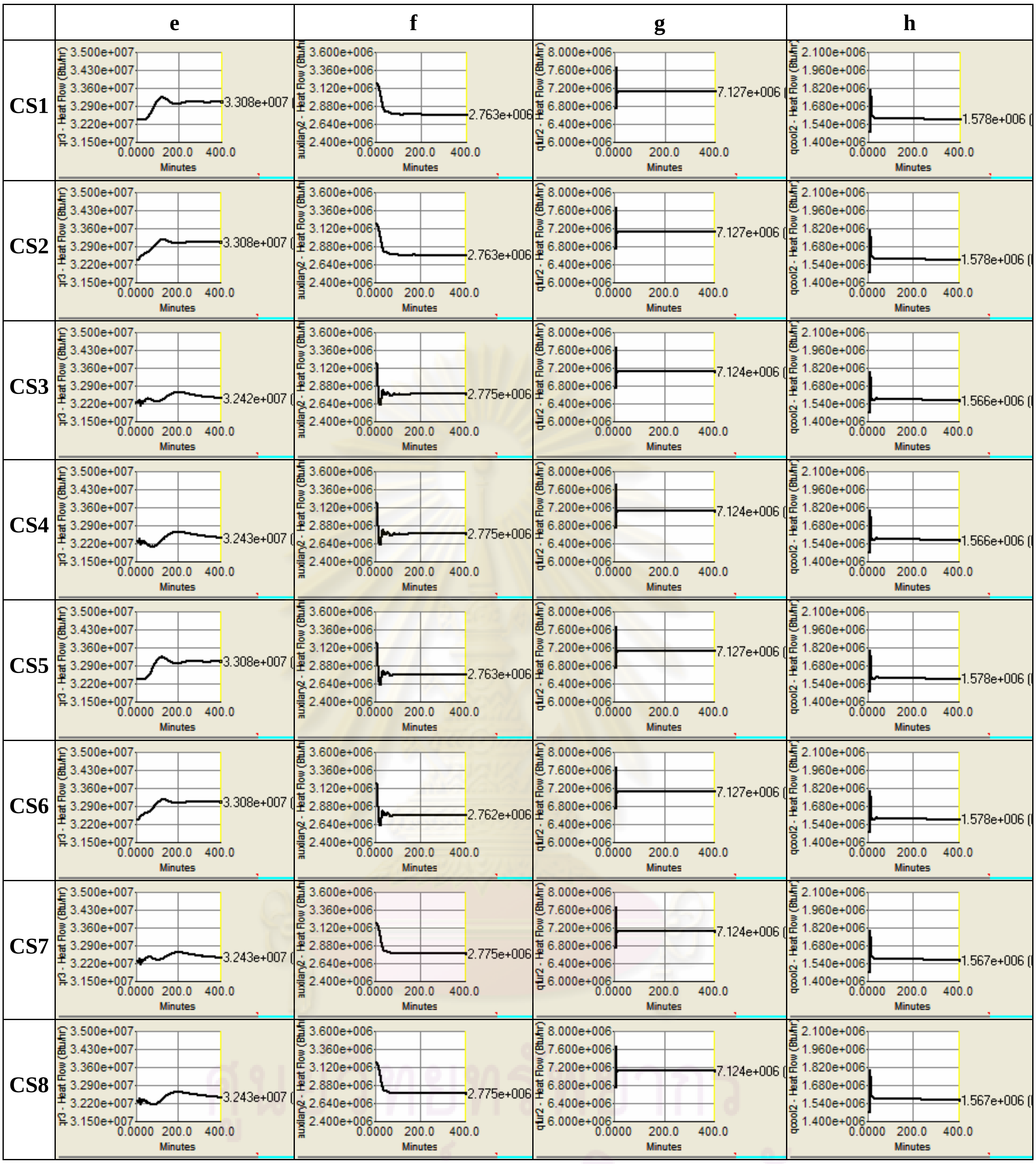


Figure 5.44 Continued Dynamic responses of the Butane Isomerization plant HIP4 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

5.4.3 Change in the heat load disturbance of hot stream for CS1 to CS8 in the new designed heat integration plants: HIP5, HIP6 and HIP7

Figure 5.77-5.79 show dynamic responses of the Butane Isomerization plant HIP5 to HIP7 when the reactor inlet temperature is changed by step increases 10^{o}F (from 390.1^{o}F to 400.1^{o}F), occurring at time equals 10 minutes.

The heat integration structure HIP5, normal butane (nC4) of all control structures decrease at first and rises to its set point over 300 minutes for CS1, CS2, CS5 and CS6 while CS3, CS4, CS7 and CS8 is going to its set point over 100 minutes as show in Figure 5.77a. For CS1 and CS2, changes for impurity of nC4 in product are larger than CS3 and CS4 because there is the base temperature controller of DIB column for CS3 and CS4 to fix the bottom composition. As for CS5 and CS6 are the same as CS1 and CS2 while CS7 and CS8 are the same as CS3 and CS4 respectively, because the control structures of DIB column are similar. The product flowrates increase due to reaction rate increases as temperature increases and fresh feed flowrates are fed to the column increasingly. As a result, CS1, CS2, CS5 and CS6 increase more CS3, CS4, CS7 and CS8 as show in Figure 5.77b-d.

Similar conclusion can be drawn for heat integrated structure HIP6 and HIP7 with control structures CS1-CS8. (see Figure 5.78-5.79).

In this work, we design 3 new heat integrated plants (HIP5, HIP6 and HIP7) in order to be more choice by improved from earlier work (Kunajitpimol, 2006), there is 1 FEHE for them and large area FEHE to completely exchange. This is to achieve energy saving by reduce amount using FEHE and can operation in good order which the control structure is evaluated with each heat integrated plant. The thermal disturbance for HIP5 is shifted to DIB column reboiler so the energy consumption decrease from normal case, the result is shown negative in Table 5.5. For HIP6 and HIP7 shift the thermal disturbance to cooler, the heat from hot stream (reactor product stream) is enough to supply to cold stream (feed reactor stream for HIP6 and purge column reboil stream for HIP7) therefore the furnace and the Purge reboiler is not implemented for HIP6 and HIP7 respectively.

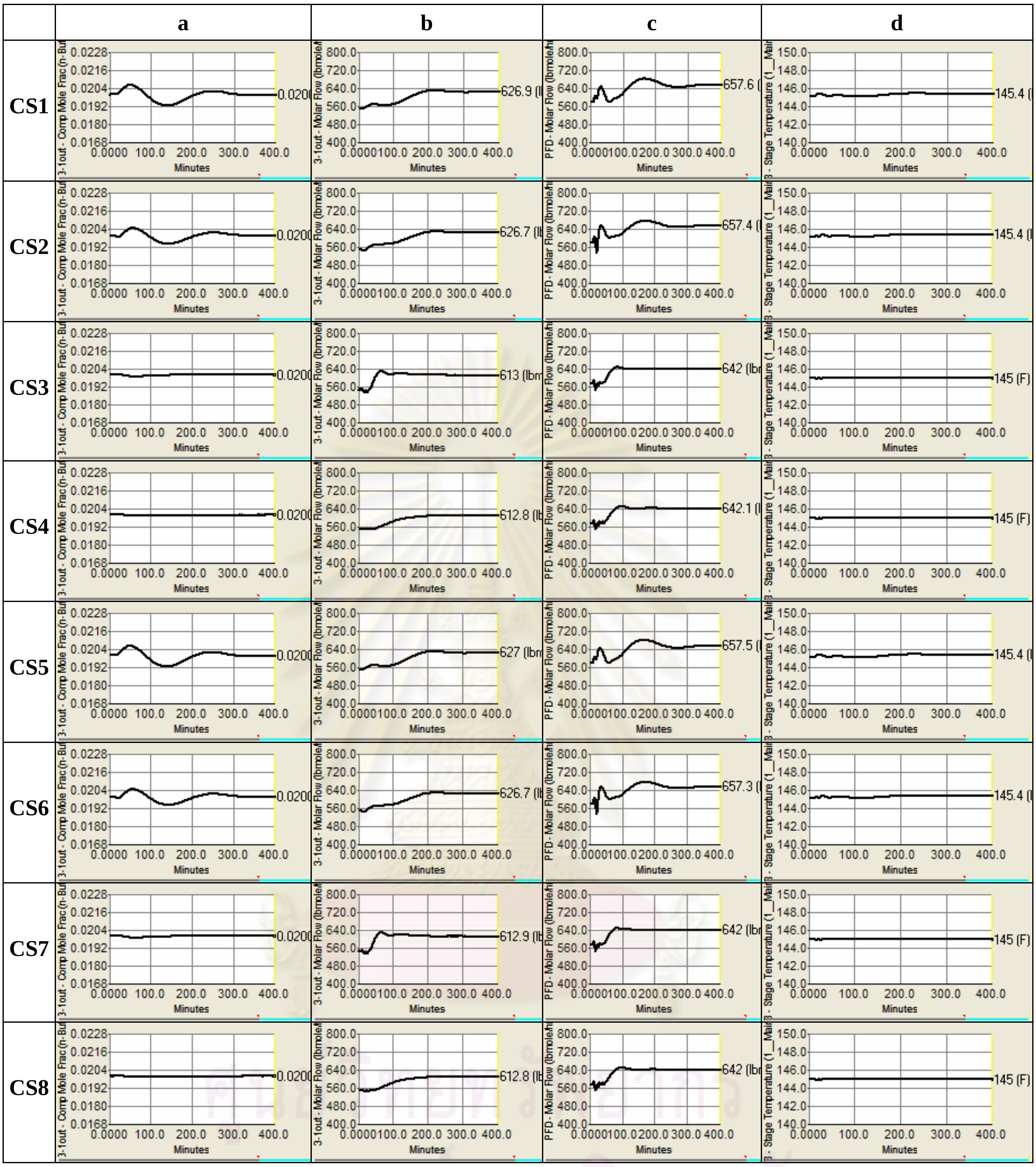


Figure 5.45 Dynamic responses of the Butane Isomerization plant HIP5 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

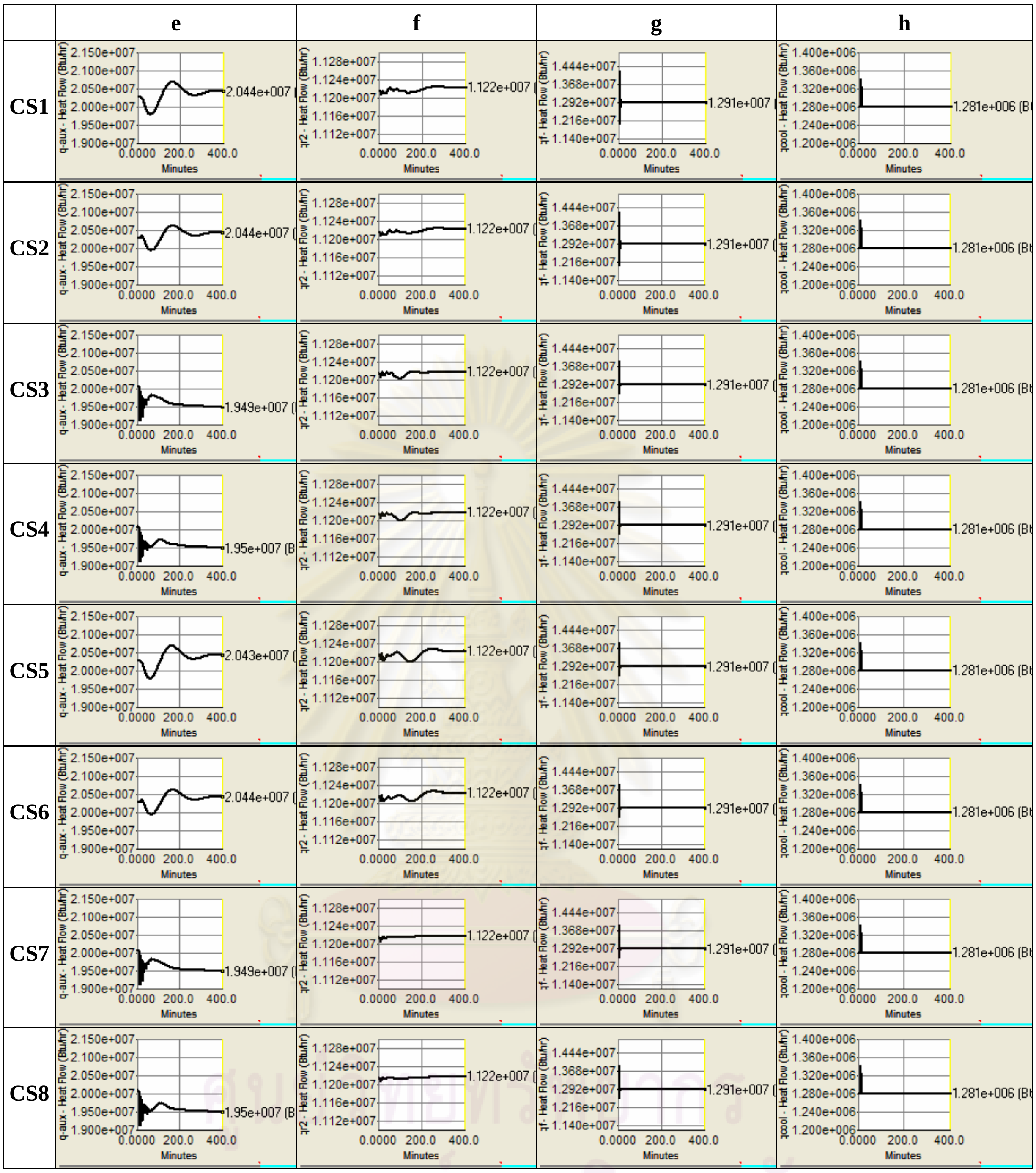


Figure 5.45 Continued Dynamic responses of the Butane Isomerization plant HIP5 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

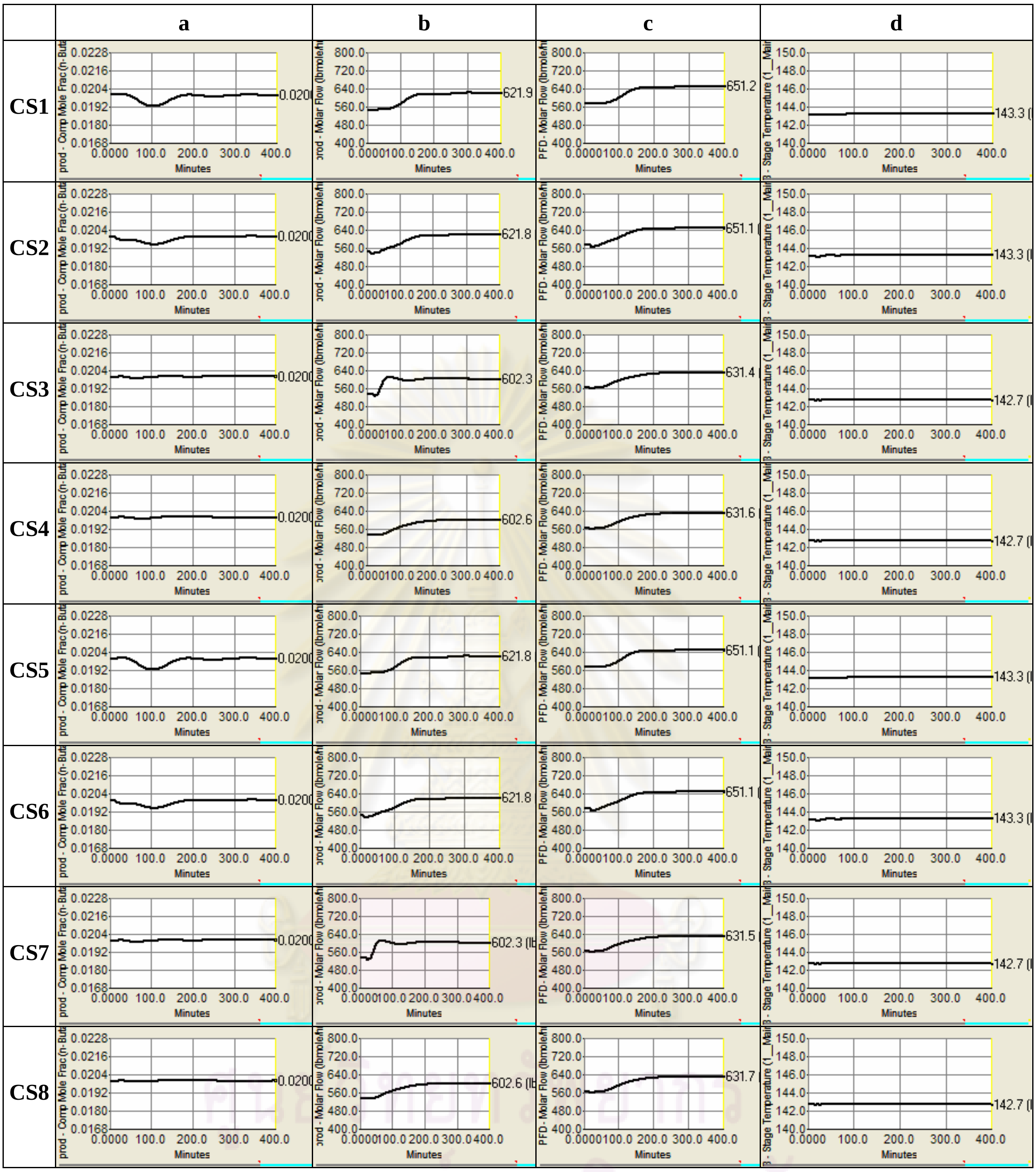


Figure 5.46 Dynamic responses of the Butane Isomerization plant HIP6 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

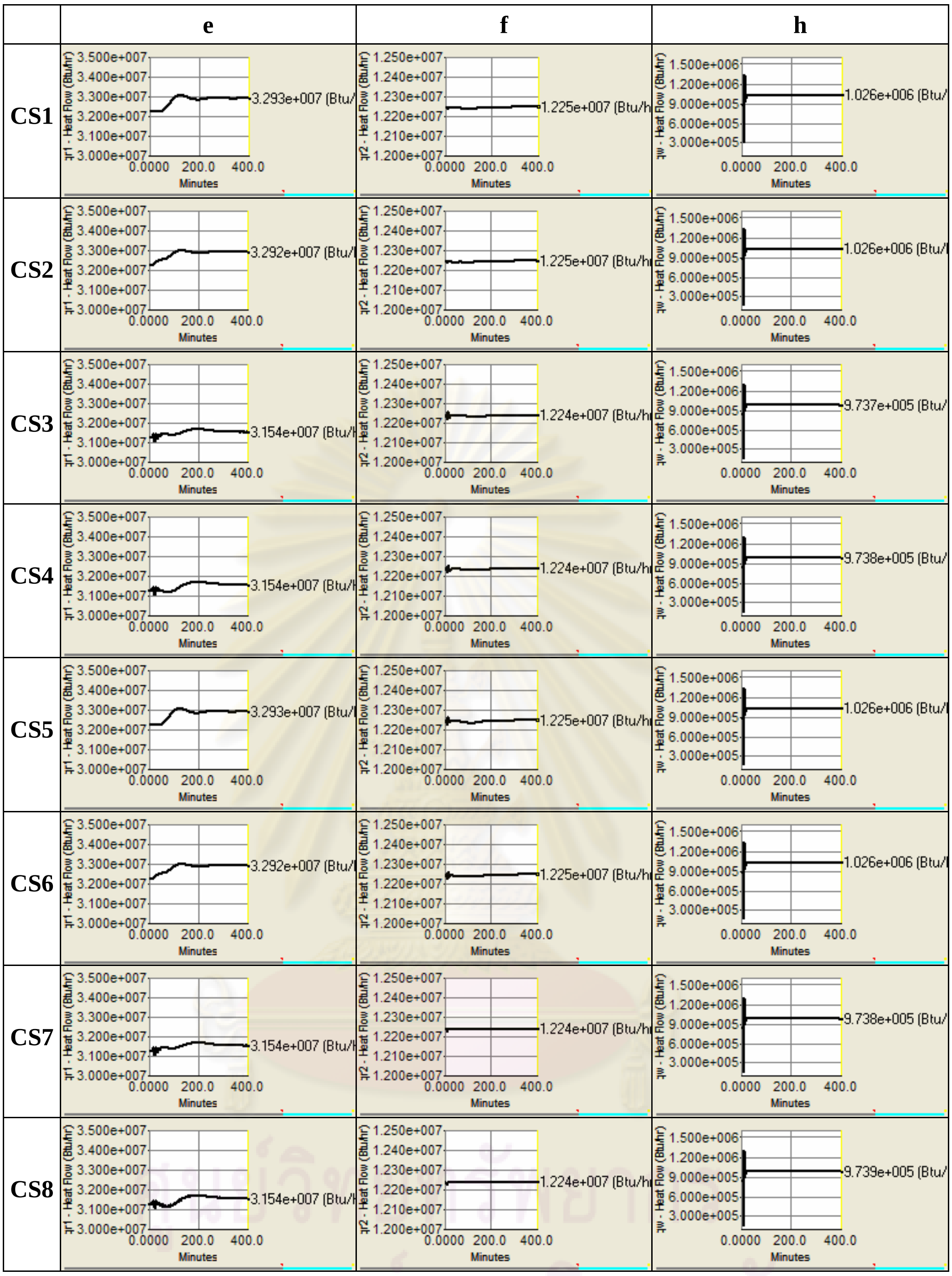


Figure 5.46 Continued Dynamic responses of the Butane Isomerization plant HIP6 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column and (h) cooler duty.

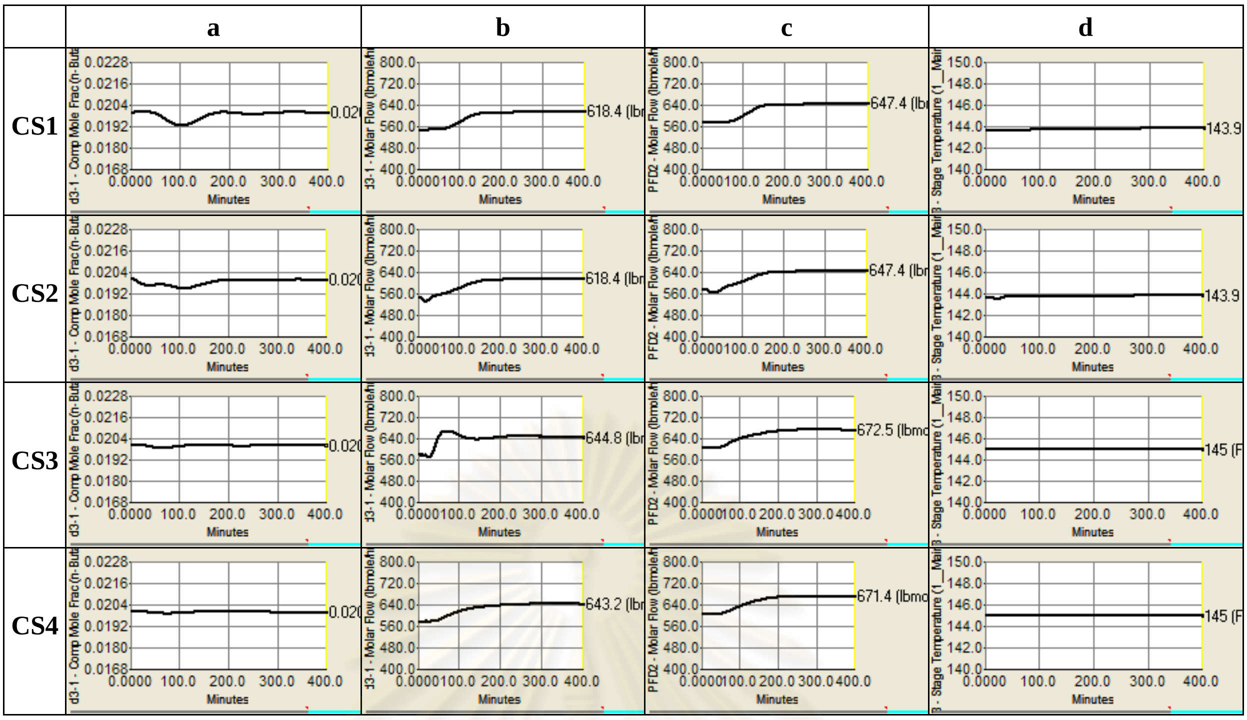


Figure 5.47 Dynamic responses of the Butane Isomerization plant HIP7 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column

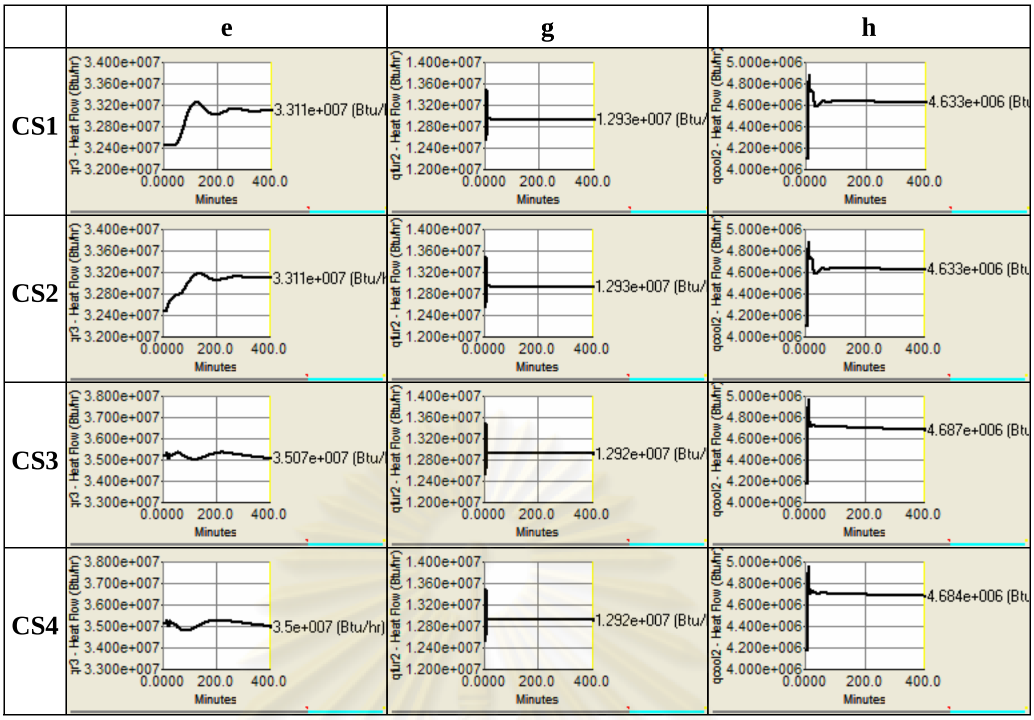


Figure 5.47 Continued Dynamic responses of the Butane Isomerization plant HIP7 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream), where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

5.4.4 Comparison of control structures for change the heat load disturbance of hot stream: CS1, CS2, CS4, CS4, CS5, CS6, CS7 and CS8 with HIP1-7

From the dynamic simulation results for all heat integration processes with control structures CS1-CS8 when the reactor inlet temperature is changed, we can conclude that CS3, CS4, CS7 and CS8 are the proper control structures. Because their feature structures are the base DIB column temperature is controlled so they reject the thermal disturbance by keeping the base temperature constant. The product flowrate increases due to reaction rate increases as temperature increasingly. The nC4 impurity in product are slowly decreased and return to its set point which CS3, CS4, CS7, and CS8 are less variable than CS1, CS2, CS5 and CS6. Also the control structures CS3, CS4, CS7 and CS8 give the better economic response for all heat integrated structure, i.e. they require less total reboiler utility consumptions. Control structure CS3, CS4, CS7 and CS8 control the base temperature of DIB column. The DIB reboiler duty compensates the thermal disturbance entering the column base. Therefore it is not propagated to the downstream unit as show in Table 5.5.

Control structures CS3 control the process is as good as CS4. The thermal disturbed stream, resulting in decreasing impurity, enters DIB column and causes direct increasing in reflux flow and then decrease in distillate flow (CS4) is just slightly better than indirect adjusting reflux flow resulting from the reflux drum level increase (CS3) which is lags behind by the drum size and lose control action. Because of DIB column was controlled with control structure CS7 and CS8 to be alike CS3 and CS4 respectively, the result is shown similarly.

Note that, control structures CS7 (temperature at purge column is not controlled) is better control than CS3 (controls purge column temperature) for structures with heat integration at DIB column, namely HIP1, HIP2 and HIP5 while CS7 is worse control for structures with heat integration at Purge column. Because the thermal disturbed stream, resulting in increasing disturbance so the temperature column need to control when there is the heat integration at its column. We can to neglect the purge column temperature control for structure with heat integration at DIB column.

5.4.5 Comparison of heat integrated structures for change the heat load disturbance of hot stream: HIP1, HIP2, HIP3, HIP4, HIP5, HIP6 and HIP7 with CS1-CS8

Figures 5.80-5.87 compare the performance of heat integration structure HIP1-7 with all control structures CS1-8. HIP1 and HIP3 with CS1 direct the thermal disturbance -62,201.14 and -74,924.69 Btu/hr to the furnace respectively. The DIB reboiler duty changes are 589,901.30 and 703,917.05 Btu/hr. HIP2 and HIP4 shift the thermal disturbance to the DIB and Purge auxiliary reboiler respectively and their change in duties are 289,196.12 and -416,340.00 Btu/hr. The DIB column requires more heat than the amount shifted to put out more boilup, so the positive figure. HIP5 with CS1shift the thermal disturbance to the DIB auxiliary reboiler as good as HIP2 and its change in duties is 157,353.08 Btu/hr., this duty is less change than HIP2 due to the large area heat exchanger of HIP5. HIP6 with CS1 shift the thermal disturbance to the cooler. Because the furnace is not used so the thermal disturbance is propagated to cooler. The change in duties is 151,287.03 Btu/hr. The Purge column reboiler is not used for HIP7 therefore the thermal disturbance is shifted to the cooler as good as HIP6. The change in duties of HIP7 with CS1 is 526,819.73 Btu/hr.

Similar trends are observed for control structures CS2-8 but with different amount of energy used. Table 5.5 and Figures 5.90-5.92 show the energy increases when the temperature of reactor feed increases 10 ^{o}F.

Integrated structure HIP5 (FEHE is placed before DIB column reboiler) with control structure CS7 (top composition is adjusted by manipulating distillate flow) has the most favourable thermal effect of -208,415.89 Btu/hr. HIP5-CS3, HIP5-CS8, HIP5-CS4, HIP1-CS3, HIP1-CS4 are the next best pairs (see Figure 5.88). In summary, integrated structure has more pronounced effect on the energy utilization. Control structure, while less dominating, is also play an important decision making role, as seen here.

Table 5.4 and Figures 5.88-5.89 show the control performances for Base Case and HIP1-7 with CS1-CS8. The top ten performers are HIP5-CS7, HIP5-CS3, HIP5-CS8, HIP2-CS7, HIP5-CS4, HIP2-CS3, HIP2-CS8, HIP1-CS7, HIP2-CS4 and HIP1-CS8. This illustrates that the plants with higher level of heat integration HIP5, HIP2 and HIP1 can have control performance as good as the base case plant with the right control structures, CS3, CS4, CS7 and CS8. We assert that the control difficulties associated with heat integration is a structural problem. They can be reduced or, in some case, eliminated by using right heat integration structure with correct control structure. As illustrated here, HIP5 with CS7 (total IAE =1.7509) is even better than BC with CS3 (total IAE = 5.8045).

Table 5.4 The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

Integral Absolute Error											
		BC	**HIP1**	**HIP2**	**HIP3**	**HIP4**	**HIP5**	**HIP6**	**HIP7**	**AVG**	**SD**
CCDt	CS1	1.5993	2.4432	2.2231	1.6398	1.5927	2.5345	1.5815	1.5162	1.8913	0.4035
	CS2	1.5688	1.9769	1.5938	1.6246	1.5673	2.5345	1.5536	1.4841	1.7380	0.3319
	CS3	0.2129	0.2129	0.2104	0.2509	0.2591	0.2129	0.2104	0.2833	0.2316	0.0268
	CS4	0.2180	0.1597	0.1597	0.2534	0.2458	0.1774	0.3750	0.2825	0.2339	0.0683
	CS5	1.6031	2.4270	2.2210	1.6408	1.5875	2.4942	1.5794	-	1.9362	0.7383
	CS6	1.5679	1.9664	1.5904	1.6217	1.5657	1.8755	1.5554	-	1.6776	0.5739
	CS7	0.2114	0.2125	0.2114	0.2490	0.2531	0.2119	0.2098	-	0.2227	0.0756
	CS8	0.2196	0.1600	0.1615	0.2488	0.2245	0.1784	0.1937	-	0.1981	0.0718
TCR	CS1	1.2255	0.9852	1.0169	1.0197	0.8368	0.9877	1.6289	0.5780	1.0348	0.2834
	CS2	1.2251	0.9603	1.0165	1.0091	0.8365	0.9877	1.7159	0.5783	1.0412	0.3075
	CS3	1.1868	0.9193	0.9203	0.9713	0.8365	0.6255	1.7138	0.5739	0.9684	0.3352
	CS4	1.1865	0.9283	0.9529	0.9722	0.8672	0.6257	1.7138	0.5745	0.9777	0.3330
	CS5	1.1896	0.9833	0.9210	0.9831	0.8369	0.6257	1.7157	-	1.0365	0.4545
	CS6	1.1897	0.9586	0.9205	0.9743	0.8371	0.6259	1.7158	-	1.0317	0.4539
	CS7	1.1865	0.9192	0.9202	0.9658	0.8360	0.6252	1.7178	-	1.0244	0.4535
	CS8	1.1864	0.9236	0.9202	0.9663	0.8359	0.6253	1.7176	-	1.0250	0.4535
TCC	CS1	2.7242	0.0922	0.2533	0.1630	0.6848	0.1262	2.8377	1.4758	1.0446	1.0900
	CS2	2.7192	0.0925	0.2541	0.1628	0.6831	0.1262	2.9081	1.4701	1.0520	1.1034
	CS3	2.6787	0.0788	0.2529	0.1530	0.7107	0.1249	2.8833	1.3895	1.0340	1.0886
	CS4	2.6780	0.0788	0.2536	0.1536	0.7129	0.1250	2.8824	1.4011	1.0357	1.0886
	CS5	2.7277	0.0790	0.2527	0.1657	0.7102	0.1246	2.9139	-	0.9962	1.1440
	CS6	2.7177	0.0794	0.2534	0.1657	0.7112	0.1249	2.9063	-	0.9941	1.1401
	CS7	2.6816	0.0789	0.2529	0.1581	0.6805	0.1250	2.8920	-	0.9813	1.1308
	CS8	2.6805	0.0788	0.2528	0.1585	0.6808	0.1250	2.8920	-	0.9812	1.1306
PCD	CS1	1.3364	1.7058	1.7069	1.2696	1.2297	1.9078	1.3131	1.1904	1.4575	0.2551
	CS2	1.3446	1.3242	1.4564	0.4356	1.2706	1.5797	1.3271	1.2602	1.2498	0.3232
	CS3	0.4005	0.3068	0.3046	0.5285	0.4866	0.3655	0.3994	0.7503	0.4428	0.1374
	CS4	0.9168	0.7689	0.7765	0.9698	0.9259	0.7725	0.8931	0.9663	0.8737	0.0817
	CS5	1.3387	1.7050	1.7096	1.2727	1.2328	1.9069	1.3109	-	1.4966	0.5473
	CS6	1.3446	1.3219	1.4541	1.2857	1.2698	1.5786	1.3287	-	1.3691	0.4627
	CS7	0.4021	0.3061	0.3037	0.5280	0.4420	0.3647	0.4002	-	0.3924	0.1466
	CS8	0.9184	0.7720	0.7766	0.9529	0.9101	0.7733	0.8999	-	0.8576	0.2919
PCP	CS1	0.0657	0.1028	0.1609	1.8230	4.4858	0.2483	0.0537	2.8012	1.2177	1.5627
	CS2	0.1057	0.0663	0.1174	5.2696	4.3860	0.1987	0.0815	2.7982	1.6279	2.0527
	CS3	0.2102	0.3610	0.3516	1.1147	2.9442	0.3063	0.2132	1.1971	0.8373	0.8784
	CS4	0.2146	0.3946	0.3874	1.1084	2.9405	0.3384	0.2107	1.1836	0.8473	0.8692
	CS5	0.2323	0.5481	0.4730	1.0561	2.7342	0.5484	0.2285	-	0.8315	0.8130
	CS6	0.2429	0.4871	0.4018	1.0875	2.7349	0.4451	0.2518	-	0.8073	0.8208
	CS7	0.0580	0.1012	0.0825	1.7961	4.4936	0.0454	0.0826	-	0.9514	1.4960
	CS8	0.0547	0.0791	0.0893	1.8103	4.4896	0.0488	0.0558	-	0.9468	1.4986

Table 5.4 Continued The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

Integral Absolute Error											
		BC	**HIP1**	**HIP2**	**HIP3**	**HIP4**	**HIP5**	**HIP6**	**HIP7**	**AVG**	**SD**
Total	CS1	6.950958	5.3292	5.3612	5.9151	8.8298	5.8045	7.4150	7.5616	6.6459	1.1683
	CS2	6.963459	4.4202	4.4382	8.5017	8.7434	5.4267	7.5862	7.5909	6.7088	1.6203
	CS3	4.689058	1.8788	2.0397	3.0184	5.2371	1.6351	5.4200	4.1941	3.5140	1.4607
	CS4	5.213912	2.3303	2.5301	3.4574	5.6923	2.0390	6.0750	4.4081	3.9683	1.4954
	CS5	7.091384	5.7424	5.5771	5.1183	7.1017	5.6999	7.7483	-	6.2970	2.2542
	CS6	7.06281	4.8134	4.6203	5.1349	7.1188	4.6500	7.7580	-	5.8798	2.2777
	CS7	4.539534	1.6179	1.7707	3.6970	6.7051	1.3722	5.3024	-	3.5721	2.1456
	CS8	5.059626	2.0135	2.2004	4.1368	7.1409	1.7509	5.7590	-	4.0087	2.2474
TCDb	CS1	-	-	-	-	-	-	-	-	-	-
	CS2	-	-	-	-	-	-	-	-	-	-
	CS3	0.3040	0.6090	0.4709	0.2314	0.2647	1.2658	0.3533	0.3374	0.4796	0.3181
	CS4	0.3382	0.7195	0.5413	0.2635	0.2924	1.2998	0.3455	0.3418	0.5177	0.3281
	CS5	-	-	-	-	-	-	-	-	-	-
	CS6	-	-	-	-	-	-	-	-	-	-
	CS7	0.2989	0.6212	0.4558	0.2286	0.2748	1.2668	0.3442	-	0.4986	0.3555
	CS8	0.3426	0.7478	0.5475	0.2636	0.2808	1.3042	0.6277	-	0.5877	0.3714
TCPb	CS1	-	-	-	-	-	-	-	31.1428	31.1428	10.2995
	CS2	-	-	-	-	-	-	-	30.9412	30.9412	10.2329
	CS3	0.1803	0.3510	0.2310	2.2038	6.2213	0.2348	0.0867	23.2511	4.0950	7.5024
	CS4	0.1901	0.3749	0.2506	2.1953	6.2380	0.2532	0.0922	23.3407	4.1169	7.5270
	CS5	0.1435	0.2615	0.2158	2.1135	5.7363	0.2434	0.1030	-	1.2596	1.8662
	CS6	0.1297	0.2155	0.1806	2.1245	5.7948	0.2078	0.0997	-	1.2504	1.8930
	CS7	-	-	-	-	-	-	-	-	-	-
	CS8	-	-	-	-	-	-	-	-	-	-

Note CC = Composition Control, TC = Temperature Control, PC= Pressure Control, R = Reactor, D = DIB, P = Purge, C = Cooler, t = top, b = bottom

Table 5.5 Energy increases as the temperature of reactor feed increase 10 °F

Energy Increase (10^5, Btu/ hr)											
		BC	**HIP1**	**HIP2**	**HIP3**	**HIP4**	**HIP5**	**HIP6**	**HIP7**	**AVG**	**SD**
Furnace	CS1	1.2064	-0.6220	3.6875	-0.7492	3.7394	3.7382	-	3.8799	2.1257	1.9791
	CS2	1.2058	-0.6285	3.6880	-0.7503	3.7391	3.7380	-	3.8800	2.1246	1.9803
	CS3	1.2635	-0.3025	3.6363	-0.5310	3.7053	3.7732	-	3.9358	2.2115	1.8938
	CS4	1.2636	-0.2993	3.6359	-0.5411	3.7055	3.7731	-	3.9396	2.2111	1.8955
	CS5	1.2064	-0.6262	3.6873	-0.7491	3.7376	3.7375	-	-	1.8323	1.8986
	CS6	1.2055	-0.6287	3.6881	-0.7506	3.7380	3.7375	-	-	1.8316	1.8993
	CS7	1.2637	-0.3022	3.6357	-0.5350	3.7055	3.7729	-	-	1.9234	1.8204
	CS8	1.2637	-0.3015	3.6355	-0.5354	3.7056	3.7730	-	-	1.9235	1.8203
DIB column reboiler	CS1	6.7213	5.8990	2.8920	7.0392	6.7959	1.5735	6.6875	6.4244	5.5041	1.9424
	CS2	6.7063	5.8912	2.8917	7.0205	6.7568	1.5470	6.6359	6.3909	5.4800	1.9368
	CS3	2.3472	-1.3447	-3.9949	1.6709	1.5140	-5.8827	2.4755	-1.5221	-0.5921	2.9285
	CS4	2.2974	-1.3441	-3.9881	1.2767	1.4650	-5.8572	2.4820	-1.4410	-0.6387	2.8733
	CS5	6.7370	5.8552	2.8797	7.0338	6.8022	1.5650	6.6657	-	5.3627	2.6087
	CS6	6.7032	5.8997	2.8615	6.9862	6.7624	1.5405	6.6323	-	5.3408	2.6008
	CS7	2.3404	-1.3161	-3.9661	1.5314	1.4670	-5.8957	2.4592	-	-0.4828	2.8900
	CS8	2.3093	-1.3170	-4.0026	1.5772	1.5113	-5.8608	2.4701	-	-0.4732	2.8926
Purge column reboiler	CS1	0.0609	0.0815	0.0736	-0.7461	-4.1634	0.0823	0.0581	-	-0.6504	1.3841
	CS2	0.0612	0.0813	0.0743	-0.7472	-4.1590	0.0822	0.0599	-	-0.6496	1.3828
	CS3	0.0333	0.0377	0.0359	-0.7950	-4.0410	0.0378	0.0307	-	-0.6658	1.3348
	CS4	0.0357	0.0373	0.0348	-0.7929	-4.0314	0.0383	0.0331	-	-0.6636	1.3318
	CS5	0.0616	0.0916	0.0808	-0.7299	-4.1555	0.0871	0.0578	-	-0.6438	1.3825
	CS6	0.0643	0.0856	0.0793	-0.7358	-4.1509	0.0781	0.0666	-	-0.6447	1.3808
	CS7	0.0356	0.0333	0.0364	-0.8112	-4.0359	0.0386	0.0312	-	-0.6674	1.3334
	CS8	0.0348	0.0284	0.0308	-0.8110	-4.0343	0.0355	0.0342	-	-0.6688	1.3322
Hot utilities usage, (Btu/hr)	CS1	7.9886	5.3585	6.6531	5.5438	6.3719	5.3940	6.7456	10.3044	6.7950	1.5626
	CS2	7.9733	5.3441	6.6540	5.5230	6.3369	5.3673	6.6958	10.2709	6.7707	1.5599
	CS3	3.6440	-1.6096	-0.3228	0.3450	1.1783	-2.0717	2.5062	2.4137	0.7604	1.9123
	CS4	3.5967	-1.6061	-0.3174	-0.0573	1.1391	-2.0458	2.5151	2.4987	0.7154	1.9228
	CS5	8.0050	5.3207	6.6478	5.5547	6.3843	5.3896	6.7235	-	6.2894	2.2408
	CS6	7.9730	5.3567	6.6289	5.4999	6.3495	5.3561	6.6989	-	6.2661	2.2315
	CS7	3.6397	-1.5850	-0.2940	0.1851	1.1366	-2.0842	2.4904	-	0.4984	1.8096
	CS8	3.6077	-1.5901	-0.3364	0.2308	1.1826	-2.0523	2.5043	-	0.5067	1.8035
Cold utilities usage, (Btu/hr)	CS1	2.7355	-0.0040	1.1966	0.0143	1.0258	-0.0049	1.5129	5.2682	1.4681	1.6870
	CS2	2.7360	-0.0040	1.1968	0.0143	1.0254	-0.0049	1.5163	5.2699	1.4687	1.6876
	CS3	2.6115	-0.0049	1.1246	0.0135	0.9756	-0.0060	1.3408	5.2144	1.4087	1.6639
	CS4	2.6114	-0.0049	1.1294	0.0136	0.9767	-0.0060	1.3408	5.1407	1.4002	1.6427
	CS5	2.7347	-0.0040	1.1998	0.0145	1.0269	-0.0050	1.5132	-	0.9257	0.9363
	CS6	2.7364	-0.0040	1.1998	0.0144	1.0277	-0.0050	1.5166	-	0.9266	0.9371
	CS7	2.6112	-0.0048	1.1317	0.0137	0.9743	-0.0059	1.3409	-	0.8659	0.8842
	CS8	2.6112	-0.0049	1.1313	0.0137	0.9747	-0.0060	1.3409	-	0.8659	0.8842

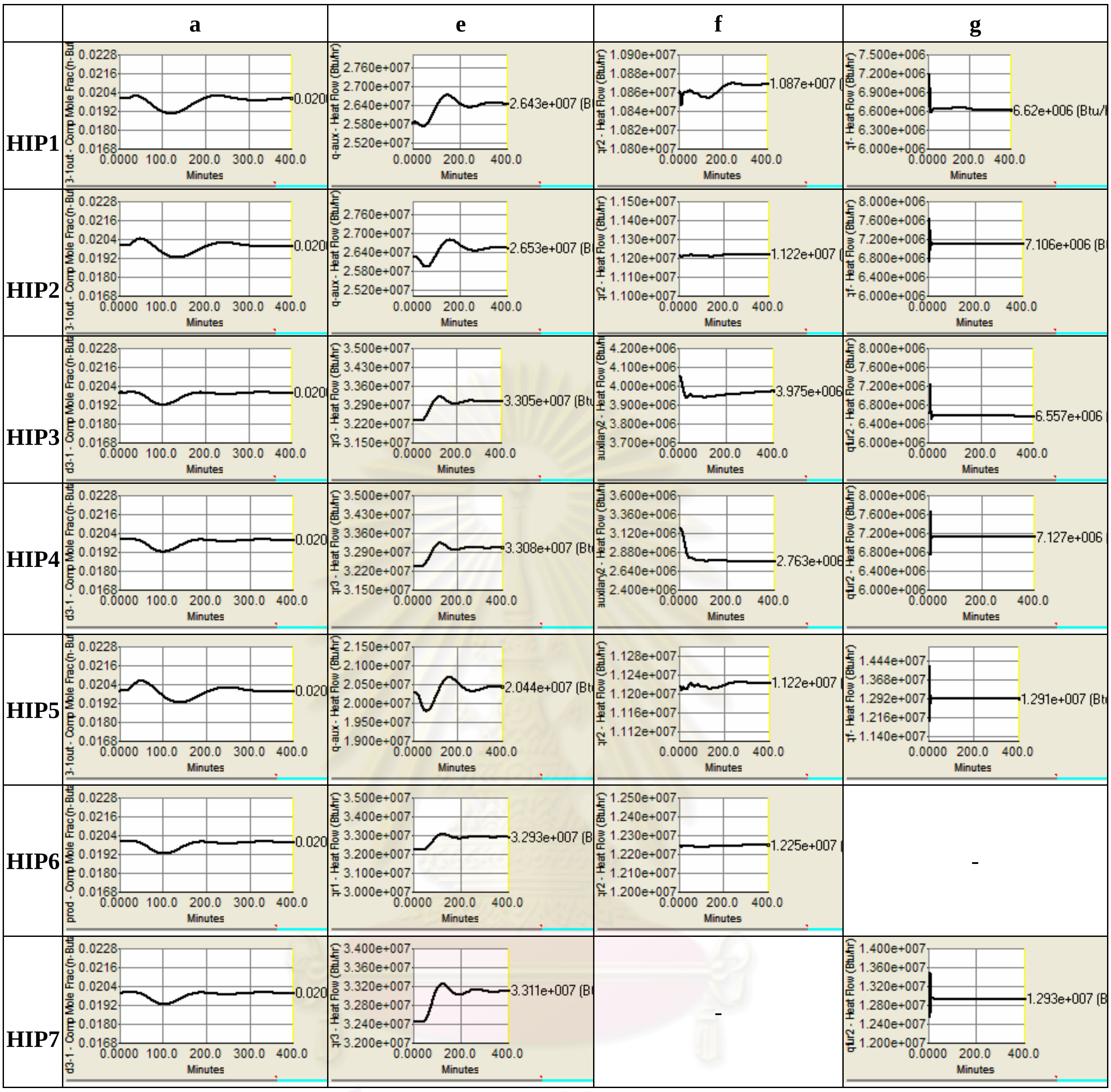


Figure 5.48 Dynamic responses of the Butane Isomerization plant CS1 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

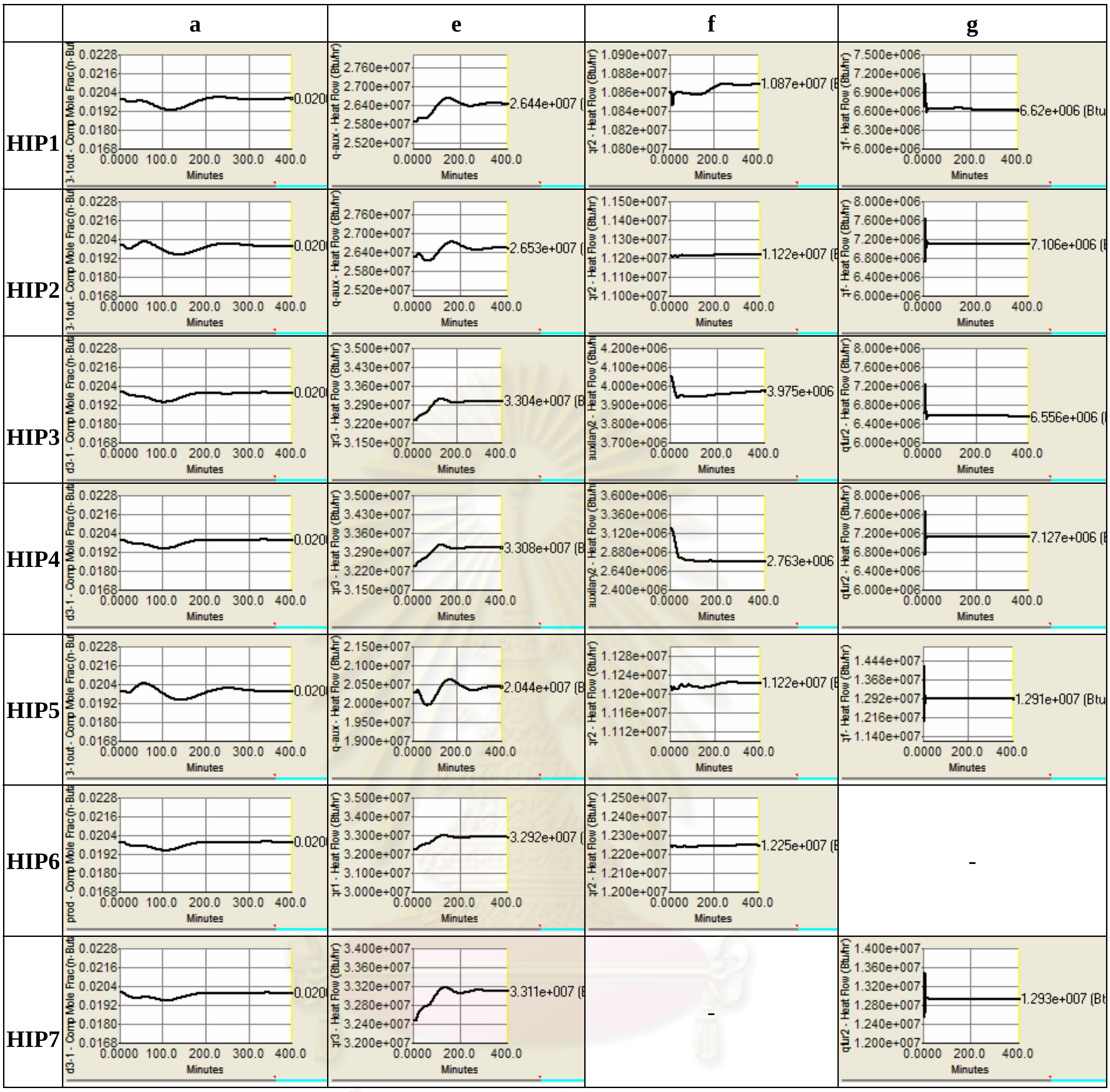


Figure 5.49 Dynamic responses of the Butane Isomerization plant CS2 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

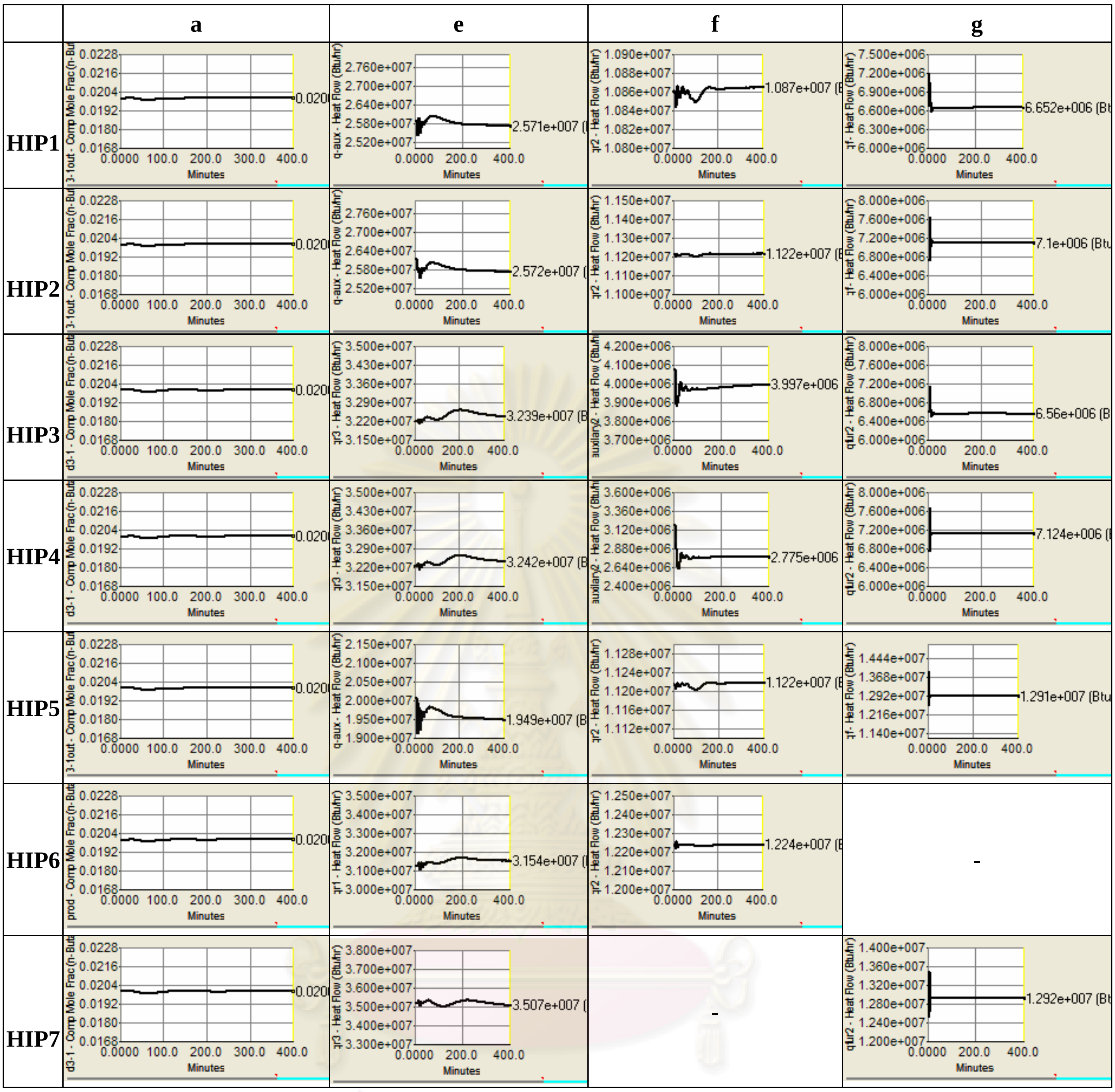


Figure 5.50 Dynamic responses of the Butane Isomerization plant CS3 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

Figure 5.51 Dynamic responses of the Butane Isomerization plant CS4 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

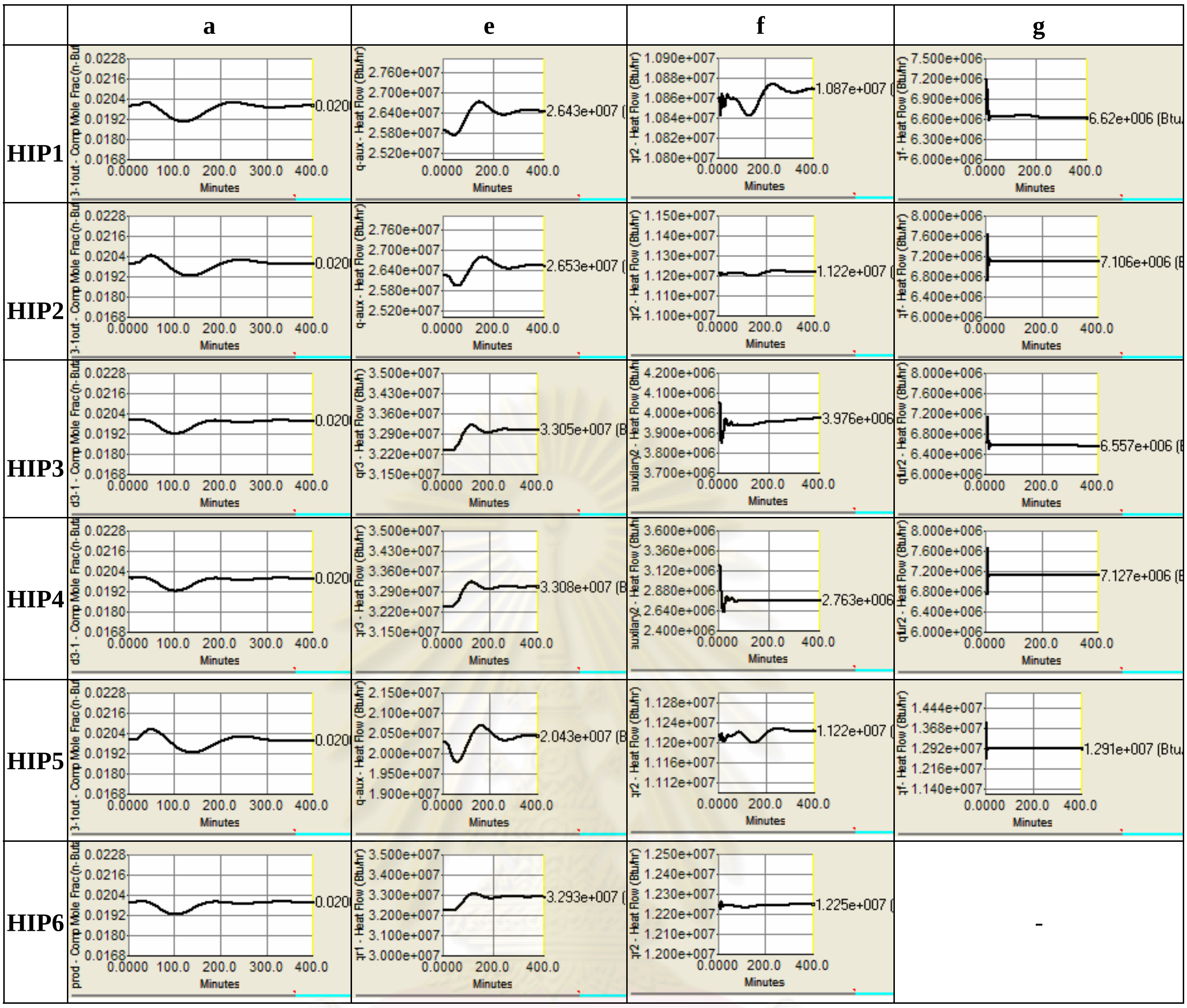


Figure 5.52 Dynamic responses of the Butane Isomerization plant CS5 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

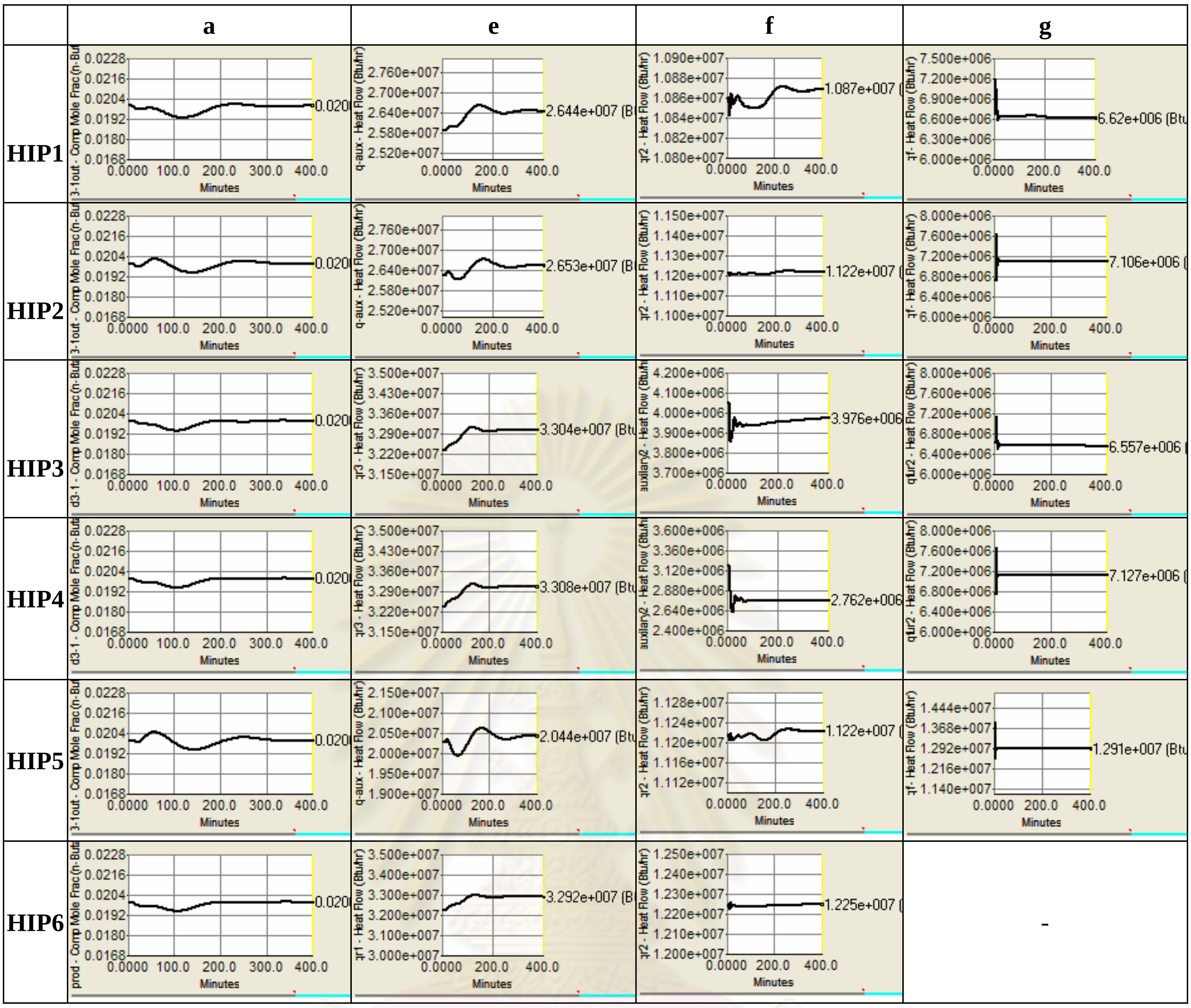


Figure 5.53 Dynamic responses of the Butane Isomerization plant CS6 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

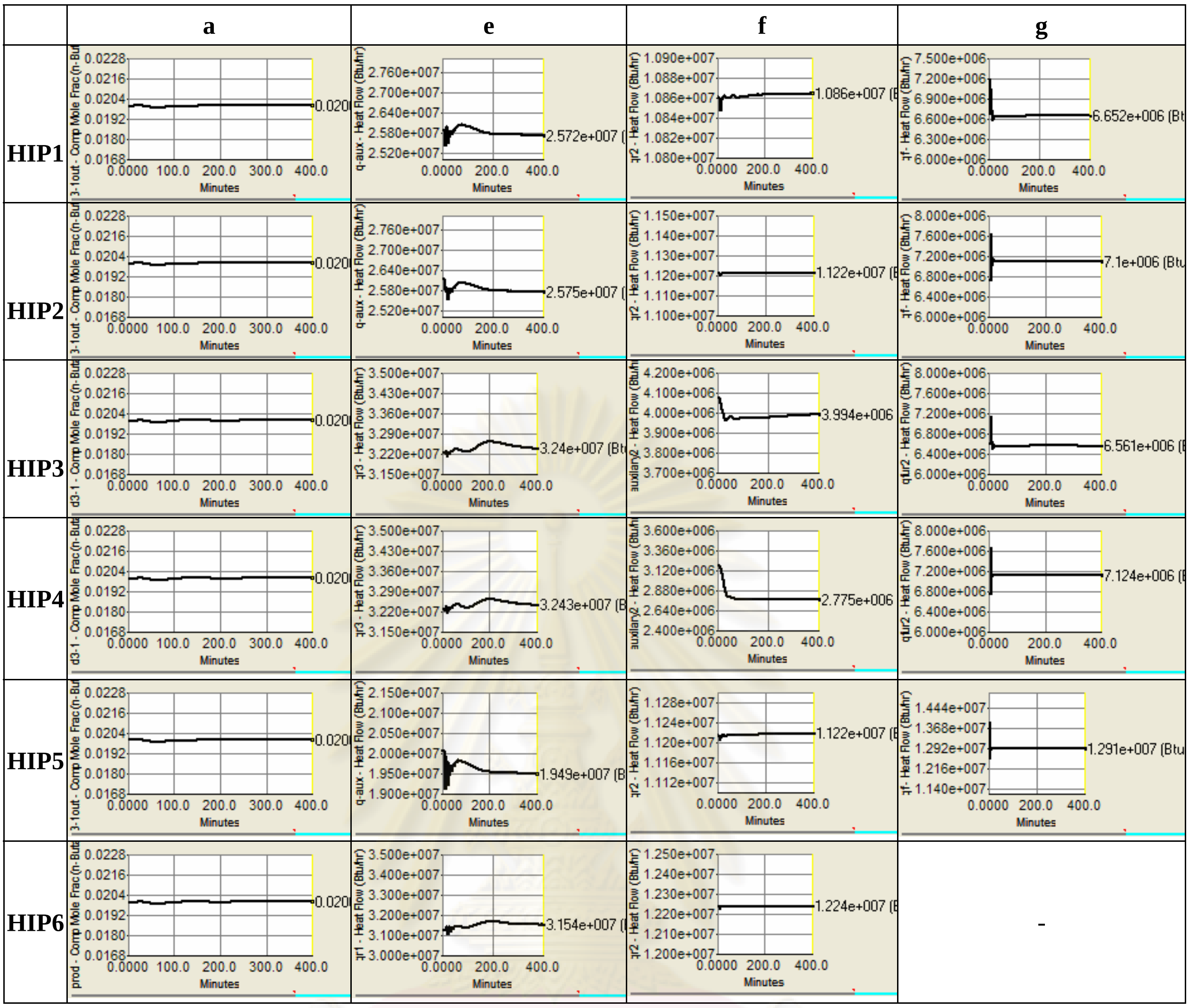


Figure 5.54 Dynamic responses of the Butane Isomerization plant CS7 to 10 ^{o}F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

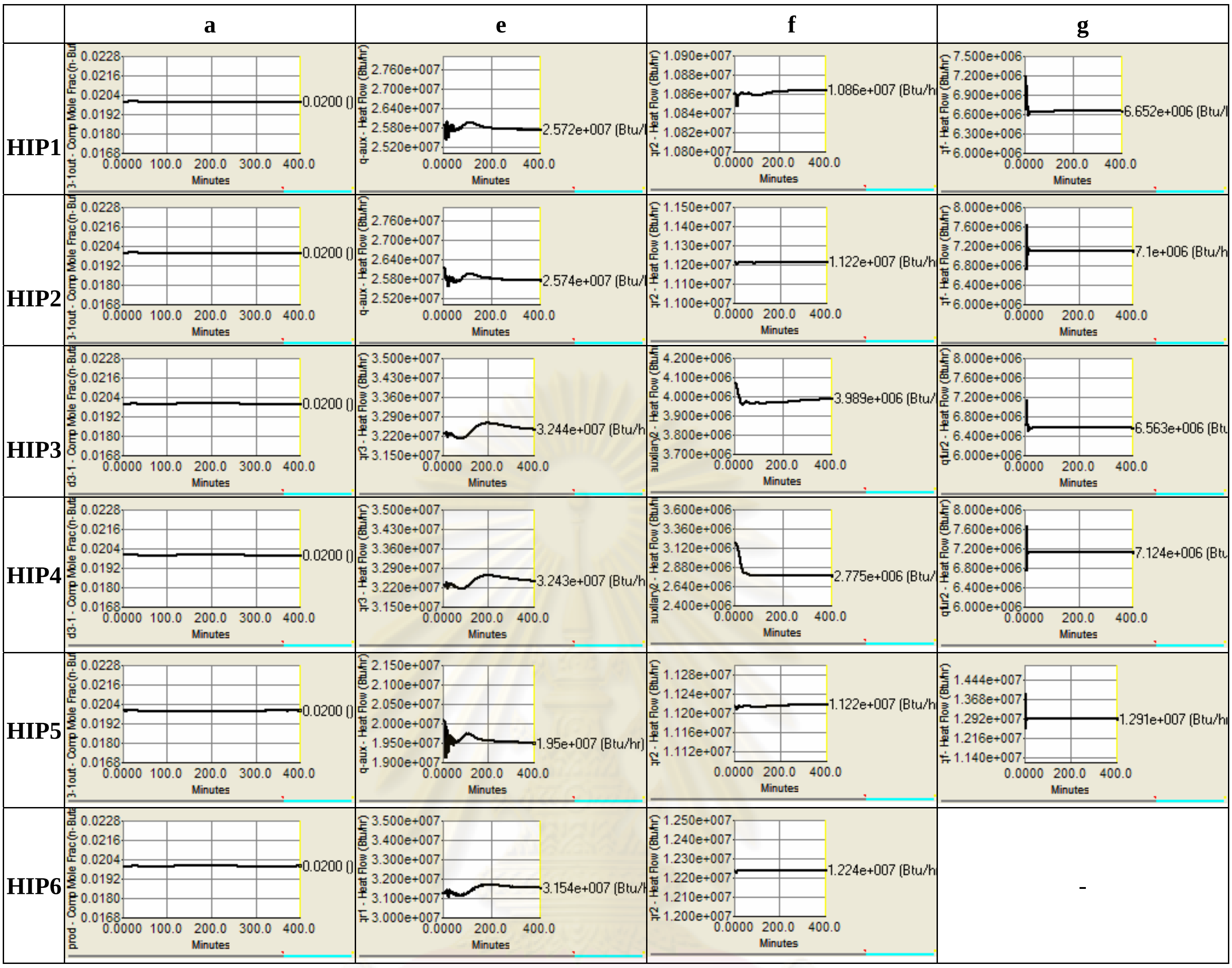


Figure 5.55 Dynamic responses of the Butane Isomerization plant CS8 to 10 °F increase in reactor inlet temperature (the heat load disturbance of hot stream); comparison between HIP1, HIP2, HIP3, HIP4, HIP5, HIP7 and HIP7, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

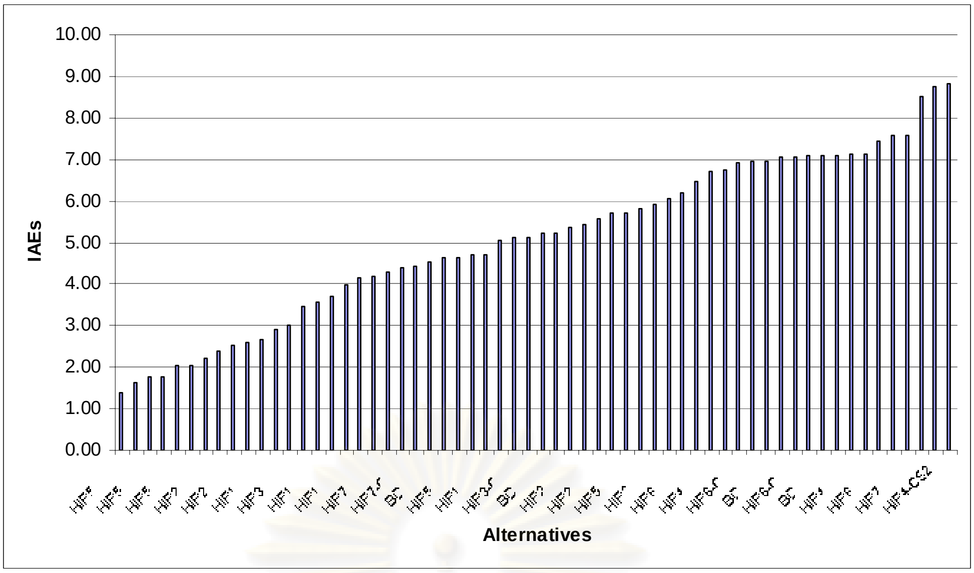


Figure 5.56 The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

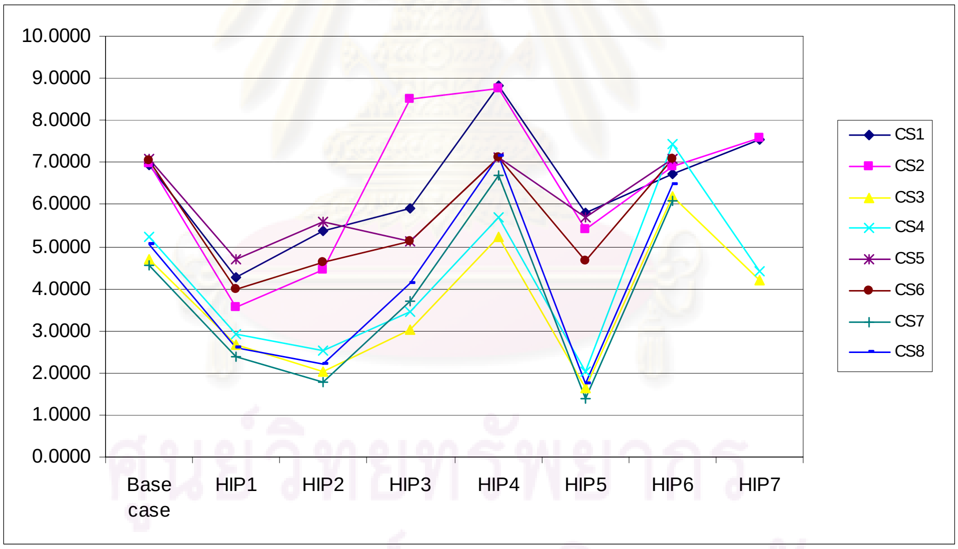


Figure 5.57 The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

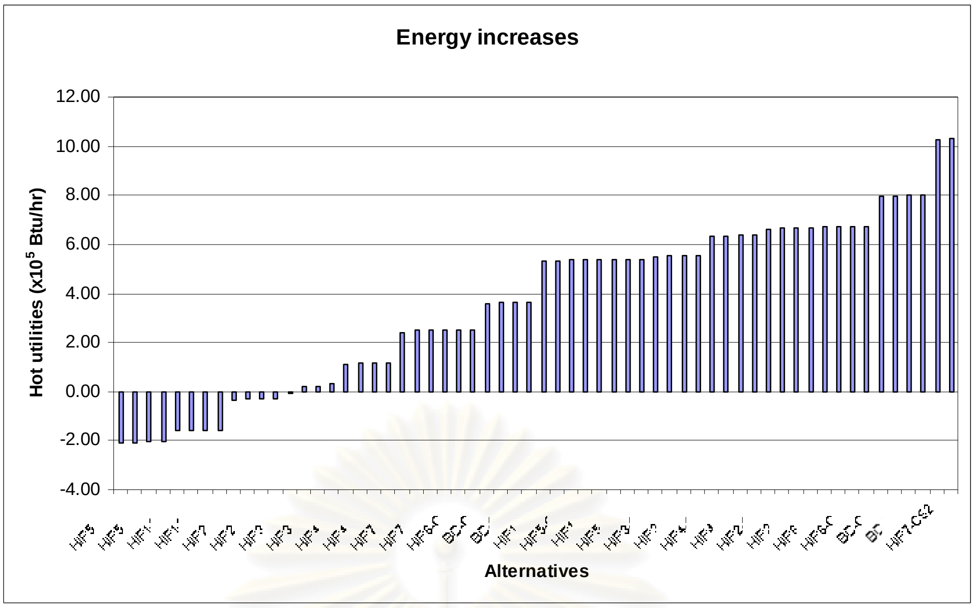


Figure 5.58 Energy increases as the temperature of reactor feed increase 10 °F

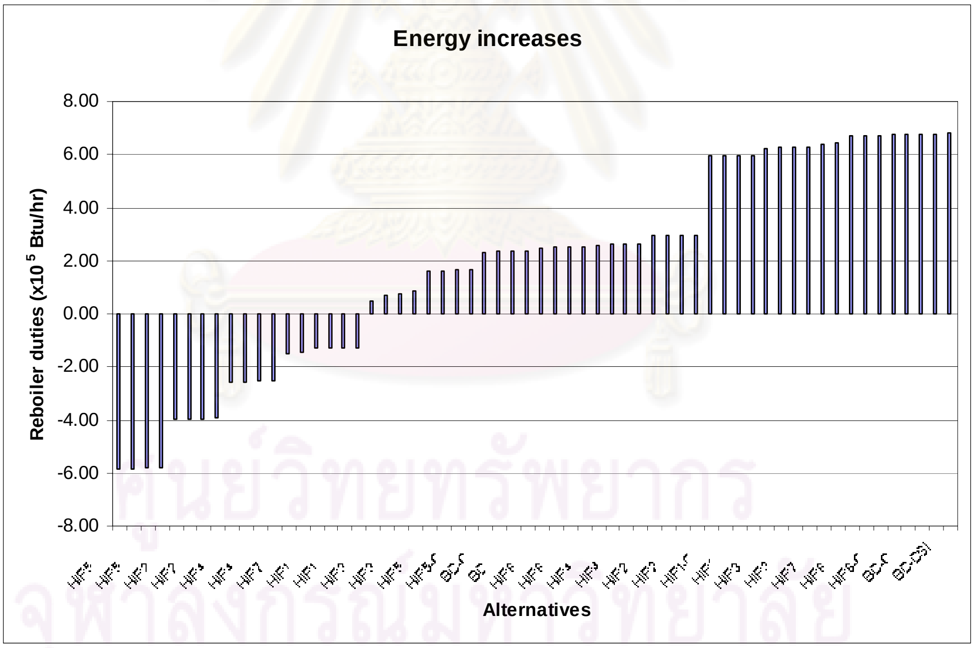


Figure 5.59 Reboiler duty increases as the temperature of reactor feed increase 10 °F

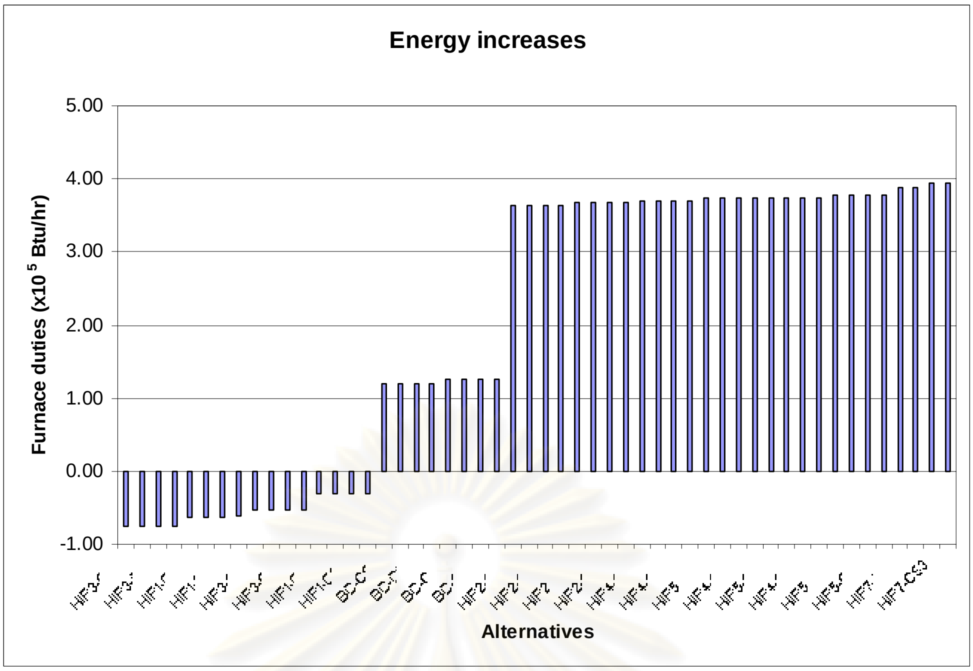


Figure 5.60 Furnace duty increases as the temperature of reactor feed increase 10 °F

5.4.6 Change in the recycle flowrates for CS1 to CS8 in Butane Isomerization plant for Base Case (Luyben et al., 1999)

On the other case, a disturbance in the production rate is also made for this study. Figure 5.93 shows the dynamic responses of CS1-CS8 for the effect of increasing the recycle flowrate from 870 to 940.1lb-mol/h. In CS1 and CS5, the effect is ‘counter-intuitive’ decrease in production rate from 548 to 531.8 lb-mol/h (Luyben, 1998), because of the reactor inlet composition to have a lower nC_4 concentration. In CS2 and CS6, a ratio controller is used to increase the reflux flowrate in the DIB column as its feed increases. The fractionating capability of the column increases as the load is increased, so the bottom does not contain less nC_4 reactant. Product rate increases from 548 to 562.9 lb-mol/h. In CS3, CS4, CS7 and CS8, the base temperature is controlled, in effect; iC_4 in the bottom is slightly changed, so the production rate increases from 536 to 544.7 lb-mol/hr. The reboiler duty in the case of CS2, CS3, CS4, CS6, CS7 and CS8 is higher than of CS1 and CS5 but the furnace duty of CS1 and CS5 is higher, since it has more recycle flow to heat. In terms of impurity, the control structures CS3 and CS4 would be the promising candidate, see Table 5.6.

Table 5.6 IAEs of *n*C4 composition loops at the top of DIB column

Integral Absolute Error							
CS1	CS2	CS3	CS4	CS5	CS6	CS7	CS8
0.2400	3.1231	0.1169	0.1928	0.2265	3.1308	0.1159	0.1938

It is interesting that the energy used in the reboiler of DIB column and furnace of each case is resulted from the control structure implemented, see Table 5.7. Figure 5.93d shows dynamic responses of the temperature on tray1of DIB column.

Table 5.7 Energy increases as the recycle flowrates increase from 870 to 940.1 lb-mol/h.

Energy Increase (10^5,Btu/ hr)				
	DIB column reboiler	Purge column reboiler	Total reboiler	Furnace
CS1	-1.1871	4.8844	3.6973	8.7471
CS2	23.1846	5.0398	28.2244	8.4880
CS3	18.0950	5.0011	23.0961	8.5295
CS4	18.0306	5.0038	23.0345	8.5296
CS5	-1.1733	4.8842	3.7109	8.7472
CS6	23.1854	5.0459	28.2313	8.4878
CS7	18.0946	5.0044	23.0989	8.5294
CS8	18.0551	5.0031	23.0582	8.5295

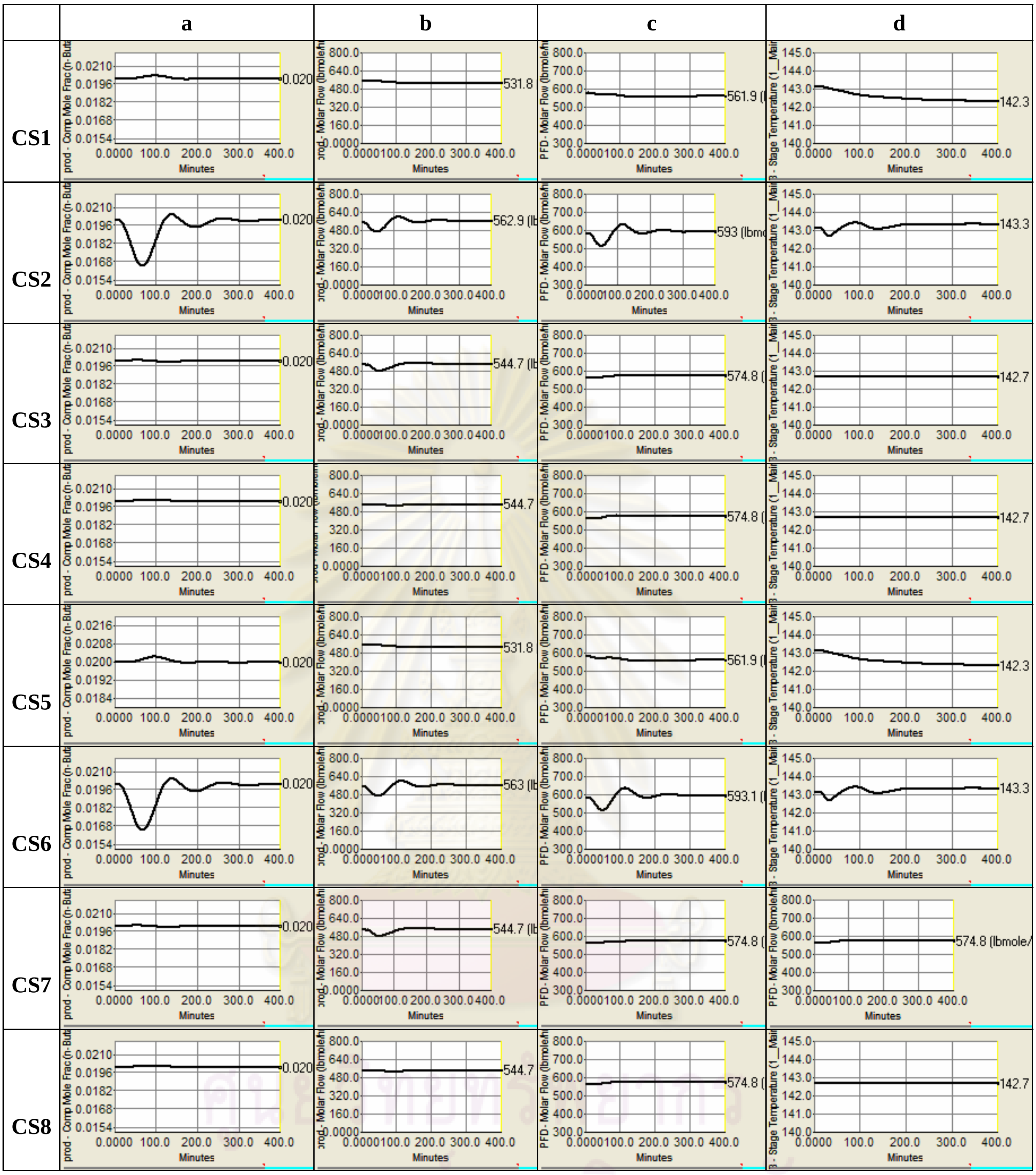


Figure 5.61 Dynamic responses to increase in recycle flows from 870.1 to 940.1 lb-mol/hr of Base case, where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

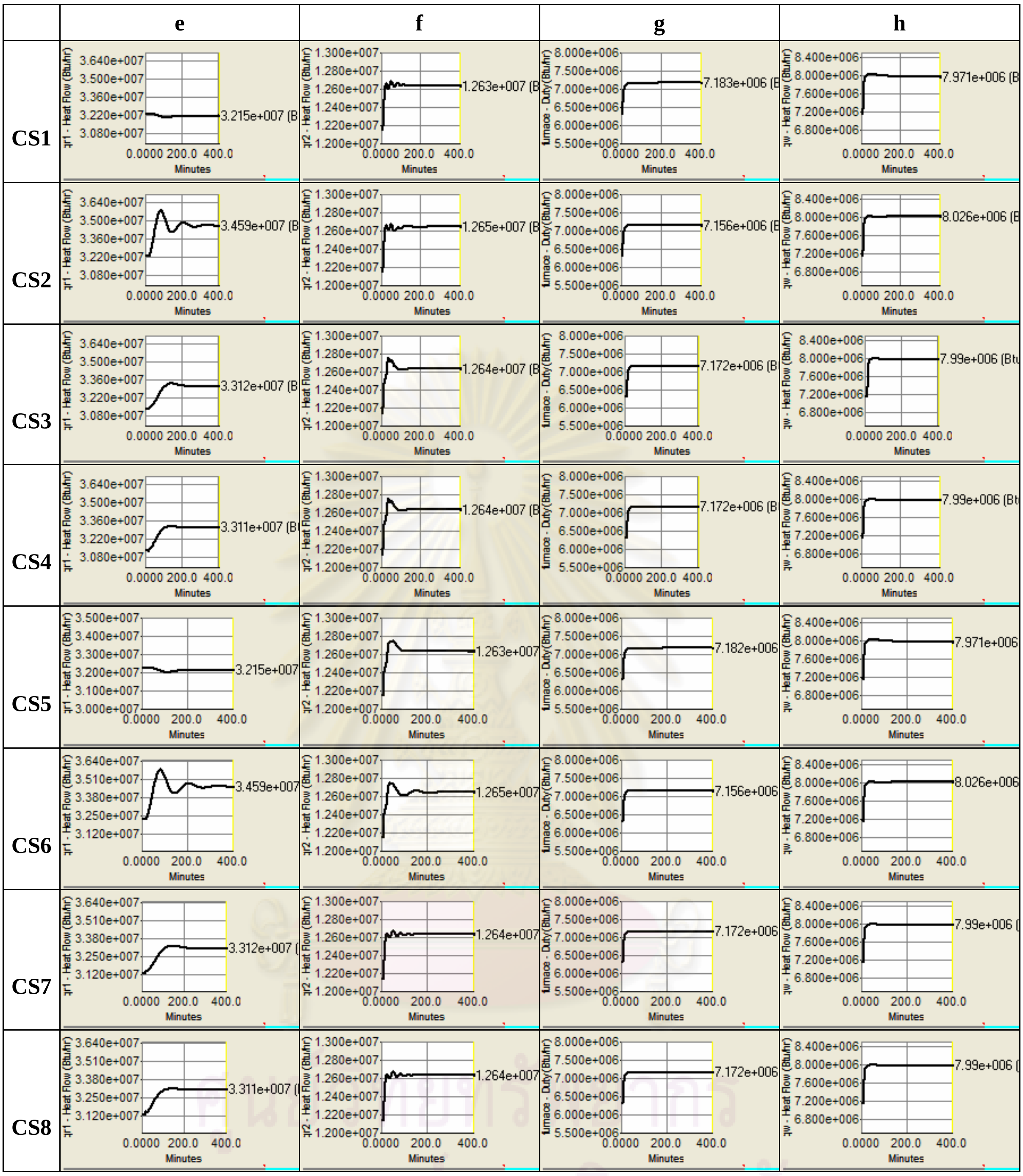


Figure 5.61 Continued Dynamic responses to increase in recycle flows from 870.1 to 940.1 lb-mol/hr of Base Case, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

5.4.7 Change in the recycle flowrates for CS1 to CS8 in Butane Isomerization plant for HIP1, HIP2, HIP3 and HIP4 (Kunajitpimol, 2006)

Figure 5.94-5.97 show dynamic responses of the Butane Isomerization plant HIP1 to HIP4 when the recycle flowrates is changed from 870.1 to 940.1 lb-mol/hr, occurring at time equals 10 minutes. For CS1 and CS5, if no other change is made, the effect is a slight decrease in production rate! This is certainly not what we would intuitively expect. This unusual behavior can be explained by considering what happens in the DIB column. More recycle means a higher feed rate in this column, which increases the column load. Since we are holding the purity of the distillate, the variability is all reflected in the bottoms stream. Thus there is an increase in the iC4 impurity in the bottoms while nC4 impurity in the product increases at first and into its set point after 200 minutes for HIP. This changes the reactor inlet composition to have a lower reactant (nC4) concentration. The result is a slight drop in the overall reaction rate, the heat integration structure HIP1 with CS1 and CS5 decrease from 538 to 530 lb-mol/hr. For CS2 and CS6, the ratio controller is used to increase the reflux flowrate in the DIB column as the recycle flowrate is increased. Now the fractionating capability of the column increases as the load is increased, so the bottoms does not contain less nC4 reactant and the top contain nC4 decrease at first and rise to its set point after 200 minutes, product rate increases from 548 to 560 lb-mol/hr. CS3 and CS7 use reflux valve to control reflux drum level so there is an increase in product flow as increase reflux drum level, increase from 538 to 554 lb-mol/hr. For CS4 and CS8 control the top composition by casecade with reflux flowrate controller so the recycle flowrate increases, the product increase from 538 to 540 lb-mol/hr. Because the base column temperature is controlled for CS3, CS4, CS5 and CS6, the variable nC4 impurity in the product is less than CS1, CS2, CS5 and CS6, see Figure 5.94.

The energy consumption of DIB column reboiler decreases as the product decreases with CS1 and CS5 while the other control structures are increases. The energy consumption of Purge column reboiler, funace and cooler increases as the recycle flowrate increases.

Similar conclusion can be drawn for heat integrated structure HIP2, HIP3 and HIP4 with control structures CS1-CS8. (see Figure 5.95-5.97).

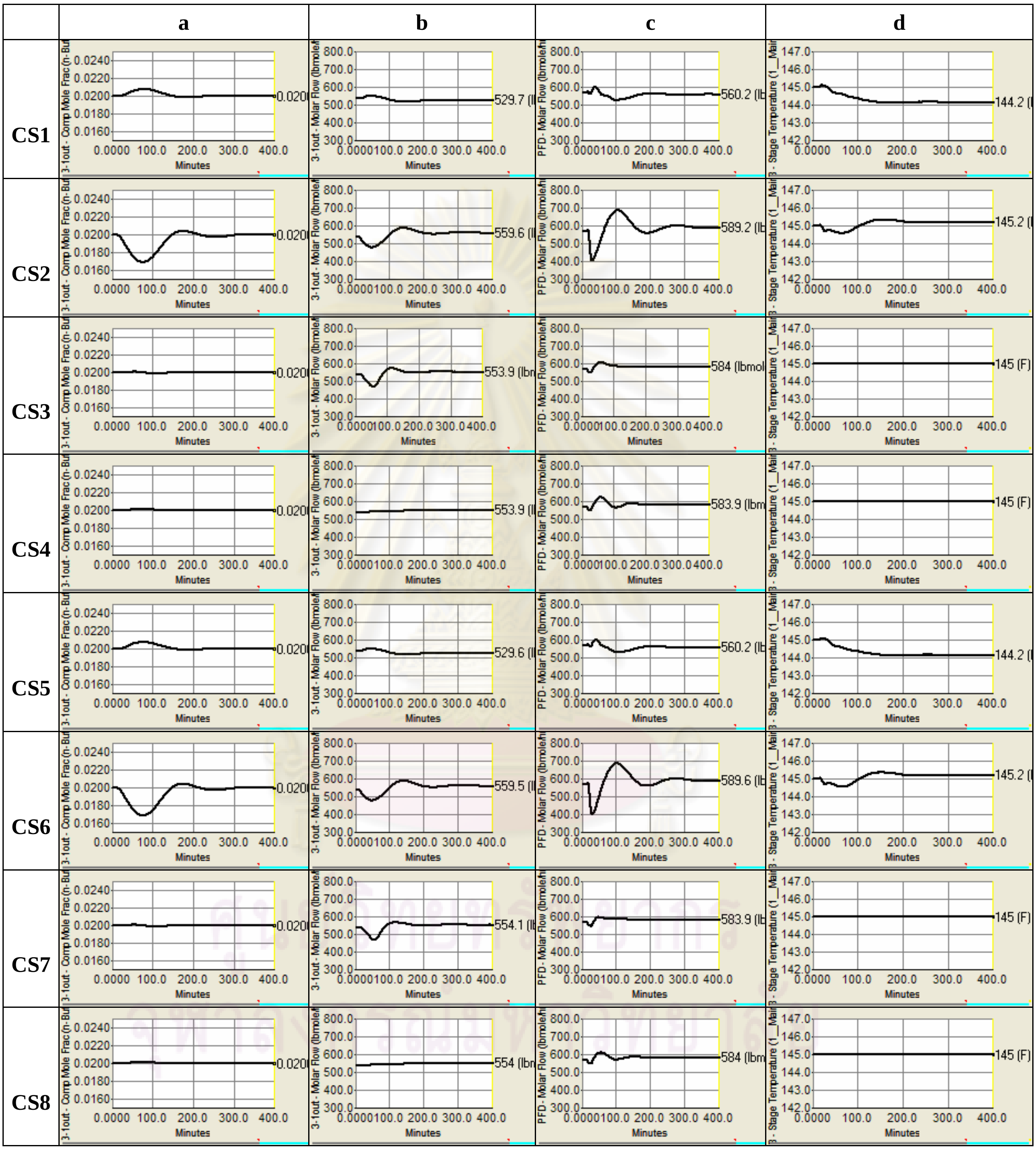


Figure 5.62 Dynamic responses of the Butane Isomerization plant HIP1 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

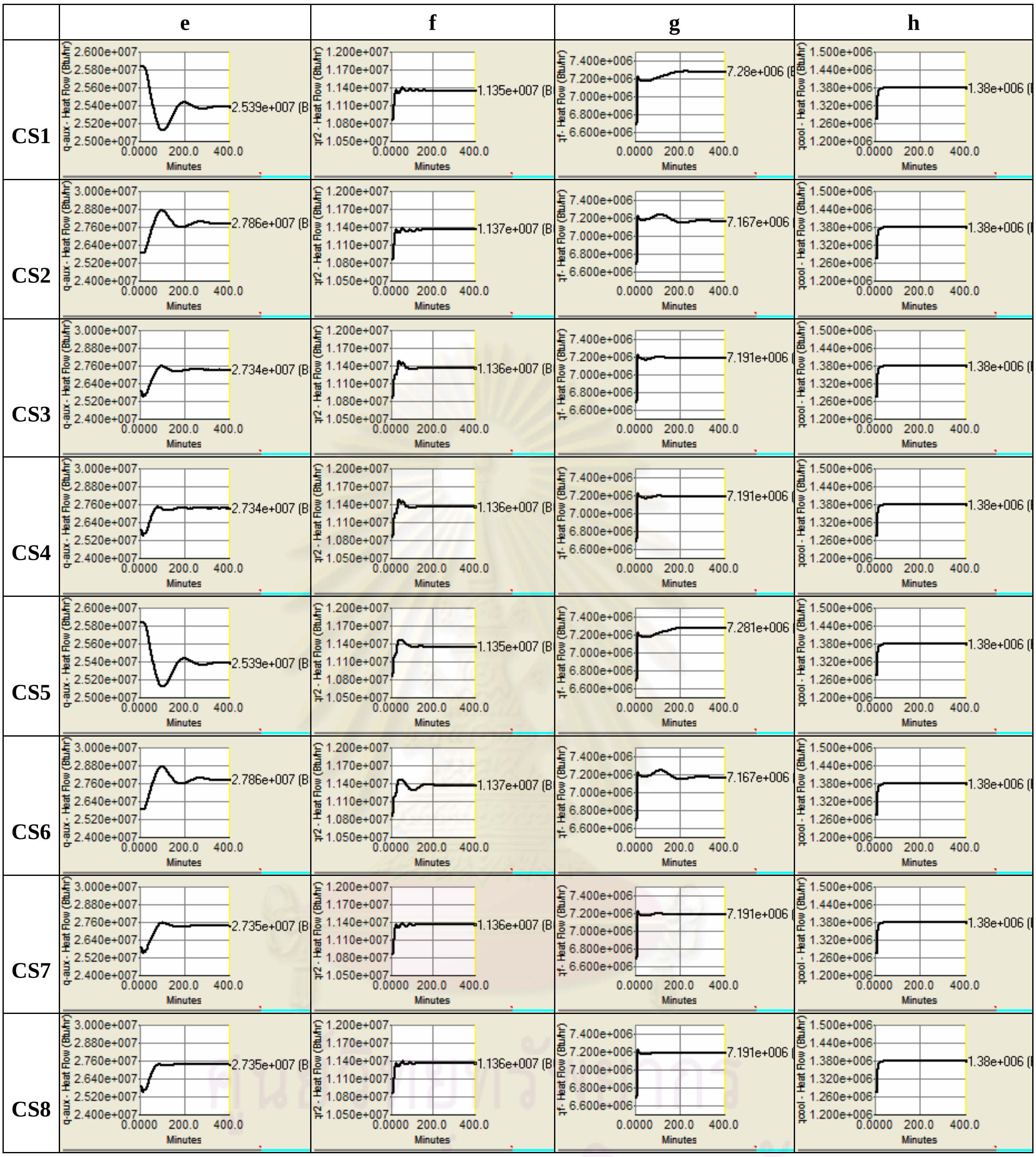


Figure 5.62 Continued Dynamic responses of the Butane Isomerization plant HIP1 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

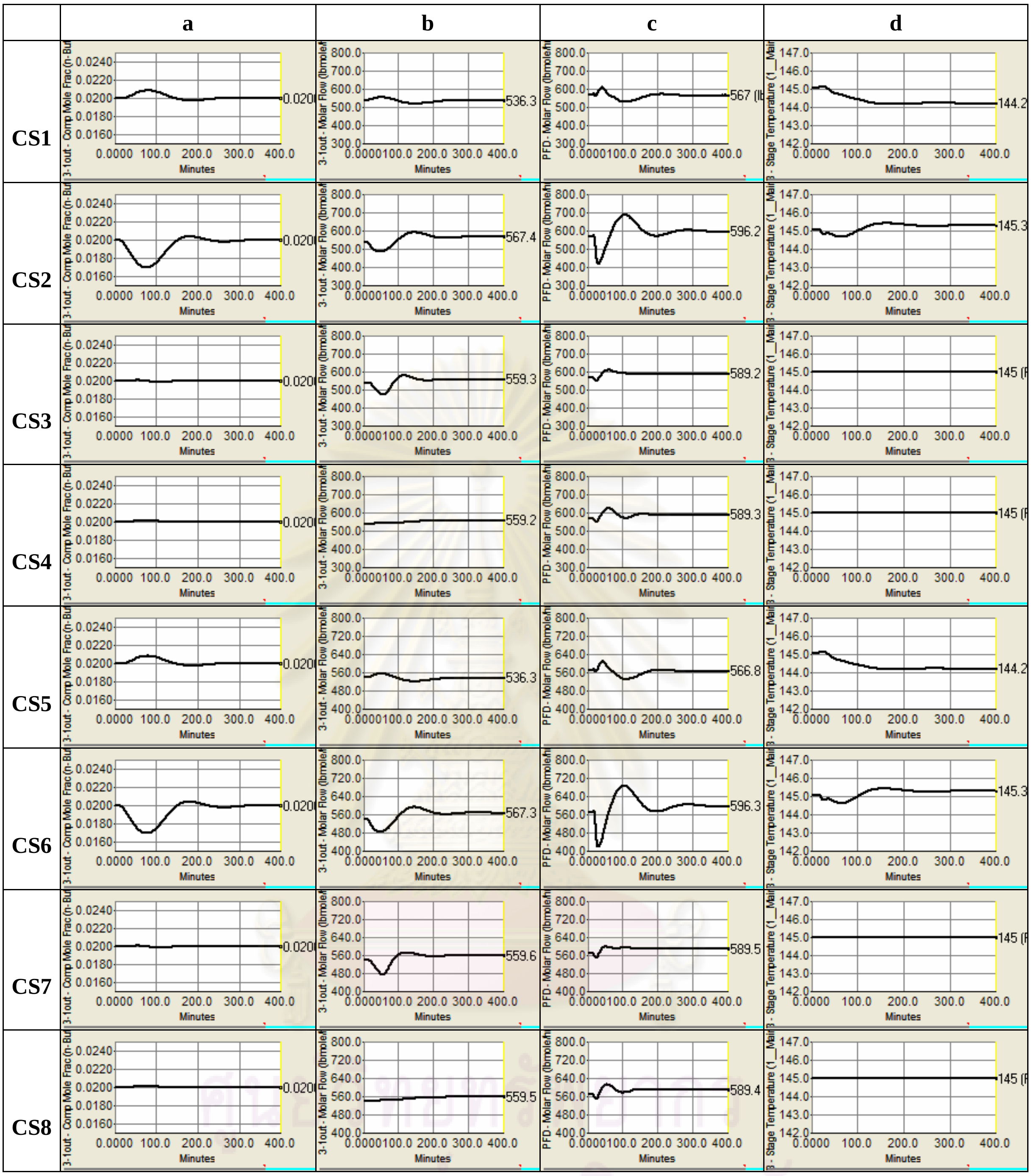


Figure 5.63 Dynamic responses of the Butane Isomerization plant HIP2 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

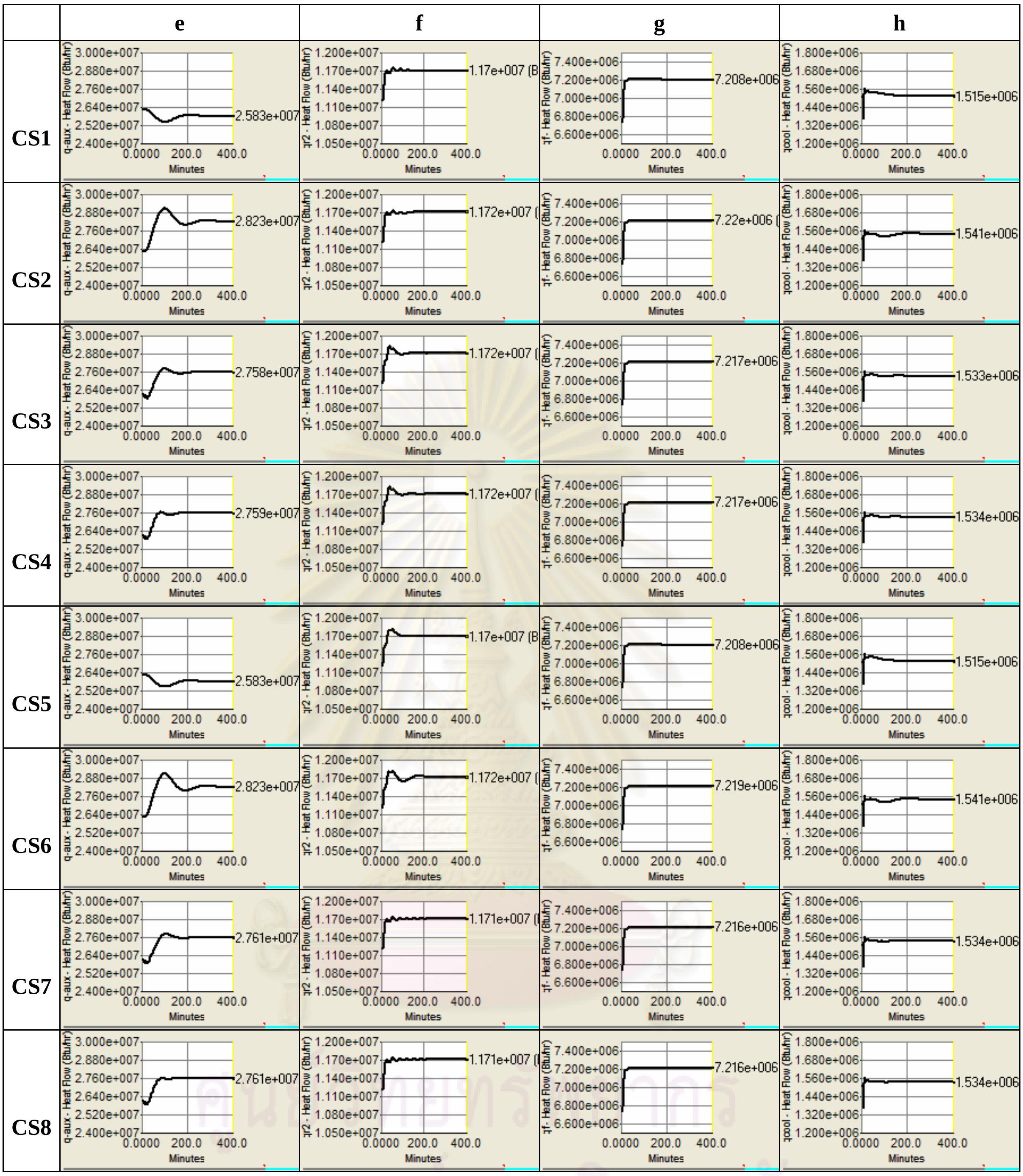


Figure 5.63 Continued Dynamic responses of the Butane Isomerization plant HIP2 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

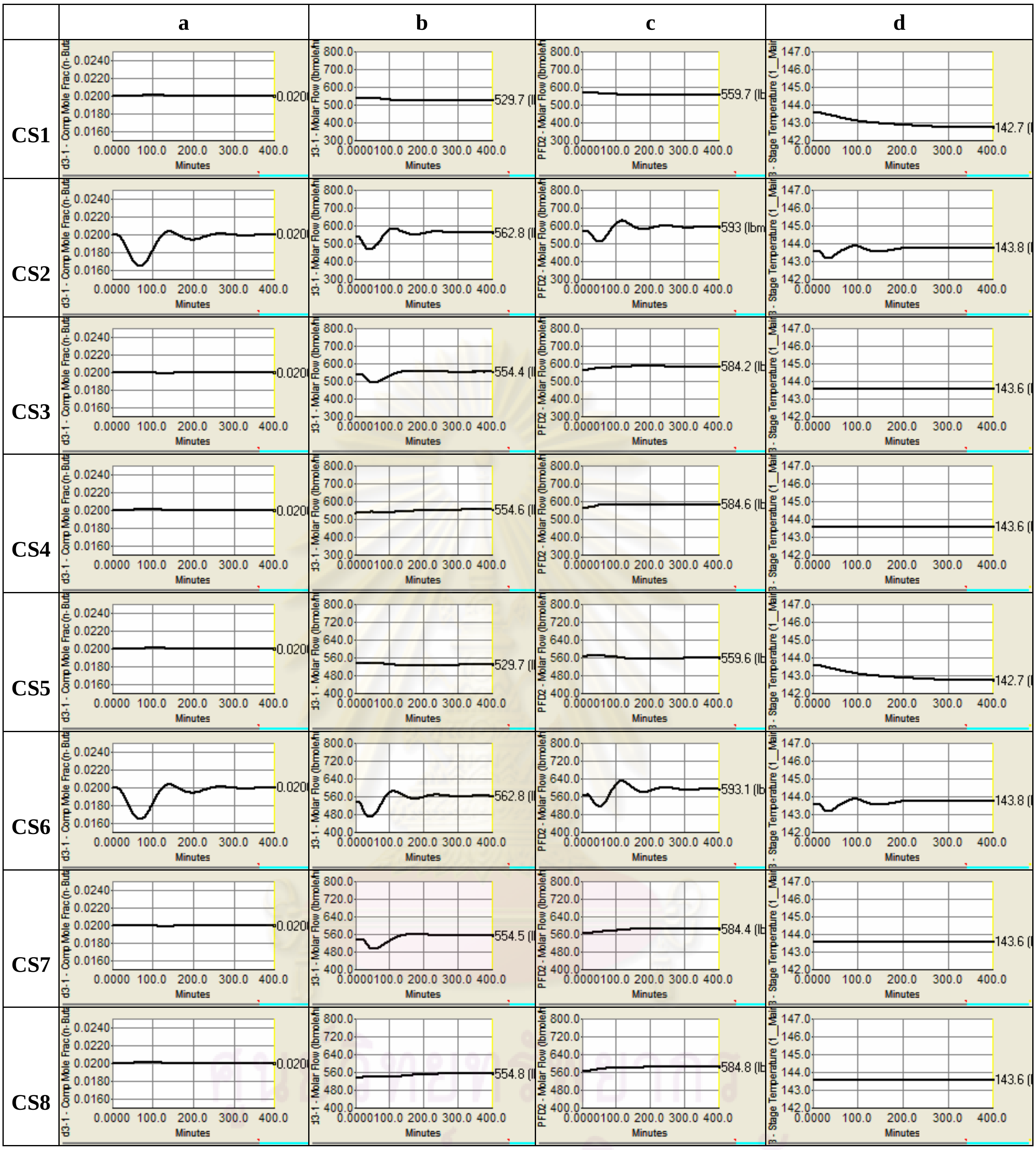


Figure 5.64 Dynamic responses of the Butane Isomerization plant HIP3 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

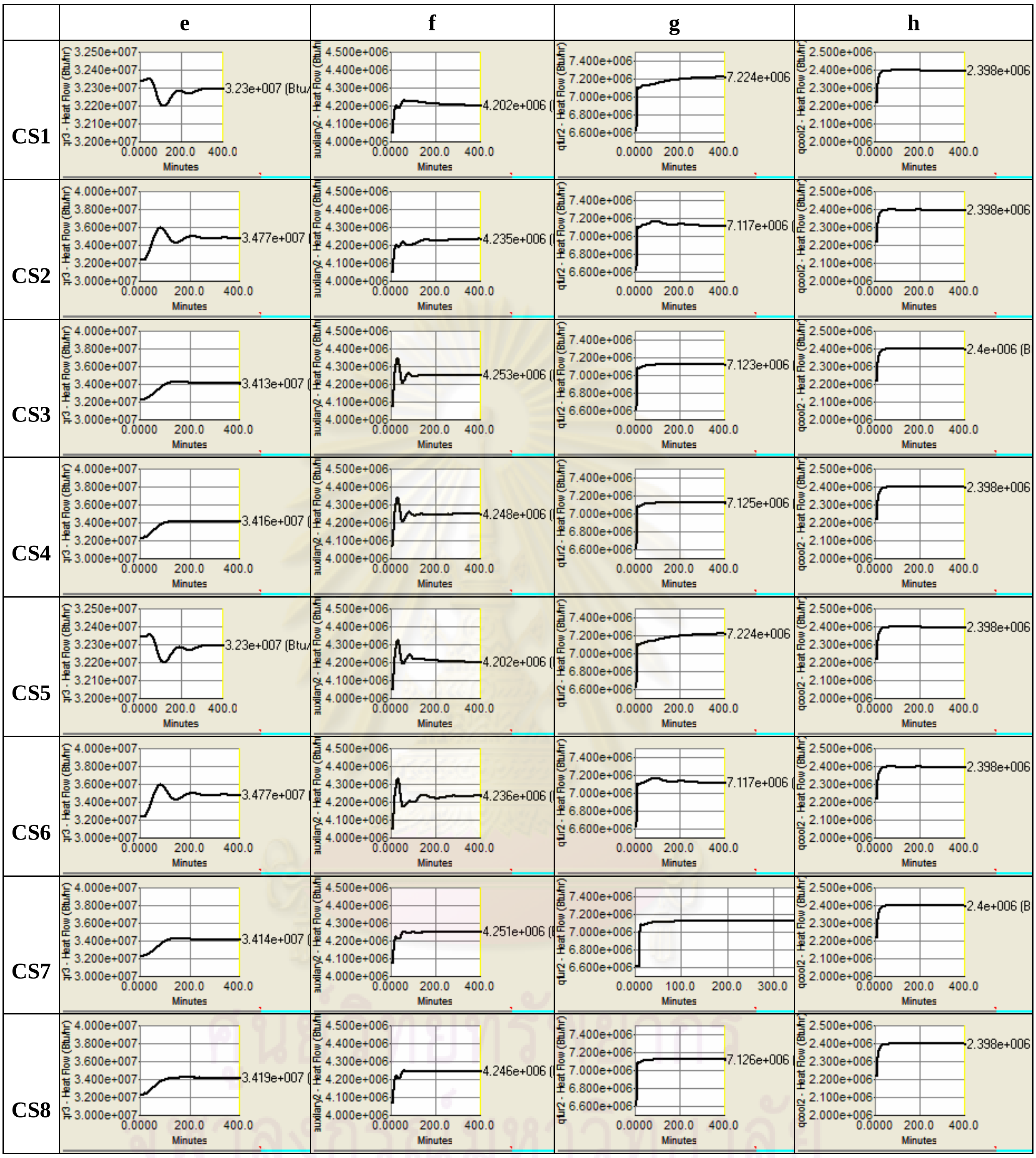


Figure 5.64 Continued Dynamic responses of the Butane Isomerization plant HIP3 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

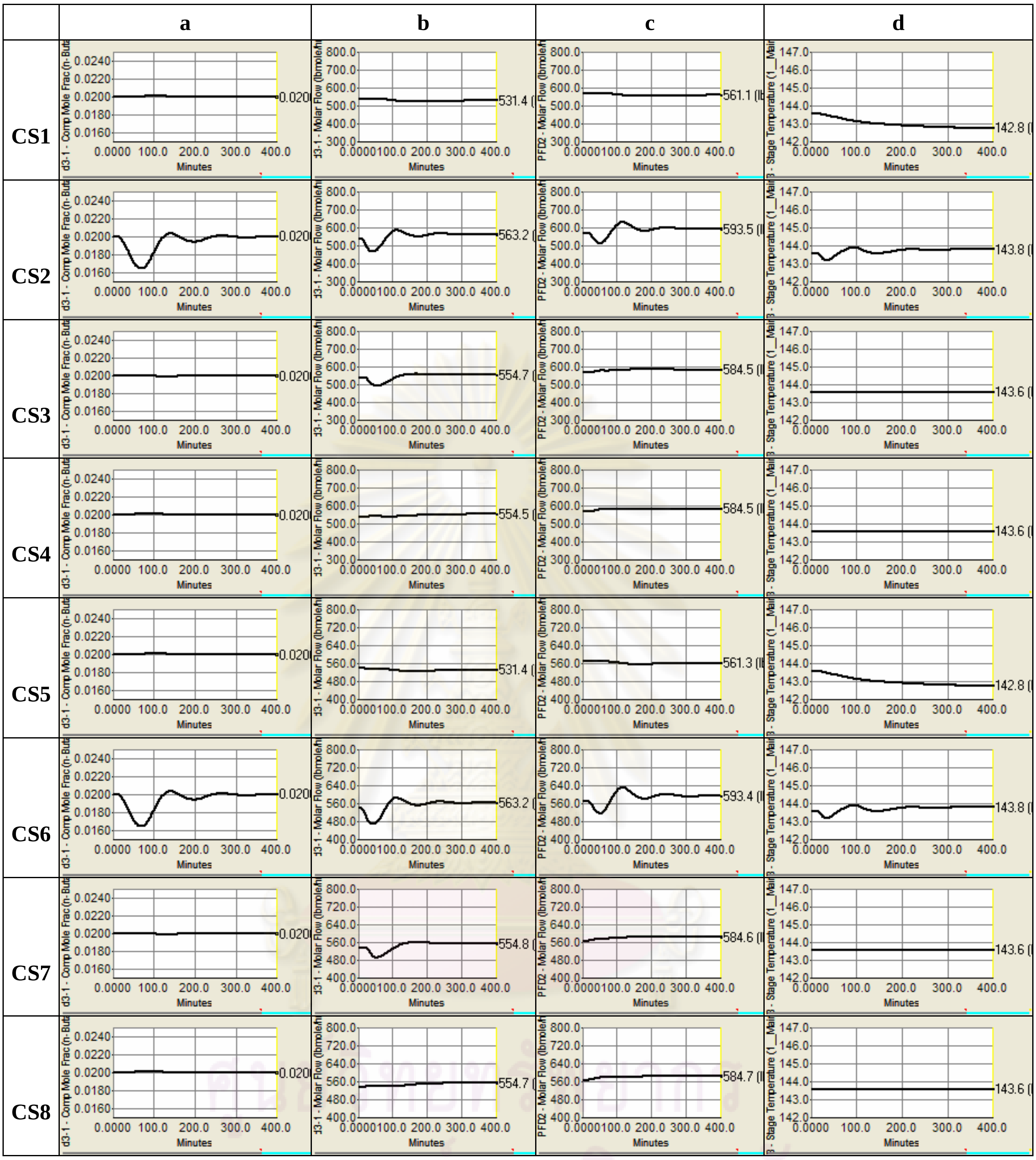


Figure 5.65 Dynamic responses of the Butane Isomerization plant HIP4 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

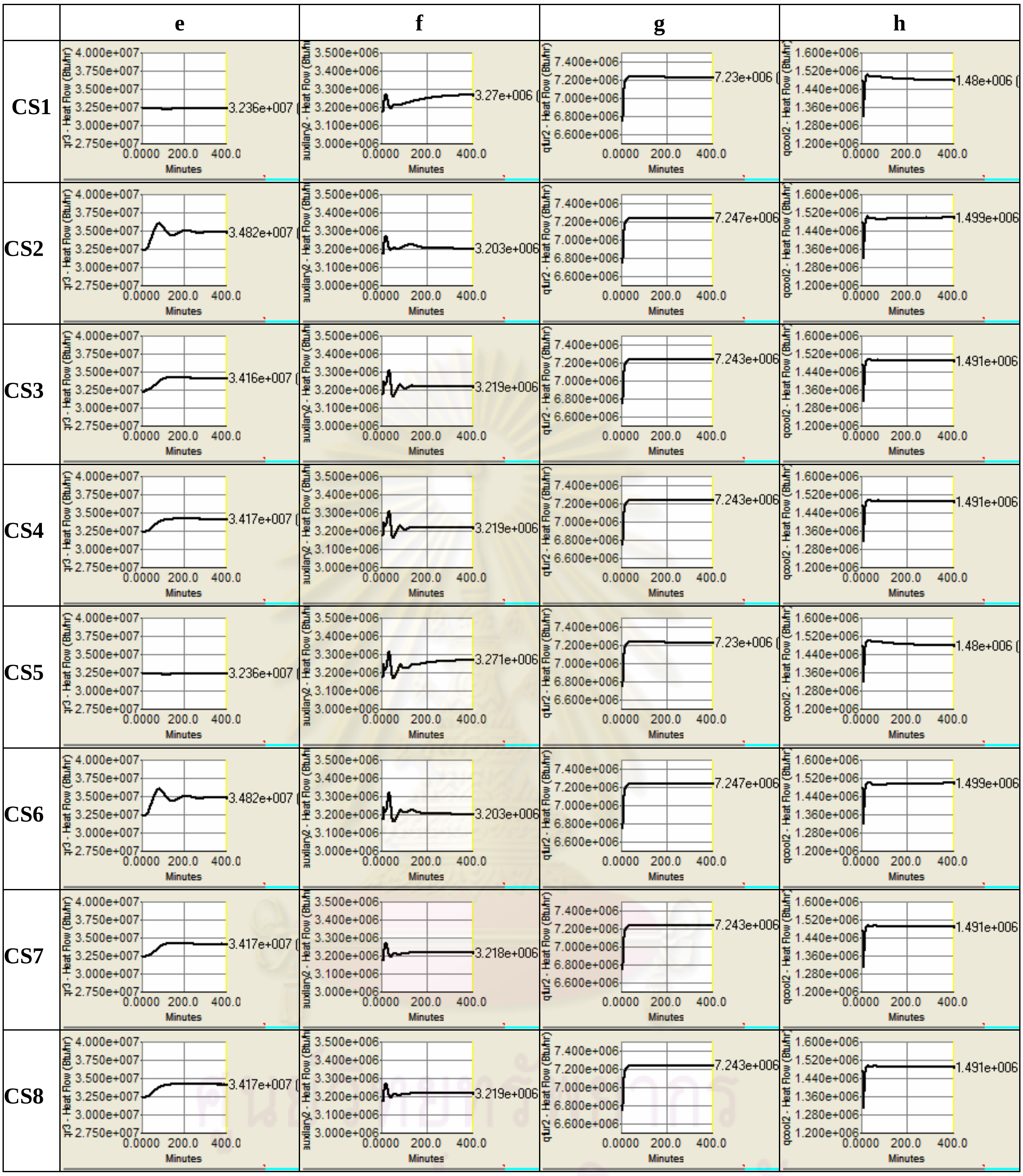


Figure 5.65 Continued Dynamic responses of the Butane Isomerization plant HIP4 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

5.4.8 Change in the recycle flowrates for CS1 to CS8 in the new designed heat integration plants: HIP5, HIP6 and HIP7

Figures 5.98-5.100 show dynamic responses of HIP5-7 when the recycle flowrate is changed from 870.1 to 940.1 lb-mol/hr occurring at time equals 10 minutes. The results of CS1-8 show the same as HIP1-4 that is the product decreases from 550 to 535 lb-mol/hr for CS1 and CS5 with network structure HIP5 while the other control structures are increases, i.e., CS2 and CS6 increases from 550 to 563.6 lb-mol/hr, CS3, CS4, CS7 and CS8 increases from 548 to 555.7 lb-mol/hr.

The energy consumption of DIB column reboiler decreases as the product decreases with CS1 and CS5 while the other control structures are increases. The energy consumption of Purge column reboiler, funace and cooler increases as the recycle flowrate increases.

Similar conclusion can be drawn for heat integrated structure HIP6 and HIP7 with control structures CS1-CS8. (see Figure 5.99-5.100).

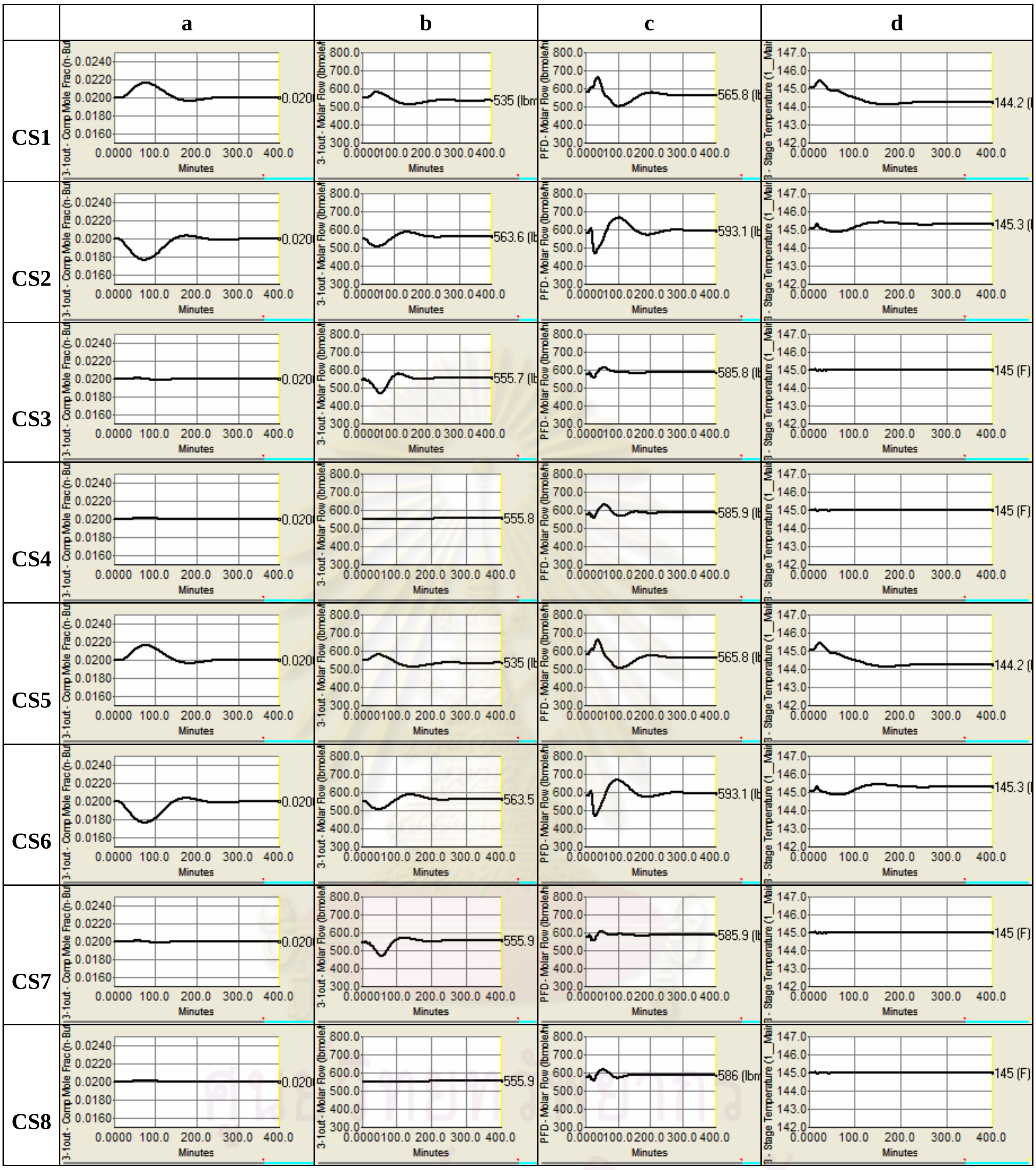


Figure 5.66 Dynamic responses of the Butane Isomerization plant HIP5 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

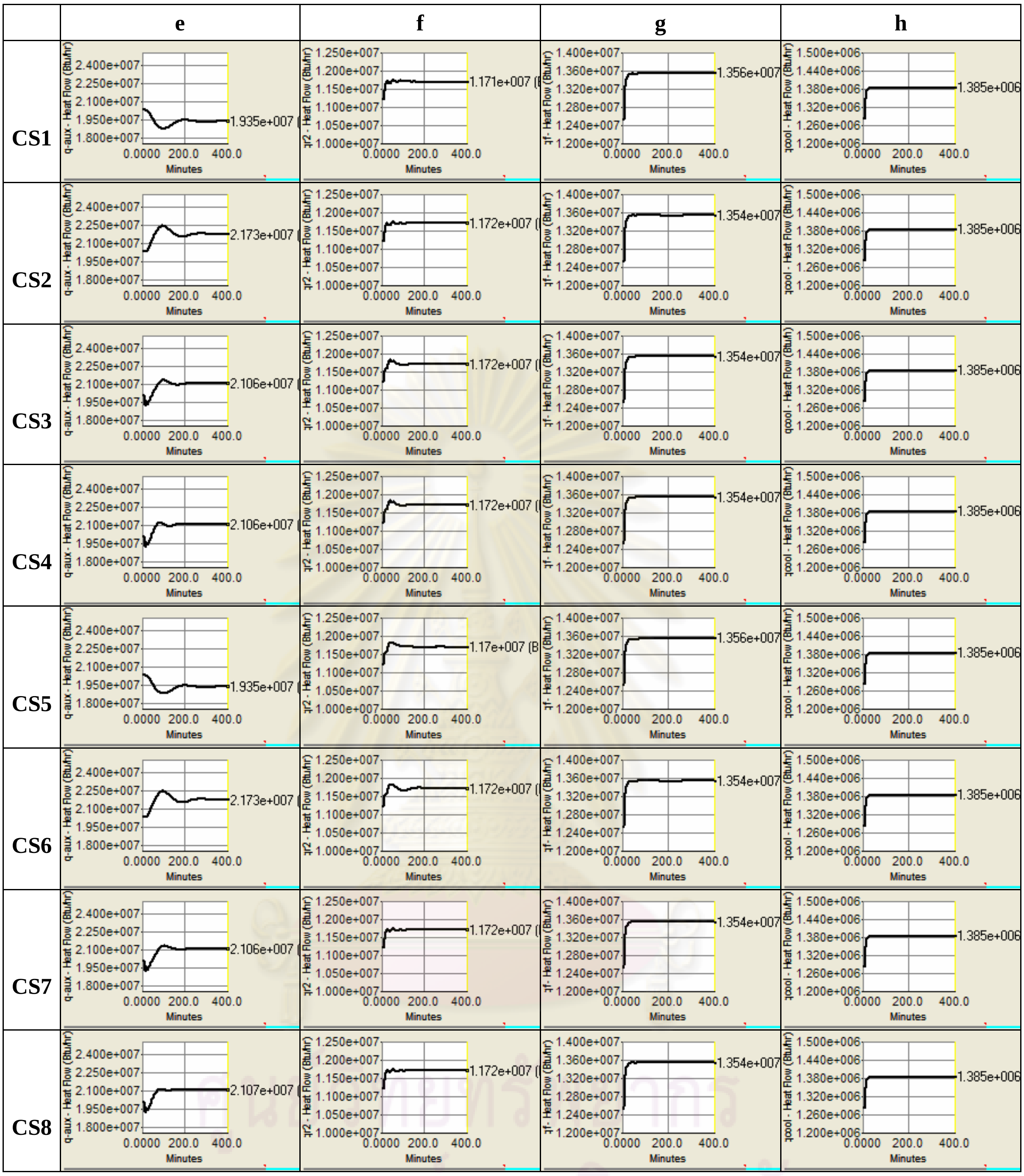


Figure 5.66 Continued Dynamic responses of the Butane Isomerization plant HIP5 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

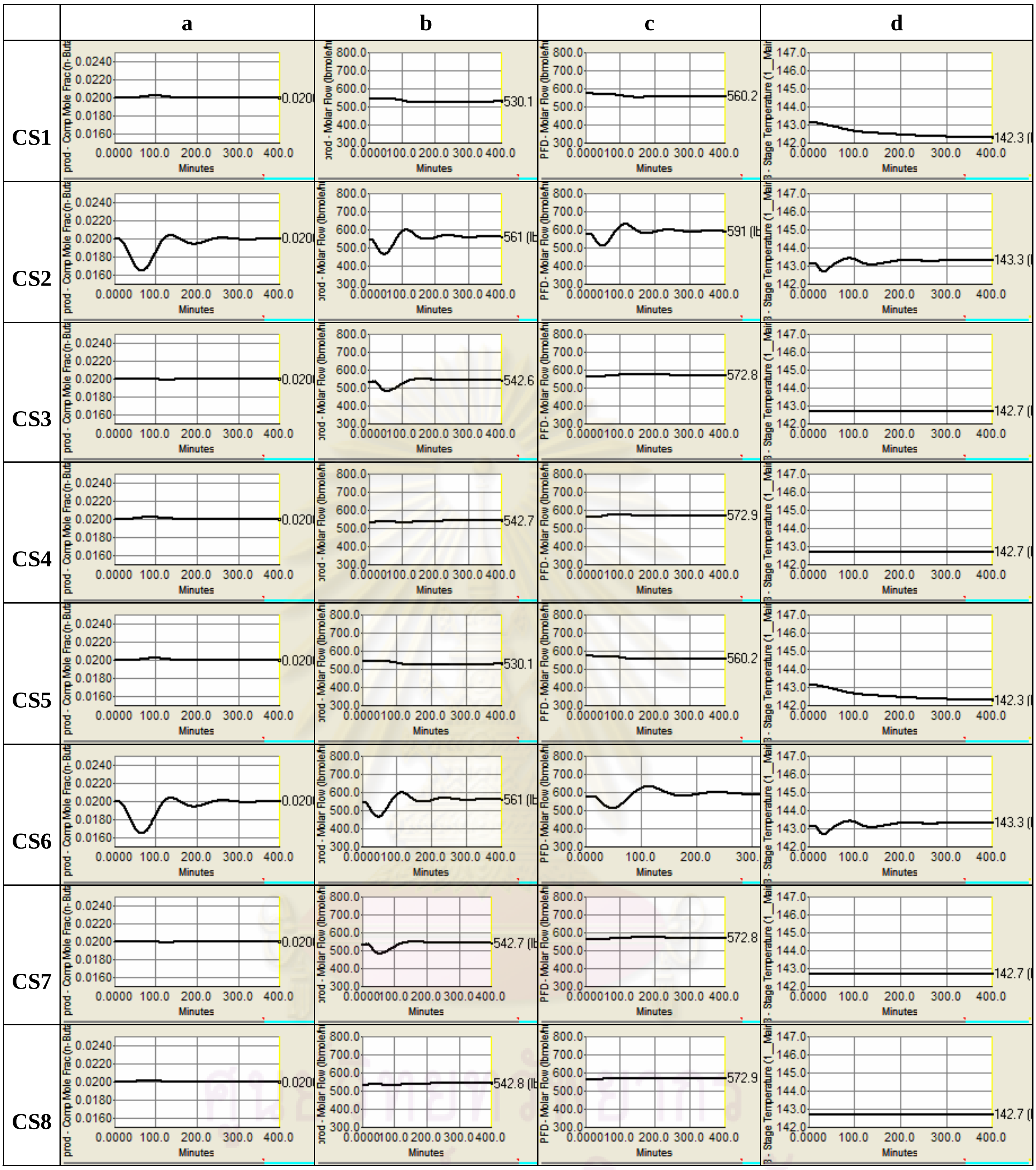


Figure 5.67 Dynamic responses of the Butane Isomerization plant HIP6 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

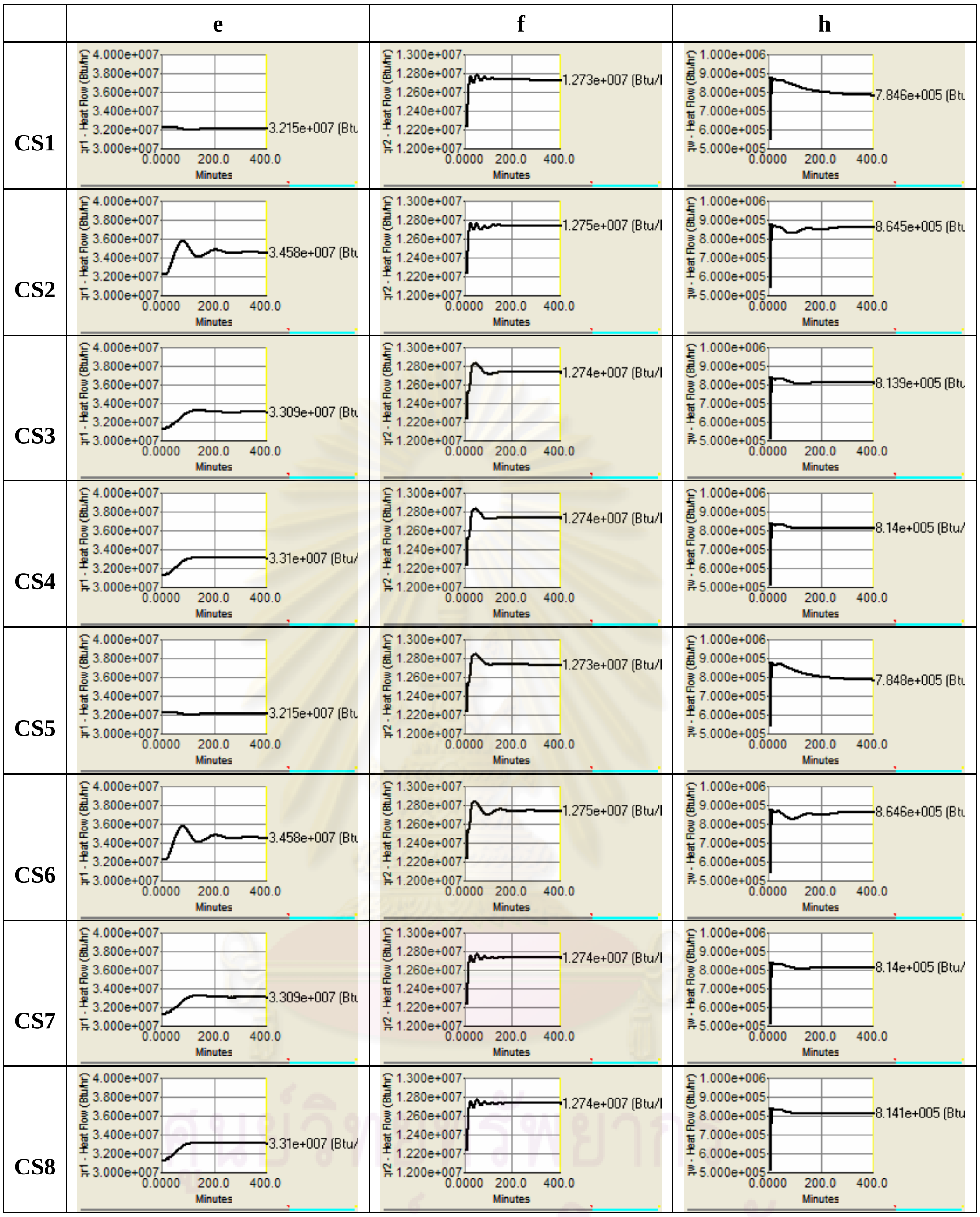


Figure 5.67 Continued Dynamic responses of the Butane Isomerization plant HIP6 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty, and (h) cooler duty.

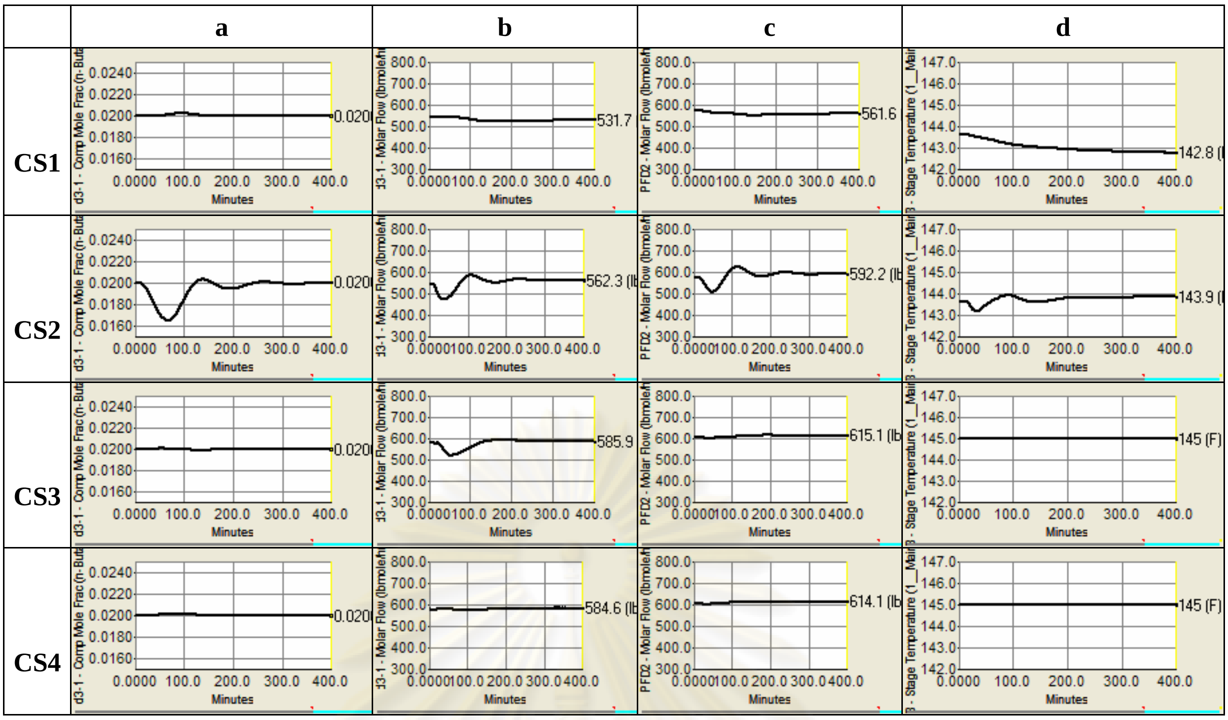


Figure 5.68 Dynamic responses of the Butane Isomerization plant HIP7 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8. , where (a) product composition, (b) product flowrate, (c) fresh feed flowrate and (d) tray 1 temperature of DIB column.

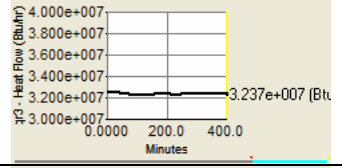
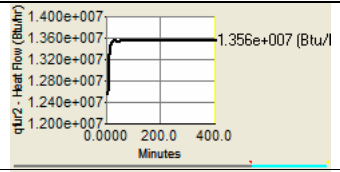
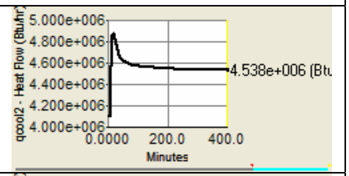
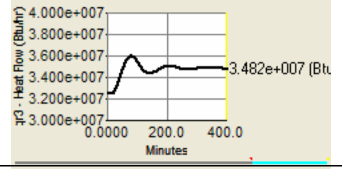
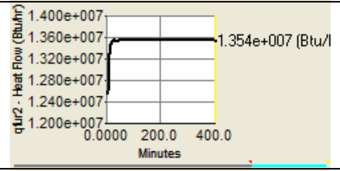
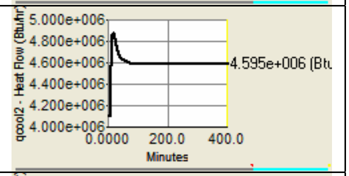
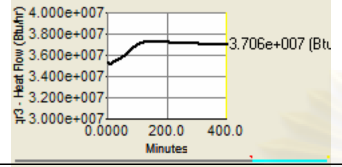
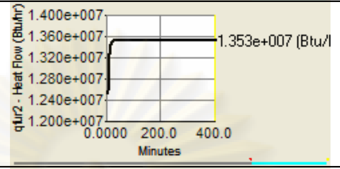
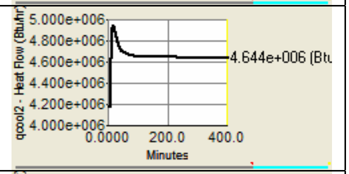
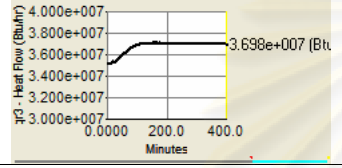
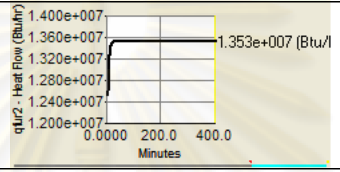
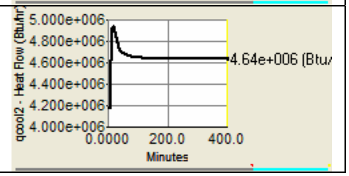
	e	g	h
CS1			
CS2			
CS3			
CS4			

Figure 5.68 Continued Dynamic responses of the Butane Isomerization plant HIP7 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (e) reboiler duty of DIB column, (f) reboiler duty of Purge column and (h) cooler duty.

5.4.9 Comparison of control structures for change in the recycle flowrates: CS1, CS2, CS4, CS4, CS5, CS6, CS7 and CS8 with HIP1-7

From Fig. 5.94-5.100 show the effect of increasing the recycle flowrate from 870 to 940.1 lb-mol/h at time equals 10 minutes of HIP1-7 with CS1-8. As discussed before the effect is a slight decrease in production rate with CS1 and CS5. CS2 and CS6 with the ratio controller are used to adjust reflux as the feed changes, product increases. CS3, CS4, CS7 and CS8 with base temperature controlled give the similar increase in product rate as CS2 and CS6 do, without the ratio controller! (see Table 5.10). The base temperature control holds the impurity of the bottoms, thus there is an increase *i*C_4 in the distillate. To some degree the fractionation capability of the column increases as the load is increased. We can conclude that control structures

CS3, CS4, CS7 and CS8 control the top composition and Tray 1 temperature better than CS1, CS2, CS5 and CS6, and consequently resulting in keep the other variables such as product flowrate, heat duties at good values, see Table 5.10. The dynamic responses of control structures CS4 and CS8 are better than CS3 and CS7 in this case because the material disturbance entered is immediately directed out of the DIB column through distillate. The reflux flow is keep at value required set by the composition set point.

Note that similar DIB column control structures of CS1, CS2, CS3 and CS4 with CS5, CS6, CS7 and CS8 respectively, their responses give the same results.

It is interesting that CS3, CS4, CS7 and CS8 control better than CS1 and CS5 for process with heat integration at DIB column, i.e., HIP1, HIP2 and HIP5, while they have worse control with heat integration at Purge column. It means that the temperature column control is necessary for structure with heat integration at column to reduce complication from disturbed stream.

The integral absolute errors of important control loops are shown in Table 5.8. The total IAEs for HIP1 with CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8 are 3.63, 7.25, 3.14, 2.77, 3.47, 7.77, 3.04 and 2.71 respectivery. Figures 5.101-5.108 show the responses of all HIP-CS structures. Table 5.8 and Figures 5.109-5.110 show the IAEs. Control structures CS5 and CS1with HIP3 are the best structures because the material disturbance is rejected out of the material loop. Control structures CS4 and CS8 are better than CS2, CS3, CS6 and CS7 when the plant disturbance is material recycle.

5.4.10 Comparison of heat integrated structures for change in the recycle flowrates: HIP1, HIP2, HIP3, HIP4, HIP5, HIP6 and HIP7 with CS1-8

Figures 5.101-5.108 compare the performance of heat integration structure HIP1-7. HIP1 and HIP3, with CS1, direct the thermal disturbance 598,108.21 and 592,360.74 Btu/hr. to the furnace respectively. The DIB reboiler duty changes are -457,974.51 and -43,323.96 Btu/hr. HIP2 and HIP4 shift the induced thermal disturbance to the DIB and Purge auxiliary reboiler respectively and their change in duties are -407,950.93 and 90,906.46 Btu/hr. HIP5 with CS1shift the thermal disturbance to the DIB auxiliary reboiler and its change in duties is -928,215.45 Btu/hr. Because the furnace is not used for HIP6 so the thermal disturbance is propagated to cooler. The change in duties of HIP6 with CS1 is -89,864.01Btu/hr. The Purge column reboiler is not used for HIP7 therefore the thermal disturbance is shifted to the cooler as good as HIP6. The change in duties of HIP7 with CS1 is 431,648.14 Btu/hr. The DIB reboiler decreases because the production decreases for all heat integration structures with CS1. However for control structure CS2-4, the DIB reboiler duty increases because the production increases. The DIB column requires more heat than the amount shifted to put out more boilup, so the positive figure. Table 5.9 and Figures 5.111-5.113 show the increases of energy usage when the recycle flowrate increases. Integrated structure HIP6, the large FEHE is placed before furnace, with control structure CS5 has the least energy consumption (CS5 decreases the production rate). HIP6-CS1, HIP4-CS1, HIP4-CS5, HIP2-CS5, HIP2-CS1, HIP5-CS5 and HIP5-CS1 are the next best (see Figure 5.111).

It is interesting to note that control structure CS5 is less energy consumption than CS1due to the Purge column temperature is not controlled.

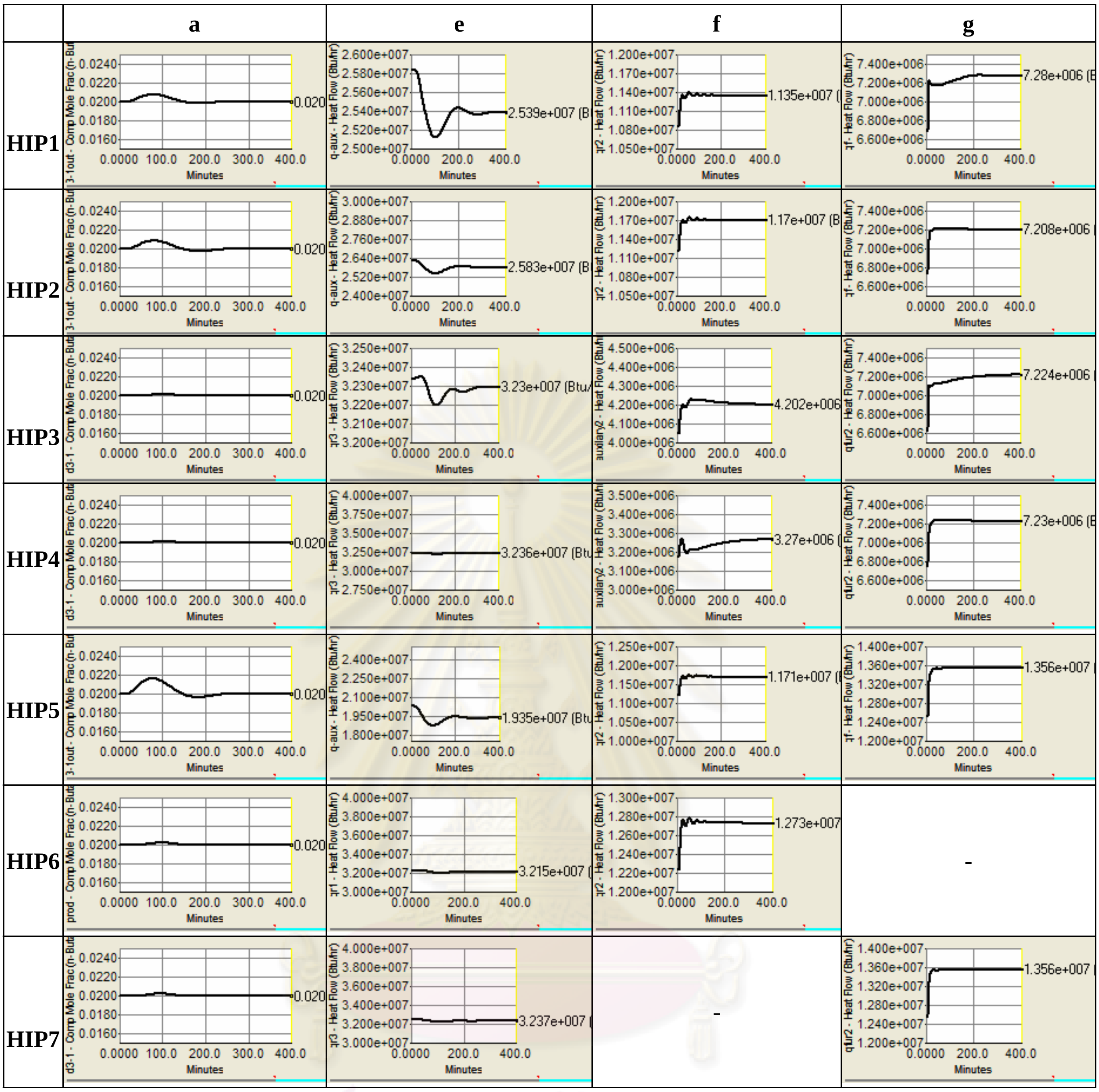


Figure 5.69 Dynamic responses of the Butane Isomerization plant CS 1 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

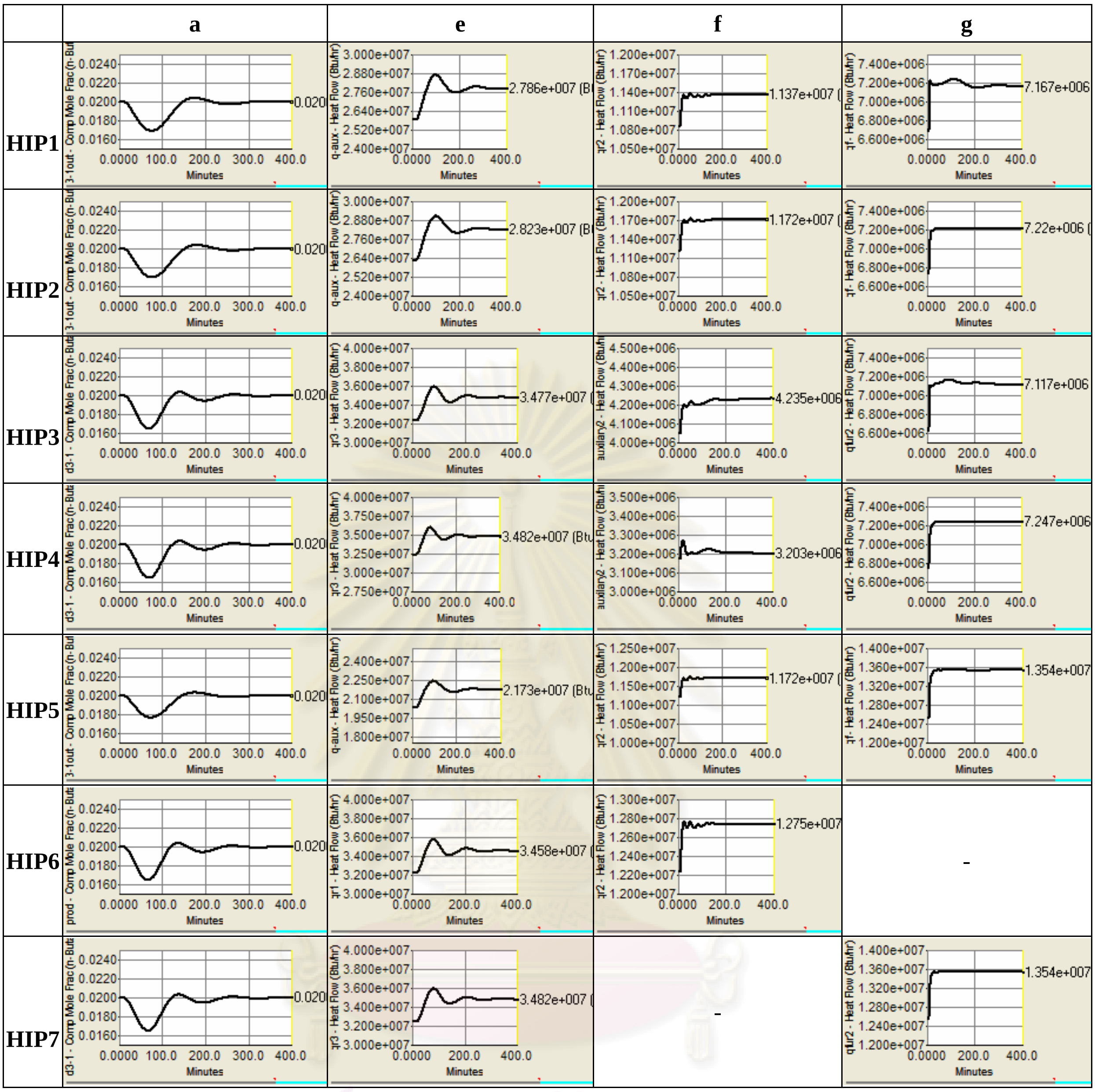


Figure 5.70 Dynamic responses of the Butane Isomerization plant CS2 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

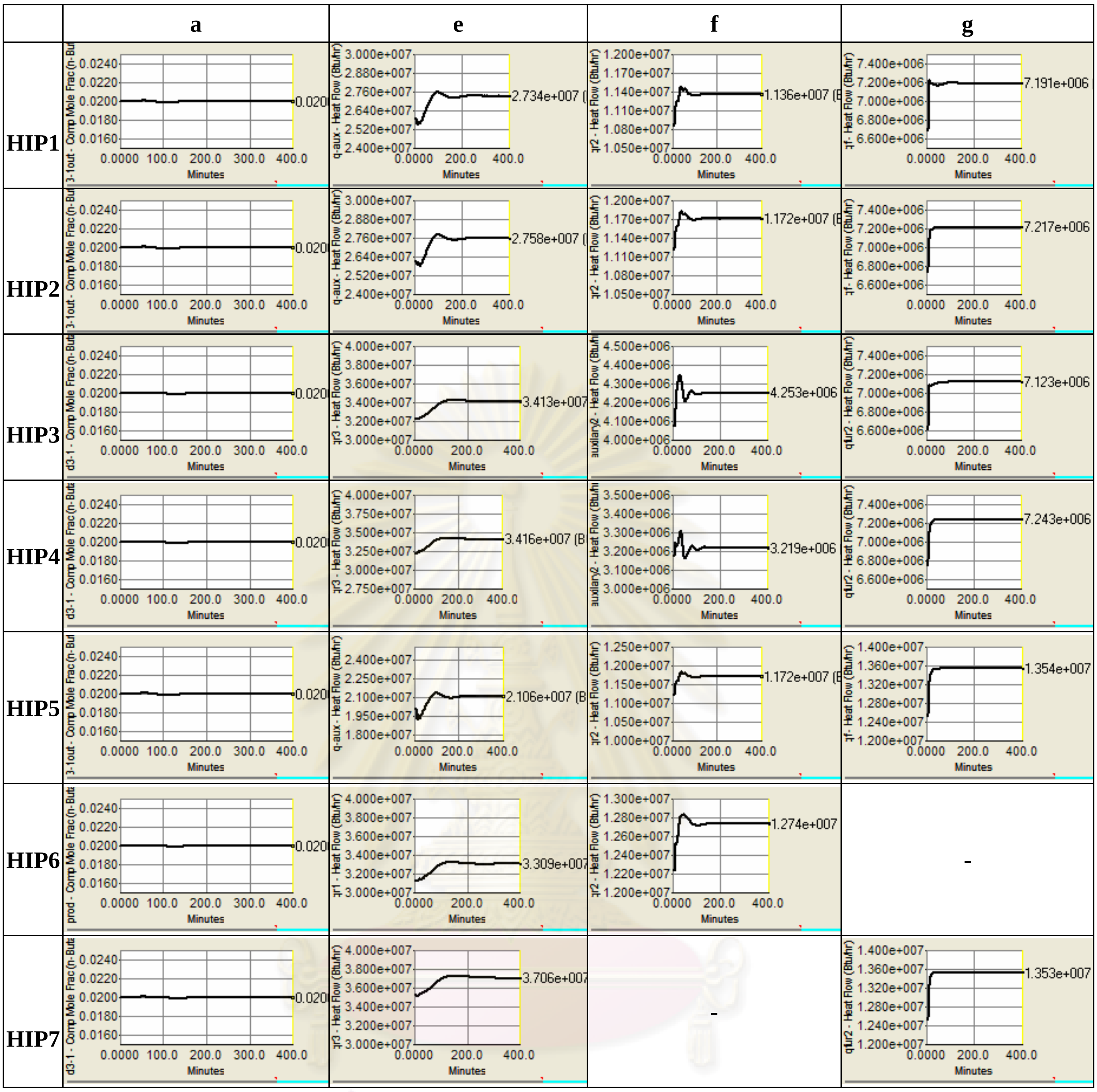


Figure 5.71 Dynamic responses of the Butane Isomerization plant CS3 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

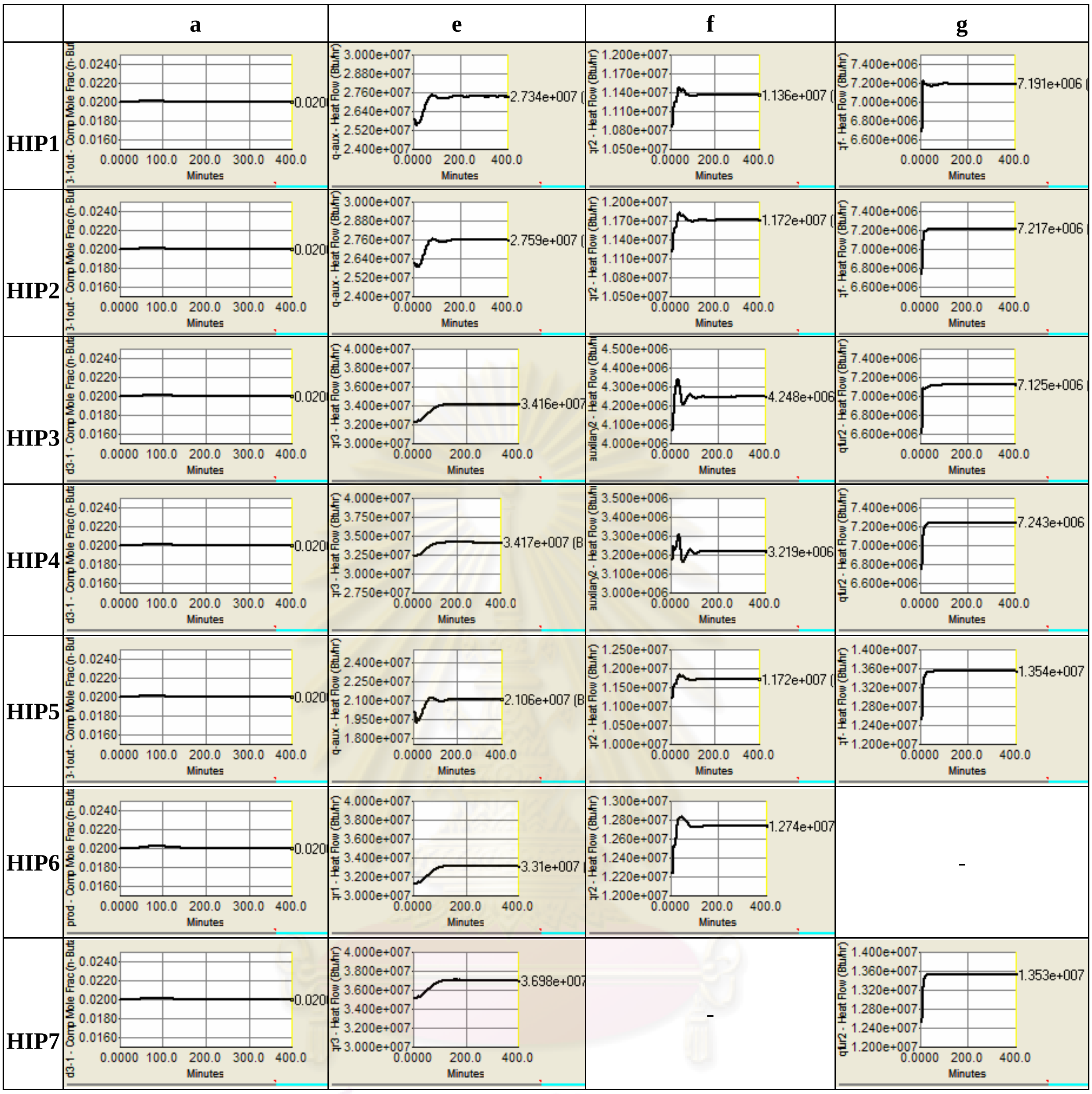


Figure 5.72 Dynamic responses of the Butane Isomerization plant CS4 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

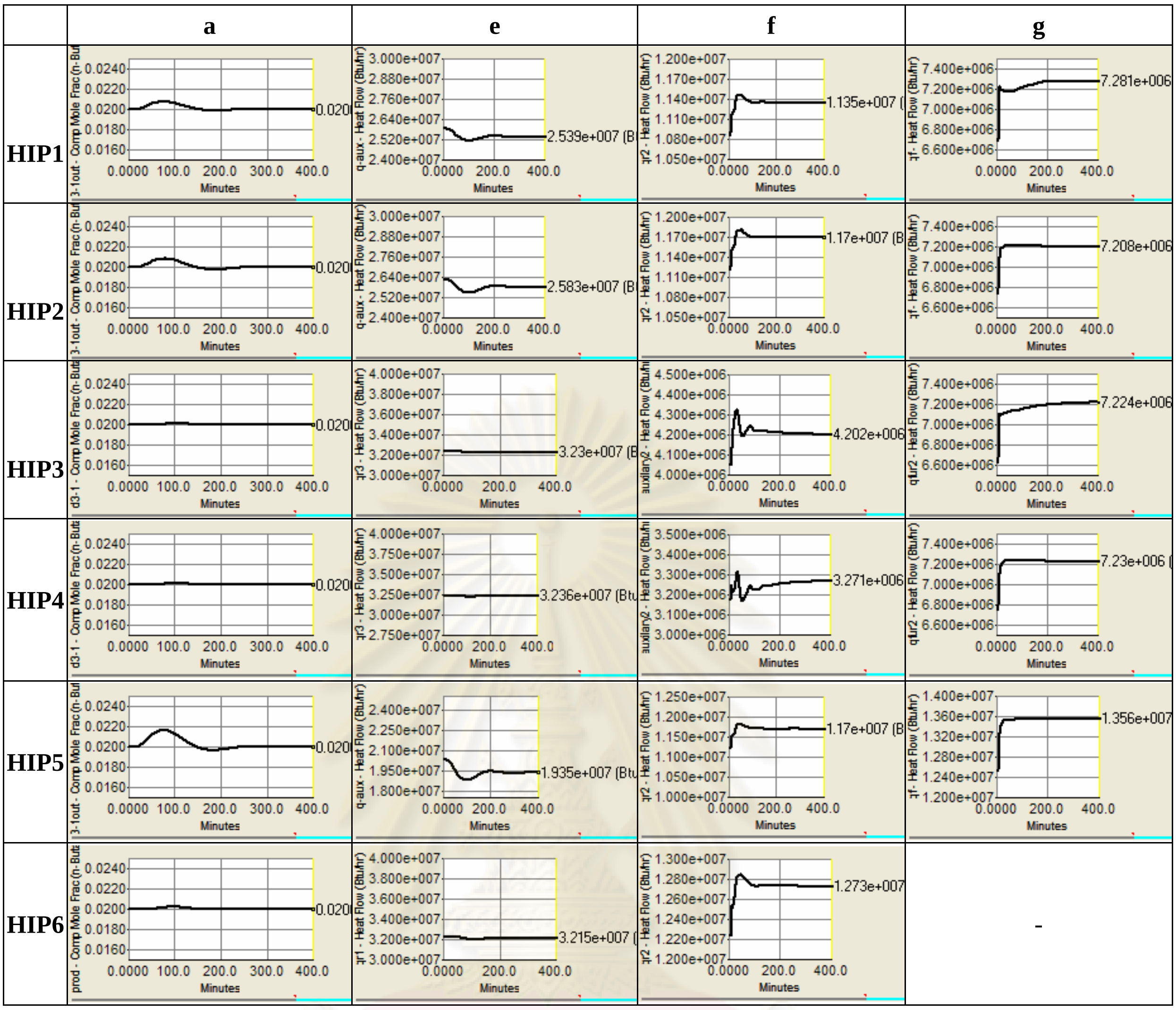


Figure 5.73 Dynamic responses of the Butane Isomerization plant CS5 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

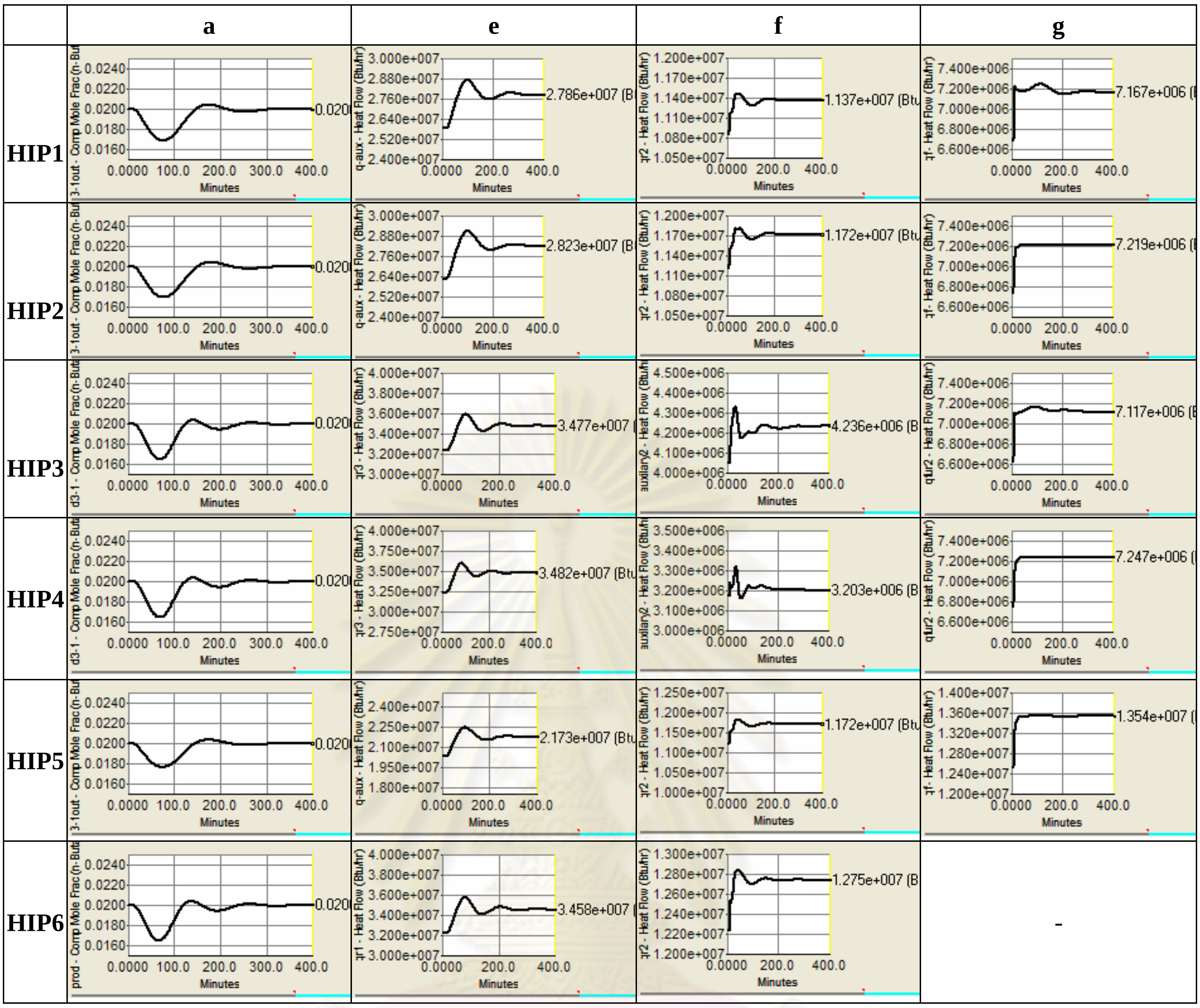


Figure 5.74 Dynamic responses of the Butane Isomerization plant CS6 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

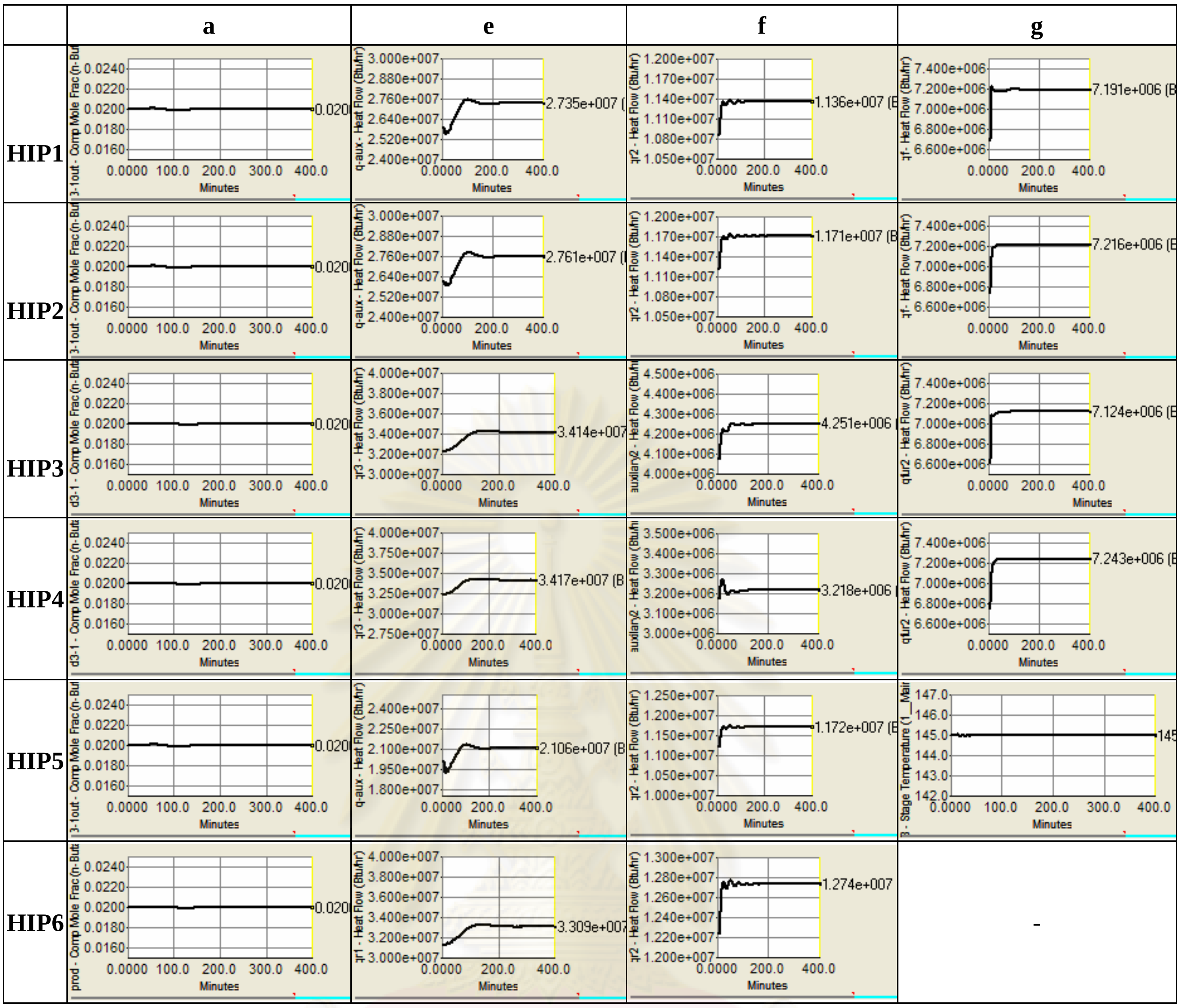


Figure 5.75 Dynamic responses of the Butane Isomerization plant CS7 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison between CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

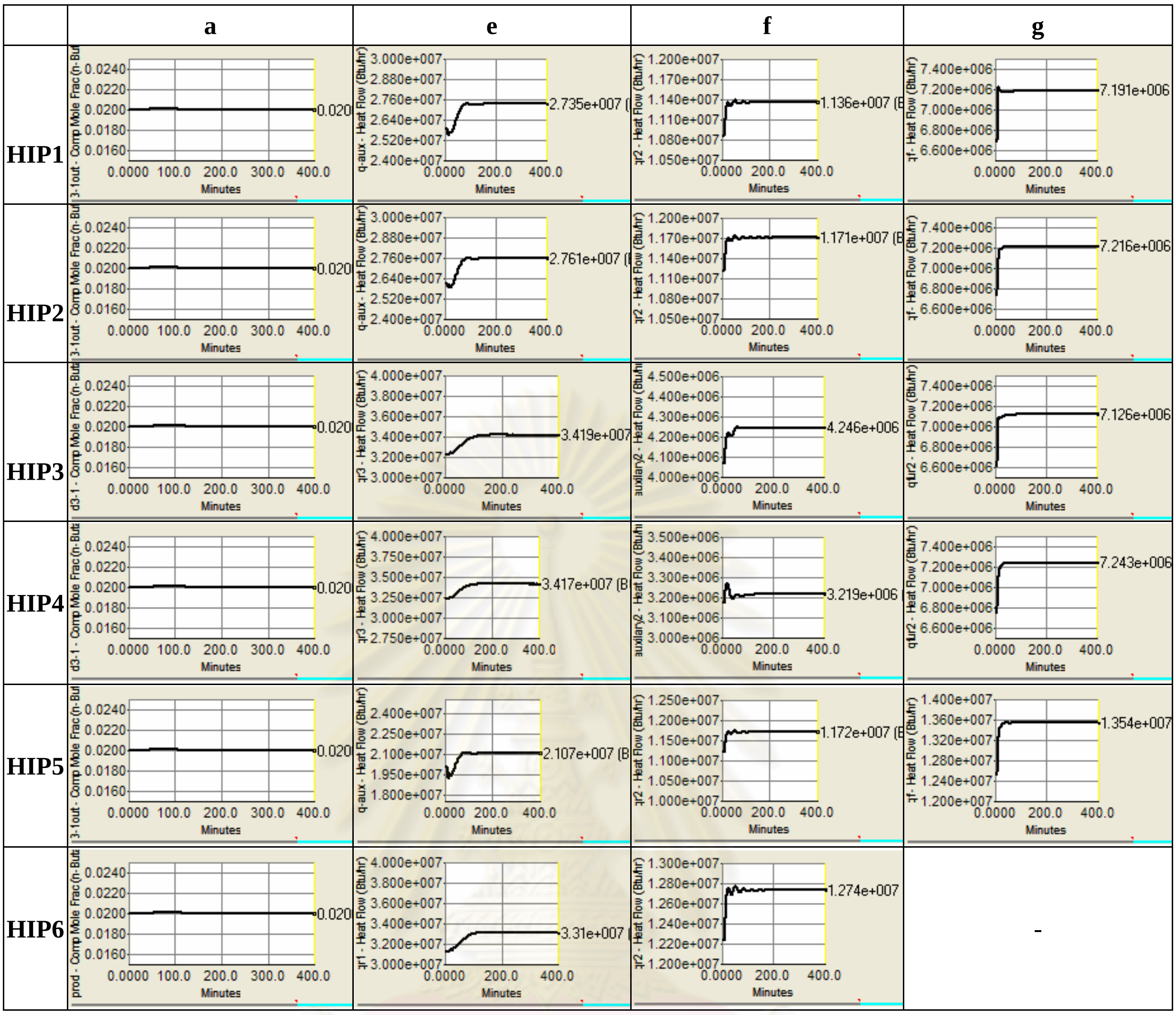


Figure 5.76 Dynamic responses of the Butane Isomerization plant CS8 to increase in recycle flow from 870 to 940.1 lb-mol/h; comparison CS1, CS2, CS3, CS4, CS5, CS6, CS7 and CS8, where (a) product composition, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, and (g) furnace duty.

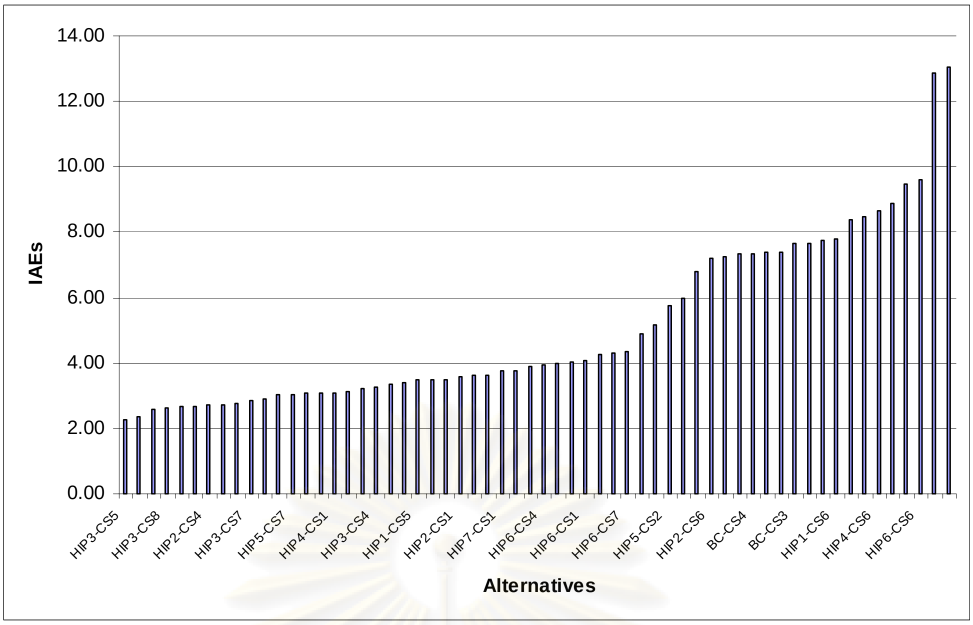


Figure 5.77 The IAE results of the control systems to increase in recycle flowrates from 870.1 to 940.1lb-mol/h.

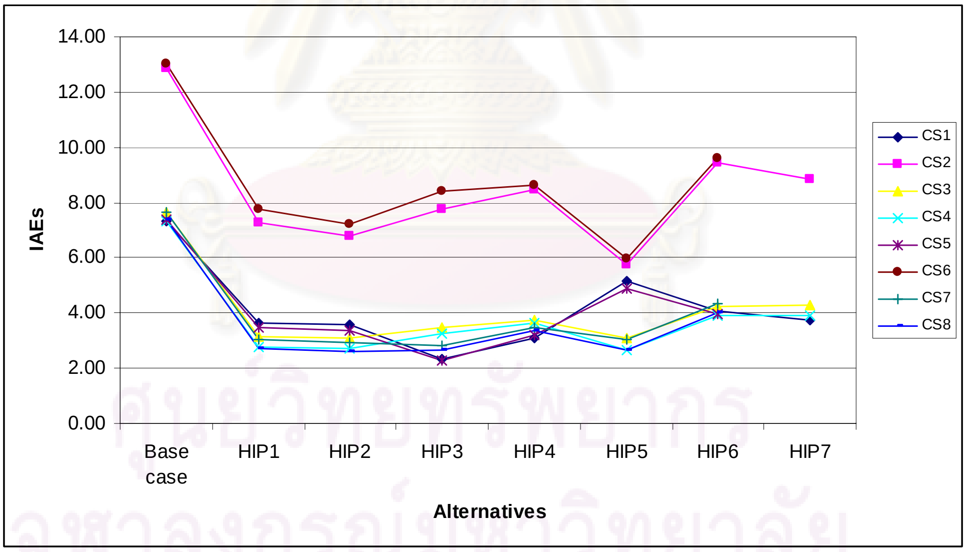


Figure 5.78 The IAE results of the control systems to increase in recycle flowrates from 870.1 to 940.1lb-mol/h.

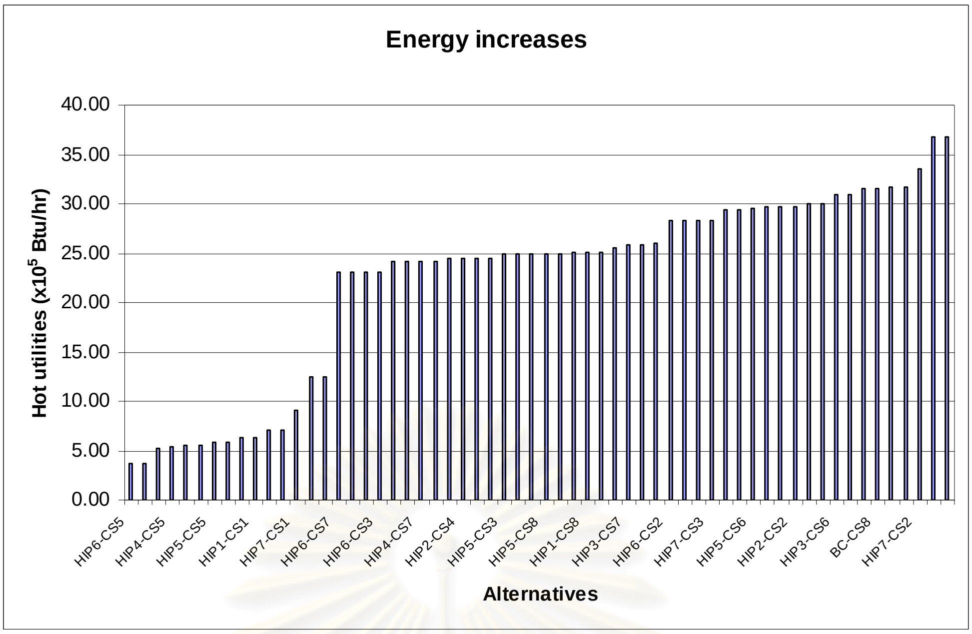


Figure 5.79 Energy increases as the recycle flowrates increase from 870 to 940.1lb-mol/h.

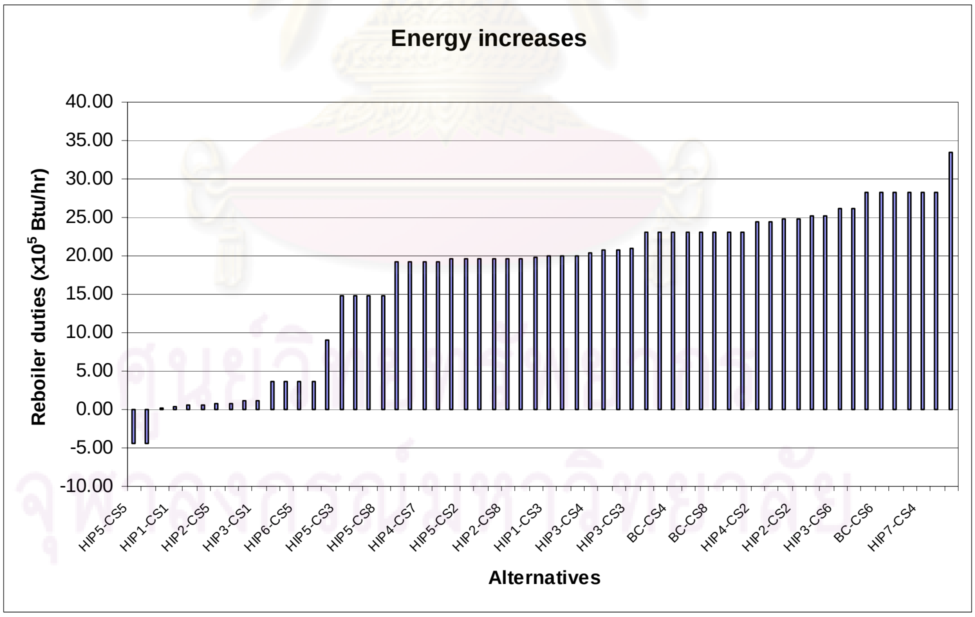


Figure 5.80 Reboiler duty increases as the recycle flowrates increase from 870 to 940.1lb-mol/h.

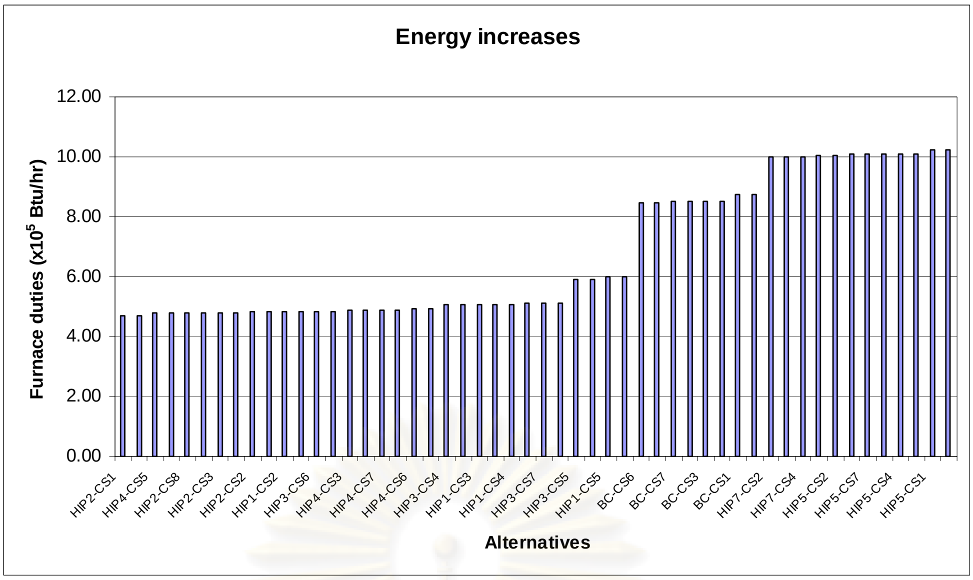


Figure 5.81 Furnace duty increases as the recycle flowrates increase from 870 to 940.1lb-mol.

Table 5.8 The IAE results of the control systems to increase in recycle flowrates from 870.1 to 940.1lb-mol/h.

Integral Absolute Error											
		BC	**HIP1**	**HIP2**	**HIP3**	**HIP4**	**HIP5**	**HIP6**	**HIP7**	**AVG**	**SD**
CCDt	CS1	0.2400	0.8719	1.0045	0.1505	0.1617	1.8569	0.2438	0.2357	0.5956	0.5702
	CS2	3.1231	3.3880	3.3464	3.1716	3.1716	2.5124	3.1181	3.0627	3.1117	0.2501
	CS3	0.1169	0.1580	0.1597	0.1231	0.1306	0.1692	0.1169	0.1394	0.1392	0.0193
	CS4	0.1928	0.1978	0.2015	0.1915	0.1928	0.2114	0.1688	0.2105	0.1959	0.0126
	CS5	0.2265	0.8419	0.9714	0.1474	0.1609	1.8184	0.2322	-	0.6284	0.5802
	CS6	3.1308	3.4262	3.3803	3.1856	3.1748	2.5461	3.1270	-	3.1387	1.0675
	CS7	0.1159	0.1468	0.1494	0.1205	0.1296	0.1570	0.1159	-	0.1336	0.0466
	CS8	0.1938	0.1949	0.2013	0.1959	0.1946	0.2053	0.1698	-	0.1937	0.0648
TCR	CS1	2.4659	1.0238	0.6808	0.8966	0.6951	0.7735	0.8090	0.7720	1.0146	0.5582
	CS2	2.4664	1.1376	0.7060	0.9078	0.6987	0.7724	0.8097	0.7746	1.0341	0.5573
	CS3	2.4198	0.9393	0.6799	0.7854	0.6933	0.7646	0.7811	0.7696	0.9791	0.5494
	CS4	2.4149	0.9289	0.6804	0.7816	0.6933	0.7640	0.7792	0.7696	0.9765	0.5482
	CS5	2.4664	1.0303	0.6818	0.8988	0.6954	0.7598	0.8108	-	1.0490	0.6514
	CS6	2.4701	1.1683	0.7136	0.9091	0.6997	0.7659	0.8103	-	1.0767	0.6552
	CS7	2.4218	0.9228	0.6787	0.7853	0.6931	0.7703	0.7817	-	1.0077	0.6383
	CS8	2.4192	0.9086	0.6787	0.7820	0.6928	0.7703	0.7795	-	1.0044	0.6375
TCC	CS1	3.6140	0.0600	0.1868	0.2188	0.6661	0.1531	1.8399	-	0.9627	1.1897
	CS2	3.5924	0.0608	0.1887	0.2193	0.6743	0.1534	1.8398	2.0230	1.0940	1.1924
	CS3	3.4855	0.0601	0.1895	0.2201	0.6553	0.1536	1.7359	2.0225	1.0653	1.1573
	CS4	3.4594	0.0603	0.1866	0.2198	0.6523	0.1537	1.7133	2.0168	1.0577	1.1487
	CS5	3.6294	0.0601	0.1891	0.2198	0.6588	0.1532	1.8693	-	0.9685	1.1972
	CS6	3.6229	0.0612	0.1948	0.2208	0.6585	0.1536	1.8683	-	0.9686	1.1947
	CS7	3.4634	0.0595	0.1852	0.2189	0.6558	0.1536	1.7146	-	0.9216	1.1332
	CS8	3.4476	0.0597	0.1843	0.2189	0.6553	0.1538	1.7008	-	0.9172	1.1272

Note CC = Composition Control, TC = Temperature Control, PC= Pressure Control, R = Reactor, D = DIB, P = Purge, C = Cooler, t = top, b = bottom

Table 5.8 Continued The IAE results of the control systems to increase in recycle flowrates from 870.1 to 940.1lb-mol/h.

		Integral Absolute Error									
		BC	HIP1	HIP2	HIP3	HIP4	HIP5	HIP6	HIP7	AVG	SD
PCD	CS1	0.3622	0.6261	0.6710	0.1568	0.1723	1.2867	0.2538	0.2345	0.4704	0.3592
	CS2	2.7922	1.6141	1.5264	2.5039	2.5466	1.2082	2.7800	2.4903	2.1827	0.5876
	CS3	0.7967	0.9910	1.0091	0.7949	0.7357	1.0283	0.7836	0.9554	0.8868	0.1121
	CS4	0.4593	0.6075	0.6122	0.5429	0.5305	0.6034	0.4523	0.4949	0.5379	0.0613
	CS5	0.2481	0.6079	0.6571	0.6093	0.1671	1.2720	0.2461	-	0.5439	0.3756
	CS6	2.7895	1.6611	1.5579	2.5238	2.5485	1.2347	2.7793	-	2.1564	0.9082
	CS7	0.7885	0.9070	0.9338	0.7947	0.7349	0.9416	0.7742	-	0.8392	0.2872
	CS8	0.4607	0.5417	0.5565	0.5362	0.5316	0.5198	0.4539	-	0.5143	0.1737
PCP	CS1	0.6425	1.0486	1.0349	0.9177	1.3972	1.1012	0.9335	0.4265	0.9378	0.2758
	CS2	0.8711	1.0539	1.0373	0.9640	1.3619	1.1057	0.9008	0.5094	0.9755	0.2264
	CS3	0.8213	0.9894	1.0364	1.5556	1.5556	0.9601	0.8251	0.4084	1.0190	0.3588
	CS4	0.8158	0.9737	1.0181	1.5460	1.5470	0.9509	0.8184	0.3999	1.0087	0.3583
	CS5	0.8154	0.9282	0.8882	0.3875	1.5210	0.8743	0.8136	-	0.8897	0.4118
	CS6	1.0168	1.4563	1.3617	1.5465	1.5644	1.2630	1.0133	-	1.3174	0.4795
	CS7	0.8814	1.0069	0.9727	0.9177	1.2786	1.0265	0.9457	-	1.0042	0.3509
	CS8	0.8805	1.0065	0.9675	0.9081	1.2729	1.0233	0.9335	-	0.9989	0.3493
Total	CS1	7.3247	3.6304	3.5781	2.3404	3.0924	5.1715	4.0799	3.7627	4.1225	1.4257
	CS2	12.8451	7.2543	6.8048	7.7665	8.4531	5.7521	9.4484	8.8598	8.3980	2.0107
	CS3	7.6402	3.1377	3.0745	3.4792	3.7704	3.0758	4.2427	4.2952	4.0895	1.4191
	CS4	7.3422	2.7681	2.6988	3.2818	3.6158	2.6834	3.9320	3.8917	3.7767	1.4316
	CS5	7.3858	3.4684	3.3876	2.2628	3.2032	4.8777	3.9719	-	4.0796	1.9705
	CS6	13.0301	7.7731	7.2083	8.3858	8.6459	5.9633	9.5982	-	8.6578	3.4606
	CS7	7.6710	3.0430	2.9198	2.8370	3.4920	3.0491	4.3322	-	3.9063	1.9835
	CS8	7.4018	2.7115	2.5883	2.6411	3.3473	2.6724	4.0375	-	3.6286	1.9306
TCDb	CS1	-	-	-	-	-	-	-	-	-	-
	CS2	-	-	-	-	-	-	-	-	-	-
	CS3	0.4380	0.7323	0.6993	0.4136	0.4017	1.2017	0.4345	0.4584	0.5974	0.2591
	CS4	0.3819	0.6644	0.6329	0.3711	0.3581	1.0991	0.4238	0.3909	0.5403	0.2395
	CS5	-	-	-	-	-	-	-	-	-	-
	CS6	-	-	-	-	-	-	-	-	-	-
	CS7	0.4273	0.6866	0.6653	0.4163	0.4081	1.1434	0.4372	-	0.5977	0.3055
	CS8	0.3852	0.6077	0.5767	0.3675	0.3622	0.9965	0.4110	-	0.5295	0.2648
TCPb	CS1	-	-	-	-	-	-	-	26.0094	26.0094	8.6018
	CS2	-	-	-	-	-	-	-	25.0817	25.0817	8.2950
	CS3	2.7782	3.2872	3.0529	2.5443	2.5123	3.2018	0.7026	25.4746	5.4442	7.6093
	CS4	2.7609	3.2685	3.0371	2.5305	2.4882	3.1895	0.6995	25.3366	5.4138	7.5684
	CS5	1.5472	1.7009	1.6168	2.5851	2.7273	1.6605	0.6983	-	1.7908	0.8383
	CS6	1.8091	2.1368	2.0015	2.8545	2.8866	1.9656	0.8070	-	2.0659	0.9151
	CS7	-	-	-	-	-	-	-	-	-	-
	CS8	-	-	-	-	-	-	-	-	-	-

Note CC = Composition Control, TC = Temperature Control, PC= Pressure Control, R = Reactor, D = DIB, P = Purge, C = Cooler, t = top, b = bottom

Table 5.9: Energy increases as the recycle flowrates increase from 870 to 940.1lb-mol/h.

Energy Increase (10^5, Btu/ hr)											
		BC	**HIP1**	**HIP2**	**HIP3**	**HIP4**	**HIP5**	**HIP6**	**HIP7**	**AVG**	**SD**
Furnace	CS1	8.7471	5.9811	4.7021	5.9236	4.7748	10.2108	-	10.1055	7.2064	3.1737
	CS2	8.4880	4.8377	4.8204	4.8556	4.9437	10.0643	-	9.9868	6.8566	3.1577
	CS3	8.5295	5.0895	4.8019	5.1037	4.8974	10.0945	-	10.0029	6.9313	3.1511
	CS4	8.5296	5.0923	4.8019	5.0878	4.8976	10.0946	-	10.0033	6.9296	3.1517
	CS5	8.7472	5.9781	4.7026	5.9197	4.7737	10.2111	-	-	6.7221	3.4118
	CS6	8.4878	4.8402	4.8215	4.8517	4.9434	10.0641	-	-	6.3348	3.3052
	CS7	8.5294	5.0897	4.8019	5.1000	4.8980	10.0941	-	-	6.4188	3.3203
	CS8	8.5295	5.0894	4.8018	5.0937	4.8980	10.0939	-	-	6.4177	3.3202
DIB column reboiler	CS1	-1.1871	-4.5797	-4.0795	-0.4332	-0.3979	-9.2822	-1.1429	-0.9870	-2.7612	2.8900
	CS2	23.1846	20.1350	19.8351	24.2739	24.2090	14.5037	23.2056	23.5348	21.6102	3.1306
	CS3	18.0950	14.9076	14.5675	19.1099	18.8876	9.7464	18.0441	18.3325	16.4613	3.0139
	CS4	18.0306	14.8721	14.5881	18.6333	18.8282	9.7785	18.0263	18.3177	16.3844	2.9437
	CS5	-1.1733	-4.6083	-4.1023	-0.4445	-0.3839	-9.2836	-1.1582	-	-3.0220	2.9870
	CS6	23.1854	20.1799	19.8009	24.2365	24.2188	14.5071	23.2001	-	21.3327	7.6823
	CS7	18.0946	14.9703	14.6503	18.9961	18.9176	9.7767	18.0081	-	16.2020	6.0923
	CS8	18.0551	14.9682	14.6398	19.0202	18.9193	9.7932	18.0375	-	16.2048	6.0924
Purge column reboiler	CS1	4.8844	4.8948	4.8929	1.5275	0.9091	4.8906	4.8856	-	3.8407	2.0110
	CS2	5.0398	5.0473	5.0511	1.8571	0.2421	5.0493	5.0387	-	3.9036	2.1635
	CS3	5.0011	5.0168	5.0197	1.7634	0.4007	5.0166	5.0068	-	3.8893	2.1281
	CS4	5.0038	5.0171	5.0191	1.7730	0.4057	5.0162	5.0077	-	3.8918	2.1266
	CS5	4.8842	4.8937	4.8938	1.5265	0.9222	4.8887	4.8812	-	3.8415	2.0085
	CS6	5.0459	5.0418	5.0560	1.8630	0.2512	5.0399	5.0470	-	3.9064	2.1617
	CS7	5.0044	5.0142	5.0201	1.7596	0.4038	5.0200	5.0065	-	3.8898	2.1283
	CS8	5.0031	5.0092	5.0152	1.7665	0.4037	5.0170	5.0085	-	3.8890	2.1265
Hot utilities usage, (x105Btu/hr)	CS1	12.4444	6.2961	5.5155	7.0178	5.2860	5.8192	3.7427	9.1185	6.9050	2.5421
	CS2	36.7124	30.0200	29.7066	30.9865	29.3947	29.6172	28.2443	33.5215	31.0254	2.5909
	CS3	31.6256	25.0139	24.3892	25.9769	24.1856	24.8575	23.0509	28.3354	25.9294	2.5962
	CS4	31.5641	24.9816	24.4091	25.4942	24.1315	24.8893	23.0339	28.3211	25.8531	2.5865
	CS5	12.4581	6.2634	5.4941	7.0017	5.3120	5.8161	3.7230	-	6.5812	3.2454
	CS6	36.7191	30.0620	29.6784	30.9512	29.4135	29.6111	28.2471	-	30.6689	10.4259
	CS7	31.6283	25.0742	24.4723	25.8558	24.2193	24.8908	23.0146	-	25.5936	8.8048
	CS8	31.5877	25.0668	24.4569	25.8805	24.2210	24.9041	23.0460	-	25.5947	8.8008
Cold utilities usage, (x105Btu/hr)	CS1	7.8476	1.0292	0.0363	1.7852	0.0518	1.0338	-0.8986	4.3165	1.9002	2.6799
	CS2	8.3946	1.0319	0.2924	1.7866	0.2415	1.0372	-0.0969	4.8905	2.1972	2.7670
	CS3	8.2743	1.0314	0.2247	1.7879	0.2197	1.0364	-0.2576	4.7801	2.1371	2.7449
	CS4	8.2741	1.0314	0.2282	1.7867	0.2206	1.0364	-0.2575	4.7063	2.1283	2.7357
	CS5	7.8473	1.0292	0.0359	1.7853	0.0513	1.0338	-0.8966	-	1.5552	2.5712
	CS6	8.3951	1.0319	0.2924	1.7866	0.2421	1.0372	-0.0962	-	1.8127	2.6419
	CS7	8.2744	1.0314	0.2298	1.7878	0.2191	1.0364	-0.2568	-	1.7603	2.6217
	CS8	8.2741	1.0313	0.2293	1.7866	0.2190	1.0364	-0.2573	-	1.7599	2.6217

Table 5.10 Results of the control systems to increase in recycle flowrates from 870.1 to 940.1lb-mol/h.

		BC	HIP1	HIP2	HIP3	HIP4	HIP5	HIP6	HIP7	AVG	SD
Product flowrate increase (lb-mol/h.)	CS1	-16.5413	-7.9356	-4.6770	-8.1617	-7.8854	-15.2129	-16.3921	-14.6228	-11.4286	4.4203
	CS2	14.5000	21.8462	26.3585	24.9952	24.1163	13.3922	14.4164	15.9727	19.4497	5.0574
	CS3	8.2937	16.2316	19.8778	18.3691	17.5042	8.0651	8.1133	7.2807	12.9669	5.1228
	CS4	8.2682	15.6780	19.6404	19.5752	17.5081	7.5759	8.1691	6.4078	12.8528	5.3997
	CS5	-16.6303	-8.1021	-4.7563	-8.2667	-7.6351	-15.1996	-16.4561	-	-11.0066	5.6045
	CS6	14.5937	21.7628	26.2250	24.8788	24.1532	13.3006	14.4838	-	19.9140	8.1731
	CS7	8.3683	16.1542	19.9490	17.8597	17.3183	8.2295	8.2122	-	13.7273	6.4111
	CS8	8.3279	16.2313	19.6836	17.6350	17.1820	8.0219	8.1977	-	13.6113	6.3566
Fresh feed flowrate increase (lb-mol/h.)	CS1	-16.4620	-7.2721	-4.0794	-8.0875	-8.5665	-14.3750	-16.3084	-14.6219	-11.2216	4.4487
	CS2	14.5456	21.3039	25.1424	25.1856	24.3655	12.7936	14.4960	16.0571	19.2362	4.9617
	CS3	8.3828	16.2541	19.5351	17.9032	17.2151	8.0165	8.1798	7.0664	12.8191	4.9930
	CS4	8.3761	16.0646	19.6567	18.4363	17.2695	8.1458	8.2366	7.1414	12.9159	5.0421
	CS5	-16.5263	-7.6384	-4.3046	-8.2566	-7.5929	-14.3927	-16.3479	-	-10.7228	5.5451
	CS6	14.6941	21.8626	25.0644	25.2870	24.3994	12.8863	14.6118	-	19.8294	8.1245
	CS7	8.3331	16.0659	19.7404	17.8506	16.9949	8.1993	8.1811	-	13.6236	6.3448
	CS8	8.3668	16.2095	19.6469	17.5752	17.0078	8.1838	8.2457	-	13.6051	6.3062
Hot utilities usage ($x10^5$, Btu/hr)	CS1	12.4444	6.2961	5.5155	7.0178	5.2860	5.8192	3.7427	9.1185	6.9050	2.5421
	CS2	36.7124	30.0200	29.7066	30.9865	29.3947	29.6172	28.2443	33.5215	31.0254	2.5909
	CS3	31.6256	25.0139	24.3892	25.9769	24.1856	24.8575	23.0509	28.3354	25.9294	2.5963
	CS4	31.5641	24.9816	24.4091	25.4942	24.1315	24.8893	23.0339	28.3211	25.8531	2.5865
	CS5	12.4581	6.2634	5.4941	7.0017	5.3120	5.8161	3.7230	-	6.5812	3.2454
	CS6	36.7191	30.0620	29.6784	30.9512	29.4135	29.6111	28.2471	-	30.6689	10.4259
	CS7	31.6283	25.0742	24.4723	25.8558	24.2193	24.8908	23.0146	-	25.5936	8.8048
	CS8	31.5877	25.0668	24.4569	25.8805	24.2210	24.9041	23.0460	-	25.5947	8.8008

CHAPTER VI

CONCLUTIONS AND RECOMMENDATIONS

6.1 Conclusion

This study considers the heat integrated process design altogether with plantwide control structure selection for reduction of energy consumption and maintaining good control performance. We look at 8 alternatives of various heat integrated processes (base case by Luyben's design, 4 complex heat integration by Kunajitpimol, and 3 new designs) and 8 plantwide control structures (2 Kunajitpimol's designs, 2 Luyben's previous designs and 4 new designs). Two kinds of disturbances are used: thermal disturbance (transformed into composition disturbance after leaving the reactor) and the material flow disturbance. The HEN design follows Wongsri's resilient HEN synthesis method (1990). The energy saved is 16.36-25.24% from the base case. The thermal load management of the resilient HEN, in and out, and to thermal sinks and sources uses Heat Pathway Heuristics (Wongsri and Hermawan, 2005). As selective controller LSS and HSS are employed to select an appropriate heat pathway through the network. In general the HPH is very useful in terms of heat load or disturbance management to achieve the highest possible dynamic MER.

Butane isomerization plant is selected to illustrate the concepts, the design procedures and the analysis is illustrated using time domain simulation-based approach through HYSYS rigorous dynamic simulator. Although heat integration process is difficult to control, but proper control structure can reduce complication for complex heat integration process control and achieve to design objectives. However, the energy usage is important to consider because the good control structure with heat integration process is less energy consumption, namely decreasing operation cost.

For the thermal disturbance, CS3, CS4, CS7 and CS8 are the proper control structure with all HIPs, their feature structure are the base DIB column temperature is controlled so they reject the thermal disturbance by keeping the base temperature constant.

For the material flow disturbance, we can conclude that the control structures CS1 and CS5 are better than other structures because of the material disturbance entered is immediately directed out of the DIB column through distillate. The reflux flow is keep at value required set by the composition set point. However, the reactor inlet composition to have a lower nC_4 concentration, this effect is decrease production rate. Therefore HIP3, HIP4, HIP6 and HIP7 with control structures CS1 and CS5 result dynamic responses from material flow disturbance test are better than other structures. It is interesting that HIP1, HIP2 and HIP5 have heat integration at reboiler DIB column, the proper control structures are CS8 and CS4 because reboil stream is divided to increase temperature with heat exchanger so the base temperature control is supportable to reject disturbance.

We can conclude that the plants with higher level of heat integration can have control performance as good as the base case plant with lower heat integration if the suitable control structures can be found. More importantly, we can avoid the control difficulties associated with the heat integration by choosing the suitable heat integrated structure and proper control structures. This can be done in real time domain using rigorous process dynamic simulator like HYSYS.

6.2 Recommendations

1. Study and design the control structure of complex heat-exchanger networks of the other process in plantwide control point of view.

2. Study the controllability characteristics of energy-integrated Butane Isomerization plant.

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APPENDICES

APPENDIX A

Table A.1 Data of Butane Isomerization process (Base Case) for simulation

Name	Fresh feed	Product	Purge	Reactor inlet	Reactor outlet
Temperature [F]	90	124	178	390	422
Pressure [psia]	300	100	66	665	650
Molar Flow [lbmole/hr]	580	550	30	870	870
Mole Frac (C_3)	0.02	0.02	0	0	0
Mole Frac (iC_4)	0.24	0.96	0	0.17	0.62
Mole Frac (nC_4)	0.69	0.02	0.01	0.81	0.36
Mole Frac (iC_5)	0.05	0	0.99	0.02	0.02

Table A.2 Equipment data

Butane isomerization process (Base Case)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE	UA (Btu/F-hr)	3.96E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-1)

Kinetics	A_f [1/h]	4x10^{8}
	E_f [Btu/lb-mol]	3x10^{4}
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	4.08E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21
FEHE 2	UA (Btu/F-hr)	1.14E+05
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-2)

Kinetics	A_f [1/h]	4x10^{8}
	E_f [Btu/lb-mol]	3x10^{4}
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	2.18E+05
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21
FEHE 2	UA (Btu/F-hr)	3.20E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-3)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	4.08E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21
FEHE 2	UA (Btu/F-hr)	1.18E+05
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-4)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	2.18E+05
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21
FEHE 2	UA (Btu/F-hr)	3.87E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-5)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	2.86E+05
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-6)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	1.80E+06
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Butane isomerization process (RHEN-7)

Kinetics	A_f [1/h]	$4x10^8$
	E_f [Btu/lb-mol]	$3x10^4$
Fooded Condensor	Holdup[ft^3]	340
Reactor	ID [ft]	7
	Length [ft]	18
	Holdup[ft^3]]	693
FEHE 1	UA (Btu/F-hr)	7.87E+04
	Shell holdup [ft^3]	21
	Tube hold[ft^3]	21

Table A.3 Column specifications

Column specifications	DIB column	Purge column
Total trays	50	20
Feed tray	20/30	10
Diameter (ft)	16	6
Refulx drum holdup(ft^3)	1700	370
Base holdup (ft^3)	2000	400
Reflux ratio	7.30	0.80
Specification 1 fraction in overhead	nC_4 = 0.0200	iC_5 = 0.0030
Specification 2 fraction in bottom	iC_4 = 0.2548	nC_4 = 0.0295

APPENDIX B

PARAMETER TUNEING OF CONTROL STRUCTURES

B.1 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 tuned with a small integral or reset time constant τ_I. A value of τ_I = 0.3 minutes works in most flow controllers. The value of controller gain should be kept modest because flow measurement signals are sometime noisy due to the turbulent flow through the orifice plate. A value of controller gain of Kc = 0.5 is often used. Derivative action should not be used.

In a real plant application, filtering of the flow signal is also recommended because of the noise. So filter is put at controller output signal and a good number to use for a flow loop is τ_F = 0.1 minute.

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 Kc = 2.

Pressure Controllers

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. Of course the gain used depends on the span of the pressure transmitter. Some simple step tests can be used to find the value of controller gain that yields satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are Kc = 2 and τ_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 and 10, the integral time, τ_I, should set between 2 and 10 minutes, and the derivative time $\tau_{d,}$ should be set between 0 and 5 minutes.

B.2 Relay-Feedback Testing

If we have a controller that needs tuning and after we have inserted reasonable lags and deadtimes, we need a quick and simple method for identifying the dynamic parameters that are important for designing a feedback controller. The relay-feedback test is a tool that serves this purpose well. 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 high of the relay h. This height is typically 5 to 10% of the controller-output scale. The loop starts to oscillate around the setpoint, with the controller output switching every time the process variable (PV) signal crosses the setpoint.

The maximum amplitude of the PV signal is used to calculate the ultimate gain K_u from the equation:

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

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 or PID controllers, using a variety of tuning methods proposed in the literature that require only the ultimate gain and ultimate frequency, e.g., Ziegler-Nichols, Tyreus-Luyben, etc.

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 closed loop, 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 (the ultimate frequency).
5. The impact of load changes that occur during the test can be detected by a change to asymmetric in the manipulated variable.

All these features make relay-feedback testing a useful identification tool.

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

$$Kc = \frac{Ku}{2.2}$$

$$\tau_I = \frac{Pu}{1.2}$$

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

$$Kc = \frac{Ku}{3.2}$$

$$\tau_I = 2.2Pu$$

Relay feedback testing can be done in HYSYS. It simply click the Tuning botton on the controller faceplate, select Autotuning and click the Start Autotuning botton. The loop will start to oscillate. After several cycles, the tuning is stopped and some ewcommended settings for a PID controller are suggested.

B.3 Inclusion of Lag

These are typically temperature and composition controllers. These loops have significant dynamic lags/or deadtimes. Realistic dynamic simulations require that we explicitly include lags and/or deadtimes in all the important loops. Usually this means controllers that affect product quality (temperature or composition) or process constraint (safety, environmental, etc.).

In this plant, a 3-minute deadtime is assumed in the product composition measurement (CC1). We include lags at temperature control loops of TC1,TC2, TCR and TC-cool. Some lags are recommended in table below:

Table B.1 Typical measurement lags

		Number	Time constant (minutes)	Type
Temperature	Liquid	2	0.5	First-Order Lag
	Gas	3	1	First-Order Lag
Composition	Chromatograph	1	3 to 10	Dead time

Table B.2 Parameter tuning of Butane Isomerization process

Controller	controlled variable	manipulated variable		Control action	Set point		K_C	τ_i (min)	τ_d (min)
CC1	product composition	CS1, CS5	DIB reboiler duty (qr1)	Reverse	0.02	-	1	60	-
		CS2, CS6	DIB reboiler duty (qr1)	Reverse	0.02		1	60	-
	DIB column top stage composition	CS3, CS7	DIB product valve (V3)	Reverse	0.259	-	23	25	0.1
		CS4, CS8	DIB column reflux valve (V1-reflux)	Direct	0.259	-	5.7	28	0.1
TCR	reactor inlet temperature	furnace duty (qfur)		Reverse	390.1	^oF	1.5	0.4	0.09
TC-cooler	cooler outlet temperature	cooler duty (qcooler)		Direct	125	^oF	5	0.4	0.095
LC11	DIB column base level	PFD feed valve (V1)		Direct	50.00	%	2	2	-
LC12	DIB column reflux drum level	CS1, CS5	DIB product valve (V3)	Reverse	50.00	%	2	-	-
		CS2, CS6	DIB product valve (V3)						
		CS3, CS7	DIB column reflux valve (V1-reflux)						
		CS4, CS8	DIB product valve (V3)						
LC21	Purge column reflux drum level	CS1,CS2, CS7, CS8	Purge reboiler duty (qr2)	Reverse	50.00	%	2	-	-
		CS3 ,CS4, CS5, CS6	TC2				0.36	10.11	-
LC22	Purge column base level	Purge feed valve (V2)		Reverse	50.00	%	2	2	-
PC1	DIB column pressure	DIB condenser duty (qc1)		Direct	100.0	psia	2	2	2
PC2	Purge column pressure	Purge condenser duty (qc2)		Direct	65.0	psia	2	2	2
PC3	FEHE1 hot stream outlet pressure	Rliquid feed valve (V6)		Direct	671.0	psia	2	2	10
TC1	DIB column stage1 temperature	DIB reboiler duty (qr1)		Reverse	142.7	^oF	14.5	0.9	0.196
TC2	Purge column average temperature	Purge reboiler duty (qr2)		Reverse	149.9	^oF	7.74	3.0	0.676

APPENDIX C

DYNAMIC RESPONSES

The dynamic responses and table of the IAE results of the control structures in the Butane Isomerization plant for Base Case and RHEN1-7 when changes in the reactor inlet and the recycle flowrates by decreasing are shown in this section.

These disturbances are made as fowrow:

- Step change of -10^{o}F decreasing at time 10 minute are made in the reactor inlet stream temperature.
- Step change of decreasing the recycle flowrates before entering the DIB column from 870.1 to 800.1 lb-mol/hr.

The disturbance testing is used to compare the dynamic response of heat integration processes (Base case, RHEN1-7) with control structures (CS1-8). Because the responses of RHEN1-7 trend to similar Base case, the dynamic response of Base case is shown in Figure C.1 and C.2 when changes in the reactor inlet and recycle flowrate by decreasing.

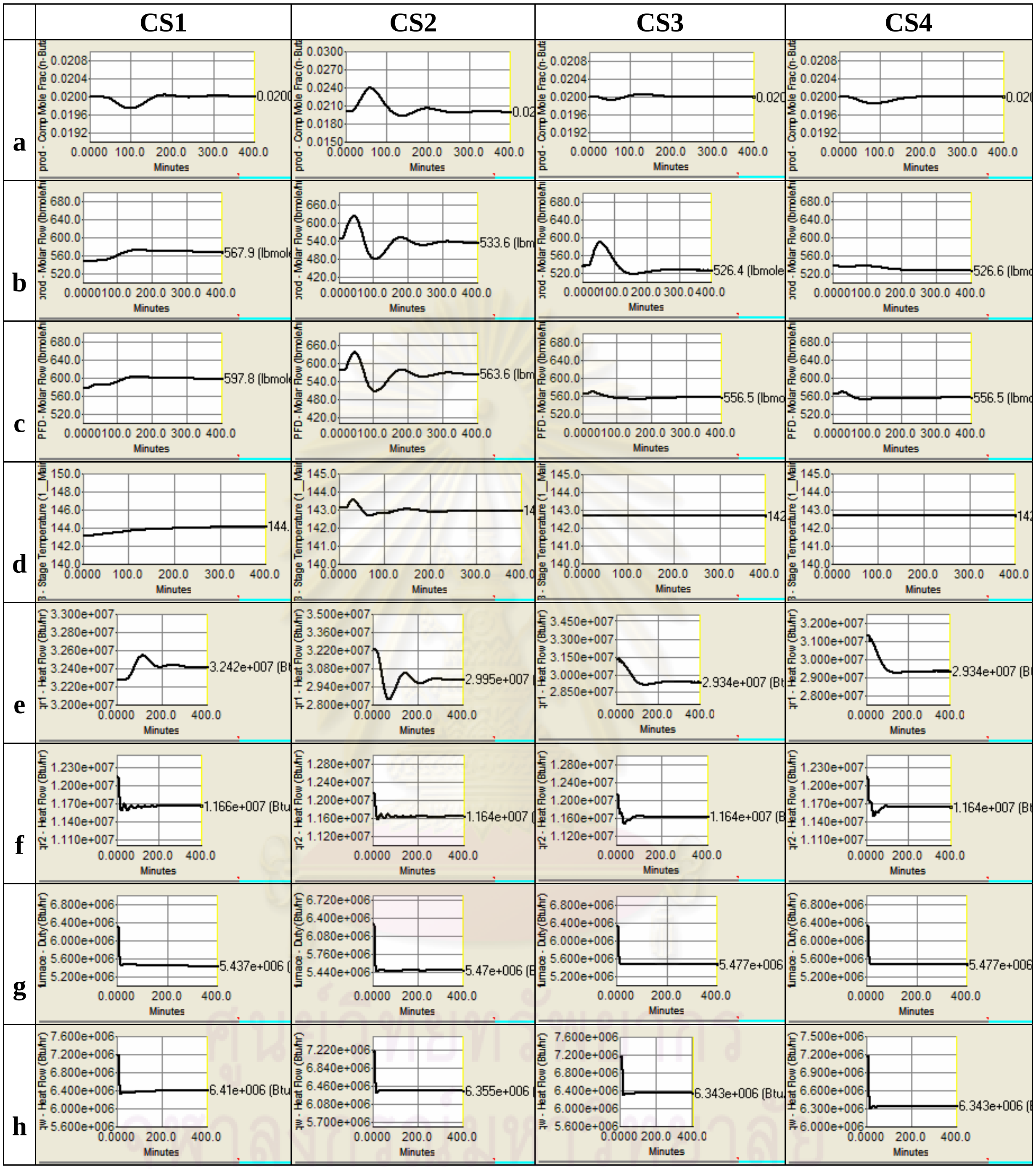

Figure C.1 Dynamic responses to increase in recycle flow from 870 to 940.1lb-mol/h., where (a) product composition, (b) product flowrate, (c) fresh feed flowrate, (d) tray 1 temperature of DIB column, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

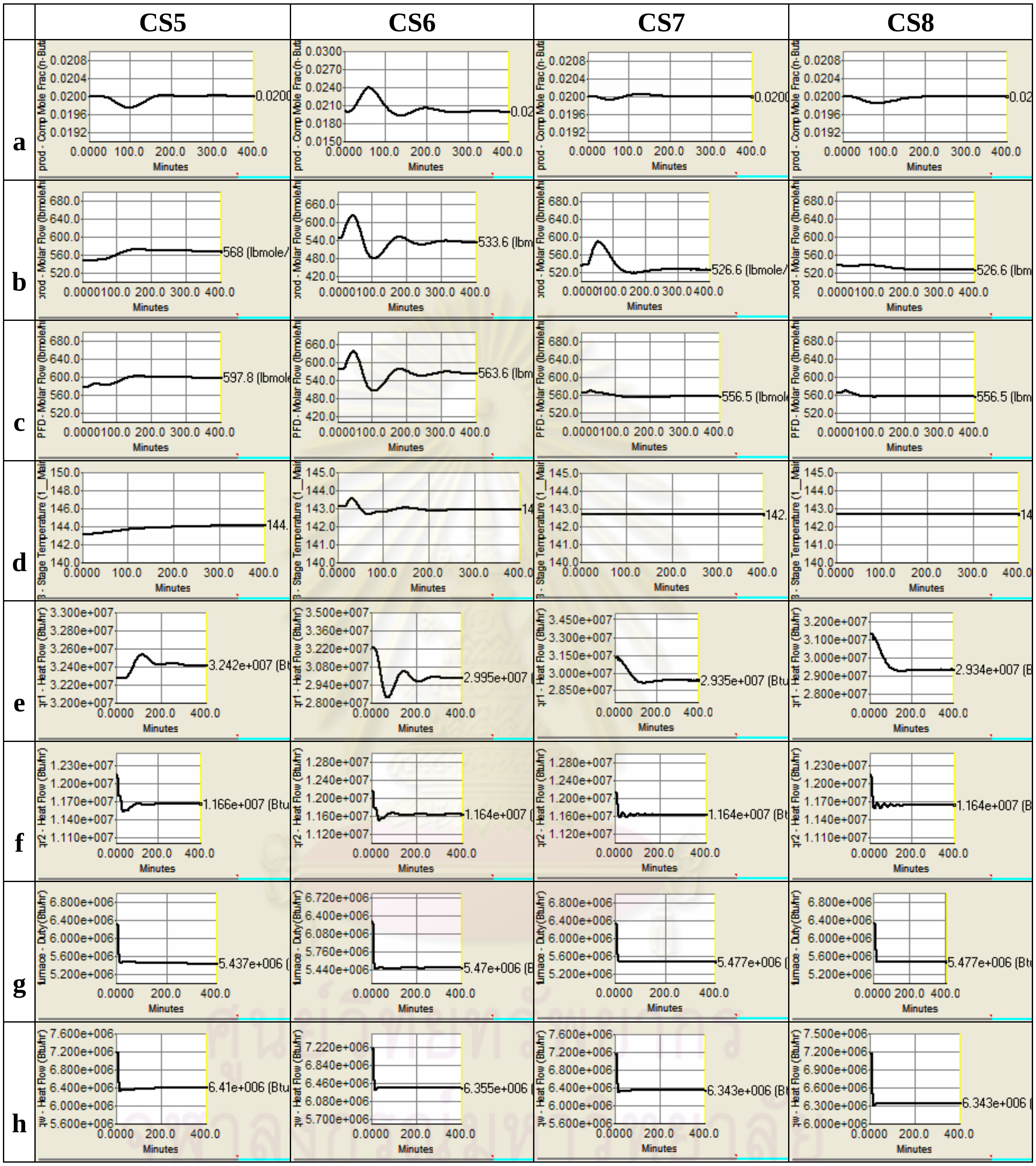

Figure C.1 Continued Dynamic responses to increase in recycle flow from 870 to 940.1lb-mol/h., where (a) product composition, (b) product flowrate, (c) fresh feed flowrate, (d) tray 1 temperature of DIB column, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

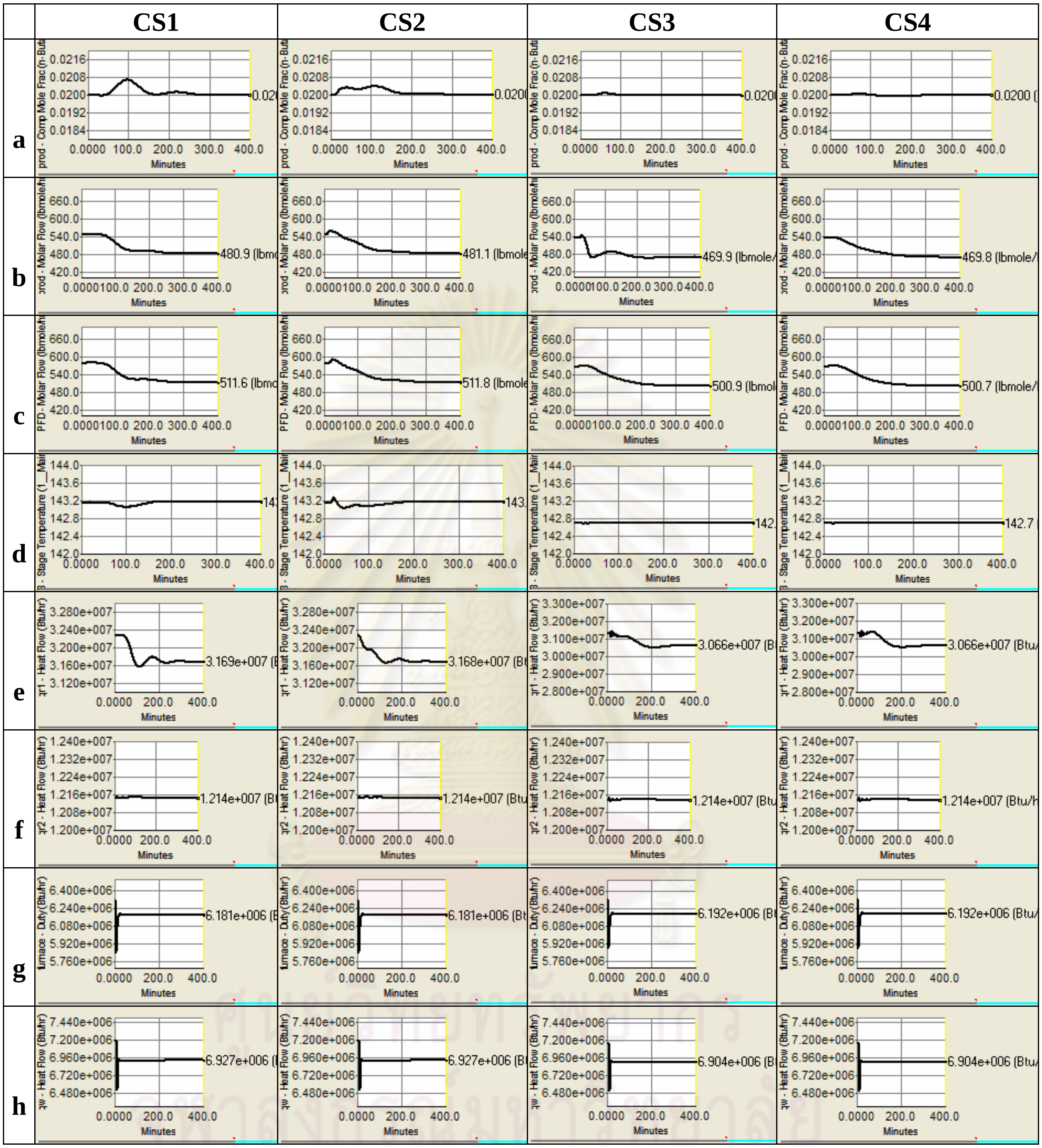

Figure C.2 Dynamic responses to 10 °F increase in reactor inlet temperature, where (a) product composition, (b) product flowrate, (c) fresh feed flowrate, (d) tray 1 temperature of DIB column, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

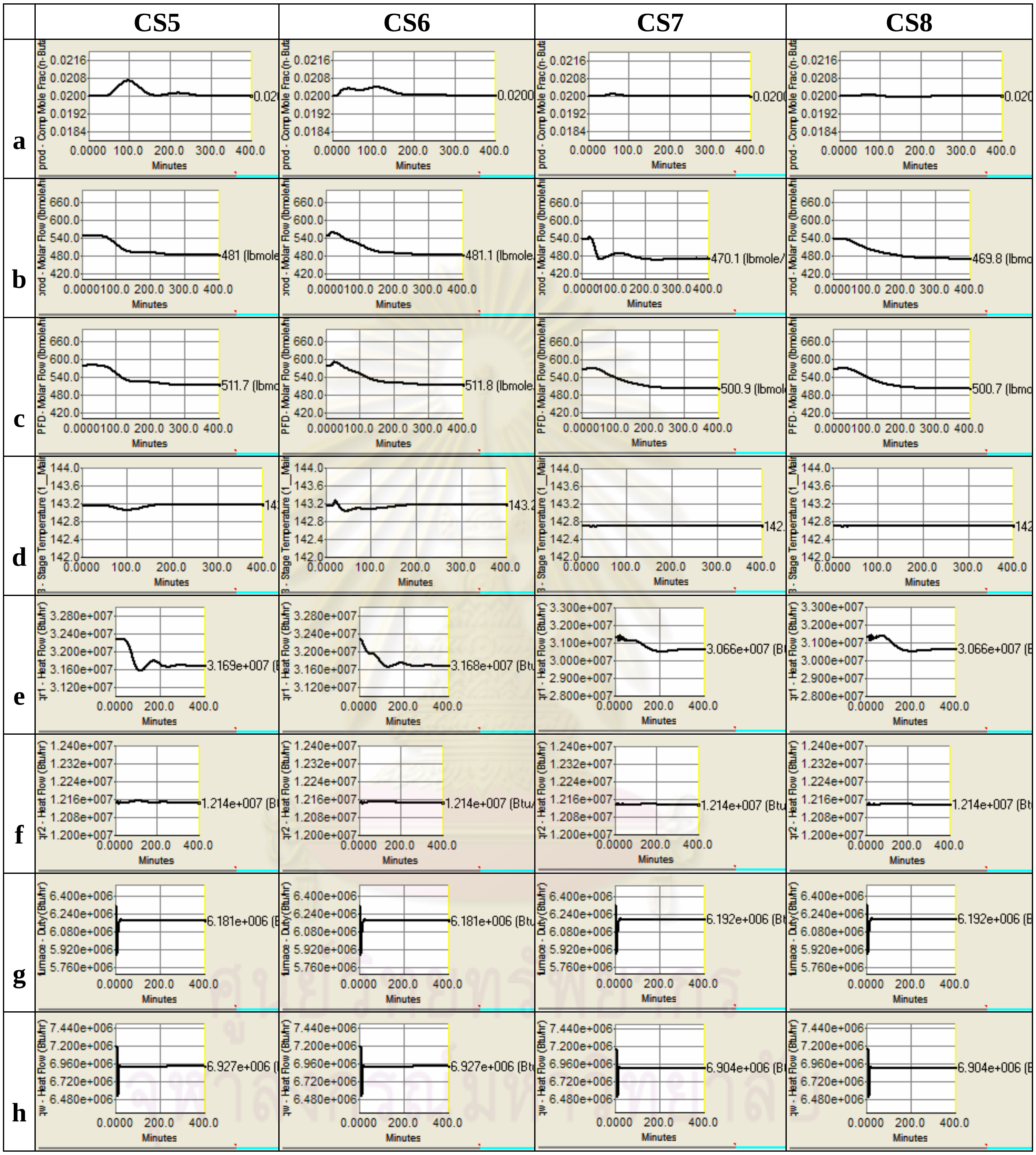

Figure C.2 Continued Dynamic responses to 10 °F increase in reactor inlet temperature, where (a) product composition, (b) product flowrate, (c) fresh feed flowrate, (d) tray 1 temperature of DIB column, (e) reboiler duty of DIB column, (f) reboiler duty of Purge column, (g) furnace duty and (h) cooler duty.

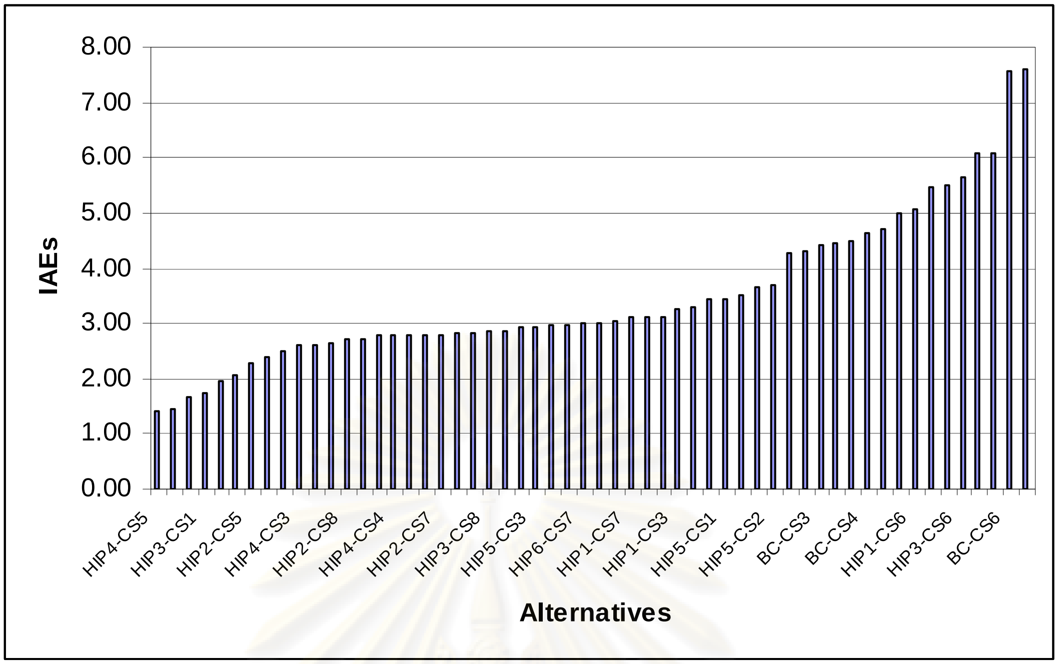

Figure C.3 The IAE results of the control systems to decrease in recycle flowrates from 870.1 to 800.1lb-mol/h.

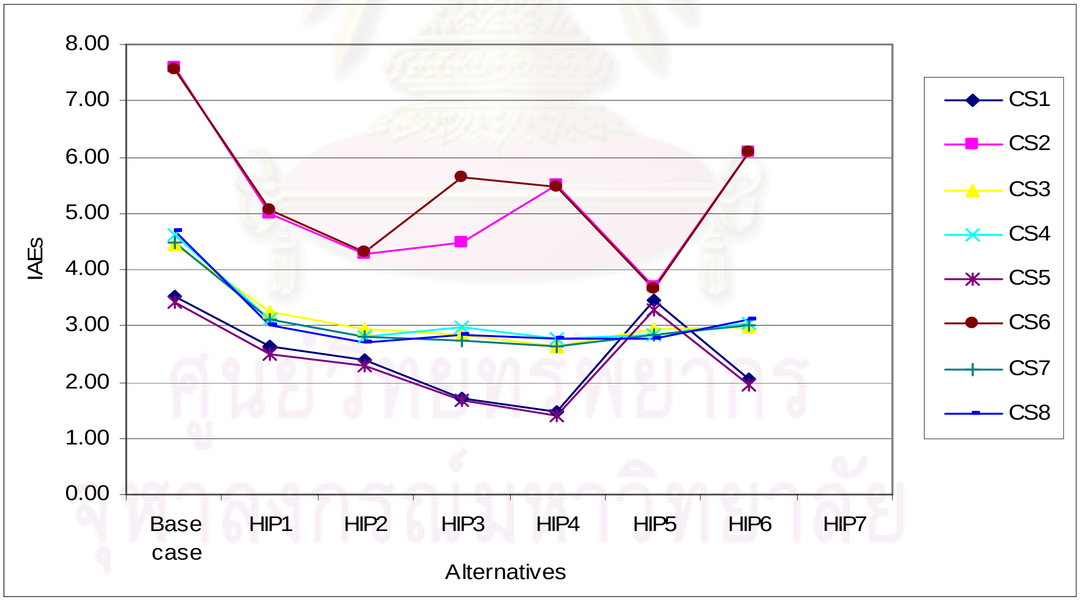

Figure C.4 The IAE results of the control systems to decrease in recycle flowrates from 870.1 to 800.1lb-mol/h.

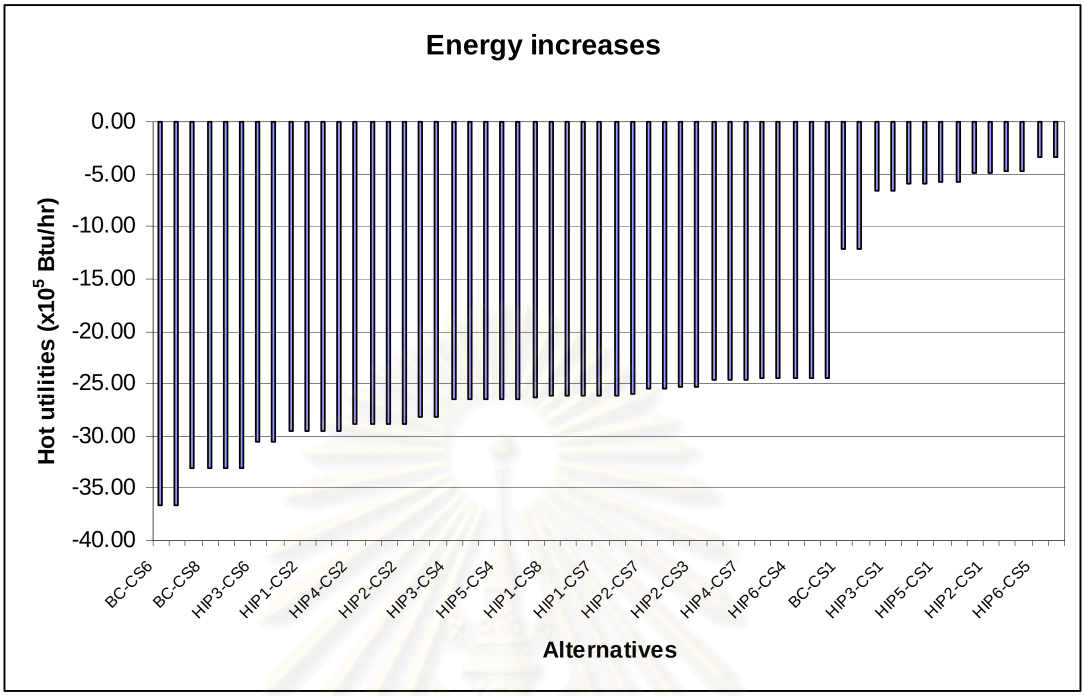

Figure C.5 Energy increases as the recycle flowrates increase from 870 to 940.1lb-mol/h.

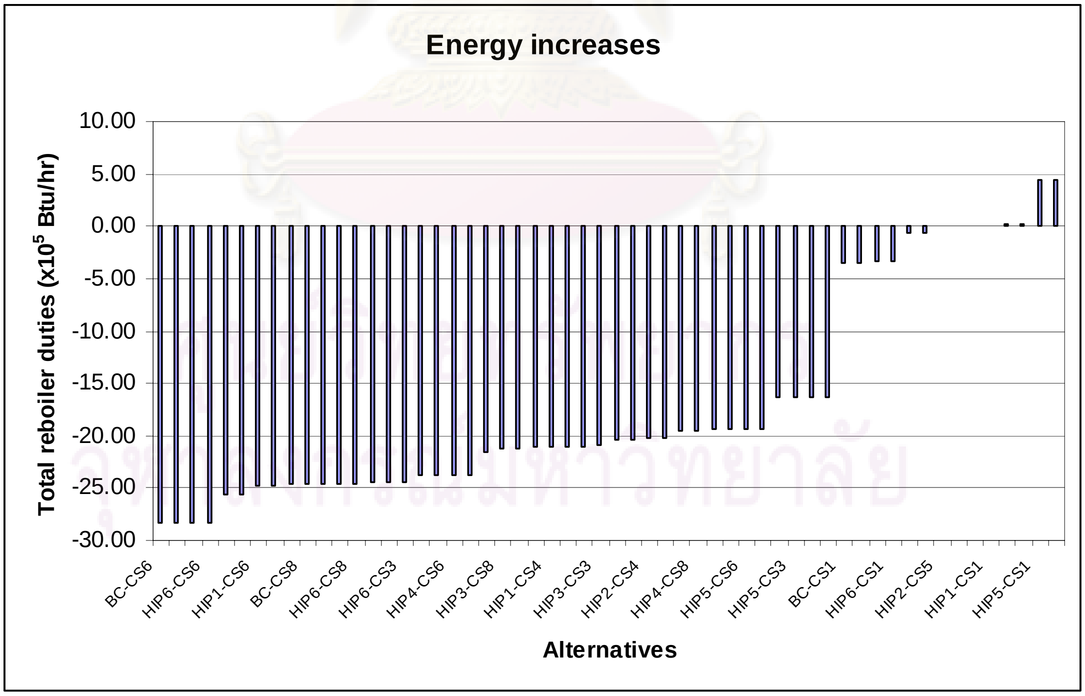

Figure C.6 Reboiler duty increases as the recycle flowrates increase from 870 to 940.1 lb-mol/h.

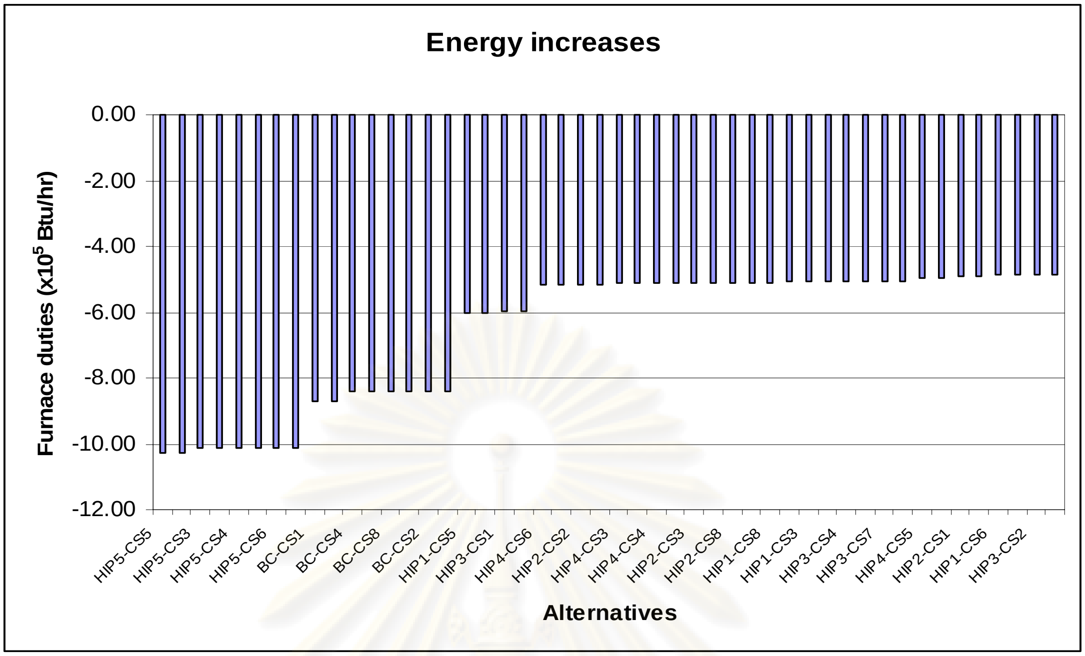

Figure C.7 Furnace duty increases as the recycle flowrates increase from 870 to 940.1 lb-mol.

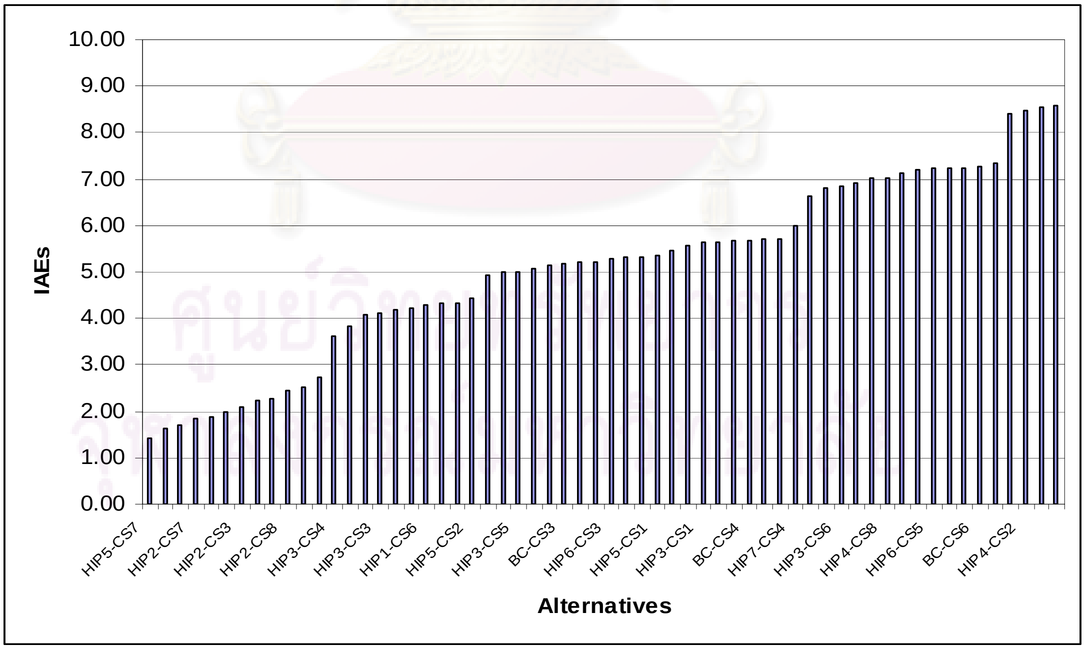

Figure C.8 The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

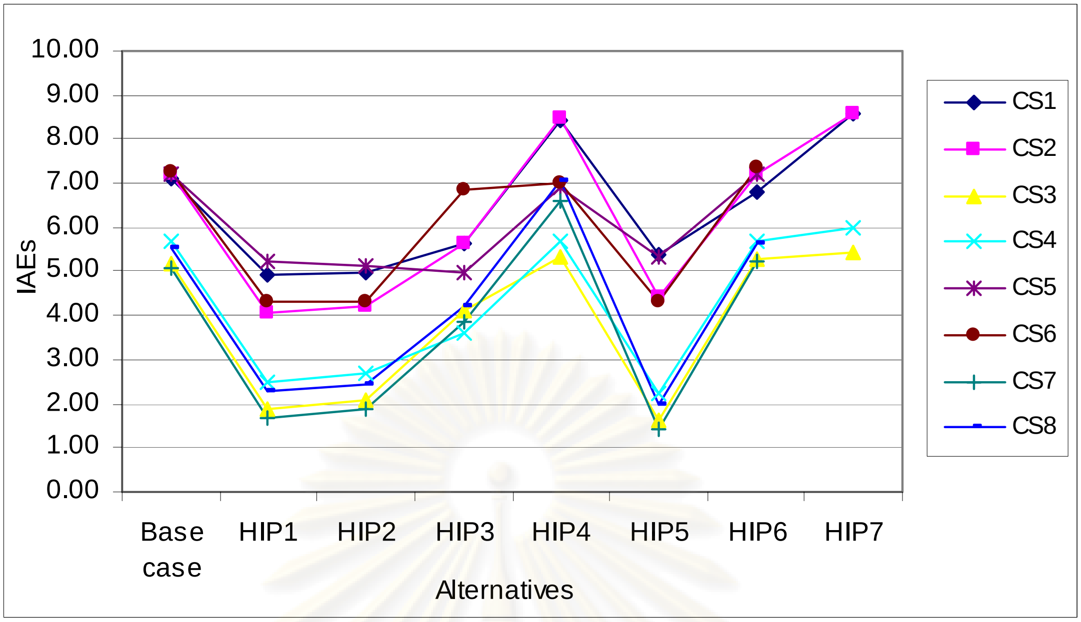

Figure C.9 The IAE results of the control systems to 10 °F increase in the setpoint of the reactor inlet temperature

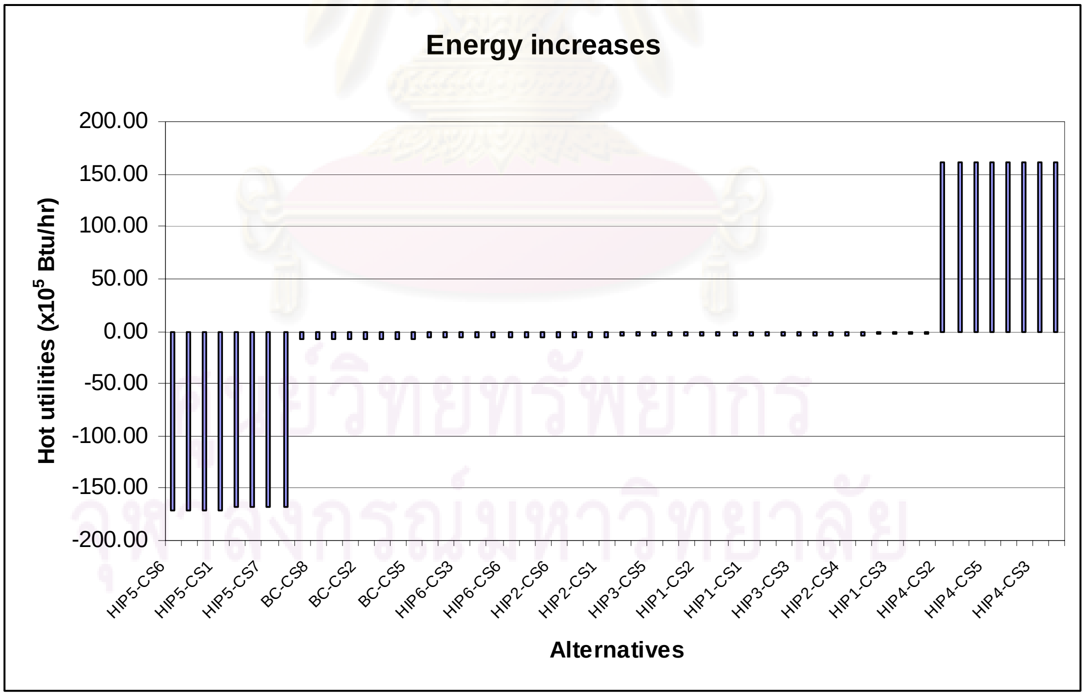

Figure C.10 Energy increases as the temperature of reactor feed increase 10 °F

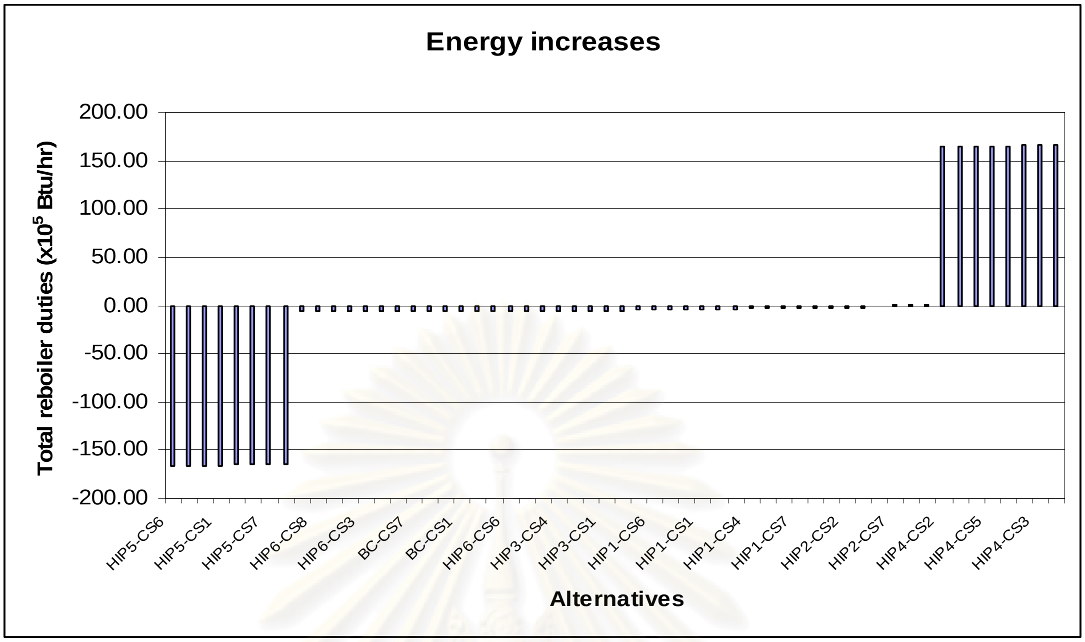

Figure C.11 Reboiler duty increases as the temperature of reactor feed increase 10 °F

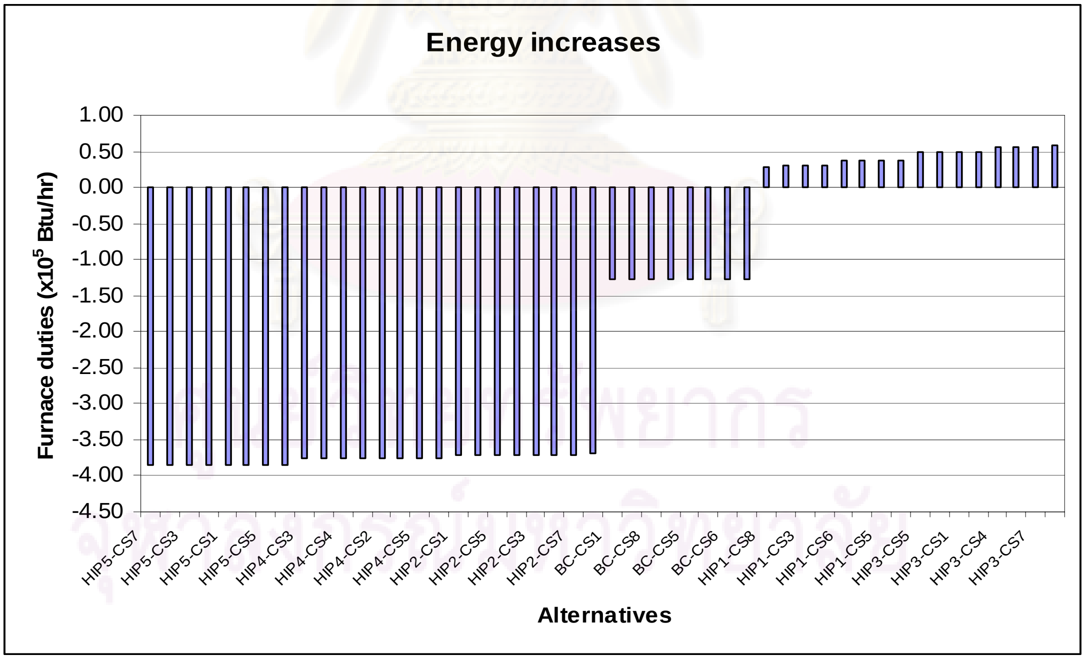

Figure C.12 Furnace duty increases as the temperature of reactor feed increase 10 °F

VITA

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